Theory, Design and Operation of Nutrient Removal Activated Sludge Processes

University of Cape Town, City Council of Johannesburg and CSIR

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THEORY, DESIGN AND OPERATION OF NUTRIENT REMOVAL ACTIVATED SLUDGE PROCESSES

A collaborative information document prepared for the

WATER RESEARCH COMMISSION

by the

UNIVERSITY OF CAPE TOWN,

CITY COUNCIL OF JOHANNESBURG and the

NATIONAL INSTITUTE FOR WATER RESEARCH OF THE CSIR

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PRETORIA, 1984

FOREWORD

The Republic of South Africa is a water-scarce country as poignantly accentuated by the recent (1983) drought. It is therefore obvious that everything possible must be done to safeguard our water resources against pollution as well as to ensure their optimal use. Wastewater treatment has an important role to play in both these objectives.

The discharge of treated effluents back to the river of origin as required by the Water Act, No 54 of 1956, has necessitated a high standard of wastewater treatment. Although this requirement has been in effect since 1962 when the General and Special Standards were promulgated, the General Standard did not require that phosphates and nitrogen be removed. As a consequence our water bodies have become enriched with these plant nutrients, and this has resulted in the proliferation of water plants and algae which in turn has given rise to problems with water purification processes, as well as aesthetic and suspected health problems. In an effort to control eutrophication the authorities have promulgated an effluent standard for dissolved ortho-phosphate of 1 mg/ℓ (as P) which is to be enforced in certain critical catchments as from August 1985.

In order to meet the challenge of producing effluents which comply with the phosphate standard as cost-effectively as possible, as well as producing effluents which are eminently suitable for re-use, the Water Research Commission initiated research on nutrient removal from wastewaters in 1974. After nearly ten years' research and development work, significant progress has been made in this field. Results and findings of the research were progressively published via a number of routes, i.e. technology transfer seminars, workshops and scientific publications. To assist the end user to effectively remove nutrients from effluents it was decided to publish a comprehensive information publication directed specifically at user organisations such as local authorities and consulting engineering firms. This publication presents practical guidelines for the planning, design and operation of biological nutrient removal plants. Possibly the best advertisement for the processes described in this publication is Johannesburg's 150 Mℓ/d Goudkoppies sewage treatment works where the biological nutrient removal process is producing an effluent which consistently meets the General and Phosphate Standards.

I recommend this publication to you as a basis for consideration for the design and operation of new plants as well as the modification and operation of existing plants to meet the General and Phosphate Standards. I further extend an invitation to you to consult with experts in this field in the RSA on any site specific problems you may be encountering in process design and operation to meet effluent standards.

M R HENZEN
CHAIRMAN: WATER RESEARCH COMMISSION

ACKNOWLEDGEMENTS

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ABSTRACT

Two summaries of case studies of biological nutrient removal are presented, i.e. full-scale experience at Johannesburg's Goudkoppies and Northern Works and pilot scale experience at the National Institute for Water Research's Daspoort experimental pilot plant. Reference is made to reports in which detailed information on the abovementioned case studies is given, as well as to other case studies which have been reported in the literature.

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INTRODUCTION

The Republic of South Africa is rapidly approaching the point of maximum economic exploitation of conventional water resources. In addition, the country is faced with deteriorating water quality which may become a limiting factor in water resources development prior to limitations on water quantity.

The accelerated deterioration of water quality is a result of, amongst others, the discharge of ever-increasing quantities of treated effluents to rivers and streams. Effective removal of pollutants from wastewaters, particularly the nutrients carbon, nitrogen and phosphorus, is therefore of the utmost importance. Furthermore, high quality effluents may serve as an economic source to augment dwindling water resources.

The need for pollution control and indirect reuse of treated wastewater is legally expressed in the Water Act (54 of 1956) and standards promulgated in terms of the Act. The General and Special Standards which were promulgated in 1962, emphasized the adequate removal of carbonaceous material and the transformation of ammonia to nitrate. More recently accelerated eutrophication of many of South Africa's impoundments, with attendant health, water quality and aesthetic problems, has become apparent and resulted in the promulgation in 1980 of an upper limit for effluent phosphates for a number of critical catchments. Furthermore, limiting the total dissolved solids content of treated effluents is now also considered of prime importance. The need to remove nutrients and reduce the increase in dissolved solids during treatment, places difficult demands on future wastewater treatment plants.

Technology for the removal of the nutrients nitrogen and phosphorus from wastewater has been developed to a high level in the Republic of South Africa. Many organizations have been instrumental in this development, including research organizations, universities, consulting engineering firms and local authorities. A wide range of physical, chemical and biological processes and combinations of these are now available both for upgrading of existing works and design of new works. Methods available are: *Physical-chemical*, of particular use for removal of phosphates from the effluents of existing works – the only viable option, in fact, for dealing with biological filter effluents*; *physical-chemical-biological***, for the removal of nitrogen and phosphates, of use where the characteristics of the wastewater are unfavourable for the biological removal of the nutrients; *biological*, for the removal of all or major fractions of the nitrogen and phosphates in the influent, no or only minimal addition of chemicals is needed to produce the required effluent phosphate quality. Biological nutrient removal is particularly attractive for the treatment of municipal wastewaters because treatment costs are generally less than with physical-chemical methods, the salt concentration of the effluent is not raised as with chemical addition and the characteristics of the wastewater usually are amenable to this type of treatment. In consequence, the Water Research Commission has been greatly interested in developing this method of nutrient removal.

The Water Research Commission has stimulated, co-ordinated and financed research and development work in the field of biological nutrient removal in the activated sludge process from about 1973. Since that time it has contracted the National Institute for Water Research of the Council for Scientific and Industrial Research, the Universities of Cape Town and Pretoria and the City Council of Johannesburg to do research and development work in this field on a co-ordinated joint venture basis. A significant amount of information on processes for biological nutrient removal is now available. Much of this information has already been published in local and international journals as well as having been presented at a number of conferences, seminars and open days. However, in the light of the urgent need of local authorities and others who have to meet ever stricter effluent standards, and in particular the effluent phosphate standard of 1 mg/ℓ soluble orthophosphate which is to be strictly enforced in a number of critical catchments from August 1985, it was decided to compile a comprehensive information document on the biological removal of nitrogen and phosphates covering relevant results and findings made under Water Research Commission sponsored projects.

This publication has been compiled as a self-contained document which does not require reference to other publications on wastewater treatment. It is intended primarily for the design engineer and management staff responsible for operation and control of wastewater purification works. The level of presentation assumes that the reader has had tertiary training and/or considerable practical experience in the field of wastewater treatment.

The basic process for the simultaneous biological removal of phosphates and nitrogen was proposed by Barnard in 1976 and is called the Phoredox process in South Africa, and the Bardenpho or Modified Bardenpho in the United States. This process belongs to the single sludge, multi-reactor group of processes. For the phosphate removal aspect the fundamental principle embodied in this process is that an anaerobic state needs to be created at some point in the process in such a way that phosphate is released, a consequence of which is that biological uptake of phosphate in excess of normal metabolic requirements is induced when the sludge is aerated subsequently. In the Phoredox process the endeavour is to create the requisite anaerobic state by mixing the influent waste stream with the

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*Phosphate removal by chemical addition: the reader is referred to the Water Research Commission publication "Guidelines for chemical removal of phosphates from municipal effluents."

**For example, the LFB process, the Lime-Flotation-Biological process developed by the National Institute for Water Research, Pretoria, RSA.

sludge recycled from the secondary settling tank without aeration in an anaerobic tank at the head of the works.

This method for promoting the anaerobic state appears in a number of processes that developed from the Phoredox process in order to accommodate wastewaters with unfavourable characteristics or in which nitrification-denitrification is not required. One such development, known as the UCT process, is described in detail in this publication. Some of the other processes which also incorporate the Phoredox principle, are not dealt with, for example, the AO process is not considered since this process is designed to remove phosphate in short sludge age systems where nitrification is prevented from occurring, and is unlikely to find significant application in the RSA. The process developed by Roberts and Kerdachi at Pinetown, Natal, also is not dealt with. This process differs from the others in that nitrification, denitrification and the anaerobic state, to induce P release, all take place in one reactor. Also, the process is oxygen-limited, in consequence, its biological response with regard to sludge production, oxygen demand, nitrification, etc. is not amenable to description by the kinetic models governing non-oxygen-limited processes such as the multireactor Phoredox, UCT and AO processes. The writers of this manuscript felt therefore that the Roberts-Kerdachi process falls outside their ambit of competence.

All the above processes are of the main-stream kind. Side-stream processes, of which the Phostrip process is currently the most widely used, are not discussed in this publication. In side-stream processes the anaerobic state is created by passing a fraction of the secondary settler underflow through an anaerobic reactor to release phosphates through endogenous respiration (instead of by the addition of feed waste flow) and precipitating the released phosphates by lime addition. Insufficient data on the behaviour of this process was available to the authors to allow directives to be put forward for design and operation of this process for producing effluents complying with South African effluent regulations with regard to both nitrogen and phosphates.

Guidelines for design set out in this publication are based primarily on about ten years of extensive laboratory investigations at the University of Cape Town on biological nutrient removal from predominantly domestic wastewaters. These guidelines were verified on data generated at pilot scale on the National Institute for Water Research's facility at Daspoort, Pretoria, and at full-scale on the Johannesburg Goudkoppies and Northern Works. The mathematical model for carbonaceous material oxidation, nitrification and denitrification have been shown to give accurate predictions of full scale plant response. The theory for biological phosphate removal is currently still in a development stage with research in this area continuing. Nontheless, tentative but conservative guidelines for process selection and design of biological phosphate removal aspects are presented.

Biological nutrient removal in the activated sludge process has only a short history of application. Currently in South Africa there are of the order of thirty plants designed for and/or being operated to obtain biological nutrient removal, with varying degrees of success. Practical experience on these plants should increase knowledge on the application of the process to a variety of wastewaters under different operating conditions and effluent requirements. Further research is still needed and is continuing in a number of areas where knowledge is still inadequate, for example, the mechanism(s) and factors controlling excess phosphate removal, the role of the anaerobic stage in the process, settling characteristics of the sludge formed in the process and factors which control these, as well as scum formation. This publication therefore does not pretend to be the final word on biological nutrient removal; it is an information document with interim guidelines for design and operation of biological nutrient removal plants.

H N S WIECHERS
WATER RESEARCH COMMISSION
PRETORIA, 1984

CHAPTER ONE

FUNDAMENTALS OF BIOLOGICAL BEHAVIOUR

G.v.R. Marais and G.A. Ekama

1. INTRODUCTION

Life, in order to persist, requires a continuous throughput of (1) matter and (2) energy:

(1) The most prominent elements passing through the life system are hydrogen (H), oxygen (O), carbon (C), nitrogen (N), phosphorus (P) and sulphur (S), plus a host of minor elements.

(2) Energy is derived principally from three sources and these form a convenient criterion to categorise the organisms implicated:

(a) Solar radiation: photo-synthetic autotrophs

(b) Organic compounds: heterotrophs

(c) Inorganic compounds: chemical autotrophs.

(a) In the biosphere the fundamental source of energy is solar radiation. The photo-synthetic autotrophic cells fix a small fraction of the solar energy by forming complex high energy organic (H, C, O) compounds and producing oxygen. Of the matter requirements the elements H and O are obtained from H_2O; carbon from CO_2; and phosphorus, nitrogen, sulphur and the minor elements from the dissolved salts of these elements. Both H_2O and CO_2 are readily available; sulphur is usually in adequate supply; however the availability of phosphorus and nitrogen usually is limited – phosphorus does not occur readily in soluble form and nitrogen is useful to the organism principally in the NH_4^+ and NO_3^- forms. The restricted availability of these two elements usually constitutes the limiting factor to the mass of autotrophic life a body of water can generate. For this reason, phosphorus and nitrogen are termed eutrophic (life-giving) substances. Phosphorus is limited by the mass available or entering a particular ecosystem, but ammonia can be generated from dissolved molecular nitrogen by certain micro-organisms. In this respect, the control of phosphorus in the body of water takes on a greater significance than the control of nitrogen.

Whereas the eutrophic elements N and P normally are the principal agents of pollution in bodies of water, by stimulating autotrophic growth, the action associated with this growth, of producing oxygen, is utilized in the oxidation pond system of wastewater treatment – the oxygen is utilized by heterotrophic organisms to metabolize the carbonaceous material in the influent to the pond, in this fashion to assist in destroying organic energy, as described in (b) below.

(b) The high energy organic compounds synthesized by the autotrophs form the basic source of energy for the heterotrophic cells to synthesize the more complex molecules that constitute the cell mass, including proteins. Only a fraction of the energy utilized is incorporated in the cell mass generated, the balance is lost as heat. The mass of organisms thus generated in turn form the matter and energy sources (prey) for other organisms that live on them (predators), and these in turn become prey to yet other predators. Each prey-predator transformation is accompanied by a substantial loss of energy; in this fashion through the sequential chain of life there is a continuous reduction of the organically bound energy originally fixed by the photo-synthetic autotrophs. When the organic energy reduces to zero heterotrophic life ceases. In wastewater treatment plants the full chain of life does not develop so that a substantial fraction of the input energy is retained in the life mass in the system but this fraction is physically separated from the liquid in the sludge wasted daily (for further treatment) leaving very little organic energy in the effluent.

Presence of organic energy in the receiving body of water gives rise to fungal and other heterotrophic growths, deoxygenation of the water (due to the metabolic activities of these growths) thereby rendering the water unsuitable for higher life forms such as fish, anaerobic conditions causing fermentation and redissolution of heavy metal salts, and so on. These effects reduce the aesthetic appearance, recreational use and the re-use of the water.

(c) Chemical autotrophs derive their energy for growth from the oxidation of inorganic compounds. In wastewater microbiology the principal species of interest in this genera are the nitrifying bacteria. These bacteria are obligate aerobes (i.e. can grow only when oxygen is present), and derive energy by oxidizing saline ammonia to nitrite and nitrate. In wastewater treatment the nitrifiers form a vital link in the removal of the autotrophic element nitrogen from the water by converting ammonia nitrogen to nitrate nitrogen when, by appropriate design, the heterotrophs can be forced to utilize the nitrate in the heterotrophic metabolism of carbonaceous material, in which action the nitrate is reduced to molecular nitrogen and escapes as nitrogen gas. (see later).

1.1 Objectives of wastewater treatment

From the considerations above the objectives of wastewater treatment are seen to be twofold:

1. Reduction of organic-bound energy to a level such that the heterotrophic growth and associated deoxygenation effects in the receiving body of water are acceptably low.
2. Reduction of the autotrophic substances, phosphates, ammonia and nitrates, to levels such that photo-synthetic autotrophic growths and their capacity to fix solar energy as organic energy in the receiving body of water are acceptably low.

2. BIOLOGICAL BEHAVIOUR

Biological metabolism is principally a process of energy conversion. In biological wastewater treatment processes, the input energies are present in the wastewater in the form of carbonaceous and nitrogenous organic compounds. These are converted into other energy forms during which some of the energy is lost as heat.

The carbonaceous energy is utilized by the heterotrophic group of organisms. This is a wide spectrum group, which, given sufficient time and appropriate environmental conditions, will utilize every type of organic material. The group is ubiquitous and in any given situation those members of the group that obtain maximum benefit from the specific organic material and environmental conditions will develop. As the organic source of the environmental conditions change, so associated changes in the heterotrophic organism species take place.

The nitrogenous material serves as prime energy source in the free and saline ammonia form. Proteinaceous nitrogen is broken down into its ammoniacal and carbonaceous constituents by heterotrophs. The free and saline ammonia is utilized as an energy source by two specific organisms species, the *Nitrosomonas* and *Nitrobacter*. The former converts ammonia to nitrite and the latter nitrite to nitrate, these two organism species jointly being called the nitrifiers. Compared to most of the members of the heterotrophic group, the nitrifiers are slow growing. Because the group consists of only two species, each performing a particular function in the nitrification process, the environmental conditions must be conducive to both species, otherwise one or the other will not propagate and nitrification will be adversely affected. In general the environmental conditions for the nitrifiers are much more circumscribed than those for the heterotrophs. With the heterotrophs, the variety of organisms is so large that some group of the species nearly always will adapt to the biodegradable organic material in the influent or changes in the environmental conditions. With the nitrifiers, once the environmental conditions change to beyond a rather restricted range, they cease to grow.

2.1 Electron donors and acceptors

For all the organisms the energy source serves two functions:

(1) for the supply of material which is transformed to new cell material i.e. synthesis of new cell mass;

(2) for the supply of energy to effect the transformation.

For heterotrophic growth energy is obtained from the carbonaceous material, as follows: Each organic molecule is split to give hydrogen ions, carbon dioxide molecules and electrons. Because it releases electrons the organic molecule is called an *electron donor* and on yielding electrons the molecule is said to be *oxidized*. The electrons (and hydrogen ions) are eventually transferred to a molecule that can receive them and this molecule is called an *electron acceptor*; on receipt of elecrons the molecule is said to be *reduced*. In this oxidation-reduction (redox) reaction *free energy* is released, that is, energy that can be utilized to do work.

Under *aerobic* conditions the electron acceptor is oxygen (O_2) and it is reduced to water (H_2O). If no oxygen is available but nitrate (NO_3^-) or nitrite (NO_2^-) is, an *anoxic* state exists; the NO_3^- and NO_2^- serve as electron acceptors and both are reduced to nitrogen gas and water. If no O_2, NO_2^- or NO_3^- is present an *anaerobic* state exists and the organism generates its own electron acceptor internally (endogenously), in contrast to externally provided (exogenous) electron acceptors (O_2, NO_2^- and NO_3^-), provided in the aerobic and anoxic states.*

The quantity of free energy released in a redox reaction depends on the electron donor and the acceptor. When carbonaceous material is the donor and oxygen is the final acceptor the free energy is relatively high; with NO_2^- and NO_3^- as acceptor the free energy is about 5% less than with oxygen (McCarty, 1964). Internally generated electron acceptors give rise to redox reactions that yield very little free energy.

Taking oxygen as the electron acceptor the free energy per electron transferred for different configurations of carbonaceous molecules lies within a relatively narrow band of values. For sewage which contains an innumerable number of different carbonaceous substances, the mean free energy per electron transferred is virtually constant, McCarty (1964).

In aerobic synthesis of organism mass a *fraction* of the carbonaceous material (with oxygen as electron acceptor) is oxidized via a complex set of redox reactions, to yield free energy, see Figure 1.1. The free energy (via ADP-ATP exchanges) is utilized to reorganise the remaining carbonaceous molecules to protoplasmic material, the free energy being lost as heat in this reorganization.

In the qualitative description above the question that arises is: "How does one quantify this synthesis reaction?" The quantification rests on the indestructability (conservation) of matter in chemical reactions.

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*In Bacteriology if the terminal hydrogen ion and electron acceptor is molecular dissolved oxygen, then the process of substrate oxidation is termed aerobic; if it is chemically bound oxygen, i.e. nitrate, nitrite or sulphate, the process is termed anaerobic; if the terminal electron acceptors originate outside the cell, i.e. exogenous, the process is termed one of respiration; if the terminal electron acceptors originate inside the cell, i.e. endogenous, the process is termed fermentation. In Sanitary Engineering, the usage of some of these terms differs from that in Bacteriology. In particular if no dissolved oxygen is present but nitrate is, the process of environment is called *anoxic*; if neither dissolved oxygen nor nitrate is present, the process or environment is called anaerobic.

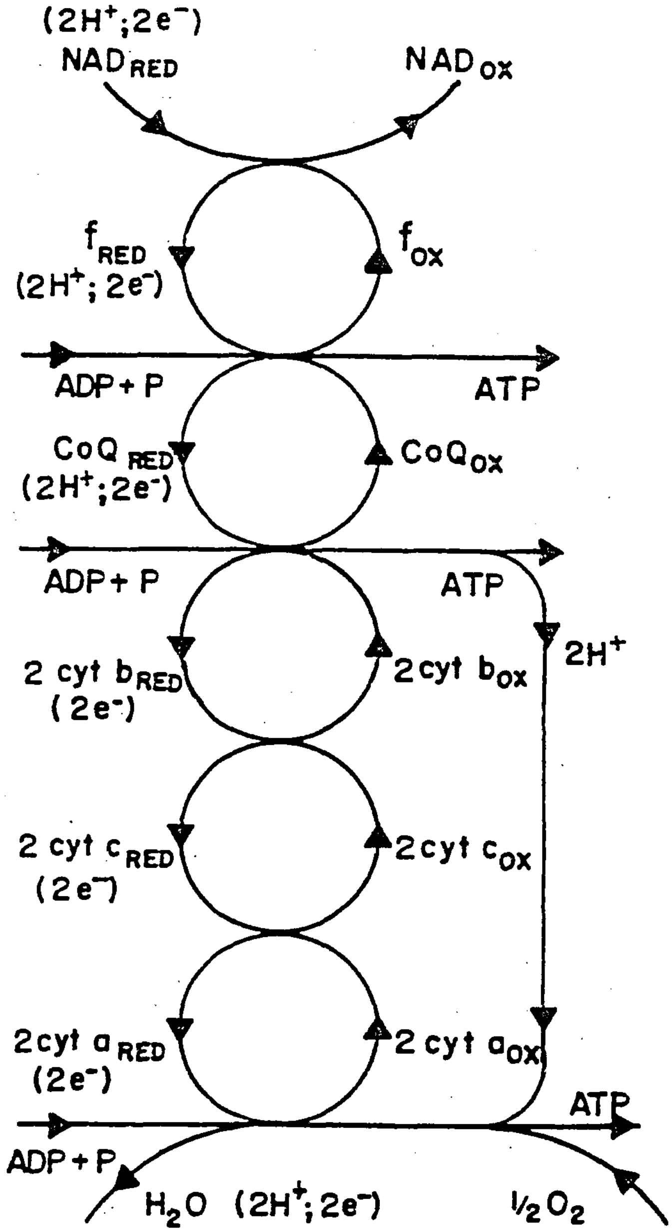

FIGURE 1.1a: ATP (biological energy) formation by sequential oxidation-reduction reactions (e^- transport) in the cytochromes of the micro-organism

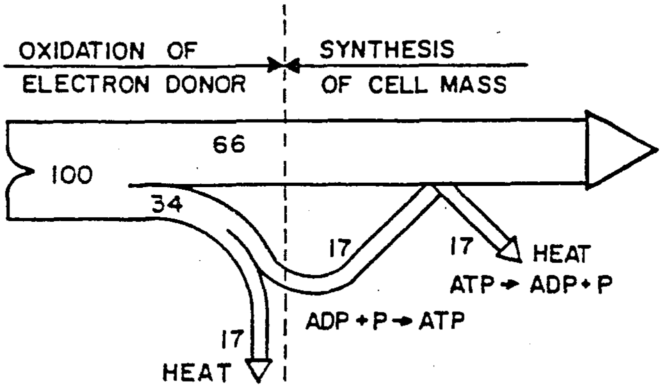

FIGURE 1.1b: Energy change during synthesis

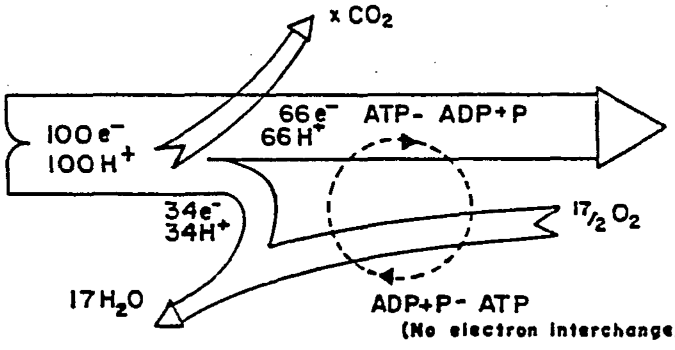

FIGURE 1.1c: Electron changes during synthesis

During synthesis, in the set of redox reactions, electrons and hydrogen ions are transferred from the donor to the acceptor (see Figure 1.1) but, the electrons can be neither created nor destroyed; however, they can be measured, as follows: The number of electrons accepted in the reduction of a molecule of oxygen to water is a defined physical-chemical quantity. If the carbonaceous material is oxidized completely to CO_2, with oxygen as acceptor, the mass of oxygen consumed gives a direct stoichiometric measure of the electrons transferable from the organic molecules, and, as the free energy per electron (with O_2 as acceptor) is virtually constant for all carbonaceous materials, indirectly a measure of the free energy also is obtained. Thus, in a synthesis reaction, by (1) oxidizing completely the input biodegradable material and (2) oxidizing completely the synthesized material, the ratio

$$\frac{\text{oxygen demand of synthesized material}}{\text{oxygen demand of input material}}$$

$$= \frac{O_2 \text{ equivalent of synthesized material}}{O_2 \text{ equivalent of input material}}$$

$$= \frac{\text{electrons available in synthesized material}}{\text{electrons available in input material}}$$

= specific *yield* coefficient of the reaction.

Very closely the specific yield coefficient above also reflects the ratio

$$\frac{\text{free energy in the synthesized material}}{\text{free energy in the input material.}}$$

The difference between (O_2 equivalent of input material) and (O_2 equivalent of synthesized material), reflects the electrons transferred to oxygen in order to release free energy to drive the synthesis reaction i.e.,

$$O_2 \text{ demand} = (O_2 \text{ equivalent of input material}) - (O_2 \text{ equivalent of synthesized material}). \quad (1.1)$$

This equation is stoichiometrically exact and forms the basis for fundamental investigations into the description of the activated sludge process. Very closely, in fact for all practical purposes exactly, it also reflects the energy changes in aerobic and anoxic redox reactions. Thus, by the developments above, we can trace the energy changes by measuring the electrons (as equivalent oxygen) in the input material and the electrons contained in the synthesized material, the difference between these two being equal to the oxygen utilized during change in state of the material. Electrons are indestructible and can and always should be accounted for in any investigation; this brings to the fore the importance of electron *mass balances* to check the validity of investigations both in research and in the operation of a plant.

During nitrification, ammonia serves as the electron donor and oxygen as the acceptor, to convert NH_4^+ to NO_2^- and NO_3^-. Ammonia is present in the influent either as an NH_3 radical in protein molecules (albuminoid ammonia) or as dissolved ammonia (NH_3) and ammonium ions (NH_4^+) (free and saline ammonia respectively). On conversion of the proteinaceous ammonium radical to the free and saline form (by the heterotrophs) it becomes available as an electron donor with oxygen as acceptor to form NO_3^- and NO_2^-. As the molecular forms of both NH_4^+ and NO_3^- are fixed, once the total mass of nitrogen available for nitrification is known the oxygen necessary to oxidize it is also known, and hence the electron donor capacity.

One last point: The yield as defined earlier, is always an experimentally derived value. Biochemistry is now sufficiently advanced to allow the yield to be calculated theoretically fairly accurately by taking into account the stoichiometry of the electron donor and acceptor, the thermodynamic free energy released per electron transferred and the biochemical pathways through which the transfer takes place (for example, the Embden-Meyerhof pathway – Krebs cycle sequence; Payne, 1971). Such an approach, although of importance conceptionally, is not so valuable practically because in wastewater treatment plants, the microbiological mass does not consist of pure culture but comprises a complex ecological system of primary organisms and ones that prey on them.

3. CARBONACEOUS ENERGY MEASUREMENT

3.1 Chemical Oxygen Demand test

The discussion in the previous section, in which it was shown that the energy changes in biological reactions is reflected in the number of electrons transferred, led to the conclusion that the electron donor capacity can be measured in terms of the oxygen required to oxidize the carbonaceous matter to CO_2. Such a measurement is available in the Chemical Oxygen Demand (COD) test.

In the COD test the electron donor capacity of carbonaceous material is measured by oxidizing the material by a strong oxidant i.e. a hot dichromate sulphuric acid solution, in which the electron acceptor is oxygen. This oxidant is so strong that oxidation of the carbonaceous matter is complete, or very near complete. From the half reaction for the reduction of oxygen,

$$4e^- + 4H^+ + O_2 \rightarrow H_2O$$

it can be seen that 1 mole of molecular oxygen (32 g) is equivalent to 4 electron equivalents. Stoichiometrically in the COD test when the equivalent of 1 g O_2 is used up then the mass of COD oxidized is alslo 1 g, i.e.

$$1 \text{ g COD} \equiv 1 \text{ g } O_2 \equiv 1/8 \text{ electron equivalent.}$$

The COD test, therefore, supplies the information required to trace the electron and associated energy changes in a process.

Ammonia is not oxidized in the test so that the test reflects only the electron donor capacity of carbonaceous material.

In practical application some organic material is not oxidized by the COD test under any circumstances, such as the aromatic hydrocarbons and pyridenes, although these are utilized by micro-organisms. Poor estimates of COD can also arise with lower fatty acids such as acetate because these compounds are in the unionized form at the low pH at which the test is performed – the unionized molecules can be extremely difficult to oxidize; in this regard addition of a catalyst silver sulphate overcomes the problem to a large degree to give virtually 100% oxidation.

In performing a COD test it should be remembered that the test involves a time dependent reaction and the specified refluxing period of 2 hours must be adhered to to ensure complete or near complete oxidation. Furthermore, the degree of oxidation is subject to mass action effects – equivalent masses of the same organic compound refluxed at different final test volumes yield different results for the COD; the temperature at which the refluxing step takes place also affects the oxidation rate, the refluxing temperature in turn is dependent on the concentration of sulphuric acid in the test. These factors have all been considered in the development of the COD test procedure. Hence, the COD test will give reliable results only if the test is done in strict accordance with the set procedures, for example, those in Standard Methods (1981). If these procedures are strictly followed our experience is that the COD test adequately reflects the electron donor capacity of the sample. Application of COD test results to carbonaceous material mass balances on activated sludge plants at steady state also indicate that with careful experimental work mass balances between 98 and 102 per cent are obtainable i.e. the mass of influent COD per day can be accounted for by the sum of the daily mass of oxygen utilized and the daily masses of COD in the waste sludge and effluent. Extensive investigations into COD balances by Schroeter, Dold and Marais (1983) have indicated that where poor balances are obtained, invariably the causes can be traced to errors in oxygen demand for nitrification (to correct the total oxygen demand to the carbonaceous oxygen demand) and/or errors in the oxygen utilization rate measurement.

3.2 Total Oxygen Demand test

In the Total Oxygen Demand (TOD) test the sample is oxidized at high temperature in a combustion oven. The TOD is the mass of oxygen required for the oxidation of all oxidizable material contained in a unit volume of wastewater sample; both carbonaceous and nitrogenous compounds are oxidized. A problem with this test is that the amount of oxygen available in the combustion oven influences the degree of oxidation of ammonia and organic nitrogen – with oxygen in excess there is a conversion of nitrogenous material to nitric oxide, whereas when oxygen is not in excess, oxidation of the nitrogenous material may be partial only. Furthermore, nitrate, nitrite and dissolved oxygen in the sample influence the TOD test result. Because the test determines both the carbonaceous and nitrogenous oxidation potential, an additional TKN test is required to isolate the carbonaceous fraction of the TOD. At the present stage of development of the test, difficulty is found in obtaining representative and reproducible readings for samples which contain particulate matter – these samples must be thoroughly homogenised to obtain representative readings as the sample volume is extremely small (10 – 20 μl). Sample homogenisation also increases the dissolved oxygen contamination. The difficulties of applying the TOD test to wastewater can be overcome, but the resulting test procedure reduces its attraction for general routine use.

3.3 Biochemical Oxygen Demand test

This test provides a measure of the oxygen consumed in the biological oxidation of carbonaceous material in a sample over 5 days at 20°C. According to Gaudy (1972) the origins of this test can be traced back about a hundred years to Frankland who appears to have been amongst the first to recognise that the observed oxygen depletion in a stored sample is due to the activity of micro-organisms. The Royal Commission on Sewage Disposal appeared to have developed the concept of using the oxygen depletion as a measure of the strength of pollution. Over the past century extensive research towards elucidating the meaning of the test, quantifying it and integrating it in theoretical descriptions of wastewater treatment processes, has been undertaken. To date the Biological Oxygen Demand (BOD) test is still the most popular parameter for assessing the pollution strength of influents and effluents and for describing the behaviour of treatment processes. Its continued popularity seems to stem mainly from the body of experience that has built up in its use. Whereas there is justification for retaining it in effluent descriptions for regulatory purposes, in so far as present day scientific descriptions of treatment process is concerned few cogent arguments for its use over that of the COD test can be produced.

The main deficiency of the BOD test arises from the inadequacy with which it assesses the electron donor capacity of the carbonaceous organic material in a sample. At the end of 5 days some of the biodegradable carbonaceous material originally present in the sample remains as unbiodegradable material. Consequently, it is not possible to perform a mass balance even with respect to the biodegradable input material. Furthermore unbiodegradable material originally present in the sample remains unaffected by the test so that there is no procedure, related to the test, whereby an assessment of this fraction can be obtained. (In contrast in the COD test, because the total electron donor capacity is measured, even though the unbiodegradable fractions are not isolated in the test, it is possible to estimate these fractions from the test results and the kinetic behavioural patterns of the treatment process).

Despite the deficiencies listed above, if the BOD_5 value constituted a *consistent* estimate it could still have provided a relative measure of practical usefulness, but even this cannot be guaranteed. Gaudy (1972) in a thought provoking review on the BOD gave a summary of the work on the BOD test (in which he took a leading part). He concluded that the standard approach to formulating the BOD time curve as a first order reaction is quite invalid: The actual curve arises from two main effects, (1) bacterial synthesis from the biodegradable input – a reaction usually complete in one to two days, and (2) predator growth using the synthesized bacteria as substrate. The second phase, being dependent on the first, usually lags the first so that a distinct "pause" or "plateau" in the time curve is exhibited (see Figure 1.2). The occurrence and duration of this pause is dependent on the bacteria-prey relationship initially pre-

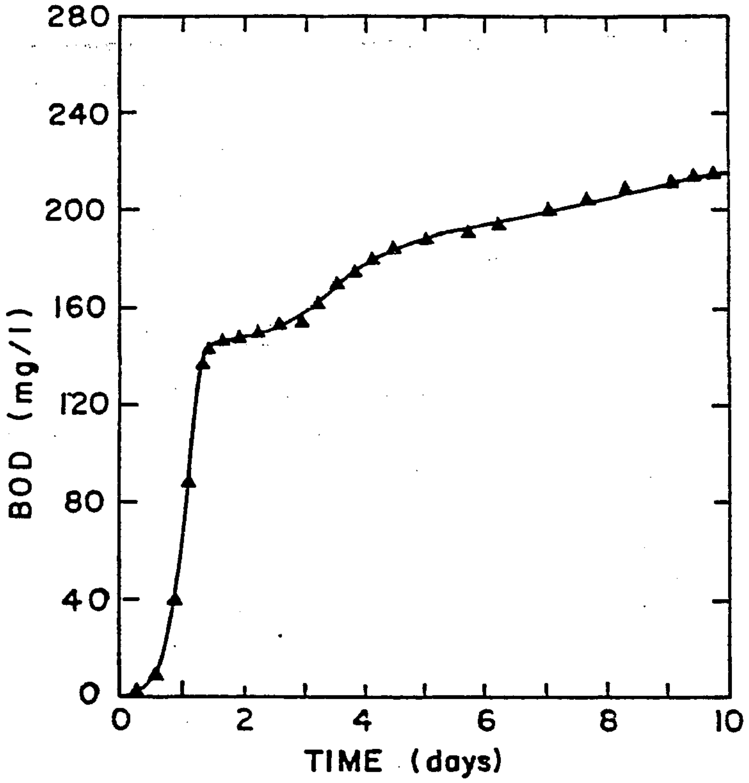

FIGURE 1.2: Typical experimental BOD time curve on wastewater sample with heterogeneous seed clearly showing the "plateau" behaviour after about 1,5 days

sent in the sample, the "strength" of the organic material, presence of inhibitory substances and so on. These compound effects certainly are not reflected by a first order formulation. Furthermore, should nitrification occur during the 5 days the BOD value can be completely misleading; this may not be apparent unless procedures to suppress nitrification are imposed on the test. Evidently the BOD_5 value is subject to a number of influences which can be present in varying degrees resulting in unknown variable effects on the observed value.

Perhaps, the neatest summing up of the use of the BOD test is that due to Hoover, Jasewicz and Porges (1953): "The BOD test is paradoxical. It is the basis of all regulatory actions and is used routinely in almost all control and research studies on sewage and industrial waste treatment. It has been the subject of a tremendous amount of research, yet no one appears to consider it adequately understood or well adapted to his own work". Considered in the light of the biochemical descriptions of biological growth these remarks are still pertinent today.

3.4 Total Organic Carbon test

In the Total Organic Carbon (TOC) test the sample is oxidized in a combustion chamber in the presence of oxygen and the carbon dioxide content of the burnt gas is measured. If the organic carbon content only is required, then it is necessary to have a pre-treatment stage in sample preparation by acidification and CO_2 purging by sparging, or, to do an additional test in which the inorganic carbon only is measured. The older instruments can only measure the carbon content of soluble organic material but newer models can deal with particulate organic material.

As a procedure to estimate the electron donor capacity of the sample the TOC test can be very misleading. This arises from the fact that the ratio of organic carbon to electrons is not constant for different organic materials. For example taking glucose and glycerol:

The number of available electrons may be determined by

(i) measuring the number of moles of O_2 consumed in the complete combustion of one mole of the compound of interest, and

(ii) multiplying by four (which is the relative number of electrons required to reduce one mole of oxygen).

e.g. (a) Glucose, $C_6H_{12}O_6$

$C_6H_{12}O_6 + 6\,O_2 \rightarrow 6\,CO_2 + 6\,H_2O$

6 moles $O_2 \Rightarrow 6 \times 4 = 24\ e^-$ compound

i.e. $24/6 = 4\ e^-$ available per unit organic C.

(b) Glycerol, $C_3H_8O_3$

$C_3H_8O_3 + 7/2\,O_2 \rightarrow 3\,CO_2 + 4\,H_2O$

7/2 moles $O_2 \Rightarrow \frac{7}{2} \times 4 = 14\ e^-$ compound

i.e. $14/3 = 4{,}66\ e^-$ available per unit organic C.

Comparing glucose and glycerol the electron donor capacity per unit C differs by 17%.

The analysis above points to the inadvisability of utilizing the TOC test to describe the energy based behaviour of wastewater treatment processes. This does not mean that the test does not have other uses; it serves a valuable function as control parameter or as a parameter in tertiary treatment where the carbon content in the effluent may be of great importance.

4. COD/VSS RATIO

Consider a completely biodegradable soluble substrate, for example glucose. If a sample of glucose solution is innoculated with bacteria, syntheses of new bacterial mass will occur. The change in soluble COD, ΔCOD (soluble), will be reflected in an increase in the COD of the bacterial mass, ΔCOD(bacteria), and oxygen utilized to generate free energy, ΔO_2(utilized),

$$\Delta\text{COD(soluble)} = \Delta\text{COD(bacteria)} + \Delta O_2\text{(utilized)} \tag{1.2}$$

Equation (1.2) reflects the destination of the electrons from the donor to the synthesized material and the electron acceptor, oxygen. From extensive work reported in bacteriology, reviewed by Payne (1971), the fraction of ΔCOD(soluble) appearing as ΔCOD(bacteria) appears to be nearly constant. Usually the fraction is expressed as a ratio called the specific yield coefficient Y_{COD} (see Eq 1.1) i.e.

$$\Delta\text{COD(bacteria)}/\Delta\text{COD(soluble)} = Y_{COD} \tag{1.3}$$

Equation 1.2 can be rewritten in terms of Y_{COD}

$$\Delta\text{COD(soluble)} = Y_{COD}.\Delta\text{COD(soluble)} + \Delta O_2 \tag{1.4}$$

and recasting Eq (1.4)

$$\Delta O_2 = (1 - Y_{COD})\,\Delta\text{COD(soluble)} \tag{1.5}$$

In wastewater treatment kinetics the usage has arisen, probably from the work of bacteriologists, to measure the mass of *volatile solids generated* in the synthesis reaction, X_a, rather than the COD of the volatile solids generated, ΔCOD(bacteria). To retain the usefulness of Eq (1.4), in particular in the form given by Eq (1.5), it is necessary to relate ΔX_a to ΔCOD(bacteria). The usage has developed to express the relationship between ΔX_a and ΔCOD(bacteria) by the ratio of the two parameters, called the COD/VSS ratio, defined by the symbol f_{cv},

$$f_{cv} = \text{COD/VSS} = \Delta\text{COD(bacteria)}/\Delta X_a$$

Using this definition the yield coefficient Y_{COD} is expressed in terms of a specific yield coefficient with respect to volatile solids Y_h, and the COD/VSS ratio, f_{cv},

$$Y_{COD} = f_{cv}.Y_h \tag{1.6}$$

and inserting in Eq (1.5),

$$\Delta O_2 = (1 - f_{cv}Y_h)\,\Delta\text{COD(soluble)} \tag{1.7}$$

Equation (1.7) is very important in biological wastewater treatment kinetics.

Considerable research has been undertaken to establish a value for f_{cv} and to determine the factors affecting it. Theoretically f_{cv} has been evaluated by accepting an empirical stoichiometric formulation for biological sludge, for example that of Hoover and Porges (1952),

$$C_5H_7O_2N + 5O_2 \rightarrow 5CO_2 + 2H_2O + NH_3$$

Stoichiometrically,

$$113 \text{ g VSS} \rightarrow 160 \text{ g O} \rightarrow 160 \text{ g COD}$$
$$\text{i.e. } 1 \text{ g VSS} \rightarrow 1{,}42 \text{ mg COD}$$
$$\text{i.e. COD/VSS} = 1{,}42 \text{ mg COD/mg VSS}$$

Eckenfelder and Weston (1956) presented experimental data that supports this theoretical ratio. However, other theoretical ratios have been proposed and each find some experimental confirmation, see Table 1.1. The basic difficulty in evaluation of f_{cv} arises from the fact that f_{cv} is a ratio so that the small random errors in either the COD or the VSS magnify the dispersion of the ratio; a reasonably stable mean estimate of the f_{cv} value of a sludge requires at least thirty COD and VSS pair determinations. With smaller numbers the mean f_{cv} value can still vary significantly.

TABLE 1.1 THEORETICAL COD VALUES FOR VARIOUS EMPIRICAL FORMULATIONS FOR MICROBIAL SLUDGE (AFTER McCARTY, 1964).

Bacterial formulation	Molecular weight	Theoretical COD in grams			
		Per mole	Per g TSS	Per g VSS	Per g carbon
$C_5H_7O_2N$	113	160	1,28	1,42	2,67
$C_5H_9O_3N$	131	160	1,10	1,22	2,67
$C_7H_{10}O_3N$	156	232	1,33	1,48	2,76
$C_5H_8O_2N$	114	168	1,32	1,47	2,80

Considering the COD/VSS ratio of sludges derived from activated sludge processes treating municipal effluents, it should be remembered that not only is active volatile mass generated but volatile mass is also present in unbiodegradable particulate form in the influent (which accumulates in the sludge mass) and inert biological residue generated due to endogenous respiration (which also accumulates in the sludge mass). Consequently the sludge normally consists of three fractions, active, inert influent and endogenous. Each of these could have a different f_{cv} value, a factor not considered in deriving the theoretical ratios discussed earlier. The proportions of the three fractions in a sludge depend on the sludge age, the active fraction being relatively small at long sludge ages and high at short sludge ages. For practical purposes an important aspect, therefore, is whether f_{cv} changes with sludge age or whether a mean experimentally derived value can be assigned to all activated sludges. Accordingly Marais and Ekama (1976) and Schroeter, Dold and Marais (1982) investigated the COD/VSS ratio of activated sludge on laboratory scale units treating municipal wastewater for sludge ages ranging from 2 to 30 days.

Their data indicate that despite the significant changes in proportion of the three fractions with changes in sludge age, f_{cv} appeared to remain substantially constant, at 1,48 mgCOD/mgVSS. This finding is very important because it allows that COD balances generally can be performed using VSS measurements on the waste sludge.

5. REFERENCES

AMERICAN PUBLIC HEALTH ASSOCIATION, AWWA and WPCF (1981) Standard methods for the examination of water and wastewater, 15th ed. Published by APHA, Washington.

ECKENFELDER, W.W. and WESTON, R. F. (1956) Kinetics of biological oxidation. In *Biological treatment of sewage and industrial wastes* Vol. 1. Eds. B.J. McCabe and W.W. Eckenfelder, Rheinhold Publ. Co. New York, 18-34.

GAUDY, A.F. (1972) *Water pollution microbiology*. Ed. R. Mitchell, Wiley Interscience, New York.

HOOVER, S.R., JASEWICZ, L. and PORGES, N. (1953) An interpretation of the BOD Test in terms of endogenous respiration of bacteria. *Sewage and Industrial Wastes*, 25, 1163-1173.

HOOVER, S.R. and PORGES, N. (1952) Assimilation of dairy wastes by activated sludge. II. The equations of synthesis and rate of oxygen utilization. *Sew. and Ind. Wastes J.*, 24, 306-312.

MARAIS, G.v.R. and EKAMA, G.A. (1976) The activated sludge process Part 1 – Steady state behaviour, *Water SA*, 2, 164-200.

McCARTY, P.L. (1964) Thermodynamics of biological synthesis and growth, *2nd Int. Conf. Wat. Poll. Research*, Pergamon Press, New York, 169-199.

PAYNE, W.J. (1971) Energy yields and growth of heterotrophs. Annual Review, Microbiology, 17-52.

SCHROETER, W.D., DOLD, P.L. and MARAIS, G.v.R. (1982) The COD/VSS ratio of the volatile solids in the activated sludge process, Res. Rept. No. W 45, Dept. of Civil Eng., University of Cape Town.

CHAPTER TWO

NATURE OF MUNICIPAL WASTEWATERS

G.A. Ekama and G.v.R. Marais

1. WASTEWATER CHARACTERIZATION

In so far as it affects biological treatment, municipal waste flows can be characterized chemically and physically.

Chemical characterization comprises identification and measurement of: (a) carbonaceous and nitrogenous constituents in their various fractions, soluble and particulate, biodegradable and unbiodegradable; and (b) inorganic constituents principally total alkalinity (Alkalinity), total acidity, pH and phosphorus. Inorganic constituents such as calcium, sodium, magnesium, chloride and sulphate usually are of minor importance with the exception of calcium which, in certain circumstances, can be of significance in phosphorus removal by calcium phosphate precipitation.

Physical characterization comprises separation of the wastewater into dissolved, suspended and settleable constituents.

2. CHEMICAL CHARACTERIZATION

2.1 Carbonaceous materials

Characterization of the carbonaceous material in the influent is done via the Chemical Oxygen Demand (COD) test. For activated sludge process design it is necessary to identify and know the magnitudes of the various fractions of the influent COD as these significantly affect the response of the process. Research at the University of Cape Town has indicated that the subdivisions shown diagrammatically in Figure 2.1 are essential for accurate description of the behaviour of the process.

2.1.1 Biodegradable and unbiodegradable fractions

From Figure 2.1 the first major subdivision is into biodegradable COD (S_{bi}) and unbiodegradable COD (S_{ui}) fractions. This division is important for the following reasons: In design, knowing biodegradable COD concentration and the flow per day gives the biodegradable COD load on the plant; knowing the mass of biodegradable COD load per day and selecting a sludge age, the daily carbonaceous oxygen requirements and the active mass in the process can be estimated from the kinetic equations governing the process (Eq 4.10, Chapter 4). Conversely the fraction of the total influent COD which is biodegradable can be estimated from the *carbonaceous* oxygen consumption rate in a steady state laboratory scale completely mixed activated sludge process run at one or more sludge ages. The mass of S_{bi} is then estimated by inserting the measured oxygen demand into the steady state equa-

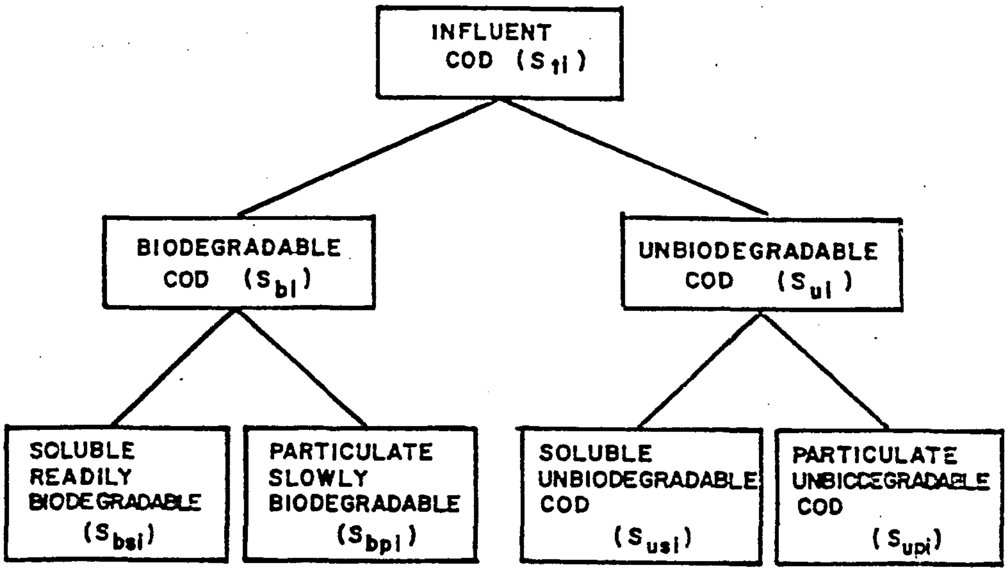

FIGURE 2.1: Division of the total influent COD in municipal wastewaters into its various constituent fractions

tion for the carbonaceous oxygen demand and by trial and error finding the value of S_{bi} that balances the equation, (Eq 4.15, Chapter 4). This approach will be adequate for developing information for design. It presumes that the various constants in the kinetic equations, such as the specific yield constant Y_h and the specific endogenous mass* rate constant b_h have been correctly evaluated from research studies, (Marais and Ekama, 1976). It presumes also that all the biodegradable COD has been metabolized, a presumption justified from research findings if the sludge age is greater than 3 days at 20°C, (Dold, Ekama and Marais, 1980). If nitrification occurs in the process this either has to be allowed for when determining the carbonaceous oxygen demand from the measured oxygen demand (see Chapter 5) or, nitrification must be suppressed by addition of a nitrification inhibitor to the process, for example by addition of thio-urea at intervals to the process.

2.1.2 *Unbiodegradable fractions*

The next major subdivision is to divide each of the unbiodegradable and biodegradable COD concentrations, S_{ui} and S_{bi} respectively, into their relevant subconcentrations.

Consider first the unbiodegradable COD concentration, S_{ui}. This concentration consists of two fractions, an unbiodegradable soluble COD, S_{usi} and an unbiodegradable particulate COD, S_{upi}. Both fractions are hypothesised to be unaffected by biological action so that at steady state the masses that enter equal the masses that leave the process. The S_{usi} passes out in the liquid effluent discharge. The S_{upi} is enmeshed in the sludge, accumulates in the system to a mass equal to that entering in the influent per day, multiplied by the sludge age, but eventually at steady state the masses entering and leaving the system are the same. The mass leaves the system via the daily sludge mass wasted from the plant. Thus the S_{upi} has the principal effect of increasing the mixed liquor concentration.

Usually it is more convenient to express the unbiodegradable particulate influent fraction not in terms of its COD but in terms of its influent volatile solids concentration, X_{ii}. This is readily accomplished by noting in Chapter 1 that the COD and volatile solids are defined as follows:

$$X_{ii} = S_{upi}/f_{cv}$$

where f_{cv} = COD/VSS ratio = 1,48 mgCOD/mgVSS.

Measurement of S_{usi} is relatively simple; a laboratory scale unit is run at say two different sludge ages both greater than 3 days. The COD of the centrifuged effluent or the filtrate through a fine filter, estimates S_{usi}. The balance of the influent COD, after subtracting S_{bi} and S_{ui} from the total influent COD, S_{ti}, is the unbiodegradable particulate COD, S_{upi}, i.e.

$$S_{upi} = S_{ti} - S_{bi} - S_{usi}$$

The reasonableness of the S_{upi} estimate is checked by running a laboratory scale unit at a long sludge age, say, 15 days or longer, calculating the total volatile solids concentration using the input values for S_{bi} and S_{upi} (i.e. X_{ii}) (Eq 4.13, Chapter 4) and checking this value against the volatile mass observed. A long sludge age is selected because the total volatile mass of sludge becomes more sensitive to X_{ii} as the sludge age increases. Overall the S_{upi}, S_{ui} and S_{bi} must give rise to consistency at the different sludge ages. Once this consistency is judged adequate the COD fractions are acceptable for design.

2.1.3 *Biodegradable fractions*

The data obtained above are all on steady state aerobic activated sludge units. Subdivision of S_{bi} into readily biodegradable soluble COD, S_{bsi}, and slowly biodegradable particulate COD, S_{bpi}, (Fig. 2.1) can be obtained only by dynamic loading of the units, for the following reasons: The fraction S_{bsi} is rapidly taken up by the sludge in a matter of minutes and metabolized, giving rise to a high unit rate of oxygen demand for synthesis. (The kinetics of utilization are in accordance with Monod's equation). The fraction S_{bpi} requires to be adsorbed and stored on the organism, broken down to simpler chemical units by extracellular enzymes and then absorbed and metabolized by the organism. The extracellular breakdown is slow and forms the limiting rate in the synthesis reaction, the rate being about 1/7 to 1/10 that for the readily biodegradable COD, S_{bsi}. (The kinetics for synthesis of particulate COD is formulated via Levenspiel's active site surface reaction theory). Under steady state (i.e. constant flow and load conditions) the stored S_{bp} is negligibly small at sludge ages greater than about 2 to 3 days. However at short sludge ages (2-3 days) under cyclic flow conditions under square wave loading, 12 h on, 12 h off, during the loading period the rate of S_{bp} adsorption and storage is so large that the organism cannot deplete it concurrently, even at the maximum utilization rate. As a consequence, after the loading period terminates the utilization rate of the stored S_{bp} continues at the same maximum rate for a while, and then slowly declines. This behaviour allows the readily biodegradable COD to be determined; measure the oxygen utilization rate for a period before loading terminates and a period after, the difference in the rates defines the rate due to the S_{bs}. Details of the test procedure are given in Appendix A2.

The importance of the S_{bsi} fraction will become evident when denitrification and biological excess phosphorus removal are discussed in Chapters 6 and 7 respectively.

2.1.4 *Analytic formulation*

For analysis the relationship indicated in Figure 2.1 can be expressed as follows:

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*Y_h = 0,45 mgVASS/mgCOD utilized, independent of temperature
b_h = 0,24 mgVASS/mgVASS/day at 20°C and
b_{hT} = 0,24 $(1{,}029)^{(T-20)}$ at temperature T°C, see Chapter 4.

Biodegradable and unbiodegradable COD fractions:

$$S_{ti} = S_{ui} + S_{bi} \tag{2.1}$$

where

S_{ti} = total influent COD concentration (mgCOD/ℓ)
S_{ui} = unbiodegradable COD concentration (mgCOD/ℓ)
S_{bi} = biodegradable COD concentration (mgCOD/ℓ)

Each of the two fractions on the right hand side of Eq 2.1 is again subdivided, Figure 2.1.

Unbiodegradable COD fractions:

The unbiodegradable COD concentration consists of two components, soluble and particulate, i.e.

$$S_{ui} = S_{usi} + S_{upi} \tag{2.2}$$

where

S_{usi} = unbiodegradable soluble COD concentration (mgCOD/ℓ)
S_{upi} = unbiodegradable particulate COD concentration (mgCOD/ℓ).

It is convenient to express S_{usi} and S_{upi} empirically in terms of the total COD concentration S_{ti}, i.e.

$$S_{usi} = f_{us} S_{ti} \tag{2.3}$$

$$S_{upi} = f_{up} S_{ti} \tag{2.4}$$

where

f_{us} = unbiodegradable soluble COD fraction with respect to the total COD (mgCOD/mgCOD)
f_{up} = unbiodegradable particulate COD fraction with respect to the total COD (mgCOD/mgCOD).

Hence from Eq (2.2)

$$S_{ui} = (f_{us} + f_{up}) S_{ti} \tag{2.5}$$

The particulate unbiodegradable COD concentration may be expressed also in terms of volatile solids, X_{ii}:

$$\begin{aligned} X_{ii} &= S_{upi}/f_{cv} \\ &= f_{up} S_{ti}/f_{cv} \end{aligned} \tag{2.6}$$

where

X_{ii} = unbiodegradable particulate volatile solids concentration in the influent (mgVSS/ℓ)
f_{cv} = COD to VSS ratio of the solids
= 1,48 mgCOD/mgVSS.

Biodegradable COD fractions:

The biodegradable COD concentration is found from Eq 2.1 as follows:

$$S_{bi} = S_{ti} - S_{ui}$$

and from Eq 2.5

$$\begin{aligned} &= S_{ti} - S_{ti}(f_{up} + f_{us}) \\ &= S_{ti}(1 - f_{up} - f_{us}) \end{aligned} \tag{2.7}$$

From Figure 2.1 the *biodegradable* COD, S_{bi}, is divided into *readily biodegradable soluble* COD, S_{bsi}, and *slowly biodegradable particulate* COD, S_{bpi}. Each can be expressed in terms of S_{bi} as follows:

$$S_{bsi} = f_{bs} S_{bi} \text{ and} \tag{2.8a}$$

$$S_{bpi} = (1 - f_{bs}) S_{bi} \tag{2.9a}$$

where

f_{bs} = readily biodegradable COD fraction with respect to the *biodegradable* COD.

The readily biodegradable COD can also be expressed in terms of the *total* COD, S_{ti}, i.e. substituting for S_{bi} from Eq 2.7 in Eq 2.8 yields

$$S_{bsi} = f_{bs}(1 - f_{up} - f_{us}) S_{ti} \tag{2.8b}$$

$$= f_{ts} S_{ti} \tag{2.9b}$$

where

f_{ts} = readily biodegradable COD fraction with respect to the total COD.

2.2 Nitrogenous materials

Characterization of the nitrogenous material in the influent is with the Total Kjeldahl Nitrogen (TKN) and Kjeldahl free and saline ammonia tests. As for the carbonaceous material, the nitrogenous material also is subdivided into different fractions but in a different fashion from that for the carbonaceous material. The subdivision is shown in Figure 2.2.

In certain wastewaters nitrate and nitrite might be present; the TKN test does not include these. The vast majority of municipal effluents will not contain nitrate or nitrite because in most sewerage systems the wastewater will be in an unoxygenated state and any nitrate entering the system is likely to be denitrified before it reaches the wastewater treatment plant.

2.2.1 Free and saline and proteinaceous fractions

The free and saline ammonia fraction, N_{ai}, is determined by a test of this name. The proteinaceous or organic fraction is determined by difference between the TKN, N_{ti}, and free and saline ammonia values, N_{ai}.

The free and saline ammonia is immediately available for incorporation into the bacterial protoplasm and for conversion to nitrite or nitrate, if the process is appropriately designed.

The proteinaceous fraction, from Figure 2.2, consists of a number of sub-fractions, *unbiodegradable soluble* and *unbiodegradable particulate organic nitrogen* fractions, N_{ui} and N_{pi} respectively, and a *biodegradable organic nitrogen* fraction, N_{oi}. The unbiodegradable soluble fraction passes unaffected through the system and is discharged in the effluent similar to the soluble unbiodegradable COD. The unbiodegradable particulate organic nitrogen by implication must be part of the unbiodegradable particulate COD (as X_{ii}), and is treated as such; hence, this fraction leaves the process via the sludge wasted daily. The

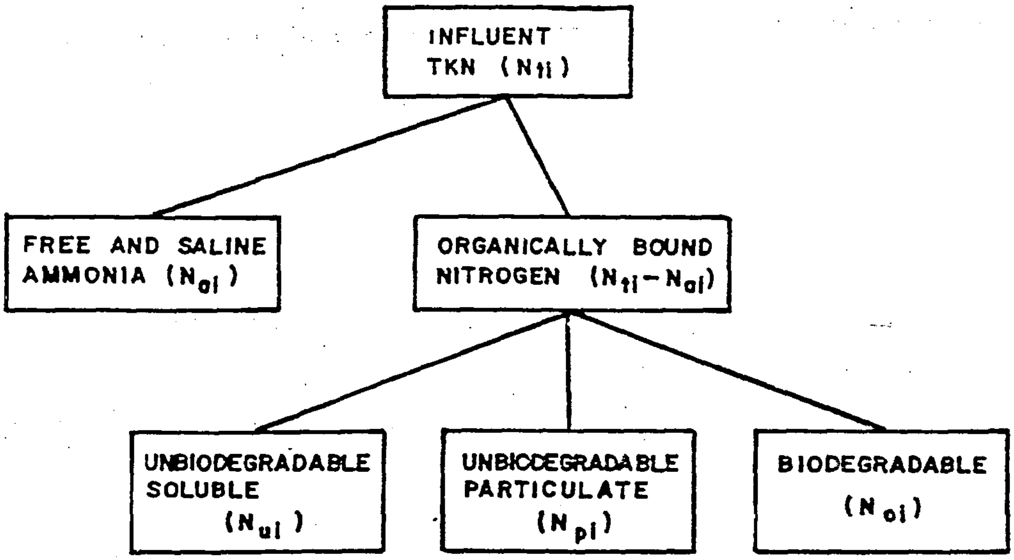

FIGURE 2.2: Division of the total influent TKN in municipal wastewater into its various constituent fractions

biodegradable organic nitrogen is broken down by the heterotrophs to free and saline ammonia which with its influent counterpart participates in any further biologically mediated reactions. The breakdown occurs very readily and normally it can be expected to be virtually complete for all sludge ages greater than about 3 days.

With regard to measurement of these fractions, the free and saline ammonia (N_{ai}) is measured directly by the test bearing this name; the total organic nitrogen is found from the difference between the TKN and free and saline ammonia test concentrations. Difficulties in quantification arise when subdividing the total organic nitrogen into its biodegradable and unbiodegradable fractions, N_{oi}, N_{ui} and N_{pi}, (see Figure 2.2). N_{ui} and N_{pi} are postulated by association to their counterparts in the COD sub-division. Estimates of their magnitudes can be obtained only by comparing the observed response of laboratory scale processes with that predicted by the theoretical model incorporating these fractions. From a large number of such comparisons on processes operated within the range of sludge ages from 2 to 30 days, under cyclic and steady states in the temperature range 12 to 25°C, it would appear that N_{ui} and N_{pi} are necessary in the model to obtain good and consistent correlation between observation and prediction. However, the fractions are small and it is not necessary therefore to have accurate estimates. Concurrently N_{pi} is expressed in terms of the unbiodegradable particulate volatile solids in the influent (X_{ii}) i.e. N_{pi} is 10% of X_{ii} (see Eq 2.13): from experimental data on the TKN/VSS ratio of the MLVSS in the biological reactor, using the same influent wastewater source it was found that the TKN/VSS ratio remained the same irrespective of sludge age from 3 to 30 days despite the fact that the various VSS fractions change with sludge age. Hence it is reasonable to accept that the TKN/VSS ratio of the inert organic solids is equal to that of the other VSS fractions at 0,10 mgN/mgVSS.

TABLE 2.1 APPROXIMATE AVERAGE MUNICIPAL WASTEWATER CHARACTERISTICS FOR RAW AND SETTLED WASTEWATERS IN SOUTH AFRICA

Wastewater Characteristic	Raw	Settled
1. Influent COD (mgCOD/ℓ)	500-800	300-600
2. Influent BOD_5 (mgBOD/ℓ)	250-400	150-300
3. Influent TKN (mgN/ℓ)	35-80	30-70
4. Influent Phosphorus (mgP/ℓ)	8-18	6-15
5. Total Suspended Solids (mg/ℓ)	270-450	150-300
6. Settleable Solids in (mg/ℓ)	150-350	0-50
in (mℓ/ℓ)**	6-14	0-2
7. Non-Settleable Solids (mg/ℓ)	100-300	100-300
8. TKN/COD ratio (mgN/mgCOD)	0,07-0,10	0,09-0,12*
9. Total P/COD ratio (mgP/mgCOD)	0,015-0,025	0,020-0,030*
10. Unbiodegradable particulate COD fraction (f_{up})	0,07-0,20	0,00-0,10
11. COD/VSS ratio of unbiodegradable particulate COD (f_{cv}) (mgCOD/mgVSS)	1,45-1,50 ave. 1,48	1,45-1,50 ave. 1,48
12. TKN/VSS ratio of unbiodegradable particulate COD (f_n) (mgN/mgVSS)	0,09-0,12 ave. 0,10	0,09-0,12 ave. 0,10
13. Unbiodegradable soluble COD fraction (f_{us})	0,04-0,10	0,05-0,20*
14. Fraction biodegradable COD (S_{bi}/S_{ti})	0,75-0,85	0,80-0,95
15. Readily biodegradable COD with respect to total (f_{ts})	0,08-0,25	0,10-0,35*
16. Maximum specific growth rate of the nitrifiers (/d)	0,20-0,70	0,20-0,70
17. Ammonia/TKN fraction (f_{na})	0,60-0,80	0,70-0,90*
18. Unbiodegradable soluble TKN fraction (f_{nu})	0,00-0,04	0,00-0,05*
19. Minimum temperature (°C)	10-15	10-15
20. Maximum temperature (°C)	20-30	20-30
21. Alkalinity (mg/ℓ as $CaCO_3$)	200-300	200-300

*Note that the characteristics of settled wastewater should be selected in conformity with the characteristics of the original raw wastewater and primary settling tank behaviour (see Chapter 4, Section 12.1).

**Volume settled in a 1 ℓ Imhoff cone in 2 hours.

Furthermore the unbiodegradable soluble organic nitrogen concentration (N_{ui}) apparently is very low, between 0 and 5% of the influent TKN for raw wastewaters. Later it will be shown (Chapter 5) that the N_{ui} concentration plays a very minor role in design. The suggested fractional values are listed in Table 2.1.

2.2.2 *Analytical formulation*

The total TKN, N_{ti} is divided into its constituents as follows, (see Figure 2.2):

$$N_{ti} = N_{ai} + N_{ui} + N_{pi} + N_{oi} \qquad (2.10)$$

where

N_{ti} = total influent TKN (mgN/ℓ)
N_{ai} = influent ammonia (mgN/ℓ)
N_{ui} = soluble unbiodegradable organic nitrogen in the influent (mgN/ℓ)
N_{pi} = particulate unbiodegradable organic nitrogen in the influent (mgN/ℓ)
N_{oi} = biodegradable organic nitrogen in the influent (mgN/ℓ)

It is convenient to express some of the fractions of nitrogen in terms of the total TKN concentrations

$$N_{ai} = f_{na}\,N_{ti} \qquad (2.11)$$

$$N_{ui} = f_{nu}\,N_{ti} \qquad (2.12)$$

where

f_{na} = ammonia fraction of the influent TKN (mgN/mgN)
f_{nu} = unbiodegradable soluble organic nitrogen fraction (mgN/mgN)

The unbiodegradable organic nitrogen in the particulate material, N_{pi}, is best expressed in terms of the unbiodegradable influent volatile material, X_{ii}, or in terms of its COD counterpart, i.e.

$$N_{pi} = f_n\,S_{upi}/f_{cv} = f_n\,X_{ii} \qquad (2.13)$$

where

f_n = nitrogen fraction of the influent biodegradable volatile particulate material $\cong$ 0,10 mgN/mgX$_{ii}$

From Eqs 2.10 to 2.13, the biodegradable organic nitrogen (N_{oi}) can be found by subtracting N_{ui}, N_{ai} and N_{pi} from N_{ti}

$$N_{oi} = N_{ti}\,(1 - f_{na} - f_{nu}) - f_n\,f_{up}S_{ti}/f_{cv} \qquad (2.14)$$

The significance of the various influent TKN fractions on the design of nutrient removal activated sludge processes is discussed in Chapter 3, Section 3.2 and in Chapter 5, Section 5.1.

2.2.3 *Specific growth rate of nitrifiers*

Experimental observations on a number of different municipal wastewater flows have indicated that the maximum specific growth rate constant of the nitrifiers, μ_{nm}, can differ greatly in value and appears to be specific to each waste flow. The magnitude of μ_{nm} can have a significant influence on the design of nutrient removal processes, see Chapters 3, 4, 5 and 6. Therefore, for optimal design it is most desirable to have an estimate of μ_{nm}. This estimate is obtained by running a completely mixed single reactor at a sludge age of about 8 days, at 20°C imposing a sequence of aerobic and anoxic periods on it and measuring the rate of nitrate increase (or the rate of Alkalinity decrease) during the aerobic period. The data thus obtained can be analysed manually to give the μ_{nm} at 20°C, i.e. μ_{nm20}. By applying the well established relationship between μ_{nm} and temperature the value at any temperature can be determined. Details of the test procedure and analysis of the data are given in Appendix 3.

2.3 Phosphorus

The total phosphorus concentration in municipal wastewaters consists mainly of two fractions, a soluble orthophosphate (PO_4^{3-}) fraction and an organically bound phosphorus fraction which may be soluble or particulate in form. In both settled and raw municipal wastewaters, the orthophosphate fraction predominates, ranging between 70 to 90 per cent of the total phosphorus.

The orthophosphate concentration* can be measured by a colorimetric test of that name. However, this measure records only the orthophosphate concentration of the influent, which may underestimate the total phosphorus in the influent by about 10 to 20 per cent or more. The organically bound phosphorus is converted to orthophosphate in the activated sludge process so that the orthophosphate concentration in the influent underestimates the phosphorus to be removed in the process. For example, if the orthophosphate concentration is 8 mgP/ℓ and the total-P is 10 mgP/ℓ then for an effluent standard of 1 mgP/ℓ orthophosphate, if the plant is designed to remove 7 mgP/ℓ on the basis of the ortho-P influent concentration, the effluent actually will contain 3 mgP/ℓ ortho-P. The process should be designed to remove 9 mgP/ℓ on the basis of the total-P influent concentration. Consequently, for the influent the only reliable test is the total-P test. For the effluent, the filtered ortho-P concentration very closely approximates the filtered total-P concentration and this ortho-P test would appear to be adequate. However, in performing the ortho-P test, a problem sometimes is encountered in that the natural colour in some wastewaters interferes with the colorimetric test giving rise to spurious results.**

3. PHYSICAL CHARACTERIZATION

Physical characterization of the influent exists mainly in estimating the expected performance of the primary settling tank. The function of the settling tank is to separate in some degree the settleable particulate

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*All phosphorus and phosphate concentrations are measured relative to the phosphorus atom i.e. mgP/ℓ.

**This is a problem encountered in treated effluents from the Western Cape, but apparently is absent from treated effluents of Johannesburg and Pretoria.

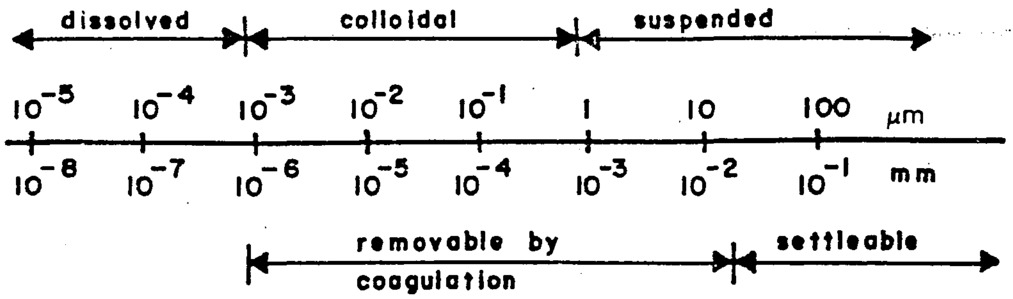

FIGURE 2.3: Particle size range and classification

material from the influent flow to give a supernatant overflow having reduced values of COD, TKN and phosphorus.

Physical characterization basically is made on the relative size of the particles in the wasteflow. Here three broad divisions are recognized, *dissolved*, *colloidal* and *suspended solids*, see Figure 2.3. Generally material of particle size smaller than 10^{-3} to 1 µm is accepted as colloidal and material with particle size larger than 1 µm as suspended. To obtain estimates of the suspended material pore size membrane filters other than 1 µm are in use which can lead to significantly different results so that when reporting or interpreting data, careful note should be made of the pore size of the filter membranes (and papers) on which the data are based.

The fraction of *suspended* solids (i.e. retained on a 1 µm filter) that can be removed by gravity settling is called the *settleable* solids and usually ranges in equivalent pore size from 10 to 50 µm depending on particle density. However, a more practical procedure to estimate the mass of settleable solids in a sample of wastewater, is to use the Imhoff cone: A one litre Imhoff cone is filled with a sample and allowed to settle for two hours. The volume of solid material settled into the bottom of the cone gives a measure (in m*l*/*l*) of the quantity of material that can be removed by gravity sedimentation. From a large number of tests it has been found that 1 m*l*/*l* of settleable solids is approximately equivalent to a dry mass* of 25 mg/*l*; with this equivalence it is possible to obtain an estimate of the mass of settleable solids using the Imhoff test.

The settleable solids can be measured either as *total solids* where total solids include both the organic and inorganic fractions, or, as *volatile or organic solids* where the volatile solid is the difference between the total solids and the ash remaining after the total solids are incinerated in an oven at 600°C for half an hour. At this temperature most inorganic (mineral) salts, except magnesium carbonate, do not decompose or volatilize. The ratio Volatile Solids:Total Solids of the settleable solids from municipal wasteflows is about 0,75 i.e. 75% volatile, 25% mineral so that Volatile Solids ≅ 0,75 Total Solids. The COD associated with the volatile solids can be estimated by multiplying the volatile solid mass by 1,48 i.e. COD ≅ 1,48 Volatile Solids.

The Imhoff cone test serves as a means whereby the percentage removal of Total Phosphorus and TKN in the primary settling tank can be estimated. This is done by determining total-P and TKN concentrations in the waste sample before settling and in the supernatant after settling in the cone.

In a primary settling tank the degree of COD and settleable solids removed by gravity sedimentation depends on the upflow velocity (overflow rate) and retention time. For relatively deep tanks (>3 m) with mean retention times of 2 to 3 h and mean overflow rates of 1,5 to 2 m/h between 50 and 80% of the settleable solids (as given by the Imhoff cone test) and 30 to 50% (mean of 40%) of the total COD will be removed. Increasing the retention time above 3 h does not improve the removals significantly. However, lowering the retention time to say one hour (equivalent to 3 m/h mean overflow rate) reduces the removal to about half of that at 3 h retention time.

With regard to the TKN and P removal in primary tanks this can vary depending on state of the waste flow. It seems that in fresh wastewater a larger fraction of the TKN and P may be bound to the settleable solids than if the wastewater is old and solubilization of the TKN and total P fractions has occurred. Unfortunately insufficient information on these aspects is available to lay down criteria for design. For design it would seem preferable to assume that solubilization does occur so that only a minor fraction of the TKN and P respectively is removed by primary sedimentation, between 15 and 25%, even though there are reported removals of up to 40%.

The discussion above indicates that with primary sedimentation a significant reduction is obtained in COD (40%) but a smaller reduction of TKN and total P (15 to 20%). This has the effect that settled wastewater has increased TKN/COD and P/COD ratios compared to the raw wastewaters. Note also that the readily biodegradable COD concentration (S_{bs}) which is completely in the dissolved form is not reduced so that the concentrations in the raw and settled flow are the same,

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*Mass retained on filter paper (Whatmans No. 1) after drying for 12 hours at 105°C. Measurements taken from raw wastewater at the Cape Flats treatment works by Ekama, Marais and coworkers.

hence the fraction S_{bi}/S_{ti} increases for settled sewage. With regard to the unbiodegradable particulate COD this fraction appears to be largely removed by primary sedimentation so that settled wastewaters contain low fractions of unbiodegradable particulate COD.

4. QUANTIFICATION OF WASTEWATER CHARACTERISTICS

In the above section the various chemical and physical wastewater characteristics were defined. In this section these characteristics are quantified to facilitate their use in design. The influence of the various characteristics on the design of an activated sludge process is discussed in Chapter 3.

The ranges of the magnitudes of the various chemical and physical characteristics of approximately normal *raw* and *settled* municipal wastewaters is such that it is difficult to give smaller ranges for these characteristics. Generally it has been found that the greater the diversity and larger the community from which the wastewater originates, the less the characteristics will vary during the life of the plant. Also

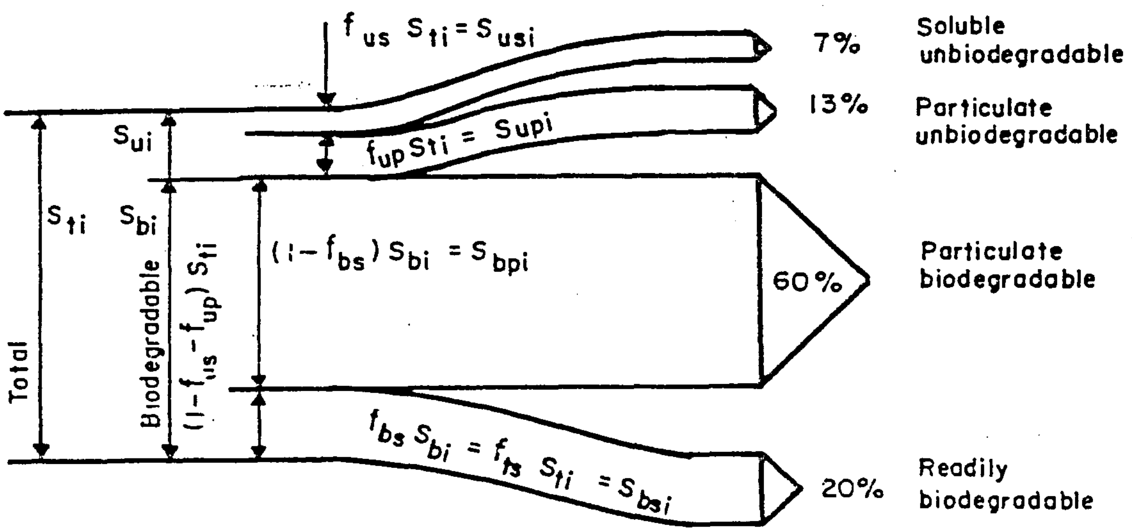

FIGURE 2.4: Diagrammatic representation of the different fractions of the total influent COD concentration. Numerical values apply to raw municipal wastewater

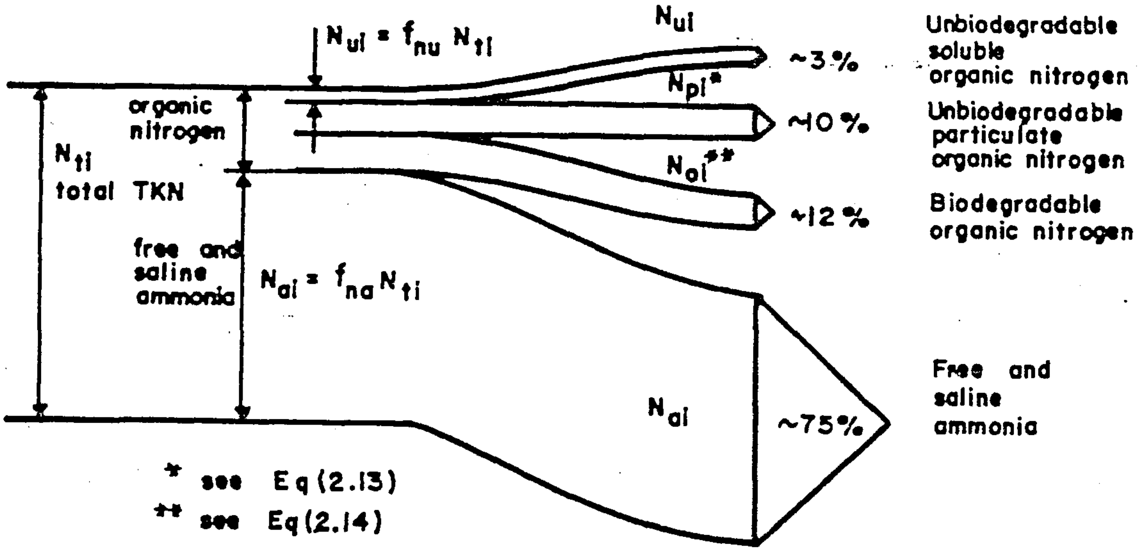

FIGURE 2.5: Diagrammatic representation of the various fractions of the influent TKN in municipal wastewater. Numerical values apply to raw municipal wastewater

the greater the diversity and larger the community, the smaller the daily variations in wastewater characteristics and flow.

A diagrammatic representation of the magnitudes of the different COD and TKN fractions in an approximately normal *raw* municipal wastewater is shown in Figures 2.4 and 2.5 respectively. With regard to the COD fractions, Figure 2.4 shows that about 7% of the total COD is soluble unbiodegradable, 13% particulate unbiodegradable, 60% particulate slowly biodegradable and 20% soluble readily biodebradable. With regard to the TKN fractions, Figure 2.5 shows that about 3% of the total TKN is soluble unbiodegradable organic nitrogen, 10% particulate unbiodegradable organic nitrogen associated with the particulate unbiodegradable COD, 12% biodegradable organic nitrogen and 75% free and saline ammonia.

Caution should be exercised when selecting an unbiodebradable particulate COD fraction (f_{up}): social organization of the community for whom the wastewater treatment process is envisaged must be critically examined. For example, analyses of data reported by Sutton *et al.* (1979) from Canada, indicate that the f_{up} fraction is 0,25 for raw wastewater. Such a high value probably arises from the use of garbage grinders. A similarly high f_{up} is used in the design procedure followed by Black and Veatch (1979) [based on McKinney's activated sludge model] which includes an unbiodegradable organic material fraction which is equivalent to an f_{up} value of 0,23. Wastewaters with high f_{up} fractions will result in greater sludge production than those with low fractions and selecting too low a value will result in underdesign of the treatment process and sludge disposal facilities. This demonstrates the pitfalls in selecting wastewater characteristics uncritically and without recognition of the factors which contribute to their nature and composition. It also demonstrates the dangers of applying uncritically, technology developed overseas to local conditions and *vice versa*.

5. REFERENCES

BLACK and VEATCH (1979) Design procedure for complete mixing activated sludge process. Civil-Environmental Division, P.O. Box 8405, Kansas City, Mo.64114.

DOLD, P.L., EKAMA, G.A. and MARAIS, V.v.R. (1980) A general model for the activated sludge process, *Prog. Wat. Tech.*, **12**, 47-77.

MARAIS, G.v.R. and EKAMA, G.A. (1976) The activated sludge process part I – Steady state behaviour, *Water SA*, **2**, 4, 163-200.

LEVENSPIEL, O. (1972) Chemical Reaction Engineering 2nd Ed., John Wiley, New York, 465-469.

SUTTON, P.M., JANK, B.E., MONAGHAN, B.A. and MURPHY, K.L. (1979) Single sludge nitrogen removal systems. Research Report No 88, Water Technology Centre, Burlington, Canada.

CHAPTER THREE

INFLUENCE OF WASTEWATER CHARACTERISTICS ON PROCESS DESIGN

by

G.A. Ekama and G.v.R. Marais

1. INTRODUCTION

In this chapter considerations in the design of nutrient removal activated sludge processes are qualitatively described to provide a background for the subsequent chapters in which the process design for COD removal only (Chapter 4), nitrification (Chapter 5), nitrification-denitrification (Chapter 6) and biological phosphorus removal (Chapter 7) are quantitatively set out – in these chapters detailed equations, design charts, sample calculations and recommendations for design are presented.

2. RELATIVE IMPORTANCE OF CHARACTERISTICS

In the design of activated sludge processes the characteristics of the wastewater are of prime importance. These characteristics govern both the selection of the process and the removals of nitrogen and phosphorus attainable in the process.

Influent wastewater characteristics of crucial importance are:

(1) Mean influent COD and TKN concentrations (S_{ti} and N_{ti} respectively) and mean daily flow (Q)
(2) Influent readily biodegradable COD fraction (f_{bs})
(3) Influent TKN/COD concentration ratio (N_{ti}/S_{ti})
(4) Influent (Total P)/COD concentration ratio (P_{ti}/S_{ti})
(5) Maximum specific growth rate of the nitrifiers at the reference temperature of 20°C (μ_{nm20})
(6) Average minimum and maximum temperatures (T_{min} and T_{max}) in the process.

Influent wastewater characteristics of lesser importance are:

(7) Unbiodegradable soluble COD fraction (f_{us})
(8) Unbiodegradable particulate COD fraction (f_{up}).

Influent characteristics normally not important in design are:

(9) Unbiodegradable soluble organic nitrogen fraction (f_{nu}) and
(10) Ammonia fraction of the TKN (f_{na}).

3. EFFECTS OF CHARACTERISTICS ON DESIGN

3.1 Influent COD and influent flow

The mean influent COD concentration, S_{ti},* and mean flow per day, Q, define the mass COD load day on the plant, $M(S_{ti})$, by

$$M(S_{ti}) = S_{ti}.Q$$

This COD mass consists of various fractions, as set out in Chapter 2, and these fractions are different for unsettled and settled municipal waste flows. However, to indicate the effect of the COD on the plant design, accept a specific waste flow influent. Then, for a specified sludge age, once $M(S_{ti})$ is known, theoretically the mass of sludge in the reactor, $M(X_v)$ or $M(X_t)$, and the mass of oxygen required daily for carbonaceous degradation, $M(O_c)$, also are known. The sludge age must be specified as the mass is very sensitive to the sludge age. Knowing the mass of sludge, the volume of the reactor is found from the specified mixed liquor concentration, X_v or X_t, i.e. the mass of sludge is "diluted" to the required concentration by providing the appropriate volume,

$$V_p = M(X_v)/X_v \quad \text{or} \quad M(X_t)/X_t$$

Knowing the volume, one can now calculate, if so desired (it is not essential) the nominal hydraulic retention time, R_{hn}, of the reactor,

$$R_{hn} = V_p/Q$$

From a process design point of view the hydraulic retention time is of little interest: For two waste flows, having the same COD fractional constitution, provided $M(S_{ti})$ and X_v are the same for both, the volumes of the two plants and the oxygen demand per day will be identical irrespective of whether the first flow consists of a low S_{ti} and high Q and the second of a high S_{ti} and low Q. The nominal hydraulic retention time will differ, the first plant having a short and the second a long retention time respectively but this does not affect the degradation action, except insignificantly. The influent flow is important in the design of the settling tanks – for the same upflow velocity (overflow rate) in the secondary

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*The mean influent COD, S_{ti}, is the *weighted* mean concentration; it is *not* the mean of the COD concentrations measured over the day on the influent. The magnitude of S_{ti} is found from measuring every hour, say, the COD, $S_{ti}(t)$, and flow, Q(t), and forming

$$S_{ti} = [\sum_{t=1}^{24} S_{ti}(t).Q(t)]/[\sum_{t=1}^{24} Q(t)]$$

If the wastewater is settled, S_{ti} refers to the measurements on the effluent from the primary settling tank.

settling tank, the size for the first flow (high) will be larger than that for the second flow (low). Thus knowing only the COD load (and its fractional components), and the sludge age, the volume of the plant and the mean carbonaceous oxygen requirements can be estimated rapidly, also, the mass of sludge to be wasted per day. Knowing the mean flow, the size of the settling tanks can be estimated. Once these basic estimates are available, the effects of cyclicity in the flow and load on the peak carbonaceous oxygen demand and peak flow through the settling tank, then can be superimposed.

It was stated above that the mass of sludge, $M(X_v)$, depends on $M(S_{ti})$ and the sludge age. As the sludge age increases so the $M(X_v)$ increases likewise, the mass being roughly proportional to the sludge age; this is illustrated in Figure 3.1 where the mass of sludge in the process per unit influent COD, for unsettled and settled "normal" municipal waste flows, are shown plotted versus sludge age. Consequently the longer the sludge age, the bigger the reactor volume required to deal with the same $M(S_{ti})$. In Figure 3.1 are also shown plotted the carbonaceous oxygen demands per unit COD input for settled and unsettled waste flows. As for $M(X_v)$ the respective magnitudes of $M(O_c)$ do not differ greatly between settled and unsettled wastes. The difference arises principally from the fraction of unbiodegradable particulate COD, f_{up}, in the two wastes being higher in unsettled than in settled waste. However, the total masses of sludge and carbonaceous oxygen demands for the two flows will differ appreciably because approximately 40% of the COD is removed in the primary settling tanks.

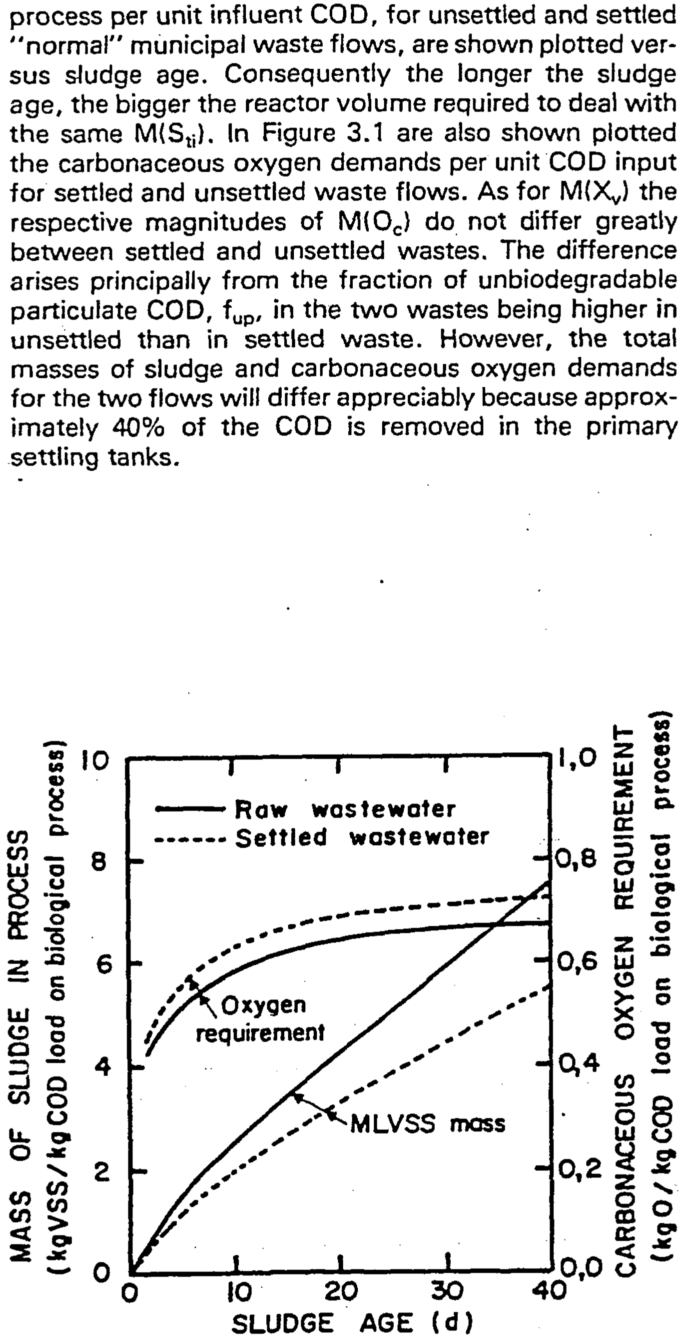

FIGURE 3.1: Mass of volatile solids and mass of carbonaceous oxygen demand per day per kg COD load per day versus sludge age for settled and unsettled municipal waste flows.

3.2 Influent TKN, μ_{nmT} and temperature

In the discussion above, to calculate the sludge mass it was stated that the sludge age must be specified. Selection of the sludge age probably is the most important decision the designer must make, particularly when the objective includes nutrient removal. If removal of carbonaceous material only is required, the sludge age can be quite short; three days will be enough for COD removal – the sludge age selected is dependent more on the expected settleability of the sludge than on the COD removal. If nitrification only is required the sludge age must be sufficiently long to allow the nitrifying organisms to grow *and* metabolize virtually all the available nitrogen. If denitrification is required in addition to nitrification, a fraction of the sludge mass must be kept unaerated to denitrify the nitrate generated. As the nitrifiers are obligate aerobes, this implies that in nitrification-denitrification plants the sludge age must be increased above that where nitrification only is required, to ensure that the aerobic sludge age is still adequate for nitrification.

The maximum specific growth rate constant of the nitrifiers, μ_{nm}, defines the minimum aerobic sludge age for the initiation of nitrification. This minimum sludge age must be increased by a further 25 per cent to ensure that nitrification is near complete. Now μ_{nm} tends to be specific to the waste flow and can range from 0,20 to 0,70/day at 20°C. Also μ_{nm} is very sensitive to temperature, the value halving for every 6°C drop in temperature. The "aerobic" sludge age is given by $1/\mu_{nm}$; it is clear therefore that, in a nitrification plant, a low μ_{nm} value for the waste flow coupled with a low temperature during the winter, will require a long sludge age to ensure efficient nitrification throughout the year. In nitrification-denitrification plants the process sludge age needs to be increased further to accommodate the effects of the unaerated fraction of the sludge mass. These factors in combination can cause that whereas a sludge age of 4 days for a $\mu_{nm} = 0{,}3$ at 20°C will be adequate theoretically for a purely nitrifying aerobic plant, for a nitrifying-denitrifying plant with $\mu_{nm20} = 0{,}3$, $T_{min} = 12°C$ and an unaerated sludge mass fraction of 0,5, the minimum process sludge age will be about 15-20 days to ensure a high efficiency of nitrification.

Provided the sludge age is adequate, the mass of nitrate formed by a process is given by the mass of TKN in the influent minus the mass of nitrogen incorporated in the sludge mass wasted every day. This gives the *nitrification capacity* of the plant (usually expressed as nitrate formed/unit influent flow). The higher the TKN concentration in the influent so, in general, the higher the nitrification capacity – we can form the rough rule that the nitrification capacity is proportional to the TKN concentration in the influent.

3.3 Readily biodegradable COD, slowly biodegradable COD

3.3.1 Nitrogen removal

In the previous section it was mentioned that a fraction of the process sludge mass must be unaerated to denitrify the nitrate generated in the aerobic fraction, but the magnitude of this unaerated fraction was still undetermined. It is the intention now to describe qualitatively how this unaerated fraction is determined and how the influent characteristics affect the determination.

In the absence of oxygen, nitrate serves as an electron acceptor. With oxygen, the process designer has the problem of determining the oxygen demand for known COD load; in denitrification the problem is to determine the COD to denitrify a known mass of nitrate. The COD for denitrification comes from three sources (1) influent readily biodegradable soluble COD, (2) influent slowly biodegradable particulate COD and (3) particulate COD derived from the death of organisms. Usually (2) and (3) are lumped together in formulating the reduction of nitrate.

In the nitrification-denitrification process because the same sludge mass generates the nitrate and denitrifies it (called the single sludge process), the sludge mass normally is subdivided into two reactors in series, the one aerated (to nitrify) and the other unaerated (to denitrify).* Two basic configurations have been developed, the Modified Ludzack-Ettinger (MLE) configuration, Figure 3.2, and the Wuhrmann configuration, Figure 3.3. In the MLE configuration, the first reactor is left unaerated, the second is aerated. The nitrate generated in the second reactor is transferred to the first via the underflow and an inter-reactor (internal) recycle. In the Wuhrmann configuration the nitrate is generated in the first reactor and transferred by the normal serial flow to the second unaerated reactor – an internal recycle is not required, the only recycle being the underflow recycle to transfer the sludge from the secondary settling tank back to the first reactor.

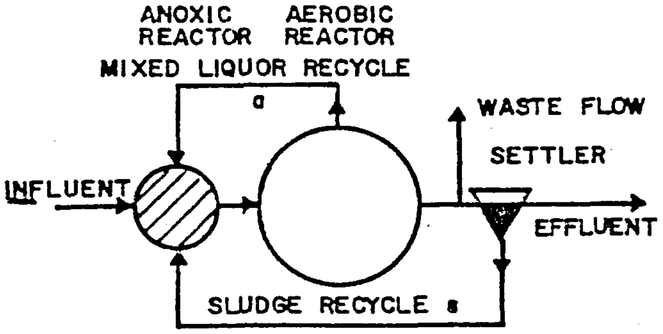

FIGURE 3.2: Modified Ludzack-Ettinger nitrification-denitrification activated sludge process.

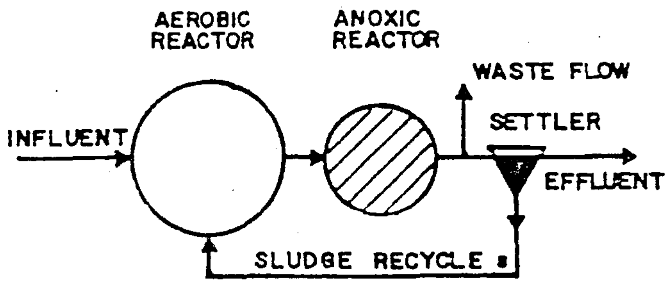

FIGURE 3.3: Wuhrmann nitrification-denitrification activated sludge process.

The denitrification action can best be explained by considering the unaerated reactors as plug flow reactors. (This can be shown to provide valid explanation also for completely mixed reactors). In the unaerated reactor of the MLE configuration (called the primary anoxic reactor), the denitrification takes place at two simultaneously occurring rates, the first due to the influent readily biodegradable COD, the second due to the particulate COD derived principally from the influent, see Figure 3.4a. The first rate is very fast, the reaction being complete in 6 to 20 minutes and it utilizes all the readily biodegradable COD. The second rate is slower, at about 1/7 of the first rate, and continues for the total time in the reactor. In the unaerated reactor of the Wuhrmann configuration, (called the secondary anoxic reactor), a single denitrification rate is observed, due principally to the slowly biodegradable particulate COD derived from death of the organisms, (Figure 3.4b). This source of COD becomes available at a low rate and denitrification rate accordingly is low, about 2/3 that due to particulate COD in the MLE primary anoxic reactor.

The nitrate removal can be expressed either as total mass removed per day or as mass removed per litre of flow, i.e. as a concentration relative to the flow mg$\Delta(NO_3\text{-}N)/\ell$ flow. The latter method is the most convenient because it allows ready comparison with the TKN concentration in the influent flow, mgTKN/ℓ flow. We have stated that the removal is due to the complete utilization of the readily biodegradable COD and the partial utilization of the biodegradable particulate COD. The denitrification achievable due to the latter can be formulated in terms of a denitrification rate constant, K, the *active* mass of sludge generated per litre of influent and the fractional mass of the process sludge contained in the specific anoxic reactor, as follows: (See Chapter 6, Section 4).

In the primary anoxic reactor

ΔN_1 = C × (mass of readily biodegradable COD per litre of influent) + K_2 × (mass of active sludge generated by one litre of influent) × (mass fraction of sludge in the primary anoxic reactor) (3.1)

In the secondary anoxic reactor

ΔN_3 = K_3 × (mass of active sludge generated per litre of influent) × (mass fraction of sludge in the secondary anoxic reactor) (3.2)

The values of K_2 and K_3 have been determined experimentally and found to remain virtually constant for sludge ages between 10 and 30 days, at a fixed

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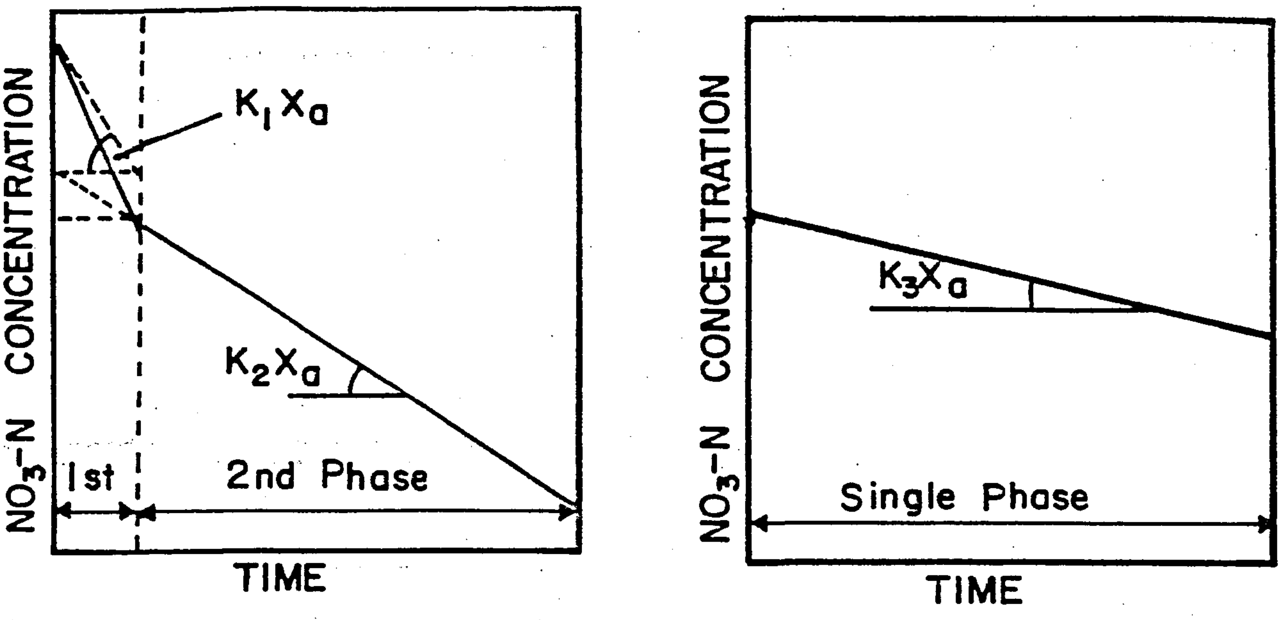
*Other processes, in which the sludge is sequentially aerated and left unaerated, also have been proposed and built, either sequentially in time in a single completely mixed reactor or sequentially in space as, for example, in the Carrousel system.



(a) Primary anoxic

(b) Secondary anoxic

FIGURE 3.4: Denitrification time behaviour in plug flow anoxic reactor, (a) primary anoxic and (b) secondary anoxic reactors.

temperature. This is also indicated by simulation using the general activated sludge model of Dold, Ekama and Marais (1980) as extended to include denitrification by van Haandel, Ekama and Marais (1981). The two equations developed above, expressed in technical terms are for primary and secondary anoxic reactors respectively

$$\Delta N_1 = f_{bs}\,(1 - f_{up} - f_{us})\,S_{ti}/8{,}6 + K_2 \frac{Y_h R_s}{1 + b_h R_s} S_{ti}\,(1 - f_{us} - f_{up}) \frac{R_{hn1}}{R_{hn}} \qquad (3.3)^*$$

$$\Delta N_3 = K_3 \frac{Y_h R_s}{1 + b_h R_s} S_{ti}\,(1 - f_{us} - f_{up}) \frac{R_{hn3}}{R_{hn}} \qquad (3.4)^*$$

R_{hn1}/R_{hn} and R_{hn3}/R_{hn} are the ratios of the nominal retention times of the primary and secondary anoxic reactors respectively with respect to the total. The ratios respectively are equal to the primary and secondary anoxic sludge mass fractions. Here as for aerobic biological degradation, the nominal retention is not the basic parameter even though one may formulate the equations to include it. From Eqs. 3.3 and 3.4 the denitrification potential is a function of the influent COD, S_{ti}, but is approximately so only, being modified by the magnitude of the unbiodegradable fractions of the influent COD, f_{us} and f_{up}.

In Eq. 3.1, all the readily biodegradable COD is utilized; the denitrification attained from this source in the influent is very effective and constitutes about 50% denitrification achieved in "normal" municipal wastewater. The removal due to the particulate biodegradable COD typified by K_2 and K_3 in Eqs. 3.1 and 3.2 *inter alia*, is directly proportional to the anoxic sludge mass fractions of the respective anoxic reactors (i.e. f_{x1} and f_{x3}). These fractions cannot be increased *ad lib*; for a fixed sludge age the magnitude of the sum of these fractions is subject to the proviso that efficient nitrification must be maintained and this in turn is fixed by the aerobic sludge age. In this way there is an upper limit to the sum of anoxic sludge mass fractions which limits the nitrate removal achievable. Furthermore because K_3 is roughly 2/3 of K_2, the nitrate removal due to K_3 in the secondary anoxic reactor per unit volume is only about 2/3 as effective as that in the primary anoxic reactor so that per unit sludge mass fraction the primary anoxic reactor is more efficient than the secondary anoxic reactor.

The nitrate removals, ΔN_1 and ΔN_3, are called the *denitrification potentials of the respective reactors*. The term *potential* is used because in the primary anoxic reactor (as in an MLE process), for example, the denitrification potential can be achieved only if sufficient nitrate, equal to the potential, is recycled to this reactor – if less is recycled the full potential of the reactor is not realized and the *denitrification performance* (or *denitrification capacity*) is less than the denitrification potential. When the performance is less than the potential then the performance can be increased by increasing the recycle. Now the recycle contains dissolved oxygen in addition to nitrate, and the oxygen utilizes the COD preferentially; in consequence, recycle ratios in excess of 4 to 6 usually do not significantly improve the performance, but even can cause a *decrease* in performance due to the high mass of DO transported to the anoxic reactor via the recycle flow. If more nitrate is recycled than that defined by the potential, only the potential value is removed. The excess is returned from the anoxic reactor to the aerobic reactor; in such a situation the recycle can be reduced appropriately without affecting the nitrogen removal by the process.

The combined primary and secondary anoxic sludge mass fraction i.e. the total unaerated sludge

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*These equations are for illustrative purposes only, the meaning of the symbols can be found in Chapters 4 to 6.

mass fraction, cannot be increased *ad lib.*, for two reasons:

(1) Nitrification requires a minimum aerobic sludge age and to preserve this, as the unaerated mass fraction is increased so also the process sludge age must be increased to preserve the minimum aerobic sludge age; increasing the sludge age, however, increases the total mass of sludge and hence the total volume of the reactor system. Furthermore, any increase has a relatively small effect on nitrate removal because the increase contributes to denitrification only through the K_2 or K_3 components and these are relatively inefficient. Practically, for South African conditions, the upper limit to a sludge age appears to be about 30 days.

(2) When the nitrification rate is high the aerobic sludge age needed is small so that with a long sludge age it would be possible to have a large unaerated sludge mass fraction. Experimental data, supported by simulation studies, show that for sludge ages between 20 and 30 days, if the unaerated fraction of a process increases above about 70% at 20°C and about 50% at 12°C the time available for aerobic degradation is insufficient to process all the particulate influent biodegradable COD and the process acquires a degradation pattern still ill understood. Consequently for design under South African conditions for sludge ages of 20 to 30 days, the unaerated mass fraction should not exceed about 55 to 60%.

We have noted earlier that the denitrification potentials are approximately proportional to the influent COD concentration, S_{ti}. Now previously it was pointed out that the nitrification capacity is approximately proportional to the influent TKN. Consequently the TKN/COD ratio of the influent is a good parameter in terms of which to assess process nitrification-denitrification behaviour because it gives approximately the ratio of the nitrate generated to the denitrification potential. Now accepting that

- μ_{nm} will vary around 0,4 per day,
- the sludge age is in the range 20 to 30 days,
- "normal" municipal wastewater flows contain readily biodegradable COD fractions of about 20% with respect to the total influent COD,
- the internal recycle ratio is limited to a maximum of 6:1, and
- the minimum winter temperature is 14°C,

it would appear that for these conditions, if the influent TKN/COD is greater than about 0,10, the denitrification potential is such that it is not possible to denitrify all the nitrate generated even if all the unaerated fraction of the process is located in the primary anoxic reactor. Consequently in this situation the MLE process is the most efficient. If the influent TKN/COD ratio is less than about 0,09 then complete denitrification is possible. Even if complete denitrification is possible this cannot be accomplished by using an MLE process only for the following reason:

Say the steady state nitrate concentration in the aerobic reactor is N mg/ℓ; if an internal recycle ratio of a is imposed and an underflow recycle ratio of s then $(a+s)/(a+s+1)$ of this nitrate is recycled and $1/(a+s+1)$ goes out with the effluent. To deal with the $1/(a+s+1)$ fraction a secondary anoxic reactor needs to be incorporated also in the process, to give the Bardenpho process first proposed by Barnard, 1973 (Figure 3.5). Of course the secondary anoxic reactor per unit volume is not as efficient as the primary because $K_3 < K_2$ so that the denitrification potential for the same unaerated total mass fraction will be lower than for an MLE process only, but this must be sacrificed and is the reason why complete denitrification is achievable only for a TKN/COD $\leqq$ 0,09 instead of 0,10. This significant reduction in the upper limit of the TKN/COD is principally because the a-recycle should not be increased above $a=6$ (to limit the feedback of oxygen to the primary anoxic reactor). If the oxygen in the recycle could have been reduced to zero then very high a-recycles could be utilized and it might not be necessary to employ a secondary anoxic reactor.

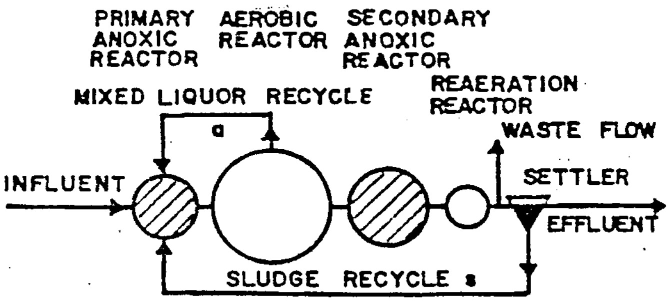

FIGURE 3.5: Bardenpho process for biological nitrogen removal – a combination of the modified Ludzack-Ettinger and Wuhrmann processes.

As the TKN/COD ratio decreases below 0,09 so the size of the primary anoxic reactor may be decreased and the second anoxic reactor increased in order to maintain a low recycle, illustrated in Figure 3.6.

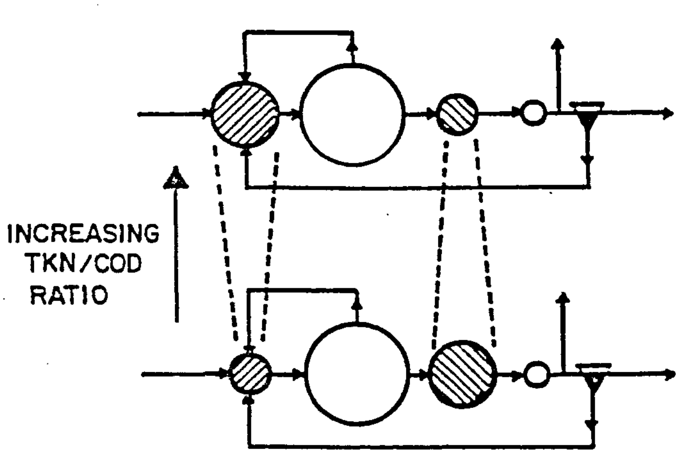

FIGURE 3.6: Effect of TKN/COD ratio on relative sizes of primary and secondary anoxic reactors in the Bardenpho process.

3.3.2 Excess phosphorus removal

There is considerable support for the view that under certain conditions some species of organisms, for example, the *Acinetobacter spp*, can take up phosphorus in excess of its metabolic requirements. Some of the acinetobacter species are obligate aerobes, others can utilize nitrate as electron acceptor. The species is slow growing with rather specific requirements to the type of nutrient it can utilize; it can utilize, for example, many of the lower fatty acids, but not sugars, such as glucose. Because it is slow growing, in aerobic systems, with mixed cultures of organisms, its growth is suppressed by faster growing organisms, because the faster growing organisms utilize most of the substrate and eventually dominates over the slow growing ones. This is the reason why it is not easy to find evidence of acinetobacter in purely aerobic processes.

From the discussion above, how then does it happen that the poly-P organism proliferates in some systems containing unaerated fractions? Assume that the organism already contains some excess phosphate. This phosphate is present as the radical PO_4^- but bound together to form polyphosphate chains. The bonds between the radicals in the chain contain appreciable energy. If the acinetobacter is discharged to an anaerobic reactor (i.e. a reactor that has neither nitrate nor oxygen present), in which substrate is present suitable for growth of *Acinetobacter spp*, it breaks down some of the bonds in the polyphosphate chain, releasing phosphate radicals to the surrounding liquid. The bond energy thus released is utilized to absorb, complex and store the substrate in the organism – we say it sequesters the substrate. In contrast, in the anaerobic zone, because no external electron acceptor (oxygen or nitrate) is available, the non poly-P organisms cannot utilize the substrate. On leaving the anaerobic zone and entering an anoxic zone (with nitrate present) or an aerobic zone (with oxygen present), the poly-P organisms now has electron acceptors available and metabolizes the stored substrate sequestered for itself in the anaerobic zone i.e. it can grow without being in competition with the non poly-P organisms for the balance of the energy still available in the surrounding medium.

In the anoxic and aerobic zones the poly-P organisms, beside using the sequestered substrate for growth, also use it to take up phosphate radicals (PO_4^-) from the surrounding medium, again to form phosphate chains for use subsequently when returned to the anaerobic zone. Thus, by having the propensity to store polyphosphate the acinetobacter and other poly-P organisms have a positive advantage over the non poly-P organisms to obtain energy for themselves in the apparently inhospitable anaerobic reactor. This explains the apparently anomolous growth response of the poly-P organisms.

The behaviour of the poly-P organisms in the anaerobic reactor is not yet well understood but a summary of the behaviour appears to be as follows: When a substrate is added that can be utilized *directly* by the poly-P organisms, such as acetate or butyric acid, the release of P (to obtain energy for sequestering the substrate) is very rapid, and the release is proportional to the mass of substrate added. Also, the net removal of P from the anaerobic-anoxic-aerobic system increases as the substrate added increases, but insufficient data has been collected to express these responses quantitatively.

When sewage is added to the anaerobic reactor the behaviour differs in two aspects from that when acetate is added: (1) only a fraction of the COD added is implicated in the P release and P removal; (2) the rate of P release is significantly slower with sewage than with acetate addition and the release is of a different kinetic order.

(1) The influent COD that induces P release seems to be associated with the readily biodegradable COD fraction. It has been observed consistently in laboratory scale processes that the release and removal of P increases and decreases as this fraction increases and decreases (see Figure 3.7) although it is not yet clear, and highly unlikely, that all the readily biodegradale COD is implicated in this action.

    With regard to the particulate biodegradable COD, this fraction appears to have little effect on P release. This is not an unreasonable conclusion because the particulate COD requires adsorption, storage and extracellular breakdown prior to absorption by the organisms; even under aerobic conditions the rate of metabolism of the particulate COD is only at 1/7 to 1/10 the rate at which readily biodegradable COD is metabolized, Dold, Ekama and Marais (1981).

(2) The slower rate of P release with sewage addition compared to that with acetate addition seems to indicate that the substrate suitable for P release becomes available indirectly to the poly-P organisms. This is further supported by the observation that whereas with acetate addition the P release appears to be zero order with respect to acetate concentration, (Figure 3.8), with sewage addition the release appears to conform to a first order reaction; in a plug flow anaerobic reactor receiving sewage the rate of release is high initially but dies away steadily along the reactor until it is virtually zero. This is also indicated in a series system of anaerobic reactors (Figure 3.9). The first order nature of the reaction in the series system is verified by plotting the log (release/reactor) versus the cumulative retention time, to give a linear plot, (Figure 3.10). Comparison between the release in a plug flow reactor and a completely mixed reactor also shows that for the same anaerobic mass fraction if the mass fraction is so small that all the P is not released in the plug flow reactor, the release in the completely mixed reactor is less than in the plug flow one, a behavioural pattern in conformity to that expected where a first order reaction controls.

According to the observed behaviour described above, at sludge ages about 20 to 25 days, the anaerobic mass fraction of a completely mixed reactor should be about 0,2 to ensure that 90% or

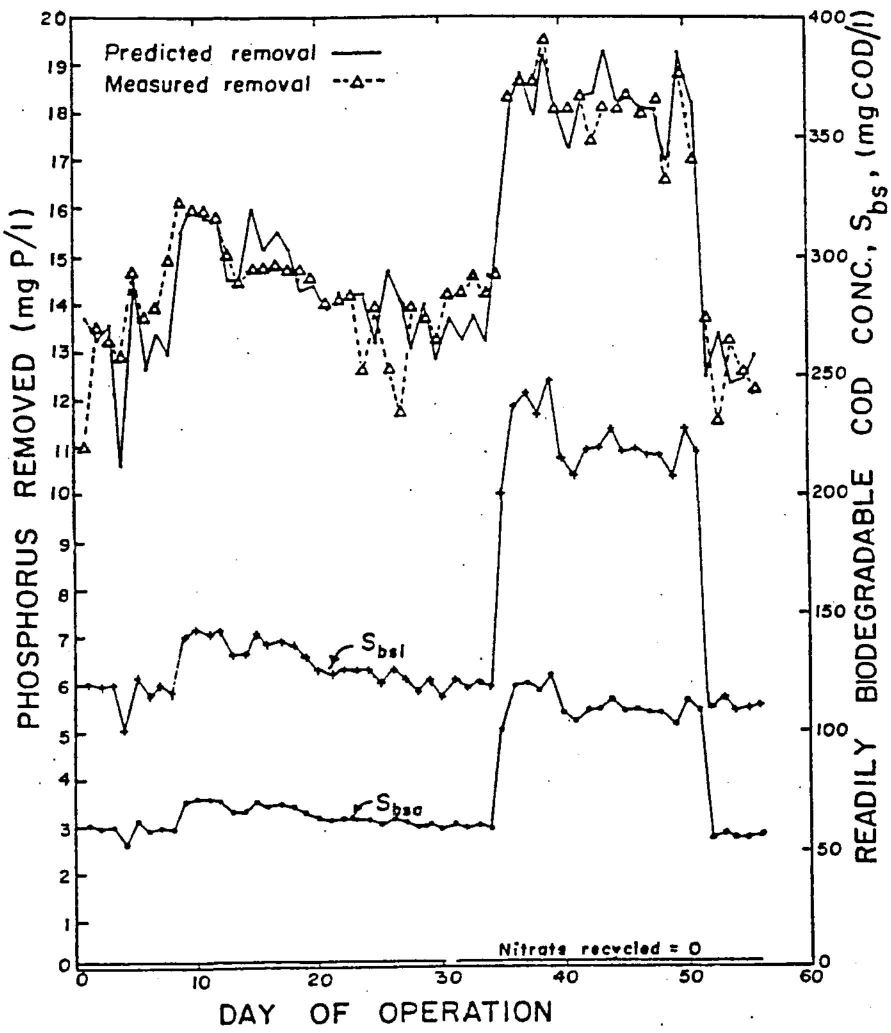

FIGURE 3.7: Experimental data showing the strong dependence of excess P removal (ΔP) on readily biodegradable COD in the influent (S_{bsi}).

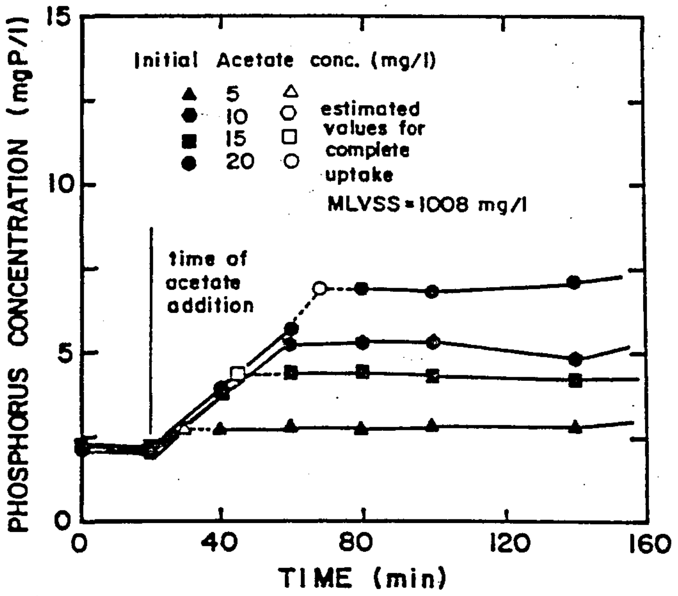

FIGURE 3.8: Phosphorus release in a batch anaerobic reactor with different concentrations of acetate added. Note the magnitude of the release is proportional to mass of acetate added but the rate of release is independent of the acetate concentration.

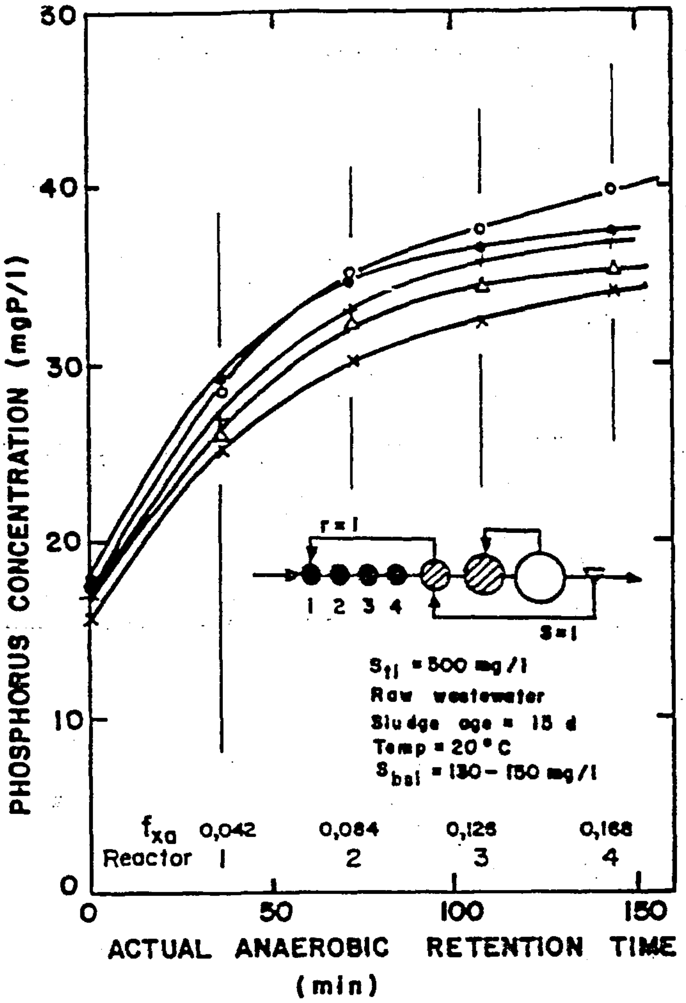

FIGURE 3.9: Phosphorus concentration in a four in series anaerobic reactor system, in a modified UCT process, receiving unsettled municipal sewage influent.

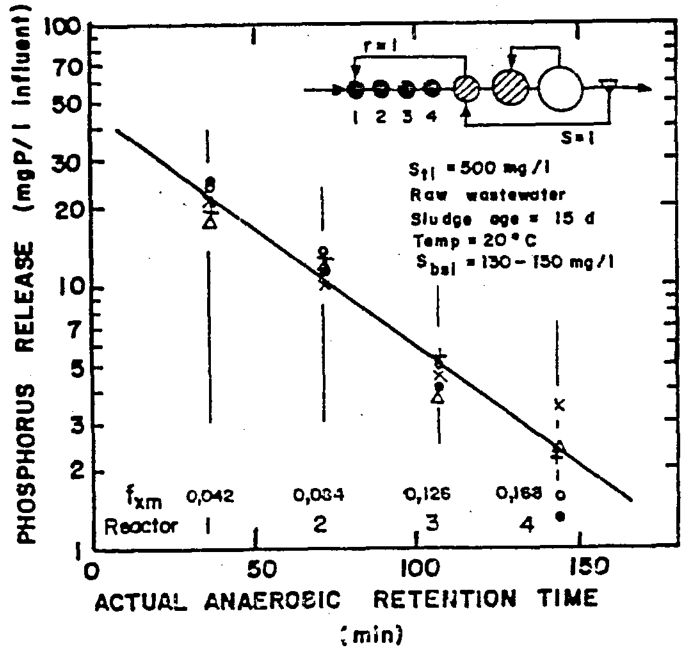

FIGURE 3.10: Phosphorus release per reactor in a four in series anaerobic reactor system in a modified UCT process receiving unsettled municipal sewage influent (see also Figure 3.9)

more of the P is released; if it were a plug flow reactor the mass fraction needs to be about 0,15 to give the same percentage release. Because the release rate is first order there is little to be gained by making the anaerobic mass fraction greater than about 20%. Mass fraction of 10% or less again will be too small allowing only about 50 to 60% of the potential release to take place. Then again, subdividing the anaerobic reactor into say two reactors in series gives improved usage of the anaerobic reactor mass fraction and requires a mass fraction of about 15% to give the same release as a completely mixed reactor of 20%. These fractions apply only for sludge ages of 20 days and longer.

The reasons for the different responses observed with acetate and sewage readily biodegradable COD additions are not yet clear but the following explanation could be valid: Most of the fractions of readily biodegradable COD, apparently, are not in a chemical form suitable for utilization by the poly-P organisms. In the anaerobic zone the non poly-P organisms, for *survival*, because no external electron acceptors are available, partially utilize the readily biodegradable COD, via the Embden-Meyerhof or other equivalent pathways, to generate a small amount of energy, by creating internal or endogenous electron acceptors: The outcome of this action is a breakdown of the original COD molecules to lower fatty acids and other similar forms. These are rejected by the organism and become available to the poly-P organisms for sequestration. Thus, where substrate is present in a form unavailable to the poly-P organism, the presence of non poly-P organisms can modify some of the COD fractions to forms suitable for use by the poly-P organisms. The kinetics of this transformation will govern the kinetics of sequestration by the poly-P organisms because the COD breakdown is the rate limiting step. The non poly-P organisms are, in terms of this hypothesis, essential to the growth of the poly-P ones.

3.3.3 Implication on process development

The rationalization of the poly-P behaviour leads to a number of conclusions that are of significance to designers of P removal processes:

(1) In terms of the hypothesis set out above, the generation of the lower fatty acids by acid digestion of primary settling tank underflow, as proposed and implemented by the Municipality of Johannesburg, makes good sense and merits serious research attention.

(2) Any external electron acceptor entering the anaerobic reactor, like nitrate or oxygen, will be utilized immediately by the non poly-P organisms acting on the readily biodegradable COD and thereby reduce the COD concentration available for the generation of substrate suitable for sequestration by the poly-P organisms.

(3) The fact that the whole sequestration action is slow means that adequate anaerobic mass must be

made available to process as much as possible of the readily biodegradable COD source. However, because the reaction is first order it also implies that there is little merit in enlarging the anaerobic mass *ad lib.*, each increase giving a decreasing return on the previous increase. Also, series anaerobic reactors are more efficient and hence smaller anaerobic mass is necesary, to achieve the same effect than with a single completely mixed anaerobic reactor.

With regard to design implications, the basic requirement in a design is to provide an anaerobic zone which receives the influent containing the readily biodegradable COD fraction and so becomes available, directly or indirectly, to the poly-P organisms for sequestration. This zone must be large enough to allow the sequestration action to be as near complete as possible in order to obtain maximum effect. Furthermore, the zone should not receive any form of the external electron acceptor (nitrate, nitrite or dissolved oxygen) as these cause the fast growing non poly-P organisms to utilize the substrate before it can be sequestered by the poly-P organisms.

The first explicit design of a configuration, in which an anaerobic reactor is proposed, was due to Barnard* (1976). He accepted the Bardenpho process (Figure 3.5) as the basic configuration to nitrify and denitrify. A fraction of the primary anoxic reactor was isolated to serve as an anaerobic reactor. This reactor was placed at the head of the process and received the underflow recycle and the influent flow (see Figure 3.11). The reasoning behind this configuration was that the Bardenpho section of the process would give complete nitrification and denitrification so that the underflow would contain little or no nitrate, in this fashion allowing optimal conditions in the anaerobic reactor and maximum release of P. The process configuration has been variously called the Modified Bardenpho, Phoredox or 5 stage Phoredox process.

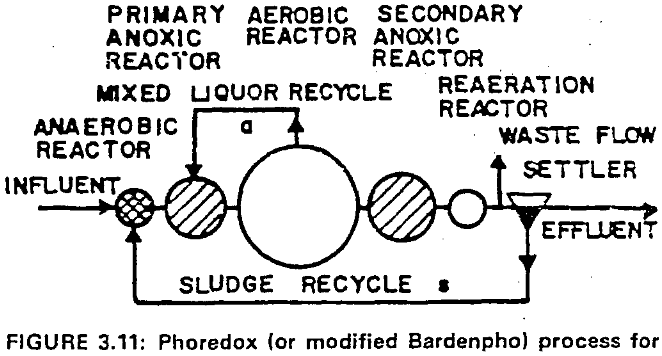

FIGURE 3.11: Phoredox (or modified Bardenpho) process for nitrification-denitrification-biological excess phosphorus removal.

In assessing the functioning of the Phoredox process, the reservation of a portion of the unaerated mass fraction to the anaerobic reactor concomitantly reduces the denitrification potential of the process. Thus, compared to the Bardenpho process for the same total unaerated mass fraction, the Phoredox has a lower denitrification potential. Consequently the maximum TKN/COD ratio the Phoredox process can denitrify completely is reduced. For a "standard" raw sewage influent, i.e. 20% of the influent COD is readily biodegradable, at sludge ages of approximately 20 days, maximum unaerated mass fraction of 0,55 and an anaerobic mass fraction of 0,15, it would seem that the maximum TKN/COD ratio for complete denitrification should not exceed 0,08. For settled sewage the readily biodegradable COD is higher ≅ 25% of the total influent COD and the maximum TKN/COD ratio is about 0,09. This ratio is very dependent on the fraction of readily biodegradable COD because this fraction contributes significantly to the denitrification. Consequently the values given here are relative ones only – it is important to estimate the true readily biodegradable fraction experimentally.

As the TKN/COD ratio increases from low values, so it will be found that the relative volumes of the primary and secondary anoxic reactors change accordingly. To achieve optimum denitrification response at low TKN/COD ratios the primary reactor is small and the secondary reactor large; at high TKN/COD ratios, near or at the maximum allowed, the primary reactor is large and the secondary very small.

Should the TKN/COD ratio exceed the maximum, nitrate will appear in the effluent. In the Phoredox process this nitrate is recycled in the underflow to the anaerobic reactor where it utilizes preferentially some, or all (depending on the concentration of nitrate) of the readily biodegradable COD, thereby reducing the magnitude of the COD available for sequestration by the poly-P organisms. Now the concentration of nitrate in the underflow recycle is the same as that in the effluent and 1 mg (NO_3-N) removes 8,6 mg readily biodegradable COD, consequently, a concentration of 5 mg/ℓ (NO_3-N) in the underflow with a recycle ratio of 1:1 will remove 43 mg/ℓ of this COD fraction which, for an influent of 500 COD/ℓ will reduce the readily biodegradable COD from approximately 100 to 60 mg/ℓ. The only operational means for reducing the mass of nitrate transferred to the anaerobic reactor is to reduce the underflow recycle ratio but this procedure has only limited possibilities because low recycle ratios may give rise to difficulties with the secondary settling tank – at low recycles the sludge mass builds up in the tank and may cause failure of the tank when the SVI is high, or loss of sludge in the effluent may occur from rising sludge due to denitrification in the sludge mass.

The principal difficulty with the Phoredox configuration, under high TKN/COD ratios, arises from the fact that the anaerobic reactor performance is dependent on the nitrate in the effluent. To make the anaerobic reactor independent of the effluent nitrate, the University of Cape Town (UCT) process configuration was developed, Figure (3.12). The basic denitrification configuration is the Modified Ludzack-Ettinger process (Figure 3.2). The anaerobic reactor is, like the

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*The basis for proposing the anaerobic reactor, by Barnard, differs fundamentally from the hypothesis suggested in the previous sections. He suggested that the *redox potential* need to be lowered sufficiently to induce P release. He noted the adverse effect on P removal by nitrate discharged to the anaerobic reactor but suggested that this arose from the nitrate limiting the depression of the redox potential. These aspects are dealt with in greater detail in Chapter 7.

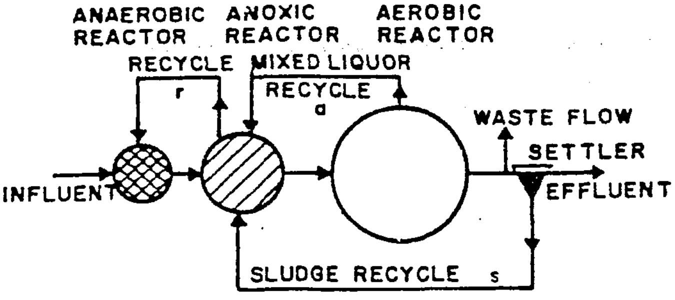

FIGURE 3.12: University of Cape Town (UCT) process for nitrification-denitrification-biological excess phosphorus removal.

Phoredox, placed at the head of the works but the recycle flow to this reactor, the r-recycle, is taken from the primary anoxic reactor. By regulating the main internal recycle from the aerobic to the anoxic reactor, the a-recycle, so that the denitrification potential of the primary anoxic reactor is not exceeded, a nitrate-free effluent is assured to the anaerobic reactor.

In the UCT process, if, say, an r-recycle of 1:1 is used the concentration in the anaerobic reactor is half that in the rest of the plant and to get the required anaerobic mass fraction the volume of this reactor must, in consequence, be increased to twice the volumetric size of that in the Phoredox for the same anaerobic mass fraction. The main advantage of the UCT process is that by operational means the anaerobic reactor can be made independent of the nitrate in the effluent, and this can be maintained up to a TKN/COD ratio of about 0,14. As an alternative to the Phoredox process, at low TKN/COD ratios, i.e. TKN/COD < 0,08, the UCT process is also satisfactory although the nitrate removal will not be as high as in the Phoredox as it is not possible ever to get zero nitrate in the effluent from a UCT process.

To overcome the need for explicit control of the a-recycle, the Modified UCT process was developed, (Figure 3.13). In this process the primary anoxic reactor is divided into two fractions, the first having an anoxic mass fraction of about 0,1. The a-recycle is discharged to the second anoxic reactor and remains fixed at say 4:1 ratio irrespective of whether nitrate is recycled back to the aerobic reactor or not, provided the a-recycle ratio is so large that it brings enough nitrate to the second anoxic reactor to equal or exceed its denitrification potential. The underflow recycle (s-recycle) discharges to the first anoxic reactor, to reduce the nitrate to zero. By means of this configuration zero nitrate in the r-recycle can be maintained for TKN/COD ratios up to about 0,11.

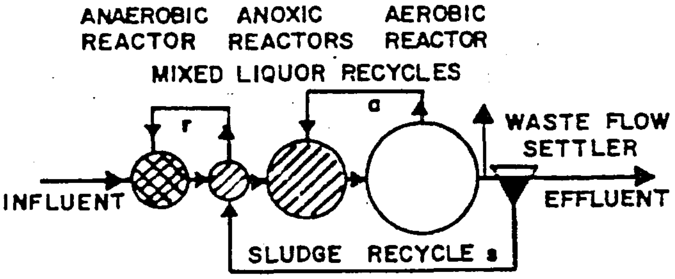

FIGURE 3.13: Modified UCT process.

4. CLOSURE

There are numerous aspects and details in the design of nutrient removal plants that have not been dealt with in this short review – attention was focussed on the major aspects of a fundamental nature only. For details of design reference needs to be made to Chapters 4 to 7. Designs that conform to the recommendations laid down in these chapters, for treating normal municipal wasteflows, settled or unsettled, will have a high expectation of phosphorus removal of about 1,7 to 2 mg PO_4-P per 100 mg influent COD, i.e. a ΔP/COD of 0,017 to 0,02 mgP/mg influent COD. This removal ratio can vary, depending principally on the fraction of readily biodegradable COD in the influent. Should the influent P/COD concentration ratio exceed 0,017 to 0,02, very likely all the phosphorus will *not* be removed biologically from the wastewater and provision necessarily will have to be made for chemical precipitation of the remaining phosphorus; should the influent P/COD be *less* than 0,017 to 0,02, then very likely removals down to 0,5 mgP/*l* can be achieved for a high proportion of the time.

The above biological P removal propensity in ΔP/COD forms an approximate basis for judging the performance of any designed plant. *The plant should not be assessed on the basis of the effluent P concentration only*, but on the ΔP per unit influent COD. If the ΔP/COD achieved is less than 0,017 to 0,02, the cause for the lower removal should be sought in the wastewater characteristics, process design or operation of the plant, or in combinations of these three. The results of such investigations would be of great value to research workers and designers in the endeavour to build plants that perform optimally.

5. REFERENCES

DOLD, P.L., EKAMA, G.A. and MARAIS, G.v.R. (1980) A general model for the activated sludge process, *Prog. Wat. Tech.*, **12**, 47-77.

VAN HAANDEL, A.C., EKAMA, G.A. and MARAIS, G.v.R. (1981) The activated sludge process part 3 – Single sludge denitrification, *Water Research*, **15**, 1135-1152.

CHAPTER FOUR

CARBONACEOUS MATERIAL REMOVAL

by

G.A. Ekama and G.v.R. Marais

1. INTRODUCTION

Basically all aerobic biological treatment processes operate on the same principles: the various treatment processes i.e. trickling filters, aerated lagoons, contact-stabilization, extended aeration, etc. differ only in the conditions under which the biological reactions are constrained to operate. The activated sludge process comprises the flow régime in the reactor, the sizes, number and configuration of the reactors, recycle flows, influent flow and other features incorporated either deliberately, or present inadvertently or unavoidably. Whereas the response of the organisms is in accordance with their nature, that of the process is governed by both the organism characteristics and the physical features that define the process.

The ecology of the suspended sludge mass in the reactors of the activated sludge processes is complex comprising numerous species of bacteria and many species of higher organisms that prey on the bacteria. Despite the complex nature of the ecosystem, it would appear that the kinetic behaviour of the process, insofar as it reflects the energy removal from the wastewater, can be modelled as if the system were governed by an equivalent or surrogate bacterial mass, having characteristics derived in part from the observed response of the treatment processes. It must be emphasised that this surrogate mass has characteristics very different from those of a pure bacterial culture. This point is mentioned because criticism of this approach has often hinged on the incompatibility of the behaviour of the equivalent bacterial mass with that observed in a pure bacterial culture.

In this chapter the basic kinetics of the carbonaceous material removal, and its incorporation in the process equations are briefly stated, and application to design is set out in detail. The design equations presented are for the steady state completely mixed single reactor activated sludge process. The steady state solution is of prime importance because (1) it forms the basis for preliminary design of any activated sludge process be it aerobic, anoxic-aerobic or anaerobic-anoxic-aerobic; (2) the relationships also are useful in indicating operational control procedures for the process.

If certain minimum conditions are imposed on the process, nitrification will occur. With nitrification the oxygen demand can show a considerable increase, of up to 40% more than that required for carbonaceous material oxidation. Nitrification is not dealt with in this chapter but is considered in detail in Chapter 5.

2. BIOLOGICAL KINETICS

When a heterotrophic organism population under aerobic conditions is brought into contact with biodegradable organic material* consisting of soluble readily biodegradable and particulate slowly biodegradable COD fractions (see Chapter 2), its response may be described qualitatively as follows:

(1) The soluble readily biodegradable COD passes directly through the cell wall and is metabolized at a high rate.

(2) The particulate slowly biodegradable COD is adsorbed onto the organisms and storage of COD takes place. This reaction is very rapid and effectively removes all the particulate and colloidal COD from the wastewater. The stored COD is broken down by extracellular enzymes and transferred through the cell wall and metabolized in the same manner as the readily biodegradable COD fraction in (1) above. The rate of enzymatic breakdown is relatively slow and constitutes the limiting rate in the overall synthesis reaction, at about one tenth of the rate for the readily biodegradable COD.

(3) A fraction of the COD metabolized is converted into new cell material, the balance is utilized to generate energy to bring about the conversion and eventually, lost as heat. The energy is generated by supplying an electron acceptor, oxygen, so that the oxygen utilized is directly related to the COD lost. There is strong experimental evidence that the fraction of COD converted to metabolic material is constant with respect to the COD utilized, Payne (1971). This fraction is called the specific growth yield coefficient, Y_{COD} (mg COD/mg COD). Usually the coefficient is expressed in terms of the volatile solids formed per unit of COD utilized, Y_h, (mg VSS/mg COD) and is related to Y_{COD} by Eq. 1.6, i.e. $Y_h = Y_{COD}/f_{cv}$.

(4) Concomitantly with (3) above, but distinct from it, there is a *net* loss of live mass called endogenous mass loss. Not all the live mass that disappears *per se* is lost as energy, a fraction remains (17 to 20 per cent, McKinney and Symons, 1964) as unbiodegradable organic residue, called endogenous residue. The mass of oxygen to be supplied for en-

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*Also called substrate or biodegradable COD.

dogenous mass loss is proportional to the volatile mass that disappears from the system. This *net* behaviour is believed to be the outcome of a more complex set of mechanisms involving the growth of primary organisms (e.g. bacteria) and predators on these organisms (e.g. protozoa). Dold, Ekama and Marais (1980) proposed and formulated the net behaviour as a uniform organism mass having a death-regeneration component, an approach which approximates the prey-predator system under certain circumstances.

Using the death-regeneration approach Dold *et al* (1980) in their development of the general activated sludge process model, and van Haandel and Marais (1981) in their extension of this model to denitrification, have shown that the response of all the known process types (including nutrient removal processes) can be simulated very closely, even under cyclic flow and load conditions. Under constant flow and load, the death-regeneration approach reduces to the endogenous mass loss approach. This latter approach leads to relatively simple process equations (Marais and Ekama, 1976) and constitutes the basic design equations in this monograph, see Eqs. 4.6 to 4.19.

3. PROCESS KINETICS

3.1 Mixing régimes

In the activated sludge process, the mixing régime in the reactor and the sludge return influence the response of the process – therefore consideration must be given to reactor kinetics.

There are two extremes of mixing; completely mixed and plug flow.

In the completely mixed régime the influent is instantaneously and thoroughly mixed (theoretically) with the reactor contents. Hence the effluent flow from the reactor has the same constitution as the reactor contents. The reactor effluent flow passes to a settling tank; the overflow from the tank is the stabilized waste stream, the underflow is concentrated mixed liquor and recycled back to the reactor.

In the completely mixed system the rate of return of the underflow has no effect on the reactor process except if an undue sludge build-up occurs in the settling tank. The shape of the reactor is approximately square or circular in plan, and mixing is usually by mechanical aerators but can also be effected by bubble aeration. Examples are extended aeration plants, aerated lagoons, Pasveer ditch and completely mixed activated sludge plants.

In a plug flow régime, the reactor usually is a long channel type basin. The influent is introduced at one end of the channel, flows along the channel axis and is mixed by air spargers set along one side of the channel. Theoretically each volume element of liquid along the axis is assumed to remain unmixed with the elements leading and following. Discharge to the settling tank takes place at the other end of the channel. To innoculate the influent waste flow with organisms, the underflow from the settling tank is returned to the influent end of the channel. This creates an intermediate flow régime deviating from true plug flow conditions depending on the magnitude of the recycled underflow. Conventional activated sludge plants are of the intermediate flow régime type with recycle ratio varying from 0,25 to 3 times the influent flow rate. If the recycle rate is raised very high, the mixing régime approaches that of complete mixing.

Intermediate flow régimes are also achieved by having completely mixed reactors in series, or by step-aeration. In the latter, the influent is fed at a series of points along the axis of the plug flow type reactor. Both configurations require, for inoculation purposes, recycling of the sludge from the settler to the influent point.

The mean kinetic response of an activated sludge process, i.e. sludge mass, daily sludge production, daily oxygen demand and effluent quality is adequately, indeed accurately, given by assuming the process is completely mixed and the influent flow and load are constant. This allows the plant volume, the mass of sludge wasted daily and mean daily oxygen demand to be determined by relatively simple formulations. Peak oxygen demands which arise under cyclic flow and load conditions can be estimated subsequently quite accurately by applying an adjustment to the mean daily demand, so also for other parameters. These adjustments have been developed from simulation studies, using the general model of Dold, Ekama and Marais (1980), on processes operated under cyclic and under constant flow and load conditions. For details see Chapter 6, Section 7.6.2.

3.2 Sludge age

A diagrammatic sketch of the completed mixed process with sludge recycle is given in Figure 4.1. This sketch differs from that usually presented in that the sludge wasted each day is abstracted directly from the reactor – the common practice is that the waste sludge is abstracted from the secondary settler underflow. Sludge abstraction directly from the reactor leads to a method of control of the sludge age, called the *hydraulic control of sludge age*, which has significant advantages compared to control by wastage via the underflow, see Section 10.

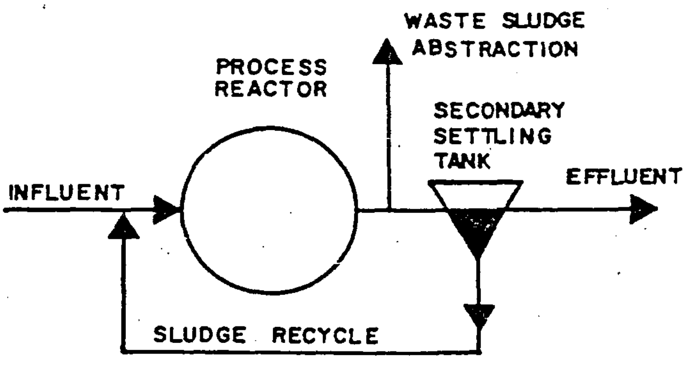

FIGURE 4.1: A diagrammatic representation of the completely mixed activated sludge process with hydraulic control of sludge age and sludge recycle.

The sludge age, R_s, in days, is defined by:

$$R_s = \frac{\text{mass of sludge in reactor}}{\text{mass of sludge wasted per day}} \text{ (days)} \qquad (4.1)$$

By abstracting the sludge directly from the reactor, the wasted liquor and reactor liquor concentrations are the same. If a sludge age of, say, 10 days is required, one tenth of the *volume of the reactor* must be wasted every day.* This can be achieved by a constant draw-off at flow rate, q, per day where q.1 day = v, the volume to be wasted,
hence

$$R_s = \frac{XV_p}{Xq} = \frac{V_p}{q} \qquad (4.2)$$

where

V_p = volume of the process reactor (ℓ)

3.3 Nominal hydraulic retention time

In activated sludge theory the volume of the process per unit of volume of influent flow is known as the nominal process hydraulic retention time i.e.

$$R_{hn} = V_p/Q \qquad (4.3)$$

where

R_{hn} = average nominal process hydraulic retention time (d)
Q = daily average influent flow rate (ℓ/d)

3.4 Effluent COD concentration

Under normal activated sludge process operating conditions in South Africa, where the sludge ages are in excess of 10 days (to ensure nitrification *and* nutrient removal) the nature of the influent COD in municipal wastewaters is such that the COD concentration in the effluent is inconsequential in the process design – the soluble readily biodegradable COD fraction is completely utilized in a very short period of time (less than 1 hour) and the particulate COD, whether biodegradable or unbiodegradable, is adsorbed or enmeshed in the sludge flocs, and settles out with the sludge in the secondary settling tanks. Consequently, *the effluent COD concentration is comprised virtually wholly of the soluble unbiodegradable COD* (from the influent) plus the COD of the sludge particles which escape with the effluent due to imperfect operation of the secondary settling tank (see Chapter 3, Section 3.3). Hence the effluent COD concentration, S_{te}, is approximately given by

$$S_{te} = S_{us} \qquad (4.4a)$$

for filtered samples, or,

$$S_{te} = S_{us} + f_{cv}.X_{ve} \qquad (4.4b)$$

for unfiltered samples, where

S_{us} = unbiodegradable COD in the effluent
= $S_{usi} = f_{us}\, S_{ti}$ (mgCOD/ℓ) (see Eq. 2.3)
X_{ve} = volatile solids concentration in the effluent (mgVSS/ℓ)
f_{cv} = COD/VSS ratio of the volatile solids
= 1,48 mgCOD/mgVSS.

4. PROCESS DESIGN EQUATIONS

Once it is recognized that all the COD in the influent, except the soluble unbiodegradable COD, is either utilized by the micro-organisms to form micro-organism mass, or remains in the process and accumulates as inert sludge mass, it follows that the mass of sludge produced and the carbonaceous oxygen demand in the process are functions of the mass of COD to be treated daily; the greater the daily COD mass load, the greater the sludge production and carbonaceous oxygen demand.

The equations** given below give the masses of sludge generated in the process and the oxygen demand for COD removal as a function of the total COD load, the wastewater characteristics i.e. the unbiodegradable soluble and particulate COD fraction (f_{us} and f_{up}) and the sludge age. The constants in the equations i.e. the specific yield coefficient Y_h, the specific endogenous mass loss rate b_h, the unbiodegradable fraction of the active mass f, and the COD/VSS ratio of the sludge f_{cv}, as well as their temperature dependencies are given in Table 4.1.

TABLE 4.1 KINETIC CONSTANTS AND THEIR TEMPERATURE DEPENDENCY FOR THE STEADY STATE CARBONACOUS DEGRADATION ACTIVATED SLUDGE MODEL (*ex* MARAIS AND EKAMA, 1976)

Constant	Symbol	Temperature Dependency Equation	θ	Standard Value (20°C)	Eq. No.
Yield coefficient (mg VSS/mg COD)	Y_h	remains constant	1,000	0,45	
Endogenous respiration rate (/d)	b_h	$b_{hT} = b_{h20}\theta^{(T-20)}$	1,029	0,24	(4,5)
Endogenous residue (mg VSS/mg VSS)	f	remains constant	—	0,20	
COD/VSS ratio (mg COD/mg VSS)	f_{cv}	remains constant	—	1,48	

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*This assumes that the mass of sludge in the secondary settling tanks is negligible with respect to that in the process reactors. This assumption is valid when the process is operated at high underflow recycle ratios (≅1:1) and when the sludge age is longer than about 3 days, (see Section 10 below).

**The derivation of these equations is given by Marais and Ekama (1976) and a list of symbols used in the equations is given in Appendix 1.

In the equations below, the prefix M denotes the mass of the parameter which follows in brackets.

(1) Influent

$$M(S_{ti}) = Q.S_{ti} \qquad \text{mg COD/d} \qquad (4.6)$$

$$M(S_{bi}) = Q.S_{bi} \qquad \text{mg COD/d} \qquad (4.7a)$$

$$= Q.S_{ti}(1-f_{us}-f_{up}) \qquad \text{mg COD/d} \qquad (4.7b)$$

$$= M(S_{ti})(1-f_{us}-f_{up}) \qquad \text{mg COD/d} \qquad (4.7c)$$

$$M(X_{ii}) = Q.X_{ii} \qquad \text{mg VSS/d} \qquad (4.8a)$$

$$= Q.f_{up}\,S_{ti}/f_{cv} \qquad \text{mg VSS/d} \qquad (4.8b)$$

$$= M(S_{ti})\,f_{up}/f_{cv} \qquad \text{mg VSS/d} \qquad (4.8c)$$

(2) Process

$$M(X_a) = V_pX_a \qquad \text{mg VSS} \qquad (4.9a)$$

$$M(X_e) = V_pX_e \qquad \text{mg VSS} \qquad (4.9b)$$

$$M(X_i) = V_pX_i \qquad \text{mg VSS} \qquad (4.9c)$$

$$M(X_v) = V_pX_v \qquad \text{mg VSS} \qquad (4.9d)$$

$$M(O_c) = V_pO_c \qquad \text{mg O/d} \qquad (4.9e)$$

(a) The Active Volatile Solids Mass (mg VSS)

$$M(X_a) = M(S_{bi})\,Y_hR_s/(1+b_hR_s) \quad \text{(mg)}$$

$$= (1-f_{us}-f_{up})\,M(S_{ti})\,Y_hR_s/(1+b_hR_s) \qquad (4.10)$$

(b) The Endogenous Residue Volatile Solids Mass (mg VSS)

$$M(X_e) = f\,b_hR_s\,M(X_a) \qquad \text{(mg)} \qquad (4.11)$$

(c) The Inert Volatile Solids Mass (mg VSS)

$$M(X_i) = M(X_{ii})\,R_s \qquad \text{(mg)}$$

$$= M(S_{ti})\,(f_{up}/f_{cv})\,R_s \qquad (4.12)$$

(d) The Total Volatile Suspended Solids Mass (mg VSS)

$$M(X_v) = M(X_a)+M(X_e)+M(X_i)$$

$$= \frac{Y_h\,R_s\,M(S_{bi})}{(1+b_h\,R_s)}(1+f\,b_h\,R_s)+M(X_{ii})\,R_s \quad \text{(mg)}$$

$$= M(S_{ti})\,R_s\left\{\frac{(1-f_{us}-f_{up})\,Y_h}{(1+b_hR_s)}(1+fb_hR_s)+\frac{f_{up}}{f_{cv}}\right\} \qquad (4.13)$$

(e) The Total Suspended Solids Mass (mg TSS)

$$M(X_t) = M(X_v)/f_i \qquad (4.14)$$

where

f_i = MLVSS/MLSS ratio of the sludge.

(f) The Carbonaceous Oxygen Demand (mg O/d)

$$M(O_c) = M(\text{O synthesis}) + M(\text{O endogenous mass loss})$$

$$= (1-f_{cv}Y_h)\,M(S_{bi})+f_{cv}(1-f)\,b_h\,M(X_a) \quad \text{(mg O/d)}$$

$$= M(S_{ti})(1-f_{us}-f_{up})\,\{(1-f_{cv}Y_h)$$

$$+f_{cv}(1-f)b_h\frac{Y_hR_s}{(1+b_hR_s)}\} \qquad (4.15)$$

Knowing the mass of mixed liquor (MX_t or MX_v) in the process, the volume of the process is determined from the value specified for the MLSS or MLVSS concentration, X_t, or X_v respectively,

$$V_p = M(X_t)/X_t = M(X_v)/X_v \qquad (4.16)$$

Knowing the volume V_p, the total nominal hydraulic retention time, R_{hn}, then is found from the specified daily average flow rate Q, using Eq. 4.3.

The above design equations lead to the following important conclusions:

The mass of volatile solids in the reactor is a function only of the mass of COD utilized per day and the sludge age. Consequently, insofar as the mass of sludge is affected, it is immaterial whether the mass of COD utilized arises from a low daily flow with a high COD concentration, or a high daily flow with a low COD concentration; provided $M(S_{ti})$ is the same in both instances, the masses of sludge will be identical. However, the hydraulic retention times will differ, being long in the first and short in the second instance, respectively. The hydraulic retention time, therefore, is incidental to the COD mass utilized, the MLVSS and the daily flow – it serves no basic kinetic function in steady state operation.

The active volatile mass $M(X_a)$ in the process is the live organism mass which performs the biodegradation processes. The remaining volatile masses, $M(X_e)$ and $M(X_i)$ are inactive and do not serve any function insofar as the biodegradation mechanisms in the process are concerned. The active fraction of the sludge mass with respect to the volatile solids f_{av} is given by

$$f_{av} = M(X_a)/M(X_v) \qquad (4.17)$$

substituting for $M(X_a)$ and $M(X_v)$ and noting that

$$M(S_{bi}) = (1-f_{us}-f_{up})\,M(S_{ti}) \text{ and}$$

$$M(X_{ii}) = (f_{up}/f_{cv}).M(S_{ti}) \text{ yields}$$

$$\frac{1}{f_{av}} = 1+fb_hR_s+\frac{f_{up}\,(1+b_h\,R_s)}{f_{cv}Y_h(1-f_{us}-f_{up})} \qquad (4.18)$$

If the total suspended solids mass (TSS) is used as the basis for determining the active fraction, the the active fraction of the sludge mass with respect to the total suspended solids f_{at} is given by

$$f_{at} = f_i\,f_{av} \qquad (4.19)$$

where

f_i = MLVSS/MLSS ratio of the sludge.

The design equations set out above form the starting point for all the activated sludge processes considered in this manuscript, from the relatively simple single reactor completely mixed aerobic process, to the more complex multi-reactor anoxic-aerobic and the anaerobic-anoxic processes. For these more complex systems the basic equations apply only if the constraints imposed on the processes are adhered to. Provided this is done the effects of nitrification and nitrification-denitrification and the associated oxygen demands can be formulated as additional to the basic equations. That this "simplified" approach is adequate for design has been established by the close correlation achieved when comparing the mean response of the process predicted by the general theory (and verified experimentally) with that predicted by the basic and additional equations.

It is important to take cognizance of the constraints. For example, in anoxic-aerobic processes an upper limit is imposed on the fractional mass of sludge that may exist unaerated. If exceeded, then *inter alia* the mass of sludge produced per day increases and the carbonaceous oxygen demand decreases below that predicted by the basic equations. The reason for these deviations lies in the kinetics of degradation of the slowly biodegradable particulate material – if the aerobic fraction of the sludge mass is too small the particulate material is only partially metabolized and accumulates in the process as additional volatile solids; concomitantly the carbonaceous oxygen demand is reduced. Evidently where such conditions apply the basic equations become inappropriate; however approximate solutions are sometimes obtainable by simulation using the general model, (see Arkley and Marais (1982) and Sehayek and Marais (1981) for a detailed discussion).

5. STEADY STATE DESIGN CHART

Plots of Eqs. 4.10 to 4.15 and 4.18 to 4.19 are shown in Figure 4.2 for a unit mass of $M(S_{ti})$. The values of the kinetic constants, Y_h, b_h, f_{cv} and f are those listed in Table 4.1 for 20°C and have been verified in extensive laboratory and pilot scale investigations (Ekama and Marais, 1978). The values of the unbiodegradable particulate and soluble COD fractions (f_{up} and f_{us} respectively) are listed in Table 4.2 and are for raw and settled municipal wastewaters.

The plots in effect give the masses $M(X_a)$, $M(X_e)$, $M(X_i)$, $M(X_v)$ and $M(X_t)$ contained in the reactor for a unit mass of COD applied per day for sludge ages up to 30 days. Also shown is the associated mass of oxygen to be supplied per day for carbonaceous material degradation and the active fraction of the sludge with respect to VSS (X_v) and TSS (X_t). The volumes of the reactors at two different sludge ages will be in direct proportion to the masses $M(X_t)$ or $M(X_v)$ if the mixed liquor concentrations, X_t or X_v are specified to be the same in both plants.

TABLE 4.2 INFLUENT WASTEWATER FRACTIONS FOR SETTLED AND UNSETTLED SEWAGE

Sewage Fraction	Sewage Unsettled	 Settled
Soluble unbiodegradable fraction, f_{us} (mg COD/mg COD)	0,05	0,08
Particulate unbiodegradable fraction, f_{up} (mg COD/mg COD)	0,13	0,04
MLVSS/MLSS ratio (f_i)	0,75	0,83

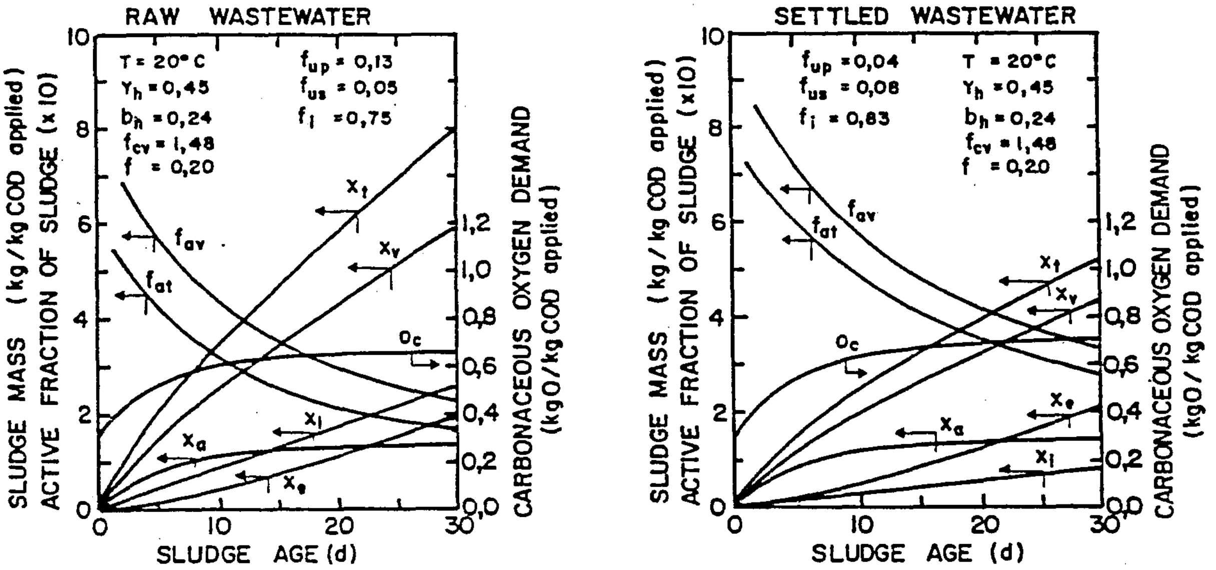

FIGURE 4.2: Active (X_a), endogenous (X_e), inert (X_i), volatile (X_v), total (X_t), and carbonaceous oxygen demand (O_c) per kg COD applied to the biological process, and active fraction with respect to volatile solids (f_{av}) and total solids (f_{at}) versus sludge age for raw (Figure 4.2a, left) and settled (Figure 4.2b, right) wastewaters at 20°C

The diagrams show that the active mass increases fairly rapidly with an increase in sludge age up to about 10 days sludge age, whereafter it increases only marginally. The carbonaceous oxygen demand shows a similar behaviour; at zero sludge age the oxygen demand is that due to synthesis, additional oxygen demand at sludge ages greater than zero is due to endogenous respiration which is proportional to the active mass. In contrast the fractions of endogenous and inert solids increase rapidly relative to the active mass at sludge ages greater than 10 days and consequently in processes treating raw wastewater at sludge ages longer than 10 days, only a small fraction of the sludge mass is active.

In design, the following sequence of calculation is suggested:

Select values for f_{up} and f_{us} which are believed to best represent these fractions of the wastewater. Select the MLVSS/MLSS ratio of the sludge produced. Determine:

(a) S_{upi} (Eq. 2.4), S_{usi} (Eq. 2.3), X_{ii} (Eq. 2.6)

(b) $M(S_{ti})$ and/or $M(S_{bi})$

(c) $M(X_a)$(Eq. 4.10), $M(X_e)$(Eq. 4.11), $M(X_i)$(Eq. 4.12), $M(X_v)$(Eq. 4.13), $M(O_c)$(Eq. 4.15) and $M(X_t)$(Eq. 4.14)

(d) V_p (Eq. 4.16)

(e) R_{hn} (Eq. 4.3)

(f) S_{te} (Eqs. 4.4a or b)

In the design sequence set out above the input COD and its characteristics will be governed by the specific waste flow. The parameter that requires selection is the sludge age; this will depend on the objectives for the plant, for COD removal only, nitrification etc. Specification of the sludge age, therefore, is an important design decision and requires special consideration, see Section 11. For our general purpose however, we will proceed to show how the various parameters such as reactor volume, oxygen demand, etc. are affected by the sludge age.

6. PROCESS VOLUME REQUIREMENTS

In Section 4 above a procedure is set out whereby the volume of the reactor is estimated. This involves determining the mass of sludge accumulated for a specified sludge age and a specified COD mass loading per day; the volume is then determined by diluting the mass of sludge accumulated to a specified mixed liquor concentration. Once the volume is fixed, the retention time, or aeration time, also is fixed (by Eq. 4.3). Hence retention time, or aeration time, is immaterial in the design and it is merely a consequence of the kinetics governing the process. This point is mentioned specifically because some design procedures lay stress on the aeration time as a design parameter, a procedure that can result in serious miscalculation of the volume requirements. Compare, for example, two plants operating at the same sludge age, both receiving the same mass of COD per day but the first at high concentration and low flow and the second at a low concentration and high flow. If designed on a specified retention time, the volume of the first will be much smaller than that of the second whereas the sludge mass contained in the reactors will be identical. Consequently the first plant may have an inordinately high MLSS concentration which may cause problems in the secondary settling tank.

Retention time therefore is a completely inappropriate parameter for design and other purposes such as a criterion for comparing the tankage requirements of different plants.

To form an overall impression of the volume requirement and in order to make valid comparisons on the volume requirements between different plants, the volume per unit COD input is a much superior parameter. This method involves the specification of an MLVSS or MLSS concentration (X_v or X_t) from which the reactor volume is determined from the sludge mass produced per unit COD load at the required sludge age (see Figure 4.2). Furthermore, in order to assess the effects of primary sedimentation it is preferable to relate the volume requirements to the input COD to the treatment works; for example for settled wastewater, the COD load on the process is reduced i.e. $S_{ti(process)} = S_{ti(works)}\ (1-f_{rps})$ where f_{rps} = fraction of COD removed by primary sedimentation. It is $S_{ti(process)}$ which fixes the sludge mass and because for settled wastewater S_{ti} is up to about 40% less than for raw wastewater, the process volume for settled wastewater accordingly will be much lower than that for raw wastewater.

From Eqs. 4.13, 4.14 and 4.16 the process volume requirements on the basis of process COD load is found by calculating the mass of sludge, $M(X_t)$, generated from the total influent COD load $M(S_{ti})$ and assuming an MLVSS/MLSS ratio for the sludge mass i.e.

$$V_pX_t\,f_i = Q.S_{ti}\,R_s\left\{\frac{(1-f_{us}-f_{up})Y_h(1+fb_hR_s)}{(1+b_hR_s)} + \frac{f_{up}}{f_{cv}}\right\} \tag{4.20}$$

and hence the process volume requirements per *unit* COD load to the works for a specified MLSS concentration X_t and assumed MLVSS/MLSS ratio f_i is given by

$$\frac{V_p}{Q.S_{ti}} = \frac{R_s}{X_tf_i}\left\{\frac{(1-f_{us}-f_{up})Y_h(1+fb_hR_s)}{(1+b_hR_s)} + \frac{f_{up}}{f_{cv}}\right\} \tag{4.21a}$$

Taking the wastewater characteristics for raw wastewater given in Table 4.2, and the values of the kinetic constants given in Table 4.1, a plot of process volume requirements in m^3/kg COD applied per day versus sludge age for different MLSS concentrations in kg/m^3 is given in Figure 4.3. Clearly, the lower the specified MLSS concentration and the longer the sludge age, the larger the process volume per unit COD load.

The process volume requirements may also be determined from the equivalent COD load per capita. The vertical axis on the right hand side of Figure 4.3 is

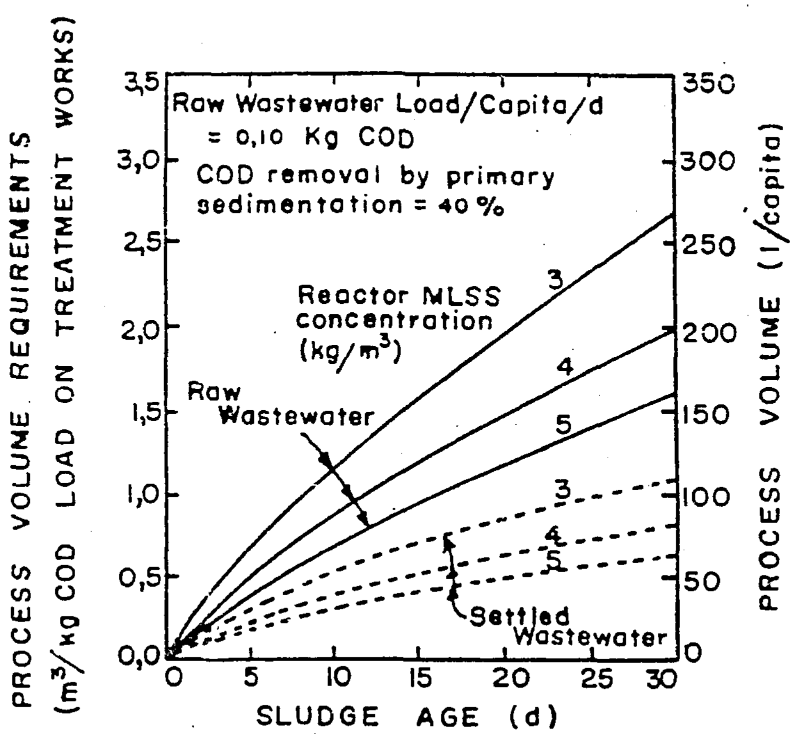

FIGURE 4.3: Process volume requirements in m³/kg COD raw wastewater load versus sludge age at different average process MLSS concentrations for raw and settled wastewater (assuming 40% COD removal by primary sedimentation). Process volume requirements in *l*/capita also given based on a raw wastewater COD contribution of 0,10 kg COD/capita/d.

scaled on the basis that the equivalent COD load per capita for raw wastewater is 0,10 kg COD/d. Hence treating raw wastewater at a sludge age of 25 days and an MLSS concentration of 5 kg/m³, a process volume of 145 *l* per capita is required or 1,45 m³/kg COD applied per day.

The comparative process volume requirements for settled wastewater per unit COD load on the treatment works also can be determined from Eq. 4.21a, provided the COD fraction removed by primary sedimentation is incorporated in Eq. 4.21a by multiplying right-hand side of Eq. 4.21a by $(1-f_{rps})$ and inserting the appropriate f_{up} and f_{us} values for settled wastewater i.e.

$$\frac{V_p}{S_{ti}Q} = \frac{R_s}{X_t f_i}\left\{\frac{(1-f_{us}-f_{up})Y_h(1+fb_hR_s)}{(1+b_hR_s)} + \frac{f_{up}}{f_{cv}}\right\}(1-f_{rps}) \tag{4.21b}$$

where S_{ti} = total COD concentration to treatment works.

The process volume requirements for settled wastewater are shown also in Figure 4.3 for the f_{up}, f_{us} and f_i values given in Table 4.2 assuming primary sedimentation removes 40% of the COD load (f_{rps} = 0,40). From Figure 4.3, treating settled wastewater at a sludge age of 25 days and an MLSS concentration of 5 kg/m³, a process volume of 0,60 m³/kg raw COD load to the works or 60 *l* per capita is required. Comparing the process volume treating raw or settled wastewaters it can be seen that a significant reduction in process volume can be obtained through primary sedimentation.

7. CARBONACEOUS OXYGEN DEMAND

The mean daily carbonaceous oxygen demand per unit COD discharged to the *process* is shown plotted versus sludge age at 20°C for raw and settled wastewater in Figures 4.2a and 4.2b respectively (calculated from Eq. 4.15) – for treatment of raw wastewater S_{ti} = COD of the raw wastewater; for treatment of settled wastewater S_{ti} = (COD of raw wastewater)$(1-f_{rp})$. These diagrams show that for sludge ages longer than 15 days the increase in carbonaceous oxygen demand per unit process COD load $\{M(O_c)/M(S_{ti})\}$ is marginal with increase in sludge age for both raw and settled wastewater. The carbonaceous oxygen demand per unit process COD load for raw and settled wastewater is usually within 10% of each other, with the demand for settled wastewater being the higher value. This is because compared to raw wastewater, more of the total COD in settled wastewater is biodegradable, i.e. $(1-f_{us}-f_{up})$ usually is greater for settled wastewater than for raw wastewater (see Eq. 2.7) but the difference is small. This can be seen in Figures 4.2a and 4.2b – at 20 days sludge age, the carbonaceous oxygen demand per unit COD load is 0,635 kgO/kgCOD for raw wastewater and 0,682 kgO/kgCOD for settled wastewater. Although there is only a small difference in carbonaceous oxygen demand *per unit process COD load* between raw and settled wastewaters there is a substantial difference in the O_2 demand per unit COD load to the works. For settled wastewater, this is given by 0,682 $(1-f_{rps})$ which for f_{rps} = 0,40 gives 0,41 kgO/kgCOD load on the works.

Clearly primary sedimentation will lead to significant aeration energy savings – because primary settling tanks remove about 30 to 40% of the raw influent COD, the carbonaceous oxygen demand for settled wastewater will be about 30 to 40% lower than that for raw wastewater.

The carbonaceous oxygen demand is the oxygen demand for the oxidation of the influent COD only. In nutrient removal processes oxygen is also required for nitrification i.e. the oxidation of ammonia to nitrate, and some oxygen is recovered during denitrification i.e. reduction of nitrate to nitrogen gas. The total oxygen demand for a nutrient removal process therefore is given by the sum of the carbonaceous and nitrification demands less that recovered by denitrification. The procedures for calculating the oxygen demand for nitrification and the oxygen recovered by denitrification are discussed in Chapters 5 and 6 respectively.

Owing to the daily cyclic nature of the COD load the carbonaceous oxygen demand will vary concomitantly. The TKN load will also vary daily cyclically; in nitrification-denitrification plants the oxygen demand for nitrification and oxygen recovery with denitrification conjointly with the carbonaceous demand give rise to a net cyclic total demand. From simulations using the general model of Dold *et al* empirical rules for estimating the daily peak and trough net oxygen demands have been developed and are discussed in Chapter 6, Section 7.6.2.

8. DAILY SLUDGE PRODUCTION

The mass of sludge produced per day by the activated sludge process is equal to the mass of sludge wasted per day and is called waste activated sludge or secondary sludge. From the definition of sludge age (see Eq. 4.1), the mass of sludge produced per day $M(\Delta X_t)$ is given by the mass of sludge in the process $\{M(X_t)\}$ divided by the sludge age (R_s) i.e.

$$M(\Delta X_t) = M(X_t)/R_s \quad \text{mg TSS/d}$$

Substituting Eq. 4.14 for $M(X_t)$ and simplifying, yields the sludge produced per day per mg COD load on the biological process i.e.

$$\frac{M(\Delta X_t)}{M(S_{ti})} = \frac{1}{f_i}\left\{\frac{(1-f_{up}-f_{us})Y_h}{(1+b_hR_s)}(1+fb_hR_s) + \frac{f_{up}}{f_{cv}}\right\} \tag{4.22}$$

A plot of the daily total sludge mass (TSS) produced per unit COD load on the biological process (Eq. 4.22) versus sludge age (R_s) is shown in Figure 4.4 for raw and settled wastewater (see Table 4.2 for wastewater characteristics). From Figure 4.4, it can be seen that the mass of sludge produced in the activated sludge process (per unit COD load on the biological process) decreases as the sludge age increases for both raw and settled wastewater but the rate of decrease is negligible at sludge ages longer than 20 days. Treating settled wastewater results in lower secondary sludge production per unit COD load on the biological process than treating raw wastewater. This is because the unbiodegradable particulate COD fraction (f_{up}) in settled wastewater is lower than that in raw wastewater.

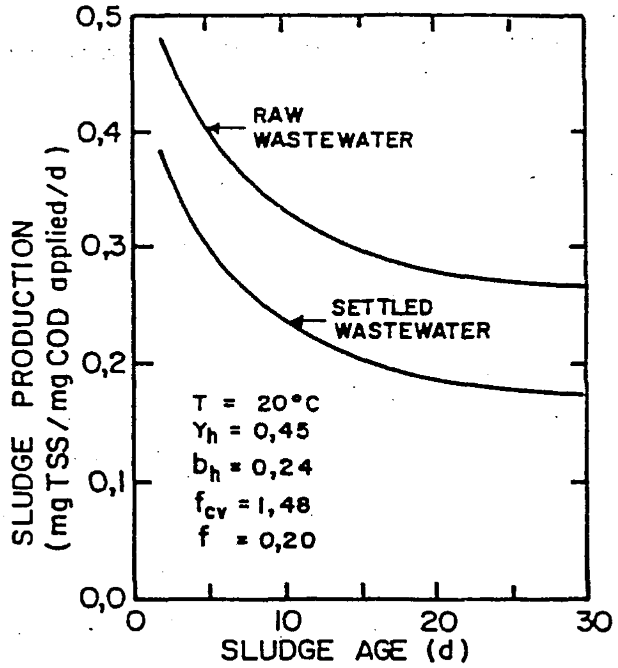

FIGURE 4.4: Daily total suspended solids (TSS) sludge production in the activated sludge process (secondary sludge) per unit COD load on the process versus sludge age for raw ($f_{up}=0.13$, $f_{us}=0.05$ and $f_i=0.75$) and settled ($f_{up}=0.04$, $f_{us}=0.08$ and $f_i=0.83$) wastewaters at 20°C

Temperature effects on secondary sludge production are small – sludge production at 14°C* is about 5% greater than at 20°C, a difference which is completely masked by the uncertainty in the estimates of the wastewater characteristics f_{up} and f_{us} and the MLVSS/MLSS ratio (f_i) of the sludge.

Although the sludge production of a plant treating settled wastewater is lower than that of a plant treating raw wastewater, the *total* sludge mass of sludge to be disposed of for a plant treating settled wastewater is higher because the total sludge includes both primary and secondary sludge – in plants treating raw wastewater, only secondary sludge is generated. The *total* sludge produced treating settled wastewater is slightly higher (by 10 to 20%) than that treating raw wastewater – this is because some of the settleable solids are biodegradable and are oxidized in the activated sludge process.

The daily production of secondary and primary sludges (when primary settling tanks are included in the treatment works) is the mass of sludge that needs to be disposed of by sludge handling methods. Sludge handling and disposal is not dealt with in this manual but the design of nutrient removal activated sludge processes should not be seen as separate from the design of the sludge disposal unit processes, in fact all unit processes of the treatment works from raw water pumping to ultimate disposal of sludge, should be viewed as an integrated system where the design of one unit process depends on the unit processes before it, and decisions

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*Sludge production at 14°C can be calculated by adjusting b_h for temperature using Eq. (4.5) in Table 4.1.

on its design may affect the design of unit processes following it. The following examples will illustrate this – (1) in sludge handling practice there is evidence that a mixture of primary and secondary sludges thickens and dewaters better than secondary sludge alone; however, in biological nutrient removal practice it has been found that when treating raw wastewater, better removals of nitrogen and phosphorus are obtained than when treating settled wastewater – ease of sludge thickening and dewatering therefore need to be weighed against the higher nutrient removal; (2) when phosphorus-rich waste biological sludges are subjected to anaerobic conditions, the phosphorus is released from the solid phase to the liquid phase; consequently, *supernatant* from waste sludge gravity thickeners may be phosphate rich and should not be discharged to the activated sludge process – in design consideration should be given to chemically treating the gravity thickener supernatant to precipitate the phosphorus or, thickening the sludge by dissolved air flotation to prevent phosphorus release; (3) to improve biological excess P removal, the Johannesburg Municipality specifically operates one of the anaerobic digesters in an acid condition for the production of short chain organic compouds which are then discharged to the anaerobic reactor of the activated sludge process (see Chapter 3, Section 3.3) – consideration must be given at the design stage to the desirability of providing a specially designed facility for this purpose.

9. NUTRIENT REQUIREMENTS FOR SLUDGE PRODUCTION

All live biological material and many inert organic materials contain nitrogen and phosphorus. In the volatile suspended solids in the activated sludge process, the nitrogen content (as N with respect to VSS, f_n) ranges between 9% and 12% with an average of about 10%, and the phosphorus content (as P with respect to VSS, f_p), under purely aerobic conditions, ranges between 1% and 3% with an average of about 2,5%.

Under daily average conditions, the mass of nitrogen incorporated into the sludge mass is equal to the nitrogen content of the daily waste sludge. Accepting that the nitrogen content of the active, endogenous and inert volatile fractions are equal (which in most instances is a reasonable assumption), and equating the *concentration* of nitrogen required for daily sludge production per litre influent, N_s, to the nitrogen content of the waste sludge yields

$$Q.N_s = f_n\, M(X_v)/R_s$$

Substituting for $M(X_v)$ from Eq. 4.10 and solving for the nitrogen requirements per unit COD, N_s/S_{ti} yields

$$\frac{N_s}{S_{ti}} = f_n\left[\frac{Y_h(1-f_{us}-f_{up})}{(1+b_hR_s)}(1+fb_hR_s)+\frac{f_{up}}{f_{cv}}\right] \tag{4.23}$$

where

f_n = nitrogen content of the volatile solids = 0,10 mg N/mg VSS

To estimate the P requirement for daily sludge production, it is not correct to assume that the P content of the active, endogenous and inert volatile fractions is the same. The P content of the endogenous and inert volatile fractions (f_p) remain approximately constant at 1,5% whereas the P content of the active mass (γ) can vary from 3% to as high as 35% depending on the different conditions in the process. Indeed, it is the propensity of the active mass to accumulate large quantities of P that is exploited in biological P removal processes. Accepting the different P contents of the active fraction and the endogenous and inert fractions, the concentration of P per ℓ influent (P_s) required for daily sludge production per mg/ℓ COD load (S_{ti}) is given by

$$\frac{P_s}{S_{ti}} = \left\{\frac{(1-f_{up}-f_{us})Y_h}{(1+b_hR_s)}(\gamma+f_p\, fb_h\, R_s)+f_p\,\frac{f_{up}}{f_{cv}}\right\} \tag{4.24}$$

where

γ = P content of the active mass (mg P/mgVASS)
f_p = P content of the inert and endogenous mass (mg P/mg VSS) = 0,015

The P incorporated into the sludge mass per unit COD (P_s/S_{ti}) is also the P removal from the wastewater per unit COD,$\Delta P/S_{ti}$. In the activated sludge process, under purely aerobic conditions the γ coefficient is about 0,03 mgP/mgVASS: when an anoxic reactor is included in the process, the coefficient is increased to about 0,06 mgP/mgVASS; when an anaerobic reactor is incorporated in the process, then, depending, amongst others, on the readily biodegradable COD concentration in the anaerobic reactor, the γ coefficient can vary between 0,06 and 0,35 mgP/mgVASS (see Chapter 7, Section 3.3). By stimulating high γ coefficients, with the aid of the anaerobic reactor, high P removals may be effected. This is the basis of biological P removal and is discussed in detail in Chapter 7.

A plot of Eqs. 4.23 and 4.24 is given in Figure 4.5 for $f_n = 0{,}10$ mg N/mg VSS, $\gamma = 0{,}03$ mgP/mgVASS, $f_P = 0{,}015$ mgP/mgVSS, $Y_h = 0{,}45$ mgVSS/mgCOD, $b_h = 0{,}24$/d at 20°C, $f_{cv} = 1{,}48$ mgCOD/mgVSS, and $f = 0{,}20$, versus sludge age (R_s) for raw ($f_{us} = 0{,}05$ and $f_{up} = 0{,}13$) and settled ($f_{us} = 0{,}08$ and $f_{up} = 0{,}04$) wastewaters.

From Figure 4.5, it is evident that greater quantities of nitrogen and phosphorus are required for sludge production from raw than from settled wastewaters. This is because greater quantities of sludge are produced per mgCOD process load when treating raw wastewaters. Furthermore, the nutrient requirements decrease as the sludge age increases because nett sludge production decreases as sludge age increases. Generally, for sludge ages greater than 10 days, the nitrogen removal attributable to nett sludge production is less than 0,025 mgN/mgCOD applied. As the influent TKN/COD ratio for domestic wastewater is in the approximate range 0,07 to 0,12, only a minor fraction of the influent nitrogen is removed by sludge wastage; additional nitrogen removal is obtained only by nitrification and denitrification, the kinetics of which are discussed in Chapters 5 and 6. With regard to P removal, as the total P/COD ratio of domestic wastewaters usually ranges

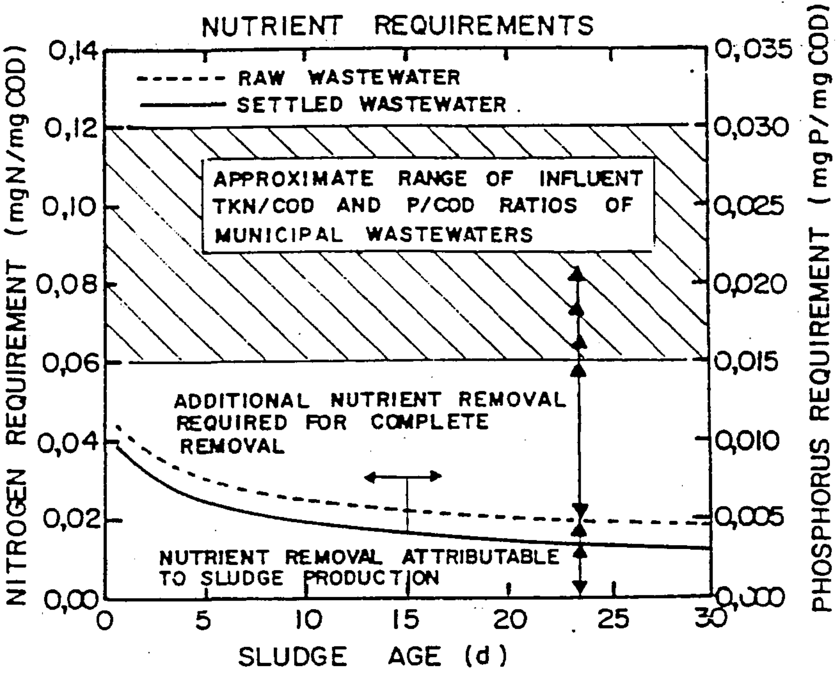

FIGURE 4.5: Normal nutrient nitrogen and phosphorus requirements per kgCOD load in the activated sludge process versus sludge age together with influent nutrient loads per kgCOD.

between 0,015 to 0,030, only a minor fraction of phosphorus, 0,005 mgP/mgVSS, can be removed by the aerobic activated sludge process; excess biological P removal is obtained by stimulating high γ coefficients in processes by incorporating an anaerobic reactor. Details of excess biological P removal are considered in Chapter 7.

10. PROCESS DESIGN AND CONTROL

The parameter of fundamental importance in the design and control of the activated sludge process is the sludge age. It is instructive to consider how this parameter influences both the control and design of the process and to contrast it to other design and control parameters that have been implemented in the past.

10.1 Sludge age versus load factor as control parameter

A widely accepted basis for design (especially in North America and Europe) is the Load Factor (LF). It is also known as the Sludge Loading Rate (SLR) and has been used principally when the BOD_5 forms the influent energy measurement parameter. The objective of the load factor was to form an estimate of the Food/Microorganism (F/M) ratio and as a result the load factor was defined as

$$LF = \frac{\text{mass of BOD load/d}}{\text{mass of sludge in process}} = \frac{M(S_{BOD})}{M(X_v) \text{ or } M(X_t)} \qquad (4.25)$$

where

$M(S_{BOD})$ = mass of BOD load per day (mg)

$M(X_v)$,$(X(X_t)$ = mass of volatile (mgVSS) or total (mgTSS) suspended solids in process respectively.

A difficulty with the approach is that both the $M(X_v)$ and $M(X_t)$ measurements include unbiodegradable and inert material. To represent the F/M ratio correctly, the LF should be defined as follows:

$$LF = F/M = M(\Delta S_{BOD})/M(X_a) \qquad (4.26)$$

where

$M(\Delta S_{BOD})$ = mass change in BOD across plant per day (mgBOD/d)

$M(X_a)$ = active mass in process (mgVASS)

As a practical control parameter, the F/M ratio as defined by Eq. 4.26 is virtually unusable principally because $M(X_a)$ cannot be measured directly. It is for this reason that the LF was *approximated* by Eq. 4.25* with $M(X_v)$ or $M(X_t)$ as the solids parameter. In this form, however, the LF is an uncertain parameter against which to correlate the response of the process because $M(X_v)$ depends on both the biodegradable COD and the inert particulate material in the influent. For example, for the same biodegradable COD concentration, the LF will differ between settled and raw wastewaters whereas the oxygen demand and $M(X_a)$ will be the same for both. Furthermore the LF will be different depending on whether the settleable solids are measured in terms of MLVSS or MLSS. When the COD forms the influent energy measurement parameter, the LF is defined by

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*Because the effluent BOD usually is very small with respect to the influent BOD, the change in BOD mass load per day $M(\Delta S_{BOD})$ is closely approximated by the BOD mass load per day $M(S_{BOD})$.

Eq. 4.27 which is even further out of place because the COD measure includes the unbiodegradable soluble and particulate COD fractions

$$LF = \frac{M(S_{COD})}{M(X_v) \text{ or } M(X_t)} \qquad (4.27)$$

where

$M(S_{COD})$ = mass of COD load per day.

If Eq. 4.27 is written in terms of the mass change of COD per day across the plant i.e. $M(\Delta S_{COD})$, then account of only the soluble unbiodegradable COD is taken. The unbiodegradable particulate COD affects the LF by increasing both the $M(\Delta S_{COD})$ and $M(X_v)$ or $M(X_t)$ so that different load factors can be obtained depending on the unbiodegradable particulate COD fraction.

To overcome the problem of the unbiodegradable COD fractions on the LF, Marais and Ekama (1976) defined the F/M ratio in terms of COD as the mass change of biodegradable COD per day per unit *active* volatile mass and, to distinguish this parameter from the LF, called it the Substrate Utilization Rate with respect to the active mass, SUR_a i.e.

$$SUR_a = F/M = M(\Delta S_{COD})/M(X_a) \qquad (4.28)$$

where

$M(\Delta S_{COD})$ = mass change in biodegradable COD per day.

Again the SUR_a parameter is of little practical value because neither the $M(\Delta S_b)$ (due to the unknown unbiodegradable particulate COD content of the wastewater) nor the $M(X_a)$ can be measured. However, using Eq. 4.28 as a basis, Marais and Ekama (1976) showed that the Substrate Utilization Rates with respect to the active mass (SUR_a), volatile mass (SUR_v) and total mass (SUR_t) can all be written in terms of sludge age, but different SUR's are obtained for a fixed sludge age depending on the parameters for measuring the sludge concentration (X_a, X_v or X_t) and the influent energy (COD or BOD) (see Figure 4.6).* Similarly for the same sludge age, different values of the LF or SLR can be obtained depending on the measurement parameters for the sludge concentration and influent energy.

The dependency of the SUR's, LF's or SLR's on the measurement parameters for the sludge concentration and influent energy is undesirable because it provides a permanent source of error and misunderstanding at, and between all levels of technical staff concerned with the process. But perhaps of greater importance is the fact *that all the different SUR's, LF's or SLR's are functionally related to the sludge age – no additional information can be derived from the SUR's, LF's or SLR's that cannot be derived from the sludge age.*

The sludge age therefore can replace completely the LF, SLR and SUR as a reference parameter. Furthermore, the sludge age can be fixed by a simple process control procedure if the process is appropriately designed. This control procedure is much more practical than procedures based on the LF approach.

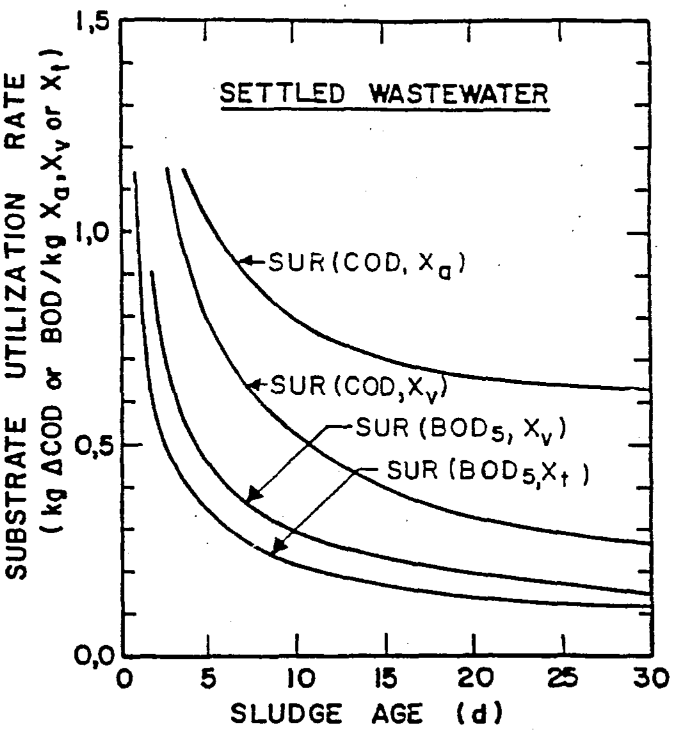

FIGURE 4.6: Comparison between various substrate utilization rates (SUR) and sludge age.

10.2 Process control

The usual procedure of the activated sludge process involves keeping the sludge concentration at some value either specified from design considerations or established from experience on the plant behaviour. Knowing the sludge concentration (X_v or X_t), to check that the LF is within the desired limits requires extensive testing throughout the day of the influent COD (or BOD) concentration and flow pattern to determine the daily COD (or BOD) *mass* load. To calculate the corresponding sludge age, a history of the daily sludge mass wastage needs to be kept. Often in conventional plants, the waste sludge is abstracted from the secondary settling tank underflow to benefit from the thickening function of the secondary settlers. However, the sludge concentration of the underflow varies considerably with the daily cyclic flow through the plant (see Figures 4.7 and 4.8). Hence, each time sludge is wasted it is necessary to measure the underflow concentration and calculate the sludge age.

From the above, it is clear that in order to know the sludge age or the LF accurately, intensive testing of the plant and influent is required. This is manageable on large plants where the technical supervision is adequate, but on small plants both the Load Factor and sludge age usually are unknown.

The discussion above demonstrates that wasting sludge from the secondary settler underflow does not result in a simple and accurate control of either the sludge age or the LF. However, by using a different sludge wasting procedure, an accurate and simple con-

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*The derivation of some of these SUR values is given by Marais and Ekama and those for well settled wastewater (i.e. f_{up} = 0,000) is shown in Figure 4.6. It should be noted that if volatile inert material is present in the influent i.e. f_{up} > 0,00, e.g. in partially settled and raw wastewaters, a set of SUR values can be calculated for each selected value of f_{up}.

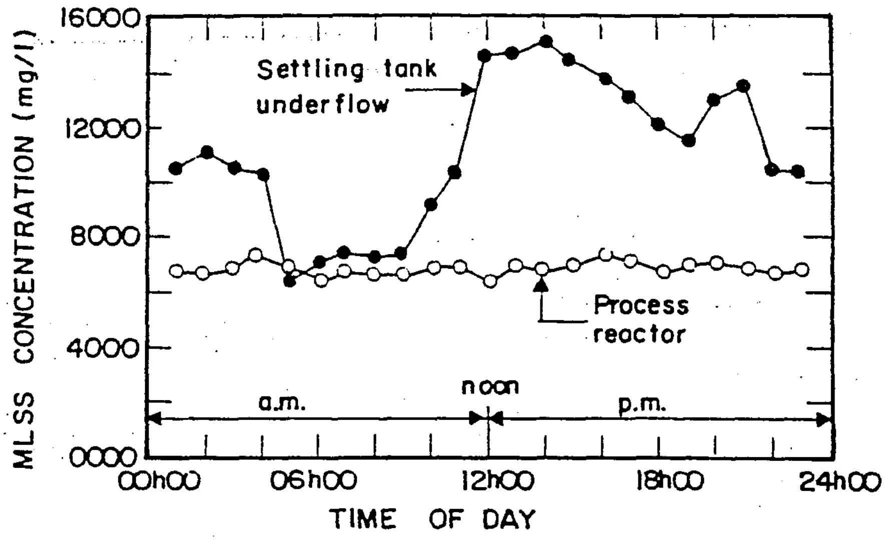
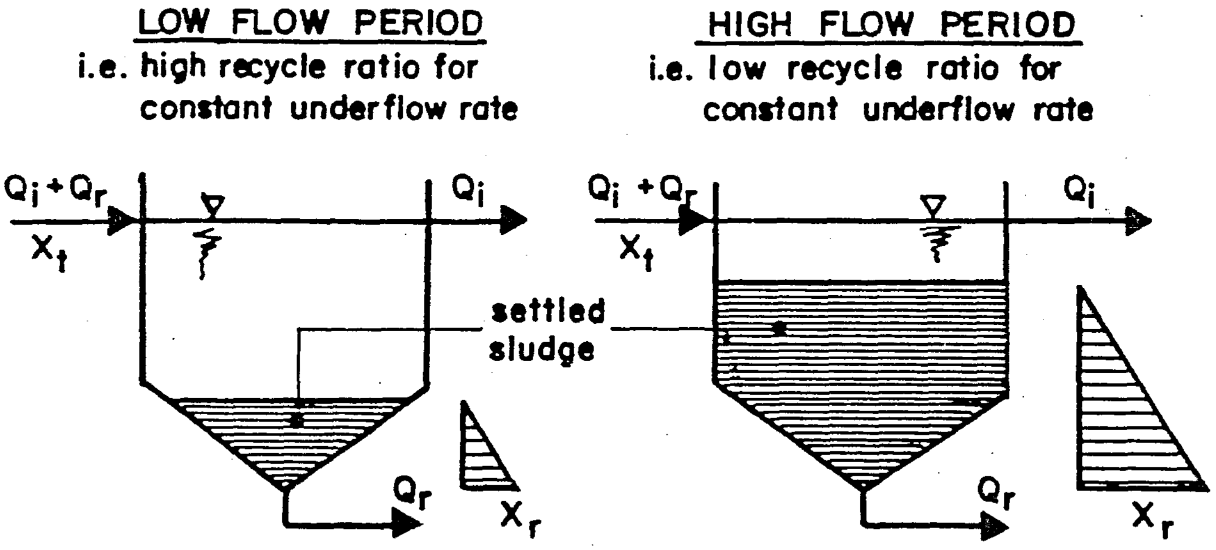


FIGURE 4.7: Experimental data from a full scale activated sludge plant illustrating stability of the reactor MLSS concentration compared with that in the secondary settler underflow.



Q_i = INFLUENT FLOW RATE (m^3/d)

Q_r = UNDERFLOW RATE (m^3/d)

X_t = SLUDGE DENSITY IN REACTOR (mg MLSS/l)

X_r = SLUDGE DENSITY IN UNDERFLOW (mg MLSS/l)

$$X_r = \frac{(Q_i + Q_r)}{Q_r} \cdot X_t = \frac{s+1}{s} \cdot X_t$$

$s = Q_r / Q_i$ = UNDERFLOW RECYCLE RATIO

FIGURE 4.8: Increased sludge accumulation and higher underflow sludge concentration for (i) decreased recycle flow for fixed influent flow and (ii) increased influent flow for fixed recycle flow.

trol of the sludge age is possible. This different sludge wastage procedure is called the hydraulic control of the sludge age.

10.3 Hydraulic control of sludge age

Hydraulic control of sludge age was first proposed and implemented in a generalized form by Garrett in 1958, based on a method of "modified sewage aeration" implemented by Setter, Carpenter and Winslow (1945).* It operates as follows: Suppose a sludge age of 10 days is specified. A satellite settling tank or dissolved air flotation unit is provided, completely independent of the secondary settling tank, to which 1/10 of the reactor volume is pumped every day. The supernatant is discharged to the effluent channel (or pumped back to the reactor),** the thickened sludge is pumped to the digesters or drying beds. That this method gives the correct sludge age can be seen from Eq. 4.2 and the procedure works because the mixed liquor concentration changes insignificantly over the day. This is verified by comparing the stability of the concentration of mixed liquor in the reactor and the concentration of sludge in the underflow over the day in one of the major plants of Johannesburg, Figure 4.7 (Nicholls, 1975).

Controlling the sludge age in small works may require pumping at regular times, every day or every few days, provided the volume pumped is such that the required volume, v, per day is abstracted. On large plants the satellite settling tank (or flotation unit) may be designed for continuous operation receiving a constant flow, or, operated over a part of the day.

An important point about hydraulic control of the sludge age is that *irrespective of the flow through the plant* if a fixed fraction of the volume of the reactor is removed and wasted every day, the sludge age is fixed. If the COD mass load per day on the plant remains constant, the sludge concentration will remain constant automatically. If the COD mass load increases, the sludge concentration will increase automatically, to maintain the same sludge age. Thus, by monitoring the MLSS concentration and its changes for a fixed sludge age an indirect measure is obtained of the COD load and load change.

By means of the hydraulic control procedure, the sludge age may be changed by simply changing the volume wasted per day. If say, the sludge age is reduced from 25 days to 20 days by hydraulic control, the full effect of the change will become apparent only after about 15 days. Thus the organisms have an opportunity to adapt gradually to the change in load.

Hydraulic control of sludge age is particularly relevant to plants with sludge ages longer than about 4 days because for these plants the mass of sludge contained in the secondary settling tanks is a minor fraction of the total mass of sludge in the system. At sludge ages shorter than 4 days the mass of sludge in the secondary settling tanks becomes appreciable with respect to the total mass of sludge in the system. Hydraulic control will have to take cognizance of this and accuracy of the control will require additional testing. It should be noted that irrespective of the sludge age of the plant, the satellite settling tank required for hydraulic control will always be very small with respect to the secondary settling tanks – at 2 days sludge age the retention time is usually about 4 to 6 hours and hence the satellite settling tank has to accommodate 1/12 to 1/8 of the daily average flow; at 25 days sludge age the retention time is about $\frac{3}{4}$ to $1\frac{1}{4}$ days and the satellite settling tank has to accommodate about 1/20 of the daily average flow; in contrast the secondary settling tank has to accommodate 2 to 4 times the daily average flow.

With regard to design, hydraulic control of the sludge age has the following implications: It devolves a greater onus on the designer and removes responsibility from the plant operator: It becomes essential that the designer calculates the sludge mass more exactly, to provide sufficient reactor volume under the design load to allow for the required concentration of MLSS in the reactor at the specified sludge age. Also, the settling tank design, recycle and aeration capacities must be adequate. If these aspects are catered for adequately, then with hydraulic control of the sludge age, plant control is simplified and, on small scale plants, may even do away with the requirements for solids and SVI tests except at long intervals. Hydraulic control of sludge age makes the Sludge Loading Rate, SLR, (or Substrate Utilization Rate, SUR), redundant and introduces an entirely different attitude to process control.

11. SELECTION OF SLUDGE AGE

Selection of the sludge age is the most fundamental and important decision in the design of an activated sludge process. The sludge age of a plant depends on many factors *inter alia,* stability of the process, sludge settleability, whether or not the waste sludge should be suitable for direct discharge to drying beds, and most important of all, the quality of effluent required i.e. is COD removal only acceptable, must be effluent be nitrified, is nitrogen and phosphorus removal required. Although the discussion below will be aimed at South African conditions, where nitrification is obligatory and stringent standards for the effluent phosphate concentration have been promulgated, other situations will be discussed briefly so as to place South African activated sludge process technology in perspective with other technologies, for example those of Europe and North America.

11.1 Short sludge ages (1-5 days)

11.1.1 Conventional plants:

These plants are operated in the conventional con-

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*For a review of the development of hydraulic control of sludge age see Ekama and Marais (1978).

**It should be noted that in a gravity thickener for waste sludge from a nutrient removal plant, considerable concentrations of phosphorus may be released in the thickener, which, when discharged to the activated sludge process will be counter-productive for excess biological P removal. The thickener supernatant may have to be treated chemically to precipitate the P before discharge to the activated sludge process (see Section 8 above).

figuration i.e. a semi plug flow configuration, but modified systems such as contact stabilization, step aeration, step feed and others are also implemented.

Short sludge age plants are extensively used in Europe and North America but are virtually unknown in South Africa principally because the promulgated standards for effluents cannot be satisfied, i.e. that of obligatory nitrification.

Short sludge age plants are for the purpose of COD removal only, towards which objective sludge ages of 1 to 3 days are sufficient. BOD or COD reductions, for the conventional system range from 75 to 90%; the actual removal achieved depends on the influent, the state of the sludge and the efficiency of the settling tanks. Predatory activity of the micro-organisms in the sludge is relatively low which causes turbidity and high effluent COD.

Nitrification should be absent, or the plant can be operated to prevent nitrification; this however may be difficult to achieve with high water temperatures in the tropics.

Advantages with low sludge age plants are: small process volumes for unit input COD (Figure 4.3) with corresponding low carbonaceous oxygen demand (see Figure 4.2); nitrification is unlikely, giving a saving in the nitrification oxygen demand (see Figure 6.1); oxygen demand under cyclic flow and load tends to be heavily damped making control of the oxygen easier than with longer sludge age plants. Disadvantages are: the mass of sludge produced by unit COD input is high (Figure 4.4) and the active fraction of the sludge is high (Figure 4.2). The sludge treatment facilities accordingly constitute an appreciable fraction of the treatment costs although the running costs can be reduced by energy recovery from anaerobic digestion of the sludge (see below); the process tends to be unstable and sensitive to mechanical breakdown.

Because of the relatively low oxygen demand and the high sludge production per unit input COD it is possible to make these high rate processes near self sufficient with respect to energy requirements – sufficient methane gas can be produced by anaerobic digestion of the waste sludge to generate electricity for oxygen supply and process operation.

11.1.2 Aerated lagoons:

In aerated lagoons, (as opposed to aerated oxidation ponds which supplement the oxygenation by algae) the oxygen requirement in the process is supplied wholly by aerators, usually floating mechanical ones. There are essentially two types of aerated lagoons, suspension mixed and facultative.

Suspension mixed aerated lagoons have sufficient energy input per unit volume by the aeration equipment to keep the sludge in suspension. In facultative lagoons this energy input is insufficient and settlement takes place on the lagoon floor, to form a sludge layer which decomposes anaerobically, as in an oxidation pond.

Kinetically, suspension mixed lagoons are members of the activated sludge process, the only difference being in that the former is a flow through system i.e. the sludge age equals the nominal hydraulic retention time, whereas the latter has a sludge recycle from the settling tank making the sludge age longer than the hydraulic retention time. Accordingly the volume of the aerated lagoon per unit input COD is large whereas it is small for the conventional short sludge plant. (For a sludge age of, say, 1 day the aerated lagoon will have an hydraulic retention time of one day whereas the short sludge age conventional plant (with a one day sludge age) will have a hydraulic retention time of about 1/8 of a day.

The effluent from a suspension mixed aerated lagoon has the same constitution as the mixed liquor in the basin. The COD removed from the system via the oxygen demand is relatively small so that the COD in the effluent is quite unacceptable for discharge to receiving waters. In fact the principal objective of all short age plants is to act as biological flocculators, to physically change the influent COD, soluble and particulate, to a form that allows effective liquid-solid separation. In the conventional short sludge age plant the waste sludge is transferred to the sludge treatment facility; in the aerated lagoon usually a second pond is provided i.e. an oxidation pond or a facultative aerated lagoon, to allow the now readily settleable particulate material to settle as a sludge layer and thereby produce a relatively solids-free effluent.

Aerated lagoons find application principally in industrial waste treatment and seasonal waters. Design considerations are not dealt with in this manual; these are discussed by Marais and Ekama (1980) and reference should be made to that report.

11.2 Intermediate sludge ages (10-15 days)

Where nitrification is obligatory, the sludge ages required are 5 to 8 times longer than those for COD removal only, depending on the temperature. In the temperate regions where water temperatures may fall below 12°C to 14°C the sludge age is unlikely to be less than 10 to 15 days. In this range of sludge age, the effluent COD concentration no longer plays a role in the design because for sludge ages longer than about 4 days, the effluent COD or BOD concentrations remain approximately constant (see Section 3.4 above). The effluent ammonia concentration also plays a minor role in design because nitrification kinetics are such that once nitrification is achieved, it is virtually complete; even though the effluent standards may require an effluent ammonia concentration of less than, say, 10 mgN/ℓ, once nitrification takes place the concentration is likely to be less than 4 mgN/ℓ. Consequently for nitrification, the sludge age of the process is fixed completely by the requirement for nitrification. (The method for calculating the minimum sludge age for nitrification is given in Chapter 5, Section 3.)

With low alkalinity wastewaters (like those encountered in the Western and Southern Cape), nitrification can cause a significant reduction in effluent pH, often as low as 5. This not only causes problems with the nitrification process itself in that only partial nitrification is achieved, with the likelihood of non-compliance of the 10 mgN/ℓ effluent ammonia standard (see Chapter 5, Section 4.3), but it also tends to favour the development of poor settling sludges and to produce aggressive effluents that can do considerable damage to concrete surfaces. To reduce these problems (and to derive other advantages cited later in this section) the policy of deliberate biological denitrification has evolved. However, once biological denitrification is incorporated in the process, sludge ages longer than 10 to 15 days are required and the process falls into the long sludge age category.

Comparing intermediate sludge age plants with high rate plants, the oxygen demand per kgCOD (including nitrification) is doubled (see Figure 6.1), the process volume is 3 to 4 times larger (see Figure 4.3), the daily sludge mass wasted is reduced by 40% (see Figure 4.4) and active fraction is much lower (see Figure 4.2). Intermediate sludge age plants are much more stable than high rate plants, requiring less sophisticated control techniques or operator intervention thereby making these plants more suitable for general application.

At intermediate sludge ages, the active fraction of the waste sludge is still too high for direct discharge to drying beds. Consequently some form of waste sludge stabilization would need to be incorporated in the treatment works, i.e. either aerobic or anaerobic digestion. The former has the advantage of ease of operation but the disadvantage of energy costs for oxygen supply; the latter has the advantage of energy production but the disadvantage of complexity of operation. Even with energy recovery by anaerobic digestion of waste sludge, because of the low mass of sludge wasted from the activated sludge plant and high oxygen demand per kgCOD load, energy self-sufficiency at intermediate sludge ages is impossible. However, on large plants (approximately 500 000 person equivalent) where technical supervision and operator expertise are of a high level, energy cost can be reduced by gas production from anaerobic digesters and probably can be justified economically, particularly if energy costs continue to increase as they have over the past decade.

In nitrifying aerobic activated sludge plants, there is always the possibility that denitrification can take place in the secondary settling tank. This problem is exacerbated by the process control procedure of abstracting the waste sludge from the settling tank underflow (see Section 10 above). This procedure is used to benefit from the thickening function of the settling tank: the lower the underflow recycle ratio, the higher the underflow concentration, the lower the volume of waste sludge. However, low underflow recycle ratios result in large sludge accumulations in the secondary settling tank (see Figure 4.8). Hence the sludge might be retained in the settling tank for a long period of time and this is likely to lead to denitrification. The degree of denitrification is increased as:

(i) the residence time in the settling tank increases – this depends on the recycle ratio and peak flow conditions;

(ii) the active fraction of the sludge increases, i.e. is greater at shorter sludge ages, Figure 4.2;

(iii) the temperature increases; and

(iv) the mass of biodegradable COD adsorbed on the organism mass increases (van Haandel, Ekama and Marais, 1981) – this is generally greatest at the peak load condition.

If denitrification in the settling tank takes place to a sufficient degree, attachment of sludge particles to the nitrogen gas bubbles causes buoying of the particles to the surface. The flotation effect can cause considerable loss of sludge via the settling tank overflow. This behaviour has been observed, for example, at Bellville, Cape Town, where sludge losses have been observed during that latter part of summer when the water temperatures are around 22°C. Sludge loss occurs daily in the afternoon after the peak flow and load when both the sludge accumulation in the settling tank and stored COD adsorbed on the sludge mass are at a maximum. Although the underflow recycle ratio is quite high ($s = 2$) sufficient sludge still accumulates to make denitrification significant at high temperatures.

The above discussion on experience with nitrifying activated sludge plants points to the following conclusions:

The secondary settling tank no longer should serve the dual purpose of solid-liquid separation and thickening. The sludge residence time needs to be minimized, to reduce denitrification, by having a high underflow recycle ratio, of 1 to 2:1, but this will result in inadequate thickening of the waste sludge withdrawn from the underflow. The thickening aspect can be solved by implementing hydraulic control of the sludge age, by providing a small independent satellite settling tank for gravity thickening of the waste sludge (the supernatant being returned to the reactor so that inadequate clarification due to denitrification flotation does not affect effluent quality) or by replacing the gravity satellite tank with a dissolved air flotation unit.* These modifications, however, ameliorate the effects on the secondary settling tank but do not positively remove the root cause, i.e. the high nitrate concentration in the mixed liquor. This can be done only by including denitrification in the process, which then shifts the process to that of the long sludge age type.

11.3 Long sludge ages (20 days or more)

11.3.1 Aerobic plants:

Long sludge age aerobic plants are usually called extended aeration plants. The sludge age is chosen to be of such a length that the active mass fraction of the waste sludge is sufficiently reduced to allow discharge directly to sludge drying beds after thickening. What this sludge age should be will depend to a degree on the climatic conditions i.e. whether the sludge can be dried before it starts smelling, but probably exceeds 30 to 40 days – research into the sludge ages necessary under South African conditions is necessary.

Compared to intermediate sludge age plants the total oxygen demand (i.e. carbonaceous plus nitrification) is about equivalent (Figure 6.1); the process

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*Thickening by dissolved air flotation now is both technically and economically feasible, Bratby and Marais (1976, 1977).

volume requirement is 50 to 60% higher and the active fraction 50% lower. As nitrification always occurs, problems with rising sludge in the settling tank are to be expected. Control of the sludge age is best via the hydraulic sludge age control method, so that high recycles from the secondary settling tank can be imposed to reduce rising sludge in the tank, as described under intermediate sludge age processes above. Problems of low pH when treating low Alkalinity wastewater are to be expected.

Extended aeration plants are very stable in operation and require probably less supervision than any other activated sludge process. Although the volume requirements and the oxygen demand per unit input COD is large, the relative ease of operation makes this process the indicated one for small communities. However, it must be remembered that this process will discharge a low COD effluent but with high nitrate and phosphate concentrations. The nitrate concentration can be drastically reduced and low pH effluent prevented by incorporating an anoxic zone in the process without basically disturbing the ease of operation. This modification always should be kept in mind when considering extended aeration as a possible process.

Processes that can be operated in the extended aeration modes are the completely mixed single reactor, the Orbal multi-channel, oxidation ditch and Carrousel processes.

11.3.2 Anoxic-aerobic process

Once sludge ages in the range greater than 20 days are accepted it is possible usually to incorporate denitrification in the process without upsetting stability in nitrificaion. The reactor is suitably subdivided into unaerated (anoxic) and aerated zones in a variety of ways, the denitrification taking place in the unaerated zones and the nitrification in the aerated zones to give the so-called nitrification-denitrification process. The advantages are substantial: (1) Nitrate concentration in the effluent is reduced, the nitrate utilized in the unaerated zones lowering the oxygen demand of the process accordingly; with complete denitrification the total oxygen demand can be reduced by approximately 15 to 20% compared with the requirements for a nitrifying process. (2) Loss of sludge in the overflow from the settling tank due to rising sludge is largely eliminated. (3) Alkalinity recovery with denitrification causes pH instability with low influent alkalinity wastewaters to be very unlikely. Normally the sludge ages in nitrification-denitrification plants are insufficient to allow direct discharge of the waste sludge to the drying beds.

Process configurations of anoxic-aerobic systems are the Modified Ludzack-Ettinger process in which a primary anoxic reactor is provided, the Bardenpho which incorporates both primary and secondary anoxic reactors, the Orbal in which anoxic zones are created by reducing the aeration energy input in channels, Carrousel and oxidation ditch in which the aeration at the aerators is controlled to induce anoxic operation in the same reactor, or in two interlinking parallel reactors.

Incorporation of biological denitrification in a process imposes additional constraints on the design. It requires the selection of a process sludge age *and* an unaerated mass fraction, (to achieve the required denitrification or the highest denitrification possible), yet ensuring efficient nitrification under all expected conditions. (See Chapter 5, Section 4.4, and Chapter 6, Section 6).

11.3.3 Anaerobic-anoxic-aerobic systems

These plants are needed where biological removal of both nitrogen and phosphorus is required and are called biological excess P removal plants. The phosphorus removal propensity is created by providing a zone in the process that receives the influent wastewater but not oxygen and nitrate (via the recycles of mixed liquor). The incorporation of this zone imposes further constraints on the design of the system: A sludge age must be selected that will ensure efficient nitrification at all times, provide for anoxic zones that will give complete or maximal removal of nitrate and an anaerobic zone that receives the influent flow and recycles of mixed liquor not containing nitrate generated in the system. The critical aspects of the design are the assurance of no discharge of nitrate to the anaerobic zone. The success of the design is sensitive to a number of factors: the influent wastewater characteristics, such as the readily biodegradable COD concentration, the TKN/COD and P/COD ratios, the maximum specific growth rate of the nitrifiers, maximum and minimum temperatures, severity of cyclic load and flow, control of aeration and control of nitrate and dissolved oxygen in some of the recycles.

Aeration control is a particularly vexing problem under cyclic load and flow conditions because the process is affected by too high or too low oxygen concentrations in the aerobic zone. Too high concentrations cause oxygen to be recycled to the anoxic zone (and the anaerobic zone in the 5 stage Phoredox process), thereby reducing the potential for P and N removal; too low oxygen concentrations cause the nitrification efficiency to decline and poor settling sludges to develop.

Oxygen control setups that have been developed still require considerable operational attention, particularly the oxygen probes, and at present their installation in smaller plants is to be questioned. These difficulties have prompted research into an alternative solution – load and flow equalization. In this approach an equalization tank is provided upstream of the plant. The effluent flow from the tank is controlled in such a manner that the cyclic fluctuations in the flow *and* load are damped to very small values. The control is via a microcomputer that calculates the discharge flow setting. The essential inputs to the computer are the depth of liquor in the equalization tank and the flow rate out of the tank; both these measurements can be readily monitored automatically and operated realiably over long periods of time. The microcomputer also has reached a stage of development that ensures its long term trouble-free service. The equalization approach with the microcomputer control, has been tested at Goudkoppies plant, Johannesburg and shows great potential for reducing aeration and other control problems in nutrient removal plants. Detailed information on equalization tanks is available in a report by Dold, Buhr and Marais (1982). Design details of the P removal processes, considered in this monograph, are given in Chapter 7.

It should be noted that the waste sludge from a biological excess P removal activated sludge process contains high concentrations of P. If the sludge is allowed to become anaerobic or is anaerobically digested much of the phosphorus removed from the wastewater will be released from the sludge mass to the bulk liquid; consequently seepage from drying beds may have to be chemically treated to precipitate the phosphorus before recycling to the plant. Failure to recognize this behaviour of phosphorus-rich biological sludges may lead to inadvertent re-contamination of the effluent with high concentrations of phosphorus (see Section 8 above).

Examples of N and P removal processes are the multi-reactor Phoredox and UCT processes in which the anaerobic, anoxic and aerobic conditions are induced in separate reactors interlinked by recycle flows. Other removal systems have been proposed, for example that by Kerdachi and Roberts (1982) in which these conditions are imposed within the same reactor.

Where P removal without obligatory nitrogen removal is required the sludge age can be selected to prevent nitrification and in the multi-reactor type plant theoretically only anaerobic and aerobic reactors are needed, i.e. no anoxic reactor. The sludge age is reduced substantially, probably to less than eight days, so that the reactor volume per input COD is much smaller than where nitrification-denitrification is a requirement. However, even with these short sludge ages it would be wise, always, to make provision for denitrification of any nitrate generated and recycled to the anaerobic zone. This provision is of particular relevance in warm climates or where the maximum specific growth rate of the nitrifiers is very high so that nitrification can occur at *aerobic* sludge ages of, say, 2 days. These short sludge age P removal plants are not considered in this monograph.

12. DESIGN EXAMPLE

In this section, the procedure for design of organic material (COD) degradation in purely aerobic processes is demonstrated with numerical examples. Assuming constant flow and load conditions, calculations are presented showing how estimates of the process volume requirements, average daily carbonaceous oxygen demand and daily sludge production are obtained for the treatment of raw or settled wastewaters based on the wastewater characteristics.

12.1 Influent Wastewater Characteristics

The daily average characteristics for a particular raw wastewater are given in Table 4.3. In the design example it is assumed that primary sedimentation tanks remove 40% of the total COD load, 15% of the TKN and total P and 10% of the readily biodegradable COD. The settled wastewater characteristics are also given in Table 4.3. It should be noted that the *soluble concentrations* in the settled wastewater should be approximately equal to those in the raw wastewater. For example, from Table 4.3 the soluble unbiodegradable COD concentration in the raw wastewater is 0,05 mg COD/mg COD, i.e. 0,05.600 = 30 mg COD/ℓ (Eq. 2.3). As the soluble unbiodegradable constituents are not affected by primary sedimentation, the unbiodegradable soluble COD *fraction* of the settled wastewater accordingly will be approximately 30/360 = 0,08 mg COD/mg COD. Similarly, the readily biodegradable COD fraction (f_{bs}) of the raw wastewater is given as 0,24. This is equivalent to a concentration of $0{,}24\,(1-f_{up}-f_{us})S_{ti} = 0{,}24(0{,}82)600 = 118$ mg COD/ℓ (see Eqs. 2.8 and 2.9). Primary sedimentation is assumed to remove 10% of the readily biodegradable COD (due to biodegradation), the f_{bs} *fraction* in the settled wastewater is given by $\{(1{,}0-0{,}10)118\}/\{(1-0{,}08-0{,}04)360\} = 0{,}33$ (see Eqs. 2.8 and 2.9). The examples show that care must be taken that settled wastewater characteristics are not fixed at values that are inconsistent with the raw wastewater characteristics and primary settling tank behaviour.

12.2 Temperature effects

From Table 4.1, the only constant in the organic material degradation theory that is affected by

TABLE 4.3 RAW AND SETTLED WASTEWATER CHARACTERISTICS (SEE ALSO TABLES 5.2 AND 6.1)

Parameter	Symbol	Value Raw	 Settled	Units
Influent COD concentration	S_{ti}	600*	360*	mgCOD/ℓ
Influent TKN concentration	N_{ti}	48*	41*	mgN/ℓ
Influent P concentration	P_{ti}	10*	8,5*	mgP/ℓ
TKN/COD ratio	—	0,080	0,114	mgN/mgCOD
P/COD ratio	—	0,017	0,024	mgP/mgCOD
Unbiodegradable soluble COD	f_{us}	0,05	0,08	mgCOD/mgCOD
Unbiodegradable particulate COD	f_{up}	0,13	0,04	mgCOD/mgCOD
MLVSS/MLSS ratio of sludge produced	f_i	0,75	0,83	mgVSS/mgTSS
Temperature – Summer	T_{max}	22	22	°C
– Winter	T_{min}	14	14	°C
pH of wastewater	—	7,5	7,5	—
Influent flow	Q	13,33	13,33	Mℓ/d

*Influent concentrations appropriately weighted to take account of daily cyclic flow variations temperature is the specific endogenous respiration rate, b_h. This rate reduces by about 3% every 1°C drop in temperature – from Eq. 4.5, (given in Table 4.1), the rate at 14°C is 0,20/d and at 22°C is 0,25/d. The effect of the reduction in the rate with decrease in temperature is that at reduced temperatures the daily sludge production is marginally increased and the average carbonaceous oxygen demand is marginally decreased – the differences in sludge production and oxygen demand are less than 5% for a 6°C change in temperature. Consequently, the average carbonaceous oxygen demand should be calculated at the maximum temperature and the process volume and sludge production at the minimum temperature in order to find the maximum values of these parameters.

12.3 Calculations for organic material degradation

The example will demonstrate the effect of temperature and sludge age on (i) the mass of sludge in the process, (ii) the average daily carbonaceous oxygen demand, (iii) the active fraction of the sludge and (iv) mass of sludge wasted daily. These four parameters will be calculated for the raw and settled wastewater at 14°C and 22°C for sludge ages ranging from 5 to 30 days. The values of the constant Y_h, f_{cv}, f and b_h (appropriately adjusted for temperature) are given in Table 4.1.

Mass of COD treated per day $= M(S_{ti}) = Q.S_{ti}$ (4.6)

Mass of biodegradable COD $= M(S_{bi}) = (1 - f_{up} - f_{us})M(S_{ti})$ (4.7c)

Mass of unbiodegradable particulate solids $M(X_{ii}) = M(S_{ti})f_{up}/f_{cv}$ (4.8c)

Hence for raw wastewater:

$$M(S_{ti}) = 13,33.10^6.600 \text{ mg COD/d} = 8000 \text{ kg COD/d}$$

$$f_{us} = 0,05 \text{ mg COD/mg COD}$$

$$f_{up} = 0,13 \text{ mg COD/mg COD}$$

$$M(S_{bi}) = (1 - 0,05 - 0,13)8000 \qquad = 6560 \text{ kg COD/d}$$

$$M(X_{ii}) = 8000.0,13/1,48 \qquad = 702,7 \text{ kg VSS/d}$$

and for settled sewage:

$$M(S_{ti}) = 60\% \text{ of } 8000 \qquad = 4800 \text{ kg COD/d}$$

$$f_{us} = 0,08 \text{ mg COD/mg COD}$$

$$f_{up} = 0,04 \text{ mg COD/mg COD}$$

$$M(S_{bi}) = (1 - 0,08 - 0,04)4800 \qquad = 4224 \text{ kg COD/d}$$

$$M(X_{ii}) = 4800.0,04/1,48 \qquad = 129,7 \text{ kg VSS/d}$$

From Eq. 4.13 the mass of volatile solids in the process is given by

$$M(X_v) = \frac{0,45\, R_s\, M(S_{bi})}{(1 + b_{hT}\, R_s)}(1 + 0,2\, b_{hT}\, R_s) + M(X_{ii})\, R_s$$

$$= \{\frac{2952\,(1 + 0,2\, b_{hT}\, R_s)}{(1 + b_{hT}\, R_s)} + 702,7\}\, R_s \tag{4.29a}$$

for raw wastewater; and

$$M(X_v) = \{\frac{1901\,(1 + 0,2\, b_{hT}\, R_s)}{(1 + b_{hT}\, R_s)} + 129,7\}\, R_s \tag{4.29b}$$

for settled wastewater.

From Eq. 4.14 the mass of total suspended solids in the process is given by

$$M(X_t) = M(X_v)/0,75 \text{ for raw wastewater} \tag{4.30a}$$

$$= M(X_v)/0,83 \text{ for settled wastewater} \tag{4.30b}$$

From Eq. 4.15, the average daily carbonaceous oxygen demand is given by

$$M(O_c) = M(S_{bi})\{(1 - f_{cv}Y_h) + f_{cv}(1 - f)b_{hT}\frac{Y_h\, R_s}{(1 + b_{hT}\, R_s)}\}$$

$$= 6560\,\{(0,334) + 0,533\frac{b_{hT}\, R_s}{(1 + b_{hT}\, R_s)}\}\text{ (kg O/d)} \tag{4.31a}$$

for raw wastewater; and

$$= 4224\,\{(0,334 + 0,533\frac{b_{hT}\, R_s}{(1 + b_{hT}\, R_s)}\}\text{ (kg O/d)} \tag{4.31b}$$

for settled wastewater.

From Eq. 4.18, the active fraction with respect to the volatile suspended solids (f_{av}) is given by

$$f_{av} = 1/\{1 + 0,2\, b_{hT}\, R_s + 0,238\,(1 + b_{hT}\, R_s)\}$$

$$= 1/\{1,238 + 0,438\, b_{hT}\, R_s\} \text{ for raw wastewater.} \tag{4.32a}$$

$$f_{av} = 1/\{1 + 0,2\, b_{hT}\, R_s + 0,068\,(1 + b_{hT}\, R_s)\}$$

$$= 1/\{1,068 + 0,268\, b_{hT}\, R_s\} \text{ for settled wastewater} \tag{4.32b}$$

From Eq. 4.22, the mass of sludge (as total suspended solids) produced per day is given by

$$M(\Delta X_t) = \frac{8000}{0,75}\{\frac{0,82.0,45}{(1 + b_{hT}\, R_s)}(1 + 0,2\, b_{hT}\, R_s) + \frac{0,13}{1,48}\}$$

$$= 10667\,\{\frac{0,369\,(1 + 0,2\, b_{hT}\, R_s)}{(1 + b_{hT}\, R_s)} + 0,09\} \tag{4.33a}$$

for raw wastewater; and

$$M(\Delta X_t) = \frac{8000}{0,83}\{\frac{0,88.0,45}{(1 + b_{hT}\, R_s)}(1 + 0,2\, b_{hT}\, R_s) + \frac{0,04}{1,48}\}$$

$$= 5783\,\{\frac{0,396\,(1 + 0,2\, b_{hT}\, R_s)}{(1 + b_{hT}\, R_s)} + 0,03\} \tag{4.33b}$$

for settled wastewater.

Substituting the b_{hT} value for 14°C (i.e. 0,202/d) and 22°C (i.e. 0,254/d) into Eqs. 4.25 to 4.29 the parameters $M(X_v)$, $M(X_t)$, $M(O_c)$, f_{av} and $M(\Delta X_t)$ can be calculated for sludge ages 5 to 30 days. The results are shown plotted in Figure 4.9.

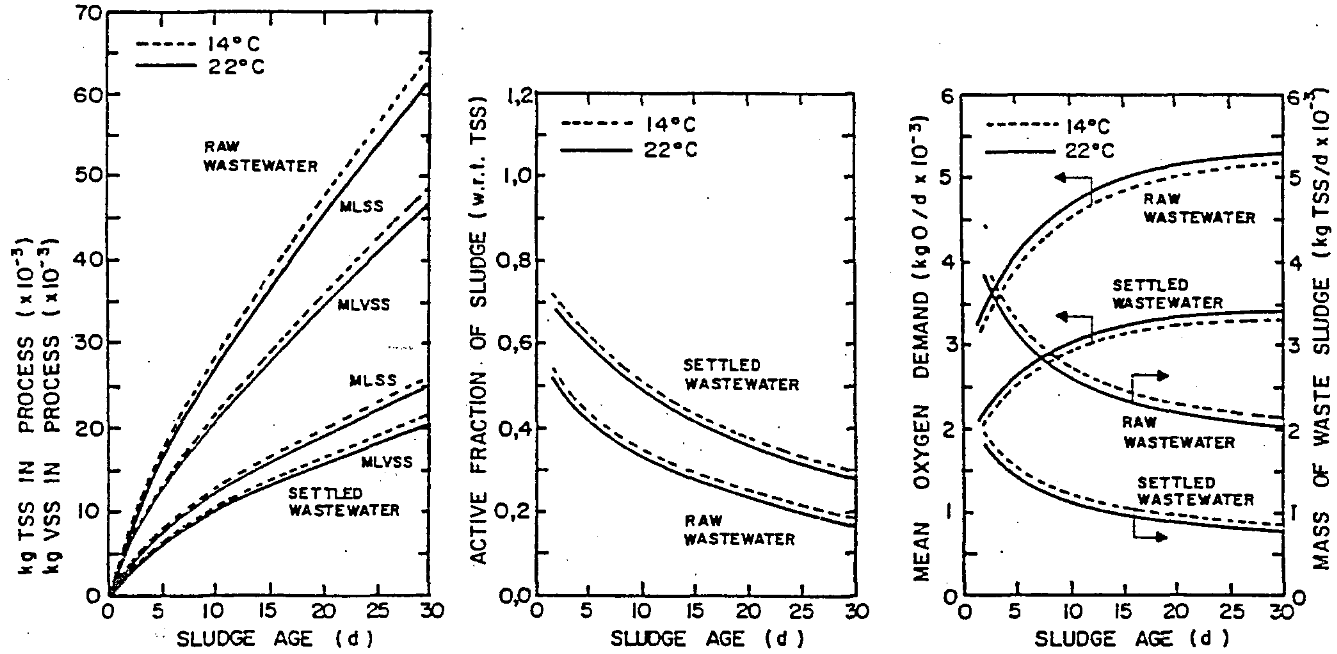

FIGURE 4.9: Mass of sludge (MLSS and MLVSS), active fraction (with respect to TSS), average carbonaceous oxygen demand and mass of sludge (as TSS) produced daily versus sludge age for raw and settled wastewaters at 14°C and 22°C. Wastewater characteristics given in Table 4.3.

Figure 4.9 confirms that the mass of sludge in the process (MLSS or MLVSS), the average carbonaceous oxygen demand and the active fraction (with respect to MLSS or MLVSS) are only marginally affected by temperature and for these parameters, insofar as design is concerned, are not really of consequence. However, the influent waste type i.e. raw or settled wastewater, has a significant effect; raw wastewater results in more sludge in the process, a higher oxygen demand and a lower active fraction of the sludge than settled wastewater. The difference in effects between raw and settled wastewater depends wholly on the efficiency of the primary settling tanks – the differences apparent in Figure 4.9 arise from a 40% COD removal in primary sedimentation. The greater the efficiency of primary sedimentation the greater the difference between the parameters shown in Figure 4.9.

From Eq. 4.16 for the same MLSS concentration, the process volume is proportional to the mass of sludge contained in it. Hence for the same MLSS concentration, the volume of the process treating settled wastewater will be only 40% of that treating raw wastewater at 25 days sludge age. Also, the settled wastewater plant requires only 64% of the oxygen the raw wastewater plant requires. However, the active fraction of the sludge in the settled wastewater plant is 34%, probably too high for direct discharge to drying beds, whereas that from the raw wastewater plant is 22%, probably sufficiently low for discharge to drying beds without further stabilization. Clearly the choice of treating settled sewage as against raw sewage requires weighing the advantages and disadvantages of each against the other, i.e. settled sewage (smaller process volume, smaller oxygen demand and smaller secondary sludge production, but, with primary sludge production and its disposal) against raw sewage (greater process volume, higher oxygen demand and higher secondary sludge production, but no primary sludge production). The choice is made more difficult by the fact that higher biological removals of N and P are achieved with raw wastewater than with settled wastewaters (see Chapter 3, Section 3). These aspects are discussed in detail in Chapters 6 and 7 and will be demonstrated by extending the above design example to include biological nitrogen and phosphorus removal.

13. CONCLUSIONS

The calculations for the daily sludge production, daily average carbonaceous oxygen demand, process volume and nitrogen and phosphorus requirements for sludge production etc. were demonstrated for a raw and a settled sewage with selected wastewater characteristics for the raw sewage and assumed effects of primary settlement on the raw sewage characteristics. Selection of the numerical magnitudes of these characteristics was based on experience with five municipal waste flows so that the data should not be taken to apply generally. Whenever possible, the characteristics should be determined experimentally, or estimated by some semi-experimental procedure supplemented by experience. Where design information is inadequate it will be necessary to select ranges of values for some parameters, ranges that appear to span across the true values, and to repeat the design a number of times using various combinations of the limiting values. In this way a picture will be built up of the sensitivity of the plant process response so that appropriate provisions can be made in the design. There are so many interactive effects that ample design provision in one aspect, for example, does not necessarily resolve that problem in terms of the whole design as it may create problems in other aspects.

14. REFERENCES

ARKLEY, M. and MARAIS, G.v.R. (1982) Effect of the anoxic sludge mass fraction on the activated sludge process. Research Report No. W 38, Dept. of Civil Eng., Univ of Cape Town.

BRATBY, J. and MARAIS, G.v.R. (1976) A guide for the design of dissolved-air (pressure) flotation systems for activated sludge processes. *Water SA*, **2**, 2, 86-100.

BRATBY, J. and MARAIS, G.v.R. (1978) Aspects of sludge thickening by dissolved-air flotation. *Wat. Pollut. Control*, **77**, 3, 421-432.

DOLD, P.L., EKAMA, G.A. and MARAIS, G.v.R. (1980) A general model for the activated sludge process. *Prog. Wat. Tech.*, **12**, 47-77.

DOLD, P.L., BUHR, H.O. and MARAIS, G.v.R. (1982) Design and control of equalization tanks. Research Report No. W 42, Dept. of Civil Eng., Univ of Cape Town.

EKAMA, G.A. and MARAIS, G.v.R. (1978) The dynamic behaviour of the activated sludge process. Research Report No. W 27, Dept. of Civil Eng., Univ. of Cape Town.

GARRETT, M.T. (1958) Hydraulic control of activated sludge growth rate. *Sew. and Ind. Wastes*, **30**, 253.

KERDACHI, D. and ROBERTS, M. (1982) Full scale phosphate removal experiences in the Umhlatuzana works at different sludge ages. *Wat. Sci. Tech.*, **15**, 3/4, 262-282.

MARAIS, G.v.R. and EKAMA, G.A. (1976) The activated sludge process part 1 – Steady state behaviour. *Water SA*, **2**, 4, 163-200.

MARAIS, G.v.R. and EKAMA, G.A. (1980) Aerated lagoons. Research Report, Dept. of Civil Eng., Univ. of Cape Town.

McKINNEY, R.E. and SYMONS, J.M. (1964) Growth and endogenous metabolism – a discussion. Proc. 1st Int. Conf. on Wat. Pollut. Control, London, 1962, Pergamon Press, Oxford.

NICHOLLS, H.A. (1975) Internal Thesis Report, Dept. of Civil Eng., Univ. of Cape Town.

PAYNE, W.J. (1970). Energy yields and growth of heterotrophs. Annual reviews – microbiology, 17-51.

SEHAYEK, E. and MARAIS, G.v.R. (1981) Kinetics of biological nitrogen removal in the single sludge activated sludge process. Research Report No. W 41, Dept. of Civil Eng., Univ. of Cape Town.

SETTER, L.R., CARPENTER, W.T. and WINSLOW, G.C. (1945) Practical application of modified sewage aeration. *Sew. Works Jour.*, **17**, 4, 669.

VAN HAANDEL, A.C., EKAMA, G.A. and MARAIS, G.v.R. (1981) The activated sludge process – 3 – Single sludge denitrification. *Water Research*, **15**, 1135-1152.

VAN HAANDEL, A.C. and MARAIS, G.v.R. (1981) Nitrification and denitrification in the activated sludge process. Research Report No. W 39, Dept. of Civil Eng., Univ. of Cape Town.

CHAPTER FIVE

NITRIFICATION

by

G.A. Ekama and G.v.R. Marais

1. INTRODUCTION

The term nitrification describes the biological process whereby free and saline ammonia is oxidized to nitrite and nitrate. Nitrification is mediated by *specific autotrophic organisms* with behavioural characteristics that differ significantly from the heterotrophic ones. The objectives in this chapter are to review briefly the kinetics of nitrification, to highlight the factors that influence this biological reaction and set out the procedure for designing a nitrifying-aerobic activated sludge process.

It has been well established that nitrification is due to two specific genera of autotrophic bacteria, the *Nitrosomonas* and *Nitrobacter*. Nitrification takes place in two sequential oxidation steps: (1) *Nitrosomonas* convert free and saline ammonia to nitrite, and (2) *Nitrobacter* convert nitrite to nitrate. The nitrifiers utilize ammonia for their synthesis nitrogen requirements, and as a source of energy to bring about synthesis. The ammonia requirement for synthesis, however, is a negligible fraction of the total ammonia processed by the organisms, at the most 2% (van Haandel and Marais, 1981). Consequently it is usual to neglect the metabolic nitrogen requirements* of the nitrifiers and to consider the nitrifiers simply to act as catalysts for the ammonia-nitrite-nitrate reactions – the reactions are taken as stoichiometric. This greatly simplifies the description of the kinetics of the process.

Consider the two basic redox reactions in nitrification,

$$NH_4^+ + 3/2\ O_2\ (\textit{Nitrosomonas}) \rightarrow NO_2^- + H_2O + 2H^+ \qquad (5.1a)$$

$$NO_2^- + 1/2\ O_2\ (\textit{Nitrobacter}) \rightarrow NO_3^- \qquad (5.1b)$$

Stoichiometrically the oxygen requirements for the first and second reactions are 3,43 and 1,14 mgO/mgN [also written as mgO/mg(NH$_3$-N)]. Hence the conversion of ammonia to nitrate, both expressed as N, requires 4,57 mgO/mgN utilized. Taking into account the ammonia utilized for metabolic purposes of the nitrifiers, the oxygen requirements per mg(NH$_3$-N) processed are slightly less, with reported values down to 4,3 mgO/mg (NH$_3$-N). However, such a sophisticated approach is more of academic than practical interest and we will accept the stoichiometric value, 4,57 mgO/mg(NH$_3$-N).

2. BIOLOGICAL KINETICS

2.1 Growth

In order to formulate the nitrification behaviour it is necessary to understand the basic biological growth kinetics of nitrification. Here one has to consider the two stage nitrification behaviour. The rate of conversion of ammonia to nitrate, by the *Nitrosomonas* is much slower than that of nitrite to nitrate, by the *Nitrobacter*. In fact, under most circumstances in municipal wastewater treatment plants, any nitrite that is formed is converted virtually immediately to nitrate. As a consequence very little nitrite is observed in the effluent from a plant operating on an influent that does not contain substances that inhibit the *Nitrobacter*. The limiting rate in the nitrification sequence, therefore, is that due to the *Nitrosomonas* – one needs to consider the kinetics of this organism only. Because the nitrite produced is virtually immediately converted to nitrate in terms of the growth rate of the *Nitrosomonas*, it can be assumed that the conversion is from ammonia to *nitrate* directly – on this basis the kinetics of nitrification reduce to the kinetics of the *Nitrosomonas*.

Experimental investigations by Downing, Painter and Knowles (1964) showed that the nitrification rate can be formulated in terms of Monod's relationship. Monod found that: (1) The mass of organisms generated is a fixed fraction of the mass of substrate (in this case ammonia) utilized. (2) The *specific rate of growth*, that is, the rate of growth per unit mass of organisms per unit time, is related to the concentration of substrate surrounding the organisms.

From (1):

$$M(\Delta X_n) = Y_n\, M(\Delta N_a) \qquad (5.2)$$

where

$M(\Delta X_n)$ = mass of nitrifiers generated, (mgVSS)

$M(\Delta N_a)$ = mass of ammonia as N utilized, [mg(NH$_3$-N)]

Y_n = mass of nitrifiers generated per unit mass of ammonia utilized, [mgX$_n$/mg(NH$_3$-N)]

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*In nitrification processes, where the influent energy source consists principally of ammonia, this simplification no longer can be applied and the determination of the nitrifying sludge mass is necessary to design the process volume (Neytzel-de Wilde, 1977). With municipal wastes under *cyclic* flow conditions the nitrifying mass also needs to be calculated.

Consequently one can write

$$\frac{dX_n}{dt} = Y_n \frac{dN_a}{dt} \tag{5.3}$$

From (2) Monod developed the following relationship:

$$\mu_{nT} = \frac{\mu_{nmT}\, N_a}{K_{nT} + N_a} \tag{5.4}$$

where

μ_n = specific growth rate observed at concentration N_a, $(mgX_n/mgX_n/d)$

μ_{nm} = maximum specific growth rate possible $(mgX_n/mgX_n/d)$

K_n = half saturation constant, i.e. the concentration at which $\mu_n = \frac{1}{2}\mu_{nm}$, $[mg(NH_3\text{-}N)/\ell]$

N_a = concentration of ammonia surrounding the organisms, $[mg(NH_3\text{-}N)/\ell]$

T subscript refers to temperature, °C

The *growth rate* is given by the product of the specific growth rate and the organism mass

$$\frac{dX_n}{dt} = \mu_{nT} \cdot X_n = \frac{\mu_{nmT}\, N_a}{K_{nT} + N_a} . X_n \tag{5.5}$$

Instead of expressing the growth rate in terms of the mass of organisms, it is often useful also to express it in terms of the depletion rate of ammonia, as follows: From Eq. 5.3 substituting for dX_n/dt in Eq. 5.5

$$\frac{dN_a}{dt} = \frac{(\mu_{nmT}/Y_n) N_a}{K_{nT} + N_a} X_n \tag{5.6}$$

Equation 5.6 is sometimes written in an equivalent form as follows: Define the maximum specific substrate utilization rate, $K_{mT} = \mu_{nmT}/Y_n$ then the specific substrate utilization rate K_T becomes from Eq. 5.4

$$K_T = \frac{K_{mT}\, N_a}{K_{nT} + N_a} \tag{5.7}$$

and the rate of substrate utilization, from Eq. 5.6

$$\frac{dN_a}{dt} = \frac{K_{mT}\, N_a}{K_{nT} + N_a} . X_n \tag{5.8}$$

This form is mentioned here because it is the one often used to express heterotrophic growth, and in fact is the form utilized in all the work published by the Marais group, because of its greater convenience for some purposes. The point to note is that Eqs. 5.6 and 5.8 are of equal merit.

The ammonia nitrogen that disappears in the nitrification reaction, reappears as nitrate nitrogen because the stoichiometric approach is adopted, i.e.

$$\frac{dN_n}{dt} = \frac{dN_a}{dt} \tag{5.9}$$

Consequently

$$\frac{dN_n}{dt} = \frac{(\mu_{nmT}/Y_n) N_a}{K_{nT} + N_n} . X_n \tag{5.10}$$

where

N_n = nitrate concentration, $(mg(NO_3\text{-}N)/\ell)$.

The oxygen consumption associated with nitrification is based on the stoichiometric assumption mentioned earlier, i.e. conversion of 1 $mg(NH_3\text{-}N)$ to 1 $mg(NO_3\text{-}N)$ requires 4,57 mgO. Consequently

$$\frac{dO_n}{dt} = 4{,}57 \frac{dN_a}{dt} \tag{5.11}$$

2.2 Endogenous respiration

Up to this stage attention has been focussed on the growth aspect. However, Herbert (1958) showed that generally an organism mass undergoes a continuous mass loss, called endogenous mass loss. Furthermore from the great amount of research undertaken by the Marais group it seems that this loss occurs independently of the growth. This allows straightforward formulation of endogenous mass loss as,

$$\frac{dX_n}{dt} = -b_{nT} X_n \tag{5.12}$$

where

b_{nT} = specific endogenous mass loss rate for *Nitrosomonas* (mg/mg/d)

2.3 Behavioural Characteristics

In Figure 5.1 the relationship between the specific growth rate, μ_n, the specific substrate utilization rate, K, and the substrate concentration, N_a is shown, as described by Monod's equation Eqs. 5.4 and 5.7 respectively. The rate constants selected are $\mu_{nm20} = 0{,}33$, $Y_n = 0{,}10$ (hence $K_m = 3{,}3$) and $K_{n20} = 1{,}0$. The interesting feature of this plot is that, because K_n is so small $\cong 1$ $mg(NH_3\text{-}N)/\ell$, the specific rate is virtually at a maximum for concentrations of 2 $mg(NH_3\text{-}N)/\ell$ and higher. However at concentrations less than two the rate rapidly declines to zero so that the ammonia is not readily reduced to zero.

3. PROCESS KINETICS

As indicated in Chapter 4 the basic process unit is the completely mixed activated sludge unit. The process under steady state provides the information necessary for design.

With regard to nitrification, for design, the principal steady state solution required is that of the effluent ammonia concentration from a completely mixed aerobic reactor. This solution forms the basis for extensive analysis of the process behaviour. A comprehensive analysis is not required here; for those interested reference should be made to Ekama, van Haandel and Marais (1979) and van Haandel and Marais (1981).

3.1 Effluent ammonia concentration

Consider nitrification in a completely mixed reactor, under constant flow and load conditions (see Figure 4.1). A mass balance on the accumulation of nitrifier mass $M(\Delta X_n)$ is given by

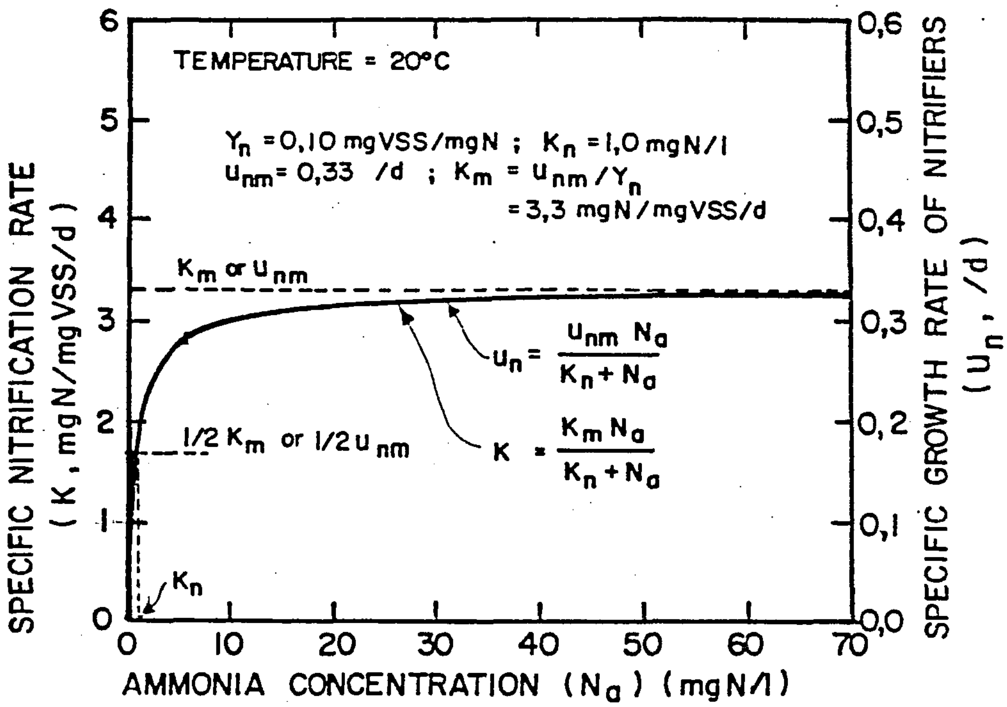

FIGURE 5.1: Monod's relationship for nitrification at 20°C.

$$M(\Delta X_n) = V_p\,\Delta X_n = [\mu_{nmT}.N_a/(K_{nT}+N_a)].X_n.V_p\,\Delta t - b_{nT}.X_n.V_p.\Delta t - X_n.q.\Delta t$$

where

V_p = process volume (ℓ)

q = volumetric withdrawal rate of waste sludge *from the reactor* (ℓ/d)

Dividing by $V_p\,\Delta t$

$$\Delta X_n/\Delta t = [\mu_{nmT}\,N_a/(K_{nT}+N_a)]X_n - b_{nT}\,X_n - X_n\,q/V_p$$

Now

$$\frac{V_p X_n}{q\,X_n} = \frac{\text{Mass of nitrifier sludge in process}}{\text{Mass of nitrifier sludge wasted per day}} = \frac{V_p}{q} = R_s$$

where

R_s = sludge age (d) (see Eq. 4.2).

Under constant flow and load, when steady state is achieved, $\Delta X_n/\Delta t = 0$ and solving for N_a yields,

$$N_a = \frac{k_{nT}\,(b_{nT} + 1/R_s)}{\mu_{nmT} - (b_{nT} + 1/R_s)} \qquad (5.13)$$

It is of interest to note that in Eq. 5.13 the ammonia concentration (N_a) in the reactor (and effluent) is independent of the specific yield constant (Y_n) and the influent ammonia concentration (N_{ai}). Using the same nitrification kinetic constants as those in Figure 5.1 and taking $b_{nT} = 0{,}0$, a plot of Eq. 5.13 with N_a versus sludge age R_s is given in Figure 5.2. Note that at long sludge ages N_a is very low and remains so until the sludge age is lowered to about 3,5 days below which, N_a increases rapidly until, in terms of this equation, N_a can exceed the influent concentration, N_{ai} – evidently the limit of application of the equation is when $N_a = N_{ai}$. Substituting N_{ai} for N_a in Eq. 5.13 and solving for R_s gives the minimum sludge age, R_{sm}, below which theoretically, no nitrification can be achieved. This minimum sludge age varies slightly with the magnitude of N_{ai}, see Figure 5.2 – higher N_{ai} gives a *slightly* lower R_{sm}. Now the effective influent N_{ai} (see later) will rarely be less than about 20 mg/ℓ. Noting that $K_{nT} \cong 1$ mgN/ℓ, then

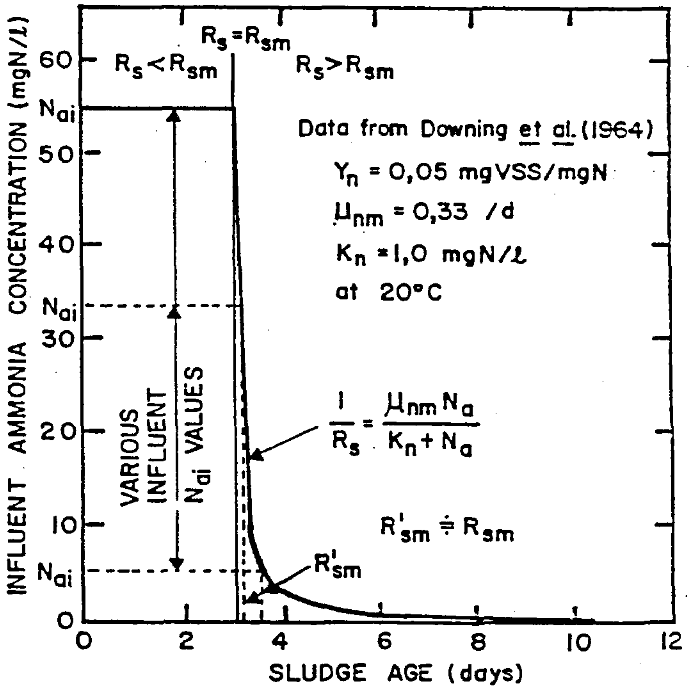

FIGURE 5.2: Relationship between sludge age for nitrification and influent ammonia concentration, N_{ai}.

K_{nT}/N_{ai} is negligibly small with respect to unity. Substituting zero for K_{nT}/N_{ai} in Eq 5.13 and solving for R_{sm}, yields

$$R_{sm} = 1/(\mu_{nmT} - b_{nT}) \qquad (5.14)$$

For practical application, Eq. 5.14 adequately defines the minimum sludge age for all N_{ai} greater than about 5 mgN/ℓ.

The virtually constant value for R_{sm} (for the accepted values of μ_{nmT} and b_{nT}) and the rapid attainment of high nitrification efficiency at sludge ages slightly greater than R_{sm} causes that in a particular plant, as the sludge age is increased, once R_s exceeds R_{sm}, high efficiency of nitrification will be observed. Consequently, under steady state conditions one would expect a plant either not to nitrify, or, to nitrify with high efficiency depending on whether R_s is shorter or longer than R_{sm} respectively. This behaviour is well attested by observation on laboratory scale plants.

From the discussion above clearly the magnitudes of μ_{nmT} and b_{nT} affect R_{sm}; it is of importance, therefore, for practical application, to enquire into those factors that influence these constants and the nitrification phenomena in general.

4. FACTORS INFLUENCING NITRIFICATION

A number of factors affect the nitrification rate constants, efficiency of nitrification and the maximum sludge age. These are (1) wastewater source, (2) temperature, (3) pH, (4) unaerated zones, (5) dissolved oxygen concentration, and (6) cyclic flow and load.

4.1 Influent source

The maximum specific growth rate constant μ_{nmT} has been observed to be specific to the source of the waste flow and, even then, to vary between different batches from the same source. This specificity is so marked that μ_{nmT} should be classified as a wastewater characteristic. The effect appears to be of an inhibitory nature due to some substance(s) in the wastewater. It is not toxicity because high efficiency of nitrification can be achieved even with a low μ_{nm} value if the sludge age is increased sufficiently. These inhibitory substances are more likely to be present in effluents having some industrial waste components and in general, the higher this fraction, the lower μ_{nmT} tends to be, but the specific waste fractions that cause the reduction of μ_{nmT} have not been clearly delineated. Taking the μ_{nmT} value at 20°C as the reference value, μ_{nm20} values have been reported ranging from 0,33 to 0,65 per day. These two limits will have a significant effect on the minimum sludge age: Two plants, having these respective μ_{nm20} values, will have R_{sm} values differing by hundred per cent. Clearly due to the link between the waste flow and μ_{nmT}, the latter's value should always be estimated experimentally for optimal design. In the absence of such a measurement, a low value for μ_{nmT} necessarily will need to be selected which, for a particular waste flow, should the actual μ_{nm} be higher, will result in a non-optimal design. An experimental procedure to determine μ_{nm20} is given in Appendix 3.

With regard to b_{n20}, this value empirically is taken as constant for all waste flows, at $b_{n20} = 0{,}04/d$. Its effect is small so that there is no need to enquire closely into all the factors affecting it.

With regard to K_{nT} little information on effects of inhibitory agents is available; very likely K_{nT} will increase with inhibition.

4.2 Temperature

The μ_{nm} and K_n constants are sensitive to temperature with a high temperature coefficient (see Eq. 5.15); semi-empirically b_{nT} is taken also to have temperature dependency at the same rate as that for heterotrophs (see Eq. 5.16).

$$\mu_{nmT} = \mu_{nm20}(1{,}123)^{(T-20)} \qquad (5.15a)$$

$$K_{nT} = K_{n20}(1{,}123)^{(T-20)} \qquad (5.15b)$$

$$b_{nT} = b_{n20}(1{,}029)^{(T-20)} \qquad (5.16)$$

The effect of temperature on μ_{nmT} is particularly dramatic, for every 6°C drop in temperature the μ_{nmT} value will halve which means that the minimum sludge age for nitrification will double. Design of plants for nitrification, therefore, should be based on the minimum expected process temperature. Note that the temperature sensitivity of K_{nT} does not affect the minimum sludge age, but it does affect the efficiency of nitrification – the higher the K_n the higher also the effluent ammonia at $R_s \gg R_{sm}$ (See Section 4.3).

4.3 pH and Alkalinity

The specific growth rate of the nitrifiers μ_n is extremely sensitive to the pH of the culture medium. It seems that the activities* of both the hydrogen and hydroxyl ions, (H^+) and (OH^-), act inhibitorily when their respective concentrations increase unduly. This happens when the pH increases above 8,5 [increasing (OH^-)] or decreases below 7 [increasing (H^+)]; optimal nitrification rates are expected for $7 < pH < 8{,}5$ with sharp declines outside this range.

From Eq. 5.4, μ_n is a function of both μ_{nm} and K_n. From an analysis of Eq. 5.13 it was shown earlier that the minimum sludge age is dominated by the magnitude of μ_{nmT}; it is only very weakly influenced by K_{nT}. At $R_s \gg R_{sm}$ however, the effluent ammonia concentration (N_a), although low, is relatively speaking, significantly higher for larger K_{nT} values: For example if K_{nT} increases by a factor of two, the effluent ammonia concentration will increase correspondingly by the same factor (see Eq 5.13). Consequently the value of K_{nT} is significant in so far as it controls the effluent ammonia concentration at $R_s \gg R_{sm}$.

With regard to experimental estimates of μ_{nmT} and K_{nT}, a great amount of research has gone into tracing the effect pH has on μ_{nmT}. These investigations generally have not separated out the effects of μ_{nmT} and K_{nT} so that most data are in effect lumped parameter estimates of μ_{nmT}. Almost no information is available on the effect of pH on K_{nT}. The effect of pH on K_{nT} can be surmised from data obtained in a laboratory scale completely mix-

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*Equivalent to concentration.

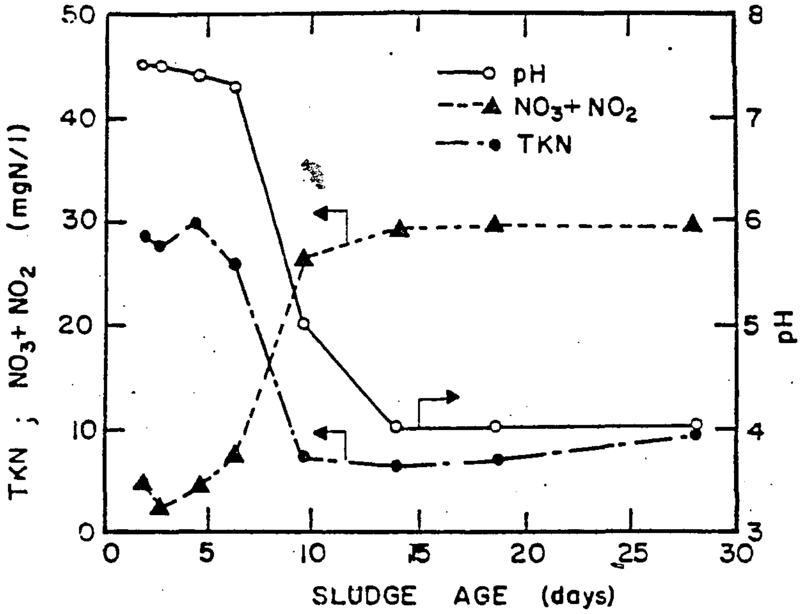

FIGURE 5.3: Relationship between TKN and nitrate concentrations and pH with sludge age in an activated plant treating weakly buffered influent sewage from the Strand-Somerset West area, Cape.

ed reactor aerobic activated sludge investigation on a waste flow from Somerset West in the Cape Province. The plant was operated at 20°C over a range of sludge ages from 2 to 30 days. The response of the plant in terms of pH, filtered effluent TKN and nitrate plus nitrite is shown in Figure 5.3. Concomitantly with nitrification the pH declined, stabilizing at about 4 when the sludge age exceeded 15 days. The TKN stabilized at 10 mgN/ℓ. Up to 10 days sludge age, the pH and nitrate concentration fluctuated in a see-saw fashion. The fact that stable nitrification was achieved only at 10 days sludge age when the pH was low, indicated that μ_{nmT} had decreased to a concomitantly low value. The fact that the effluent TKN ($\cong$ ammonia) stabilized at the high value of 10 mg/ℓ indicated that K_{nT} had increased. On controlling the pH at 7,0 nitrification commenced at about 3 days sludge age and the effluent TKN ($\cong$ ammonia) stabilized at about 4 mg/ℓ for $R_s > 3$ days. Evidently the pH and μ_{nmT} reacted interactively so that the minimum sludge age would fluctuate with the pH and achieve stability only at the long sludge age of 10 to 15 days. Evidently also the K_{nT} value is increased at low pH values. Due to the lack of information it is difficult to model quantitatively the effect of pH on μ_{nmT} and K_{nT}. The following analysis, based on various data sources and empirical assumptions on the behaviour, demonstrates the general expected trends even though the predicted values cannot be taken as quantitatively correct. Accepting that μ_{nm} remains constant for $7,2 < pH < 8,5$ and decreases as the pH decreases below 7,2 (Downing *et al.* 1964, Loveless and Painter, 1968) this behaviour can be modelled as follows:

For $7,2 < pH < 8,5$,

$$\mu_{nmpH} = \mu_{nm7,2} \quad (5.17a)$$

For $5 < pH < 7,2$,

$$\mu_{nmpH} = \mu_{nm7,2}\,\Phi_{ns}^{(pH-7,2)} \quad (5.17b)$$

where

Φ_{ns} = pH sensitivity coefficient
$\cong 2,35$.

In Figure 5.4 the predictions, using Eq. 5.17 are compared with experimental data on μ_{nm} collected by Malan and Gouws (1966). This comparison is illustrative only as it is not clear from the data of Malan and Gouws whether the effect of pH on K_n was taken into account when calculating μ_{nmpH}.

Changes in the value of K_n with pH are not at all documented in the literature. The best than can be done is to accept a dependency on pH and check the predictions against ammonia effluent concentrations reported at the different pH values. On this basis the following equations are suggested,

For $7,2 < pH < 8,5$

$$K_{npH} = K_{n7,2} \quad (5.18a)$$

For $5 < pH < 7,2$

$$K_{npH} = K_{n7,2}\,\Phi_{ns}^{(7,2-pH)} \quad (5.18b)$$

where

Φ_{ns} = pH sensitivity coefficient
$\cong 2,35$.

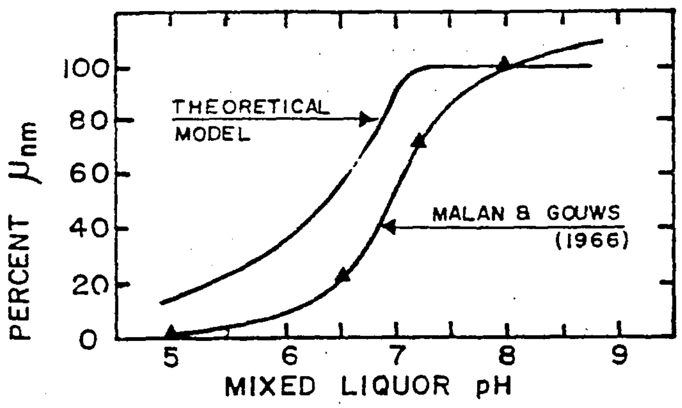

FIGURE 5.4: Relationship between maximum specific growth rate for *Nitrosomonas*, μ_{nm}, and mixed liquor pH. Experimental data after Malan and Gouws (1966).

The overall effect of pH or $5 < pH < 7,2$ and temperature on μ_n can be modelled by substituting Eqs. 5.15, 5.17 and 5.18 into Eq. 5.4 i.e.

μ_{nTpH} = μ_n at temperature T°C and pH equal to pH.

$$\mu_{nTpH} = \frac{\mu_{nm20}(2,35)^{(pH-7,2)}(1,123)^{(T-20)}.N_a}{(K_{n20}(2,35)^{(7,2-pH)}(1,123)^{(T-20)} + N_a)} \qquad (5.19)$$

where

μ_{nTpH} = specific growth rate at temperature T and pH = pH

μ_{nm} = maximum specific growth rate at temperature 20°C and pH = 7,2

K_{n20} = saturation coefficient at temperature 20°C and pH = 7,2.

Although Eq. 5.19 is qualitative in prediction only, it serves an important function in illustrating the adverse effects that temperature and pH can have on μ_n: A plot of μ_{nTpH} is shown in Figure 5.5 for selected values of pH and temperature accepting $\mu_{nm20(7,2)} = 0,33/d$ and $K_{n20(7,2)} = 1,00$ mgN/ℓ. Clearly low pH and low temperature can have significant effects on design.

In design the temperature effects have to be accepted as restrictions imposed by the environment. With regard to the effects of low pH, these may, or may not, occur depending on the alkalinity of the influent. Low influent Alkalinity is likely to give rise to low pH in the purely aerobic activated sludge process. However, it is possible to limit, or completely obviate, pH reduction by operating the plant as an anoxic-aerobic process. The reasons for this are as follows:

From the overall stoichiometric equations for nitrification (Eqs. 5.1a and 5.1b), nitrification releases hydrogen ions which in turn decreases Alkalinity* of the mixed liquor. For every 1 mg (NH_4^+-N) that is nitrified 7,14 mg Alkalinity (as $CaCO_3$) is destroyed. Based on equilibrium chemistry of the carbonate system (Loewenthal and Marais 1977), equations linking the pH with Alkalinity for any partial pressure of carbon dioxide have been developed by van Haandel and Marais (1981). These relationships are shown plotted in Figure 5.6. When the Alkalinity falls below about 40 mg/ℓ as $CaCO_3$ then, irrespective of the partial pressure of carbon dioxide, the pH becomes unstable and tends to decrease to low values. Generally, if nitrification causes the Alkalinity to drop below about 40 mg/ℓ (as $CaCO_3$) problems associated with low pH very likely will arise in the plant, such as poor nitrification efficiency, tendency for developing bulking sludges, corrosive effluent, etc.

For any particular wastewater the possible effect of pH can be readily assessed, as follows: Take the exam-

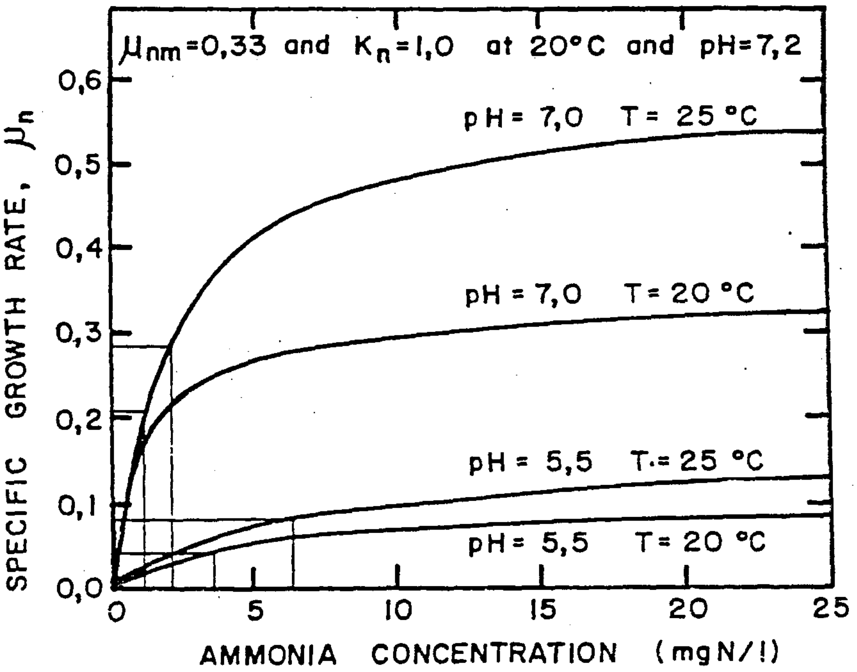

FIGURE 5.5: Diagrammatic representation of Eq. 5.19 – relationship describing the specific growth rate of *Nitrosomonas* for changes in temperature and mixed liquor pH.

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*Alkalinity signifies the Total Alkalinity obtained by titrating the sample to an equivalent carbonic acid solution at pH approximately 4,3.

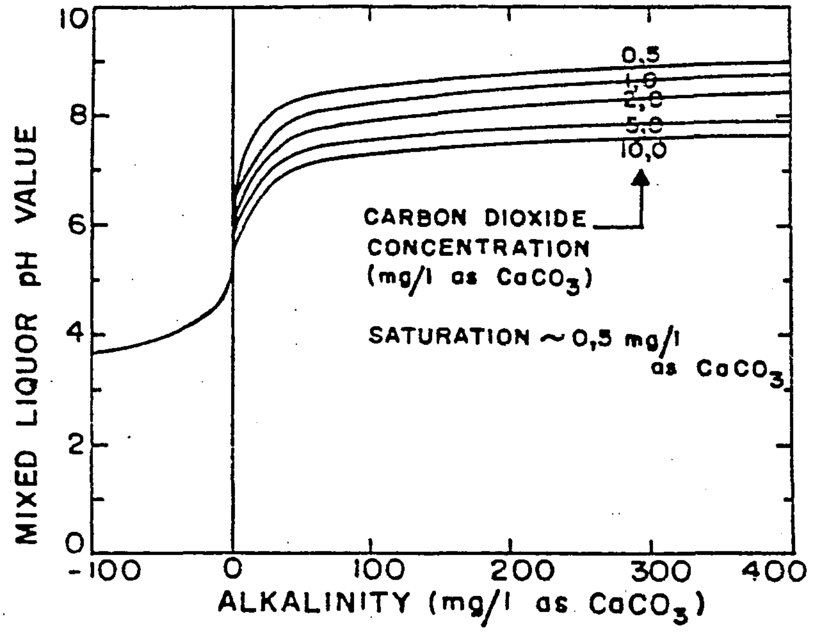

FIGURE 5.6: Mixed liquor pH versus alkalinity for different concentrations of carbon dioxide.

ple of an influent having an Alkalinity of 200 mg/ℓ as $CaCO_3$; expected production of nitrate is 24 mgN/ℓ. The expected Alkalinity in the effluent will be (200 − 7,14.24) = 29 mg/ℓ as $CaCO_3$. From Figure 5.6, such an effluent is likely to have a pH < 7,0.

Wastewaters having low Alkalinity are often encountered where the municipal supply is drawn from areas underlain with sandstone.

The abovementioned problems are to be expected in aerobic nitrifying plants treating municipal effluents of towns located in the coastal areas of South Africa, from Cape Town to, and including, Natal. The effluent from Somerset West mentioned earlier (see Figure 5.3) is a typical example. The only practical approach to treating such effluents is to create an anoxic zone in the process to denitrify some or all of the nitrate generated. In contrast to nitrification, denitrification takes up hydrogen ions which is equivalent to generating Alkalinity. By considering nitrate as electron acceptor, it can be shown that for every mg (NO_3-N) denitrified, there is an increase of 3,57 mg Alkalinity as $CaCO_3$. Hence incorporating denitrification in a nitrification process causes the net loss of Alkalinity to be reduced usually sufficiently to maintain the Alkalinity above 40 mg/ℓ and consequently the pH above 7.*

In the example above, where the Alkalinity in the process is expected to decline to 29 mg/ℓ as $CaCO_3$, if 25% of the nitrate could be denitrified the gain in Alkalinity would be (0,25.29.3,57) = 26 mg/ℓ as $CaCO_3$ and will result in an Alkalinity of (29 + 26) = 55 mg/ℓ as $CaCO_3$ in the process. In this event the pH should remain above 7. For low Alkalinity wastewaters it is imperative, therefore, that denitrification be built into nitrifying plants.

Incorporation of unaerated zones in the process influences the sludge age of the process at which nitrification takes place so that cognizance must be taken of the effect of an anoxic or unaerated zone in establishing the sludge age of a nitrifying-denitrifying plant.

4.4 Unaerated zones

The effect of unaerated zones on nitrification can be readily formulated if the following assumptions are made:

(1) Nitrifiers, being obligate aerobes, can grow only in the aerobic zones of a process.

(2) Endogenous mass loss of the nitrifiers occurs under both aerobic and unaerated conditions.

(3) The concentration of nitrifiers in the unaerated and aerated zones is essentially equal.

With these assumptions, Ekama, van Haandel and Marais (1979) showed that if a fraction f_{xt} of the total sludge mass is unaerated, i.e. $(1 - f_{xt})$ is aerated, the effluent ammonia is given by

$$N_a = \frac{K_{nT}\,(b_{nT} + 1/R_s)}{\mu_{nmT}\,(1 - f_{xt}) - (b_{nT} + 1/R_s)} \qquad (5.20)$$

Equation 5.20 is identical in structure to Eq. 5.13, if one views the effect of the unaerated mass (f_{xt}) as

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*Quantitative estimation of the denitrification that can be achieved in an anoxic-aerobic process is considered in Chapter 6.

reducing the value of μ_{nmT} to $\mu_{nmT}(1-f_{xt})$.

Following the same reasoning as that preceding Eq. 5.14 one can show that the minimum sludge age for nitrification R_{sm} in a process having an unaerated mass fraction, f_{xt}, is

$$R_{sm} = \frac{1}{\mu_{nmT}(1-f_{xt})-b_{nT}} \qquad (5.21)$$

Alternatively, if R_s is specified the minimum aerobic sludge mass fraction $(1-f_{xm})$ that must be present for nitrification is found by substituting R_s for R_{sm} and f_{xm} for f_{xt} in Eq. 5.21 and solving for $(1-f_{xm})$, i.e.

$$(1-f_{xm}) = (b_{nT}+1/R_s)/\mu_{nmT} \qquad (5.22)$$

or equivalently the maximum allowable unaerated sludge mass fraction is given by solving for f_{xm} in Eq. 5.22,

$$f_{xm} = 1-(b_{nT}+1/R_s)/\mu_{nmT} \qquad (5.23)^*$$

For a fixed sludge age, R_s, the design value for the minimum aerobic sludge mass fraction $(1-f_{xm})$ always should be significantly higher than that given by Eq. 5.22, because nitrification becomes unstable when the aerated sludge mass fraction decreases to near the minimum value as given by Eq. 5.22. Consequently to ensure *efficient* nitrification (>90%) the *minimum aerobic sludge mass fraction* $(1-f_{xm})$ must be increased by a factor of safety, S_f to give the *minimum design aerobic sludge mass fraction*; from Eq. 5.22,

$$(1-f_{xm}) = S_f(b_{nT}+1/R_s)/\mu_{nmT} \qquad (5.24a)$$

and the corresponding *maximum design unaerated sludge mass fraction*, from Eq. 5.23 is

$$f_{xm} = 1-S_f(b_{nT}+1/R_s)/\mu_{nmT} \qquad (5.24b)$$

With the aid of the temperature dependency equations for nitrification (Eqs. 5.15 and 5.16), a diagrammatic representation of Eq. 5.24 is given in Figure 5.7 for $S_f = 1{,}25$, $\mu_{nm20} = 0{,}65$ and 0,33/d at 14°C and 20°C. The two μ_{nm20} values are the approximate extremes of the range of μ_{nm20} values reported in the literature. The diagram shows that f_{xm} is very sensitive to the maximum specific growth rate of the nitrifiers (μ_{nmT}): unless a sufficiently large *aerobic* sludge mass fraction is provided (i.e. $1-f_{xm}$, see right hand of Figure 5.7), nitrification will not take place and consequently nitrogen removal by denitrification is not possible.

If one fixes the minimum design aerobic sludge mass fraction, the sludge age can be determined by solving for $(1-f_{xm})$ in Eq. 5.24a. Knowing R_s the value of the effluent ammonia concentration (N_a) can be determined from Eq. 5.20. By doing a number of solutions of this kind, at 14°C, it will become apparent that if S_f is selected at 1,25 or greater the effluent quality will have stability and range from 2 to 4 mg $(NH_3\text{-}N)/\ell$ depending on the other constants in the formulations. Accepting the sludge age as determined at 14°C, then at 20°C the

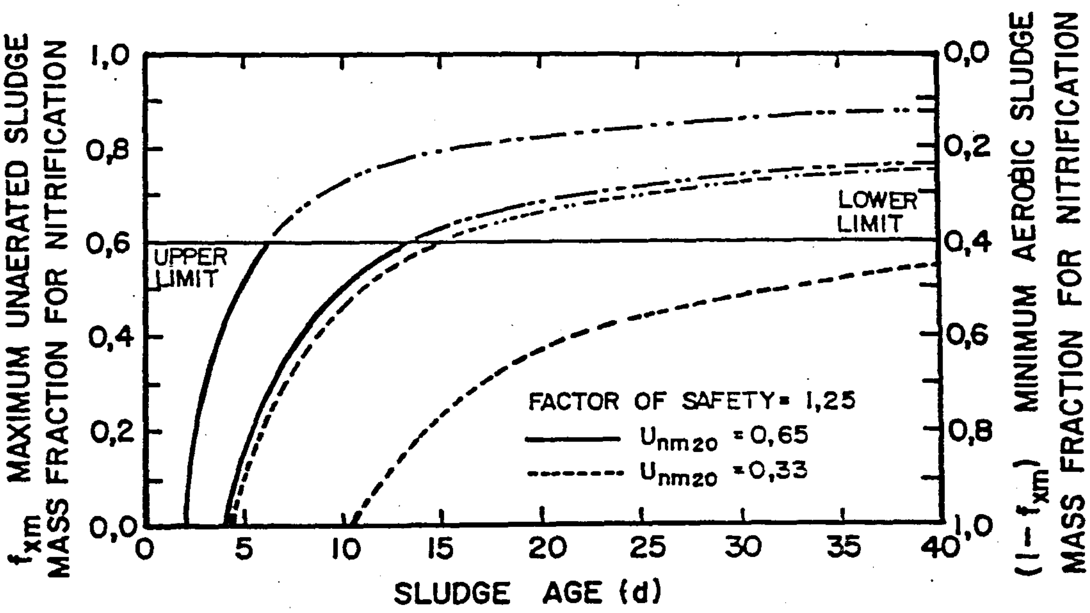

FIGURE 5.7: Maximum allowable unaerated sludge mass fraction and minimum aerated sludge mass fraction required to sustain nitrification versus sludge age for different maximum specific growth rates of the nitrifiers at 20°C and 14°C and $S_f = 1{,}25$.

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*The usual approach for determining f_{xm} is based on the concept of "aerobic sludge age" i.e. equating the minimum sludge age for nitrification R_{sm} to the aerobic sludge age of the process R_{sa}. This leads to $f_{xm} = 1-1/\{R_s(\mu_{nmT}-b_{nT})\}$. This equation overestimates the maximum unaerated sludge mass f_{xm} at a particular sludge age because the approach ignores the endogenous mass loss of the nitrifiers in the unaerated zones. If the endogenous mass loss can be ignored in nitrification kinetics i.e. $b_n = 0{,}0$, then the two approaches are equivalent.

value of S_f will be found to have increased significantly and the effluent ammonia to have decreased to 2 mg/ℓ or less. Consequently, for design the lower expected temperature should be selected to determine the sludge age and the aerobic mass fraction. If this is done, using $S_f = 1,25$, one can accept an effluent ammonia value of 2-4 mgN/ℓ at the lowest temperature and 1-3 mgN/ℓ at 20°C, that is, Eq. 5.20 is bypassed by the approximation above, provided R_s at least equals 1,25 R_{sm}. This approximation is sufficient for design and will be the approach followed in the design example.

4.4.1 Maximum allowable unaerated mass fraction

The equations above allowed the maximum *design* unaerated sludge mass fraction to be determined for any selected sludge age (Eq. 5.24b). If one accepts $S_f = 1,25$ then f_{xm} or $(1 - f_{xm})$ can be plotted versus R_s for any selected μ_{nmT}. This is illustrated in Figure 5.7 where the design (f_{xm}) and $(1 - f_{xm})$ values are plotted versus R_s for 20°C and 14°C for two selected μ_{nm20} reference values of 0,65 and 0,33 (b_{n20} was fixed at 0,04/d). Evidently from Figure 5.7, for $\mu_{nm20} = 0,65$ the unaerated mass fraction at 14°C can be *approximately* 0,8 at sludge age of 30 days. Such a high unaerated mass fraction is the indicated one theoretically at $R_s = 10$ days or longer at 20°C. One may now ask: Is this mass fraction acceptable for design or are other constraints present that may limit the fraction to lower values? To answer this, two aspects appear to be pertinent:

(1) Experience with laboratory scale nitrification denitrification plants tends to support the opinion that at high unaerated mass fractions, the process is prone to bulking particularly at low temperatures. Processes at unaerated mass fractions of 0,5 or less show much less tendency in this regard.

(2) An upper limit to the unaerated mass fraction is evident from experimental work on the Ludzack-Ettinger process, and theoretical predictions using the general activated sludge model. Experimentally at 20°C with $R_s = 20$ days, if $f_{xm} > 0,70$, the mass of sludge generated is found to increase sharply. This observation is indicated also by theoretical simulation. Theoretically, this happens for $f_{xm} > 0,60$ at $T = 14°C$ and $R_s = 20$ days. The reason is that for such high f_{xm}, the exposure of the sludge to aerobic conditions becomes insufficient to utilize the adsorbed and enmeshed particulate material. This leads to a decrease in active mass and oxygen demand and a build-up of enmeshed material, which causes a progressive decline in process activity because of the decline in active mass. When this happens, the process still *functions* in that the COD is removed from the wastewater, but the degradation of the COD is reduced; the process begins to assume the role of a contact reactor of a contact-stabilization process, i.e. a bio-flocculator with minimal degradation. This critical state occurs at lower f_{xm} as the temperature is decreased and the sludge age is reduced (Arkley and Marais, 1982).

From the discussion above it would appear that the unaerated mass fraction should not be allowed to increase above a certain upper limit of about 50%, but certainly not above 60%, as indicated in Figure 5.7.

In design of N and P removal plants the unaerated sludge mass fraction f_{xm} usually needs to be high i.e. ≥40%. If the μ_{nm20} value is low i.e. <0,40/d (which will be the usual case in design where insufficient information on the μ_{nm20} is available) the necessary high f_{xm} magnitudes will be obtained only at long sludge ages (see Figure 5.7). For example, if $\mu_{nm20} = 0,35$/d, then with $S_f = 1,3$ at $T_{min} = 14°C$, an $f_{xm} = 0,45$ (Eq. 5.24b) gives a sludge age of 25 days and for $f_{xm} = 0,55$ a sludge age of 37 days. Long sludge ages require large process volumes – increasing R_s from 25 to 37 days increases the process volume by 40% (see Figure 4.3) whereas f_{xm} increased only 22%. Also, for the same P content in the sludge mass, the P removal is reduced as the sludge age increases because the mass of sludge wasted daily decreases as the sludge age increases (see Figure 4.5). Consequently, for low μ_{nm20} values, the increase in N and P removal that can be obtained by increasing the unaerated sludge mass fraction above 0,50 to 0,60 might not be economical due to the large process volumes this will require, and might even be counter productive insofar as it affects P removal. A sludge age of 35 days probably is near the limit of eonomic practicality which, for low $\mu_{nm14} = 0,16$ values will limit the unaerated mass fraction to about 0,5. At higher μ_{nm14} values, the sludge ages allowing 50 per cent unaerated mass fractions decrease significantly again indicating the advantages of determining experimentally the value of μ_{nm20} to check whether a higher value is acceptable.

4.5 Dissolved oxygen concentration

High dissolved oxygen concentrations, up to 33 mg/ℓ, do not appear to affect nitrification rates significantly. Low oxygen concentrations however do reduce the nitrification rate. Stenstrom and Poduska (1980) have suggested formulating this effect as follows:

$$\mu_{no} = \mu_{nmo} \frac{O}{K_o + O} \qquad (5.25)$$

where

O = oxygen concentration in bulk liquid (mgO/ℓ)
K_o = half saturation constant (mgO/ℓ)
μ_{nmo} = maximum specific growth rate (/d)
μ_{no} = specific growth rate at O mg/ℓ oxygen(/d).

The value attributed to K_o has ranged from 0,3 to 2 mgO/ℓ, i.e. at DO values below K_o the growth rate will decline to less than half the rate where oxygen is present in adequate concentrations.

The wide range of K_o probably has arisen because the concentration of DO in the bulk liquid is not necessarily the same as inside the biological floc where the oxygen consumption takes place. Consequently the value will depend on the size of floc, mixing intensity and oxygen diffusion rate into the floc. Furthermore in a full-scale reactor the DO will vary over the reactor volume due to the discrete points of oxygen input (with mechanical aeration) and the impossibility of achieving instantaneous and complete mixing. For these reasons it is not really possible to establish a generally applicable minimum oxygen value – each reactor will have a value specific to the conditions prevailing in it. In nitrifying reactors with bubble aeration a popular DO lower limit, to ensure unimpeded nitrification, is 2 mgO/ℓ at the surface of the liquid.

Under cyclic flow and load conditions the difficulties of ensuring an oxygen supply matching the oxygen demand and a lower limit for the DO have been discussed previously (see Chapter 4, Section 11.3). Equalization of the load probably presents the best practical means to facilitate control of the oxygen level in the reactor. In the absence of equalization, mitigation of the adverse effects of low oxygen concentration during peak oxygen demand periods is by having sludge ages significantly longer than the minimum necessary for nitrification.

4.6 Cyclic flow and load

It is well attested experimentally that under cyclic flow and load conditions the nitrification efficiency of an activated sludge process decreases compared to that under steady conditions. This behaviour is also apparent in simulation studies using the general activated sludge model. From the simulation studies, during the high flow and/or load period, even though the nitrifiers are operating at their maximum rate, it is not possible to oxidize all the ammonia available, and an increased ammonia concentration is discharged in the effluent. This in turn reduces the mass of nitrifiers formed in the process – equivalently the effect of cyclicity in flow and load is to reduce the aerobic sludge age. The adverse effect of the cyclicity becomes more marked as the fractional amplitude of the flow and load cycles increase, and, is ameliorated as the aerobic sludge age increases. Simulation studies of the cyclic effect have indicated that an approximate relationship can be established between the effluent ammonia concentration under steady state and that under cyclic conditions (see van Haandel and Marais 1981). However if the factor of safety on nitrification, S_f, is 1,25 or greater the cyclic effect is relatively small and for design usually can be neglected.

At the risk of being tedious, it should again be pointed out that if the value of μ_{nm} is selected higher than the actual value, even with a sludge age of 1,25 to 1,35 times the theoretical minimum, the plant is likely to give rise to a fluctuating nitrate effluent with reduced mean efficiency in nitrification. Hence a conservative estimate of μ_{nm} is essential for a safe design.

Under cyclic flow and load conditions, it is necessary to know the value of both the specific yield coefficient of the *Nitrosomonas*, Y_n, the maximum specific growth rate constant, μ_{nmT}, and the half saturation constant, K_n; in contrast, only μ_{nmT} and K_n are necessary in steady state calculations. The values of these constants have been determined by a variety of methods but the constants have a conjoined effect that may cause that in a particular method of evaluation different sets of values may be obtained. For practical purposes it is not possible to obtain a complete set of values so that one approach is to accept the value of one constant and in any particular experimental evaluation determine the other constants from the experimental data using this value. The estimates for Y_n and K_n accepted in this monograph are listed in Table 5.1; experimental μ_{nm} determination, therefore, is relative to the value of Y_n and K_n listed.

TABLE 5.1 KINETIC CONSTANTS FOR *NITROSOMONAS* ACCEPTED IN THIS MONOGRAPH

Constant	Symbol	Value at 20°C	Temperature Coefficient	Eq No.
Specific yield coefficient	Y_n	0,10	1,000	–
Endogenous respiration rate	b_n	0,04	1,029	5.16
Half saturation coefficient	K_n*	1,00	1,123	5.15b

*Note that K_n increases with increase in temperature.

5. DESIGN CONSIDERATIONS

5.1 Fate of the influent TKN

The various fractions of the influent TKN have been described in Chapter 2. Of these the following are of importance to our discussion: (1) free and saline ammonia concentration (N_{ai}); (2) biodegradable organic nitrogen (N_{oi}); and (3) the soluble unbiodegradable organic nitrogen (N_{ui}).

The biodegradable organic nitrogen is decomposed in the process by heterotrophic action to free and saline ammonia. Furthermore, on death of the organisms, by endogenous mass loss, the nitrogen released is in the organic bound form and the biodegradable fraction also is broken down to free and saline ammonia. These two sources of free and saline ammonia together with that in the influent are utilized by the organism mass first for the formation of new organisms and the balance for generation of nitrate, if the conditions are favourable.

The 'exact' formulations describing the interactions between the free and saline ammonia and the organic bound nitrogen are complex and can be developed effectively (at present) only in terms of the growth-death-regeneration model of Dold, Ekama and Marais (1981).* However under steady state conditions the kinetics simplify and the formulations reduce to describing net effects. These can be developed readily in terms of the synthesis – endogenous respiration approach already used to develop the steady state equations for carbonaceous material degradation, in Chapter 4. For design there is even a further simplification possible: In nitrification systems if, for example, the constraints are put on the system that the minimum design sludge age must be at least 1,25 to 1,35 times the theoretical

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*A more elaborate discussion is available by van Haandel and Marais (1981).

minimum to initiate nitrification, and, that the pH remains above 7, some of the steady state kinetic equations predict effluent concentrations that fall within a narrow band of values, sufficiently narrow that one can accept the effluent parameter virtually as constant.* Below are set down some of these steady state equations of importance in developing design procedures. Simple approximations adequate for design are indicated and some general response graphs are derived.

In the design of those aspects of a process effecting nitrification, basically the following information is needed: (1) effluent TKN concentration; and (2) effluent nitrate concentration.

5.2 Effluent TKN

The effluent TKN will consist of free and saline ammonia (N_a), biodegradable soluble organic N (N_o), unbiodegradable soluble organic N (N_u) and TKN in the effluent volatile solids. Equations for the fractions are set out below:

(1) *Free and saline ammonia (N_a):*

N_a is given by Eq. 5.20, i.e.

$$N_a = \frac{K_{nT}\,(b_{nT}+1/R_s)}{\mu_{nmT}\,(1-f_{xt})-(b_{nT}+1/R_s)} \qquad (5.20)$$

Note that N_a is independent of the influent ammonia concentration. Note also that the equation applies only if the sludge age $R_s > R_{sm}$, however this condition always should be satisfied in design.

(2) *Biodegradable organic nitrogen concentration (N_o)*

The biodegradable organic nitrogen is broken down to free and saline ammonia by heterotrophic organisms. Although a complex reaction, involving a feedback of this nitrogen form from death of the organisms, the prediction simulated by the general model indicates that the following equation, in terms of the influent biodegradable organic nitrogen concentration, is adequate

$$N_o = \frac{N_{oi}}{1+K_r X_a R_{hn}} \qquad (5.26)$$

where

N_{oi} = influent biodegradable organic nitrogen mg(TKN-N)/ℓ. Its value is given by Eq. 2.14.

K_r = kinetic constant for degradation of N_o. Its value and temperature dependency is given in Table 5.3.

X_a = active mass concentration in aerobic reactor.

R_{hn} = nominal hydraulic retention time in process.

(3) *Soluble unbiodegradable organic nitrogen (N_u)*

$$N_u = N_{ui} \qquad (5.27)$$

where

N_{ui} = soluble unbiodegradable organic nitrogen, mg(TKN-N)/ℓ. Its value is given by Eq. 2.12.

All these effluent fractions are soluble; consequently the total soluble TKN in the effluent (N_t) is given by the sum of these, i.e.

$$N_t = N_a + N_o + N_u \quad \text{(filtered } N_t\text{)} \qquad (5.28a)$$

The experimental value of N_t will be given by the TKN of the *filtered* effluent sample. If the sample is not filtered the theoretical TKN of the effluent will be greater by the concentration of TKN in the volatile solids, i.e.

$$N_{tv} = N_a + N_o + N_u + f_n . X_{vf} \quad \text{(unfiltered } N_t\text{)} \qquad (5.28b)$$

where

X_{vf} = effluent volatile solid concentration (mgVSS/ℓ)

f_n = fraction of TKN in VSS ≅ 0,1 mg(TKN-N)/mgVSS.

5.3 Nitrification capacity

The concentration of nitrate generated in the system (N_c) is given by the TKN in the influent (N_{ti}) less the soluble TKN in effluent (N_t) less the concentration of total influent TKN incorporated in the net sludge generated each day (N_s) i.e.

$$N_c = N_{ti} - N_t - N_s^{**} \qquad (5.29)$$

The value of N_s is determined from the mass of nitrogen incorporated in the volatile solids per day. It is given by f_n times the mass of volatile solids wasted per day ($M(X_v)/R_s$), divided by the mean daily flow, i.e.

$$N_s = \frac{f_n . M(X_v)}{R_s . Q} \quad \text{(see Chapter 4, Section 9).}$$

where

$M(X_v)$ is given by Eq. 4.13.

Note that $M(X_v)$ does not include the volatile mass of nitrifiers as the masses of these, we have stated previously, are negligible.

In Eq. 5.29, N_c defines the *nitrification capacity* of the process. A formal definition of nitrification capacity (N_c) is the mass of nitrate produced per unit volume of flow, i.e. mg(NO_3-N)/ℓ of flow.

Eq. 5.29 shows that the nitrification capacity (N_c) is the difference between the influent TKN concentration (N_{ti}) and the sum of the effluent TKN concentration (N_t) and the nitrogen required for sludge production (N_s). The effluent TKN concentration (N_t) depends on the efficiency of nitrification. In calculations, with maximum unaerated sludge mass fraction (f_{xm}), (for a selected sludge age), if the factor of safety (S_f) is selected greater than 1,25 to 1,35 at the *lowest expected temperature* (T_{min}), the efficiency of nitrification always will be high (>90%) and N_t generally will be less than 3 to 4 mgN/ℓ. Also, with $S_f \geq 1{,}25$ at T_{min}, N_t will be vir-

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*The same approach also applied in the design equations for carbonaceous material degradation that, provided $R_s > 3$ days usually one can accept the filtered *biodegradable* COD in the influent as zero (see Chapter 4, Section 3.4).

**This equation presumes that no nitrate is present in the influent and this is true, virtually always, for municipal flows in South Africa.

tually independent of both the process configuration and the subdivision of the sludge mass into aerated and unaerated fractions. *Consequently, for design N_t usually need not be calculated explicitly from Eqs. 5.20, 5.26 to 5.28, but can be taken to be approximately 3 to 4 mgN/ℓ* provided that there is reasonable assurance that the actual μ_{nm20} value will not be less than the value accepted for design and that there is sufficient aeration capacity so that nitrification is not inhibited by an insufficient oxygen supply. Accepting the calculated f_{xm} and selected sludge age (R_s) at the lower temperature, then at higher temperatures the nitrification efficiency and the factor of safety (S_f) both will increase so that at summer temperatures (T_{max}), N_t will be lower, approximately 2 mgN/ℓ.

Dividing Eq. 5.29 by the total influent COD concentration (S_{ti}) yields the nitrification capacity per mgCOD applied to the biological process, N_c/S_{ti},

$$N_c/S_{ti} = N_{ti}/S_{ti} - N_s/S_{ti} - N_t/S_{ti} \qquad (5.30)$$

where

N_c/S_{ti} = nitrification capacity per mgCOD applied to the process (mgN/mgCOD)

N_{ti}/S_{ti} = TKN/COD ratio of the wastewater

N_s/S_{ti} = nitrogen required for sludge production per mgCOD applied (see Eq. 4.23).

In Eq. 5.30 the nitrification capacity/COD ratio (N_c/S_{ti}) of a process can be estimated approximately by evaluating each of the terms in the RHS of the equation as follows:

N_t/S_{ti}: Provided the constraint for efficient nitrification is satisfied at the lowest temperature (T_{min}) the TKN in the effluent (N_t) can be assumed at T_{min} to be about 4 mgN/ℓ i.e. N_t/S_{ti} will range from 0,005 to 0,01 for S_{ti} ranging from 800 to 400 respectively. At T_{max}, $N_t \cong 2$ mgN/ℓ.

N_s/S_{ti}: This ratio gives the TKN concentration in the influent abstracted for sludge production. Its value will depend on whether settled or raw wastewater is being treated, and the sludge age. The ratio, N_s/S_{ti}, for this purpose can be read from Figure 4.5, or calculated from Eq. 4.23.

N_{ti}/S_{ti}: This value is obtained by measurement of the TKN and COD of the influent and can range from 0,07 for unsettled municipal waste flows to 0,12 for settled flows.

A graphical exposition of the relative importance of these three ratios on N_c/S_{ti} is obtained by plotting N_c/S_{ti} versus sludge age for selected influent TKN/COD (N_{ti}/S_{ti}) ratios ranging from 0,07 to 0,09 for raw wastewater and 0,09 to 0,11 for settled wastewater. The N_c/S_{ti} values were found from Figure 4.5 (or Eq. 4.23) and the N_t/S_{ti} values selected constant at 0,007 (this is for S_{ti} of approximately 600 mgCOD/ℓ). The plots for 14°C and 20°C are shown in Figures 5.8a and 5.8b respectively. Also shown plotted on the diagrams are the minimum sludge ages for unaerated mass fractions of 0,0, 0,2, 0,4 and 0,5, at a selected μ_{nm20} value of

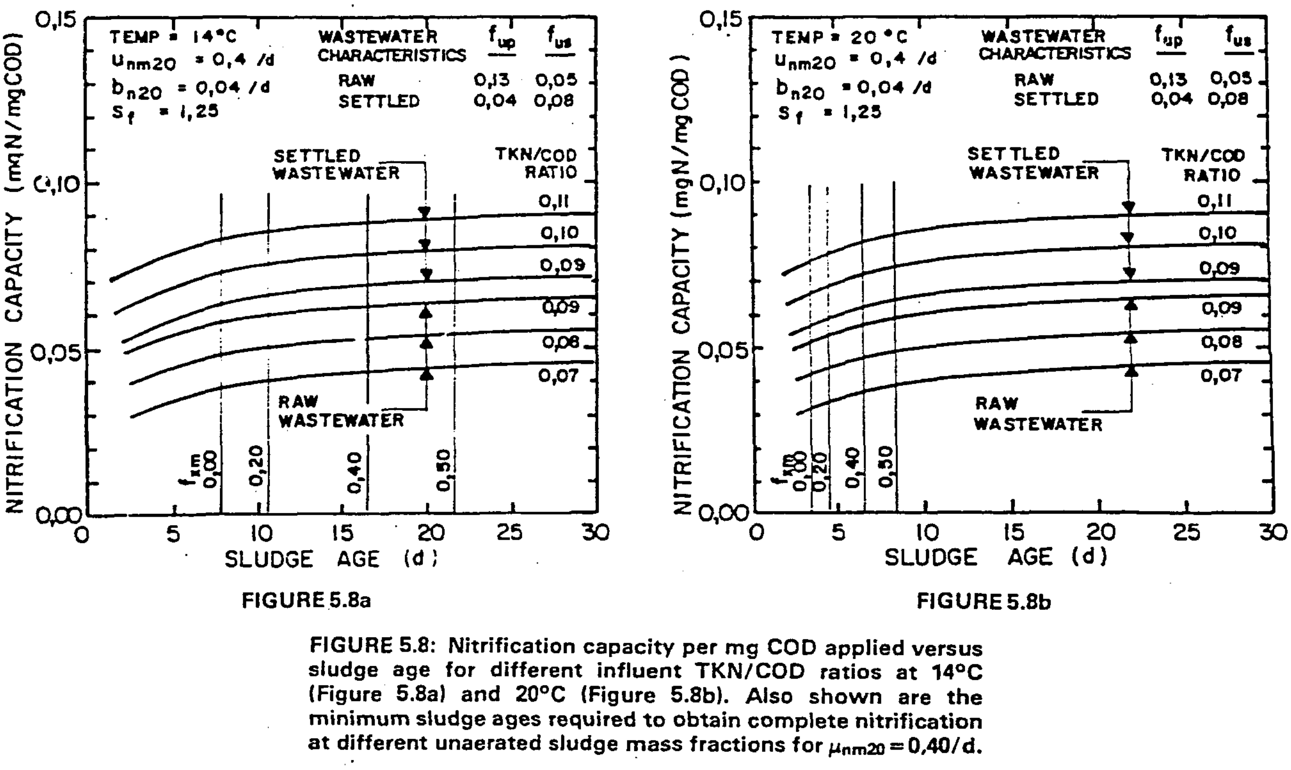

FIGURE 5.8a

FIGURE 5.8b

FIGURE 5.8: Nitrification capacity per mg COD applied versus sludge age for different influent TKN/COD ratios at 14°C (Figure 5.8a) and 20°C (Figure 5.8b). Also shown are the minimum sludge ages required to obtain complete nitrification at different unaerated sludge mass fractions for $\mu_{nm20} = 0{,}40$/d.

0,4. For any particular unaerated mass fraction the plotted values of N_c/S_{ti} are valid only at sludge ages greater than the corresponding minimum sludge age. These plots allow one to assess the effects of the three ratios and temperature versus sludge age.

Considering the effects of temperature, to obtain complete nitrification at 14°C (for a selected f_{xm}), the sludge age required is more than double that at 20°C. The corresponding nitrification capacities per influent COD at 14°C show a marginal reduction to those at 20°C (because the sludge production at 14°C is slightly higher than at 20°C).

Considering the effect of sludge age for a selected N_{ti}/S_{ti} value, the N_c/S_{ti} increases as the sludge age increases because the nitrogen needed for sludge production decreases with sludge age and thus becomes available for nitrification. However, the increase is marginal for $R_s > 10$ days.

Considering the influent TKN/COD ratio (N_{ti}/S_{ti}) evidently for both settled and unsettled wastewater, at any selected sludge age, N_c/S_{ti} is very sensitive to the N_{ti}/S_{ti} value; an increase of 0,01 in N_{ti}/S_{ti} causes an increase of about 0,01 in N_c/S_{ti} (0,01 mgN/mgCOD). For the same N_{ti}/S_{ti} ratio for raw or settled wastewater (compare TKN/COD ratio of 0,09 in Figures 5.8a and b), N_c/S_{ti} for unsettled is less than for settled wastewater because more volatile solids are produced per unit COD load from unsettled than from settled wastewater; however, leaving aside this point of difference, the increase in nitrate produced per influent COD will, when the influent TKN/COD ratio increases, increase the difficulties, or, make it impossible, to obtain complete denitrification. This will become clear when denitrification is considered in Chapter 6. As primary settling increases the influent TKN/COD ratio difficulties in obtaining adequate denitrification are always greater with settled than unsettled wastewater; however, because primary settling removes COD, then for the same flow, the mass of volatile solids and the oxygen demand in the biological process will be lower for settled than for raw wastewater thereby giving rise to savings in plant volume and oxygenation costs.

6. DESIGN EXAMPLE

Design of a nitrification process without denitrification will be considered. When denitrification is required constraints additional to those imposed by nitrification must be applied and these cannot be incorporated in the design until the theory of denitrification is dealt with.

For the purpose of constructive comparison the nitrifying activated sludge plant will be designed for the same waste flows and characteristics as that accepted for the design of a process for carbonaceous material removal (see Chapter 4, Section 12). The wastewater characteristics, for raw and settled wastewaters are listed in Table 4.3 and the additional characteristics needed for nitrification are listed in Table 5.2.

6.1 Effect of nitrification on mixed liquor pH

An important initial consideration is the possible effect of mixed liquor pH on the μ_{nm20} value. In Section 4.3 above it was stated that nitrification consumes alkalinity – 7,14 mg/ℓ as $CaCO_3$ for every 1 mgN/ℓ ammonia converted to nitrate – and if there is insufficient alkalinity in the influent, the mixed liquor pH is depressed below 7,0 causing a reduction in the μ_{nm20} value (see Eq 5.17).

The influent TKN/COD ratio of the raw wastewater is 0,08 mgN/mgCOD (see Tables 4.3 and 5.2). With a relatively low μ_{nm20} value of 0,36/d, the sludge age needs to be 10 days or longer to ensure nitrification

TABLE 5.2 RAW AND SETTLED WASTEWATER CHARACTERISTICS FOR NITRIFICATION (SEE ALSO TABLES 4.3 AND 6.1)

Parameter	Symbol	Value Raw	 Settled	Units
Influent TKN concentration	N_{ti}	48	41	mgN/ℓ
Influent TKN/COD ratio	—	0,080	0,114	mgN/mgCOD
Ammonia fraction of TKN	f_{na}	0,75*	0,83*	mgN/mgN
Unbiodegradable soluble organic N fraction	f_{nu}	0,03*	0,04*	mgN/mgN
Nitrogen fraction of volatile solids in influent	f_n	0,10	0,10	mgN/mgVSS
pH of wastewater	—	7,5	7,5	—
Total alkalinity	alk	200	200	mg/ℓ as $CaCO_3$
Maximum specific growth rate of nitrifiers	μ_{nm20}	0,36	0,36	/d.
Influent flow	Q	13,3	13,3	Mℓ/d

*See Chapter 3, Section 8. Note that settled wastewater characteristics must be selected so that they are consistent with primary sedimentation tank behaviour and the raw wastewater characteristics.

($S_f = 1{,}3$) at a minimum temperature of 14°C in a purely aerobic process ($f_{xm} = 0{,}0$) (see Eq. 5.24). At this sludge age the nitrification capacity is about 0,05 mgN/mgCOD for a TKN/COD ratio of 0,08 mgN/mgCOD, (see Figure 5.8a). Hence the nitrate concentration produced (per ℓ influent) should be about $= 0{,}05.600 = 30$ mgN/ℓ. This will cause an alkalinity reduction of $7{,}14.30 = 214$ mg/ℓ as $CaCO_3$. Because the influent alkalinity is only 200 mg/ℓ as $CaCO_3$, the mixed liquor alkalinity will drop below 40 mg/ℓ as $CaCO_3$ causing the mixed liquor pH value to drop below 7 (see Figure 5.6). The low mixed liquor pH value will cause unstable and incomplete nitrification (see Section 4.3 above) and produce an aggressive water (which over a few years can cause considerable damage to the concrete surfaces of the treatment works) and can be the cause of bulking sludges and high solids content in the settling tank effluent due to denitrification action in the secondary settling tank.

The simple approximate calculation above, therefore, can make the designer aware, at an early stage of possible adverse consequences to be expected from a proposed design. In the instance above the designer should give consideration to a nitrification-denitrification process, to recover some of the alkalinity loss due to nitrification, thereby to maintain the pH, reduce the possibility of bulking and eliminate (usually) rising sludge problems in the settling tank. (See Chapter 6, Section 1.4). In general a high TKN/COD ratio coupled with low alkalinity in the influent are reliable indicators warning of problems to be expected in aerobic nitrifying processes.

6.2 Minimum sludge age for nitrification

For the purposes of demonstrating nitrification under purely aerobic conditions, it will be accepted that the influent alkalinity is sufficiently high to maintain an effluent-Alkalinity above 40 mg/ℓ as $CaCO_3$. Hence no adjustment to μ_{nm20} and K_{n20} for pH needs to be made as the pH is assumed to remain above 7,2. The adjustment of the kinetic constants of nitrification for temperature is given in Table 5.3.

For a completely aerobic process i.e. $f_{xm} = 0$, with $\mu_{nm20} = 0{,}36$/d and $S_f = 1{,}3$, the minimum sludge age for nitrification (R_{sm}) is found from Eq. 5.24 i.e.

$$R_{sm} = S_f/(\mu_{nmT} - b_{nT})$$

$$= 3{,}2 \text{ days at } 22°\text{C } (2{,}5 \text{ days with } S_f = 0{,}0)$$

$$= 8{,}9 \text{ days at } 14°\text{C } (7{,}5 \text{ days with } S_f = 0{,}0)$$

Clearly, to ensure nitrification throughout the year for the relatively low μ_{nm20} value of 0,36/d, the sludge age of a purely aerobic process should be about 10 to 12 days.

6.3 Raw wastewater

The influent TKN concentration of the raw wastewater is 48 mgN/ℓ (see Table 5.2 for wastewater characteristics). Accepting an ammonia fraction of the influent TKN (f_{na}) of 0,75* and an unbiodegradable soluble organic nitrogen fraction (f_{nu}) of 0,03* for the raw sewage (from Table 5.2), gives the influent ammonia concentration (N_{ai}) (from Eq. 2.11) as

$$N_{ai} = 0{,}75.48 = 36 \text{ mgN}/\ell$$

and the unbiodegradable soluble organic nitrogen concentration (N_{ui}), (from Eq. 2.12) as

$$N_{ui} = 0{,}03.48 = 1{,}50 \text{ mgN}/\ell.$$

Accepting the nitrogen content of the volatile solids in the influent (f_n) as 0,10 mgN/mgVSS (see Table 5.2), then (from Eq. 2.13) the nitrogen concentration associated with the unbiodegradable particulate COD (N_{pi}) is

$$N_{pi} = 0{,}10\,(0{,}13.600)/1{,}48 = 5{,}3 \text{ mgN}/\ell.$$

Hence (from Eq. 2.14), the biodegradable organic nitrogen concentration in the influent which can be converted to ammonia (N_{oi}) is

$$N_{oi} = 48\,(1 - 0{,}75 - 0{,}03) - 5{,}3 = 5{,}3 \text{ mgN}/\ell.$$

TABLE 5.3 ADJUSTMENT OF KINETIC CONSTANTS OF NITRIFICATION FOR TEMPERATURE

Constant	Symbol	20°C	θ	14°C	22°C	Eq. No.
Max. specific growth rate	μ_{nm20}	0,36	1,123	0,180	0,454	5.15a
Saturation coefficient	K_{n20}	1,0	1,123	0,50	1,26	5.15b
Endogenous respiration rate	b_{n20}	0,04	1,029	0,034	0,042	5.16
Organic N conversion rate	K_r	0,015	1,029	0,013	0,016	5.31

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*The actual choice of these values makes relatively little difference to the final results of the nitrification calculations: in the case of the former (f_{na}), most of the organic nitrogen is converted to ammonia in the process and in the case of the latter (f_{nu}), this value merely increases the effluent TKN concentration by a fixed value.

6.4 Settled wastewater

Following the above procedure for settled wastewater, i.e. $f_{na}=0{,}83$, $f_{nu}=0{,}04$ (see Table 5.2) yields the following:

$N_{ti} = 41{,}0 \qquad = 41{,}0 \text{ mgN}/\ell$

$N_{ai} = 0{,}83.41{,}0 \qquad = 34{,}0 \text{ mgN}/\ell$

$N_{ui} = 0{,}04.41{,}6 \qquad = 1{,}50 \text{ mgN}/\ell$

$N_{pi} = 0{,}10\,(0{,}04.360)/1{,}48 = 1{,}0 \text{ mgN}/\ell$

$N_{oi} = 41{,}0-34{,}0-1{,}5-1{,}0 = 4{,}5 \text{ mgN}/\ell$

6.5 Nitrification process behaviour

For any sludge age longer than about 2 days the effluent biodegradable organic nitrogen concentration (N_{oe}) is given (from Eq. 5.26)

$$N_{oe}{}^{*} = N_{oi}/(1 + K_{rT}X_aR_{hn}) \quad (\text{mgN}/\ell)$$

where from Eq. 4.9a, $R_{hn}X_a = M(X_a)/Q$. Hence

$$N_{oe} = N_{oi}/(1 + K_{rT}M(X_a)/Q) \quad (\text{mgN}/\ell) \qquad (5.31)$$

From Eq 5.27, the unbiodegradable organic nitrogen in the effluent is

$$N_{ue} = N_{ui} = 1{,}50 \text{ mgN}/\ell \text{ for raw and settled wastewater} \qquad (5.32)$$

The nitrogen required for sludge production (N_s) is given by Eq 4.23 i.e.

$$N_s = f_n\{\frac{(1-f_{us}-f_{up})Y_h}{(1+b_{hT}R_s)}(1+fb_{hT}R_s)+\frac{f_{up}}{f_{cv}}\}\,S_{ti} \qquad (5.33)$$

The influent TKN concentration available for nitrification (N_{ti}') is the difference between the total TKN concentration in the influent (N_{ti}) and that required for sludge production (N_s) (from Eq. 5.19)

$$N_{ti}' = N_{ti} - N_s \qquad (5.34)$$

If the sludge age of the process is longer than the minimum required for nitrification ($R_s > R_{sm}$ for $S_f=1{,}00$, Eq. 5.14), the TKN concentration available for nitrification will be partially or almost wholly nitrified to nitrate and the effluent nitrate concentration (N_{ne}) is the difference between the nitrogen available for nitrification (N_{ti}', Eq. 5.34) and the effluent TKN concentration (N_{te}). If $R_s < R_{sm}$, no nitrification takes place; the effluent nitrate concentration (N_{ne}) is zero and the effluent ammonia concentration (N_{ae}) is equal to the nitrogen available for nitrification (N_{ti}', Eq. 5.34) minus the sum of the unbiodegradable and biodegradable TKN concentrations in the effluent ($N_{ue}+N_{oe}$). For both $R_s < R_{sm}$ and $R_s > R_{sm}$, the effluent TKN concentration (N_{te}) is the sum of effluent unbiodegradable and biodegradable organic nitrogen concentrations (N_{ue} and N_{oe} respectively) and the effluent ammonia concentration (N_{ae}).

For $R_s > R_{sm}$:

From Eq 5.13, the effluent ammonia concentration is given by

$$N_{ae} = K_{nT}(b_{nT} + 1/R_s)/\{\mu_{nmT} - (b_{nT} + 1/R_s)\} \qquad (5.35a)$$

Hence the effluent TKN concentration (N_{te}) is

$$N_{te} = N_{ae} + N_{oe} + N_{ue} \quad (\text{mgN}/\ell) \qquad (5.36a)$$

and the effluent nitrate concentration (N_{ne}) is

$$N_{ne} = N_{ti}' - N_{te} \quad (\text{mgN}/\ell)$$

$$= N_{ti} - N_s - N_{te} \quad (\text{from Eq. 5.34}) \qquad (5.37a)$$

Analogous to the concentration of active heterotrophic organisms (see Eq. 4.10), the nitrifier organism mass is given by

$$M(X_n) = M(N_{ne})\, Y_nR_s/(1+b_{nT}R_s) \quad (\text{mgVSS}) \qquad (5.38a)$$

where

$M(N_{ne})$ = mass of nitrate generated per day $= Q.N_{ne}$

where N_{ne} is given by Eq. (5.37a).

The oxygen demand for nitrification is simply 4,57 mgO/mgN times the mass of nitrate produced per day, i.e.

$$M(O_n) = 4{,}57\, M(N_{ne}) \quad (\text{mgO/d}) \qquad (5.39a)$$

For $R_s < R_{sm}$:

For $R_s < R_{sm}$, no nitrification takes place. Hence the effluent nitrate concentration is zero i.e.

$$N_{ne} = 0{,}00 \quad (\text{mgN}/\ell) \qquad (5.37b)$$

The effluent ammonia concentration is

$$N_{ae} = N_{ti}' - N_{oe} - N_{ue} \quad (\text{mgN}/\ell)$$

$$= N_{ti} - N_s - N_{oe} - N_{ue} \quad (\text{from Eq. 5.34}) \qquad (5.35b)$$

The effluent TKN concentration is

$$N_{te} = N_{ae} + N_{ue} + N_{oe} \qquad (5.36b)$$

The nitrifier sludge mass and the nitrification oxygen demand are zero, i.e.

$$M(X_n) = 0{,}0 \quad (\text{mgVSS}) \qquad (5.38b)$$

$$M(O_n) = 0{,}0 \quad (\text{mgO/d}) \qquad (5.39b)$$

Substituting the respective influent nitrogen concentrations for raw and settled wastewaters and the values of the kinetic constants at 14°C into Eqs. 5.31 to 5.39 and plotting the results at different sludge ages yields Figure 5.9. In Figure 5.9a, the different concentrations of nitrogen in the process versus sludge age for raw and settled wastewater at 14°C are shown. In Figure 5.9b the nitrifier sludge mass and nitrification oxygen demand for raw and settled wastewater at 14°C are shown. Also shown in Figure 5.9b are the carbonaceous and total oxygen demands for raw and settled wastewater at 14°C. The calculations were repeated for 22°C and are shown in Figure 5.10a and b.

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*The subscript e used here and later refers to the effluent. However, $N_o = N_{oe}$, $N_a = N_{ae}$, etc. because completely mixed conditions are assumed.

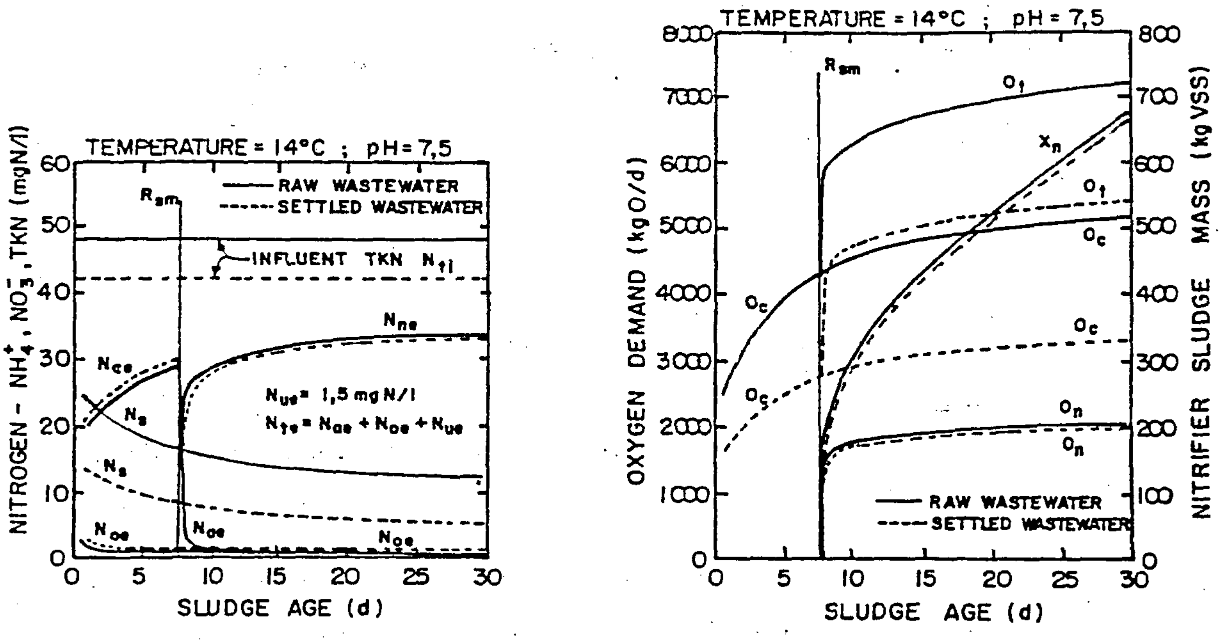

FIGURE 5.9a

FIGURE 5.9b

FIGURE 5.9: Effluent ammonia (N_{ae}), organic nitrogen (N_{oe}), and nitrate (N_{ne}) concentrations and nitrogen required for sludge production (N_s) (Figure 5.9a), carbonaceous (O_c), nitrification (O_n), and total (O_t) oxygen demand and the nitrifier sludge mass (X_n) (Figure 5.9b) versus sludge age for raw and settled wastewater at 14°C.

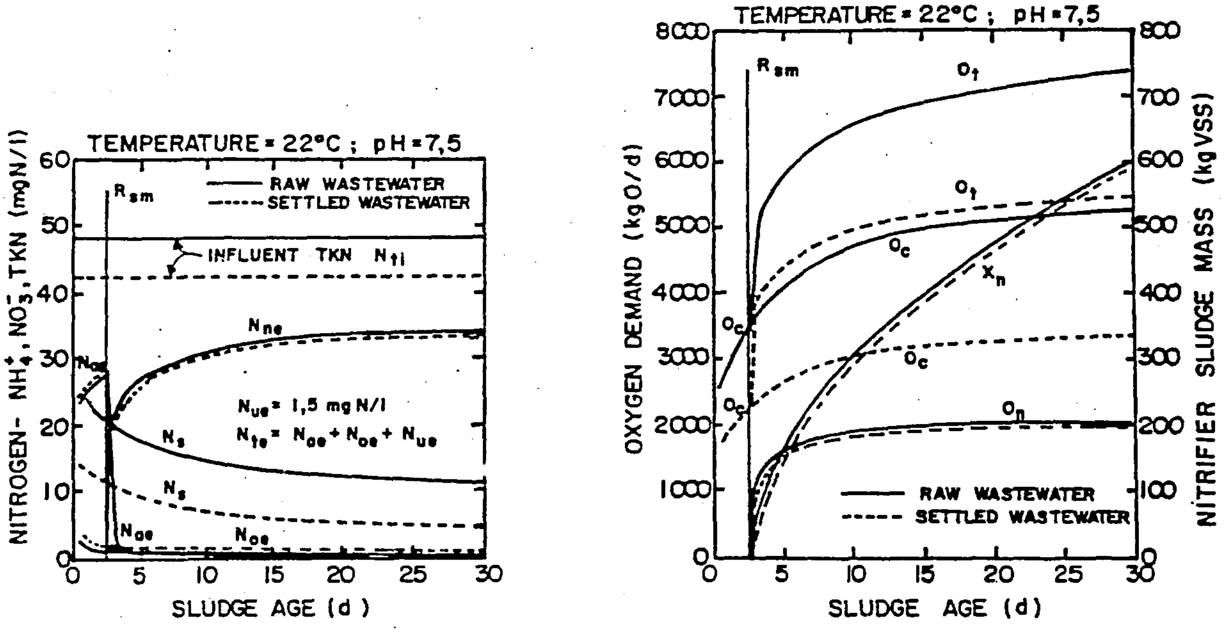

FIGURE 5.10a

FIGURE 5.10b

FIGURE 5.10: Effluent ammonia (N_{ae}), organic nitrogen (N_{oe}), and nitrate (N_{ne}) concentrations, nitrogen required for sludge production (N_s) (figure 5.10a) carbonaceous (O_c), nitrification (O_n), and total (O_t) oxygen demand and the nitrifier sludge mass (X_n) (Figure 5.10b) versus sludge age for raw and settled wastewater at 22°C.

Figures 5.9 and 5.10 show that once the sludge age is approximately 25 per cent longer than the minimum required for nitrification, nitrification is complete and comparing the results for raw and settled wastewater, there is little difference between the nitrification oxygen demand and the concentrations of ammonia, nitrate and TKN in the effluent. The reasons for this similar behaviour are: (1) the primary settling tank removes only a small fraction of the influent nitrogen*; and (2) settled wastewater results in lower sludge production, so that the nitrogen available for nitrification in raw and settled wastewater are approximately equal. Temperature has little effect on the different concentrations of nitrogen. However, a change in temperature causes a significant change in the minimum sludge age for nitrification.

Considering Figures 5.9a and 5.10a, the effluent organic nitrogen concentration (N_{oe}) is less than 2 mgN/ℓ for all sludge ages greater than about 3 days. For sludge ages about 25 per cent longer than the minimum for nitrification, the effluent ammonia concentration (N_{ae}) is less than 1 mgN/ℓ so that for $R_s > 1{,}25\ R_{sm}$ the effluent TKN concentration can be taken to be approximately 4 mgN/ℓ. The increase in nitrate concentration (N_{ne}) with increase in sludge age is principally attributable to the reduction in the nitrogen required for sludge production (N_s). This is important for nutrient removal plants as it indicates that increasing the sludge age of a process, increases the nitrification capacity (see section 5.2 above) even though the effluent TKN concentration remains constant at about 4 mgN/ℓ (see Figure 5.8). For $R_s < R_{sm}$, the effluent ammonia concentration (N_{ae}) and hence the effluent TKN concentration (N_{te}), increases with increasing sludge age up to R_{sm} because N_s reduces for increases in R_s.

Figures 5.9b and 5.10b show that the nitrification oxygen demand increases rapidly once $R_s > R_{sm}$ but if $R_s > 1{,}25\ R_{sm}$, the increase is marginal irrespective of the temperature or wastewater type, i.e. between sludge ages 10 and 30 days about 2000 kgO/d are required for nitrification. This nitrification oxygen demand represents an increase of 40% and 60% respectively** above the carbonaceous oxygen demand for the raw and settled wastewater. However, the *total* oxygen demand for treating settled wastewater is only 75% of that for treating raw wastewater.

In order that nitrification can proceed without inhibition by oxygen limitation, it is important that the aeration equipment is adequately designed to supply the total oxygen demand; generally heterotrophic organism growth takes precedence over nitrifier growth when oxygen supply becomes insufficient. This is because heterotrophic organisms can grow adequately with dissolved oxygen concentrations of 0,5 to 1,0 mgO/ℓ whereas nitrifiers require a minimum concentration of 1 to 2 mgO/ℓ.

The nitrifiers sludge mass given in Figures 5.9b and 5.10b shows a rapid increase once $R_s > R_{sm}$ and is approximately the same for raw or settled sewage, but is slightly higher at 14°C than at 22°C. Comparing the nitrifier sludge mass to the heterotrophic sludge mass, even at high TKN/COD ratios, the nitrifier sludge mass comprises only about 2 per cent of total volatile mass and may be neglected in the determination of the volatile solids concentration in the activated sludge process treating domestic wastewaters.

It is worth repeating that primary sedimentation removes only a minor fraction of the TKN but a significant fraction of COD. Even though the settled wastewater has a lower TKN than the raw wastewater the process effluent nitrate concentration will not reflect this difference because the nitrogen removed for sludge production is lower for settled than for raw wastewater. Consequently the nitrate generated is approximately the same for both waste flows. In contrast, the maximum denitrification possible (denitrification potential) in the process is dependent principally on the influent COD concentration and this concentration, we have seen, is significantly reduced by primary sedimentation. This may result in a situation where it may be possible to obtain complete nitrate removal when treating raw wastewater but highly unlikely when treating settled wastewater. This difference in removal can have a profound effect on the process design for biological excess phosphorus removal.

9. REFERENCES

ARKLEY, M. and MARAIS, G.v.R. (1982) Effect of the anoxic sludge mass fraction on the activated sludge process. Research Report No W 38, Dept. of Civil Eng., Univ. of Cape Town.

DOLD, P.L., EKAMA, G.A. and MARAIS, G.v.R. (1980) A general model for the activated sludge process. *Prog. Wat. Tech.*, 12, 47-77.

DOWNING, A.L., PAINTER, H.A. and KNOWLES, G. (1964) Nitrification in the Activated Sludge Process, *J. Proc. Inst. Sew/Purif.*, 64, 2, 130-158.

EKAMA, G.A., VAN HAANDEL, A.C. and MARAIS G.v.R. (1979) The present status of research on nitrogen removal: a model for the modified activated sludge process. Presented at the Symposium on nutrient removal, SA Branch of the IWPC, Pretoria, May 1979. Proc. published by Water Research Commission of South Africa.

HERBERT, D. (1958) Recent progress in microbiology. VII. International congress on microbiology. Ed. by Tunevall, Alinquist and Wiksell, Stockholm, 1958, 381-396.

LOEWENTHAL, R.E. and MARAIS, G.v.R. (1977) Carbonate chemistry of aquatic systems: Theory and application. Ann Arbor Science, Michigan, USA.

LOVELESS, J.E. and PAINTER, H.A. (1968) The influence of metal ion concentration and pH value in the growth of a *Nitrosomonas* strain isolated from activated sludge. *J. Gen. Micro.*, 52, 1-14.

MALAN, W.N. and GOUWS, E.P. (1966) Geaktiveerde slyk vir rioolwatersuiwering op Bellville. Research Report, Council for Scientific and Industrial Research, Nov. 1966.

MONOD, J. (1950) Technique of continuous culture – Theory and application, (Translation from French), *Ann. Inst. Pasteur*, 79, 167.

NEYTZEL-DE WILDE, F.G. (1977) Treatment of effluents from ammonia plants – Part 1. Biological nitrification of an inorganic effluent from a nitrogen-chemicals complex. *Water S.A.*, 3(3) 113-122.

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*This depends on the state of the sewage; if the sewage is "fresh" a larger fraction of the TKN can still be in particulate proteinaceous form.

**Roughly for $R_s > 1{,}25\ R_{sm}$, the increase in the total oxygen demand due to nitrification is given by $M(O_t) = (1 + 5\ f_{ns})M(O_c)$ where f_{ns} is the TKN/COD ratio of the wastewater. For processes including denitrification see Chapter 6, Section 7.6.2.

STENSTROM, M.K. and PODUSKA, R.A. (1980) The effect of dissolved oxygen concentration on nitrification. *Wat. Research*, **14**, 6, 643-649.

VAN HAANDEL, A.C. and MARAIS, G.v.R. (1981) Nitrification and denitrification in the activated sludge process. Research Report W 42, Dept. of Civil Eng., Univ. of Cape Town.

CHAPTER SIX

BIOLOGICAL NITROGEN REMOVAL

by

G.A. Ekama and G.v.R. Marais

1. PROCESS FUNDAMENTALS

In this chapter the kinetics of the single sludge nitrification-denitrification process will be considered in detail. It will be shown that for municipal wastewaters, significant or even complete, removal of nitrate is possible with this process.

There are two main biological activities whereby nitrigen is removed from the wastewater: (1) sludge production and (2) respiration for energy abstraction from organic compounds. These two reactions differ fundamentally in their biological objectives.

1.1 Nitrogen removal by sludge production

The nitrogen removal attributable to sludge production was considered in detail in Chapter 4, Section 9, and it was concluded that for most municipal wastewaters only a minor fraction of nitrogen could be removed by sludge production (see Figure 4.5).

1.2 Biological denitrification

This biological reaction is known as dissimilative reduction of nitrogen or denitrification and involves the reduction of nitrate or nitrite present in the wastewater to gaseous nitrogen which escapes to the atmosphere. Removal of nitrogen via this process is a consequence of biological redox reactions, to obtain energy from the organic material. In these reactions nitrate and nitrite serve the same function as oxygen, i.e. that of a hydrogen ion/electron acceptor (see Chapter 1).

Nitrate readily replaces oxygen as the terminal electron acceptor because the electron pathway is very similar to that of oxygen. The difference is in the electron transfer from the cytochromes where the specific enzyme reductase is replaced by the enzyme nitrate reductase which catalyses the final electron transfer to nitrate instead of oxygen (see Chapter 1). Studies of pure cultures of denitrifying organisms indicate that the presence of dissolved oxygen prevents the formation of the enzyme necessary for the final electron transfer to nitrate (Chang and Morris, 1972). (See Section 2.2 below).

From stoichiometry, it can be shown that when nitrate serves as the terminal electron acceptor, the equivalent mass of oxygen, (as O) if oxygen were the terminal electron acceptor, is

$$1\ \text{mgNO}_3 \text{ as N} \equiv 2{,}86\ \text{mgO as O} \qquad (6.1)$$

Knowing the equivalence between nitrate and oxygen, nitrate consumption for synthesis of new organism mass can be formulated with the equations for oxygen consumption by only slight modification. It also allows COD balances to be properly executed on denitrification processes.

1.3 Effect of denitrification on oxygen demand

In Chapter 5, Section 1, it was shown that when 1 mgN ammonia or TKN is converted to nitrate 4,57 mgO are required. In the section above it was shown that when 1 mgN nitrate is denitrified it is equivalent to 2,86 mgO being supplied. Thus for nitrification, 4,57 mgO/mgN are required, but in denitrification 2,86 mgO/mgN can be recovered, i.e. with denitrification 2,86/4,57.100 = 63% of the oxygen demand for nitrification can be recovered, if complete denitrification is achieved. Because the oxygen requirement for nitrification is about 25 to 35% of the total oxygen requirement, a saving of 15 to 20% of the total oxygen requirement can be made by including biological denitrification. This is demonstrated in Figure 6.1 where the oxygen demand for (i) carbonaceous, (ii) total including nitrification and (iii) total including nitrification and denitrification (per kgCOD load) is plotted versus sludge age for a typical raw wastewater (see Tables 4.3, 5.2 and 6.1).

Other advantages of processes including denitrification are discussed in Chapter 4, Section 11.2; see also Chapter 5, Section 4.3 and Chapter 6, Section 1.4.

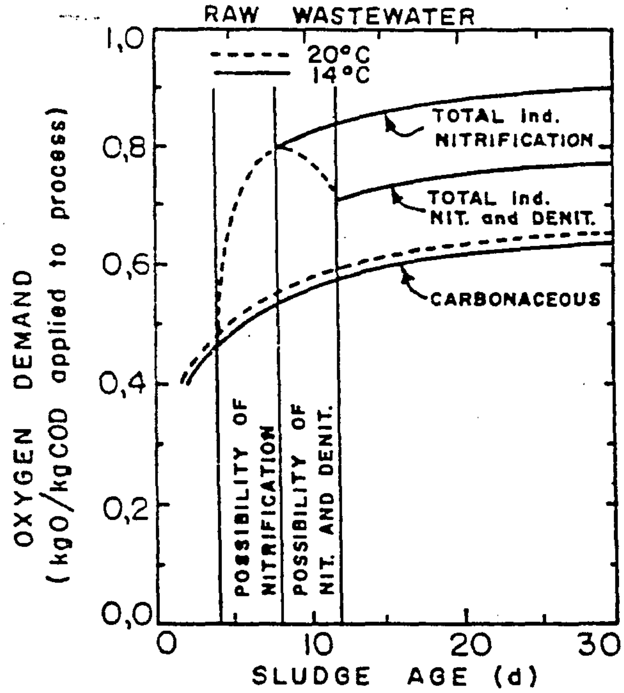

FIGURE 6.1: Carbonaceous, total including nitrification and total including nitrification and denitrification oxygen demand per unit COD load versus sludge age for raw wastewater.

1.4 Effect of denitrification on Alkalinity

Biological denitrification in which nitrate is reduced to nitrogen gas is accompanied by an increase in Alkalinity. From stoichiometry, for 1 mgN nitrate denitrified, 3,57 mg as $CaCO_3$ Alkalinity is produced. In Chapter 5, Section 4.3 it was shown that in the nitrification process, 7,14 mg/ℓ as $CaCO_3$ alkalinity was lost for every 1 mgN/ℓ ammonia nitrified. Consequently, through denitrification, half the decrease in alkalinity due to nitrification can be recovered.

The recovery of alkalinity by denitrification is important when treating wastewaters with low alkalinity values – less than 200 mg/ℓ as $CaCO_3$. For such wastewaters the production of only 15 mgN/ℓ nitrate without denitrification will reduce the mixed liquor pH value to below 5. In this event, because nitrification is extremely sensitive to pH values below 7, only partial nitrification will be obtained, even at long sludge ages. However, should denitrification be incorporated in the process, the recovery of alkalinity by denitrification usually will cause the mixed liquor pH value to be maintained near 7 even for low alkalinity wastewaters. This favourable effect, of maintaining the mixed liquor pH value near 7, makes a nitrification – denitrification process virtually mandatory for activated sludge processes treating wastewaters with low alkalinity values. This aspect is discussed in detail in Chapter 5, Section 4.3 and Chapter 4, Section 11.2.

2. REQUIREMENTS FOR DENITRIFICATION

For denitrification to occur, the following conditions are necessary:

(1) Presence of nitrate (or nitrite)
(2) Absence of dissolved oxygen
(3) A facultative bacterial mass
(4) Presence of a suitable electron donor (energy source).

2.1 Presence of nitrate

Normally the presence of nitrate implies nitrification as a prerequisite. The conditions for adequate nitrification in an activated sludge process have been discussed in detail in Chapter 5, Section 4. The prerequisite for nitrification involves the estimation of the maximum sludge mass fraction which may be left unaerated but still ensuring nitrification.

2.2 Absence of dissolved oxygen

The inhibitory effect of dissolved oxygen on denitrification has been extensively reported in the literature. In a comparative test series, Carlson (1972) found that at zero dissolved oxygen level the nitrate removed was 100 per cent, while at 0,2 mg/ℓ dissolved oxygen (2 to 10 per cent oxygen saturation), no significant denitrification was obtained. Furthermore, it should be remembered that in a completely mixed anoxic reactor, even if no measurable dissolved oxygen concentration (DO) is present, any oxygen entering such a reactor will be utilized preferentially by the organism mass, thereby reducing (in stoichiometric proportion) the mass of nitrate the reactor can denitrify. Possible sources of DO are: (1) high DO concentrations in the recycles from the aerobic zones to the anoxic zones – the DO concentration should be controlled to be between 1 and 2 mgO/ℓ i.e. sufficiently high so as not to inhibit nitrification in the aerobic reactor but not too high to reduce significantly the denitrification in the anoxic reactor; (2) entrainment into the anoxic reactor via the air-water interface due to unnecessarily high mixing intensities; (3) Archimedian screw pumps for recycle pumping; (4) hydraulic jumps, cascades and shooting flow conditions in interconnecting channels – closed channels and tranquil flow conditions entrain less oxygen. (See also Chapter 10 on operation of nutrient removal plants).

2.3 Facultative bacterial mass

The propensity to denitrify is widespread among bacteria. Dissimilative denitrification with end products N_2, NO and NO_2 has been established in numerous cases. The bulk of the bacterial mass in wastewater treatment is facultative and a significant fraction is capable of dissimilative denitrification. Studies on denitrification tend to support the view that there is little difference between the bacterial masses in processes where nitrification only or nitrification-denitrification takes place. There is little merit therefore in attempting to analyse the bacterial composition of the sludge in detail. A sludge generated under aerobic conditions, when subjected to the appropriate environmental conditions, will show a denitrifying capability immediately and will continue to do so, subsequently without apparent change in reactivity.

2.4 Electron donor (Energy source)

The oxidation of carbonaceous compounds with nitrate as electron acceptor i.e. denitrification, provides the energy required by the faculative heterotrophs for synthesis of new mass and endogenous respiration. The electron donor constitutes the source of the energy. The axis about which all biological denitrification investigations revolve is the energy source which serves as the electron donor in the denitrification process. Denitrification has been investigated for a variety of substances as carbonaceous energy sources. These can be categorised as follows:

(a) Energy sources not present in the wastewater, i.e. *an external carbonaceous energy source* which is added at the denitrification stage of the process. Compounds used as an external energy source include methanol, methane, ethanol, acetone and acetic acid.

(b) Energy sources present in the influent wastewater, i.e. *an internal carbonaceous energy source* which enters the system with the wastewater.

(c) Energy sources which are *self-generated* within the system by the release of nutrient by the organisms in the death phase.

The type of energy source utilised has led to different process configurations for denitrification studies. Usually in municipal wastewater treatment only energy sources (b) and (c) above are of interest as these are sources of energy available in the single sludge activated sludge process. However, in some nutrient removal plants a readily biodegradable COD source may be added to the anaerobic reactor to improve the biological P removal (see Chapter 4, Section 8 and Chapter 7, Section 3.3). This practice not only improves P removal but also the removal of nitrate by denitrification as it constitutes the addition of an external energy source (group a). Sources of readily biodegradable COD are acid digester supernatant or some form of organic wastewater with a large readily biodegradable COD fraction such as malting industry wastewater, fruit and vegetable canning factory wastewaters, juicing factory wastewaters, etc. The addition of these wastes increases the nitrate removal potential by denitrification because the additional readily biodegradable COD allows the initial rapid rate of denitrification to operate for a longer period of time (see Section 4.1.1 below).

3. DENITRIFICATION PROCESS CONFIGURATION

Two basic activated sludge process configurations have been developed for single sludge biological denitrification: (1) the Wuhrmann configuration and (2) the Ludzack-Ettinger configuration.

3.1 The Wuhrmann configuration

The single sludge nitrification-denitrification system in which endogenous energy release provides the energy source for denitrification was first proposed by Wuhrmann (1964). A schematic presentation of the process is shown in Figure 6.2. It consists of two reactors in series, the first aerobic and the second anoxic. The influent is discharged to the first reactor where aerobic growth of both the heterotrophic and nitrifying organisms takes place. Provided the sludge age is sufficiently long and the aerobic fraction of the system is adequately large, nitrification will be complete in the first reactor.

The mixed liquor from the aerobic reactor passes to the anoxic reactor, also called the *post-denitrification reactor* or *secondary anoxic reactor* where it is kept completely mixed by stirring, but with no aeration. The outflow from the anoxic reactor passes through a settling tank and the underflow is recycled back to the aerobic reactor.

Energy release by the sludge mass due to the death of organisms provides the energy source for denitrification in the anoxic reactor. However, the rate of release of energy is low, so that the rate of denitrification is also low. Consequently, in order to obtain a meaningful reduction of the nitrate concentration in the anoxic reactor, the anoxic fraction of the system must be large and this may cause a breakdown of the nitrification process. Thus, although theoretically the system has the potential to remove all the nitrate, from a practical point this usually is not possible as the anoxic volume fraction will need to be so large that the conditions for nitrification cannot be satisfied particularly if the temperatures are low, below 15°C. Furthermore, in the anoxic reactor, organic nitrogen and ammonia are released through organism death, some of which passes out with the effluent thereby reducing the total nitrogen removal of the system. To minimise the ammonia content of the effluent, a flash or reaeration reactor may be placed between the anoxic reactor and the settling tank. In this reactor the ammonia is then nitrified to nitrate thereby reducing overall efficiency of the nitrate reduction capability of the process.

3.2 The Ludzack-Ettinger configuration

Ludzack and Ettinger (1962) were the first to propose a single sludge nitrification-denitrification process utilizing the biodegradable material in the influent as an energy source for denitrification. A schematic presentation of this system is shown in Figure 6.3. It consists of two reactors in series, *partially* separated from each other. The influent is discharged to the first reactor which is maintained in an anoxic state by stirring without aeration. The second reactor is aerated and

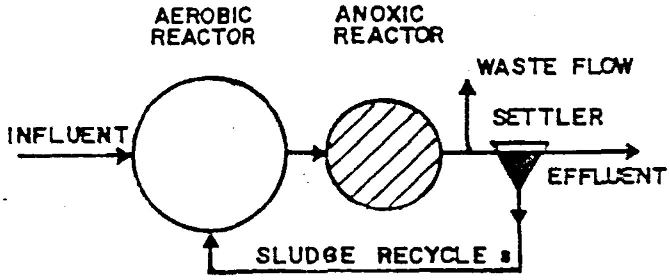

FIGURE 6.2: The Wuhrmann process for nitrogen removal.

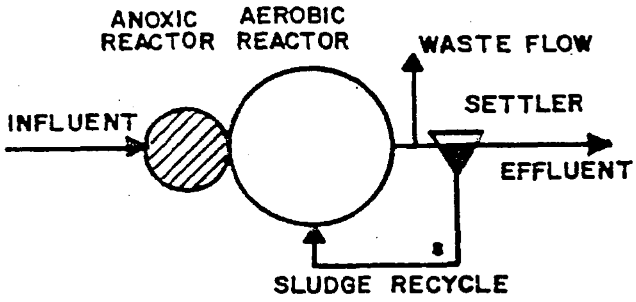

FIGURE 6.3: The Ludzack-Ettinger process for nitrogen removal.

nitrification takes place. As there is only partial separation, the mixed liquor in the first reactor is in communication with that in the second reactor. Due to the mixing action in both reactors, an interchange of the nitrified and anoxic liquors is induced, and the nitrate entering the anoxic reactor is reduced to nitrogen gas. Ludzack and Ettinger reported that the process gave variable denitrification results, probably due to the lack of control of the interchange of the contents of the two reactors.

In 1973, Barnard in developing the Bardenpho process proposed an improvement of the Ludzack-Ettinger process, by completely separating the anoxic and aerobic reactors, recycling the underflow from the settler to the anoxic reactor, and providing an additional recycle from the aerobic to the anoxic reactor (Figure 6.4).

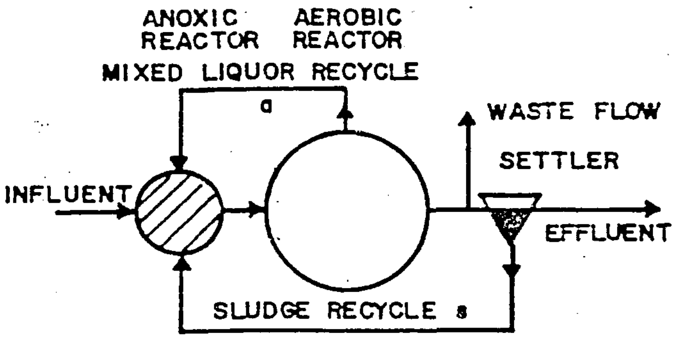

FIGURE 6.4: The modified Ludzack-Ettinger process for nitrogen removal.

These modifications allow a significant improvement in control over the process performance. The high influent energy source discharged to the anoxic reactor also called the *pre-denitrification reactor* or *primary anoxic reactor*, gives rise to a high rate of denitrification and a substantially higher reduction of nitrate than in the Wuhrmann process, even when the pre-denitrification reactor of this process is substantially smaller than the post-denitrification reactor of the Wuhrmann process. For the purposes of identification, this process will be called the modified Ludzack-Ettinger process.

With the modified Ludzack-Ettinger process, complete denitrification cannot be achieved because a part of the total flow from the aerobic reactor is not recycled to the anoxic reactor but is discharged directly with the effluent.

3.3 The Bardenpho configuration

In order to overcome the deficiency of incomplete denitrification in the modified Ludzack-Ettinger process, Barnard (1973) proposed combining this process with that of Wuhrmann and called it the Bardenpho process. A schematic presentation of the Bardenpho process is shown in Figure 6.5. Barnard considered that the low concentration of nitrate discharged from the aerobic reactor to the secondary anoxic reactor will be denitrified to produce a relatively nitrate-free effluent. To strip the nitrogen bubbles generated in the secondary anoxic reactor attached to the sludge flocs, he introduced a flash aeration reactor between the secondary anoxic and the final settling tank. The flash aeration reactor was also considered necessary to nitrify the ammonia released during the sludge residence time in the secondary anoxic reactor. In order to reduce the possibility of flotation of sludge in the settler due to denitrification of residual nitrate, the sludge accumulation in the settler was to be kept to a minimum. This was achieved by having a high recycle rate from the settler, approximately equal to the mean influent flow.

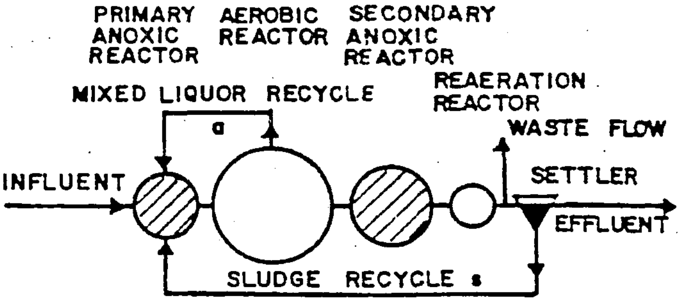

FIGURE 6.5: The Bardenpho process for nitrogen removal.

Although in concept the Bardenpho process has the potential for complete removal of nitrate, in practice this is not always possible, an aspect that will be discussed in detail in Section 6 below.

4. EXPERIMENTAL BASIS FOR DENITRIFICATION KINETICS

Carlson (1971) and Christensen and Harremoës (1977) suggested that the kinetic reaction for denitrification by activated sludge mixed liquor can be expressed by

$$dN_n/dt = KX \tag{6.2}$$

where

dN_n/dt = denitrification rate (mgN nitrate/litre/unit time)

N_n = nitrate concentration (mgN/ℓ)

t = time (hours or days)

X = volatile solids concentration (mgVSS/ℓ)

K = specific denitrification constant (mgN/mgVSS/time).

This expression for the denitrification rate was developed from batch tests. It indicates that the nitrate-time relationship is linear and independent of the nitrate concentration, i.e. the rate is zero order with respect to nitrate concentration, and a function only of the volatile sludge concentration.

In order to evaluate Eq. 6.2 as a basis for describing denitrification behaviour in the single activated sludge process, Marais and a number of co-workers (Stern, 1974; Wilson, 1976; Marsden, 1977), undertook an extensive investigation into the kinetic behaviour of the modified Ludzack-Ettinger and Wuhrmann configurations using plug flow mixing regimes for the respective primary and secondary anoxic reactors. By measuring the nitrate concentration along the length of the plug flow anoxic reactors, the denitrification kinetic behaviour was determined by plotting nitrate concentration-time profiles. Experiments were carried out using only raw and settled wastewaters at different temperatures, sludge ages, sludge concentrations, influent COD concentrations and recycle ratios.

The investigation was divided into two parts: (1) constant flow and load conditions, and (2) cyclic flow and load conditions.

4.1 Constant flow and load conditions

4.1.1 Experimental work with plug flow reactors

Typical nitrate concentration-time profiles for secondary and primary anoxic plug flow reactors are shown in Figures 6.6 and 6.7 respectively. The secondary anoxic profiles all exhibited a single linear phase that appeared to be independent of the nitrate concentration i.e. the profiles exhibited a behavioural pattern not at variance with Eq. 6.2. The primary anoxic profiles, however, exhibited profiles that all indicated a two-stage denitrification response, i.e. (1) a fast primary phase of short duration (1 to 10 mintes), and thereafter (2) a slow second-

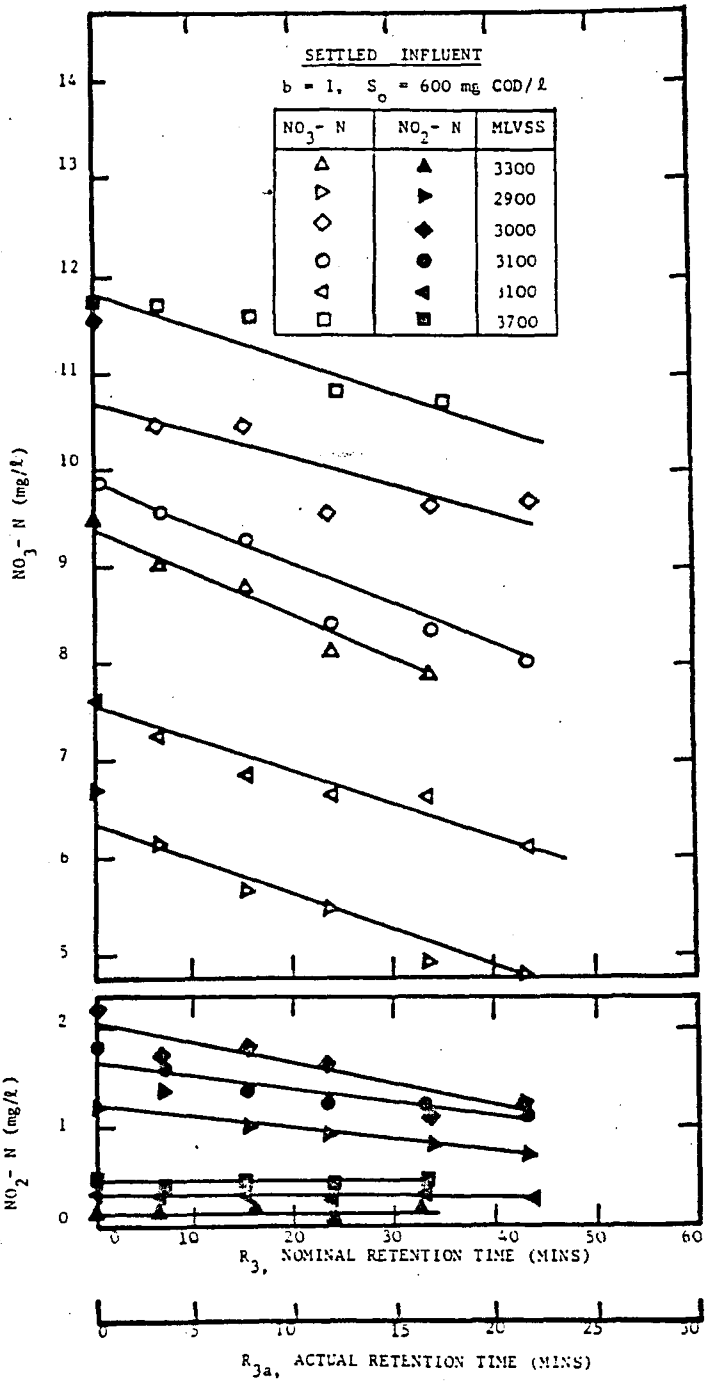

FIGURE 6.6: Nitrate concentration profiles observed in a plug-flow secondary anoxic reactor at 20°C (see Figure 6.2. After Stern and Marais, 1974).

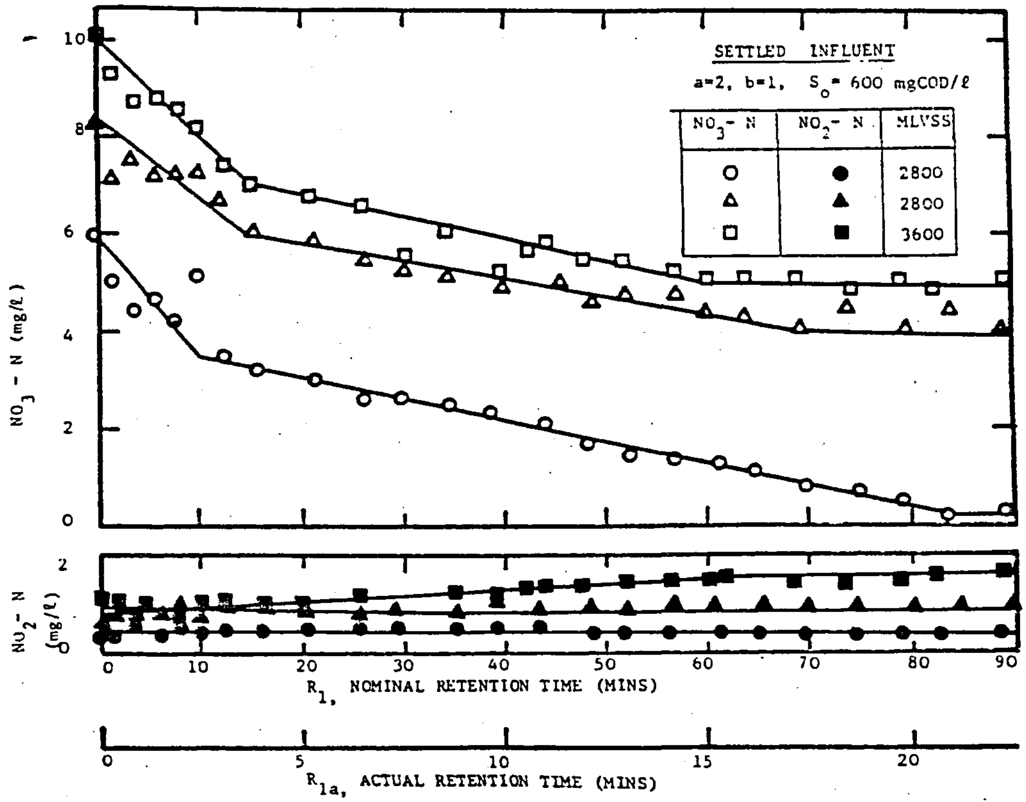

FIGURE 6.7: Nitrate concentration profiles observed in a plug-flow primary anoxic reactor at 20°C (see Figure 6.4. After Stern and Marais, 1974)

ary phase at about one seventh the rate of that of the primary phase.

In order to fit the experimental data into the functional form expressed by Eq. 6.2, it is necessary to discuss the following aspects: (i) sludge concentration, (ii) denitrification rate constants, (iii) recycle ratios, and (iv) sludge age.

(i) *Sludge concentration:* As the denitrification reaction is mediated by the active organisms, Eq 6.2 should be expressed in terms of the active mass concentration X_a, i.e.

$$\Delta N_n = K X_a \Delta t \tag{6.3}$$

In all the investigations by Marais and his co-workers it was found that the sludge production in anoxic-aerobic systems was not distinguishably different from a comparable aerobic system, provided that the anoxic volume fraction was not too great (i.e. less than about 50 per cent). Consequently, the active mass concentration X_a in the system could be expressed by Eq. 4.10 (see Chapter 4, Section 4), (Arkley and Marais, 1982).

(ii) *Denitrification rate constants, K:* If the denitrification rate constant, K, maintains the same value throughout the actual residence time in the plug-flow reactor, R_a, then the difference between the influent and effluent nitrate concentrations, N_{ni} and N_{no} respectively, can be written in terms of Eq. 6.3

$$N_{ni} - N_{no} = \Delta N_{na} = K X_a R_a \tag{6.4}$$

where

R_a = actual retention time in plug flow reactor (h or d)

Eq. 6.4 is valid only if the effluent nitrate from the reactor is greater than zero, see Eq. 6.17 below. The difference $(N_{ni} - N_{no})$ is the *actual* removal of nitrate concentration in the *reactor*. However, as the rate of flow through the reactor is Q (1+a) (where Q = units of influent flow to the plant per unit of time and a = recycle ratio), the flow passes (a+1)/1 times through the reactor. This gives the *system* removal, N_{ns}, i.e.

$$\Delta N_{ns} = (a+1)(N_{ni} - N_{no}) \tag{6.5}$$

The *nominal* retention time of the plug-flow reactor is defined by

$$R_n = V_r/Q \tag{6.6}$$

where

R_n = nominal retention time

V_r = volume of the reactor (*ℓ*)

Q = daily average influent flow to process.

Hence the relationship between the actual and nominal retention times is

$$R_a = R_n/(1+a) \tag{6.7}$$

Substituting Eq. 6.4 for $N_{ni} - N_{no}$ and Eq 6.7 for R_a into Eq. 6.5, the systen nitrate removal is given by

$$\Delta N_{ns} = K X_a R_n \qquad (6.8)$$

Comparing Eqs. 6.4 and 6.8, note that K is independent of the recycle ratio a. Hence K can be determined directly from Eq. 6.4 i.e. from the actual observed reduction in the nitrate concentration in the anoxic reactor. With these basic equations the nitrate reduction in the primary and secondary anoxic reactors now can be formulated.

- Secondary anoxic reactor:

Referring to the experimental secondary anoxic reactor profiles, Figure 6.6, only one constant linear nitrate time reduction phase was observed and the analyses above allow the denitrification rate constant in this reactor, K_3 to be determined directly from Eq. 6.4 i.e.

$$\Delta N_{nss} = K_3 X_a R_{ns} \qquad (6.9)$$

where the additional subscript s refers to the secondary anoxic reactor.

By applying Eq. 6.9 to all the profiles observed experimentally, a set of K values was obtained at 14°C and 20°C respectively and the mean value determined by a graphical statistical analysis. The statistical plot of the K_3 values obtained at 14°C is shown in Figure 6.8. From the mean K_3 values at 14°C and 20°C the effect of temperature on K_3 was formulated as

$$K_{3T} = 0{,}072\ (1{,}03)^{T-20}\ (\mathrm{mgNO_3\text{-}N/mgVASS/d}) \qquad (6.10)$$

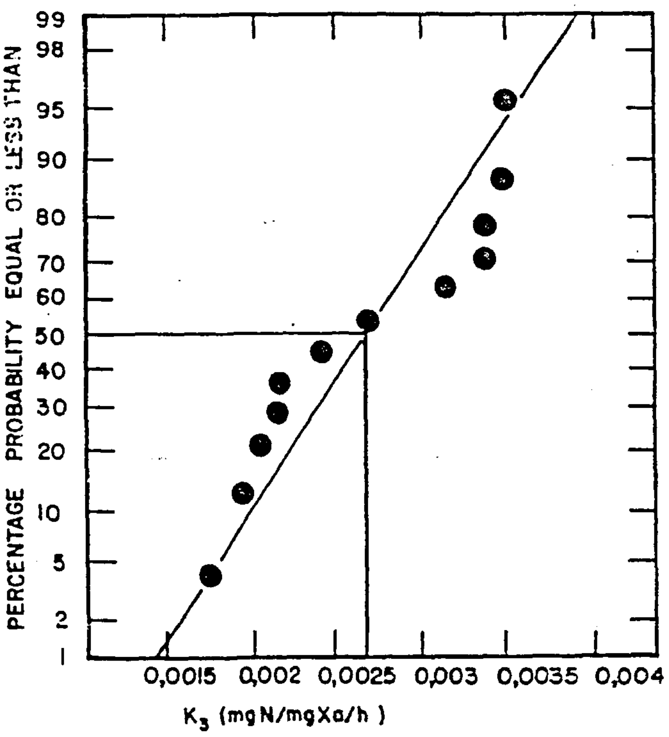

FIGURE 6.8: Statistical plot of the observed denitrification rate constants in a secondary anoxic reactor (K_3).

- Primary anoxic reactor:

Referring to the primary anoxic reactor profiles, two linear reduction phases were observed. The first phase persisted only for a limited period, the second phase persisted for the remaining retention time in the flug flow reactor, see Figure 6.7. To determine the denitrification rate constants of the two phases, it was presumed that the second phase was also operative during the first phase i.e. the slope of the nitrate-time profile of the first phase is the sum of the slopes of the two phases; the slope of the first phase is governed by $(K_1 + K_2)$. Later it will be shown that there is good reason to believe that this is indeed so – see Section 4.1.2 below. The separation of the phases and their associated K values are best appreciated by graphically setting out the nitrate-time profile as shown in Figure 6.9. Hence the *system* reduction in the pre-denitrification reactor can be written:

$$\begin{aligned}\Delta N_{nps} &= \Delta N_{n1s} + \Delta N_{n2s}\\ &= K_1 X_a t_1 (a+1) + K_2 X_a R_{ap} (a+1)\\ &= K_1 X_a t_1 (a+1) + K_2 X_a R_{np} \qquad (6.11)\end{aligned}$$

where

t_1 = duration of the first denitrification phase; subscripts 1 and 2 refer to the first and second phases of denitrification; subscript p refers to the primary anoxic or pre-denitrification reactor; subscripts a and n on the R parameter refer to actual and nominal retention times respectively.

From the many experimental profiles the values of K_1 and K_2 were calculated using the construction set out in Figure 6.9, and Eq. 6.11. The statistical plots of the K_1 and K_2 values at 14°C and 20°C are shown in Figure 6.10. From the mean values at these two temperatures, the following temperature dependency equations were obtained,

$$K_{1T} = 0{,}720\ (1{,}20)^{(T-20)}\ (\mathrm{mgNO_3\text{-}N/mgVASS/d}) \qquad (6.12)$$

$$K_{2T} = 0{,}1008\ (1{,}08)^{(T-20)}\ (\mathrm{mgNO_3\text{-}N/mgVASS/d}) \qquad (6.13)$$

for $T \geqq 13°C$

$K_{2T} = K_{3T}$ for 13°C.

The temperature sensitivity of the denitrification rate constants given above and in Eq. 6.10 can be considered reliable only in the range 12°C to 24°C.

From the experimental data as analysed above, Marais and co-workers came to the conclusion that the system reduction of nitrate in the first phase of denitrification, ΔN_{n1s}, was approximately proportional to the influent *biodegradable* COD concentration, S_{bi} (see Eq. 2.7) i.e.

$$\Delta N_{n1s} = \alpha S_{bi} \qquad (6.14)$$

Experimentally, for raw or settled wastewater, the mean value of α was found to be 0,028 mgN/mg *biodegradable* COD, and this value of α appeared to be independent of temperature. Taking account

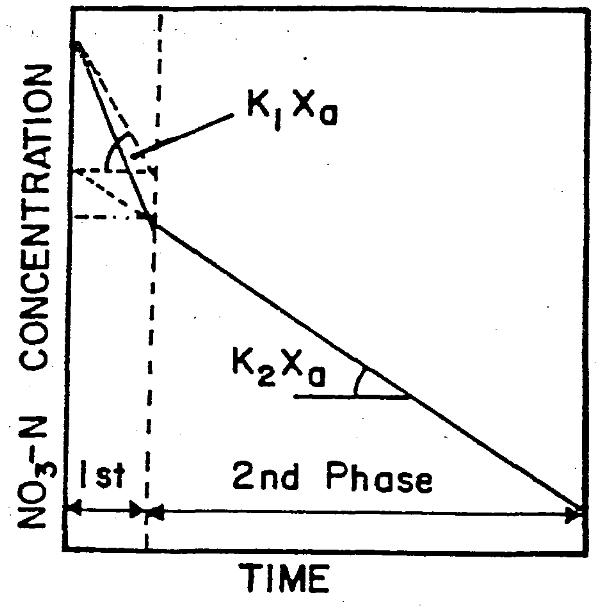

FIGURE 6.9: Denitrification rate constants K_1 and K_2 in a primary anoxic reactor.

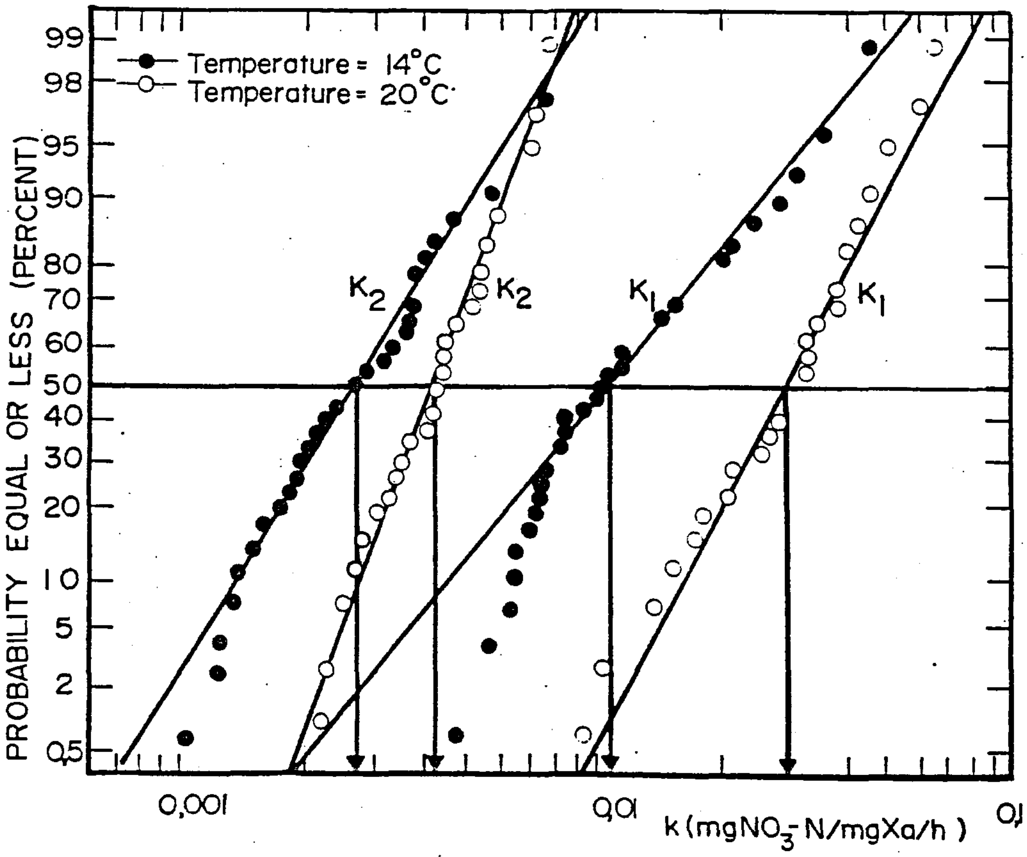

FIGURE 6.10: Statistical plots of denitrification rate constants K_1 and K_2 (pre-denitrification) at 14°C and 20°C.

of these characteristics, the system nitrate reduction in the primary anoxic reactor can be written as

$$\Delta N_{nps} = \alpha\, S_{bi} + K_2 X_a R_{np} \quad (6.15)$$

and the duration of the first phase of denitrification can be found from Eq. 6.11 as

$$t_1 = \alpha\, S_{bi}/\{K_1 X_a (1 + a)\} \quad (6.16)$$

From Eq. 6.16, the duration of the first phase of denitrification in the primary anoxic reactor, t_1, depends on the recycle ratio and the active mass concentration. Usually t_1 lies in the range 2 to 10 minutes.

(iii) *Recycle ratios:* The total system nitrate reduction of the primary and secondary anoxic reactors is the sum of Eqs. 6.9 and 6.15 i.e.

$$\Delta N_{nts} = \alpha\, S_{bi} + K_2 X_a R_{np} + K_3 X_a R_{ns} \quad (6.17)$$

where

ΔN_{nts} = total system nitrate removal.

An important point to note is that Eq. 6.17 is valid only if the nitrate is not reduced to zero in any one of the anoxic reactors. Where this condition is satisfied, then from Eq. 6.17 the magnitude of the recycle ratio does not affect the system reduction. However, there must be a minimum recycle; this is determined by the recycle value that *just* causes the nitrate to be zero at the end of a plug flow reactor. If the recycle is less, the nitrate will become zero before the end of the reactor is reached – the recycle can be increased (with concomitant increase in the nitrate removal) to the point where nitrate *just* starts to appear in the effluent. Once nitrate appears in the effluent from all the anoxic reactors, a further increase in the recycle will have no effect on the system nitrate reduction.

(iv) *Sludge age:* Marais and co-workers found experimentally that at a specific temperature the value of the respective denitrification constants appeared to remain unchanged irrespective of the sludge age for sludge ages between 10 and 25 days.

4.1.2 Nitrate removal and substrate utilization

It has been established in the biological growth kinetic theory that for the utilization of 1 mgCOD under aerobic conditions, an amount of $(1 - f_{cv}Y_h)$ mg oxygen is required. Numerically with $f_{cv} = 1{,}48$ mgCOD/mgVSS and $Y_h = 0{,}45$ mgVSS/mgCOD, the oxygen consumption per substrate utilization is 0,33 mg/mgCOD (Chapter 1, Section 4). In an anoxic reactor with substrate utilisation, nitrate is reduced instead of oxygen. As the oxygen equivalent of nitrate is 2,86 mgO/mgN (from Eq. 6.1), the nitrate consumption per mgCOD utilised is $(1 - f_{cv}Y_h)/2{,}86 = 0{,}116$ mgN/mgCOD or conversely 8,6 mgCOD are required to reduce 1 mgN nitrate. From Eq. 6.14 the nitrate concentration removed during the first phase of denitrification is αS_{bi} where $\alpha = 0{,}028$ and S_{bi} = biodegradable influent COD concentration. Thus the COD reduction during the first phase of denitrification is $0{,}028.8{,}6.S_{bi} = 0{,}24\ S_{bi}$. From measurements of the readily biodegradable COD in the influent feed, it was found from Eq. 2.8 that the readily biodegradable COD fraction f_{bs} of raw wastewater was 0,24. It was therefore hypothesized that the rapid first phase of denitrification is associated with the utilization of the readily biodegradable COD of the influent. It was also hypothesized that the second slower denitrification rate in the primary anoxic reactor and the rate in the secondary anoxic reactor are associated with the utilization of the slowly biodegradable particulate COD. From a simulation study of the denitrification behaviour in primary and secondary plug flow anoxic reactors, it was concluded that both these hypotheses appear to be valid. Details of this work (see van Haandel and Marais, 1981) are not important here, however, the following conclusions are: (i) the utilization of readily biodegradable COD and slowly biodegradable COD take place simultaneously (see Section 4.1.1 and Figure 6.9) and (ii) the second rate in the primary anoxic reactor (K_2) and the rate in the secondary anoxic reactor (K_3) both arise from the utilization of adsorbed slowly biodegradable COD but the K_2 is more rapid than K_3 because the concentration of adsorbed slowly biodegradable COD in the primary anoxic reactor is higher than that in the secondary anoxic reactor – in the primary anoxic reactor the adsorbed slowly biodegradable COD arises from that in the influent and organism death whereas that in the secondary anoxic reactor arises from organism death only.

4.1.3 Completely mixed anoxic reactors

Up to this stage, only plug flow reactors have been considered. Although this flow regime is important from a research point of view, it has little practical application in full scale nitrification-denitrification plants. Because the denitrification reaction in an anoxic reactor is approximately zero order with respect to nitrate, the reduction achieved in a reactor is virtually independent of the flow regime. Consequently, the equation developed for the nitrate reduction, Eq. 6.17 should be valid also for completely mixed reactors. Where the plug flow reactors were replaced by completely mixed reactors of the same volume, experimental data indicate that the reductions are substantially the same.

Generally, completely mixed reactors are not useful to obtain basic time behaviour characteristics as it is possible only to measure influent and effluent concentrations. However, if the reduction is a single zero order reaction (as in the case of the secondary anoxic reactor), the correct K value can be calculated from the influent and effluent nitrate concentrations (provided the latter is not zero) but if the reduction is due to two zero order reactions (as in the case of the primary anoxic reactor), the correct K values cannot be separated out; the combined effect of the two K values is to give rise to an apparent K value that varies with the retention time. This point is mentioned because in many instances the literature gives primary anoxic denitrification rate constants calculated from influent and effluent nitrate concentrations from completely mixed primary anoxic reactors. A further cause of error is that the denitrification rate constants in the literature often are given in terms of mg nitrate removed per mg *volatile suspended solids* per unit time. Constants reported in terms of total VSS should be used with caution because the denitrification rates are first order with respect to the *active* volatile suspended solids (X_a) (see Eq. 6.3), not first order with respect to the total volatile suspended solids (X_v) – the inert fraction of the VSS varies significantly with sludge age and wastewater type (see Figure 4.2), so that the denitrification rate constants in terms of total VSS cannot be applied to wastewaters and sludge ages different to those on which the rates were measured. However, if the wastewater characteristics, sludge age and temperature are known then, via the activated sludge theory given in Chapter 4, the denitrification rate constant, if given in terms of VSS, may be corrected to be in terms of the active volatile solids concentration.

4.2 Cyclic flow and load conditions

When denitrification experiments were run under cyclic loading conditions, the approach outlined above for calculating the extent of denitrification no longer appeared to be applicable – the total nitrate concentration removed was less than observed under steady state loading conditions, the removal of nitrate decreased as the severity of the cyclic load variation increased, and the TKN in the effluent showed a marked cyclic response, being high during the high loading phase and low during the low loading phase, whereas the nitrate behaved in exactly the opposite fashion. It was evident that to establish a general denitrification model, an alternative approach to that established for analysing the steady state behaviour was needed. This was sought in the general model for the activated sludge process developed by Ekama and Marais (1978). The details of this research work are not essential for using this monograph, but it can be mentioned that a large measure of success was achieved in predicting the behaviour of denitrification under dynamic loading conditions once a number of modifications to the model of Ekama and Marais (1978) were accepted. Interested readers are referred to the following publications: Dold, Ekama and Marais (1980); Ekama, van Haandel and Marais (1979); Siebritz, Ekama and Marais (1980); van Haandel and Marais (1981); van Haandel, Ekama and Marais (1981) and van Haandel, Dold and Marais (1982).

The outcome of the research work described in the references above, was a general dynamic model of the activated sludge process including nitrification and denitrification. The theoretical predictions of the general model correlated well with the experimental data observed under both constant and cyclic loading conditions at laboratory and pilot scale.

The general dynamic model, which describes the activated sludge process in basic biological mechanisms, was utilised to study the validity of the empirical denitrification constants K. A comparison of the theoretically predicted and experimentally observed K values is given in Figures 6.11 and 6.12. These figures

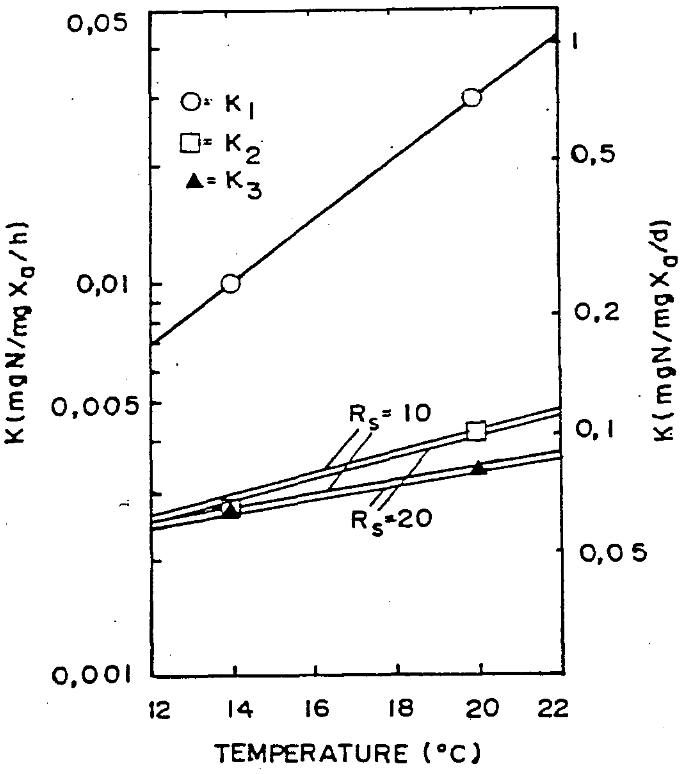

FIGURE 6.11: Simulated and experimental denitrification rate constants K_1, K_2 and K_3 versus temperature for 10 and 20 days sludge age.

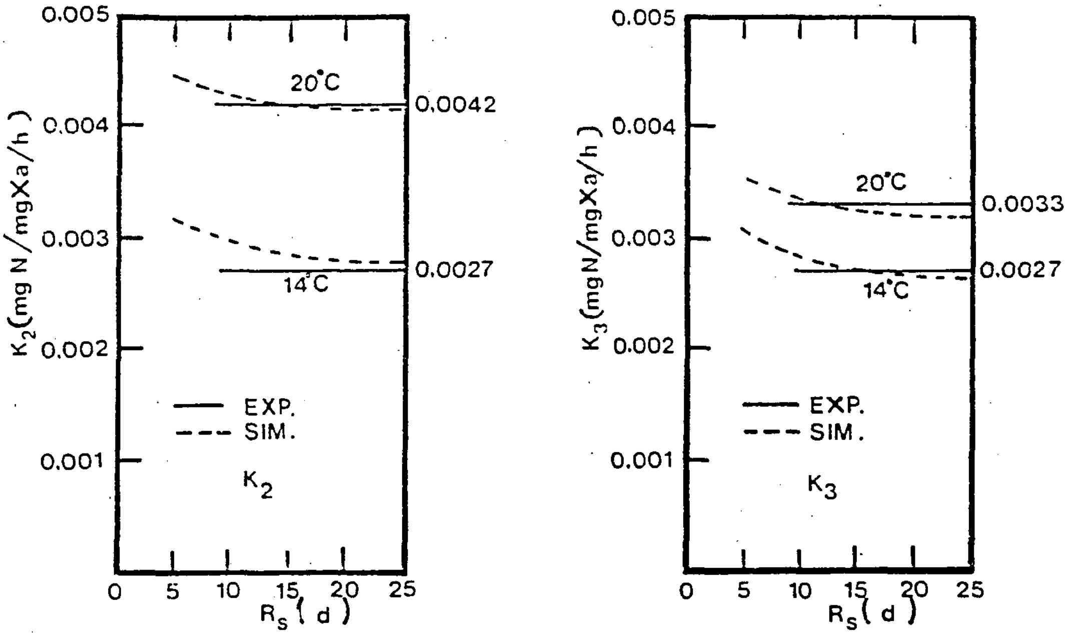

FIGURE 6.12: Simulated and experimental denitrification rate constants K_2 and K_3 versus sludge age at 14°C and 20°C.

show that for sludge ages between 10 and 30 days and temperatures between 12 and 22°C, good correlation between simulated and experimental K values is obtained. The good correlation indicates that although the K values are empirical and have no fundamental biological kinetic basis (of the kind on which the general dynamic model is based), they nevertheless appear to have such consistency that they can be applied to the design of anoxic reactors for constant loading conditions.

5. DENITRIFICATION POTENTIAL

From the empirical denitrification equation of the primary and secondary anoxic reactors, Eqs. 6.15 and 6.9, it is clear that there is a maximum concentration of nitrate that can be removed by the anoxic reactors. In terms of these equations, this maximum for the primary anoxic reactor depends on the influent biodegradable COD, the active mass concentration and the retention time, and for the secondary anoxic reactor only on the active mass concentration and the retention time. The maximum concentration of nitrate that can be removed by an anoxic reactor is known as the *denitrification potential* of the reactor.

5.1 Denitrification potential of the primary anoxic reactor

From Eq. 6.15, the denitrification potential of the primary anoxic reactor, D_{p1} is given by

$$D_{p1} = \alpha S_{bi} + K_2 X_a R_{np} \qquad (6.18)$$

Now $X_aR_{np} = X_aV_a/Q$ (where V_a = volume of the primary anoxic reactor and Q = daily average influent flow). The term X_aV_a represents the mass of sludge in the primary anoxic reactor, which can be expressed as a fraction of the total mass of sludge in the process. Let f_{x1} be the mass of sludge in the primary anoxic reactor as a fraction of the total mass of sludge in the process i.e. the primary anoxic sludge mass fraction, then from Eq. 4.10

$$X_aV_a = f_{x1}\,M(X_a)$$

$$= f_{x1}\,Y_hR_sQS_{bi}/(1+b_hR_s)$$

Dividing though by Q

$$X_aV_a/Q = f_{x1}\,Y_hR_sS_{bi}/(1+b_hR_s) \qquad (6.19)$$

Substituting the RHS of Eq. 6.19 for X_aR_{np} into Eq 6.18 yields

$$D_{p1} = S_{bi}\,\{\alpha + K_2f_{x1}Y_hR_s/(1+b_hR_s)\} \qquad (6.20)$$

where

D_{p1} = denitrification potential of the primary anoxic reactor (mgN/ℓ influent)
S_{bi} = biodegradable COD concentration of the influent
α = fraction of nitrate removed by the initial rapid phase of denitrification
= $f_{bs}\,(1-f_{cv}Y_h)/2{,}86$ (see Section 4.1.2 above)

where

f_{bs} = readily biodegradable fraction of the influent biodegradable COD
f_{x1} = primary anoxic sludge mass fraction.

In Eq. 6.20, it is assumed that the initial rapid rate of denitrification is always complete, i.e. the actual retention time in the primary anoxic reactor is always longer than the time required to utilize all the readily biodegradable COD in the influent, given by t_1 in Eq. 6.16. If t_1 is taken to be the minimum actual hydraulic retention time, it can be shown by substituting Eq. 4.10

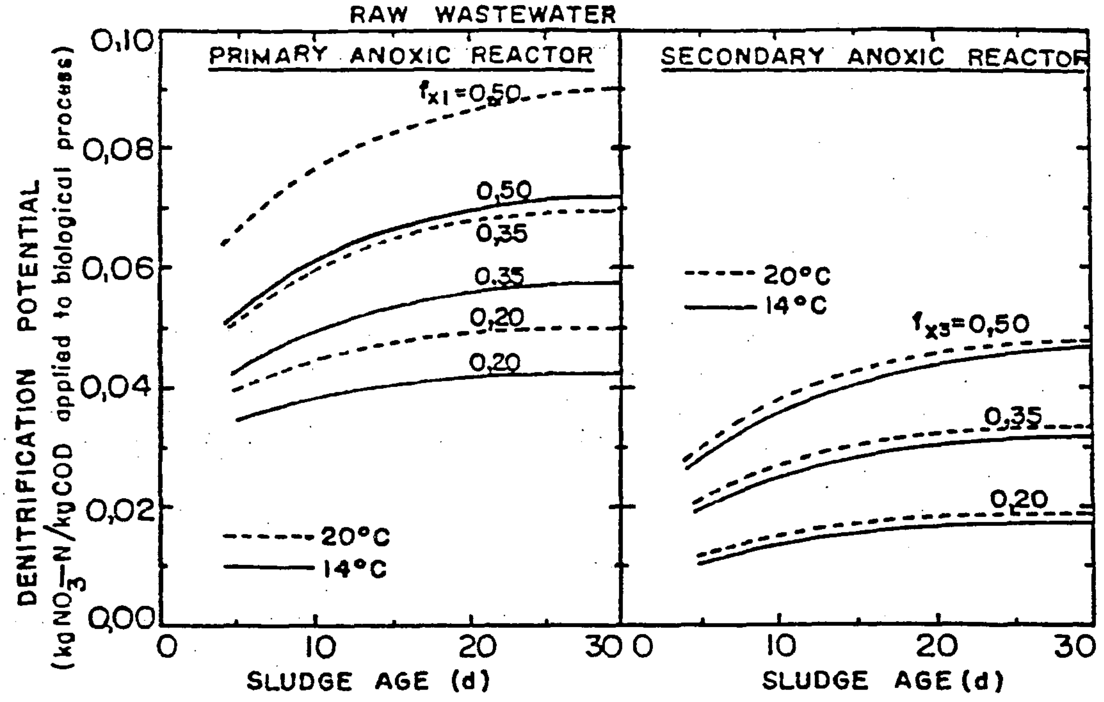

FIGURE 6.13: Denitrification potential per kgCOD applied to biological process versus sludge age for primary (left) and secondary (right) anoxic reactors of different unaerated sludge mass fractions ($f_{x1}$and f_{x3}) for raw wastewater at 14°C and 20°C.

for X_a and Eq. 6.16 for t_1, that the minimum primary anoxic sludge mass fraction f_{x1min} to deplete the readily biodegradable COD is

$$f_{x1min} = \frac{f_{bs}(1 - f_{cv}Y_h)(1 + b_hR_s)}{2{,}86\ K_1Y_hR_s} \qquad (6.21)$$

Substituting the values of the kinetic constants into Eq. 6.21, $f_{xpmin} < 0{,}08$ for $R_s > 15$ days at 14°C. This value is much lower than most practical primary anoxic reactors so that Eq. 6.20 will be valid in most cases.

A diagrammatic representation of Eq. 6.20 is given in Figure 6.13 for different primary anoxic sludge mass fractions (f_{x1}) and temperatures. The diagram shows that the denitrification potential of a primary anoxic reactor increases as sludge age, anoxic sludge mass fraction and temperature increase, but that the denitrification potential is most sensitive to the anoxic sludge mass fraction (f_{x1}).

5.2 Denitrification potential of the secondary anoxic reactor

The denitrification potential of the secondary anoxic reactor is found in a similar fashion as that for the primary anoxic reactor and is found to be given by

$$D_{p3} = S_{bi}\, f_{x3}\, K_3\, Y_h\, R_s/(1 + b_{hT}R_s) \qquad (6.22)$$

where

D_{p3} = denitrification potential of the secondary anoxic reactor (mgN/l influent)

f_{x3} = anoxic sludge mass fraction of the secondary anoxic reactor.

A diagrammatic representation of Eq. 6.22 is given in Figure 6.13 for different anoxic sludge mass fractions and temperatures. The denitrification potential increases as the sludge ages, temperature and anoxic sludge mass fraction (f_{x3}) increase, but is most sensitive to the anoxic sludge mass fraction (f_{x3}).

6. DESIGN PROCEDURE FOR N REMOVAL PROCESSES

In the preceding sections, four important parameters were defined and discussed. These are (1) maximum unaerated sludge mass fraction (f_{xm}, Chapter 5, Section 4.4), (2) nitrification capacity (N_c, Chapter 5, Section 5.3), (3) denitrification potential (Section 5 above) and the TKN/COD ratio (Chapter 3, Section 3,3). These parameters greatly facilitate the design of a nitrogen removal process. The salient aspects of these four parameters are briefly set out below.

(i) Maximum unaerated sludge mass fraction

As nitrification is a prerequisite for denitrification the fraction of the sludge mass under aerobic conditions must be sufficiently large to allow complete nitrification. At a particular sludge age, this sets a limit on the fraction of sludge that is not aerated i.e. sets a maximum unaerated sludge mass fraction (f_{xm}). This maximum unaerated sludge mass fraction is dependent principally on the maximum specific growth rate of the nitrifiers at 20°C (μ_{nm20})

attainable with the wastewater, the minimum average temperature of the wastewater, (T_{min}) and the sludge age of the process (R_s). In order to ensure nitrification, it is recommended that f_{xm} is limited at 0,50 to 0,60 and that the minimum aerobic sludge mass fraction is increased by a factor of safety of 1,25 to 1,35 (see Chapter 5, Section 4.4).

(ii) Nitrification capacity of process

The nitrification capacity is the concentration of nitrate per unit of influent flow that the process produces by nitrification of the influent TKN concentration. If the unaerated sludge mass fraction satisfies the conditions set out above, complete nitrification is likely to take place and the effluent TKN concentration will be between 2 and 3 mgN/ℓ. Under these conditions the nitrification capacity is independent of the process configuration and is the difference between the influent TKN concentration and the sum of the nitrogen required for sludge production and effluent TKN concentration (Chapter 5, Section 5.3).

(iii) Denitrification potential of process

The denitrification potential of a process is the maximum mass of nitrate per unit influent flow that the process as designed can denitrify. It is directly proportional to the influent biodegradable COD concentration and the position and sizes of the anoxic reactors because denitrification in the primary anoxic reactor is faster than in the secondary anoxic reactor.

(iv) The influent TKN/COD ratio

Because the nitrification capacity and denitrification potential are approximately proportional to the influent TKN and COD concentrations respectively, the influent TKN/COD ratio is a relative measure of the mass of nitrate generated in the process by nitrification and the mass of nitrate that can be removed in the process by denitrification.

6.1 Principles of the design procedure

From the wastewater characteristics i.e. influent TKN and COD concentrations (N_{ti} and S_{ti}), maximum specific growth rate of the nitrifiers at 20°C (μ_{nm20}), the readily biodegradable COD fraction (f_{bs}) and the minimum average wastewater temperature (T_{min}), the maximum unaerated sludge mass fraction (f_{xm}) and the nitrification capacity (N_c) can be calculated for a selected sludge age (R_s). The f_{xm} then can be subdivided into primary and secondary anoxic sludge mass fractions (f_{x1} and f_{x3}) and this division fixes the denitrification potential of these two reactors and hence also of the process. When considering the subdivision of f_{xm}, it should be remembered that the denitrification potential of the primary anoxic reactor (D_{p1}) is greater than that of the secondary anoxic reactor (D_{p3}) for equal anoxic sludge mass fractions, but that sufficient nitrate must be recycled to the primary reactor to fully utilize its denitrification potential.

The relative sizes of the primary and secondary anoxic sludge mass fractions (f_{x1} and f_{x3}), as well as whether or not a secondary anoxic reactor is to be included in the process configuration, depend on the influent TKN/COD ratio. If the objective is to remove as much nitrogen as possible, the denitrification potential of the process (which may or may not include a secondary anoxic reactor) must be as close as possible to the nitrification capacity. Complete denitrification is possible only when the process configuration incorporates a secondary anoxic reactor and if the sum of the denitrification potentials of the primary and secondary anoxic reactors is greater than the nitrification capacity, i.e. $D_{p1} + D_{p3} > N_c$ (provided D_{p1} is fully utilized by having a sufficiently high mixed liquor a-recycle).

If complete denitrification is not possible due to a high TKN/COD ratio, nitrate will be present in the effluent. If the effluent nitrate concentration is greater than about 5 to 6 mgN/ℓ, the incorporation of a secondary anoxic reactor becomes an inefficient utilization of the unaerated sludge mass fraction; an improved nitrate removal will be obtained if the secondary anoxic reactor (f_{x3}) is incorporated into the primary anoxic reactor (f_{x1}). Removal of the secondary anoxic reactor from the Bardenpho process (Figures 6.5) reduces the configuration to a Modified Ludzack-Ettinger (MLE) process (Figure 6.4).

The above discussion indicates that in order to select a nitrogen removal process configuration, it is necesary to check whether or not complete denitrification can be achieved.

6.2 Can complete denitrification be achieved?

In the nitrogen removal processes, the maximum anoxic sludge mass fraction available for denitrification, f_{xdm}, is equal to the maximum unaerated sludge mass fraction f_{xm}, i.e.

$$f_{xdm} = f_{xm} \tag{6.23}$$

where f_{xm} is given by Eq. 5.24 for selected R_s, μ_{nmT} and T_{min}.

To check whether or not complete denitrification can be obtained with the maximum anoxic sludge mass fraction, the following reasoning is applied: In the Bardenpho process (which has the potential for complete denitrification), for a fixed underflow s-recycle ratio,* the mixed liquor a-recycle ratio governs the distribution of the nitrate from the aerobic reactor between the primary and secondary anoxic reactors – the higher the a-recycle ratio with respect to the s-recycle ratio, the greater the proportion of nitrate recycled to the primary anoxic reactor. For the selected a- and s-recycle ratio, the best denitrification performance is obtained when the primary anoxic reactor is just loaded to its denitrification potential. If the primary anoxic reactor is loaded to its denitrification potential, the nitrate concentration in the outflow of the reactor is zero and

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*Recycle ratios specified with respect to the daily average influent flow rate.

the nitrate concentration in the aerobic reactor is given by $N_c/(a+s+1)$ i.e. the nitrification capacity diluted into the total flow entering the aerobic reactor. Knowing the nitrate concentration in the aerobic reactor and including the effect of the dissolved oxygen concentration (DO) in the aerobic reactor and underflow s-recycle, the equivalent nitrate load imposed on the primary and secondary anoxic reactors by the selected a- and s-recycles can be calculated. Now a useful analytical device is to increase the equivalent nitrate load on the secondary anoxic reactor by a factor K_{2T}/K_{3T} so that the reactor acts like a primary anoxic reactor except that it receives no readily biodegradable COD. Now adding the equivalent nitrate load on the primary anoxic reactor to the adjusted equivalent nitrate load on the secondary anoxic reactor, a total equivalent nitrate load is obtained. Because the secondary anoxic reactor has been transformed to act like a primary anoxic reactor, their respective contributions to total denitrification potential of the process need not be known, and the maximum anoxic sludge mass fraction (f_{xdm}) can be assumed to be in the form of a primary anoxic reactor. Consequently the process denitrification potential D_{pp} can be calculated by means of Eq 6.20 with $f_{x1}=f_{xdm}$. Now if the process denitrification potential D_{pp} of the maximum anoxic sludge mass fraction f_{xdm}, is greater or equal to the total equivalent nitrate load, complete denitrification is possible for the selected a- and s-recycle ratios. In contrast, if the D_{pp} of f_{xdm} is less than the total equivalent nitrate load, complete denitrification is not possible for the selected a- and s-recycle ratios, and nitrate will be present in the effluent.

Following the above reasoning and assuming the DO concentrations in the a- and s-recycles are O_a and O_s mgO/ℓ respectively, the following equation for the effluent nitrate concentration N_{ne} for the Bardenpho process can be developed:

$$N_{ne} = \frac{\{\frac{N_c}{a+s+1}+\frac{O_a}{2{,}86}\}\{a+\frac{K_{2T}}{K_{3T}}(s+1)\}+\frac{s.O_s}{2{,}86}-D_{pp}}{\{\frac{K_{2T}}{K_{3T}}+s(\frac{K_{2T}}{K_{3T}}-1)\}} \tag{6.24}$$

where

N_{ne} = effluent nitrate concentration (mgN/ℓ)
N_c = nitrification capacity (mgN/ℓ)
a,s = mixed liquor and sludge underflow recycle ratios repsectively
K_{2T}, K_{3T} = second denitrification rate in primary anoxic reactor and denitrification rate in secondary anoxic reactor respectively (mgN/mgVASS/d)
O_a, O_s = dissolved oxygen concentration in mixed liquor a and underflow s recycle respectively (mgO/ℓ)
D_{pp} = denitrification potential at temperature T°C on the assumption that the maximum anoxic sludge mass fraction f_{xdm} found from Eq 6.23 is all in the form of a primary anoxic reactor i.e. D_{pp} is given by Eq. 6.20 with $f_{x1}=f_{xdm}$ and K_2 adjusted to T°C with Eq. 6.13. (6.25)

Also, the optimum primary and secondary anoxic sludge mass fractions f_{x1} and f_{x3}, and the total f_{xdt} are given by

$$f_{x1} = \frac{\{(\frac{N_c}{a+s+1}+\frac{O_a}{2{,}86})a+(N_{ne}+\frac{O_s}{2{,}86})s-\frac{f_{bs}(1-f_{cv}Y_h)\,S_{bi}}{2{,}86}\}}{\{S_{bi}\,\frac{Y_hR_s}{(1+b_{hT}R_s)}\,K_{2T}\}} \tag{6.26}$$

$$f_{x3} = \frac{\{(s+1)\left[\frac{N_c}{a+s+1}+\frac{O_a}{2{,}86}-N_{ne}\right]\frac{K_{2T}}{K_{3T}}\}}{\{S_{bi}\,\frac{Y_hR_s}{(1+b_{hT}R_s)}\,K_{2T}\}} \tag{6.27}$$

$$f_{xdt} = f_{x1}+f_{x3} \tag{6.28}$$

where

f_{x1}, f_{x3}, f_{xdt} = primary, secondary and total anoxic sludge mass fractions respectively.

For a selected sludge age (R_s) and wastewater characteristics (S_{ti}, N_{ti}, f_{bs}, μ_{nm20}, T_{min}, f_{up}, f_{us}) – from which f_{xm} (Eq. 5.24), f_{xdm} (Eq. 6.23), K_{2T}/K_{3T} (Eqs. 6.13 and 6.10), N_c (Eq. 5.29) and D_{pp} (Eq. 6.25) can be calculated – and selected DO concentrations in the a-and s-recycle ratios (O_a and O_s), the only unknowns in Eqs. 6.24 to 6.28 are the a- and s-recycle ratios. The s-recycle ratio usually is specified (at say 1) to obtain satisfactory settling tank performance. Hence N_{ne}, f_{x1} and f_{x3} all depend on the a-recycle ratio only, and hence these three parameters can be calculated for various choices of the a-recycle ratio. If $N_{ne} \leqq 0{,}0$ mgN/ℓ for some specified s- and selected a-recycle ratios, then complete denitrification is possible at these recycle ratios.

In the solution procedure, if Eq. 6.24 yields $N_{ne}<0$, N_{ne} must be set equal to zero before substitution into Eqs 6.26 and 6.27. When N_{ne} has to be set equal to zero, it will be noticed that f_{xdt} obtained from Eq. 6.28 is less than the f_{xdm} assumed to calculate D_{pp}. The reason for this is that when $N_{ne} < 0$, $D_{pp} > N_c$. By inserting $N_{ne}=0$ in Eqs. 6.26 and 6.27, these equations give f_{x1} and f_{x3} that will *just* produce complete denitrification for the selected a- and specified s-recycle ratios, so that f_{xdt} turns out to be less than f_{xdm}.

For the specified s-recycle ratio, Eqs. 6.24 to 6.28 are valid only for a-recycle ratios falling between a lower and an upper limit:

(i) *Lower limit:* the a-recycle must not be lower than that which gives f_{x1} as calculated from Eq. 6.26 less than f_{x1min} as calculated from Eq 6.21 – if $f_{x1} < f_{x1min}$, the readily biodegradable COD is not completely utilized in the primary anoxic reactor leading to inefficient denitrification and also making Eq. 6.20 for D_{p1} invalid.

(ii) *Upper limit:* the a-recycle ratio must not be increased above that which gives f_{x3} from Eq. 6.27 a smaller value than that required to remove only the DO discharged to the secondary anoxic reactor, i.e. $D_{p3} \geqq (1+s)\,O_a/2{,}86$ where D_{p3} is given by Eq. 6.22. Hence the minimum secondary anoxic sludge mass fraction is given by

$$f_{x3min} = \frac{(1+s)\,O_a\,(1+b_{hT}R_s)}{2{,}86\,S_{bi}\,Y_h R_s\,K_{3T}} \qquad (6.29)$$

Only the a-recycle ratios falling between the lower and upper limits specified above are valid a-recycle ratios.

By analysing the process with a specified s-recycle ratio for increasing values of valid a-recycle ratios, then, if $N_{ne} > 0$, N_{ne} and the optimal subdivision of f_{xdm} into f_{x1} and f_{x3}, or, if $N_{ne} \leq 0$, the total anoxic sludge mass fraction to just obtain $N_{ne} = 0$, f_{xdt}, and its optimal subdivision into f_{x1} and f_{x3} can be plotted versus the a-recycle ratio.

For design purposes where the process is to operate between temperatures T_{min} and T_{max}, the following sequence of calculations needs to be made to check whether complete denitrification can be achieved and if so, to select the optimal process configuration:

Step 1: Select wastewater characteristics; S_{ti}, N_{ti}, f_{bs}, f_{up}, f_{us}, μ_{nm20}, T_{max}, T_{min}.

Step 2: Select R_s and S_f.

Step 3: Calculate f_{xm} for T_{min} from Eq. 5.24.

Step 4: With f_{xm} and R_s calculate S_f for T_{max} from Eq. 5.24.

Step 5: Estimate N_{te} for T_{max} and T_{min}.

Step 6: With f_{up} and f_{us}, calculate N_s for T_{max} and T_{min} for selected R_s from Eq. 4.23.

Step 7: Calculate N_c from Eq. 5.29.

Step 8: Calculate D_{pp} from Eq. 6.25 for T_{max} and for T_{min} with f_{xdm} given by Eq. 6.23.

Step 9: Select s, O_a and O_s.

Step 10: Calculate f_{x1min} from Eq. 6.21 for T_{max} and for T_{min}.

Step 11: Calculate f_{x3min} from Eq. 6.29 for T_{max} and for T_{min}.

Step 12: Select the a-recycle ratio and calculate N_{ne} for T_{min} from Eq. 6.24.

Step 13: If $N_{ne} < 0$, set $N_{ne} = 0$.

Step 14: With N_{ne} calculate f_{x1}, f_{x3} and f_{xdt} from Eqs. 6.26 to 6.28.

Step 15: Check that $f_{x1} \geq f_{x1min}$ and $f_{x3} \geq f_{x3min}$. If not, discard selected a-recycle ratio as invalid.

Step 16: Repeat steps 12 to 15 for different a-recycle ratios.

Step 17: Repeat steps 12 to 16 for T_{max}.

The analysis as to whether or not complete denitrification can be achieved must be undertaken at the *lowest expected temperature (T_{min})*, because the denitrification potential D_{pp} decreases with decrease in temperature. If complete denitrification can be obtained at T_{min} (i.e. $N_{ne} < 0$ from Eq. 6.24 for some valid a-recycle ratio at the specified s-recycle ratio), then f_{x1} is fixed at the *highest expected* temperature (T_{max}). In repeating the calculation at T_{max} it will be found that f_{xdt} at T_{max} is less than f_{xdt} at T_{min}, and for a certain a-recycle ratio, there will be a minimum f_{xdt}*. The f_{x1} value corresponding to the minimum f_{xdt} at T_{max} fixes the primary anoxic sludge mass fraction for both T_{max} and T_{min}. Now f_{x3} is fixed at the *minimum expected* temperature (T_{min}) at the f_{x3} value corresponding to the fixed f_{x1} value and the total anoxic sludge mass fraction f_{xdt} for both T_{max} and T_{min} is given by $f_{x1} + f_{x3}$. The f_{x3} is fixed at T_{min} because if it were fixed at T_{max}, then f_{xdt} will be insufficient to achieve complete denitrification at T_{min} because for fixed f_{x1}, f_{xdt} at T_{max} is less than f_{xdt} at T_{min}. With f_{x1} and f_{x3} fixed for both T_{max} and T_{min}, the optimum performance of the process is achieved when it is operated at the a-recycle ratio corresponding to the fixed f_{x1} value, this a-recycle ratio being the optimum a_o. It should be noted that with f_{x1} and f_{x3} fixed,

(i) a_o decreases with decrease in temperature because D_{p1} decreases with decrease in temperature, and

(ii) the use of any a-recycle ratio other than a_o results in a poorer denitrification performance; if $a < a_o$, D_{p1} is not fully utilized and if $a > a_o$, unnecessarily large quantities of DO are discharged to the primary anoxic reactor.

In the design of a Bardenpho process for a selected sludge age, if it is found that the total anoxic sludge mass fraction for complete denitrification f_{xdt} is less than the maximum allowed f_{xdm}, the following options are open:

(1) f_{x1} and f_{x3} can be increased to give an f_{xdt} equal to f_{xdm}. This introduces a factor of safety in denitrification by making an allowance for a reduction in the denitrification rates K at 20°C, a reduction in the influent readily biodegradable COD fraction f_{bs} or minimum temperature (T_{min}) or an increase in influent TKN/COD ratio.

(2) The sludge age (R_s) can be reduced so that f_{xdm} becomes equal to f_{xdt}. The lower R_s will allow a saving in process volume (see Figure 4.3). The lower R_s can be estimated as follows: Set f_{xdm} equal to the f_{xdt} required for complete denitrification at T_{min}; determine f_{xm} from Eq. 6.23; with f_{xm} calculate R_s from Eq. 5.24; to check, repeat step by step procedure with the new R_s.

Option (2) is not as good an alternative as option (1) because it exludes a factor of safety on the nitrogen

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*If T_{max} is less than about 18°C, a clear minimum cannot be discerned (see Figure 6.15 for 14°C). In this event any reasonable f_{x1} can be selected with the proviso that complete utilization of the readily biodegradable COD can be *assured* even with a significant decrease in K_{120} (see Eqs. 6.12 and 6.21). The existence of a minimum, and the temperatures above which it becomes discernible, depends much on the choice of the DO concentration in the recycles O_a and O_s.

removal. However, exclusion of a factor of safety on denitrification is not critical for N removal in the Bardenpho process because it achieves efficient N removal with respect to the other N removal process (i.e. the MLE process) up to an effluent nitrate concentration of about 5-6 mgN/ℓ. However, the exclusion of a factor of safety on denitrification for the Phoredox process may have serious consequences because the discharge of any nitrate to the anaerobic reactor severely reduces the P removal obtainable in the process. The effect of denitrification efficiency on excess biological P removal is discussed in detail in Chapter 7.

6.3 When complete denitrification cannot be achieved

In the Bardenpho process, if it is found that complete denitrification cannot be obtained, (i.e. $N_{ne} > 0$ from Eq. 6.24 for all valid a-recycle ratios at the specified s-recycle ratios) nitrate will be present in the effluent and f_{xdt} will be equal to f_{xdm}. Generally, if the objective is N removal only, the Bardenpho process will be appropriate only if N_{ne} is less than 5 to 7 mgN/ℓ; when $N_{ne} >$ 5 to 7 mgN/ℓ, – which will usually be the case if the TKN/COD ratio > 0,10 mgN/mgCOD – the MLE process (Figure 6.4) will yield a better nitrogen removal efficiency for the particular wastewater.

For the selected sludge age (R_s), factor of safety on nitrification (S_f) and wastewater characteristics (S_{ti}, N_{ti}, f_{bs}, T_{min}, T_{max}, f_{up}, f_{us}), the f_{xm} (Eq. 5.24), N_c (Eq. 5.29), f_{xdm} (Eq. 6.23) and D_{pp} (Eq. 6.25 with $f_{x1} = f_{xdm}$) for the MLE process are equal to those of the Bardenpho process and are found by the identical procedure (i.e. steps 1 to 8 given above). For the MLE process, the best denitrification performance is obtained when the anoxic reactor is just loaded with nitrate to its denitrification potential. For a specified s-recycle ratio, the mixed liquor a-recycle ratio which loads the anoxic reactor to its denitrification potential is the optimum a-recycle ratio, a_o, and will yield the minimum effluent nitrate concentration. When the nitrate load imposed on the anoxic reactor by the a- and s-recycles (including DO concentrations in the a- and s-recycles of O_a and O_s mgO/ℓ respectively), is less than or equal to the denitrification potential (D_{pp}), the nitrate concentration in the outflow of the anoxic reactor is zero and that in the aerobic reactor and effluent is $N_c/(a+s+1)$ i.e. the nitrification capacity "diluted" into the total flow entering the aerobic reactor. Equating the nitrate load on the anoxic reactor imposed by the a- and s-recycles (including the DO in the recycles) to the denitrification potential D_{pp} and solving for the a-recycle ratio yields a_o;

$$a_o = \{-B + \sqrt{(B^2 + 4AC)}\}/(2A) \qquad (6.30)$$

where

$$A = O_a/2{,}86$$

$$B = N_c - D_{pp} + \{(s+1)O_a + s.O_s\}/2{,}86$$

$$C = (s+1)(D_{pp} - sO_s/2{,}86) - sN_c$$

The effluent nitrate concentration N_{ne} for any a-recycle ratio less than or equal to a_o is given by Eq. 6.31 but will be a minimum when $a = a_o$, i.e.

$$N_{ne} = N_c/(a+s+1) \text{ mgN}/\ell \qquad (6.31)$$

In the design of an MLE process, it will be found the lower N_c is with respect to D_{pp} (i.e. the lower the influent TKN/COD ratio) the lower the N_{ne} and the higher the a_o. In practice, to operate an MLE process with an a_o-recycle ratio greater than 6:1 will be uneconomical: the increase in nitrate removal by increasing a_o from 6 to 10 is only 5% and, therefore, will not be cost-effective with respect to the increased pumping costs. Hence, the N removal of the MLE process is restricted by a maximum a-recycle ratio of say 6, which results in a minimum N_{ne} of 5 to 7 mgN/ℓ. This restriction on the MLE process will not be a problem in practice because high values for a_o are only obtained for wastewaters with low TKN/COD ratios (< 0,10 mgN/mgCOD) and these wastewaters are more effectively treated in a Bardenpho process (see section above). Alternatively, if near-complete or complete N removal is not essential, the sludge age can be reduced; for a fixed TKN/COD ratio this reduces D_{pp} (because f_{xm} is reduced) with respect to N_c and hence reduces a_o. The optimum R_s for the fixed TKN/COD ratio can be found using the procedure set out above with different choices of R_s or, alternatively, the method of van Haandel, Dold and Marais (1982) can be used.

The design procedure for N removal processes set out above will be demonstrated in the next section with the aid of worked examples.

7. DESIGN EXAMPLES

The design of biological N removal plant by means of the procedure set out above is demonstrated by continuing the numerical example given in Chapter 4, Section 12 (in which the design procedure for organic material removal was demonstrated) and Chapter 5, Section 6 (in which the behaviour of nitrification was demonstrated).

7.1 Variability of wastewater characteristics

7.1.1 Variability of nitrification rate

The nitrification rate, as expressed by the maximum specific growth rate of the nitrifiers at 20°C (μ_{nm20}) is strongly dependent on the nature and source of the wastewater, so much so that it is considered a wastewater characteristic. The value can range from 0,65/d in purely domestic wastewaters down to 0,20/d for wastewaters with a large industrial* contribution. Owing to uncertainty in the μ_{nm20} value for a particular wastewater, in the absence of any data on the μ_{nm20} value, a low value, say between 0,3 to 0,4/d, needs to be selected to provide reasonable assurance that a safe design with respect to nitrification will be obtained (see Chapter 5, Section 4.1). In this design example, a value of 0,36/d is selected to demonstrate the situation in a

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*The type of industry referred to here is that which produces effluents which inhibit biological activity, e.g. metal finishing industry, plating works, paint production plants etc.

design with a relatively low μ_{nm20} value (see Tables 6.1 and 6.2).

7.1.2 Variability in denitrification rates

The denitrification rates K at 20°C are relatively insensitive to the nature and source of the wastewater. However, different values have been observed with different wastewaters – in instances where the wastewater contains an appreciable industrial* contribution, reduced denitrification rates may be encountered. The denitrification rates used in this design example are the mean values found in the experimental work at the University of Cape Town, i.e. $K_{120} = 0{,}720$, $K_{220} = 0{,}101$ and $K_{320} = 0{,}072$ mgNO$_3$-N/(mgVASS.d) (see Table 6.2). These values were found on Strandfontein wastewater, which is an approximately average wastewater with a relative small industrial contribution. For most normal municipal wastewaters encountered in South Africa the denitrification rates given above are likely to result in a sufficiently accurate estimate of the denitrification potential of the process. Where there is doubt regarding the applicability of these rates, it is recommended that they are reduced by a factor of safety on denitrification, i.e. by about 1,1 to 1,2 times. The magnitude of the factor of safety should reflect the uncertainty in the applicability of the average denitrification rates given in this monograph.

7.1.3 Variability in readily biodegradable COD fraction

The readily biodegradable COD fraction of a wastewater can vary considerably depending on the composition and origin of the various contributions that make up the wastewater. It has been found that in approximately normal principally domestic raw wastewater, approximately 20% of the total (or 25% of the biodegradable) COD is readily biodegradable. However, where industrial wastewaters containing a high biodegradable organic material content are discharged to the municipal sewerage system, significantly higher readily biodegradable COD fractions can be encountered. For example, yeast factories, breweries, malting processes and fruit and vegetable juicing and canning industries produce wastewaters with high readily biodegradable COD fractions. Municipal wastewaters containing con-

TABLE 6.1 RAW AND SETTLED WASTEWATER CHARACTERISTICS IMPORTANT FOR DENITRIFICATION (FOR OTHER CHARACTERISTICS SEE TABLES 4.3 AND 5.2

Parameter	Symbol	Value Raw	 Settled	Units
Maximum specific growth rate of nitrifiers at 20°C	μ_{nm20}	0,36	0,36	/d
Denitrification rates at 20°C	K_1	0,720	0,720	mgNO$_3$-N/mgVASS/d.
	K_2	0,101	0,101	
	K_3	0,072	0,072	
Readily biodegradable COD fraction	f_{bs}	0,24	0,33*	

*Note increase on f_{bs} fraction in settled wastewater (see Section 7.1.3) – based on a 40% total COD decrease and a 10% readily biodegradable COD decrease in the primary settling tanks. (See Chapter 4, Section 12.1 and Table 4.3).

TABLE 6.2 TEMPERATURE ADJUSTMENT OF THE DENITRIFICATION RATES K_1, K_2 and K_3

Constant	Symbol	20°C	θ	14°C	22°C	Eq. No.
1st Denit. rate in Prim. Anoxic	K_{120}*	0,720	1,200	0,241	1,036	6.12
2nd Denit. rate in Prim. Anoxic	K_{220}*	0,101	1,080	0,0636	0,1178	6.13
Denit. rate in Second. Anoxic	K_{320}*	0,072	1,029	0,0607	0,0762	6.10
Max. Spec. growth rate of Nitrifiers	μ_{nm20}**	0,36	1,123	0,18	0,45	5.15
Endog. Resp. rate for Nitrifiers	b_{n20}**	0,04	1,029	0,034	0,042	5.16

*Units: mgNO$_3$-N/(mgVASS.d)
**Units: per day

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*Industries producing biodegradable wastes e.g. malting factories, breweries, fruit and vegetable processing factories, etc. do not fall in this group as these may *promote* biological N and P removal. (see Section 7.1.3).

tributions from these sources can have relatively high readily biodegradable COD fractions depending on the magnitude of the industrial discharges.

Many factors can reduce the readily biodegradable COD fraction of a wastewater. For example the discharge of waste biological sludge to the sewerage system for sludge handling at treatment works lower down the sewer line can deplete a wastewater of all the readily biodegradable COD, particularly if the combined flow is lifted by Archimedian screw pumps. Further examples are cited in Chapter 7, Section 9.

The above examples demonstrate that the readily biodegradable COD fraction can differ significantly between different wastewaters. Consequently, it is recommended that whenever possible, the fraction should be measured. The importance of this fraction in biological nitrogen (and phosphorus) removal is such that it is always worth the effort of measuring it.*

It should be noted that the readily biodegradable COD fraction (as defined by Eqs. 2.8 and 2.9) is affected by primary sedimentation. This is because primary sedimentation removes a large fraction of the particulate COD but very little soluble COD. Consequently, the readily biodegradable COD concentrations (S_{bsi}) are approximately the same before and after primary sedimentation but, because the total or biodegradable COD (S_{ti}, S_{bi}) has been significantly reduced in the primary settlers, the readily biodegradable COD fraction (f_{ts}, f_{bs}) of the settled wastewater is increased (see Chapter 2, Section 2.1).

For the design example it will be assumed that the readily biodegradable COD fraction with respect to the biodegradable COD f_{bs} for the raw wastewater is 0,24, i.e. the value found for approximately normal principally domestic raw wastewaters. The readily biodegradable COD fraction (f_{bs}) of the settled wastewater, assuming that primary sedimentation removes 40% of the total COD but only 10% of the readily biodegradable COD is 0,33. (see Table 6.1, Chapter 4, Section 12.1 and Chapter 2, Section 2).

7.2 Effect of temperature

From the kinetics of denitrification discussed above, it is clear that the denitrification rates K_1, K_2 and K_3 all decrease with decreasing temperature. This, combined with the fact that the maximum specific growth rate of the nitrifiers at 20°C (μ_{nm20}) also decreases with decreasing temperature has the result that the critical condition for the nitrogen removal process is the average *minimum* wastewater temperature. The design completed for the average minimum wastewater temperature (T_{min}) is checked at the average maximum wastewater temperature (T_{max}) because the oxygen demand per unit COD load increases as the temperature increases.

To facilitate the calculations of the design example at $T_{min} = 14°C$ and $T_{max} = 22°C$, the denitrification rates K_1, K_2 and K_3 are adjusted for temperature in Table 6.2. Temperature adjustment of μ_{nm20} and b_n is given in Table 5.3, but for convenience is repeated in Table 6.2.

7.3 Maximum unaerated sludge mass fraction

This aspect is discussed in detail in Chapter 5, Section 4.4 but it is briefly repeated here because sizing of the maximum unaerated sludge mass fraction is one of the most important decisions in the design of a nutrient removal process.

For nitrogen removal processes, the larger the unaerated sludge mass fraction (f_{xt}) the greater the nitrogen removal and the lower the effluent nitrate concentration. Generally at a minimum temperature of 14°C, to obtain an f_{xt} of 0,50 for favourable μ_{nm20} values (0,40/d), a sludge age of 15 to 20 days is required to ensure nitrification ($S_f = 1,25$) (see Eq. 5.24); for less favourable μ_{nm20} values (between 0,3 and 0,4/d), the required sludge age is 20 to 30 days (see Figure 6.14).

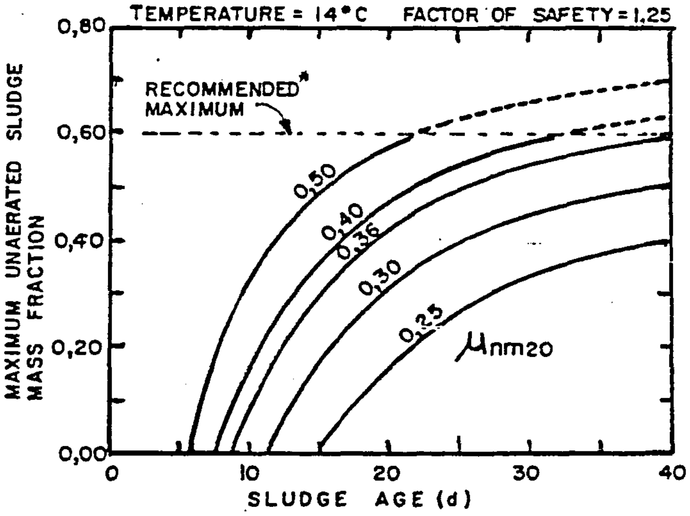

FIGURE 6.14: Maximum unaerated sludge mass fraction versus sludge age for different maximum specific growth rates of the nitrifiers at 20°C (μ_{nm20}) at 14°C and factor of safety (S_f) of 1,25. (From Eq. 5.24).

Now because the volume of the process is directly proportional to the mass of sludge in it (see Chapter 4, Section 6) a 30-day sludge age process will require a process volume about 1/3 larger than a 20 day one. Consequently, for unfavourable μ_{nm20} values it may be preferable to accept a lower f_{xm} and a higher effluent nitrate concentration – a reduction of f_{xm} from 0,50 at $R_s = 30$ d to $f_{xm} = 0,40$ at $R_s = 22$ d results in a 30% saving in process volume and an increase in effluent nitrate concentration of 6 mgN/ℓ (for an influent COD of 600 mg/ℓ and TKN/COD ratio of 0,10). The above argument applies also to favourable μ_{nm20} values because, depending on the effluent quality requirements, it may be acceptable to design a 15-day sludge age process with an effluent nitrate concentration of 5 mgN/ℓ rather than a 20-day process which can achieve complete denitrification (depending on the TKN/COD ratio). Also, for instances where both the

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*The readily biodegradable COD fraction f_{ts} (or f_{bs}) or the actual concentration S_{bsi} cannot be measured directly. Its measurement requires the operation and monitoring of a single completely mixed activated sludge process at a very short sludge age under daily cyclic square wave loading conditions (see Appendix 2).

μ_{nm20} and TKN/COD ratio are unfavourable, a 20-day sludge age process with external energy source addition, to augment denitrification, may be a more economical process than a 35-day process. *Clearly, the decision on the unaerated sludge mass fraction and sludge age of the process should not be taken too lightly, because with these two parameters fixed, the N removal of the process with the influent and self-generated energy sources is fixed, so that with fixed influent wastewater characteristics, the effluent quality will be fixed (see Chapter 3).*

For the wastewater characteristics of the design example i.e. T = 14°C and $\mu_{nm20} = 0{,}36/d$ (see Tables 4.3 and 5.2) and accepting a factor of safety on nitrification $S_f = 1{,}25$, a 25-day sludge age process is required for $f_{xm} = 0{,}50$ (see Chapter 5, Section 4.4 and Figure 6.14). As this sludge age is not excessively long for $f_{xm} = 0{,}50$, a 25-day sludge age process will be accepted for a preliminary design analysis.

7.4 Nitrification

Because a factor of safety of 1,25 was imposed in the design of the unaerated sludge mass fraction and sludge age, the minimum sludge age for nitrification is sufficiently below the actual sludge age to assure that complete nitrification will take place i.e. the effluent TKN concentration can be assumed to be 3 mgN/ℓ. (see Chapter 5, Section 4.4). This can be checked by repeating the nitrification calculations given in Chapter 5 for 25 days sludge age.

For the raw wastewater characteristics (i.e. $f_{up} = 0{,}13$ mgCOD/mgCOD, $f_{us} = 0{,}05$ mgCOD/mgCOD, $T_{min} = 14°C$, $S_{ti} = 600$ mgCOD/ℓ – see Tables 4.3 and 5.2) and 25 days sludge age, and accepting the nitrogen content of the volatile solids f_n to be 0,10 mgN/mgVSS, the nitrogen required for sludge production is given by Eq. 4.23 i.e.

$$N_s = 600.0{,}10\left[\frac{0{,}45\,(1-0{,}05-0{,}13)}{(1+0{,}20.25)}(1+0{,}2.0{,}20.25)+\frac{0{,}13}{1{,}48}\right]$$

$$= 600.0{,}021 = 12{,}6 \text{ mgN}/\ell.$$

The nitrification capacity (N_c) is found from Eq. (5.29) i.e. for the raw wastewater at 14°C

$$N_c = 48-12{,}6-3 = 32{,}4 \text{ mgN}/\ell.$$

The nitrification oxygen demand is found from Eq. 5.39 i.e.

$$M(O_n) = 4{,}57\,.\,N_c\,.\,Q$$

$$= 4{,}57.32{,}4.\,13{,}33.\,10^6 \text{ mgO/d}$$

$$= 1973 \text{ kgO/d}.$$

The above calculations for N_s, N_c and $M(O_n)$, including those for the raw wastewater at 22°C and the settled wastewater at 14°C and 22°C, are tabulated in Table 6.3.

In the design, because it is intended to reduce the nitrate concentration as much as possible, the Alkalinity change in the wastewater will be minimized: Assuming that 80% of the nitrate formed is denitrified, the alkalinity change $\Delta Alk = 7{,}14.N_c - 3{,}57.$ (Nitrate denitrified) $= -7{,}14.32{,}4 + 3{,}57.0{,}80.32{,}4 = -139$ mg/ℓ as $CaCO_3$.

TABLE 6.3 SUMMARY OF DESIGN CALCULATIONS FOR THE BARDENPHO AND MLE PROCESSES TREATING RAW OR SETTLED WASTEWATER AT 25 DAYS SLUDGE AGE. WASTEWATER CHARACTERISTICS GIVEN IN TABLES 4.3, 5.2 AND 6.1

Parameter	Symbol	Units	Raw Wastewater				Settled Wastewater			
			Bardenpho		MLE		Bardenpho		MLE	
Temperature	T	°C	14	22	14	22	14	22	14	22
Safety factor	S_f		1,25	2,7	1,25	2,7	1,25	2,7	1,25	2,7
Max. unaerated sludge mass	f_{xm}		0,50	0,50	0,50	0,50	0,50	0,50	0,50	0,50
Effluent TKN	N_{te}	mgN/ℓ	3,0	2,0	3,0	2,0	3,0	2,0	3,0	2,0
N for sludge production	N_s	mgN/ℓ	12,6	12,0	12,6	12,0	5,7	5,4	5,7	5,4
Nitrification capacity	N_c	mgN/ℓ	32,4	34,0	32,4	34,0	32,3	33,6	32,3	33,6
Nitrification oxygen demand	$M(O_n)$	kgO/d	1973	2071	1973	2071	1961	2046	1961	2046
Max. anoxic sludge mass	f_{xdm}		0,50	0,50	0,50	0,50	0,50	0,50	0,50	0,50
Max. denitrification potential	D_{pp}	mgN/ℓ	43,1	58,7	43,1	58,7	31,1	41,2	31,1	41,2
DO in a-recycle	O_a	mgO/ℓ	2,0	2,0	2,0	2,0	2,0	2,0	2,0	2,0
DO in s-recycle	O_s	mgN/ℓ	1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0
Underflow recycle ratio	s		1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0
Min. Primary anoxic	f_{x1min}		0,062	0,017	–	–	0,085	0,029	–	–
Min. Secondary anoxic	f_{x3min}		0,024	0,023	–	–	0,039	0,036	–	–
Optimum a-recycle ratio	a_o		3,3	5,0	17,2*	36,0*	2,2	5,0	4,7	13,4*
Effluent nitrate	N_{ne}	mgN/ℓ	0,0	0,0	1,7*	0,9*	4,9	1,7	4,8	2,2*
Primary anoxic	f_{x1}		0,16	0,16	0,50	0,50	0,30	0,30	0,50	0,50
Secondary anoxic	f_{x3}		0,24	0,19	–	–	0,20	0,20	–	–
Total anoxic	f_{xdt}		0,40	0,35	0,50	0,50	0,50	0,50	0,50	0,50

*These are the theoretical results from Eqs. 6.30 and 6.31. Owing to an economical upper limit to the mixed liquor a-recycle ratio, these theoretical a_o values should not be implemented in practice. (see text).

With an influent Alk of 200 mg/ℓ as $CaCO_3$ (see Table 5.2) the effluent Alk = 200 − 139 = 61 mg/ℓ as $CaCO_3$, which, from Figure 5.6, will maintain a pH above 7, (see Chapter 5, Section 4.3).

7.5 Denitrification

Accepting a sludge age of 25 days, which allows a maximum unaerated sludge mass fraction f_{xm} of 0,50, the denitrification behaviour of the Bardenpho and MLE processes will be demonstrated for the typical raw and settled wastewaters at 14°C and 22°C.

In the calculations it will be assumed that there is a dissolved oxygen concentration of 2 mgO/ℓ in the interconnecting flows from the aerobic to the anoxic reactors (i.e. $O_a = 2$ mgO/ℓ) and 1 mgO/ℓ in the underflow recycle from the secondary settling tank (i.e. $O_s = 1$ mgO/ℓ). The underflow recycle ratio (s) is assumed to be 1:1.*

For nitrogen removal processes, the maximum anoxic sludge mass fraction (f_{xdm}) is equal to the maximum unaerated sludge mass fraction (f_{xm}) i.e. $f_{xdm} = 0{,}50$ (see Eq. 6.23). The denitrification potential of the primary anoxic reactor at 14°C with $f_{x1} = f_{xdm} = 0{,}50$ for the raw wastewater is found from Eq. 6.20 (see Eq. 6.25)

$$\begin{aligned} D_{pp} &= (1-0{,}05-0{,}13)\ 600\Big\{\frac{0{,}24\ (1-0{,}45.1{,}48)}{2{,}86} + \\ &\qquad 0{,}0636.0{,}50.0{,}45.25/(1+0{,}20.25)\Big\} \\ &= 492(0{,}028+0{,}0596) \\ &= 13{,}8+29{,}3 \\ &= 43{,}1 \text{ mgN}/\ell . \end{aligned}$$

The D_{pp} values for the raw wastewater at 22°C and those for the settled wastewater at 14°C and 22°C are listed in Table 6.3.

The minimum primary and secondary anoxic sludge mass fractions (f_{x1min} and f_{x3min}) are found from Eq. 6.21 and Eq. 6.29 respectively. For example for the raw wastewater at 14°C,

$$\begin{aligned} f_{x1min} &= \frac{0{,}24(1-0{,}45.1{,}48)(1+0{,}20.25)}{2{,}86.0{,}241.0{,}45.25} \\ &= 0{,}062 \quad \text{and} \\ f_{x3min} &= \frac{(1+1)2{,}0(1+0{,}20.25)}{(1-0{,}05-0{,}13)600.0{,}45.25.0{,}0607.2{,}86} \\ &= 0{,}025 \end{aligned}$$

The values for the raw wastewater at 22°C and those for the settled wastewater at 14°C and 22°C are given in Table 6.3.

At this stage of the design procedure, we have collected all the information necessary to check whether or not complete denitrification can be achieved i.e. steps 1 to 11 of the step by step procedure have been completed.

Continuing the procedure with step 12, a value for the a-recycle ratio is selected and the effluent nitrate concentration is calculated from Eq. 6.24. Selecting a = 3,3 then for the *raw wastewater at 14°C*,

$$\begin{aligned} N_{ne} &= \frac{\{\frac{32{,}4}{3{,}3+1+1} + \frac{2}{2{,}86}\}\{3{,}3 + \frac{0{,}0636}{0{,}0607}(1+1)\} + \frac{1.1}{2{,}86} - 43{,}1}{\{\frac{0{,}0636}{0{,}0607} + 1(\frac{0{,}0636}{0{,}0607} - 1)\}} \\ &= -5{,}47 \text{ mgN}/\ell \end{aligned}$$

Because $N_{ne} < 0{,}0$, N_{ne} must be set equal to zero. Setting $N_{ne} = 0$ and substituting into Eqs. 6.26 to 6.28 yields the optimum primary (f_{x1}), secondary (f_{x3}) and total (f_{xdt}) anoxic sludge mass fractions *just* to achieve complete denitrification with an a-recycle ratio of 3,3 i.e.

$$\begin{aligned} f_{x1} &= \frac{\{\frac{32{,}4}{3{,}3+1+1} + \frac{2}{2{,}86}\}\ 3{,}3 + (0 + \frac{1}{2{,}86})1 - \frac{0{,}24.0{,}334.492}{2{,}86}}{492\ \frac{0{,}45.25}{(1+0{,}20.25)}\ 0{,}0636} \\ &= 0{,}16 \end{aligned}$$

$$\begin{aligned} f_{x3} &= \frac{\{(1+1)[\frac{32{,}4}{3{,}3+1+1} + \frac{2}{2{,}86} - 0]\frac{0{,}0636}{0{,}0607}\}}{492\ \frac{0{,}45.25}{1+0{,}20.25}\ 0{,}0636} \\ &= 0{,}24 \end{aligned}$$

$$f_{xdt} = 0{,}40$$

The above calculation is repeated for different values of a. The results of N_{ne}, f_{x1}, f_{x3} and f_{xdt} are shown plotted versus a in Figure 6.15a. Repeating the design procedure *at 22°C* (steps 12 to 15) for the raw wastewater produces the results shown in Figure 6.15b (see also Table 6.3).

Figure 6.15 shows that, *for the raw wastewater*, complete denitrification can be achieved at *14°C* for any a-recycle ratio > 1,8 (Figure 6.15a). At *22°C* (Figure 6.15b), a minimum f_{xdt}** for complete denitrification of 0,35 occurs at a = 5,0. The f_{x1} corresponding to a = 5,0 is 0,16 and this fixes the design value of f_{x1} for both 14°C and 22°C. Now if f_{x3} is fixed at 22°C at the value corresponding to f_{x1} i.e. $f_{x3} = 0{,}19$ (Figure 6.15b), then f_{xdt} will be insufficient to achieve complete denitrification at 14°C; for a fixed f_{x1}, f_{xdt} at 14°C is greater than f_{xdt} at 22°C (compare Figures 6.15a and b). Hence f_{x3} is fixed at 14°C at the value corresponding to $f_{x1} = 0{,}16$ i.e. $f_{x3} = 0{,}24$ (Figure 6.15a). Hence $f_{xdt} = 0{,}40$ for complete denitrification at 14°C and 22°C.

With f_{x1} and f_{x3} fixed for both 14°C and 22°C, the optimum performance of the process is achieved when it is operated at the a-recycle ratio corresponding to the

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*This is usually fixed at a value such that satisfactory settling tank operation is obtained. Secondary settling tank behaviour is described in Chapter 8.

**If T_{max} is less than 18°C, a clear minimum cannot be discerned (see Figure 6.15a at 14°C). In this event, any reasonable f_{x1} can be selected with the proviso that complete utilization of the readily biodegradable COD can be assured even with a significant decrease in K_{120} (see Eqs. 6.12 and 6.21). The existence of a minimum, and the temperature above which it becomes discernible, depends much on the choice of the DO concentrations in the recycles, O_a and O_s.

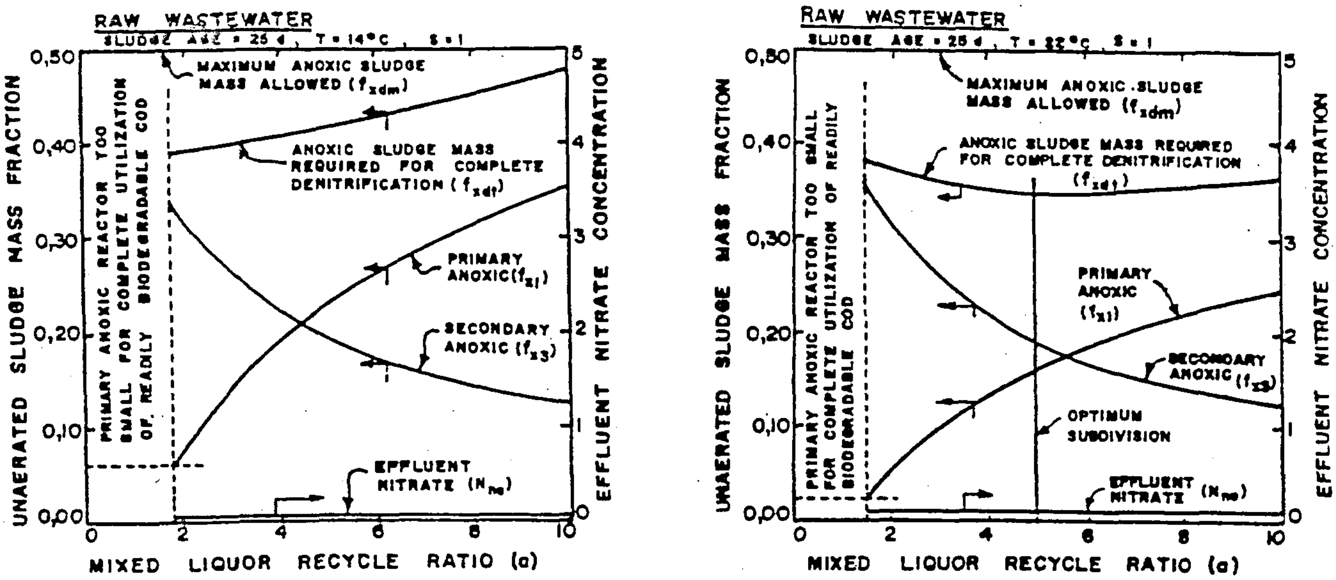

FIGURE 6.15: Primary (f_{x1}), secondary (f_{x3}), total (f_{xdt}) and maximum allowable (f_{xdm}) anoxic sludge mass fractions and effluent nitrate concentrations (N_{ne}) versus mixed liquor a-recycle ratio for the Bardenpho process treating raw wastewater at 25 days sludge age at 14°C (left, Figure 6.15a) and 22°C (right, Figure 6.15b).

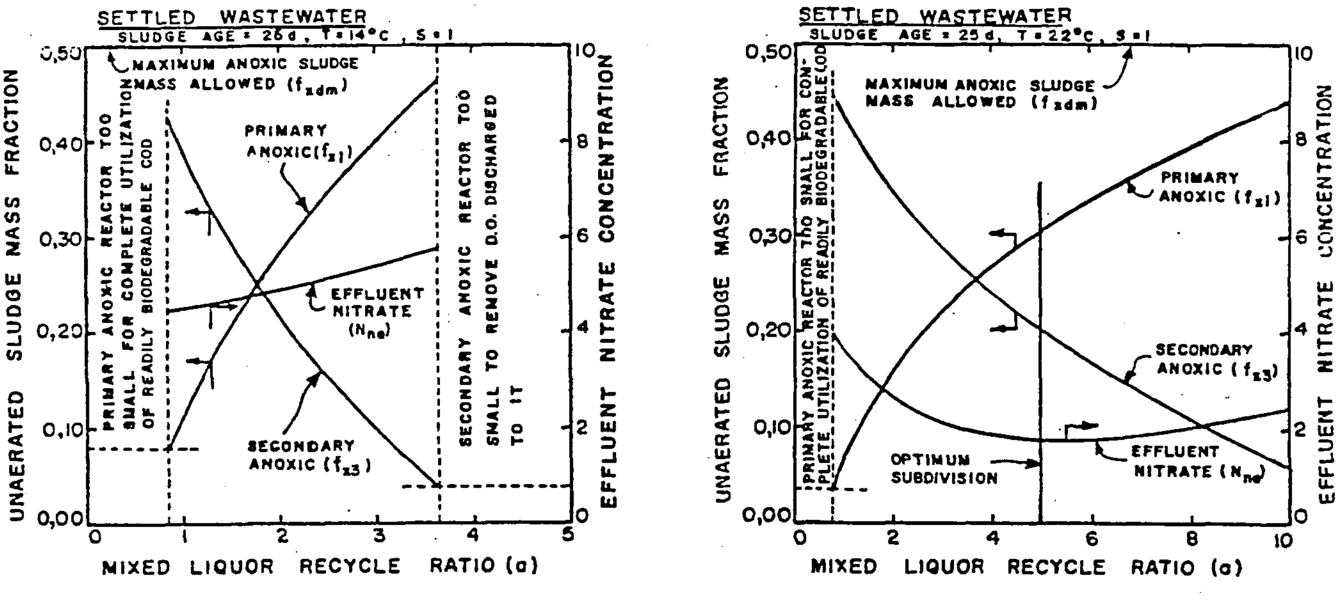

FIGURE 6.16: Primary (f_{x1}), secondary (f_{x3}), total (f_{xdt}) and maximum allowable (f_{xdm}) anoxic sludge mass fractions and effluent nitrate concentration (N_{ne}) for the Bardenpho process treating settled wastewater at 25 days sludge age at 14°C (left, Figure 6.16a) and 22°C (right, Figure 6.15b).

fixed f_{x1} value, this a-recycle ratio being the optimum a_o; at 14°C, $a_o = 3,3$ (Figure 6.15a) and for 22°C, $a_o = 5,0$ (Figure 6.15b). This demonstrates that as the temperature increases, the a-recycle ratio needs to be increased to maintain optimum performance. This is because with increase in temperature, the denitrification potential of the primary anoxic reactor (D_{p1}) increases and consequently more nitrate must be recycled to it to fully utilize its D_{p1}. It should be noted that with f_{x1} and f_{x3} fixed, the use of any a-recycle ratio other than a_o results in poorer denitrification performance: if $a < a_o$, D_{p1} is not fully utilized and if $a > a_o$, unnecesarily large quantities of DO are discharged to the primary anoxic reactor.

The design parameters calculated above are summarized in Table 6.3.

The design procedure *for the settled wastewater* at *14°C and 22°C* was repeated and the results are shown plotted in Figures 6.16a and b respectively. At 22°C (Figure 6.16b), a minimum effluent nitrate concentration[*] of 1,7 mgN/ℓ is obtained with $a = 5,0$. The f_{x1} corresponding to $a = 5,0$ is 0,30 and this value fixes f_{x1}. Now because the maximum anoxic sludge mass fraction is fully utilized i.e. $f_{xdt} = f_{xdm}$, f_{x3} is fixed at the difference between f_{xdm} and f_{x1} i.e. $f_{x3} = 0,20$. Hence f_{x1} and f_{x3} are fixed at 0,30 and 0,20 respectively for 14°C and 22°C. The optimum a-recycle ratio is 5,0 at 22°C yielding an effluent nitrate concentration of 1,7 mgN/ℓ and that at 14°C is 2,2 (i.e. the value corresponding to $f_{x1} = 0,30$, Figure 6,16a) yielding $N_{ne} = 4,9$ mgN/ℓ. The design results are listed in Table 6.3.

The design calculations above show that with the *raw wastewater*, complete denitrification can be obtained and is achieved with a total anoxic sludge mass fraction (f_{xdt}) of 0,40 whereas 0,50 is allowed (f_{xdm}). As N removal only is important in this example, the Bardenpho process is appropriate and (i) if a factor of safety for complete denitrification is required, option 1 (see Section 6.2 above) can be taken, i.e. f_{x1} and f_{x3} are increased such that $f_{x1} + f_{x3} = f_{xdt} = f_{xm} = 0,50$; or (ii) if no factor of safety for complete denitrification is required, option 2 can be taken, i.e. the sludge age can be reduced such that $f_{xm} = f_{xdt} = 0,40$.

For the *settled wastewater*, complete denitrification cannot be achieved at both 14°C and 22°C. Because the maximum anoxic sludge mass fraction is fully utilized (i.e. $f_{xdt} = f_{xdm} = 0,50$), the denitrification efficiency can be improved only by increasing the sludge age of the process. However, the increase in sludge age is at the expense of an increased process volume. A sluge age of 40 d, which allows $f_{xm} = f_{xdm} = 0,60$,[**] is required to produce complete denitrification at 14°C (found by trial and error using design procedure). From Eq. 4.21 an increase in sludge age from 25 to 40 days represents an increase in process volume of about 40%. Such a large increase in volume may not be merited for an improvement in effluent nitrate concentration from 5 to 0 mgN/ℓ. Consequently a sludge age of 25 to 30 d is considered the maximum economical sludge age beyond which there is little cost effective improvement in denitrification efficiency. If complete denitrification must be achieved, it may be preferable to treat raw wastewater at 20-day sludge age which allows an $f_{xdm} = f_{xm} = 0,40$ (see Figure 6.15a). Although this increases the process volume by 80% it removes the problem of primary sludge disposal. Alternatively, when treating settled wastewater, and complete denitrification must be achieved, a 20-day sludge age process with readily biodegradable COD addition to the primary or secondary anoxic reactor may be aconomically viable – if 60[***] mgCOD/ℓ influent readily biodegradable COD is added to the plant, complete denitrification will be obtained with settled wastewater at 14°C and 20-day sludge age. For the design flow of 13,3 Mℓ/d, the COD addition is 60.13,3 = 800 kg/d i.e. 10% of the raw wastewater COD load.

If complete denitrification is not essential, the Bardenpho process as designed for the settled wastewater i.e. sludge age = 25 d, is appropriate. Alternatively, the feasibility of an MLE proces can be checked to see whether or not this process may yield a more economical design. The design procedure for the MLE process is demonstrated below.

The design procedure for the MLE process is identical to that for the Bardenpho process up to step 9 of the step by step procedure. For the same design conditions of the Bardenpho process (i.e. $R_s = 25$ d etc), the results for the MLE process are listed in Table 6.3 and can be seen to be identical to those of the Bardenpho process. Substituting the values for the nitrification capacity N_c and denitrification potential D_{pp} into Eqs. 6.30 and 6.31, the optimum mixed liquor a-recycle ratio a_o and minimum effluent nitrate concentration N_{ne} are obtained; e.g. for the *settled wastewater at 14°C* –

$$A = 2/2,86 = 0,70$$

$$B = 32,3 - 31,1 + \{(1+1)2 + 1.1\}/2,86 = +2,95$$

$$C = (1+1)(31,1 - 1.1/2,86) - 1.32,3 = +29,2$$

Hence

$$a_o = \{-2,95 + \sqrt{2,95^2 + 4.0,70.29,2}\}/(2.0,70)$$

$$= 4,7 \text{ and}$$

$$N_{ne} = 32,3/(4,7 + 1 + 1) = 4,8 \text{ mgN}/\ell.$$

The above results, as well as those for the settled wastewater at 22°C and those for the raw wastewater at 14°C are listed in Table 6.3.

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[*] If T_{max} is less than 18°C, a clear minimum cannot be discerned (see Figure 6.16b at 14°C). In this event, any reasonable f_{x1} can be selected with the proviso that complete utilization of the readily biodegradable COD can be assured even with a significant decrease in K_{120} (see Eqs. 6.12 and 6.21). The existence of a minimum, and the temperature above which it becomes discernible, depends much on the choice of the DO concentrations in the recycles, O_a and O_s.

[**] This is the recommended upper limit for the maximum unaerated sludge mass fraction (see Chapter 5, Section 4.4).

[***] Assuming that $N_{ne} = 7,0$ mgN/ℓ at $R_s = 20$ d (i.e. 2 mgN/ℓ greater than at $R_s = 25$ d), the COD consumption for its removal is 7,0.8,6 = 60 mgCOD/ℓ influent.

The results show that for all cases except one (i.e. for settled wastewater at 14°C), the optimum mixed liquor a-recycle ratio a_o exceeds 5. Although these results include the discharge of dissolved oxygen to the anoxic reactor, recycle ratios above 4 to 5 are not cost effective – the small decrease in effluent nitrate concentration (N_{ne}) which an increase in a-recycle ratio from 5 to 10 produces does not warrant the additional pumping costs (e.g. for $N_c = 33$, $N_{ne} = 4{,}7$ for $a = 5$ and 2,7 for $a = 10$). Consequently, the a-recycle ratio is limited at 5, which results in a minimum N_{ne} from the MLE process of about 5 to 6 mgN/ℓ for most municipal wastewaters. Hence for N removal processes, if effluent nitrate concentrations lower than 5 to 6 mg/N/ℓ are required, the Bardenpho process is the indicated process because this process can achieve low N_{ne} with low a-recycle ratios.

Comparing the Bardenpho and MLE processes, (see Table 6.3) if the effluent quality standards require that complete or near complete denitrification must be produced, then the Bardenpho process is the more superior process for both the raw and settled wastewaters. However, the process needs to be operated at a sludge age of about 20 days for the *raw wastewater* (yielding complete denitrification) and at least 30 days for the *settled wastewater* (yielding an effluent nitrate concentration of about 3 mgN/ℓ) – the settled wastewater process will have a smaller volume than the raw wastewater process but will have the additional problem of primary sludge disposal. Alternatively, if complete denitrification is not necessary and an effluent nitrate concentration of 5 to 6 mgN/ℓ is acceptable, the MLE process will be an acceptable process. For the raw wastewater, a sludge age of only 15 days (yielding $f_{xm} = f_{xdm} = 0{,}32$) is sufficient to produce an effluent nitrate concentration of about 5 mgN/ℓ with the MLE process and for the settled wastewater, a sludge age of 25 days will yield an effluent nitrate concentration of about 5 mgN/ℓ.

7.6 Process volume and oxygen demand

7.6.1 Process volume

Having determined the subdivision of the sludge mass into anoxic and aerobic fractions to achieve the required nitrogen removal, the actual sludge mass in the process needs to be calculated to determine the volumes of the different reactors. The mass of total (MLSS) or volatile (MLVSS) sludge in the activated sludge process for a selected sludge age and wastewater characteristics is given by Eq. 4.21 (see Chapter 4, Section 6) i.e. for the given raw and settled wastewater characteristics, the MLSS sludge mass in the process at 25-day sludge age and 14°C is 54 200 and 21 850 kgTSS respectively. Selecting an MLSS concentration in the process of 4 000 mg/ℓ* (i.e. 4 kg/m³), the volume of the process treating raw wastewater is 13 550 m³ and that treating settled wastewater is 5 460 m³. Now, because the sludge mass in the nitrogen removal processes is uniformly distributed in the process, i.e. each reactor has the same MLSS concentration, the volume fractions of the reactors are equal to the sludge mass fractions. Accepting the Bardenpho process design for the raw and settled wastewaters (see Table 6.3) the volume of the reactors are found directly from the sludge mass fraction and total process volumes; reactor retention times, nominal and actual are found from the reactor volumes and the nominal and total flows passing through them – see Table 6.4. Note that the reactor nominal retention time is a consequence of the mass of sludge generated from the influent COD load, the selected MLSS concentration and the sludge mass fraction – the retention time *per se* has no significance in kinetics of nitrification and denitrification (see Chapter 4, Sections 6 and 12).

TABLE 6.4 REACTOR VOLUMES, NOMINAL AND ACTUAL RETENTION TIMES FOR THE BARDENPHO PROCESS TREATING THE TYPICAL RAW AND SETTLED WASTEWATER AT 25-DAY SLUGE AGE. FOR PROCESS PERFORMANCE SEE TABLE 6.3

Process Parameter	Symbol	Raw Waste-water	Settled Waste-water	Units
Sludge mass	$M X_t$	54 200	21 850	kgTSS
MLSS concentration	X_t	4 000	4 000	mg/ℓ
Total volume	V_p	13 550	5 460	m³
Total flow	Q	13 330	13 330	m³/d
Total nominal retention time	R_{nt}	24,4	9,8	h
Mixed liquor recycle (14°C)	a	3,3	2,2	
Und. recycle	s	1,0	1,0	
Primary anoxic				
Mass fraction	f_{x1}	0,20	0,30	
Volume		2 710	1 640	m³
Nom. ret. time		4,9	2,9	h
Actual ret. time		1,13	0,92	h
Main aeration				
Mass fraction*		0,45	0,45	
Volume		6 100	2 460	m³
Nom. ret. time		11,0	4,4	h
Actual ret. time		2,6	1,4	h
Secondary anoxic				
Mass fraction	f_{x3}	0,30	0,20	
Volume		4 060	1 090	m³
Nom. ret. time		7,3	2,0	h
Actual ret. time		3,7	1,0	h
Reaeration				
Mass fraction*		0,05	0,05	
Volume		680	270	m³
Nom. ret. time		1,2	0,50	h
Actual ret. time		0,6	0,25	h

*The reaeration reactor is usually taken to have a sludge mass fraction of around 0,05 to 0,07 and this fraction is deducted from the aerobic sludge mass fraction; the remaining aerobic sludge mass fraction constitutes the main aeration sludge mass fraction.

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* A method for estimating the reactor MLSS concentration for design conditions is as follows: for a selected X_t calculate the process volume requirements. For accepted settling characteristics of the sludge, calculate the required settler surface area for the selected X_t (see Chapter 8). Estimate the combined cost of the process and settling tanks. It will be found that as X_t increases the cost of the process decreases but the cost of the settling tanks increases. The X_t that gives the minimum combined cost is the indicated MLSS concentration (usually between 3 000 and 6 000 mgTSS/ℓ with settled wastewater being lower than raw wastewater).

7.6.2 Daily average total oxygen demand

The total oxygen demand in a nitrogen removal process is the sum of that required for carbonaceous material (COD) degradation and nitrification, less than recovered by denitrification. The daily average oxygen demand for (i) carbonaceous material removal (MO_c) is given by Eq. 4.15 (see Chapter 4, Section 7), and (ii) nitrification is given by Eq. 5.39 (see Section 7.4 above and Chapter 5, Section 6.5) and these oxygen demands in the Bardenpho process at 25-day sludge age for the typical raw and settled wastewaters at 14°C and 22°C are listed in Table 6.5.

TABLE 6.5 DAILY AVERAGE OXYGEN DEMAND (IN kgO/d) IN THE 25-DAY SLUDGE AGE BARDENPHO PROCESS FOR THE TYPICAL RAW AND SETTLED WASTEWATERS AT 14°C AND 22°C

Oxygen Demand kgO/d	Symbol	Raw		Settled	
		14°C	22°C	14°C	22°C
1. Carbonaceous*	$M(O_c)$	+5105	+5205	+3287	+3352
2. Nitrification**	$M(O_n)$	+1973	+2071	−1961	+2046
3. Denitrification***	$M(O_d)$	−1235	−1296	−1048	−1216
4. Total	$M(O_t)$	+5843	+5980	+4200	+4185

*See Chapter 4, Section 12.3, Eq. 4.27
**See Section 7.4 above
***From Eq. 6.32 and Table 6.3

The oxygen recovered by denitrification (MO_d) is given by 2,86 times the nitrate mass denitrified (see Section 1.3) where nitrate mass denitrified is the product of the daily average influent flow Q and the nitrate concentration denitrified. The nitrate concentration denitrified is given by the difference in the nitrification capacity N_c and the effluent nitrate concentration: Hence

$$M(O_d) = 2{,}86.(N_c\text{-}N_{ne})Q \qquad (6.32)$$

From the denitrification performance data of the Bardenpho process in Table 6.3, the oxygen recovered by denitrification for the typical raw and settled wastewaters are listed in Table 6.5.

For the raw wastewater, Table 6.5 shows that (i) the nitrification oxygen demand (MO_n) is about 40% that required for COD removal (MO_c), (ii) about 60% of MO_n can be recovered by incorporating denitrification, (iii) the additional oxygen demand by incorporating nitrification and denitrification is only 15% of that required for COD removal only and (iv) the effect of temperature on the total oxygen demand is marginal – less than 3% (see also Figure 6.1).

For the settled wastewater, Table 6.5 shows that (i) the nitrification oxygen demand is about 60% of that required for COD removal, (ii) about 50% of the nitrification oxygen demand can be recovered by denitrification, (iii) the additional oxygen demand by incorporating nitrification and denitrification is about 30% of that required for COD removal only and (iv) the effect of temperature on the total oxygen demand is marginal – less than 3% more at the lower temperature.

Comparing the oxygen demand for the raw and settled wastewaters, it can be seen that the total oxygen demand for the latter is about 30% less than that of the former. This saving is possible because primary sedimentation removes 35 to 45% of the raw wastewater COD (see Chapter 4, Section 12 and Chapter 5, Section 6.5). Furthermore, for the settled wastewater, the nitrification oxygen demand is a greater proportion of the total, and also, less of the nitrification oxygen demand can be recovered by denitrification compared to the raw wastewater. These effects are due to the higher TKN/COD ratio of the settled wastewater.

Generally for design, preliminary estimates of the total daily average oxygen demand can be roughly approximated with the following:

(i) including complete nitrification;

$$M(O_t) = \{1+5f_{ns}\}\,M(O_c) \qquad (6.33)$$

(ii) including complete nitrification, and denitrification to the degree possible depending on the TKN/COD ratio (see Figure 6.17)

$$M(O_t) = \{1+2f_{ns}\}\,M(O_c) \qquad (6.34a)$$

for $f_{ns} < 0{,}09$ (i.e. effluent nitrate < 2 mgN/ℓ)

$$= \{1+2f_{ns}\,[1+10(f_{ns}-0{,}08)]\}M(O_c) \qquad (6.34b)$$

for $f_{ns} > 0{,}09$ (i.e. effluent nitrate > 2 mgN/ℓ)

where f_{ns} = influent TKN/COD ratio.

These equations, in conjunction with the carbonaceous oxygen demand $M(O_c)$ equation in terms of the influent COD load (MS_{ti}) (see Chapter 4, Section 7) will allow the calculation of preliminary estimates of the daily average total oxygen demand in nitrogen removal processes between sludge ages of 15 to 30 d for raw and settled wastewaters for known influent COD loads and TKN/COD ratios.

Knowing the average daily total oxygen demand, the peak total oxygen demand can be roughly estimated by means of a simple design rule. From a large number of simulations with the general dynamic model, it was found that provided the factor of safety on nitrification (S_f) is greater than 1,25 to 1,35, the relative amplitude {i.e. (Peak-Average)/Average} of the total oxygen demand is a fraction 0,50 of the relative amplitude of the total oxygen demand potential of the influent COD and TKN load {i.e. $Q(S_{ti}+4{,}57.N_{ti})$}. For example, with the raw wastewater design, if the peak influent total oxygen demand potential is obtained at a time of day when the influent flow rate, COD and TKN concentrations are 23,1 Mℓ/d, 863 mgCOD/ℓ and 65 mgN/ℓ respectively – i.e. 23,1 (863+4,57.65) = 26 800 kgO/d – and the average total oxygen demand potential is 13,33 (600+4,57.48 = 10 920 kgO/d (see Table 4.3), the relative amplitude of the total influent oxygen demand potential is (26 800 – 10 920)/10 920 = 1,45; hence the relative amplitude of the total oxygen demand is approximately 0,50.1,45 = 0,72; from Table 6.5 the average daily total oxygen demand is 5 980 kgO/d (at 22°C) and hence the peak oxygen demand is 1,72.5 980 = 10 300 kgO/d. As with all simplified design rules, the above rule should be used with discretion and caution, and where possible, the peak total oxygen demand is best estimated by means of the general dynamic activated sludge model.

7.7 Closure

In the above design example, the raw wastewater TKN/COD ratio is 0,08 mgN/mgCOD. If the calculations are repeated for the raw wastewater but assuming different TKN/COD ratios and sludge ages, it will be found that (i) for a fixed sludge age, once the TKN/COD ratio is above that with which complete denitrification can be achieved, the effluent nitrate concentration increases as the TKN/COD ratio increases and (2) at a fixed TKN/COD ratio above that with which complete denitrification can be achieved, as the sludge age increases, the effluent nitrate concentration decreases. This implies that at a fixed sludge age, the N removal potential of the process is fixed so that the effluent quality depends on the magnitude of the influent TKN/COD ratio; however, the N removal potential of the process may be improved by increasing the sludge age.

Based on 14°C and the raw wastewater characteristics of the design example (except the TKN/COD ratio) the behaviour described above is shown graphically in Figure 6.17, in which the influent TKN/COD ratio is plotted versus the effluent nitrate concentration that can be achived at different sludge ages. The process volume required at the different sludge ages (as a fraction of that required at 25 d) is also given on Figure 6.17. The areas for which the Bardenpho and MLE process achieve the optimum N removal are demarcated and generally where the effluent nitrate concentration is greater than 5 to 6 mgN/*l*, the MLE process will yield an improved efficient nitrate concentration at an economical mixed liquor a recycle ratio i.e. < 4:1. It should be noted that at a fixed TKN/COD ratio, the total average daily oxygen demand does not vary significantly between different sludge ages so that aeration costs do not significantly affect the conclusions that can be drawn from Figure 6.17 (see Figure 6.1).

Figure 6.17 shows that for a TKN/COD ratio of say 0,095 mgN/mgCOD, complete denitrification can be achieved if the sludge age is 30 days in a Bardenpho process. However, if an effluent quality of 7 mgN/*l* is acceptable, then this can be achieved in an MLE process at a sluge age of 20 days. The process volume saving that can be made by reducing the effluent nitrate standard from 0 to 6 mgN/*l* is 1 − 0,83/1,16 = 0,28 i.e.

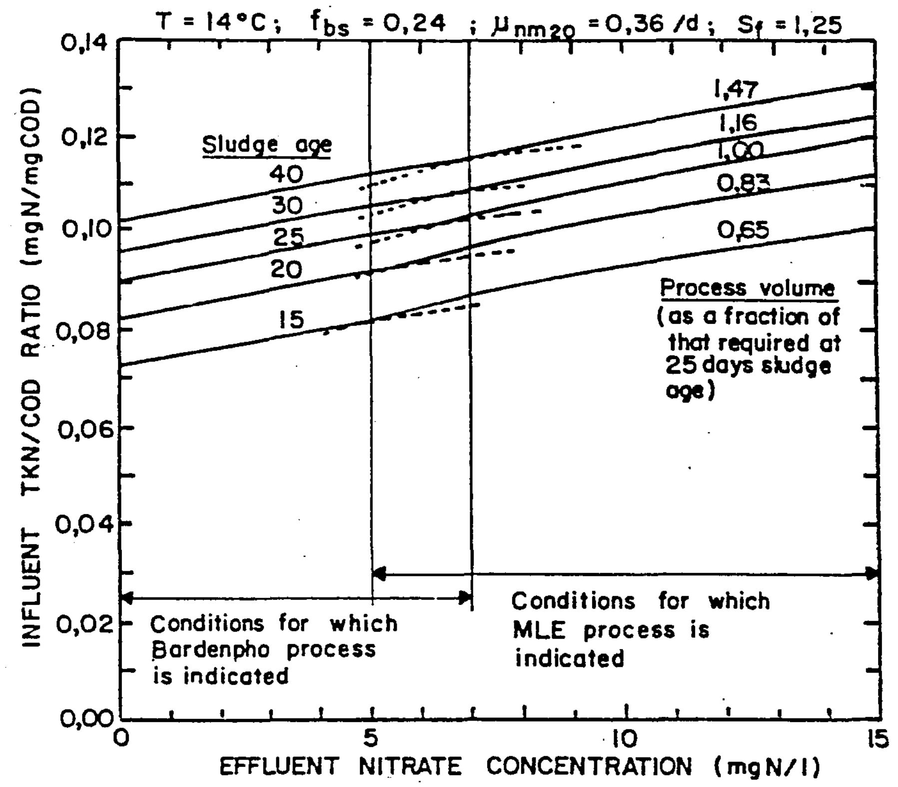

FIGURE 6.17: Effluent nitrate concentration attainable for different influent TKN/COD ratios for N removal processes operating at different sludge ages for T = 14°C, readily biodegradable COD fraction (f_{bs}) = 0,24 and maximum specific growth rate of the nitrifiers at 20°C (μ_{nm20}) = 0,36/d.

the MLE process at 20-day sludge age with 7 mgN/ℓ is about 30% smaller than the Bardenpho process at 30-day sludge age with complete denitrification. Figure 6.17 shows also that for low μ_{nm20} values, 14°C and 25-day sludge age (to avoid excessively large process volumes), the Bardenpho process can achieve (i) complete denitrification for TKN/COD ratios less than 0,090 mgN/mgCOD; (ii) near complete denitrification (i.e. effluent nitrate concentration between 1 and 5 mgN/ℓ) for TKN/COD ratios between 0,09 and 0,10 mgN/mgCOD and (iii) incomplete denitrification (i.e. effluent nitrate concentration above 5 mgN/ℓ) for TKN/COD ratios above 0,10 mgN/mgCOD; hence the MLE process is the indicated process for TKN/COD ratios above 0,10. The above limits are relatively sensitive to the μ_{nm20} value and readily biodegradable COD fraction; (i) the limits are given for a low μ_{nm20} value (around 0,36/d) because this will be the usual case in design unless sufficient information is available justifying the use of a higher one (see Appendix 3) – if μ_{nm20} is greater than 0,40/d, the results will be similar to those shown in Figure 6.17 except that they can be achieved at shorter sludge ages; (ii) the limits given are for an approximately normal raw wastewater readily biodegradable COD fraction f_{bs} of 0,24 – there are many factors which can affect the f_{bs} fraction of a wastewater (see Section 7.1.3 above) so that at the design stage, the value of f_{bs} also will be uncertain [unless the value has been measured (see Appendix 2)] and it is likely that a value of 0,24 will be used initially for want of a better estimate; if the actual f_{bs} value is greater than 0,24, lower effluent nitrate concentrations will be obtained than those given in Figure 6.17 and if f_{bs} is lower than 0,24, higher effluent nitrate concentrations will be obtained.

Although the limits given in Figure 6.17 should not be taken as general because of their dependence on the wastewater characteristics, in particular μ_{nm20} and f_{bs}, they are useful for setting approximate guidelines for N removal process selection and the sludge age at which it is to be operated to achieve a certain effluent nitrate concentration for a given influent TKN/COD ratio. The guidelines in Figure 6.17 are not intended to obviate doing detailed design calculations but serve the purpose of demonstrating the behaviour of N removal processes. Indeed it is recommended that in process design, the wastewater characteristics are changed to encompass a range of possible values for the particular wastewater and to undertake design calculation for combinations of different wastewater characteristics. This will allow the selection of a process design that can best deal with anticipated changes in wastewater characteristics and will expose weaknesses in the design so that control measures can be incorporated to deal with fluctuations in wastewater characteristics during the life of the plant.

8. REFERENCES

ARKLEY, M. and MARAIS, G.v.R. (1982) Effect of the anoxic sludge mass fraction on the activated sludge process. Research Report No. W 38, Dept. of Civil Eng., Univ. of Cape Town.

BARNARD, J.L. (1973) Biological denitrification, *Wat. Pollut. Control*, **72**, 6, 705-720.

CARLSON D.E. (1971) Nitrogen removal and identification for water quality control, OWRR, Project No. 4040, Dept. of Civil Eng., Univ. of Washington.

CHANG, J.P. and MORRIS, J.E. (1962) Studies of the utilization of nitrate by *Miaococcus Denitrificans*, *J. Gen. Microbial.*, **29**, 301.

CHRISTENSEN, M.H. and HARREMOËS P. (1977) Biological denitrification of sewage: A literature review. *Prog. Wat. Tech.*, **8**, 415, 509-555.

DOLD, P.L., EKAMA G.A. and MARAIS G.v.R. (1980) A general model for the activated sludge process. *Prog. Wat. Tech.*, **12**, 47-77.

EKAMA, G.A. and MARAIS, G.v.R. (1978) The dynamic behaviour of the activated sludge process. Research Report No. W 27, Dept. of Civil Eng., Univ. of Cape Town.

EKAMA, G.A., VAN HAANDEL, A.C. and MARAIS, G.v.R. (1979) The present status of research on nitrogen removal: a model for the modified activated sludge process. Presented at the Symposium on Nutrient Removal, SA Branch of IWPC Pretoria, May 1979. Proceedings published by Water Research Commission of South Africa.

LUDZACK, F.J. and ETTINGER, M.B. (1962) Controlling operation to minimize activated sludge effluent nitrogen *JWPCF*, **34**, 920-931.

MARSDEN, M.G. and MARAIS, G.v.R. (1977) The role of the primary anoxic reator in denitrification and biological phosphorus removal. Research Report No. W 19, Dept. of Civil Eng., Univ. of Cape Town.

SIEBRITZ, I.P., EKAMA, G.A. and MARAIS, G.v.R. (1980) Excess biological phosphorus removal in the activated sludge process at warm temperature climates. Proc. *Waste Treatment and Utilization*, Vol. 2, 233-251, Eds. C.W. Robinson, M. Moo-Young and G.J. Farquhar, Pergamon Press, Toronto.

STERN, L.B. and MARAIS, G.v.R. (1974) Sewage as the electron donor in biological denitrification. Research Report No. W 7, Dept. of Civil Eng., Univ. of Cape Town.

VAN HAANDEL, A.C. EKAMA, G.A. and MARAIS, G.v.R. (1981) The activated sludge process part 3 – Single sludge denitrification. *Water Research*, **15**, 1135-1152.

VAN HAANDEL, A.C. and MARAIS G.v.R. (1981) Nitrification and denitrification kinetics in the activated sludge process. Research Report No. W 39, Dept. of Civil Eng., Univ. of Cape Town.

VAN HAANDEL, A.C., DOLD, P.L. and MARAIS, G.v.R. (1982) Optimization of nitrogen removal in the single sludge activated sludge process. *Wat. Sci. Tech.*, **14**, 443.

WILSON, D.E. and MARAIS, G.v.R. (1976) Adsorption phase in biological denitrification. Research Report No. W 11, Dept. of Civil Eng., Univ. of Cape Town.

WUHRMANN, K. (1964) Nitrogen removal in sewage treatment processes, *Vehr. Int. Ver. Theor. Angew. Limnol*, **15** (in German).

CHAPTER SEVEN

BIOLOGICAL EXCESS PHOSPHORUS REMOVAL

by

G.A. Ekama, G.v.R. Marais and I.P. Siebritz

1. INTRODUCTION

From the first publication reporting phosphorus removal in excess of normal metabolic requirements in some activated sludge plants there has been controversy as to the mechanism whereby the excess P removal is accomplished, whether the mechanism is a precipitation of inorganic compounds, albeit biologically mediated, or biological through metabolic formation and accumulation of phosphorus compounds in or on the organisms. The objective of this chapter is not to discuss the evidence that supports the excess biological P removal hypothesis,* but to describe briefly the theory of biological excess P removal as understood by the authors and to demonstrate how this theory can be used as an aid for the design of biological P removal activated sludge processes. This does not imply that precipitation of inorganic phosphorus salts due to chemical changes resulting from biological action e.g. alkalinity, acidity and pH does not take place:** Such inorganic precipitation certainly can take place, but it would appear that in the treatment of municipal wastewaters by an appropriately designed activated sludge process, within the normal ranges of pH, alkalinity, acidity and calcium concentrations in the influent, excess P removal is principally mediated by a biological mechanism.

2. BACKGROUND

2.1 Earlier developments

From the reports of the earlier investigators into excess biological P removal (e.g. Fuhs and Min Chen, 1975 and Barnard, 1976), one conclusion emerged which now appears to be generally accepted: excess P removal is stimulated by stressing the organisms by withholding the oxygen supply. Quantification of this stress, however, presented major problems *and still does*, because the parameters in terms of which the stress is to be formulated are not yet explicitly identified.

Barnard (1976) from pilot scale studies (Barnard, 1975a, b) on the Bardenpho process (Figure 7.1) reported that excess biological P removal is induced if at some point in the process configuration the organism mass is stressed by subjecting it to an "anaerobic"*** state (i.e. a state in which neither oxygen nor nitrate is present) such that phosphorus is released by the sludge mass to the bulk liquid. He proposed to produce this stage efficiently by including an anaerobic reactor ahead of the primary anoxic reactor in the Bardenpho process, this reactor receiving the influent flow and the underflow recycle from the secondary settling tanks, (Figure 7.2). This configuration has become known in South Africa as the 5-stage Phoredox or simply the Phoredox process, and in the United States as the Modified Bardenpho process.

To explain the excess removal phenomenon, Barnard in 1976 hypothesized that it is not the release *per se* that stimulates the excess uptake mechanism but that the release indicates that a certain low redox potential has been established, i.e. that the low redox potential triggers off the release and thereby stimulates excess P uptake. In terms of this hypothesis nitrate recycled via the underflow to the anaerobic reactor will restrain, in some degree, the level to which the redox potential can be lowered and consequently, nitrate can be expected to influence excess P uptake adversely. No data on redox potential was reported. Barnard apparently accepted that the Bardenpho section of the plant should reduce the nitrate sufficiently that any nitrate in the underflow would not prevent the attainment of the low redox potential necessary for P release in the anaerobic reactor. In any event he considered that nitrate entering the anaerobic reactor could be countered by increasing the retention time of this reactor. For design of the anaerobic reactor he suggested a nominal retention time of one hour, Barnard (1976).

Barnard's work on the nitrification-denitrification excess P removal process stimulated extensive research into this process, to gain experience on its behaviour, to delineate more precisely the factors influencing excess removal and to develop criteria for design. Nicholls

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*For a review of the observations that support the biological excess P removal hypothesis, see Marais, Loewenthal and Siebritz (1982).

**Points of view with regard to biological mediated precipitation of phosphorus are presented in the Post Conference Seminar on Phosphate Removal in Biological Treatment Process, IAWPR, Pretoria 1982. Published in Water Science and Technology Vol 15 No 3/4 1983.

***Anoxic; anaerobic: The meaning we will attach to these two terms, in nitrification-denitrification processes, follows that originally used by Barnard, i.e. Anoxic: a state in which nitrate is present but no oxygen; Anaerobic: a state in which neither nitrate nor oxygen is present. The inadequacies of these definitions are apparent when attempting to compare the state of two reactors of the same size in a completely mixed and a plug flow reactor respectively. A completely mixed anaerobic reactor, for example, will have no nitrate in the reactor and effluent; the equivalent plug flow reactor however may contain nitrate for a considerable portion of the reactor length i.e. be partly "anoxic", partly "anaerobic" – the inadequacy arises in that no indication is given as to the *intensity* of the state.

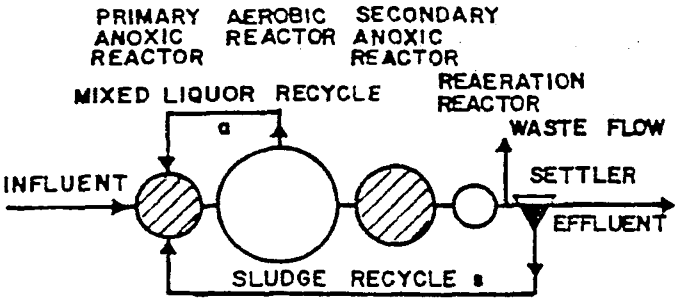

FIGURE 7.1: The Bardenpho process for biological nitrogen removal.

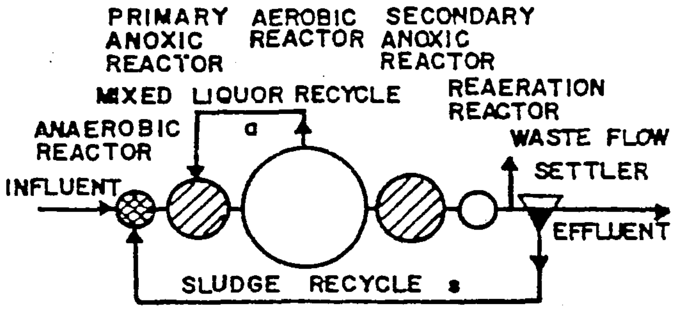

FIGURE 7.2: The Phoredox process for biological nitrogen and phosphorus removal, also called the Modified Bardenpho process.

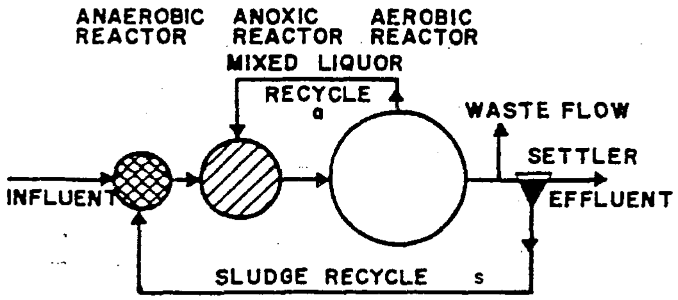

FIGURE 7.3: The 3 stage Phoredox process for biological nitrogen and phosphorus removal.

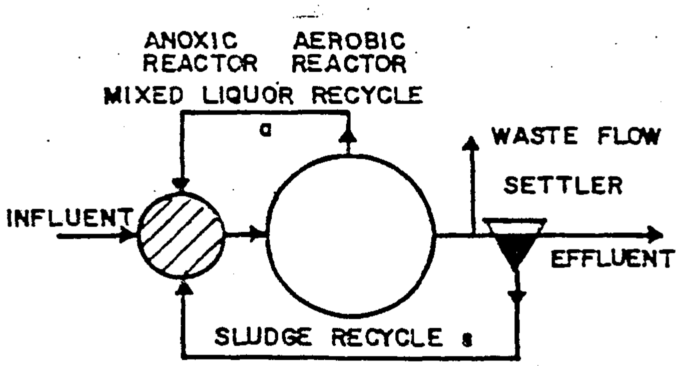

FIGURE 7.4: The modified Ludzack-Ettinger process for biological nitrogen removal.

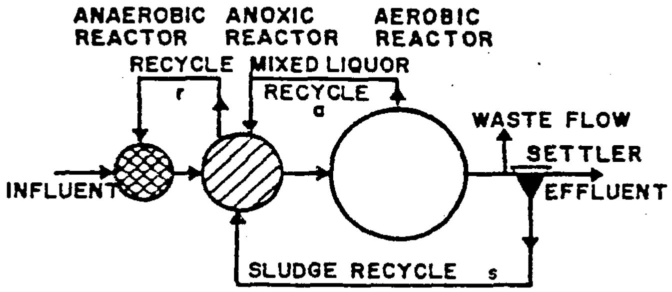

FIGURE 7.5: The UCT process for biological nitrogen and phosphorus removal.

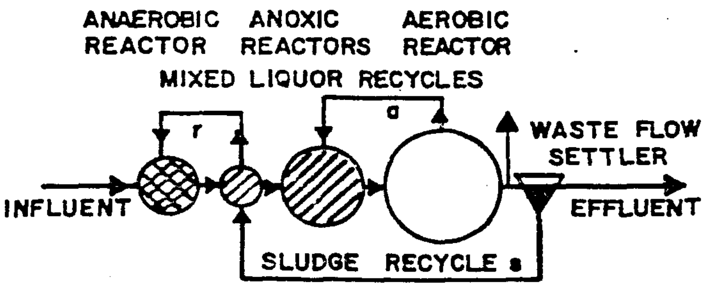

FIGURE 7.6: The modified UCT process for biological nitrogen and phosphorus removal.

(1975) at full scale, and McLaren and Wood (1976) and Simpkins and McLaren (1978) at laboratory and pilot scale applied the Phoredox process at approximately 20°C to treat effluents from Johannesburg and Pretoria, and were successful in attaining excess P removal. They were also successful in obtaining excess P removal in the 3-stage Phoredox (Figure 7.3) i.e. a Phoredox process without the secondary anoxic and reaeration reactors. Simpkins and McLaren found that (1) the anaerobic reactor was necessary for excess P removal, (2) nitrate in the recycle adversely affected P removal and (3) increasing the volume of the anaerobic reactor increased the excess P removal. These findings were in conformity with Barnard's redox potential hypothesis although again no redox potential measurements were reported.

None of the investigations above provided a reliable model to predict the magnitude of the denitrification to be expected even though it was evident that for design, evaluation of the nitrate in the recycle could be crucial in assessing the success of a process both in stimulation of the P release and magnitude of the P uptake.

Marais and his group (Stern 1974, Martin 1975, Marsden 1976, Wilson 1976 with Marais) recognized the importance of quantifying the nitrate removal. To obtain information on the magnitude and kinetics of denitrification they replaced the completely mixed reactors in the Bardenpho process by plug flow reactors and measured the nitrate along the reactor axes under constant flow and load conditions. At the same time they also monitored the phosphate behaviour in the various reactors making up the processes. Their findings on denitrification kinetics are reviewed in Chapter 6. With regard to the phosphorus behaviour, with the raw sewage they used as influent, difficulties were experienced in obtaining nitrification, even at long sludge ages of 10 to 20 days, if the unaerated mass fraction of the sludge exceeded 30 to 40%. When the plants were operated at unaerated mass fractions that allowed nitrification the denitrification obtained was insufficient and the effluent nitrate was high. Yet these plants gave very good phophorus removal. However when the sewage source was changed the good P removal previously obtained declined virtually to zero.

In a subsequent study of the Phoredox process, Rabinowitz and Marais (1980) selected the 3-stage Phoredox (Figure 7.3) as the basic configuration in preference to the 5-stage for the following reason: The wastewater source did not allow an unaerated mass fraction of greater than 40 per cent at 14°C for a sludge age (R_s) of 20 days if efficient nitrification was to be maintained; taking account of the fact that in the Phoredox process the anaerobic reactor cannot contribute to the process denitrification potential, the 5-stage process could not reduce the nitrate to zero for the measured TKN/COD ratio of the waste flow. Consequently (as discussed earlier in Chapter 6) the secondary anoxic reactor volume was added to the primary anoxic to obtain the maximum nitrate removal and hence the minimum nitrate concentration in the underflow recycle. The findings from this investigation can be summarized as follows:

(1) When the nitrate concentration in the effluent (and underflow recycle) was low usually P release and excess uptake were observed. In general there was a tendency for the excess uptake to decrease quite disproportionately as the nitrate in the recycle increased, a behaviour also noted by Simpkins and McLaren, and Barnard.

(2) With different batches of wastewater having the *same* nitrate concentration in the recycle, one batch may give high P release and excess removal whereas the next may give no (or little) release and little excess removal. No apparent reason for this behaviour could be discovered.

The overall P removal performance was disappointing; not only did the plant not remove P in excess over long periods of time but the removal was erratic due to the effects of (1) and/or (2) above. Increasing the anaerobic mass fraction during periods of low P removal was found to be counter-productive as this could be done only at the expense of the *anoxic* mass fraction which in turn gave rise to increased nitrate in the recycle. It was finally concluded that for the waste flows used in the experimental investigation, treatment by the Phoredox type process was not suitable for excess P removal; this did not imply that the process might not be suitable for other waste flows but the investigation did bring to light that there were constraints, not adequately recognized before, that may prevent high P removals:

(1) For any selected sludge age and minimum temperature the requirement for efficient nitrification imposes an upper limit on the unaerated mass fraction.

(2) The limitation on the unaerated mass fraction correspondingly limits the concentration of nitrate that can be removed. If the nitrate generated is higher than the denitrification achievable, nitrate will appear in the effluent and, in the Phoredox system, the P removal will be adversely affected.

2.2 Later developments

2.2.1 *Positive exclusion of nitrate from the anaerobic reactor*

From the findings on the Phoredox process the Marais group concluded that irrespective of other factors that may affect the excess P uptake, a major factor influencing the uptake was the presence of nitrate in the underflow recycle. If the nitrate content in the underflow to the reactor could be kept at low concentration, then there was a high expectation that excess P removal would be obtained. The principal obstacle to attaining this desirable end in the Phoredox process appeared to be that the nitrate discharged to the anaerobic reactor is linked directly to the concentration in the effluent. If for any reason the nitrate concentration increased while the COD remained constant, i.e. if the influent TKN/COD ratio increased, the process appeared to offer little option to reduce this by operational means. The only operational means available was to reduce the magnitude of the underflow recycle but this was a risky option as the settleability of the mixed liquor in the plants tended to be poorer than in pure aerobic systems. Evidently a process configuration was needed that made the anaerobic reactor *independent* of the effluent nitrate concentration. Towards this end, after a series of attempts, the configuration shown in Figure 7.5 was devised called the University of Cape Town process (UCT process).

In the UCT process, the settling tank underflow s-recycle as well as the mixed liquor a-recycle are discharged to the anoxic reactor and an additional mixed liquor r-recycle from the anoxic to the anaerobic reactor is introduced. The nitrate recycled to the anoxic reactor can be controlled by appropriately adjusting the mixed liquor a-recycle such that the nitrate concentration in the outflow of the anoxic reactor remains approximately zero. Consequently the mixed liquor r-recycle from the anoxic to the anaerobic reactor will contain very little or no nitrate and the anaerobic condition in the anaerobic reactor will be optimal. Thus, *in the UCT process by application of an appropriate operational control strategy, the anaerobic reactor can be maintained independent of the nitrate in the effluent even if the influent TKN/COD ratio to the plant varies.*

Laboratory scale tests on the UCT process using waste flows from Cape Town showed improved excess P removal in both the concentration removed and consistency in removal over those obtained in the Phoredox process. But perhaps the most important achievement from a research point of view, was that with the UCT process it was possible to eliminate the confounding effect of the nitrate in the recycle flow to the anaerobic reactor on excess P removal so that other factors influencing the excess removal could be investigated with greater ease. In the experimental response data the effects of these other factors became clearly evident:

(1) For the same influent COD, one batch of sewage gave high removals, another gave low, an observation previously surmised but not explicitly identified due to the difficulty of isolating this effect when nitrate was recycled to the anaerobic reactor.

(2) Generally as the storage time of a batch of sewage (at 5°C) increased, both the excess P removal and the nitrate removal declined. Evidently on storage a progressive change was taking place in the constitution of the sewage which acted adversely on the two phenomena; direct evidence of change was indicated by the gradual reduction in COD of the batch during storage.

2.2.2 Phosphorus removal process analyses

Owing to the importance attached to the nitrate concentration discharged to the anaerobic reactor in excess biological P removal, the nitrogen removal theory was applied to the P removal processes to investigate their propensity for removing nitrogen.

(i) *The Phoredox Process:* In this process, a certain fraction (f_{xa}) of the total unaerated sludge mass fraction (f_{xt}) is "set aside" as an anaerobic reactor to establish the prerequisites for excess phosphorus removal. Only the remaining unaerated sludge mass fraction ($f_{xdt} = f_{xt} - f_{xa}$) is available as primary and secondary anoxic reactors for nitrogen removal. If no nitrate is to be recycled to the anaerobic reactor (to ensure the most "intense" anaerobic condition) complete denitrification must be achieved in the anoxic sludge mass fraction (f_{xdt}). The denitrification behaviour of the anoxic reactors in the Phoredox process for complete denitrification is the same as those in the Bardenpho process for complete denitrification. However, due to the presence of the anaerobic reactor (f_{xa}), the anoxic sludge mass fraction (f_{xdt}) for the Phoredox process ($f_{xdt} = f_{xt} - f_{xa}$) is less than that for the Bardenpho process ($f_{xdt} = f_{xt}$). Hence, the upper limit of the TKN/COD ratio for complete denitrification in the Phoredox process is lower than that for complete denitrification in the Bardenpho process, by about 0,005, for approximately normal municipal wastewaters with $f_{xa} = 0,15$ at 14°C. Consequently, for an unsettled "normal" municipal wastewater, at 25 days sludge age and 14°C, the 5 stage Phoredox process will achieve complete denitrification only if the TKN/COD ratio is less than about 0,085.* Should the TKN/COD ratio exceed this limit, complete denitrification is unlikely resulting in nitrate in the effluent and underflow s-recycle. From a design point of view, to provide a factor of safety, the TKN/COD ratio limit should not exceed 0,07* to 0,08* at 14°C, for sludge age 20 to 30 days, to have reasonable assurance that complete denitrification will be attained. The safe upper limit, of a 0,07 to 0,08 TKN/COD ratio, for successful implementation of the Phoredox process, restricts application of the process in the treatment of municipal wastewaters because the TKN/COD ratio of raw wastewater ranges between 0,07* and 0,09* and that of settled wastewater generally is above 0,10*.

(ii) *The UCT Process:* The analysis of the UCT process (Figure 7.5) showed that for this process, there is also an upper TKN/COD ratio above which excess P removal is unlikely to be attained. For a normal municipal wastewater at a TKN/COD of 0,14* (at 14°C and 25 days sludge age) the nitrate concentration in the effluent and hence in the underflow s-recycle is so high that this recycle by itself (at s = 1) fully loads the anoxic reactor to its denitrification potential i.e. the mixed liquor a-recycle needs to be reduced to zero. Hence, for TKN/COD ratios > 0,14, nitrate will be present in the primary anoxic reactor and a discharge of nitrate to the anaerobic reactor cannot be avoided leading to a decline in excess P removal. For design, to introduce a factor of safety on the

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*These limits apply to a readily biodegradable COD fraction $f_{ts} = 0,20$ ($f_{bs} = 0,24$). For higher (lower) f_{ts} fractions these limits are higher (lower). It should be noted that throughout this monograph, where TKN/COD ratio limits are given, these apply for $f_{ts} = 0,20$ ($f_{bs} \cong 0,24$) unless otherwise stated.

denitrification, the upper limit for the TKN/COD ratio above which excess P removal is unlikely to be attained in the UCT process is 0,12 to 0,14. This limit is above that for most settled and raw municipal wastewaters.

(iii) *The Modified UCT Process:* Experience with the UCT process exposed two different types of problem that could significantly affect the successful operation of the process, these were (i) process control and (ii) sludge settleability.

(a) The mixed liquor a-recycle ratio needs to be carefully controlled so that the primary anoxic reactor is just underloaded with nitrate to avoid a nitrate discharge to the anaerobic reactor. Under full-scale operation such careful control of the a-recycle ratio is not possible due to uncertainty in the TKN/COD ratio, particularly under cyclic flow and load conditions. To simplify the operation of the UCT process a modification was sought whereby careful control of a-recycle would not be necessary.

(b) As the TKN/COD ratio increases, the a-recycle ratio needs to be decreased to avoid a nitrate discharge to the anaerobic reactor, this in turn causes an increase in the *actual anoxic retention time*. For relatively high influent COD concentrations (> 500 mgCOD/ℓ), and TKN/COD ratios $> 0,11$, the actual anoxic retention time exceeds one hour. From experimental observation, in some of the units, when the actual anoxic retention time exceeded 1 hour, the settleability of the sludge declined.* To preserve good settleability of the sludge, a modification to the UCT process was sought whereby the actual anoxic retention time could be limited at 1 hour.

Both problems above are associated with the a-recycle ratio; both problems can be accommodated by a modification of the UCT process called the Modified UCT process (Figure 7.6). In the Modified UCT process, the *anoxic* reactor is subdivided into two reactors, the first having a sludge mass fraction of about 0,10 and the second having the balance of the anoxic sludge mass fraction available. The first anoxic reactor receives the underflow s-recycle and the r-recycle to the anaerobic reactor is taken from it. The second anoxic reactor receives the a-recycle. The minimum a-recycle is that which introduces just sufficient nitrate to the second anoxic reactor to load it to its denitrification potential. Any recycle higher than the minimum will not remove additional nitrate so that at higher recycles more nitrate is introduced than removed in the second anoxic reactor and nitrate will appear in the effluent from this reactor. This however is immaterial insofar as it affects the nitrate in the aerobic reactor which remains constant once $a > a_{min}$. Consequently one can raise the a-recycle to any value greater than a_{min}, to give the required actual retention time, without affecting the nitrate recycled to the first anoxic reactor – careful control of the a-recycle is no longer necessary. This improvement however is obtained at a cost: The maximum TKN/COD ratio to give a zero nitrate discharge to the anaerobic reactor is reduced from 0,14 in the UCT process to 0,11 in the Modified UCT process. However a TKN/COD ratio of 0,11 mgN/mgCOD includes most settled and raw municipal wastewaters. Furthermore, by making provision that the r-recycle can be taken from either the first or second anoxic reactor, the process can be operated either as a modified UCT or a UCT process, as may be required.

The developments described so far were all guided by the hypothesis that excess P removal is stimulated, and is best achieved, by having an anaerobic reactor and that it is optimized by preventing nitrate from entering the reactor. No information was available to quantify firstly, the conditions in the anaerobic reactor that cause the stimulation and, secondly, the magnitude of the excess P removal to be expected. Accordingly these two aspects were further investigated.

3. EVOLUTION OF EXCESS PHOSPHORUS REMOVAL MODEL

3.1 Prerequisites for excess phosphorus removal

The improved understanding of the nitrogen removal process behaviour led to a new approach in investigating the prerequisites in the anaerobic reactor for excess phosphorus removal. Ekama, van Haandel and Marais (1979) hypothesized that if the mass of nitrate entering an unaerated reactor is less than that reactor's denitrification potential, the difference defines an "*anaerobic capacity*" in the reactor. They speculated that this anaerobic capacity could substitute for the redox potential level (suggested by Barnard, 1976) as a measure for predicting when phosphorus release would take place in the anaerobic reactor. Siebritz, Ekama and Marais (1980) intensively investigated this hypothesis. They plotted the P release in the anaerobic reactor, the P uptake in the aerobic reactor and the system P removal versus the anaerobic capacity (A_c) for data from UCT and modified Phoredox processes (Figure 7.7). This plot shows that (i) if $A_c < 10$ mgNO$_3$-N/ℓ, no P release in the anaerobic reactor is obtained; in fact, P uptake is noted; (ii) when $A_c > 10$ mgNO$_3$-N/ℓ, P release is obtained; and (iii) as A_c increases above 10 mgNO$_3$-N/ℓ so the P release in the anaerobic reactor, the P uptake in the aerobic reactor and the system P removal increase.**

The data in Figure 7.7 appeared to support the anaerobic capacity hypothesis. Consequently if the anaerobic hypothesis was correct, then provided this capacity could be induced in any system other than the

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*For reasons not yet understood the anaerobic retention time, both nominal and actual does *not* appear to affect the settleability of the sludge.

**From Figure 7.7 (and Figure 7.8) it can be seen that having stimulated P release in the anaerobic reactor, the P removal increased from 4 to only 6 mgP/ℓ. The reason for this low increase in excess P removal is that the experimental units turned out to be poorly designed to achieve high P removals because the factors affecting the magnitude of excess P removal were not understood; once these factors were understood, it became possible to design processes capable of achieving high levels of excess P removal (see Figures 7.10 and 7.11).

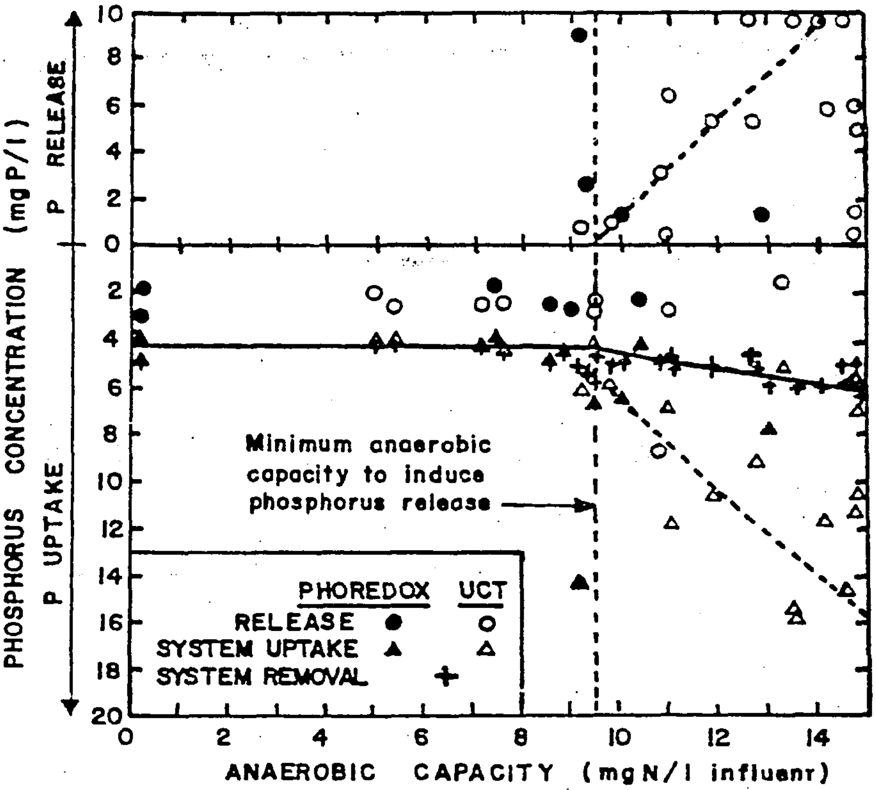

FIGURE 7.7: P release in the anaerobic reactor, process P uptake and process P removal versus the anaerobic capacity of the anaerobic reactor (A_c) for the modified Phoredox and UCT processes.

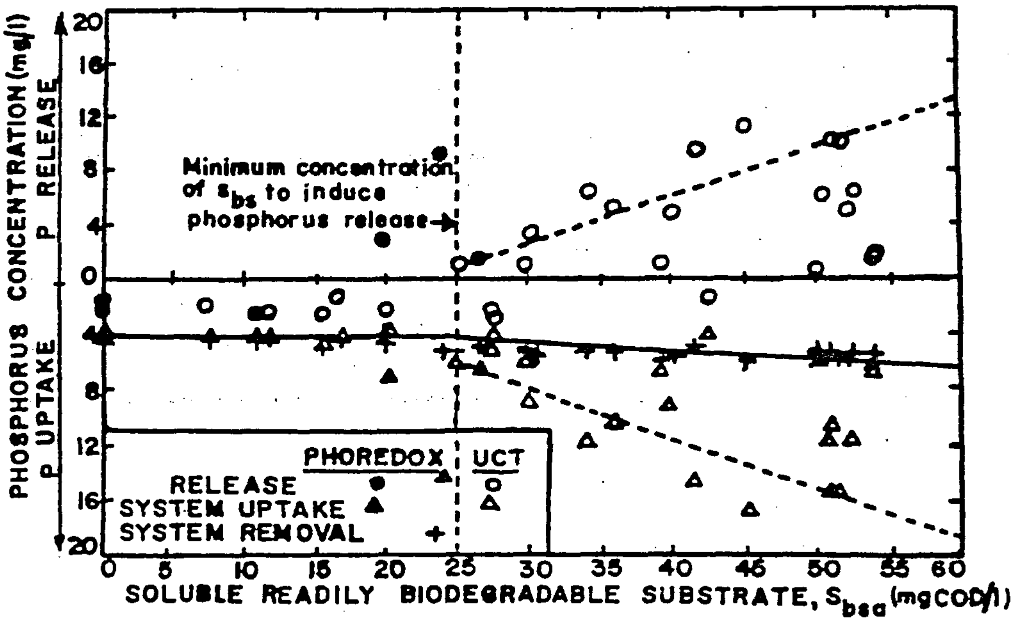

FIGURE 7.8: P release in the anaerobic reactor, process P uptake and process P removal versus readily biodegradable COD concentration in the anaerobic reactor (S_{bsa}) for the modified Phoredox and UCT processes.

UCT and Phoredox processes, P release and uptake should be observed.

To test the above conclusion experimentally, three Modified Ludzack-Ettinger (MLE) processes (Figure 7.4) and a modified UCT process were set up and fed from the same wastewater source. The three MLE units were given unaerated sludge mass fractions of 40, 55 and 70 per cent respectively and the mixed liquor a-recycle ratios were set such that the anaerobic capacities ranged from 6 to 35 mgNO_3-N/ℓ in the anoxic reactors. Over two months of operation no P release nor excess P removal were observed in any of the MLE units. In contrast, the modified UCT process with a 0,10 anaerobic sludge mass fraction, consistently gave P release and excess P removal. It was concluded from these results that the anaerobic capacity hypothesis is not sustained.

In seeking an explanation for the different P release behavioural patterns in the modified UCT and MLE processes, it was noted that the only evident difference lay in the concentration of readily biodegradable COD surrounding the organisms in the anaerobic reactor (S_{bsa}). In the modified UCT process the readily biodegradable COD concentration in the anaerobic reactor (S_{bsa}) is the maximum possible as no nitrate is recycled to the anaerobic reactor; in contrast, in the MLE process sufficient nitrate is recycled to the anoxic reactor to utilize all the readily biodegradable COD i.e. $S_{bsa} = 0{,}0$. Therefore the different behavioural patterns of the processes would be consistently described if it is assumed that the concentration of readily biodegradable COD in the anaerobic reactor (S_{bsa}) surrounding the organisms is the key parameter determining whether or not P release and excess P uptake takes place.

To test the assumption above, it was necessary to develop a general equation by means of which S_{bsa} can be calculated:

Consider the anaerobic reactor in the UCT process,

$$S_{bsa} = (f_{bs}\, S_{bi} - \Delta S_{bs})/(1+r) \quad (\text{mgCOD}/\ell) \qquad (7.1a)$$

and that in the Phoredox process

$$S_{bsa} = (f_{bs}\, S_{bi} - \Delta S_{bs})/(1+s) \quad (\text{mgCOD}/\ell) \qquad (7.1b)$$

The term ΔS_{bs} is the concentration of readily biodegradable COD utilized for synthesis of cell material with nitrate (introduced via the r-recycle) and dissolved oxygen (introduced via the influent) serving as terminal electron acceptors. Because 8,6 mgCOD are utilized for the removal of 1 mgNO_3-N and 3,0 mgCOD with 1 mgO, (see Chapter 1), ΔS_{bs} can be expressed in terms of the nitrate in the r-recycle, N_{nr}, and the DO in the influent (O_i) and r-recycle (O_r), i.e.

for the UCT process

$$\Delta S_{bs} = r(8{,}6.N_{nr} + 3{,}0.O_r) + 3{,}0\, O_i \qquad (7.2a)$$

and for the Phoredox process

$$\Delta S_{bs} = s(8{,}6\, N_{ns} + 3{,}0.O_s) + 3{,}0\, O_i \qquad (7.2b)$$

where

N_n and O refer to the nitrate and DO concentrations respectively. Symbols r and s and subscripts r,s and i refer respectively to the r- and s-recycles and the influent.

The calculation of S_{bsa} fortunately was possible because once the importance of the influent readily biodegradable COD fraction (f_{bs}) was recognized from the work on denitrification, it became standard experimental procedure to measure or determine almost daily, f_{bs}, f_{us}, f_{up}, S_{ti}, N_{nr}, O_r and O_i. Consequently Eqs. 7.1 and 7.2 were applied to the data used to plot Figure 7.7 and replotted in Figure 7.8 with S_{bsa} on the horizontal axis instead of the anaerobic capacity. From Figure 7.8 the following conclusions were drawn:

(1) The minimum readily biodegradable COD concentration in the anaerobic reactor (S_{bsa}) to stimulate phosphorus release in the reactor is about 25 mgCOD/ℓ.

(2) The degree of P release appears to increase as S_{bsa} increases above 25 mgCOD/ℓ, i.e. P release increases as (S_{bsa}-25) increases.

(3) Excess phosphorus uptake is obtained only when phosphorus release takes place, and tends to increase with (S_{bsa}-25).*

The plots in Figures 7.7 and 7.8 show a striking similarity but on closer inspection, this similarity is more apparent than real. The similarity is there only for the UCT and Phredox data: for these processes, the two sets of data fall in approximately the same position relative to the coordinate axes. This is so because in these processes, the principal contributor to A_c is the readily biodegradable COD in the influent, S_{bsi} – the contribution by the slowly biodegradable influent COD S_{bpi} is very small owing to the relatively small anaerobic sludge mass fractions (f_{xa}) – so that A_c and S_{bsa} are essentially equivalent parameters. In contrast, in the MLE process, the influent readily biodegradable COD S_{bsi} will have been completely utilized by the nitrate entering the anoxic reactor, i.e. $S_{bsa} = 0$. Nevertheless high A_c values are obtained from the S_{bpi} contribution because the unaerated sludge mass fractions are large. Consequently for the MLE process, there is no relationship between S_{bsa} and A_c. Hence in Figure 7.7, the MLE data fall in the right hand side of the plot (i.e. $A_c > 10$ mgNO_3-N/ℓ) whereas in Figure 7.8 they will fall in the left hand side of the plot (i.e. $S_{bsa} = 0$). Consequently, the MLE behaviour is incorrectly predicted by the A_c hypothesis (Figure 7.7) but correctly predicted by the S_{bsa} hypothesis (Figure 7.8). These results allowed the following statement to be made: A prerequisite for P release in the anaerobic reactor is that the concentration of readily biodegradable COD surrounding the organisms in the anaerobic reactor must exceed approximately 25 mgCOD/ℓ.

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*From Figure 7.7 (and Figure 7.8) it can be seen that having stimulated P release in the anaerobic reactor, the P removal increased from 4 to only 6 mgP/ℓ. The reason for this low increase in excess P removal is that the experimental units turned out to be poorly designed to achieve high P removals because the factors affecting the magnitude of excess P removal were not understood; once these factors were understood, it became possible to design processes capable of achieving high levels of excess P removal (see Figures 7.10 and 7.11).

3.2 Magnitude of excess phosphorus removal

A prerequisite for P release was stated above; however, the magnitudes of the P release and subsequent uptake and removal were not explicitly defined, once the prerequisite was satisfied. For this purpose an extensive and intensive investigation was inaugurated utilizing the UCT and modified UCT configurations. Series of tests were done under constant flow and load conditions: (1) at 14°C and 20°C, (2) using raw wastewater influents with COD's of 800, 500 and 300 mg/ℓ, and (3) varying the r-recycle and anaerobic mass fraction. In all the investigations the influent wastewater characteristics, i.e. COD, TKN, total P and readily biodegradable COD fraction f_{bs}, and filtered TKN, NH_4^+, NO_3^- and total P in the process reactors were measured daily. In addition, all the normal parameters such as MLVSS, oxygen utilization rate, stirred settling velocity, pH and alkalinity were measured at regular intervals.

An analysis of these data* indicated that provided $S_{bsa} > 25$ mgCOD/ℓ,

(1) for fixed r-recycle ratio and influent COD, increasing the volume of the anaerobic reactor (at the expense of the other reactor volumes) tended to increase P release and the P removal;

(2) for fixed r-recycle ratio and anaerobic reactor volume, an increase in influent COD concentration tended to increase the P release and the P removal.

Although the trends stated above were evident it was not possible to identify a consistent behavioural pattern and it was concluded that the basic parameters had not been properly identified or isolated. In part, this was due to the interactive effects between the different parameters; for example, increasing the r-recycle ratio, (i) increases the mass of sludge passing through the anaerobic reactor, (ii) decreases the time the sludge is retained in the anaerobic reactor, (iii) increases the concentration of sludge in the reactor, and (iv) decreases the concentration of S_{bsa} in the anaerobic reactor (even if no nitrate is present in the r-recycle). With so many associated changes it was difficult to see directly which parameters are the significant ones. It was decided, therefore, to investigate the system theoretically on the basis of the perceived and hypothesized behaviour.

Considering Figure 7.8, the P release, uptake and removal appeared to be linearly related and it was decided to utilize the removal as the criterion against which to judge the effect on P by any modification. As a consequence, P release *per se* was not incorporated into the hypothesis proposed to describe the P removal. The following behavioural pattern was hypothesized on the P removal based on experimental observations:

(1) Excess P removal is obtained only when $S_{bsa} > 25$ mgCOD/ℓ.**
(2) As S_{bsa} increases above 25 mgCOD/ℓ, so the P removal increases.
(3) The longer the *actual* anaerobic retention time (R_{an}), the higher the P removal.
(4) The larger the mass of sludge recycled through the anaerobic reactor, the higher the P removal;

or, transformed to a more useful equivalent:

The greater the mass of sludge recycled through the anaerobic reactor each day expressed as a fraction of the mass of sludge in the process, n, the higher the P removal.

The statements (2) and (4) above do not give an explicit quantitative expression of the expected behaviour. To obtain such an expression an hypothesis on the expected behavioural pattern of (2) to (4) has to be made. A hypothesis consistent with the observations is:

"When any one of the factors (S_{bsa}-25), R_{an}, or n, is zero, excess phosphorus removal will be zero."

From this hypothesis, the *simplest* form of an equation expressing the tendency of a process to achieve excess P removal is

$$P_f = (S_{bsa}-25).R_{an}.n \qquad (7.3)$$

where

P_f = excess phosphorus removal propensity factor.

If P_f as defined above describes the observed behaviour of the three parameters, then the P removal due to excess uptake in the sludge (P_s) should be a function of P_f, i.e.

$$P_s = f(P_f) \qquad (7.4)$$

Now $R_{an}.n$ in Eq. 7.3 can be expressed in terms of the process configuration and operational parameters, and for both the UCT and Phoredox processes this term can be shown*** to be equal to f_{xa} i.e. $R_{an}.n = f_{xa}$ where f_{xa} is the anaerobic sludge mass fraction. Substituting f_{xa} for $R_{an}.n$ in Eq. 7.3 yields

$$P_f = (S_{bsa}-25)f_{xa}; \text{ when } S_{bsa} < 25,\ P_f = 0,0 \qquad (7.5)$$

Hence for the UCT and Phoredox process, it was concluded that the propensity to give excess P removal under constant flow and load conditions is a function only of the magnitude of readily biodegradable COD concentration in the anaerobic reactor above 25 mgCOD/ℓ and the magnitude of the anaerobic sludge mass fraction.

3.3 Quantitative model for excess phosphorus removal

As excess P removal is accepted to be a biological activity, it can be assumed that, (i) only the active fraction of the sludge can take up phosphorus in excess, (ii) the phosphorus removal will increase as the proportion of phosphorus in the active mass increases, and (iii) the proportion of phosphorus in the inert fractions of the sludge will remain unchanged. Accepting these assumptions, Martin and Marais (1975) developed the following equation for phosphorus removal based on

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*Detailed description of the experimental investigation is given in a report by Siebritz, Ekama and Marais (1983).

**S_{bsa} is the concentration after allowing for removal of S_{bs} due to nitrate and oxygen entering the anaerobic reactor.

***See Siebritz, Ekama and Marais (1982).

the steady state activated sludge process model presented by Marais and Ekama (1976) (see Chapter 4, Section 9) i.e.

$$P_s = S_{ti}\,\{\frac{(1-f_{us}-f_{up})Y_h}{(1+b_{hT}R_s)}\,(\gamma + f_p\ f\ b_{hT}R_s) + f_p\,\frac{f_{up}}{f_{cv}}\} \qquad (7.6)$$

where

P_s = P removal from the wastewater by incorporation in the sludge (mgP/ℓ)

f_p = P content of the endogenous and inert fractions of the volatile mass (mgP/mgVSS)

γ = coefficient of excess P removal i.e. the P content of the active mass (mgP/mgVASS)

With the aid of Eqs. 7.1, 7.2, 7.5 and 7.6, the coefficient of excess P removal γ and the P removal propensity factor P_f were calculated for daily sets of experimental data* measured over a period of 18 months. In calculating the daily γ and P_f data pairs, those data for which there were known causes that may have adversely or favourably influenced the process behaviour were discarded.** The remaining daily γ and P_f data pairs are shown plotted in Figure 7.9.

The data in Figure 7.9 show considerable scatter although a definite trend is discernible. Some of the scatter can be attributed to the selection of the data values in the calculation of P_f: The f_{bs}, nitrate and P removal values used in the calculations were all measured on the same day, i.e. the assumption was made that the response of γ to P_f is instantaneous. This, in fact, is not so; due to the long hold up time in the system it was repeatedly observed that the P removal response is out of phase with the input by some fraction of a day. (This is apparent in Figure 7.10 which records the observed and predicted P removal in one set of experiments).

Accepting the relatively large scatter, the form of the relationship and the constraints on the $\gamma - P_f$ relationship were derived as follows: From Figure 7.9 apparently

(i) there is an upper limit to γ, of about 0,35 at very high propensity factors,

(ii) there is a lower limit to γ of about 0,06 at a propensity factor of zero,

(iii) for propensity factors ranging from infinity to zero, the maximum change in γ is $(0{,}35-0{,}06)=0{,}29$ and the decrease in γ with decrease in P_f appears to be of an exponential form.

On the basis of the above the following expression

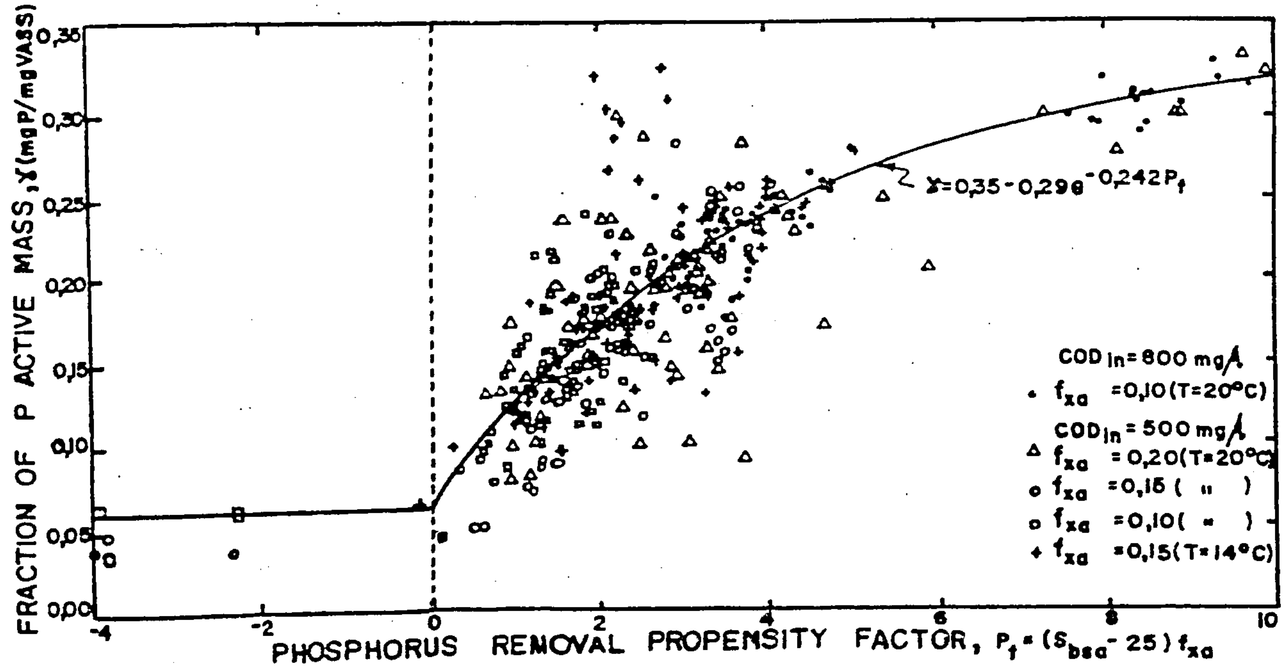

FIGURE 7.9: Coefficient of excess P removal (γ) versus P removal propensity factor (P_f) observed in the modified Phoredox, UCT and modified UCT processes at different anaerobic sludge mass fractions (f_{xa}), influent COD concentrations and temperatures.

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*The readily biodegradable COD fraction f_{bs} in the influent was measured in accordance with the procedure developed by Dold, Ekama and Marais (1980). Details of this procedure are given in Appendix 2.

**It was found that after a disruption of the process due to a mechanical breakdown, or a change in influent sewage characteristics or process parameters, the process required approximately 2 to 3 days to restabilize.

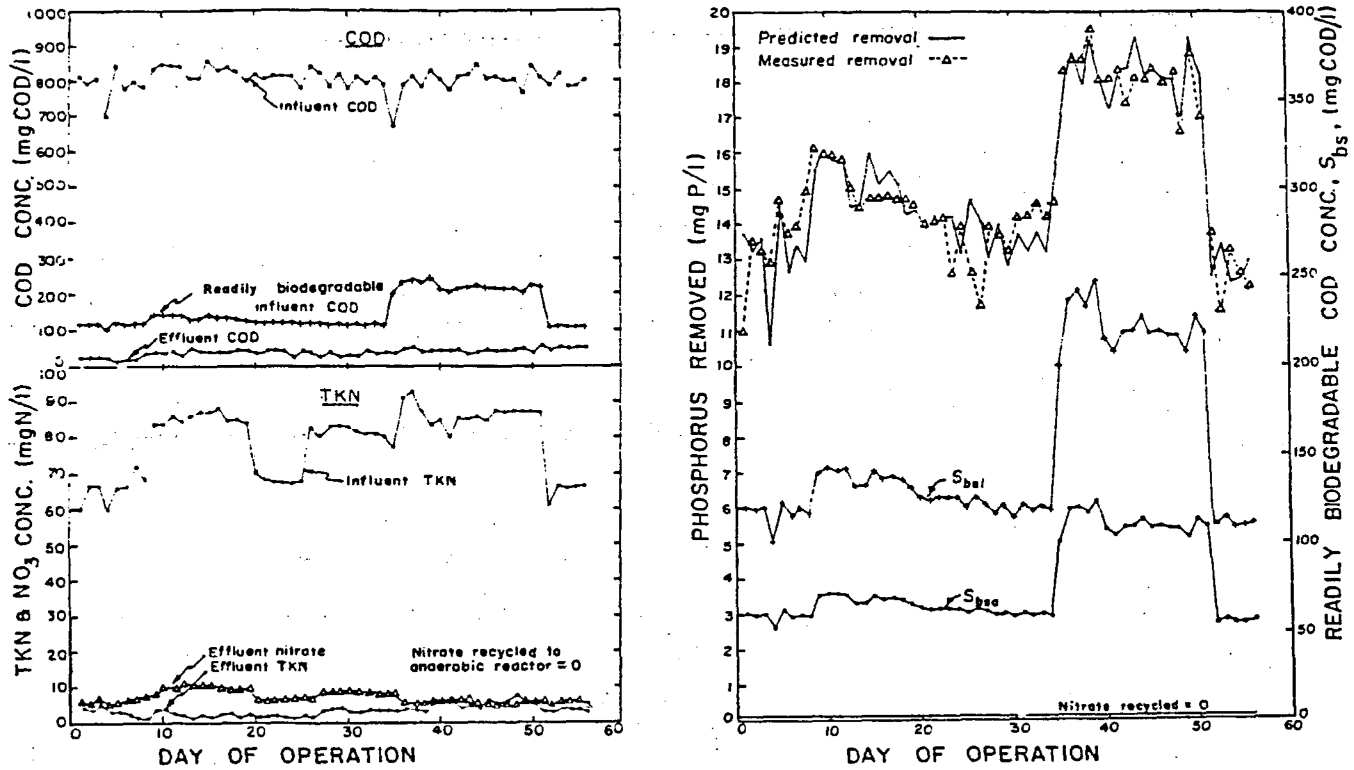

FIGURE 7.10: ***Left,*** **daily experimental data of influent COD, TKN and readily biodegradable COD concentrations and nitrate concentrations in the effluent and the recycle to the anaerobic reactor and *right,* the measured and predicted phosphorus removal for a modified UCT process with the following sludge mass fractions: anaerobic 0,10, first anoxic 0,08, second anoxic 0,22, aerobic 0,60. (Data from Siebritz, Ekama and Marais, 1983).**

was chosen to represent the trend in the plotted data

$$y = 0,35 - 0,29 \exp(-C.P_f)$$

The value of C was found by a least square analysis of the differences between the measured and predicted phosphorus removals and found to be equal to $-0,242$, i.e.

$$y = 0,35 - 0,29.\exp(-0,242\ P_f) \quad (\text{mgP/mgVASS}) \qquad (7.7)$$

To check the predictive capacity of Eq. 7.7, it was applied to the daily P_f data in all the five sets of experiments on the modified UCT, UCT and Phoredox processes in which S_{bsi} and the nitrate in the recycles were measured. (As stated above these sets of data covered influent COD values from 300 to 800 mg/*l* at 14° and 20°C under a variety of TKN/COD ratios, anaerobic mass fractions from 0,10 to 0,20 and sludge ages 12* to 25 days). As an illustration two sets of data are shown in Figures 7.10 and 7.11 together with the predicted P removal values. Evidently the model gives a good prediction of the observed P removal. In Figure 7.10, the pronounced favourable effect of an increase in the readily biodegradable COD concentration (S_{bsi}) on the P removal, and in Figure 7.11 the depressing effect of nitrate in the recycle to the anaerobic reactor, are clearly illustrated.

The theory was tested further by feeding additional readily biodegradable COD (acetate and glucose) to the anaerobic reactor. Taking due account of the increased COD, again the predictions of P removal were in satisfactory correlation with those observed.

While doing a biodegradability investigation of the waste flow from the town of Caledon in South Africa, unexpected verification of the theory was obtained. The process configuration was a Modified Ludzack-Ettinger process (Figure 7.4) with an anoxic sludge mass fraction of 0,30 operated at 20-day sludge age, at 20°C, and with a mixed liquor (a) and an underflow (s) recycle ratio of 4:1 and 1:1 respectively. The wasteflow contained an industrial fraction (from a malting factory) constituting approximately 2/3 of the total COD load. The influent COD and TKN strengths were 1300 mg/*l* and 70 mgN/*l* respectively. It was found that the malting

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*The units in the short sludge age range were non nitrification-denitrification processes of the MLE configuration (Figure 7.4) with the first reactor anaerobic and the second aerobic.

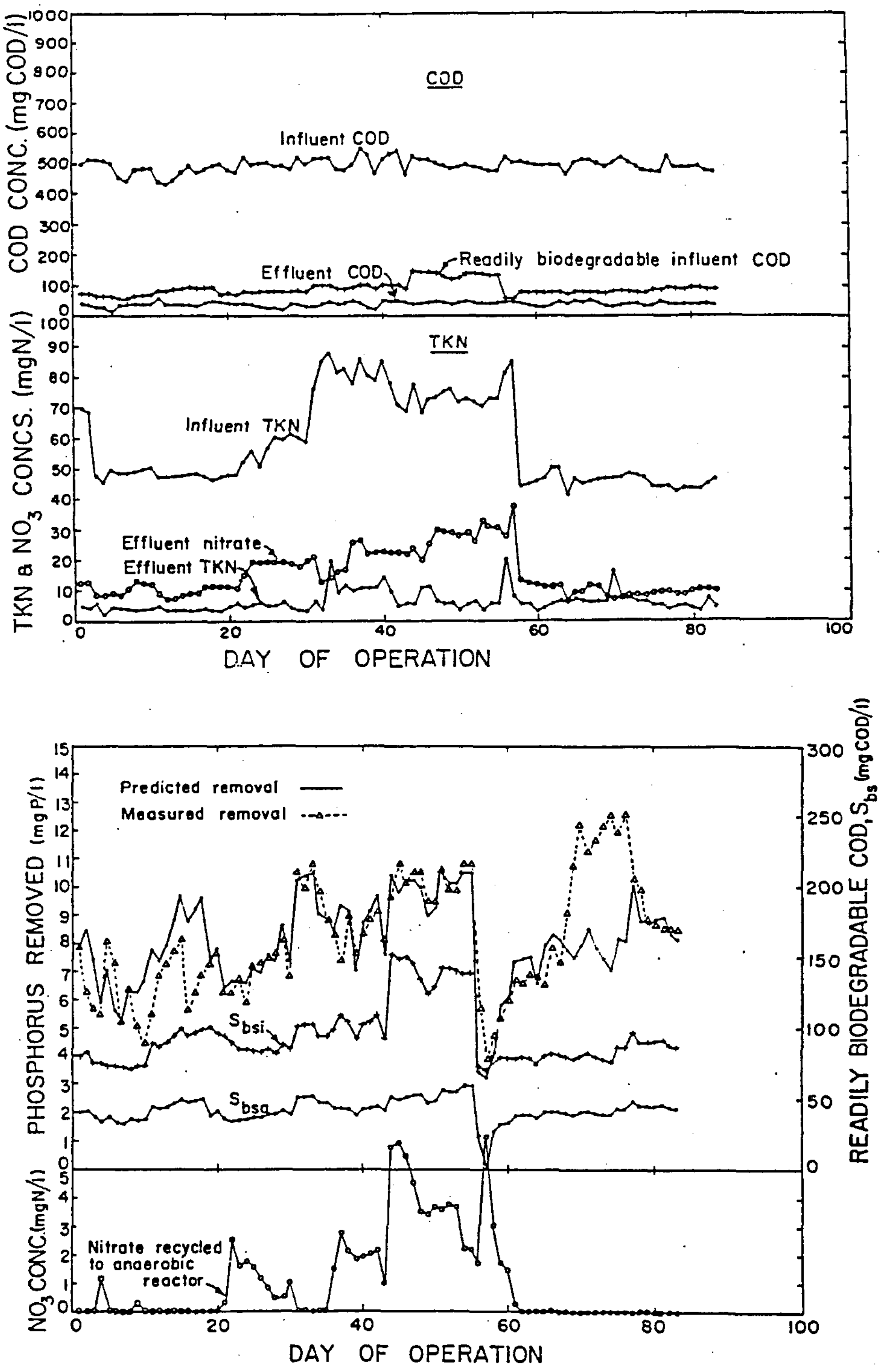

FIGURE 7.11: Response of a laboratory scale UCT process (at 14°C; anaerobic mass fraction = 0,15; anoxic mass fraction 0,30; aerobic mass fraction 0,55; sludge age 20 days;) treating unsettled flow from Mitchell's Plain. *(Data from Siebritz, Ekama and Marais, 1982b).*

waste was virtually completely readily biodegradable resulting in a very high readily biodegradable COD fraction (f_{bs}) and concentration (S_{bsi}), 0,70 and 980 mgCOD/ℓ respectively. Taking due account of the diluting effect of the total recycle flow of 5 and the reduction of the readily biodegradable COD in the "anoxic" reactor by the nitrate discharged in the recycle flows (nitrification was complete, effluent TKN = 4 mgN/ℓ), the readily biodegradable COD concentration in the "anoxic" reactor (S_{bsa}) was estimated at 101 mgCOD/ℓ, the propensity factor (P_f) at 22,9 (Eq. 7.5), the γ coefficient at 0,35 (Eq. 7.7) to give an estimated P removal of 31 mgP/ℓ influent (Eq. 7.6 with $f_{us} = 0{,}05$ and $f_{up} = 0{,}06$). The observed P removal was 29 mgP/ℓ. The results of this experiment should be contrasted with those discussed earlier on the Modified

Ludzack-Ettinger processes. In those processes, no excess P removal was observed or predicted, even though the anoxic mass fraction was as high as 0,70, principally because the readily biodegradable COD concentration in the influent (S_{bsi}) was low and completelty utilized in the anoxic reactor by the nitrate recycled to this reactor. Clearly if sufficient S_{bsi} is available and the TKN not excessive, it is possible to obtain high P removals even though the nitrate recycled to the anaerobic reactor is high, without having to resort to Phoredox or UCT configurations, provided the conditions for P release are satisfied as set out in Section 3.1 above.

At full scale Nicholls, Osborn and Marais (1982) and Nicholls (1982) tested the predictive power of the general nitrification-denitrification model on the Goudkoppies 5-stage Phoredox plant receiving an effluent that had passed through primary settling and then through an equalization tank. The P removal behaviour of the plant was also monitored by measuring P conditions in the anaerobic and aerobic reactors and in the effluent. To evaluate both the model and P removal theory, *inter alia* the readily biodegradable COD in the influent discharged to the process (S_{bsi}) was measured, as described by Dold, Ekama and Marais (1980). The S_{bsi} was found to be very low due to its utilization by an anaerobic sludge layer that formed on the bottom of the unstirred equalization tank. In terms of the theory set out here the S_{bsi} was so low that with the 1:1 underflow recycle from the secondary settling tank to the anaerobic reactor, even if no nitrate was present in the recycle, S_{bsa} would be less than 25 mg/ℓ and consequently little excess P removal was to be expected, and, virtually no excess removal was observed for the plant. Measurements of S_{bsi} on the *influent* to the equalization tank however indicated that the S_{bsi} was sufficiently high that, if the equalization tank was bypassed, excess P removal was theoretically possible provided the nitrate in the underflow recycle to the anaerobic reactor could be maintained at less than 5 mgN/ℓ. Now at this point a computer-based equalization control strategy was implemented on the equalization tank (Dold, Buhr and Marais, 1982) and operated in such a fashion that any settled material was flushed daily from the tank. The flushing action prevented build-up of a sludge layer and effectively eliminated the loss of S_{bsi} in the tank. The process recycles in the plant were changed in accordance with those predicted by the general model to achieve maximum nitrate removal. As this removal was still insufficient to reduce the nitrate in the underflow recycle to less than 5 mgN/ℓ, the aerators next to the discharge points from the primary anoxic reactor were switched off, thereby effectively increasing the anoxic zone and further reducing the nitrate in the effluent. (Sufficient aerator capacity was available in the rest of the aeration basin not to affect the nitrification efficiency). In this fashion the nitrate was kept between 3 and 5 mgN/ℓ in the effluent and the underflow recycle. Under these conditions the predicted P removal was 4 to 5 mgP/ℓ influent; observed mean removal was 5 mgP/ℓ. The results from this investigation are particularly instructive because they demonstrate that little or no P removal is achievable if S_{bsi} is too low, or, where S_{bsi} is sufficiently high, although the potential for removal is there, it will not be possible to realise this potential if nitrate in the recycle to the anaerobic reactor is so high that it reduces the S_{bsa} concentration to very low levels.

When investigating P removal behaviour in laboratory scale processes, batches of sewage usually were obtained from the outfall sewer and stored at 5°C for subsequent feed to the laboratory process units. Often it was noticed that with time the magnitudes of denitrification and P removal declined while feeding from the same raw sewage batch. Measurement of the readily biodegradable COD fraction of the daily feed taken from the cold storage batch indicated that normally there was a slow continuous loss of this fraction, giving rise to an associated reduction in denitrification and P removal. Hence, batches should not be stored longer than about one to one and a half weeks and the S_{bsi} in the feed should be measured daily. Furthermore, by monitoring the S_{bsi} on a routine basis, changes in S_{bsi} between batches are picked up that explain the "erratic" P removal behaviour noted earlier between different batches of sewage. Also, it is vitally important that the tanker transporting the batch to the laboratory, and the cold storage tank, are thoroughly cleaned between batches otherwise hydrogen sulphide generation and rapid loss of S_{bsi} will be encountered. High hydrogen sulphide concentration in the feed to a process can have adverse effects on the process response.

3.4 Comments on the parametric model and recent research developments

The parametric model described above was developed from observed data on experimental processes operated over a wide range of conditions. The conditions ranged as follows:

(i) Influent COD concentration : 250 – 800 mgCOD/ℓ

(ii) Readily biodegradable COD : 70 – 220 mgCOD/ℓ
i.e. fraction f_{ts} : 0,12 – 0,27

(iii) TKN/COD ratio : 0,09 – 0,14

(iv) Sludge age : 13 and 25 days

(v) Temperature : 12°C and 20°C

Development of the model required the identification of the principal factors and conditions that appeared to influence the excess P removal and integration of these into a consistent quantitative expression for estimating the excess P removal. Summarizing, the main factors identified are:

(i) As the concentration of readily biodegradable COD in the influent increased, so the excess P removal did likewise.

(ii) As the anaerobic sludge mass fraction increased so the excess P removal did likewise.

(iii) For identical conditions for (i) and (ii) above, as the sludge age *decreased*, so the excess P removal *increased*.

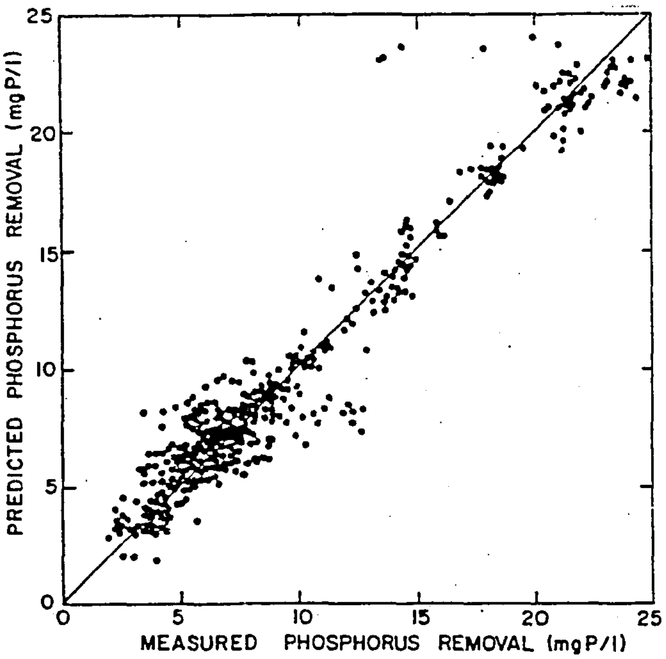

FIGURE 7.12: Comparison of measured P removals with those predicted by the parametric model for all the data observed under the different operating conditions.

(iv) Temperature reduction appeared to have a slight positive effect on excess P removal.

The above tendencies formed the basis for structuring the formulations for estimating the excess P removal and the equations so derived were calibrated against the observed data (see Section 3.3 above). A comparison of the theoretically predicted and experimentally measured process P removal data for the conditions set out above is shown in Figure 7.12: clearly a good correlation is obtained.

*It is of the greatest importance to note that the model is semi-empirical – **its use must be limited strictly to within the ranges of process parameters and wastewater characteristics listed above.** In particular, the model should not be used for design of plants with sludge ages less than about 14 days; in South Africa no research has been done on the process at short sludge ages and there is no certitude that the model predicts the behaviour adequately.*

Despite the good correlation, the parametric model is unsatisfactory from a scientific point of view because it is independent of any formal hypothesis on the biological mechanisms driving the process. Consequently the model should serve as an interim measure only for design and other practical purposes. What is required is a model based on fundamental biochemical and physical rules that will give rise to similar behavioural characteristics that formed the basis of the parametric model.

Historically important contributions towards developing such a model were made by Fuhs and Min Chen (1975) and the Johannesburg group (Nicholls, 1978). The latter group in particular, inaugurated a new approach which has influenced more recent hypothesized mechanisms proposed by Rensink (1981) and Marais *et al.* (1982). Preliminary indications are that a model based on the behaviour proposed by Rensink and Marais *et al.* has the potential to explain most of the behavioural characteristics of excess P removal and should in time allow a kinetic description to be incorporated into the general activated sludge process theory.

In essence, the basic concepts governing their model are as follows: Poly-P organisms, such as *Acinetobacter*, store phosphate internally as poly phosphate chains under aerobic conditions – some strains can do so also under anoxic conditions with nitrate serving as the electron acceptor. When poly-P organisms are discharged to an anaerobic reactor which receives the wastewater influent flow, the organisms survive the anaerobic conditions by breaking down some of the accumulated poly-P chains and the energy thereby released is utilized to complex the lower fatty acids surrounding the organisms to higher forms, such as aceto-acetate or polyhydroxybutyrate and store these compounds internally. When the organisms thereafter enter an aerobic (or anoxic)* zone, they utilize the complexed compounds with oxygen (or nitrate)* as the electron acceptor for growth and replenishment of the poly-P stores.

If lower fatty acids were to be discharged to the *aerobic* (or *anoxic*) zones, the poly-P organisms would have to compete with the heterotrophs (or facultative heterotrophs) for this substrate because now it is available to these other organisms due to the presence of oxygen (or nitrate) as electron acceptor. However, because the poly-P organisms are relatively slow growing compared to the other organisms, they will obtain only a minor fraction of the lower fatty acids in the aerobic reactor and consequently their growth will be very small. (This explains the reason why poly-P organisms are rarely found in considerable numbers in purely aerobic or anoxic-aerobic systems). Hence, by incorporating an anaerobic zone (which receives no nitrate from return recycle and to which the influent flow is discharged), the growth of the poly-P organisms is favoured, by allowing the organisms the opportunity to sequester the lower fatty acids by using their poly-P stores. In the anaerobic zone the other organisms are inoperative due to the unavailability of external electron acceptors.

With regard to the source of lower fatty acids for growth of the poly-P organisms this is present in the influent in the readily biodegradable COD fraction. However, not all the readily biodegradable COD is likely to be in a lower fatty acid form. Under these circumstances the behaviour of the facultative

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* Florentz and Granger (1982), using P^{31} nuclear magnetic resonance techniques, present evidence showing polyphosphate accumulation under anoxic conditions with nitrate serving as the electron acceptor.

heterotrophs may well be advantageous to the poly-P organisms. In the anaerobic reactor the facultative organisms can derive a small amount of energy (principally for survival) by generating internal to itself an electron acceptor to break down the higher forms of readily biodegradable material (like glucose) to lower fatty acids, via the Embden-Meyerhof pathway.

Under anaerobic conditions these acids cannot enter the Krebs cycle because of the lack of an external electron acceptor, and the lower fatty acids formed by the facultative heterotrophs are released to the bulk liquid. In this way the poly-P organisms still can sequester lower fatty acids even though none might be available in the influent. In this fashion a slow growing organism like *Acinetobacter* can flourish in an anaerobic-anoxic-aerobic mixed culture system even though it may be an obligate aerobe. Apparently within the time scale of normal anaerobic zones, i.e. actual retention times up to 2 hours, a significant reduction of slowly biodegradable particulate COD to lower fatty acids is unlikely so that where lower fatty acid production does take place in the anaerobic reactor, this is principally from the readily biodegradable COD fraction.

In terms of the biochemical model the P released in the anaerobic zone is due to the utilization of poly-P reserves to provide energy for complexing the lower fatty acids. From stoichiometry (see Marais *et al.* 1982) for every P released there is an approximately constant mass of lower fatty acid complexed – the mass of poly-P organisms that will be formed depends on the mass of lower fatty acid complexed. The excess P removal attainable depends on the mass of poly-P organisms formed* which in turn depends on the proportion of mass of lower fatty acids and readily biodegradable COD in the influent that is complexed by the poly-P organisms.

Research into P release under batch anaerobic conditions using *acetate* as lower fatty acid (Law, Dold and Marais, 1982) has supported the above conclusions that there is an approximate proportionality between the mass of P released and the mass of acetate complexed – about 1 mgP/ℓ phosphate to 2 mgCOD/ℓ acetate. The rate of acetate complexing is very fast. Now in a normal municipal wastewater the lower fatty acid fraction of the readily biodegradable COD is likely to be low. Here the presence of the facultative heterotrophs may be important for the generation of the lower acids. However, in this case the poly-P organisms can complex the lower fatty acids only as fast as these are produced by the facultative heterotrophs: batch test experiments where P release was stimulated using municipal wastewater, indicated that the rate of lower fatty acid complexing, or equivalently the rate of P release, is about 5 times slower than that with acetate. Furthermore, these tests support the earlier conclusion that only the readily biodegradable COD is reduced to lower fatty acids for complexing, by the consistent way in which the mass of P released follows the mass of readily biodegradable COD input to the batch test even with a relatively prolonged anaerobic period (about 3 hours).

In order to develop a greater understanding of the kinetics of readily biodegradable COD reduction to lower fatty acids and P release, a Modified UCT process with a plug flow anaerobic reactor treating municipal wastewater was operated. It was found that the mass of phosphorus released per unit elemental length of plug flow reactor decreased exponentially along the length of the reactor. This means that the rate of readily biodegradable COD reduction and complexing is first order with respect to the readily biodegradable COD concentration. Using this first order approach, an elementary steady state kinetic model has been developed. This model indicates that because the readily biodegradable COD is reduced and complexed according to a first order rate, greater efficiency would be derived by subdividing the anaerobic reactor into a

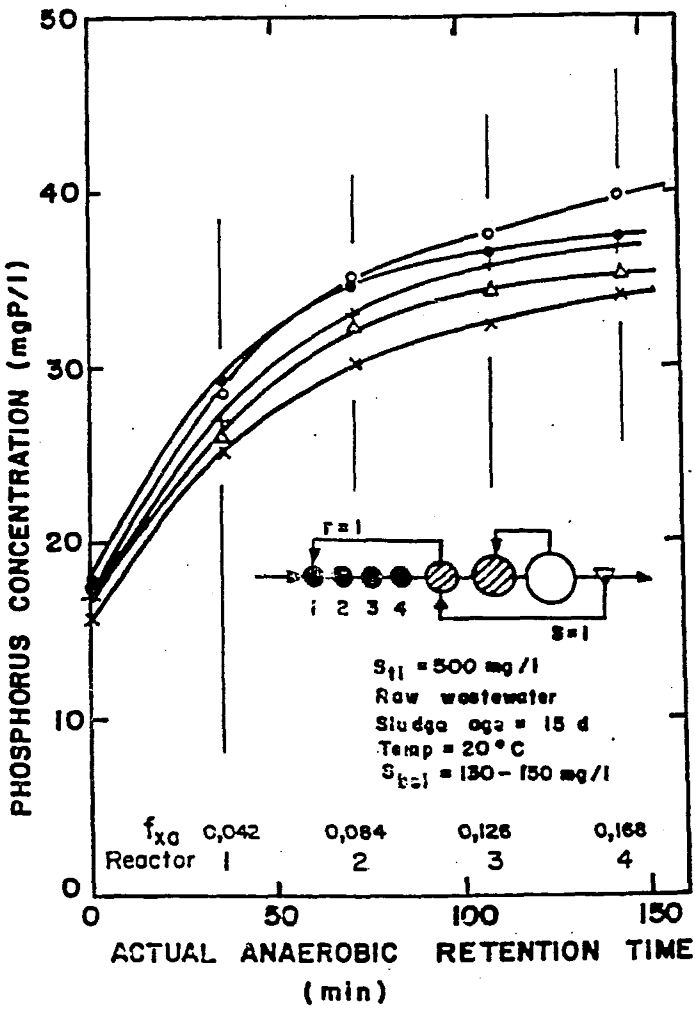

FIGURE 7.13a: Phosphorus concentration profile in a 4 in-series anaerobic reactor in a modified UCT process. S_{ti} = 500 mgCOD/ℓ raw wastewater, $S_{bsi} \cong$ 150 mgCOD/ℓ, sludge age 15 days, r-recycle ratio 1:1, temperature 20°C, MLVSS in anaerobic reactions 1450 mgVSS/ℓ.

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*At steady state, the nett number of poly-P organisms that form is equal to the number removed in the daily waste sludge. In essence excess biological P removal is the continuous formation of poly-P organisms and their removal from the process at a stage when they are filled with poly-P.

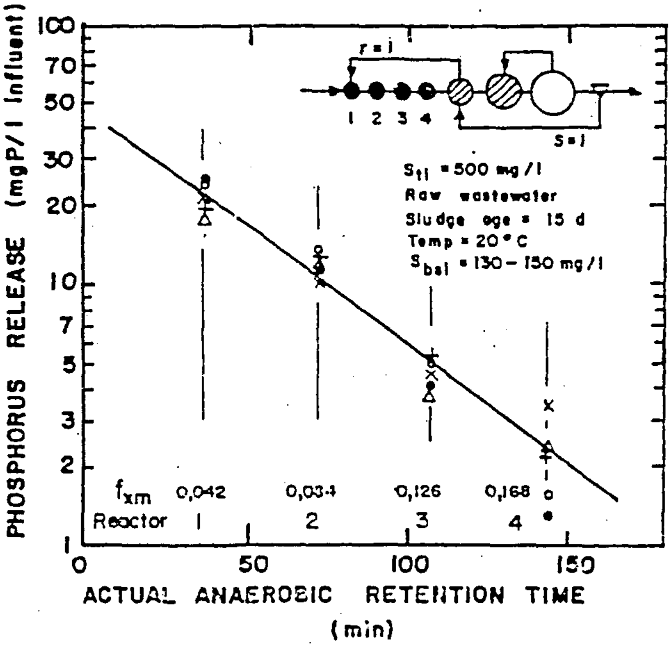

FIGURE 7.13b: Phosphorus release per reactor in 4 in-series anaerobic reactor. Data from Figure 7.13a.

number of compartments and operated in series. Experiments with continuous Modified UCT processes in which the anaerobic reactor was subdivided into 4 equal compartments, corroborated the first order approach (Figure 7.13) – for the same anaerobic sludge mass fraction and readily biodegradable COD in the influent, the mass of P release in the anaerobic reactor was greater for the series reactor anaerobic system than for the single reactor anaerobic system; the P removal too was proportionally higher in the former than in the latter.

On the strength of the results discussed above it would appear that the anaerobic reactor should be subdivided into 2 or more compartments in series to achieve improved excess P removals for a fixed anaerobic sludge mass fraction. This subdivision also allows greater process flexibility in that not only can the anaerobic sludge mass fraction be changed but also allows different nutrient removal processes to be incorporated on the same plant (see Section 5 below).

Comparing the results of the experiments under batch conditions and under continuous conditions with plug flow and multi-compartment anaerobic reactors with those predicted by the parametric model, it appeared that there is an inconsistency. The results showed that P release is obtained for acetate and readily biodegradable COD concentrations below 25 mgCOD/ℓ whereas in the continuous experiments on which the parametric model is based this was apparently not so. The causes for the difference can be ascribed to (1) measurement procedure and (2) kinetic behaviour.

(i) Because the P concentration in the influent consists of two fractions i.e. a large soluble (80%) and a small particulate (~20%), an error is made when calculating the magnitude of the P release in the anaerobic reactor. In calculating the magnitude of the P release a mass balance on P across the anaerobic reactor is made and this is done by assuming that all the P entering and leaving the reactor is in a soluble form, even that of the influent. The limitation of this method of measuring the P release was realized and is probably the principal reason why in the parametric model it was found that the P content of the active mass (γ) once anaerobic conditions were imposed is 0,06 instead of the more usual value of 0,03 for purely aerobic processes i.e. P release probably did take place for $S_{bsa} < 25$ mgCOD/ℓ and this stimulated the additional P uptake from 0,03 to 0,06.

(ii) The parametric model indicates that apparently a minimum of 25 mgCOD/ℓ readily biodegradable COD in the anaerobic reactor is needed to induce excess P removal. A reason for this behaviour now can be advanced in terms of the kinetics describing the rate of P release observed on the series anaerobic reactors. A model has been constructed incorporating the kinetic P release hypothesis and all the data observed to date processed. The model predicts P uptake that is not as good as that obtained by the parametric model; evidently it requires further development. In consequence the model will not be described here. However, on one aspect of the model predictions deserve mention. Evidently for sludge ages of 20 days and anaerobic mass fraction of 0,15 influent COD of raw sewage 500 mg/ℓ and $S_{bsi} = 100$ mg/ℓ (the "normal" conditions tested in the experimental investigation) in the anaerobic reactor only about 70 – 80 per cent of the readily biodegradable COD is sequestered i.e. ± 20 to 30 mg/ℓ COD is discharged from the anaerobic reactor – this would explain why (S_{bs}-25) appears in the parametric model. If however the anaerobic reactor is subdivided into 2 or more reactors the S_{bs} in the effluent from such a series reduces to 5 to 10 mg/ℓ. The effluent S_{bs} furthermore depends on the influent S_{bs} so that it would seem the factor (S_{bs}-25) is only an approximation.

The discrepancies discussed above are not crucial to the predictive power of the parametric model – a series of calculations using the kinetic and parametric approaches show that significant differences arise only for low strength COD wastewaters, and here the parametric model tends to predict lower removals than the kinetic model. However, the kinetic model is not yet verified or developed sufficiently to replace the parametric model.

The above discussion is not intended to cast doubt on the parametric model and its usefulness; that can clearly be seen in Figures 7.10, 7.11 and 7.12. The parametric model evolved out of an intensive research programme to develop an understanding of and design criteria for biological nutrient removal processes with little emphasis on kinetics. In the future, it is very likely that it will be superseded by a more elegant kinetic model without necessarily leading to very much better predictions. However, for the present, the parametric model is probably the best and most exhaustively tested model available and can be of great value for analysing biological P removal process behaviour as it is understood at present. The remaining part of this Chapter will demonstrate the use of the parametric model for design.

4. IMPLICATIONS OF THE MODEL ON DESIGN

The parametric model for excess biological P removal consists essentially of two concepts i.e. (i) the coefficient of excess P removal (γ) and (ii) the excess P removal propensity factor (P_f), which are linked by an empirical relationship, Eq. 7.7.

4.1 Coefficient of excess P removal (γ)

When excess P removal takes place, this will be reflected in the magnitude of the coefficient of excess P removal γ in Eq. 7.6, the value being greater than 0,02 to 0,03 mgP/mgVASS usually found for purely aerobic processes. With excess P uptake found in processes incorporating anaerobic reactors, γ can range from a minimum of about 0,06 to a maximum of about 0,35 when the appropriate anaerobic conditions are present.

A plot of Eq. 7.6 (divided through by S_{ti}) versus sludge age (R_s) is given in Figure 7.14 for raw and settled wastewater (see Tables 4.3, 5.2 and 6.1) at 14°C and 22°C assuming a constant γ value of 0,18 mgP/mgVASS, a value readily attained in appropriately designed P removal plants. Figure 7.14 shows that the mass of P removed per mgCOD in the influent, P_s/S_{ti}, decreases sharply with sludge age so that *to remove the maximum P, the sludge age should be kept as low as possible*. The P removal per unit COD load is approximately the same for raw and settled sewage. Hence, because primary sedimentation removes as much as 40% of the COD load, P removal with settled wastewater is about 40% less than with raw wastewater. For a constant γ the effect of temperature on removal is small because the nett mass of sludge produced does not change significantly with temperature.

4.2 Excess P removal propensity factor (P_f)

The magnitude of the coefficient of excess P removal γ is influenced by a number of conditions in the anaerobic reactor: (1) the magnitude of the readily biodegradable COD (S_{bsa}) in excess of 25 mgCOD/ℓ, (ii) the actual retention time and (iii) the mass of sludge (expressed as a fraction of the mass of sludge in the process) passing through the anaerobic reactor each day. These conditions were quantified for the UCT and Phoredox processes and formalized into the parameter called the excess P removal propensity factor (P_f) given by Eq. 7.5. The readily biodegradable COD concentration in the anaerobic reactor (S_{bsa}) for the UCT and Phoredox processes can be calculated from Eqs. 7.1 and 7.2. These predictions are based on the model derived from observations on processes with sludge ages between 13 and 25 days. Where the sludge ages are outside the limits, in particular where the sludge ages are less than 13 days, it would be unwise to extrapolate the model to these conditions.

4.3 Excess P removal model

The coefficient of excess P removal (γ) and the propensity factor (P_f) described above were linked with an empirical relationship i.e. Eq. 7.7 found from data given in Figure 7.9. Equation 7.7 shows that the higher the P_f

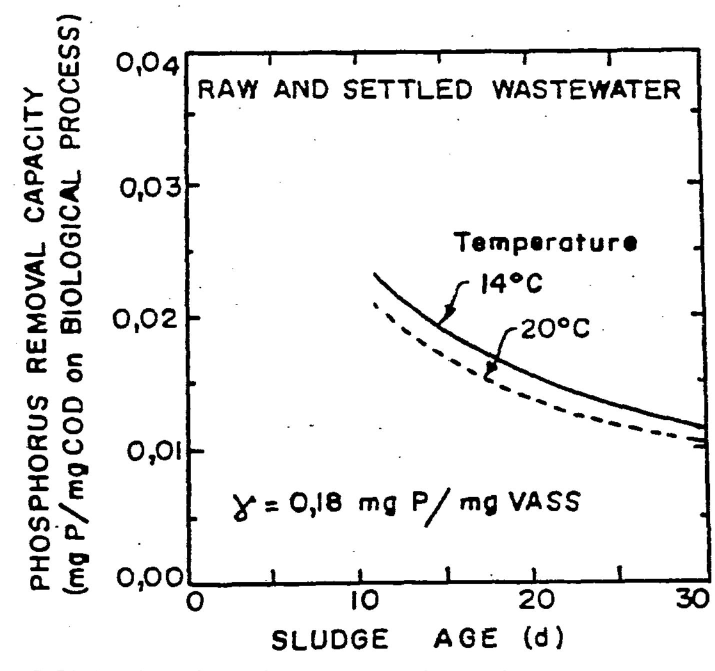

FIGURE 7.14: Phosphorus removal per kg COD load on biological process versus sludge age for an excess P removal coefficient (γ) of 0,18 mgP/mgVASS at 14°C and 20°C for raw and settled wastewaters.

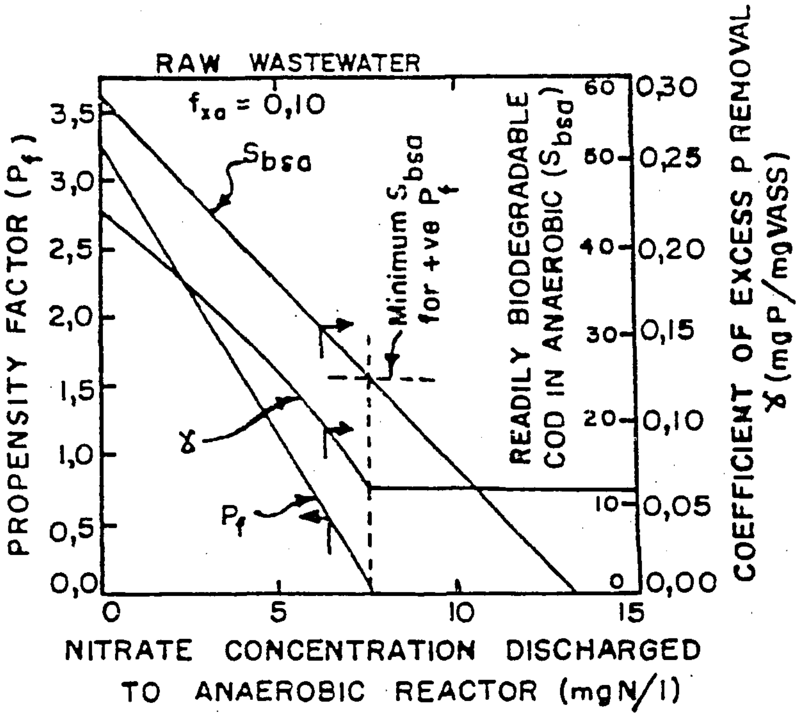

FIGURE 7.15: Propensity factor (P_f), rapidly biodegradable COD concentration in anaerobic reactor (S_{bsa}) and coefficient of excess P removal (γ) versus the nitrate concentration discharged to the anaerobic reactor for an anaerobic sludge mass fraction (f_{xa}) of 0,10 for raw wastewater.

factor, the greater the γ coefficient. Consequently, at a fixed sludge age, the higher the P_f factor in a process, the greater the P removal per unit COD load (P_s/S_{ti}). For a fixed f_{xa}, high P_f factors are obtained with (i) high influent COD concentrations (for normal f_{bs} fractions in the wastewater of about 20%), (ii) high readily biodegradable COD fractions (f_{bs}), and (iii) a zero nitrate concentration in the recycles to the anaerobic reactor (N_{nr} and $N_{ns} = 0$). The sensitivity of excess P removal to nitrate discharged to the anaerobic reactor is demonstrated in Figure 7.15. In Figure 7.15, the S_{bsa} concentration, P_f factor and γ coefficient calculated from Eqs. 7.1, 7.2, 7.5 to 7.7 for the raw wastewater characteristics given in Tables 4.3, 5.2 and 6.1 are plotted versus the nitrate concentration discharged to the anaerobic reactor for an anaerobic sludge mass fraction $f_{xa} = 0{,}10$, underflow recycle ratio s = 1, DO in s-recycle $O_s = 1{,}0$ mgO/ℓ and DO in influent $O_i = 0{,}0$ mgO/ℓ. It can be seen that for nitrate concentrations > 7 mg/ℓ, the γ coefficient ia reduced to one third of the value for a zero nitrate concentration; at 25-day sludge age and 14°C, this reduction in γ from 0,22 to 0,07 mgP/mgVASS, results in a decline in P removal from 9,3 to 3,5 mgP/ℓ. This example demonstrates that for good P removal, nitrate discharge to the anaerobic reactor should be avoided.

5. DESIGN PROCEDURE*

When selecting a Phoredox or UCT process for excess P removal, it is necessary to establish whether or not near complete denitrification can be achieved for the selected sludge age (R_s) and maximum specific growth rate of the nitrifiers at 20°C (μ_{nm20}). If near complete denitrification can be achieved for sludge ages less than about 25 days and total anoxic sludge mass fractions (f_{xdt}) less than about 0,40 at the minimum temperature (T_{min}), then a Phoredox process is indicated; if it cannot be achieved, then a UCT process is indicated.

For the P removal processes, an anaerobic reactor is required for the stimulation of excess P removal, so that the maximum *anoxic* sludge mass fraction available for denitrification (f_{xdm}), is the difference between the maximum unaerated sludge mass fraction (f_{xm}) and the selected anaerobic sludge mass fraction (f_{xa}) i.e.

$$f_{xdm} = f_{xm} - f_{xa} \tag{7.8}$$

where f_{xm} is given by Eq. 5.24 for a selected R_s, μ_{nm20}, S_f and T_{min}.

The denitrification behaviour of the Phoredox process is identical to that of the Bardenpho process, i.e. Eqs. 6.24 to 6.29 apply. Note however, that f_{xdm} for the Phoredox process is given by Eq. 7.8 whereas f_{xdm} for

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*The design procedure described in this section has been published in a shorter form by Ekama, Siebritz and Marais (1982).

the Bardenpho process is given by Eq. 6.23. Hence for the same unaerated sludge mass fraction f_{xm}, the Phoredox process has a lower f_{xdm} than the Bardenpho process, by an amount equal to f_{xa}. Now from the behaviour of the Bardenpho process (see Chapter 6, Section 7), it was shown that when the wastewater characteristics are favourable, in particular, a high readily biodegradable COD fraction (f_{bs}), a high μ_{nm20} value and a low TKN/COD ratio, near complete denitrification can be achieved at sludge ages around 20 to 25 days (see Figure 6.17). Consequently for favourable wastewater characteristics near complete denitrification, and hence excess P removal can be achieved in the Phoredox process (because the effluent nitrate concentration is low and hence little nitrate is discharged to the anaerobic reactor). The excess P removal attainable, P_s, is calculated with the aid of the parametric model described above and knowing P_s, the effluent P concentration P_{te} given by the difference between the influent P concentration P_{ti} and P_s i.e.

$$P_{te} = P_{ti} - P_s \text{ (mg P/}\ell\text{)} \tag{7.9}$$

In the design of a Phoredox process for a selected sludge age, if it is found that the total anoxic sludge mass fraction for complete denitrification f_{xdt} is less than the maximum allowed f_{xdm}, the following options are open:

(1) f_{x1} and f_{x3} can be increased to give a total anoxic sludge mass fraction equal to f_{xdm} which introduces a factor of safety for denitrification, the magnitude of which is equal to f_{xdm}/f_{xdt}.

(2) The sludge age can be reduced so that f_{xdm} becomes equal to f_{xdt}. The lower sludge age will allow a saving in process volume (see Figure 4.3). This lower sludge age R_s, can be estimated as follows: set f_{xdm} equal to f_{xdt} required for complete denitrification at T_{min}; determine f_{xm} from Eq. 7.8; with f_{xm} calculate R_s from Eq. 5.24; to check repeat step by step procedure for the new R_s.

(3) For the Phoredox process, f_{xa} can be increased to improve excess P removal, so that $f_{xdm} = f_{xdt}$ from Eq. 7.8.

Options (2) and (3) above are not recommended as these exclude a factor of safety in denitrification: Exclusion of a factor of safety on denitrification for the Phoredox process may have serious consequences because the discharge of *any* nitrate to the anaerobic reactor reduces the P removal attainable in the process (Figure 7.15).

For design purposes where the Phoredox process is to operate between temperatures T_{min} and T_{max}, the following sequence of calculations needs to be made to check whether or not complete denitrification can be achieved in a Phoredox process and, if so, to select the optimal process configuration for N and P removal:

Step 1: Select wastewater characteristics; S_{ti}, N_{ti}, P_{ti}, f_{bs}, f_{up}, f_{us}, μ_{nm20}, T_{max}, T_{min}.

Step 2: Select R_s and S_f.

Step 3: Calculate f_{xm} for T_{min} from Eq. 5.24.

Step 4: With f_{xm} and R_s calculate S_f for T_{max} from Eq. 5.24.

Step 5: Estimate N_{te} for T_{max} and T_{min}.

Step 6: With f_{up} and f_{us}, calculate N_s for T_{max} and T_{min} for selected R_s from Eq 4.23.

Step 7: Calculate N_c from Eq. 5.29

Step 8: Select f_{xa}.

Step 9: Calculate D_{pp} from Eq. 6.25 for T_{max} and for T_{min} with f_{xdm} given by Eq. 7.8.

Step 10: Select s, O_a and O_s.

Step 11: Calculate f_{x1min} from Eq. 6.21 for T_{max} and for T_{min}.

Step 12: Calculate f_{x3min} from Eq. 6.29 for T_{max} and for T_{min}.

Step 13: Select the a-recycle ratio and calculate N_{ne} for T_{min} from Eq. 6.24.

Step 14: If $N_{ne} < 0$, set $N_{ne} = 0$.

Step 15: With N_{ne} calculate f_{x1}, f_{x3} and f_{xdt} from Eqs. 6.26 to 6.28.

Step 16: Check that $f_{x1} \geq f_{x1min}$ and $f_{x3} \geq f_{x3min}$. If not, discard selected a-recycle ratio as invalid.

Step 17: Repeat steps 12 to 15 for different a-recycle ratios.

Step 18: Repeat steps 12 to 16 for T_{max}.

Step 19: Fix f_{x1} and f_{x3} by procedure given in Chapter 6, Section 6.2. This fixes N_{ne} and hence N_{ns}.

Step 20: Select O_i.

Step 21: Calculate ΔS_{bsa} from Eq. 7.2b.

Step 22: Calculate S_{bsa} from Eq. 7.1b.

Step 23: Calculate P_f from Eq. 7.5.

Step 24: Calculate γ from Eq. 7.7

Step 25: Calculate P_s from Eq. 7.6.

Step 26: Calculate P_{te} from Eq. 7.9.

Step 27: Is P_{te} sufficiently low? If not, decrease R_s or increase f_{xa} at expense of f_{xdm}; however, improvement will only be achieved if $f_{xdt} < f_{xdm}$ i.e. $N_{ne} < 0{,}0$; if $f_{xdt} = f_{xdm}$ i.e. $N_{ne} > 0$, changes will increase P_{te}.

Step 28: Repeat calculations for T_{max}.

Step 29: Can required P_{te} be achieved with the Phoredox process? If not, select a UCT process.

When the required P removal cannot be attained in the Phoredox process due to incomplete denitrification, the UCT or Modified UCT processes are the indicated processes.* The configuration of the UCT processes is such that the conditions in the anaerobic reactor are independent of the denitrification behaviour for TKN/COD ratios up to about 0,13. Hence in their design, the P removal can be estimated from the conditions in the anaerobic reactor without the need to know the extent of denitrification i.e. N and P removal can be dealt with separately in the design of these processes.

In the design for excess P removal in the UCT process, the readily biodegradable COD concentration in the anaerobic reactor (S_{bsa}) can be found directly from Eq. 7.1a and 7.2a without the need for selecting R_s or μ_{nm20}: S_{bsa} depends only on the influent COD strength (S_{ti}), the readily biodegradable COD fraction (f_{bs}) and the r-recycle ratio from the anoxic to the anaerobic reactor. The r-recycle ratio usually is selected at 1:1; higher ratios lead to greater dilution of S_{bsa} with a concomitant reduction in P removal, and lower values lead to inordinately large anaerobic reactor *volume* fractions with respect to the anaerobic sludge *mass* fractions (see Eqs. 7.10 to 7.11). Selecting $r = 1$ and the nitrate (N_{nr}) and DO (O_r) concentrations in the r-recycle of say 1 mgN/ℓ and 1 mgO/ℓ to include a factor of safety for excess P removal, then S_{bsa} can be calculated from Eqs. 7.1a and 7.2a and will remain unchanged irrespective of the temperature. Knowing S_{bsa}, the propensity factor P_f can be calculated from Eq. 7.5 for a selected anaerobic sludge mass fraction f_{xa}, and from the P_f value, the γ coefficient can be calculated from Eq. 7.7. Knowing γ, the P removal from the particular wastewater can be calculated from Eq. 7.6 for a selected sludge age (R_s). The P removal should be determined at T_{max} because, for a constant S_{bsa} or γ, the P removal at T_{max} is less than at T_{min} (Figure 7.14). Hence, for various values of f_{xa}, the P removal at different sludge ages can be calculated, and estimates of f_{xa} and R_s giving the desired or optimal P removal can be found. In selecting f_{xa}, it is recommended that f_{xa} should not be less than 0,10 and not greater than 0,25: for $f_{xa} < 0,10$, the P removal usually will be inadequate even at relatively short sludge ages for normal wastewaters, particularly for settled wastewaters; for $f_{xa} > 0,25$ no UCT experimental response data are available so that the parametric model has not been calibrated to include the process behaviour for $f_{xa} > 0,25$.** At $f_{xa} = 0,25$, the process has been found to operate satisfactorily (Siebritz *et al.* 1983). With regard to selecting R_s, cognizance must be taken of its effect on the maximum unaerated sludge mass fraction f_{xm} (Eq. 5.24) because, with f_{xa} fixed, f_{xm} fixes the maximum anoxic sludge mass fraction f_{xdm} (Eq. 7.8) which, in turn, governs the degree of denitrification that can be achieved.

Once f_{xa} and R_s are selected, the denitrification can be calculated by applying the design equations for N removal in the MLE process, except that f_{xdm} is given by the difference between f_{xm} and f_{xa} (see Eq. 7.8).

For design purposes, where the process is to operate between temperatures T_{min} and T_{max}, the following sequence of calculations is recommended to facilitate selection of the optimal process configuration for N and P removal in the UCT process:

Step 1: Select wastewater characteristics; S_{ti}, N_{ti}, P_{ti}, f_{bs}, f_{up}, f_{us}, μ_{nm20}, T_{max}, T_{min}.

Step 2: Select r (usually 1,0) and N_{nr} and O_r (say 1 mg/ℓ each).

Step 3: Calculate S_{bsa} from Eqs. 7.1a and 7.2a.

Step 4: Select f_{xa} (between 0,10 and 0,25).

Step 5: Calculate P_f from Eq. 7.5 and γ from Eq. 7.7.

Step 6: Select R_s.

Step 7: Calculate P_s from Eq. 7.6 for T_{max} and for T_{min}.

Step 8: Calculate P_{te} from Eq. 7.9 for T_{max} and for T_{min}.

Step 9: Is P_{te} adequate? If not, repeat steps 4 to 8 for different f_{xa} and/or R_s (see Design Chart Figure 7.18).

Step 10: Select S_f.

Step 11: Calculate f_{xm} for T_{min} from Eq. 5.24.

Step 12: With f_{xm} and R_s calculate S_f for T_{max} from Eq. 5.24.

Step 13: Estimate N_{te} for T_{max} and T_{min}.

Step 14: With f_{up} and f_{us}, calculate N_s for T_{max} and T_{min} for selected R_s from Eq. 4.23.

Step 15: Calculate N_c from Eq. 5.29.

Step 16: Calculate D_{pp} from Eq. 6.25 for T_{max} and for T_{min} with f_{xdm} given by Eq. 7.8.

Step 17: Select s, O_a and O_s.

Step 18: Calculate a_o from Eq. 6.30 and N_{ne} from Eq. 6.31.

Step 19: Are P_{te} and N_{ne} adequate? If P_{te} is too high, increase f_{xa} or reduce R_s – these changes will increase N_{ne}; if P_{te} is too low, decrease f_{xa} or increase R_s – these changes will decrease N_{ne}.

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*It is possible to design a Modified Phoredox/Modified UCT/UCT process combination (see later in this section).

**Laboratory and pilot scale data on process behaviour for $f_{xa} > 0,25$ are available from the NIWR (see Interim Design Guide by NIWR).

Step 20: Repeat from steps 4 to 17 until required or optimal P_{te} and N_{ne} is obtained.

Having completed the design for the UCT process, if the influent TKN/COD ratio is less than about 0,11 to 0,12 mgN/mgCOD, the process may be converted to a modified UCT process (Figure 7.6). The conversion is done by subdividing the *anoxic* sludge mass fraction (f_{xdm}) into two sub-fractions; the first is usually allocated a sludge mass fraction f_{xd1} of about 0,10, and the second, f_{xd2}, having the remaining sludge mass fraction i.e. $f_{xd2} = f_{xdm} - f_{xd1}$. Having made this subdivision, the possibility of operating the modified UCT process (Figure 7.6) as a UCT process (Figure 7.5) is not excluded; by making provision that the r-recycle flow to the anaerobic reactor can be taken from either the first or the second anoxic reactors, the process can be operated as either a modified UCT or a UCT process (Figure 7.16). The advantages of operating the process as a modified UCT process were set out in Section 2.2.3 above (see also Chapter 3, Section 3.3.3).

It is necessary to check whether or not the modified UCT/UCT process combination can be operated as a modified UCT process. For successful operation of the modified process, the denitrification potential of the first anoxic reactor must be greater than, or at least equal to the equivalent nitrate load on the reactor at the *minimum expected temperature (T_{min})*. The denitrification potential is calculated by substituting $f_{xd1} = 0,10$ for f_{x1} in Eq. 6.20. The equivalent nitrate load is found from the effluent nitrate concentration (N_{ne}) from the UCT process, the s-recycle ratio and the DO concentration in the s-recycle (O_s) i.e. the equivalent nitrate load is $s(N_{ne} + O_s/2,86)$. For TKN/COD ratios $< 0,12$ mgN/mgCOD, it will be found that the denitrification potential is sufficient to denitrify the equivalent nitrate load with a reasonable factor of safety (S_{fd}) of say 1,3, where S_{fd} is the ratio of the denitrification potential and the equivalent nitrate load on the first anoxic reactor.

By setting $f_{xd1} = 0,10$, usually it will be found that the reactor is sufficiently large for complete utilization of the influent readily biodegradable COD (S_{bsi}). This can be checked by means of Eq. 6.21. If T_{min} is less than 14°C, it may be found from Eq. 6.21 that 0,10 is insufficient for complete utilization of the S_{bsi}. In this event, f_{xd1} can be increased slightly to a maximum of 0,12. Values of $f_{xd1} > 0,12$ are not recommended because for temperatures $> T_{min}$, the reactor will be over-designed leading to a reduced nitrogen removal efficiency of the process. By designing the first anoxic reactor (f_{xd1}) to allow the possibility of complete or near complete utilization of the S_{bsi}, the process can achieve excess P removal up to the TKN/COD ratio at which S_{bsi} is completely utilized. This TKN/COD ratio can be calculated by trial and error by testing increasing TKN/COD ratios on the process.

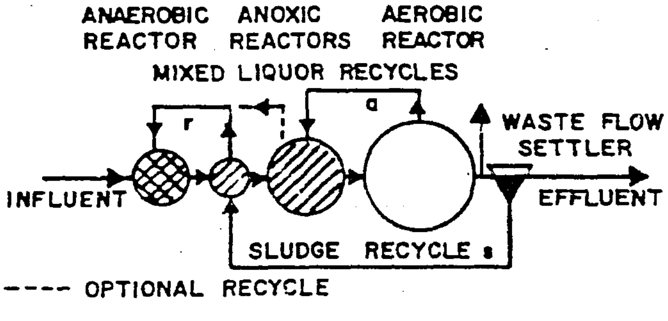

FIGURE 7.16: The modified UCT/UCT process combination for biological P and N removal.

Generally it will be found that the higher the TKN/COD ratio, the lower the factor of safety S_{fd} for complete denitrification of the equivalent nitrate load on the first anoxic reactor. When $S_{fd} < 1,1$, the prospect of operating the modified UCT/UCT process combination (Figure 7.16) as a modified UCT process has little merit. This limit is strongly dependent on the influent readily biodegradable COD fraction f_{bs}: For normal f_{bs} ratios in raw wastewaters of 0,20 to 0,25, it will be found that S_{fd} is less than 1,1 for TKN/COD ratios above 0,11 to 0,12 mgN/mgCOD at a minimum temperature of 14°C. When the f_{bs} fraction and TKN/COD ratio of the influent are such that operation of the modified UCT process is not possible then it must be operated as a UCT process. In the UCT process, if the a-recycle ratio is too low to produce an actual anoxic retention time lower than 1,5 hours, poor sludge settling characteristics may be obtained, and should be taken into account in the design of the secondary settling tanks (see Chapter 8).

When the influent readily biodegradable COD fraction and TKN/COD ratio are such that operation of a modified UCT process is possible, the mixed liquor a-recycle ratio is set at the greater value of two lower limits i.e. (1) that which loads the second anoxic reactor to its denitrification potential or (2) that which will produce an actual anoxic retention time of less than 1 hour. The first limit is calculated with the aid of Eq. 6.30 except that D_{pp} is replaced by the denitrification potential of the *second* anoxic reactor, which can be calculated by substituting the anoxic sludge mass fraction of the second anoxic reactor f_{xd2} for f_{x1} in Eq 6.20. The second limit is found by calculating the volume of the second anoxic reactor as set out below (see Section 7.5 above) and with the specified s-recycle ratio and influent flow rate, finding the a-recycle ratio which would produce an actual retention time of 1 hour.

A problem with the design of a nutrient removal process is that at the design stage the wastewater characteristics, in particular the TKN/COD ratio and maximum specific growth rate of the nitrifiers at 20°C (μ_{nm20}), are not sufficiently accurately known to design an optimal nutrient removal process. With regard to the μ_{nm20} value little can be done (besides measuring it, van Haandel (1981) and Sehayek (1981) with Marais) other than selecting a conservative value. However, with regard to the TKN/COD ratio virtually complete process flexibility can be provided for, so that depending on the TKN/COD ratio, the process can be operated as a Modified Phoredox, Modified UCT or UCT process. A schematic diagram of a process with this flexibility is given in Figure 7.17 (incorporation of the 5 stage Phoredox is difficult due to the problem of having to convert the secondary anoxic reactor to an aerobic

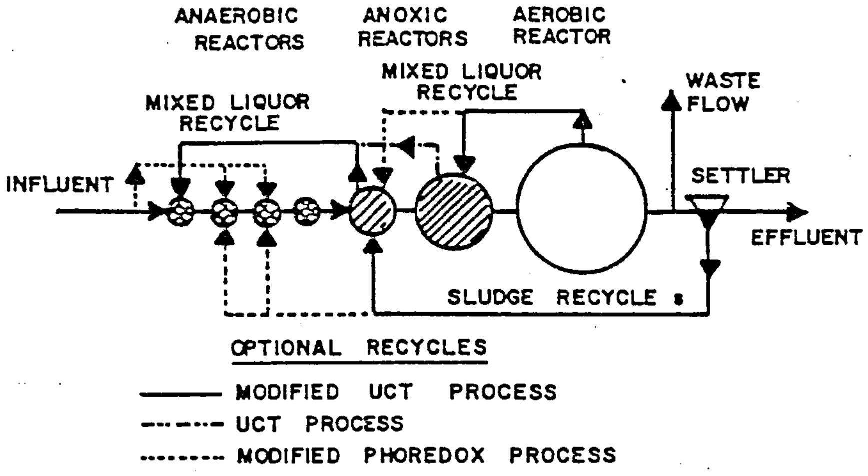

FIGURE 7.17: Schematic process configuration layout showing additional and optimal recycles and flows allowing operation as a Modified Phoredox, Modified UCT or UCT process.

reactor). Note in Figure 7.17 that the anaerobic sludge mass fraction is subdivided into 4 equal sized compartments in series. The reason for this is twofold:

(1) From a practical point of view, from the same anaerobic reactor volume, the UCT process has an anaerobic sludge mass fraction half that for the Modified Phoredox process. Consequently in the design, the anaerobic volume must be such that when the process is operated as a UCT process, the required anaerobic sludge mass fraction is obtained. If then the process is required to be operated as a Modified Phoredox, the first and second compartments can be bypassed thereby allowing the same anaerobic sludge mass fraction to be used for the Modified Phoredox process. It should be noted that if the anaerobic volume is not reduced when converting to a Modified Phoredox, the *aerobic* sludge mass fraction will be reduced, and this reduction may be detrimental for achieving nitrification, particularly at the lower temperatures. The equations for calculating the relationship between the volume fractions and sludge mass fractions for the different processes are given in Section 7.5 below.

(2) From a theoretical point of view, the kinetics of biological excess P removal are such that for the same total anaerobic sludge mass fraction subdivision into a number of equal sized completely mixed compartments in series is more efficient than a single completely mixed reactor (see Section 3.4 above). Theoretically the greater the number of compartments in series the better, but practically very little advantage is gained with more than 3. In the proposed composite configuration (Figure 7.17), the anaerobic zone comprises 4 compartments when operated as a UCT process and 2 when operated as a Modified Phoredox. With the Modified Phoredox consideration could be given to utilizing three of the four anaerobic compartments but caution should be exercised in not increasing the anaerobic sludge mass fraction too much because its increase is at the expense of the *aerobic* sludge mass fraction with the result that nitrification and/or phosphorus uptake may be adversely affected (see Chapter 10).

The design calculations for the Phoredox and UCT processes set out above will be demonstrated with the aid of worked examples after the presentation of a design chart for graphically estimating the excess P removal attainable.

6. DESIGN CHART

The excess biological P removal theory described in Section 3 above can be summarized in a design chart (see Figure 7.18). For a raw or settled wastewater, the chart is constructed as follows:

(i) In the bottom right hand quadrant, a graph of total influent COD concentration (S_{ti}) versus readily biodegradable COD concentration (S_{bsi}) for various readily biodegradable COD fractions (f_{bs} or f_{ts}) is plotted from Eqs. 2.8a or 2.9a.

(ii) In the top right hand quadrant the propensity factor P_f versus S_{bsi} is plotted for different anaerobic sludge mass fraction (f_{xa}) and fixed recycle ratio to the anaerobic (r or s) assuming say no nitrate or dissolved oxygen is discharged to the anaerobic reactor from Eqs. 7.1 or 7.2 and 7.5.

(iii) The left hand side of the P_f axis is rescaled in terms of γ by means of Eq. 7.7.

(iv) In the upper left hand quadrant a graph of P removal per unit COD (P_s/S_{ti}) versus γ is plotted for different sludge ages for fixed temperature from Eq. 7.6.

(v) In the bottom left hand quadrant, a graph of P_s versus S_{ti} is plotted for fixed removals of P in mg/ℓ.

The diagram constructed by the above procedure is valid for the selected wastewater characteristics f_{up} f_{us}. and temperature. An example of such a diagram is given in Figure 7.18 for the typical raw wastewater characteristics given in Table 4.3. In the construction of the diagram, it is assumed that no nitrate (or dissolved oxygen) is discharged to the anaerobic reactor. Consequently the chart is valid for (i) the Phoredox process, provided that complete denitrification can be achieved so that the nitrate concentration in the under flow recycle is zero; (ii) the UCT process, provided the mixed liquor a-recycle ratio is carefully controlled so that it does not overload the primary anoxic reactor with nitrate; and (iii) the modified UCT process, provided the effluent nitrate concentration is sufficiently low so that the underflow recycle does not overload the first anoxic reactor with nitrate.

The design chart is used as follows: For the raw wastewater characteristics given in Table 4.3 (i.e. $f_{up} = 0,13$, $f_{us} = 0,05$) and taking a temperature of 20°C, draw a horizontal line at the influent COD concentration S_{ti} of 600 mg COD/ℓ (line AB). Now the readily biode-

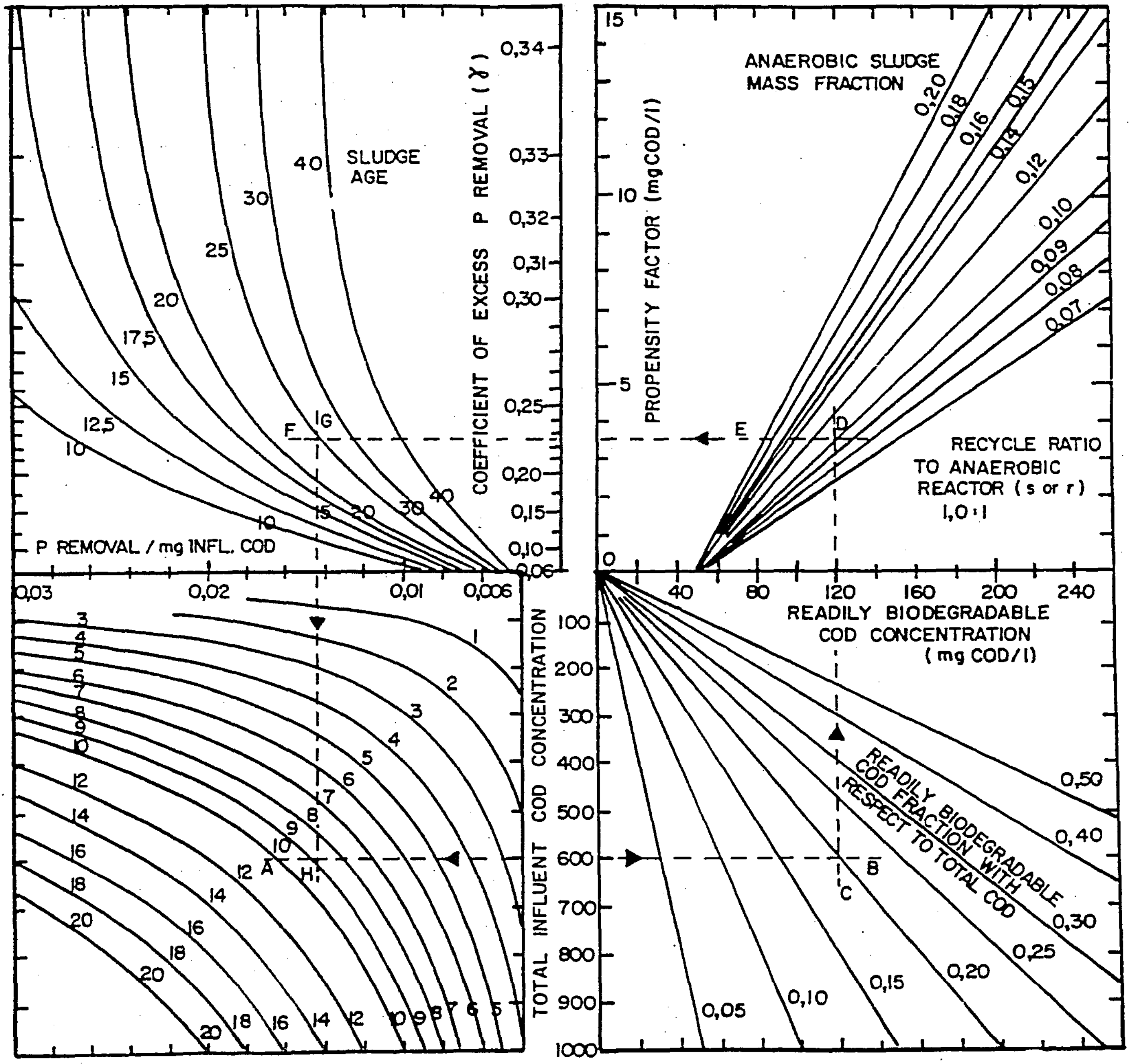

FIGURE 7.18: Design chart for graphically estimating the excess P removal attainable in nutrient removal processes at 14°C treating raw wastewater provided no nitrate is discharged to the anaerobic reactor.

gradable COD fraction with respect to total COD (f_{ts}) is 0,20 (i.e. $f_{bs} = 0,24$, the fraction with respect to the biodegradable COD).

Moving to the right: where line AB cuts the $f_{ts} = 0,20$ line, draw a vertical line (line CD). Moving upwards along line CD; where line CD cuts the selected anaerobic sludge mass fraction line (say 0,10), draw a horizontal line (line EF). Moving left along line EF: the intercepts with the P_f and γ axes give the values of these parameters. Where line EF cuts the selected sludge age line (say 25 days), draw a vertical line (line GH). Moving downwards along line GH: the intercept with the P_s/S_{ti} axis gives the value of this parameter, and moving further down, the P removal in mgP/ℓ is given at the intercept of lines GH and AB i.e. 8,8 mgP/ℓ. Hence the P removal for a raw wastewater with $S_{ti} = 600$, $f_{ts} = 0,20$ ($f_{bs} = 0,24$), $f_{xa} = 0,10$ and $R_s = 25$ d at $T = 20°C$ is 8,8 mgP/ℓ. Clearly, with the aid of the diagram, the effect of changing R_s, f_{xa} and f_{ts} (f_{bs}) on P removal can readily be ascertained (subject to the provisos mentioned above for the different processes).

7. DESIGN EXAMPLES

The design of excess P removal in nutrient removal processes by means of the procedure set out above is demonstrated by continuing the numerical example given in Chapter 4, Section 12 (in which the design procedure for organic material removal was demonstrated), Chapter 5, Section 6 (in which the behaviour of nitrification was demonstrated) and Chapter 6, Section 7 (in which nitrogen removal was demonstrated). The wastewater characteristics used in the design example, which are those of a typical raw and settled municipal wastewater, have all been defined and are given in Tables 4.3, 5.2 and 6.1. The causes and effects of variability in some of these characteristics, such as temperature, denitrification rates, readily biodegradable COD fraction etc. have been discussed in detail in Chapter 6, Sections 7.1 and 7.2.

Many of the decisions related to the design of excess P removal in a nutrient removal process are contingent upon the nitrogen removal behaviour of the process. Consequently, in the design example continuous reference to the design examples for nitrogen removal will be made, i.e. Chapter 6, Section 7 and a thorough *understanding of this section, and nitrogen removal in* general, will be presumed.

7.1 The Phoredox process

In the calculations for nitrogen removal in the Bardenpho process treating the *raw wastewater* at 25-day sludge age and 14°C, it was found that a total anoxic sludge mass fraction (f_{xdt}) of 0,40 was required to achieve complete denitrification (see Table 6.3). Now the total unaerated sludge mass fraction (f_{xm}) was found to be 0,50. Hence the largest anaerobic sludge mass fraction (f_{xa}) to effect excess P removal in a Phoredox process, (which requires near complete denitrification to achieve good excess P removal) (see Chapter 3, Section 3) is $0,50 - 0,40 = 0,10$ (from Eq. 7.8). Because complete denitrification can be achieved with $f_{xdt} = 0,40$, the nitrate concentration in the underflow recycle (N_{ns}) is zero, and, taking the DO concentration in the underflow recycle (O_s) as 1 mgO/ℓ, the readily biodegradable COD in the anaerobic reactor (S_{bsa}) for an underflow recycle ratio (s) of 1 is found from Eqs. 7.1b and 7.2b, i.e.

$$\Delta S_{bs} = 1(8{,}6.0 + 3{,}0.1) + 0$$

$$= 3$$

$$S_{bsa} = \{0{,}24(1 - 0{,}13 - 0{,}05)\,600 - 3\}/(1+1)$$

$$= 57{,}5 \text{ mgCOD}/\ell.$$

The propensity factor P_f is found from Eq. 7.5 i.e.

$$P_f = (57{,}5 - 25).0{,}10 = 3{,}25 \text{ mgCOD}/\ell$$

and the coefficient of excess P removal γ is found from Eq. 7.7 i.e.

$$\gamma = 0{,}35 - 0{,}29\exp(-0{,}242.3{,}25)$$

$$= 0{,}218 \text{ mgP/mgVASS}.$$

Hence the P removal at 25-day sludge age and 14°C is found from Eq. 7.6 i.e.

$$P_s = 600\left\{\frac{(1-0{,}05-0{,}13)0{,}45}{(1+0{,}20.25)}(0{,}218+0{,}015.0{,}20.0{,}20.25) + 0{,}015.0{,}13/1{,}48\right\}$$

$$= 600\,\{0{,}0615\,(0{,}218 + 0{,}015) + 0{,}0013\}$$

$$= 600.0{,}0156$$

$$= 9{,}4 \text{ mgP}/\ell.$$

The influent P concentration is 10 mgP/ℓ so that with 9,4 mgP/ℓ removal, the effluent P concentration P_{te} is $10{,}0 - 9{,}4 = 0{,}6$ mgP/ℓ (see Eq. 7.9). Similarly, the P removal at 22°C is found to be 8,0 mgP/ℓ. Note that S_{bsa}, P_f and γ remain unchanged at different temperatures but P_s is decreased because b_h in Eq. 7.6 is increased to 0,25/d at 22°C (see Table 4.1). The important design results are listed in Table 7.1.

In the design of the Phoredox process above, f_{xa} was set at the maximum value leaving a just sufficiently large f_{xdt} *to attain complete denitrification. This leaves the* Phoredox process in a critical state because to achieve the estimated P removals, complete denitrification should be achieved (see Figure 7.15). However, this cannot be guaranteed because there is no factor of safety on the denitrification. To introduce a factor of safety on the denitrification two options are open; (i) the primary anoxic sludge mass fraction (f_{x1}) can be increased, but this will have to be at the expense of a reduction in f_{xa} (because f_{xm} is fixed at 0,50 to achieve nitrification) but a reduction in f_{xa} will cause a reduction in P removal – from the design chart at 20°C (Figure 7.18) a reduction in f_{xa} from 0,10 to 0,07 reduces the P removal from 8,8 to 7,7 mgP/ℓ, (ii) the sludge age can be increased allowing a larger unaerated sludge mass fraction (f_{xm}), and hence, with f_{xa} fixed at 0,10, allowing a larger anoxic sludge mass fraction (f_{xdm}). However, the sludge age not only requires a larger process

TABLE 7.1 SUMMARY OF DESIGN CALCULATIONS FOR THE PHOREDOX AND UCT PROCESSES TREATING RAW OR SETTLED WASTEWATER AT 25-DAY SLUDGE AGE (WASTEWATER CHARACTERISTICS GIVEN IN TABLES 4.3, 5.2 and 6.1)

Parameter	Symbol	Units	Raw Wastewater				Settled Wastewater			
			Phoredox		UCT		Phoredox		UCT	
Temperature	**T**	**°C**	**14**	**22**	**14**	**22**	**14**	**22**	**14**	**22**
Safety factor	S_f		1,25	2,7	1,25	2,7	1,25	2,7	1,25	2,7
Max. unaerated sludge mass fraction	f_{xm}		0,50	0,50	0,50	0,50	0,50	0,50	0,50	0,50
Effluent TKN	N_{te}	mgN/ℓ	3,0	2,0	3,0	2,0	3,0	2,0	3,0	2,0
N for sludge production	N_s	mgN/ℓ	12,6	12,0	12,6	12,0	5,7	5,4	5,7	5,4
Nitrification capacity	N_c	mgN/ℓ	32,4	34,0	32,4	34,0	32,3	33,6	32,3	33,6
Anaerobic sludge mass fraction	f_{xa}		0,10	0,10	0,10	0,10	0,10	0,10	0,10	0,10
Max. anoxic sludge mass fraction	f_{xdm}		0,40	0,40	0,40	0,40	0,40	0,40	0,40	0,40
Max. denitrification potential	D_{pp}	mgN/ℓ	37,2	49,7	37,2	49,7	27,2	35,4	27,2	35,4
DO in a-recycle	O_a	mgO/ℓ	2,0	2,0	2,0	2,0	2,0	2,0	2,0	2,0
DO in s-recycle	$\left.\begin{matrix}O_s\\O_r\end{matrix}\right\}$	mgO/ℓ	1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0
Underflow recycle ratio	s		1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0
Prim. anoxic sludge mass fraction	f_{x1}		0,16	0,16	0,40	0,40	0,26	0,26	0,40	0,40
Sec. anoxic sludge mass fraction	f_{x3}		0,024	0,023	—	—	0,014	0,014	—	—
Total anoxic sludge mass fraction	f_{xdt}		0,40	0,40	0,40	0,40	0,40	0,40	0,40	0,40
Optimum a-recycle ratio	a_o		3,3	4,2	5,0*	5,0*	1,3	3,0	2,5	5,0*
Effluent nitrate	N_{ne}	mgN/ℓ	0,0	0,0	4,6	4,8	8,0	4,8	7,2	4,8
Nitrate recycled to anaerobic reactor	$\left.\begin{matrix}N_{ns}\\N_{nr}\end{matrix}\right\}$	mgN/ℓ	0,0	0,0	0,0	0,0	8,0	4,8	0,0	0,0
Readily biodegradable COD in anaerobic reactor	S_{bsa}	mgCOD/ℓ	57	57	57	57	16,5	30,0	50,8	50,8
Propensity factor	P_f	mgCOD/ℓ	3,3	3,3	3,3	3,3	0,00	0,50	2,57	2,57
γ coefficient		mgP/ mgVASS	0,22	0,22	0,22	0,22	0,06	0,093	0,195	0,195
P removal	P_s	mgP/ℓ	9,4	8,0	9,4	8,0	1,9	2,3	5,1	4,3
Effluent P concentration	P_{te}	mgP/ℓ	0,6	2,0	0,6	2,0	6,6	6,2	3,4	4,2

*These are not the calculated optimum values but are values at the practical and economical limit (see Chapter 6, Section 7.5).

volume, but also reduces the P removal – from the design chart (Figure 7.18), an increase in R_s from 25 to 30 days (which increases f_{xdt} from 0,40 to 0,43) reduces the P removal from 8,8 mgP/ℓ to 7,6 mgP/ℓ. Clearly the raw wastewater influent TKN/COD ratio (i.e. 0,08) is just on the limit for successful implementation of the Phoredox process – any increase in TKN/COD ratio will result in significant concentrations of nitrate in the effluent and underflow recycle (> 3 mgN/ℓ) which, when discharged to the anaerobic reactor, will cause a deterioration in P removal (Figure 7.15).

From the above, it is clear that the TKN/COD ratio of the settled wastewater (i.e. 0,114) is too high to attain excess P removal in the Phoredox process. This is demonstrated in the calculations below: For this process operating at 25-day* sludge age with $f_{xa} = 0,10$* and hence $f_{xdm} = 0,40$, from the denitrification theory an N_{ne} of 4,8 mgN/ℓ is obtained at 22°C with $f_{x1} = 0,26$, $f_{x3} = 0,14$, $a_o = 3,0$ and $N_{ne} = 8,0$ mgN/ℓ at 14°C with $a_o = 1,3$ (see Table 7.1). At 14°C, with 8 mgN/ℓ nitrate in the effluent and underflow s-recycle, and taking the s-recycle ratio to be specified at 1:1, and containing a DO concentration (O_s) of 1 mgO/ℓ, then from Eqs 7.1b and 7.2b, the readily biodegradable COD concentration in the anaerobic reactor (S_{bsa}) is only 16,5 mgCOD/ℓ, which is too low for stimulating excess P removal. Hence from Eq. 7.5 the propensity factor (P_f) is 0,0 and from Eq. 7.7 the γ coefficient is at its minimum of 0,06 mgP/mgVASS. From Eq. 7.6, with $\gamma = 0,06$, the P removal (P_s) at $R_s = 25$ and T = 14°C is 1,9 mgP/ℓ, which, from Eq. 7.9, with $P_{ti} = 8,5$ mgP/ℓ gives an effluent P concentration (P_{te}) of 6,6 mgP/ℓ, (see Table 7.1). A similarly poor P removal is obtained also at 22°C (see Table 7.1). Clearly when the wastewater characteristics are such that near complete denitrification cannot be obtained, the Phoredox process will not function as an efficient biological P removal facility.

7.2 The UCT processes

In the UCT processes, the complicating effect of the possibility of a discharge of nitrate to the anaerobic reactor is absent. Consequently, in their design, the nitrogen and phosphorus removal aspects can be dealt with separately and the P removal in the processes can be calculated without first having to determine the effluent nitrate concentration.

In the design of the UCT or modified UCT processes, first the UCT process (Figure 7.5) is designed and once this is complete, a check is made as to whether or not this configuration can be converted to a modified process (Figure 7.6). It is recommended that where it is possible to convert the UCT process to a modified one, the option of operating the modified process as a UCT process should always be included (see Figure 7.16). Consider also the option of incorporating

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*Selected the same as for the raw wastewater case so that a direct comparison can be made between the settled and raw wastewaters.

the Modified Phoredox/Modified UCT/UCT combination (see Figure 7.17).

At the same sludge age, anaerobic sludge mass fraction and temperature, the P removal attainable in the UCT process is the same as that attainable in the Phoredox process with complete denitrification i.e. when no nitrate is discharged to the anaerobic reactors of both processes. Consequently, with the UCT process operating at 25-day* sludge age with $f_{xa} = 0,10$,* the P removal from the *raw wastewater* is 9,4 mgP/ℓ at 14°C and 8,0 mgP/ℓ at 22°C. Note also that for the same process parameters R_s, f_{xa}, f_{xdm} and temperature, the nitrification capacity N_c and denitrification potential D_{pp} for the UCT process are identical to those for the Phoredox process. The nitrogen removal behaviour of the UCT process is identical to that of the MLE process and hence the effluent nitrate concentration N_{ne} and optimum mixed liquor a-recycle ratio a_o are found from N_c and D_{pp} with the aid of Eqs. 6.30 and 6.31; from Eq. 6.30, a_o at both 14°C and 22°C is greater than the practical limit of 5 and hence a_o is limited at 5 which gives N_{ne} at 14°C and 22°C as 4,6 mgN/ℓ and 4,8 mgN/ℓ respectively from Eq. 6.31. The important design details emerging from the above calculations are given in Table 7.1. A comparison of the Phoredox and UCT process will be given once the design of the UCT process treating the settled wastewater has been demonstrated.

With the UCT process treating the *settled wastewater* at 25-day sludge age and an anaerobic sludge mass fraction of 0,10, the P removal is calculated from the theory (note that now, the P removal is not the same as that in the Phoredox process because with the settled wastewater in this process, nitrate is discharged to the anaerobic reactor). Taking the nitrate recycled to the anaerobic reactor in the UCT process as 0,0 and an r-recycle ratio of 1:1 with a DO concentration (O_r) of 1 mgO/ℓ, the readily biodegradable COD concentration in the anaerobic reactor (S_{bsa}) is found from Eqs 7.1a and 7.2a i.e.

$$\Delta S_{bs} = 1\,(8{,}6.0+3{,}0.1)+3{,}0.0$$

$$= 3 \text{ mg COD}/\ell$$

$$S_{bsa} = \{0{,}33.(1-0{,}08-0{,}04).360-3\}/(1+1)$$

$$= 50{,}8 \text{ mgCOD}/\ell.$$

The propensity factor P_f is found from Eq. 7.5 i.e.

$$P_f = (50{,}8-25).0{,}10 = 2{,}58 \text{ mgCOD}/\ell$$

and the γ coefficient is found from Eq. 7.7 i.e.

$$\gamma = 0{,}35-0{,}29\exp\,(-0{,}242.2{,}58)$$

$$= 0{,}195 \text{ mgP/mgVASS}.$$

Hence the P removal P_s at 25 days sludge age and 14°C is found from Eq. 7.6 i.e.

$$P_s = 360\,\{\frac{(1-0{,}08-0{,}04)0{,}45}{(1+0{,}20.25)}\,(0{,}195+0{,}015.0{,}20.0{,}20.25) + 0{,}015.0{,}04/1{,}48\}$$

$$= 5{,}1 \text{ mgP}/\ell.$$

The effluent P concentration P_{te} is given by Eq. 7.9 i.e.

$$P_{te} = 8{,}5-5{,}1 = 3{,}4 \text{ mgP}/\ell.$$

Similarly, P_s and P_{te} at 22°C are 4,3 and 4,2 respectively.

The nitrogen removal behaviour of the UCT process is found from Eqs. 6.30 and 6.31 – note that the nitrification capacity N_c and denitrification potential D_{pp}, with $f_{xdm} = 0{,}40$ are identical to those of the Phoredox process. At 14°C the effluent nitrate concentration N_{ne} and optimum a-recycle ratio a_o are 7,2 mgN/ℓ and 2,5 and at 22°C, the calculated a_o is greater than the practical limit of 5 and with $a = 5$, $N_{ne} = 4{,}8$ mgN/ℓ. Important design details of the N and P removal design are given in Table 7.1.

7.3 Comparison of Phoredox and UCT processes

Comparing the Phoredox and UCT processes at 25-day sludge age and anaerobic sludge mass fraction of 0,10 (see Table 7.1), it can be seen that for the *raw wastewater*, the Phoredox and UCT processes achieve the same excess P removals, but the N removal is better in the Phoredox than in the UCT process; the former achieves complete denitrification whereas the latter produces an effluent nitrate concentration of 5 mgN/ℓ. However, the Phoredox process is in a critical state in that an unexpected increase in TKN/COD ratio above 0,08 will result in a deterioration in excess P removal. In contrast the UCT process can accommodate unexpected increases in TKN/COD ratio up to about 0,13 (0,11 for the modified process), while maintaining its calculated excess P removal of 9,4 mgP/ℓ at 14°C; this flexibility is at the cost of producing a minimum effluent nitrate concentration of 5 mgN/ℓ.

From the above results, the following general conclusion can be made: For wastewaters with TKN/COD ratios and readily biodegradable COD fractions such that the Phoredox process can achieve good excess P removals, its effluent nitrate concentration will be near zero. For the same conditions the UCT process will also achieve good excess P removal but its effluent nitrate concentration will be 5 to 6 mgN/ℓ.

In the Phoredox process, the good effluent quality is at the cost of process flexibility to accommodate unexpected TKN/COD ratio increases; this flexibility is incorporated in the UCT process but it is at the cost of a poor effluent quality. In these situations (which usually arise with wastewater of TKN/COD ratio less than 0,08 with 20 to 25% readily biodegradable COD fraction, f_{ts}), the process to select would depend on the variability in TKN/COD ratio of the wastewater.

Comparing the Phoredox and UCT processes treating the *settled wastewater* with a TKN/COD ratio of 0,114, it can be seen that both processes produce approximately similar effluent nitrate concentrations (see Table 7.1). However, with respect to excess P removal, the Phoredox process can be seen to have a poorer P removal facility than the UCT process; the difference arises from the fact that in the Phoredox process, the exclusion of nitrate from the anaerobic reactor becomes impossible once near com-

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*Selected the same so that a direct comparison between the Phoredox and UCT process can be made. A design with different R_s and f_{xa} to effect better P removal (at the cost of poorer N removal) will be presented below.

plete denitrification cannot be achieved, whereas in the UCT process, this is possible up to TKN/COD ratios of about 0,13 (depending on the f_{bs} (f_{ts}) fraction).

7.4 Redesign of the UCT process

Although in the UCT process the exclusion of nitrate from the anaerobic reactor is possible, thereby allowing the greatest potential for P removal, the actual P removal attained in the process for the settled wastewater is insufficient to reduce the effluent P concentration (P_{te}) below 1 mgP/ℓ. This is because the anaerobic sludge mass fraction (f_{xa}) and sludge age (R_s) have been incorrectly selected to attain $P_{te} < 1$ mgP/ℓ. To improve the P removal, R_s needs to be reduced and f_{xa} increased. From the design chart for raw wastewater (Figure 7.18), if f_{xa} is increased from 0,10 to 0,15 and the sludge age decreased from 25 to 20 days, the estimated P removal (at 20°C with no nitrate and DO recycled to the anaerobic reactor) is 11,5 mgP/ℓ.

Accepting an anaerobic sludge mass fraction of 0,15 and the sludge age of 20 days, the design of the UCT process for raw and settled wastewater at 14°C and 22°C was undertaken and the important design results are listed in Table 7.2. By following the step by step procedure, the calculations for the raw wastewater at 14°C and 22°C is demonstrated: Accepting an r-recycle ratio to the anaerobic reactor (r) of 1, with 1 mgN/ℓ nitrate (N_{nr}) and 1 mgO/ℓ dissolved oxygen (O_r) (to introduce a factor of safety), the readily biodegradable COD concentration in the anaerobic reactor (S_{bsa}) is found from Eqs 7.1a and 7.2a

$$\Delta S_{bs} = 1(8,6.1+3,0.0)+3,0.0 = 11,6 \text{ mgCOD}/\ell$$

$$S_{bsa} = \{0,24(1-0,13-0,05)\,600-11,6\}/(1+1)$$

$$= 53,3 \text{ mgCOD}/\ell.$$

With $f_{xa} = 0,15$, the propensity factor P_f is found from Eq. 7.5 i.e.

$$P_f = (53,3-25).0,15 = 4,25 \text{ mgCOD}/\ell$$

The γ coefficient is found from Eq. 7.7 i.e.

$$\gamma = 0,35-0,29\exp(-0,242.4,25)$$

$$= 0,246.$$

With $R_s = 20$ d, the P removal P_s at 14°C and 22°C is found from Eq. 7.6 i.e.

TABLE 7.2 SUMMARY OF DESIGN CALCULATIONS FOR THE UCT PROCESS OPERATING AT 20-DAY SLUDGE AGE WITH AN ANAEROBIC SLUDGE MASS FRACTION OF 0,15 TREATING RAW AND SETTLED WASTEWATERS (WASTEWATER CHARACTERISTICS GIVEN IN TABLES 4.3, 5.2 AND 6.1)

Parameter	Symbol	Units	Raw Wastewater		Settled Wastewater	
Temperature	**T**	**°C**	**14**	**22**	**14**	**22**
Nitrate in r-recycle	N_{nr}	mgN/ℓ	1,0	1,0	1,0	1,0
DO in r-recycle	O_r	mgO/ℓ	1,0	1,0	1,0	1,0
Influent biodegradable COD	S_{bi}	mgCOD/ℓ	492	492	317	317
Rapidly biodegradable COD fraction	f_{bs}	—	0,24	0,24	0,33	0,33
r-recycle ratio	r	—	1,0	1,0	1,0	1,0
Rapidly biodegradable COD in anaerobic	S_{bsa}	mgCOD/ℓ	53,3	53,3	46,5	46,5
Anaerobic sludge mass fraction	f_{xa}		0,15	0,15	0,15	0,15
Propensity factor	P_f	mgCOD/ℓ	4,24	4,24	3,22	3,22
γ coefficient	γ	mgP/mgVASS	0,246	0,246	0,217	0,217
P removal	P_s	mgP/ℓ	12,2	10,4	6,7	5,6
Effluent P	P_{te}	mgP/ℓ	0,0	0,0	1,8	2,9
Factor of safety	S_f	—	1,25	2,9	1,25	2,9
Max. unaerated sludge mass fraction	f_{xm}	—	0,42	0,42	0,42	0,42
Effluent TKN	N_{te}	mgN/ℓ	3,0	2,0	3,0	2,0
N for sludge production	N_s	mgN/ℓ	13,2	12,6	6,1	5,7
Nitrification capacity	N_c	mgN/ℓ	31,8	33,4	31,9	33,3
Max. anoxic sludge mass fraction	f_{xdm}	—	0,27	0,27	0,27	0,27
Denitrification potential	D_{pp}	mgN/ℓ	29,0	37,3	22,0	27,3
Underflow recycle ratio	s	—	1,0	1,0	1,0	1,0
DO in a-recycle	O_a	mgO/ℓ	2,0	2,0	2,0	2,0
DO in s-recycle	O_s	mgO/ℓ	1,0	1,0	1,0	1,0
Opt. a-recycle ratio	a_o	—	3,6	5,0*	0,93	2,25
Effluent nitrate	N_{ne}	mgN/ℓ	5,7	4,8	10,9	7,9
UCT process modification						
1st Prim. Anoxic	f_{xd1}	—	0,10	0,10	0,10	0,10
Denitrification potential	D_{pd1}	mgN/ℓ	19,6	21,8	16,0	18,0
Equiv. nitrate load		mgN/ℓ	6,1	4,9	11,3	8,3
Factor of safety	S_{fd}	—	3,2	4,5	1,4	2,2
Modification successful			Yes		Yes	

*The calculated a_o value (i.e. 9,3) is greater than the practical limit of 5. Hence a_o is set equal to the practical limit of 5.

$$P_s = 600\left\{\frac{(1-0{,}13-0{,}05)0{,}45}{(1+0{,}20.20)}(0{,}246+0{,}015.0{,}20.0{,}20.20) + 0{,}015.0{,}13/1{,}48\right\}$$

$$= 12{,}2 \text{ mgP}/\ell.$$

and at 22°C

$$P_s = 600\left\{\frac{(1-0{,}13-0{,}05)0{,}45}{(1+0{,}25.20)}(0{,}246+0{,}015.0{,}20.0{,}25.20) + 0{,}015.0{,}13/1{,}48\right\}$$

$$= 10{,}4 \text{ mgP}/\ell.$$

As both these removals are greater than the influent P concentration, the effluent P concentration can be taken to be zero.

Accepting a nitrification safety factor (S_f) of 1,25 the maximum unaerated sludge mass fraction at 20-day sludge age, 14°C and $\mu_{nm20} = 0{,}36/d$ is found from Eq. 5.24. (Temperature dependencies of nitrification constants are given in Table 6.2), i.e.

$$f_{xm} = 1 - 1{,}25\ (0{,}034 + 1/20)/0{,}18$$

$$= 0{,}42$$

With $f_{xm} = 0{,}42$, S_f at 22°C is found from Eq. 5.24, i.e.

$$S_f = (1 - 0{,}42).0{,}454/(0{,}042 + 1/20)$$

$$= 2{,}9.$$

Because a reasonable S_f at 14°C has been selected, complete nitrification can be assumed to take place and hence effluent TKN concentration of 3,0 mgN/ℓ at 14°C and 2,0 mgN/ℓ at 22°C are acceptable. The nitrogen requirement for sludge production at 14°C and 22°C is found from Eq 4.23 i.e. 13,2 at 14°C and 12,6 at 22°C. Hence the nitrification capacity N_c is found from Eq. 5.29 i.e. 31,8 at 14°C and 33,4 at 22°C.

With f_{xm} and f_{xa} fixed at 0,42 and 0,15 respectively, the maximum anoxic sludge mass fraction (f_{xdm}) is given by Eq. 7.8 i.e. $f_{xdm} = 0{,}42 - 0{,}15 = 0{,}27$. With $f_{xdm} = 0{,}27$, the denitrification potential of the primary anoxic reactor with $f_{x1} = f_{xdm} = 0{,}27$ is found from Eq. 6.20 i.e. 29,0 mgN/ℓ at 14°C and 37,3 at 22°C. With N_c and D_{pp} known and selecting the underflow s-recycle ratio as 1 and the DO in the a- and s-recycles (O_a and O_s respectively) as 2 and 1 mgO/ℓ respectively, the nitrogen performance is found from Eqs. 6.30 and 6.31 i.e. the optimum a-recycle ratio (a_o) and effluent nitrate concentration (N_{ne}) at 14°C are 3,6 and 5,7 mgN/ℓ respectively and at 22°C the calculated a_o is greater than the practical limit of 5 (i.e. 9,3) so that a_o is set at 5 giving $N_{ne} = 4{,}8$ mgN/ℓ (see Table 7.2). For design details of the above procedure see Chapter 6, Section 7.5.

The results in Table 7.2 show that for the *raw wastewater* at 14°C and 22°C, more P can be removed than is present in the influent so that an effluent P concentration of less than 1 mgP/ℓ can be obtained with a factor of safety. The effluent nitrate concentrations are 5,7 mgN/ℓ at 14°C and 4,8 mgN/ℓ at 22°C. Comparing these results with the Phoredox process at 25-day sludge age (Table 7.1), it can be seen that Phoredox process produces a better effluent quality. However, the UCT process incorporates a factor of safety on the P removal whereas the P removal in the Phoredox process depends on the attainment of near complete denitrification without a factor of safety. Consequently, although the UCT process yields a poorer N removal, the P removal can be attained with greater surety in the event of unexpected increases in the TKN/COD ratio. The upper limit of the TKN/COD ratio can be calculated by testing higher TKN/COD ratios on the design and for approximately normal wastewater is about 0,13.

For the settled wastewater, the P removal (P_s) is 6,7 mgP/ℓ at 14°C and 5,6 mgP/ℓ at 22°C, leaving 1,8 mgP/ℓ and 2,9 mgP/ℓ in the effluent at 14°C and 22°C respectively. The effluent nitrate concentrations (N_{ne}) are 10,9 mgN/ℓ and 7,9 mgN/ℓ at 14°C and 22°C respectively. If P_s is insufficient, then f_{xa} can be increased to its maximum of 0,25, but this will result in a poorer N removal i.e. with $f_{xa} = 0{,}20$, $P_s = 7{,}5$ mgP/ℓ and $N_{ne} = 12{,}2$ mgN/ℓ at 14°C and $P_s = 6{,}4$ mgP/ℓ and $N_{ne} = 9{,}7$ mgN/ℓ at 22°C. If the N removal with $f_{xa} = 0{,}20$ is insufficient, then the sludge age can be increased to improve the N removal but the increase in R_s will reduce the P removal. (see design chart). Clearly there are limits to which biological nutrient removal can be achieved even in an optimally designed process. These limits, and hence the effluent quality, are dependent on the influent wastewater characteristics, in particular the readily biodegradable COD fraction, the TKN/COD ratio and the P/COD ratio i.e. the higher the first and the lower the other two the better the effluent N and P quality. However, it should be noted that biological N and P removal can be improved by adding readily biodegradable COD to the process (see Chapter 3, Section 3). The effect of this external COD addition on the N and P removal in a process can be checked by appropriately adjusting the readily biodegradable COD fraction and influent COD concentration.

Having completed the design for the UCT process, it is necessary to check whether or not the process can be operated as a modified UCT process (Figure 7.6). The conversion to the modified UCT process is done by subdividing the anoxic sludge mass fraction in two subfractions; the first is usually allocated a sludge mass fraction of f_{xd1} of about 0,10, and the second, f_{xd2}, having the remaining sludge mass fraction i.e., $f_{xd2} = f_{xdm} - f_{xd1}$. For successful operation of the modified UCT process, the denitrification potential of f_{xd2} (D_{pd1}) must be greater than its equivalent nitrate load imposed by the underflow s-recycle at the minimum expected temperature T_{min}. The equivalent nitrate load is calculated from the effluent nitrate concentration for the UCT process before subdivision and is given by $s(N_{ne} + O_s/2{,}86)$. The factor by which D_{pd1} is greater than $s(N_{ne} + O_s/2{,}86)$ is called the factor of safety (S_{fd}) and it is recommended that this factor be greater than 1,2. (see Section 5 above).

The calculations for the modification of the UCT process treating the raw and settled wastewaters at 20 days sludge age with an anaerobic sludge mass fraction of 0,10 are given in Table 7.2. For both the raw and settled wastewaters the modification is acceptable – the factor of safety S_{fd} at 14°C for the raw wastewater is 3,2 and for the settled wastewater is 1,4. Generally it will be found that the modification is acceptable for TKN/COD < 0,12 with a readily biodegradable COD fraction (f_{bs}) around 0,25. The TKN/COD ratio limit for successful application of the modification is strongly dependent on the readily biodegradable fraction f_{bs} – the lower the fraction the lower the TKN/COD ratio limit (see Section 5 above). It should be noted also that the magnitude of the anaerobic sludge mass fraction f_{xa}, plays a rôle as to whether or not a UCT process modification will be successful. For example, at 20 days sludge age with $f_{xa} = 0,20$, the effluent nitrate concentration treating settled wastewater at 14°C is 12,2 mgN/ℓ (as previously calculated), giving an equivalent nitrate load of 12,6 mgN/ℓ; with $D_{pd1} = 16,0$ mgN/ℓ (see Table 7.2), the factor of safety $S_{fd} = 1,26$, which is lower than that obtained with $f_{xa} = 0,15$ (see Table 7.2), and near the limit where the modification would be acceptable.

When checking whether or not the modified UCT process can be successfully applied, it is necessary to check that the first anoxic reactor is sufficiently large to allow complete utilization of the readily biodegradable COD. This is checked by means of Eq. 6.21 (see Section 5 above). From Eq. 6.21, the minimum anoxic sludge mass fraction to completely utilize the readily biodegradable COD at 14°C is 0,062 and 0,085 for the raw and settled wastewaters respectively (see Table 6.3). Hence $f_{xd1} = 0,10$ is adequate. By designing f_{xd1} to allow the possibility of complete utilization of the readily biodegradable COD, the modified process can achieve excess P removal up to the TKN/COD ratio at which the effluent nitrate concentration is so high that all the readily biodegradable COD is utilized. This TKN/COD ratio can be calculated by trial and error by testing increasing TKN/COD ratios on the process.

Having made the modification to the UCT process, the possibility of operating it as a UCT process is not excluded; by making provision that the r-recycle flow can be taken from either the first or the second anoxic reactor (see Figure 7.15) the process can be operated either as a modified UCT or a UCT process. The advantages of designing the process as a modified UCT process were set out in detail in Section 2.2.3 above. Furthermore complete flexibility of process operation can be incorporated into the plant by providing optional recycles allowing the plant to be operated as a Modified Phoredox, Modified UCT or UCT process (see Figure 7.17). Also, irrespective of the type of process or combination of processes that are selected for a particular design, it is strongly recommended that the anaerobic reactor is subdivided into three or four compartments operated in series. The advantage of this series system anaerobic reactor is given in Section 3.4 above.

7.5 Process volume

The volumes of the reactors of the Phoredox process are calculated by means of the method described in Chapter 6, Section 7.6.1 and for the same sludge age, temperature and COD load, the total volume of a Phoredox process will be the same as that of the Bardenpho process, i.e. subdivision of the process volume between the different reactors is proportional to the sludge mass fractions of the different reactors. The reactor volumes and retention times for the Phoredox process at 25-day sludge age (see data in Table 7.1) are set out in Table 7.3.

With the UCT process, and its modification the MLSS (or MLVSS) is not uniform throughout the process with the result that the reactor volume fractions are not equal to the sludge mass fractions. There are two methods for determining the volume fractions (or volumes) of the reactors:

TABLE 7.3 REACTOR VOLUMES, NOMINAL AND ACTUAL RETENTION TIMES FOR THE PHOREDOX PROCESS TREATING RAW AND SETTLED WASTEWATER AT 25-DAY SLUDGE AGE. FOR PROCESS PERFORMANCE SEE TABLE 7.1

Process Parameter	Symbol	Wastewater		Units
		Raw	Settled	
Sludge mass	MX_t	54 200	21 850	kgTSS
Average MLSS conc.	$\bar{X}_t$	4 000	4 000	mg/ℓ
Total volume	V_p	13 550	5 460	m³
Total flow	Q	13 330	13 330	m³/d
Total nominal retention time	R_{ht}	24,4	9,8	h
Mixed liquor recycle (14°C)	a	3,3	1,3	–
Underflow recycle	s	1,0	1,0	–
Anaerobic				
Mass fraction	f_{xa}	0,10	0,10	
Volume		1 355	546	m³
Nominal retention time		2,4	1,0	h
Actual retention time		1,2	0,5	h
Primary anoxic				
Mass fraction	f_{x1}	0,16	0,26	
Volume		2 168	1 420	m³
Nom. retention time		3,9	2,6	h
Actual retention time		0,74	0792	h
Main aeration*				
Mass fraction		0,45	0,45	
Volume		6 098	2 457	m³
Nom. retention time		11,0	4,4	h
Actual retention time		2,1	1,34	h
Secondary anoxic				
Mass fraction	f_{x3}	0,24	0,14	
Volume		3 252	764	m³
Nom. retention time		5,9	1,4	h
Actual retention time		2,9	0,7	h
Reaeration*				
Mass fraction		0,05	0,05	
Volume		678	273	m³
Nom. retention time		1,2	0,50	h
Actual retention time		0,6	0,25	h

*The reaeration reactor is usually taken to have a sludge mass fraction of around 0,05 to 0,07 and this fraction is deducted from the aerobic sludge mass fraction; the remaining aerobic sludge mass fraction constitutes the main aeration sludge mass fraction.

(i) By calculating the reactor volume fractions from the mass fractions for a fixed total process volume; the anaerobic volume fraction is given by

$$f_{va} = f_{xa} \{(1+r)/(r+f_{xa})\} \qquad (7.10)$$

and the anoxic and aerobic volume fractions are given by

$$f_{vd}/f_{xd} = f_{vb}/f_{xb} = 1 - \{f_{xa}/(r+f_{xa})\} \qquad (7.11)$$

where

first subscript v and x refer to volume and mass fractions respectively and second subscripts a, d and b refer to anaerobic, anoxic and aerobic reactors respectively.

Knowing the volume fractions, the volumes of the reactors are calculated from the total volume of the process for a selected mean process MLSS concentration $\bar{X}_t$ (see Eq. 4.16). For the selected mean MLSS concentration $\bar{X}_t$, the MLSS concentrations in the reactors are given by the following equations; in the anaerobic reactor (X_{ta}),

$$X_{ta} = \bar{X}_t . f_{xa}/f_{va} \qquad (7.12)$$

in the anoxic (X_{td}) and aerobic (X_{tb}) reactors

$$X_{td} = X_{tb} = \bar{X}_t/[1-\{f_{xa}/(r+f_{xa})\}] \qquad (7.13)$$

From Eq. 7.13, it can be seen that the sludge concentration in the anoxic (X_{td}) and aerobic (X_{tb}) reactors is greater than the selected mean concentration $\bar{X}_t$. For $r=1$ and $f_{xa}=0,15$, X_{td} and X_{tb} are 15% higher than $\bar{X}_t$. This higher than mean aerobic reactor sludge concentration must be taken into account in the design of the secondary settling tanks, which receive the outflow from the aerobic reactor.

(ii) By calculating the reactor volumes from the sludge mass fractions for a fixed aerobic reactor sludge concentration X_{tb}; if the anaerobic sludge mass fraction is f_{xa}, then the remaining sludge $(1-f_{xa})M(X_t)$, is diluted into the anoxic and aerobic reactors i.e. the combined volumes of the anoxic and aerobic reactors is given by $(1-f_{xa})M(X_t)/X_{tb}$. The subdivision of this volume is easily found from their sludge mass fractions because the sludge concentrations in these reactors are the same. Now the sludge concentration in the anaerobic reactor X_{ta} is a fraction $r/(1+r)$ of the concentration in the aerobic reactor i.e. $X_{ta} = X_{tb}\, r/(1+r)$. Knowing the mass of sludge in the anaerobic reactor i.e. $f_{xa}\, M(X_t)$, and the concentration X_{ta} the volume is given by $f_{xa}\, M(X_t)/X_{ta}$. This method will yield a larger process volume than method (i).

Comparing the volumes of the UCT process attained with the above two methods with volume of the Phoredox process, method (i) will produce a process volume equal to that of the Phoredox for the same mean process concentration $\bar{X}_t$ but the aerobic reactor sludge concentration in the UCT process will be greater than that in the Phoredox {by a factor $(1+f_{xa})$}, and method (ii) will produce a process volume larger than that of the Phoredox {by a factor $(1+f_{xa})$} for the same aerobic reactor sludge concentration. Reactor volumes, retention times for the UCT process as found from method (ii) are give in Table 7.4.

7.6 Oxygen demand

The average daily oxygen demand for a nitrogen and phosphorus removal process is calculated in the identical fashion as that for a nitrogen removal process. This is discussed in detail in Chapter 6, Section 7.6.2.

A simple design rule by means of which the average peak daily oxygen demand can be estimated from the influent COD and TKN load variation over the day is also given in that section.

TABLE 7.4 REACTOR VOLUMES AND NOMINAL AND ACTUAL RETENTION TIMES FOR THE MODIFIED UCT PROCESS TREATING RAW AND SETTLED WASTEWATER AT 20-DAY SLUDGE AGE. FOR PROCESS PERFORMANCE SEE TABLE 7.2

Process Parameter	Symbol	Wastewater Raw	Wastewater Settled	Units
Sludge mass	$M(X_t)$	54 200	21 850	kgTSS
Aerobic reactor MLSS concentration	X_{tb}	4 000	4 000	mgTSS/ℓ
Aerobic mass fraction	f_{xa}	0,15	0,15	–
Aerobic sludge mass		8 130	3 278	kgTSS
Anoxic + Aerobic sludge mass		46 070	18 572	kgTSS
Anoxic + Aerobic volume		11 518	4 643	m³
Total anoxic mass fraction	f_{xdm}	0,27	0,27	
First primary anoxic	f_{xd1}	0,10	0,10	
Second primary anoxic	f_{xd2}	0,17	0,17	
Aerobic mass fraction	f_{xb}	0,58	0,58	
1st Anoxic volume		1 355	546	m³
2nd Anoxic volume		2 304	1 475	m³
Aerobic volume		7 860	3 168	m³
r-recycle ratio	r	1	1	
Anaerobic MLSS conc.	X_{ta}	2 000	2 000	mgTSS/ℓ
Anaerobic volume		4 065	1 639	m³
Total volume		15 583	6 282	m³
Average process conc.	$\bar{X}_t$	3 480	3 480	mgTSS/ℓ
Opt. a-recycle ratio	a_o	3,6	0,93*/ 1,5**	
Underflow recycle ratio	s	1,0	1,0	
Influent flow	Q	13 330	13 330	m³/d
Retention times				
Anaerobic: Nominal		7,3	3,0	h
Actual		3,6	1,5	h
1st Anoxic: Nominal		2,4	1,0	h
Actual		0,8	0,3	h
2nd Anoxic: Nominal		4,1	2,7	h
Actual		0,9	1,4*/ 1,1**	h
Aerobic: Nominal		14,2	5,7	h
Actual		3,1	3,0*/ 2,3**	h

*Based on the calculated optimum mixed liquor recycle ratio a_o.

**Based on a mixed liquor recycle ratio to yield an actual anoxic retention time of aobut 1 hour.

8. GUIDELINES FOR PROCESS SELECTION

From the theory of biological excess P removal and the nitrogen removal behaviour of the various processes, the following broad guidelines can be given to assist the selection of a process:

8.1 Process selection

(1) If the readily biodegradable COD concentration in the influent (S_{bsi}) is less than 60 mgCOD/ℓ (irrespective of the influent COD concentration), significant excess P removal is unlikely to be obtained in any of the processes.

(2) If $S_{bsi} > 60$ mgCOD/ℓ, excess P removal can be achieved provided nitrate can be excluded from the anaerobic reactor, the removal increasing as S_{bsi} increases. Whether or not nitrate can be excluded

from the anaerobic reactor depends on the influent TKN/COD ratio and the process type. The limits* given below are for approximately normal wastewater, (eg. with readily biodegradable COD fraction f_{ts} around 15 to 20%) and where complete nitrification is obligatory**

(i) If the TKN/COD < 0,08 mgN/mgCOD, complete nitrate removal is possible and the Phoredox process is indicated.

(ii) If 0,08 < TKN/COD < 0,11 complete nitrate removal no longer is possible, but nitrate can be excluded from the anaerobic reactor by using the modified UCT process.

(iii) If 0,11 < TKN/COD < 0,14, the modified UCT process no longer can exclude nitrate from the anaerobic reactor and the UCT process is indicated provided the a-recycle ratio is carefully controlled. However, poor settling sludges may be encountered.

(iv) If TKN/COD > 0,14 it is unlikely that biological excess P removal will be achieved with normal municipal wasteflows.

8.2 Anaerobic sludge mass fraction selection

The readily biodegradable COD fraction (f_{ts}) is approximately 0,20 for raw municipal wastewaters (i.e. f_{bs} about 0,25), and approximately 0,30 for settled municipal wastewaters (i.e. f_{bs} about 0,34) – see Table 6.2 and Chapter 3, Section 3. Consequently, for approximately normal municipal wastewaters, the lower the COD strength (S_{ti}) the lower the S_{bsi}; when S_{ti} for raw wastewater is less than 250 mgCOD/ℓ, S_{bsi} is less than 50 mgCOD/ℓ and the attainment of excess P removal for r or s-recycle ratios of 1:1 is unlikely or sporadic even if the process is correctly selected and designed. Conversely the higher the S_{ti}, the greater the S_{bsi} and the easier it is to establish the conditions for excess P removal and obtain high P contents in the active mass of the sludge γ. As γ is related to both ($S_{bsa} - 25$) and the anaerobic sludge mass fraction f_{xa}, the adverse effect of low ($S_{bsa} - 25$) at low S_{ti} can be countered to a degree by increasing f_{xa} to its maximum* of about 0,25.

The following guidelines based on raw COD strengths are not unreasonable for an initial estimate of f_{xa} to obtain good excess P removals or approximately normal municipal wastewaters:

(i) $S_{ti} < 400$ mgCOD/ℓ, $f_{xa} = 0{,}20 - 0{,}25$

(ii) $400 < S_{ti} < 700$, $f_{xa} = 0{,}15 - 0{,}20$

(iii) $S_{ti} > 700$ mgCOD/ℓ, $f_{xa} = 0{,}10 - 0{,}15$

The above initial estimates must be checked and modified depending on the actual P/COD and readily biodegradable COD fraction.

8.3 Subdivision of anaerobic reactor

Recent research has indicated that it is advantageous to subdivide the anaerobic reactor into a number of compartments operated in series. This is because the absorption of readily biodegradable COD by the poly-P accumulating organisms appears to be a first order reaction (see Section 3.4 above) so that, for the same anaerobic sludge mass fraction, a series of completely mixed reactors operates more effectively than a single completely mixed reactor. Theoretically, the greater the number of equal volumed compartments the better, but practically very little advantage is gained by having more than 3 or 4. Four compartments would appear to be the most suitable because then, by providing the required additional recycle flows, the process can be operated as a Modified Phoredox or UCT type process each with approximately the same anaerobic sludge mass fraction (see Figure 7.17).

9. CLOSURE AND CONCLUSIONS

The above calculations demonstrated that:

(1) The average wastewater characteristics govern the design of and the effluent quality attainable in an activated sludge process for biological N and P removal. The principal wastewater charactericstics are:

   (i) the COD concentration (S_{ti})
   (ii) the readily biodegradable COD fraction (f_{ts})
   (iii) the TKN/COD ratio
   (iv) the maximum specific growth rate of the nitrifiers at 20°C (μ_{nm20})
   (v) the maximum and minimum temperatures (T_{max} and T_{min})
   (vi) the P/COD ratio.

(2) Once the sludge age of a selected process is fixed the removals of N and P attainable are fixed and the N and P removal *efficiencies* then depend on the TKN/COD and P/COD ratios of the wastewater. Because for approximately normal wastewaters ($f_{ts} = 0{,}20$) complete denitrification can only be achieved for TKN/COD ratios less than 0,08 (without the addition of an external energy source), the higher the ratio above 0,08, the lower N removal efficiency.

(3) Attainment of excess P removal is crucially dependent on the readily biodegradable COD concentra-

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*Generally, the higher the f_{ts} ratio above 0,20, the higher these limits and the lower the f_{ts} ratio below 0,20 the lower these limits. These different limits can be calculated with the theory presented in this monograph using different f_{ts} ratios.

**In South Africa, complete nitrification is obligatory. Where complete nitrification is not obligatory, nitrate discharge to the anaerobic reactor can be avoided by limiting nitrification by e.g. limiting the oxygen supply to the process, but in South Africa, such a strategy falls in line more with an emergency action than a design criterion.

***Laboratory and pilot scale data on process behaviour for $f_{xa} > 0{,}20$ is available from the NIWR (see Interim Design Guide by NIWR).

tion in the influent S_{bsi}. For recycle ratios to the anaerobic reactor of 1:1, if $S_{bsi} < 50$ mgCOD/ℓ, it is highly unlikely that excess P removal can be achieved in any process; if $S_{bsi} > 50$ mgCOD/ℓ, excess P removal can be achieved provided nitrate can be excluded from the anaerobic reactor, in which event, the degree of excess P removal depends on the magnitude of S_{bsi} above 50 mgCOD/ℓ and the influent COD strength S_{ti}. The sensitivity of excess P removal to the S_{bsi} concentration is such that any practice that reduces S_{bsi} should be avoided (see Chapter 3, Section 3) and conversely, any practice which increases S_{bsi} should be encouraged (see Chapter 3, Section 3 and Chapter 6, Section 7.1.3).

(4) In general, the establishment of a readily biodegradable COD concentration in the anaerobic reactor (S_{bsa}) in excess of 25 mgCOD/ℓ, to stimulate excess P removal, becomes increasingly more difficult as the raw influent COD strength of the wastewater (S_{ti}) decreases: For average readily biodegradable COD fractions ranging between 0,20 to 0,25, the minimum S_{ti} for which S_{bsa} is greater than 25 mgCOD/ℓ is about 250 mgCOD/ℓ; as S_{ti} increases above 250 mgCOD/ℓ, the establishment of S_{bsa} greater than 25 mgCOD/ℓ becomes increasingly easier and the excess P removal per unit influent COD increases and the attainment of excess P removal becomes less sensitive to external factors, e.g. DO control in the process.

(5) Cyclic flow and load conditions do not appear to affect the mean daily excess P removal, in fact, the removal often is improved, particularly under low average influent COD conditions. The reason for this improvement is that under the cyclic conditions (with constant recycle flows) during the peak periods the peak COD is higher than the daily mean causing a higher ($S_{bsa} - 25$) condition and hence higher P removals during the associated high flow period. However, under cyclic flow conditions the DO concentration in the aerobic reactor must be closely controlled; too high DO concentration reduces the nitrate removal whereas too low concentration may inhibit nitrification, inhibit excess uptake of P and adversely affect settling. The practical difficulties associated with oxygen control are such that serious consideration should be given to equalization tanks operated under a control strategy that equalizes *both* flow and load. Such a strategy is now available (Dold, Buhr and Marais, 1982) (see Chapter 10 on the operation of nutrient removal processes).

(6) For successful excess P removal in the Phoredox process, near complete denitrification must be achieved to avoid excessive nitrate discharge to the anaerobic reactor. Complete denitrification can be achieved only for TKN/COD ratios < 0,08* mgN/mgCOD for readily biodegradable COD fractions (f_{ts}) of 0,20 at 14°C. For TKN/COD ratios > 0,08* mgN/mgCOD, complete denitrification is unlikely to be achieved and the indicated processes for obtaining excess P removal are the UCT type processes.

(7) In the UCT type processes, by appropiate control of the mixed liquor a-recycle, the anaerobic reactor can be protected against nitrate discharge for TKN/COD ratios up to 0,13 mgN/mgCOD even though nitrate will be present in the effluent. For TKN/COD ratios < 0,11, the modified UCT process can achieve excess P removal and has the advantage over the UCT process of less need for operator intervention and is more likely to maintain reasonable good sludge settling characteristics. For TKN/COD ratios > 0,11 mgN/mgCOD, the only process that can achieve excess P removal is the UCT process, but it is possible that rather poor sludge settling characteristics will be obtained if the influent COD is high.

(8) For TKN/COD ratios > 0,14 mgN/mgCOD, it is unlikely that biological excess P removal can be obtained when complete nitrification is obligatory due to the inability to achieve sufficient denitrification without the addition of an external energy source. However, such high TKN/COD ratios are unlikely to be encountered in raw or settled municipal wastewaters.

(9) Primary sedimentation is unfavourable for achieving high removals of both N and P because it significantly increases the TKN/COD and P/COD ratios and significantly reduces the COD strength of the wastewater although the readily biodegradable COD concentration is only marginally affected. However, primary sedimentation significantly reduces the process volume requirements and total oxygen demand.

(10) For design of nutrient removal plants the influent wastewater characteristics (or their ranges) need to be known with much greater surety than for conventional activated sludge plants otherwise a design may be produced that fails totally in its P removal objectives. Vital influent characteristics are listed in (1) above. Useful data are the Total Alkalinity and the unbiodegradable COD fractions f_{us} and f_{up}. In the event that the vital characteristics are known only approximately, values should be selected that lead to conservative designs, usually at the cost of reduced N and P removal.

(11) The selected process must have factors of safety built into it or have flexibility in operation, so that if the influent characteristics are more adverse than accepted for design, it will be possible to accommodate them. For example, if a Phoredox process is selected it is recommended that normally the maximum TKN/COD ratio should not exceed 0,07*

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*See * on pages 7-30 and 7-32.

mgN/mgCOD and the process should have a built-in safety factor by designing it for 0,08*.mgN/mgCOD. For TKN/COD ratios > 0,07* mgN/mgCOD, the Phoredox/modified UCT/UCT process should be selected; with this process combination, by appropriate regulation of the recycles, excess P removal can be induced (subject to S_{bsi}). Theoretically, at low TKN/COD ratios (≅ 0,08*), the Phoredox process will be the optimal one, but practically its inflexibility to accommodate TKN/COD ratio variations which exceed 0,08* is a disadvantage; with the modified UCT/UCT process excess P removal can be obtained (subject to S_{bsi}) at the expense of N removal. By providing the various optional recycles, the best results can be obtained when required during the life of the plant.

(12) The selected process should require the least amount of decision making from the operator to obtain consistent results. In this regard the modified UCT/UCT process appears to lend itself most readily to setting operational procedures to absorb adverse conditions when these arise. Also operational adjustment to mitigate the regular changes in wastewater flow and characteristics observed over different days of the week can be preprogrammed.

10. REFERENCES and BIBLIOGRAPHY

BARNARD, J.L. (1975a) Biological nutrient removal without the addition of chemicals. *Water Research*, **9**, 485-490.

BARNARD, J.L. (1975b) Nutrient removal in biological systems. *Wat. Pollut. Control*, **74**, 2, 143-154.

BARNARD, J.L. (1976) A review of biological phosphorus removal in the activated sludge process. *Water SA.*, **2**, (3) 136-144.

DOLD, P.L., EKAMA, G.A. and MARAIS G.v.R. (1980) A general model for the activated sludge process. *Prog. Wat. Tech..*, **12**, 47-77.

DOLD, P.L., BUHR, H.O. and MARAIS, G.v.R. (1982) Design and control of equalization tanks. Research Report No. W 42, Dept. of Civil Eng., University of Cape Town.

EKAMA, G.A., SIEBRITZ, I.P. and MARAIS, G.v.R. (1982) Considerations in the process design of nutrient removal activated sludge processes. *Wat. Sci. Tech.*, **15**, 3/4, 283-318.

EKAMA, G.A., VAN HAANDEL, A.C. and MARAIS, G.v.R. (1979) The present status of research on nitrogen removal: a model for the modified activated sludge process. Presented at the Symposium on nutrient Removal, SA Branch of IWPC Pretoria, May 1979. Proceedings published by Water Research Commission of South Africa.

FUHS, G.W. and MIN CHEN (1975) Microbiological basis of phosphate removal in the activated sludge process for the treatment of wastewater. *Microbial Ecology*, **2**, 119-138.

FLORENTZ, M. and GRANGER, P. (1982) Phosphorus 31 nuclear magnetic resonance of activated sludge use for the study of the biological removal of phosphates from wastewater. Submitted for publication to *Biochemical and Biophysical Research Communications*.

HOFFMAN, R. and MARAIS, G.v.R. (1977) Phosphorus removal in the modified activated sludge process. Research report W. 22, Dept. of Civil Eng., Univ. of Cape Town.

LAW, S.M, DOLD, P.L. and MARAIS, G.v.R. (1982) Phosphate release and uptake by activated sludge in batch experiments. Research Report W 44 Dept. of Civil Eng., University of Cape Town.

MARAIS, G.v.R. and EKAMA, G.A. (1976) The activated sludge process Part I – steady state behaviour. *Water S.A.*, **2**(4) 163-200.

MARAIS, G.v.R., LOEWENTHAL, R.E. and SIEBRITZ I.P. (1982) Review: Observations supporting phosphorus removal by biological uptake. *Wat. Sci. Tech.*, **15**, 3/4, 15-41.

MARSDEN, M.G. and MARAIS, G.v.R. (1977) The role of the primary anoxic reactor in denitrification and biological phosphorus removal. Research Report W 19, Dept. of Civil Eng., University of Cape Town.

MARTIN, K.A.C. and MARAIS, G.v.R. (1975) Kinetics of enhanced phosphorus removal in the activated sludge process. Research Report W 14, Dept. of Civil Eng., University of Cape Town.

McLAREN, A.R. and WOOD, R.J. (1976) Effective phosphorus removal from sewage by biological means. *Water SA.*, **2**(1) 47-50.

NICHOLLS, H.A. (1975) Full scale experimentation on the new Johannesburg extended aeration plant. *Water SA.*, **1**(3) 121-132.

NICHOLLS, H.A. (1978) Kinetics of phosphorus transformations in aerobic and anaerobic environments. Presented at 9th IAWPR post conf. seminar, Copenhagen, *Prog. Wat. Tech.* **10**.

NICHOLLS, H.A. (1982) Application of the Marais-Ekama activated sludge model to large plants. *Wat. Sci. Tech.*, **14**, 581-598.

NICHOLLS, H.A., OSBORN, D.W. and MARAIS, G.v.R. (1982) Performance of large scale nutrient removal activated sludge plants. Research Report W 43, Dept. of Civil Eng., University of Cape Town.

RABINOWITZ, B. and MARAIS, G.v.R. (1980) Chemical and biological phosphorus removal in the activated sludge process. Research Report W 32, Dept. of Civil Eng., University of Cape Town.

RENSINK, J.H. (1981) Biologische defosfatering en processbepalende faktoren. In Symp. Book – Defosfatering niewe ontwikkelingen en praktijkervaringen in Nederland and Zweden. NVA Symp. (in Dutch).

SIEBRITZ, I.P., EKAMA, G.A. and MARAIS, G.v.R. (1980) Excess biological phosphorus removal in the activated sludge process at warm temperature climates. Proc. Waste Treatment and Utilization, Vol. 2, 233-251, Eds. C.W. Robinson, M. Moo-Young and G.J. Farquhar, Pergamon Press, Toronto.

SIEBRITZ, I.P., EKAMA, G.A. and MARAIS, G.v.R. (1983) Biological phosphorus removal in the activated sludge process. Research Report W 46, Dept. of Civil Eng., University of Cape Town.

SIEBRITZ, I.P., EKAMA, G.A. and MARAIS, G.v.R. (1982) A parametric model for biological excess phosphorus removal. *Wat. Sci. Tech.*, **15**, 3/4, 127-152.

SIMPKINS, M.J.N.P. and GERBER, A. (1981) Technical guide on biological nutrient removal: The Phoredox Process. Confidential report. SAIDCOR, Pretoria.

SIMPKINS, M.J. and McLAREN, A.R. (1978) Consistent biological phosphate and nitrate removal in an activated sludge plant. *Prog. Wat. Tech.*, **10**, (5/6) 433-442.

STERN, L.B. and MARAIS, G.v.R. (1974) Sewage as the electron donor in biological denitrification. Research Report W 7, Dept. of Civil Eng., University of Cape Town.

WILSON, D.E. and MARAIS, G.v.R. (1976) Adsorption phase in biological denitrification. Research Report W 11, Dept. of Civil Eng., University of Cape Town.

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*These values are valid for a readily biodegradable COD fraction (f_{bs}) of 0,24. Generally the higher (lower) the fraction, the higher (lower) these limits.

CHAPTER EIGHT

SECONDARY SETTLING TANKS

by

G.A. Ekama, A.R. Pitman, M. Smollen and G.v.R. Marais

1. INTRODUCTION

In the activated sludge process, it is necessary to separate the treated wastewater from the biological sludge mass thereby producing a clear final effluent. This solid/liquid separation phase is traditionally achieved by gravity sedimentation tanks although flotation has also been shown to be feasible.* Only solid/liquid separation by gravity sedimention in secondary settling tanks will be considered in this chapter.

The secondary settling tank is a vital part of the activated process. It combines the function of a clarifier (producing a clarified final effluent) and a thickener (producing a continuous underflow of thickened sludge for return to the biological reactor) (Figure 8.1). Should the settling tank fail in either of these two functions, sludge will be carried over the effluent weirs and escape with the effluent. Besides delivering an effluent of poor quality, loss of sludge could affect the behaviour of the biological process by uncontrolled reduction of the sludge age to values below that required for proper plant performance; for example, if the sludge age is reduced to below that required for nitrification, nitrogen removal by denitrification will cease.

The conditions in the biological reactor affect the settling and clarification characteristics of the sludge. For example, under-aeration reduces the settleability of the sludge; over-aeration may lead to pin-point floc formation and poor clarification even though the sludge maintains good settling characteristics and often in processes where large proportions of the sludge are intentionally unaerated as in nutrient removal processes, poor settling characteristics are encountered. Thus, the functions of the biological process and that of the secondary settling tank interact upon each other and the design of the one cannot be undertaken independently of the other – failure in either of these two unit processes causes failure of design objectives.

The settling characteristics of the sludges in activated sludge processes differ widely between different process types and also vary with time in the same process depending on the composition of the influent (see Section 7 below). For the various types of wastewaters and sludges to be treated, the design procedures for the secondary settling tanks have tended to be of an *ad hoc* nature – for each type of sludge, design rules were set down that experience has shown will provide generally for adequate functioning of the settling tank over the range of sludge settling behaviour that was found could arise. These rules often do not *directly* incorporate the settling characteristics of the sludge – these are lumped together with the hydraulic characteristics, as illustrated by the design rules taken from the IWPC designers' handbook:** (1) the overflow rate must not exceed 1 $m^3/m^2/h$ at Peak Wet Weather Flow (PWWF); (2) the retention time must not be less than 1,5 h at Peak Dry Weather Flow (PDWF); (3) maximum discharge per unit length of effluent weir is 8,3 $m^3/h/m$. These design rules implicity accept that the sludge settling characteristics will not fall below a certain minimum quality, a quality unknown but in conformity with past experience of similar sludges. Where processes are developed that generally tend to affect the settling characteristics adversely these rules may result in inadequate designs resulting in failure of the settler and the process.

Considerable research has been undertaken in unravelling the behaviour of secondary settling tanks and to develop a reliable design procedure that recognizes the settling characteristics of the sludge. However, the settling process is complex and the fac-

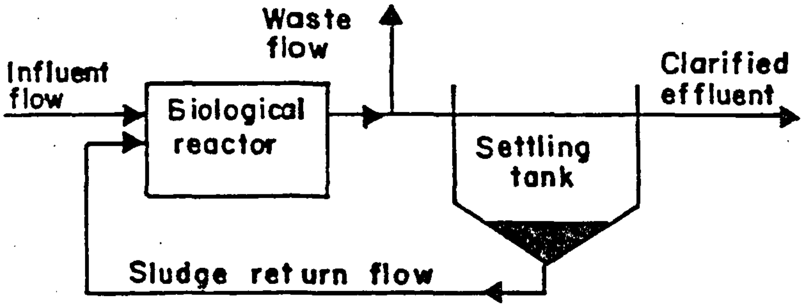

FIGURE 8.1: The activated sludge process incorporating the secondary settling tank for liquid-solid separation.

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*A design guide for solid/liquid separation and waste sludge thickening by flotation is available (Bratby and Marais, 1976, 1977).

**"Guide to the design of sewage purification works" prepared by the IWPC.

tors affecting settling so many that general procedures for design have not emerged. This is principally the case with activated sludge where constituents in the influent wastewater or conditions imposed on the process may radically change the settling characteristics of the sludge, and *ad hoc* rules specific only to certain categories of sludge have tended to be the usual method.

Lack of understanding of settling behaviour in settling tanks, and of its interaction with the process in both design and operation has been the source of considerable difficulties in the operation of activated sludge processes in South Africa. The first indication of failure has nearly always been a loss of sludge over the effluent weirs of the settling tank. This loss of sludge does not necessarily imply that the settling tank is at fault; the conditions in the biological reactor may have caused a deterioration in the settling characteristics of the sludge.

The design of the secondary settling tank based on the settling characteristics of the sludge and the interaction between the secondary settling tank and biological process are briefly discussed in this chapter.

2. SECONDARY SETTLING TANK DESIGN

The settling tank performance is not only governed by the overflow rate (clarification function) but also by the transmission of the applied solids to the bottom of the tank (thickening function). These two functions operate simultaneously but usually either one or the other will be critical during the daily cyclic operation of the settling tank. Conventional design procedures specifying only an overflow rate in effect tend to provide mainly for the clarification function. In contrast although research has been devoted to thickening, design criteria for thickening have not been developed to a large degree.

Earlier research into thickening has been mainly undertaken with non flocculant suspensions (e.g. Coe and Clevenger, 1916 and Kynch, 1952). These investigations have established the importance of the mass flux approach. Although activated sludge is a flocculant suspension, it has been shown that the mass-flux concept also can be applied to activated sludge (Dick, 1970 and 1972). Application of the mass flux concept to the settling tank design has been particularly assisted by the graphical method of Yoshioka *et al.* (1957), which allows estimation of the settling tank area for the thickening function (see Section 5.1 below).

The mass flux approach requires the measurement of the change in settling velocity of the sludge with change in sludge concentration as the sludge settles and thickens in the bottom of the settling tank. A method of assessing the settling behaviour of the sludge in the bottom of the tank is by means of the stirred batch settling test.

3. STIRRED BATCH SETTLING TEST

Mixed liquor from activated sludge plants usually show a strong flocculating tendency and even at low concentrations (1 000 mg MLSS/ℓ) exhibit a zone settling mode. In the zone settling mode, the concentration of sludge is such that the particles irrespective of size, all tend to settle at the same rate throughout the zone depth. The rate of settling is controlled by the rate at which the water passes upward through the mass and this is inversely related to the concentration. Although the mass settles as an entity leaving a clear liquid-mixed liquor interface, the particles are supported by the water between the particles, i.e. upper particles are not mechanically supported by lower ones.

Compression is observed at the bottom of the batch settling test. In the compression region each layer of particles provides mechanical support to the layers above it. Particle movement no longer is governed only by hydraulic frictional forces and the links between the particle. Forces transmitted by particle contact via the compressible properties of the sludge and the interstitial pressure caused by the squeezing out of water from the compressing particles govern the settling behaviour in the compression stage.

Consider a batch settling test tube filled with sludge of concentration such that zone settling commences immediately (see Figure 8.2). Initially the concentration is uniform throughout and region B occupies the total depth (Figure 8.2a). Immediately after settling commences, a solid-liquid interface develops and a region of clarified liquid A is formed (Figure 8.2b). In region B the particles still settle at uniform velocity under zone settling conditions, and throughout region B the concentration is constant. The interface settling velocity is a constant value and is a function of the concentration of region B, which is the same as the MLSS concentration at the start of the test. Simultaneously with the formation of region A, regions C and D are formed. Region D is one in which compression occurs and region C is the transition stage from zone settling to

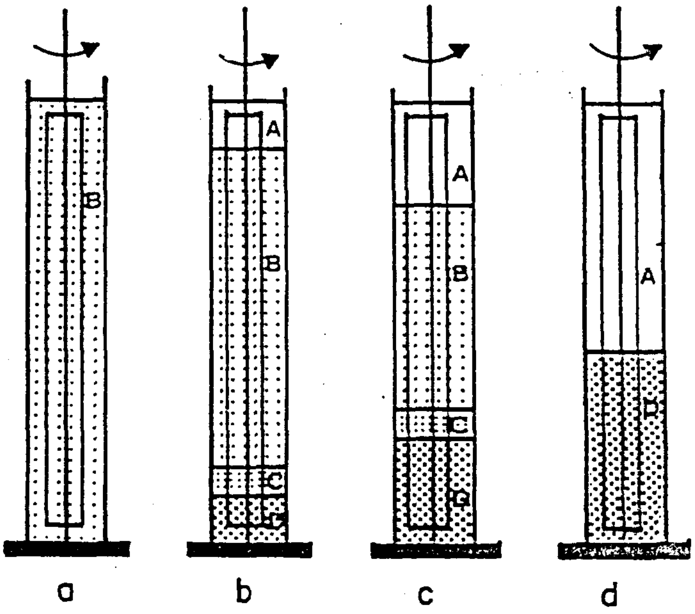


FIGURE 8.2: Chronological progress of a stirred batch settling test.

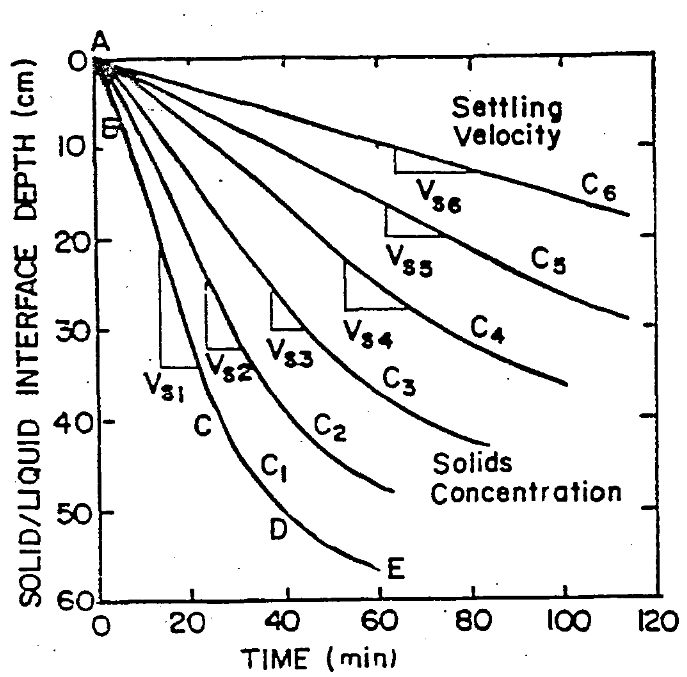


FIGURE 8.3: Solid-liquid interface height versus time observed in batch settling tests at different initial solids concentrations.

compression. Region C is still a zone settling region but the concentration increases down the zone. In a particular test, the transition zone depth generally is considered to remain constant until the solid-liquid interface reaches it.

A plot of the solid-liquid interface height versus time is given in Figure 8.3. The interface settling velocity is slow initially (A-B) because the activated sludge requires some time ro reflocculate from the disturbance of the sludge flocs during the filling of the test cylinder. To minimize this effect, it is necessary to fill the test cylinder from the bottom as slowly as practically possible so as to cause minimal disruption of the sludge flocs. After a time a constant interface velocity is observed (B-C) until region B sinks into region C. Thereafter the interface velocity decreases as denser and denser concentrations appear at the interface due to the subsidence of the less dense layers into the denser ones, until the compression region D is reached. Thereafter subsidence continues but now the velocity is governed by the compression behaviour of the sludge.

The interface velocity of the constant concentration zone settling region is given by the slope of line B-C in Figure 8.3. This velocity is defined as the settling velocity (V_s) of the sludge at the concentration equal to that of the uniform concentration in the test cylinder at the start of the test (X). The settling velocity V_s decreases as the sludge concentration increases (X) (Figure 8.3).

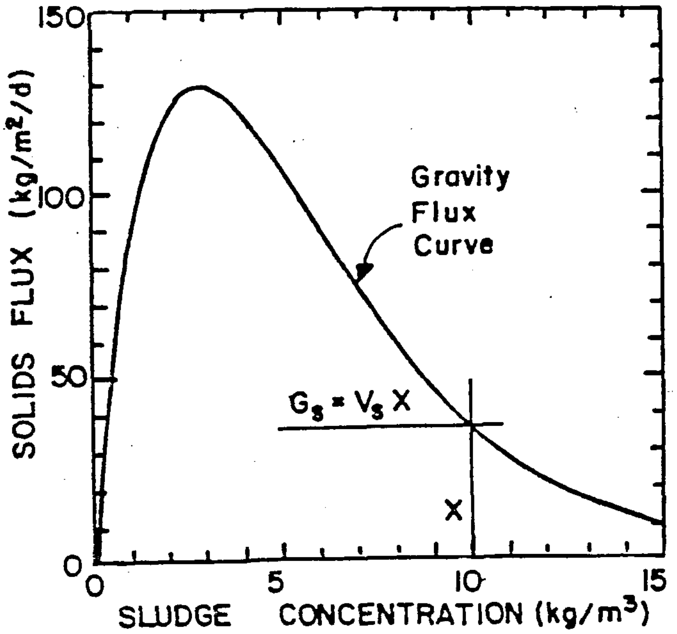


FIGURE 8.4: The gravity solids flux versus sludge concentration curve, i.e. downwards sludge transport due to gravity sedimentation.

4. SOLIDS FLUX

Solids flux is defined as the product of the settling velocity V_s and the concentration of the sludge, i.e.

$$G_s = V_s X \tag{8.1}$$

G_s = solids flux (kg/m²/d)

V_s = settling velocity (m/d)

X = sludge concentration (kgMLSS/m³)

By conducting a number of batch settling tests, the flux at different sludge concentrations can be determined. A typical flux curve for activated sludge is shown in Figure 8.4 in which solids flux (G_s) is plotted versus sludge concentration (X). At low X, although V_s is high the product G_s is low. At high X, V_s is low also giving a low G_s. At some intermediate X (about 3 to 5 kg/m³ for activated sludge), G_s is a maximum.

5. GRAPHICAL APPLICATION OF FLUX THEORY

5.1 Graphical procedure

In continuous settling tank operation (shown schematically in Figure 8.5), the sludge entering the settler is transferred to the bottom by two flux components (1) the gravity flux (G_s), and (2) the flux caused by the downdraft flow generated by the sludge abstraction flow from the bottom of the settler, called the bulk flux. The gravity flux is given by Eq. 8.1. The bulk flux (G_b) is

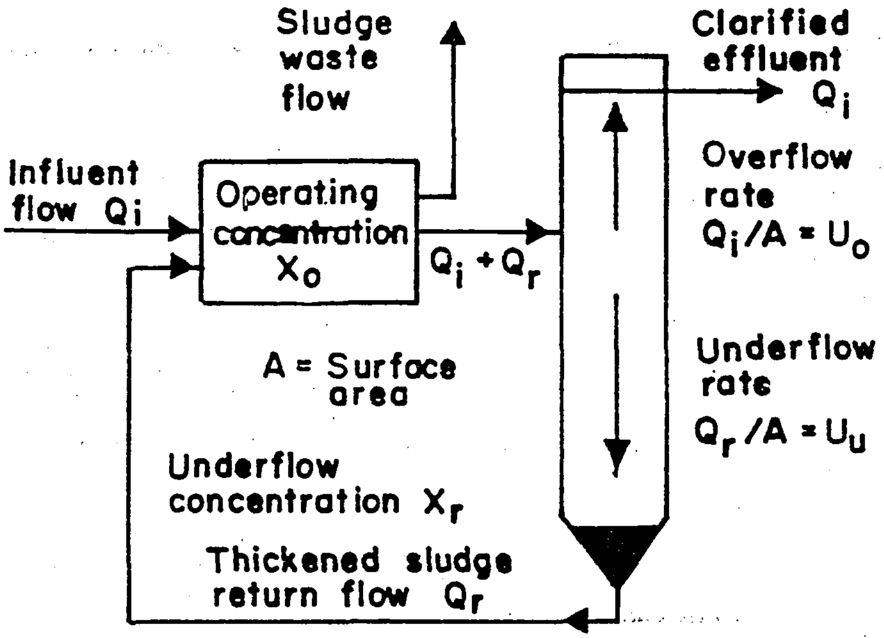


FIGURE 8.5: Idealized schematic representation of the secondary settling tank in the activated sludge process.

the product of the sludge concentration X and the underflow velocity (U_u) where the underflow velocity is the downdraft velocity caused by the underflow sludge abstraction i.e.

$$G_b = U_u X \tag{8.2}$$

and

$$U_u = Q_r/A \tag{8.3}$$

where

U_u = underflow velocity (m/d)
Q_r = underflow rate (m³/d)
A = surface area of settling tank (m²)

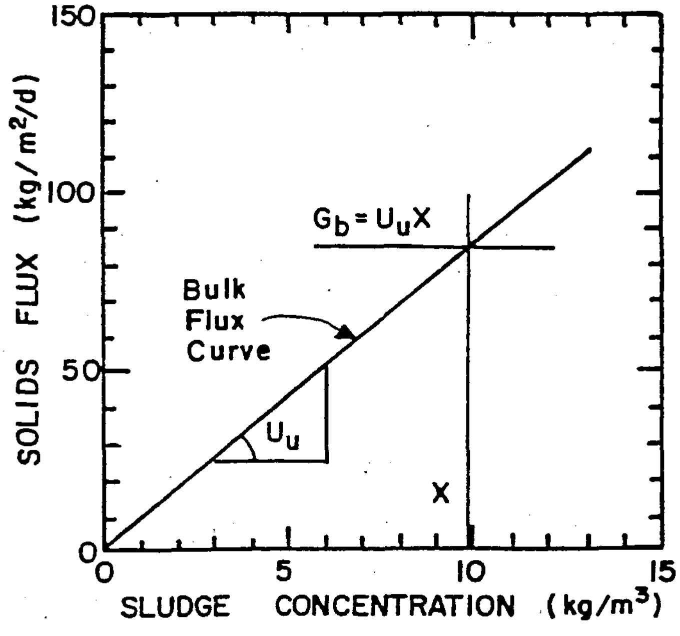


FIGURE 8.6: The bulk flux curve versus sludge concentration for fixed underflow rate (U_u) i.e. downward sludge transport due to the underflow sludge withdrawal.

For a fixed underflow rate, the bulk flux is proportional to the sludge concentration. This is depicted graphically in Figure 8.6.

The total flux (G_t) to the bottom of the settler is the sum of the two flux components i.e.

$$G_t = G_s + G_b$$

$$= X(V_s + U_u) \tag{8.4}$$

For a selected value of the underflow velocity U_u (i.e. Q_r and A), the two flux components can be added graphically (see Figure 8.7). For the particular choice of U_u, the total flux G_t attains a minimum value G_1 at a concentration X_1, and this minimum limits the rate at which sludge can reach the bottom of the settling tank. Hence to ensure that all the sludge reaches the bottom, the sludge mass applied to the settling tank per unit surface area i.e. the applied flux (G_{ap}) must be equal to or less than the limiting flux (G_1). The applied flux on the settling tank is given by the product of the sludge concentration in the reactor (X_o) and the combined recycle and influent flow ($Q_i + Q_r$) per unit area (see Figure 8.5). Hence, for safe operation of the settling tank

$$G_{ap} \leqq G_1 \tag{8.5}$$

where

$$G_{ap} = X_o(Q_i + Q_r)/A = X_o(U_o + U_u) \tag{8.6}$$

where

X_o = operating MLSS sludge concentration in biological reactor (kg/m³)

U_o = overflow rate

$$= Q_i/A \text{ (see Figure 8.5)} \tag{8.7}$$

Q_i = influent wastewater flow to process (m³/d)

G_1 = limiting flux (kg/m²/d)

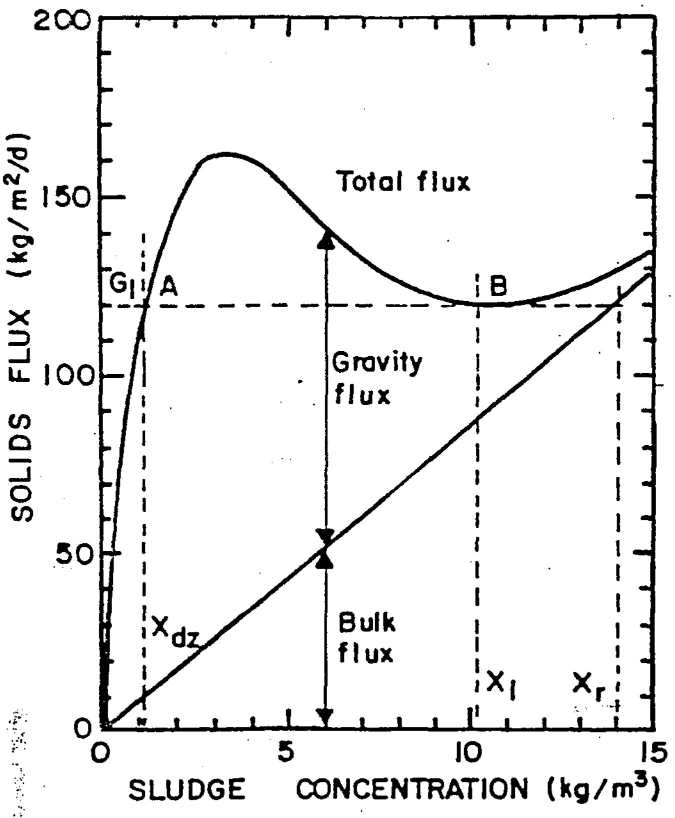


FIGURE 8.7: Total flux in settling tank versus sludge concentration i.e. downward sludge transport due to gravity sedimentation and underflow sludge withdrawal.

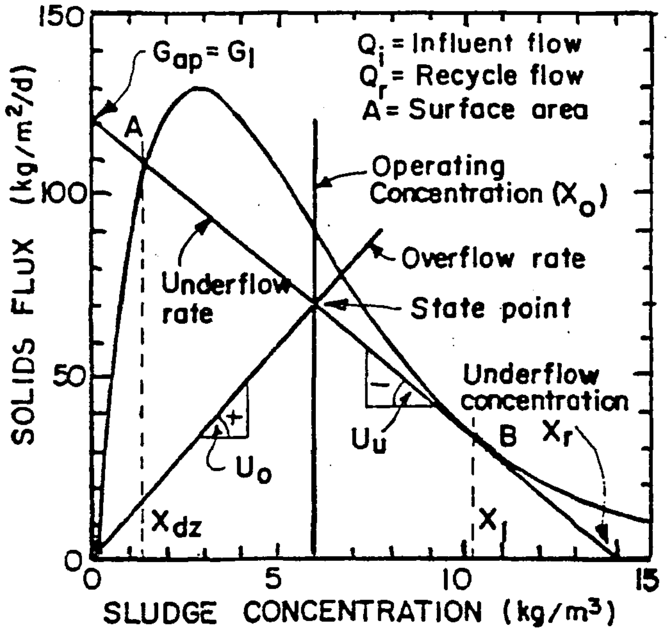


FIGURE 8.8: The gravity flux curve onto which is superimposed the settling tank behaviour; the state point defines the loading state of the tank and the underflow rate (U_u), overflow rate (U_o) and the operating solids concentration (X_o) lines intersect at the state point.

From a mass balance around the settling tank, if no sludge is lost over the effluent weirs, the mass of sludge applied is equal to the mass of sludge abstracted via the sludge underflow i.e.

$$Q_r X_r = X_o (Q_i + Q_r) \qquad (8.8)$$

where

X_r = MLSS concentration in underflow recycle (kg/m³)

For specified reactor MLSS concentration X_o and influent flow Q_i and selected underflow recycle Q_r, the area A is found by trial and error using the graphical construction shown in Figure 8.7. The correct value of A is that value which makes Eq. 8.5 an equality; values of A which make the LHS > RHS are too small which means settling tank failure in thickening will occur and values of A which make the LHS < RHS are too large which means the settling tank operates successfully but the area can be reduced.

The value of A found by the above procedure is valid only for the specified X_o and Q_i and selected Q_r. By repeating the procedure for different values of Q_r, different values of A will be obtained. For the proposed plant, the maximum area for the anticipated range of Q_r is the required design area.

The clarification criterion is satisfied if the overflow velocity (U_o) is less or equal to the settling velocity of sludge at the concentration at which it is discharged to the settling tank (X_o) i.e.

$$U_o \leqq V_s \text{ at } X_o \qquad (8.9)$$

Clearly the above procedure is extremely tedious because for every choice of Q_r, different total flux curves need to be constructed.

The above graphical design procedure has been considerably streamlined by Yoshioka *et al.* (1957), (see also Dick and Young, 1972; Pitman, 1980; and White, 1975) by introducing the concept of the *state point*. On the gravity flux plot (see Figure 8.8), the operating sludge concentration (X_o) is represented by a vertical line at the specified X_o value. The overflow rate ($U_o = Q_i/A$) is represented by a line (called the overflow line) from the origin with the slope equal to the overflow rate. The intersection point of the overflow line and the operating concentration line is called the state point. The underflow rate ($U_u = Q_r/A$) is represented by a line (called the underflow line) with the slope equal to the underflow rate passing through the state point (see Figure 8.8). The intersection of the underflow line with the horizontal axis gives the underflow concentration (X_r) (as given by Eq. 8.8) and that with the vertical axis gives the applied flux G_{ap} (as given by Eq. 8.6).

The state point defines the operating state of the settling tank and operating changes are reflected on the flux curve as follows: (i) a decrease in the underflow rate rotates the underflow line about the state point in an anticlockwise direction while the state point remains in position (see Figure 8.9a); (ii) an increase in overflow rate rotates the overflow line about the origin in an anticlockwise direction, moves the state point upwards along the operating sludge concentration line while the underflow line slope remains constant but continues to intersect the state point (Figure 8.9b), (iii) an increase in

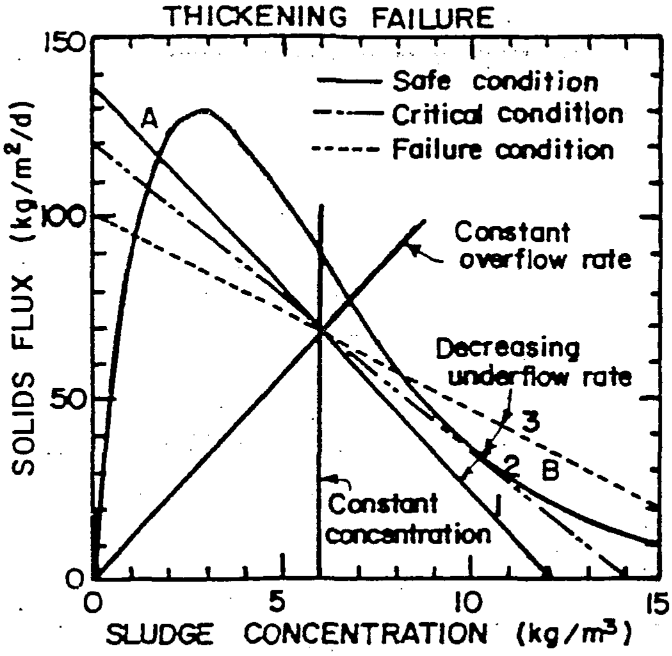


FIGURE 8.9a: Graphical representation of the settling tank operation conditions on the gravity flux curve showing decrease in underflow rate (U_u); (1) safe operating conditions, (2) critically loaded conditions and (3) overloaded conditions.

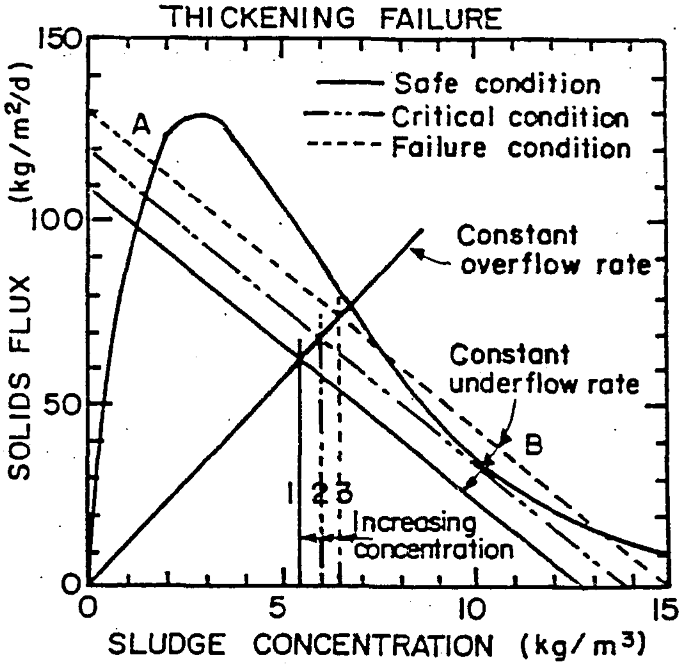


FIGURE 8.9c: Graphical representation of the settling tank behaviour on the gravity flux curve showing increase in operating solids concentration (X_o); (1) safe operating conditions, (2) critically loaded conditions and (3) overloaded conditions.

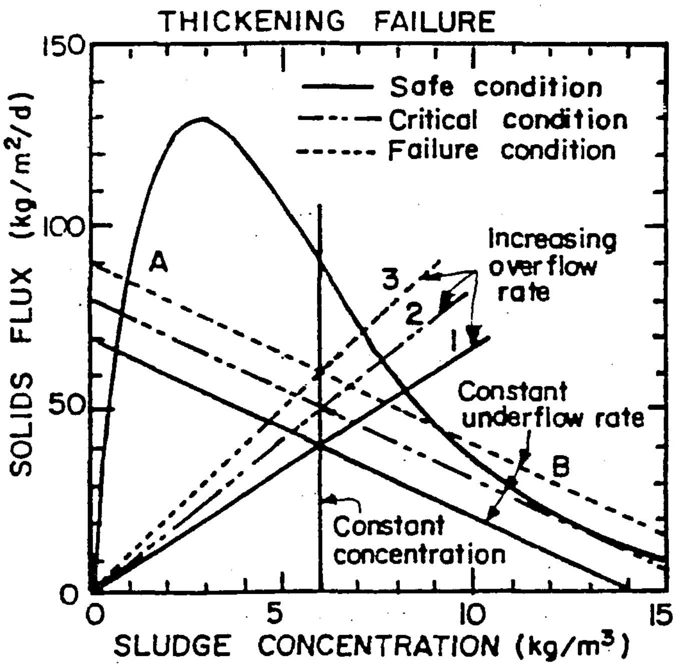


FIGURE 8.9b: Graphical representation of the settling tank operation conditions on the gravity flux curve showing increase in overflow rate (U_o); (1) safe operating conditions, (2) critically loaded conditions and (3) overloaded conditions.

operating solids concentration moves the operating solids line to the right and because the slopes of the underflow and overflow lines remain unchanged, the state point moves along the overflow line while the underflow line continues to pass through the state point (Figure 8.9c).

The state of the settling tank, i.e. whether or not it is in an under- or overloaded state, and in the overloaded state, whether failure will occur in its thickening or clarification function, can be obtained from the state point on the gravity flux curve. Referring to Figure 8.9a, b and c:

(1) When the state point is within the envelope of the flux curve, the clarification criterion is met and thickening governs the behaviour of the settling tank. When the underflow line
    (i) cuts the flux curve at one point only (i.e. near A), safe operating conditions prevail;
    (ii) cuts the flux curve at one point only (i.e. near A) and is tangential to the flux curve (i.e. near B), critically loaded conditions prevail;
    (iii) cuts the flux curve at 3 points (i.e. once near A and twice near B) overloaded conditions prevail.

(2) When the state point is on the flux curve critical conditions with respect to clarification prevail.

Under these conditions, thickening will also be critical if the underflow line is tangential to the flux curve on the state point; any underflow rate less than this will cause thickening failure.

(3) When the state point is outside the envelope of the flux curve, clarification failure conditions prevail (and thickening also because conditions 1(i) and 1(ii) above cannot be met).

(4) When the slope of the underflow line is too steep to make a tangent to the flux curve, but the state point is within the envelope of the flux curve, the thickening criterion no longer needs to be met and clarification only governs the design.

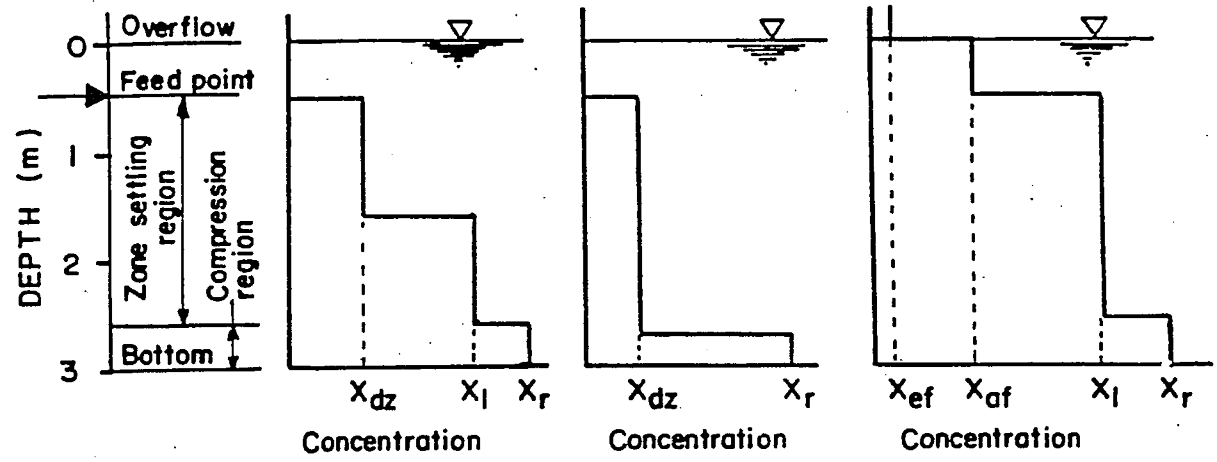
The above conditions can be readily applied to design. For given Q_i and X_o, select A which gives U_o (Eq. 8.7) and with X_o, fixes the state point. The area A must of course be selected such that the state point falls on or within the flux curve. On the state point, draw the underflow line such that it makes a tangent to the flux curve. From the slope of the underflow line and selected A, determine Q_r (Eq. 8.3). Notice that in this procedure Eqs. 8.5 to 8.9 are being solved graphically: the intercept of the underflow line with the vertical is the applied flux G_{ap} and when the underflow line makes a tangent to the flux curve, the applied flux equals the limiting flux G_l (see Eq. 8.5 and Figure 8.8).

Although the above procedure is still a trial and error one in that a repeated selection of A is made and the corresponding Q_r for safe operation calculated, it is far simpler than the total flux curve method because each selected A can be analysed on the same flux curve, thus obviating the need for constructing different total flux curves (Figure 8.7). Note that the two methods give identical results; the condition described by the horizontal line from G_l in Figure 8.7 is identical to that described by the underflow line tangential to the flux curve in Figure 8.8.

5.2 Concentration profiles

The gravity flux curve can also be used to plot idealized sludge concentration profiles in the settling tank (see Alkema, 1971; Pitman, 1978; White, 1975; Vesilind, 1968 and Dick and co-workers, 1967, 1970, 1972). The concentration profiles in the settling tank represented by the safe, critical and overloaded conditions in Figs. 8.9a, b and c are shown in Figure 8.10. The step changes in concentration shown in Figure 8.10 are idealized; in practice the changes are more gradual.*

At critically loaded conditions, three concentration zones occur in the settling tank (Figure 8.10a) (1) the underflow concentration zone in the bottom of concentration X_r where X_r is given by the intercept of the underflow line with the horizontal axis, (ii) the zone settling region where the concentration is equal to the limiting concentration X_l which causes the limiting flux G_l, where X_l is given by the concentration at the point



CRITICALLY LOADED (a) UNDERLOADED (b) OVERLOADED (c)

FIGURE 8.10: Sludge concentration versus depth profiles in the settling tank for (i) critically loaded conditions (Figure 8,10a), (ii) underloaded conditions (Figure 8,10b) and overloaded conditions (Figure 8.10c).

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*The depth required to allow these changes in concentration to take place that governs the design of the depth of the settling tank. The flux theory *per se* gives only the area requirement for thickening; the depth must be estimated from experience obtained in practice (see Section 7 below) and to allow a certain accumulation of sludge in the settling tank (see Section 5.3 below).

of tangency of the underflow line of the flux curve; and (iii) a dilute zone settling region above the zone settling region of concentration X_{dz} where X_{dz} is given by the concentration of the intersection point of the underflow line with the flux curve to the left of the turning point (i.e. near point A on Figures 8.9a, b and c. Under critically loaded conditions the fluxes of (i) the upper dilute zone, (ii) the zone settling region i.e. the limiting flux G_1, and (iii) the underflow withdrawal flux are all equal to the applied flux G_{ap} (Eq. 8.6) i.e.

$$G_{ap} = X_o (U_o + U_u) \quad (8.6)$$

$$= X_{dz} (V_{sxdz} + U_u) \quad (8.10a)$$

$$= X_1 (V_{sx1} + U_u) \quad (8.10b)$$

$$= X_r U_u \quad (8.10c)$$

where

V_{sxdz} and V_{sx1} is the settling velocity of the sludge at concentrations X_{dz} and X_1.

The zone settling region of limiting sludge concentration is not stable: the layer will contract or expand depending on whether the conditions are slightly underloaded or overloaded respectively.

When conditions change from a critically loaded to an underloaded condition (by reducing the overflow rate or operating sludge concentration or increasing the underflow rate, see Figure 8.9), the zone settling region will contract downwards until it sinks into the underflow concentration layer. The time taken for the layer to disappear depends on the rate of sludge withdrawal from the tank bottom by the underflow recycle. When steady state conditions again become established, only two concentration zones will be present, (see Figure 8.10b) i.e. the underflow concentration zone X_r and the dilute zone settling region; these concentrations can be found from Eqs 8.10a and c respectively. When conditions change from a critically loaded to an overloaded condition (by increasing the overflow rate or operating sludge concentration or reducing the underflow rate, (see Figure 8.9), the limiting concentration zone settling region will expand upwards until it reaches the height of the feed point (see Figure 8.10c). The solids transfer to the bottom of the settler is limited by the limiting flux G_1 and the difference between the applied flux G_{ap} and G_1 is the flux that causes the expansion of the zone settling layer i.e. sludge is being accumulated in the settling tank. The time taken for the zone settling layer to reach the feed point depends on the magnitude of the difference between G_{ap} and G_1, the larger the difference the faster the upward movement. After the zone settling layer reaches the feed point a sludge layer above the feed point develops (see Figure 8.10c). The concentration of this layer is different to the zone settling layer below the feed point because the direction of water flow above the feed point is upwards whereas below the feed point it is downwards (see Figure 8.5). The concentration of the sludge layer above the feed X_{af} can be found from the flux moving upwards, i.e.

$$X_{af} (U_o - V_{sxaf}) = G_{ap} - G_1 \quad (8.11)$$

where

V_{sxaf} = settling velocity of sludge at concentration X_{af}.

When this layer reaches the effluent overflow, sludge loss with the effluent will begin. The concentration that is lost with the effluent X_{ef} given by

$$X_{ef} = (G_{ap} - G_1)/U_o \quad (8.12)$$

When sludge is lost from the settling tank via the effluent flow, the operating concentration decreases. The operating concentration will continue to decrease until a new critical loaded steady state condition develops (see Figure 8.9c from overloaded to critically loaded conditions). The limiting flux G_1, limiting sludge concentration X_1 and underflow concentration X_r which define the sludge handling capacity of the settling tank are given by the vertical axis intercept, tangent point and horizontal axis intercept of the new underflow line tangential to the flux curve and parallel to the existing overload underflow line. Note that this new underflow line does not pass through the existing state point but that a new state point will develop at a lower operating concentration (see Figure 8.9c for overloaded to critically loaded conditions).

The above discussion shows that some time elapses between the onset of overloaded conditions and the time of sludge loss with the underflow. The length of time that elapses as well as the concentration of sludge that is lost depend on the severity of the overload (i.e. $G_{ap} - G_1$), the higher the overload, the shorter the time and the higher the concentration.

It should be noted that with overloaded conditions, the concentrations of the sludge profile are not given by the state point condition on the flux diagram. This is because, for the overflow, underflow and sludge operating concentrations lines to intersect at the state point a mass balance between sludge applied and sludge withdrawal by the underflow must be obtained i.e. Eq. 8.8 does not apply because sludge will eventually be lost via the overflow. If the overloaded conditions are allowed to persist sludge will be lost from the system thereby reducing the process sludge operating concentration. Consequently, for new stable conditions to develop under a continuous overload, it can be seen that a considerable mass of sludge must be lost to effect the required decrease in operating concentration.

5.3 Dynamic conditions

Under normal daily operation of the settling tank, the operation sludge concentration and underflow rate can be taken as remaining approximately constant. However, the influent flow and hence the overflow changes cyclically over the day (see Figure 8.9b). It may happen that under peak flow conditions, overloaded conditions develop, in which event the zone settling layer will begin to rise. The overload, depending on its severity, may persist for a few hours without any sludge loss, by which time the peak flow period may have passed and the zone settling layer again begins to drop. Consequently, over the day, the sludge layer will move up and down and the settler may even experience a temporary overload without sludge loss in the effluent (Tracy and Keinath, 1973).

The principal factor that prevents sludge loss during a temporary overload is sludge storage, which, once the area has been designed appropriately according to the flux theory, is directly proportional to the depth of the settler. Generally the deeper the settler, the greater the temporary overload that can be sustained without sludge loss. Unfortunately, no guidance regarding the design of the depth can be derived from the flux theory *per se* and the depth is best based on practical experience. A depth of 3 to 4 m appears to be adequate (see Section 7 below). Deep tanks also have the advantage that allowance is made for the various concentration zones such as the underflow concentration zone where sludge compression occurs. Under cyclic loading conditions the compression zone depth also varies in accordance with the load, being deeper at the peak flow period. Deep tanks are also less prone to disturbance of the settling zone sludge layers caused by hydraulic turbulence of the discharge and sludge withdrawal flows (see also Section 6.5 below).

5.4 Discussion

Clearly the graphical approach to the flux theory described above gives considerable insight into the design and operation of secondary settlers taking due consideration of the settling characteristics of the sludge. However, despite convenient developments in the graphical approach, the procedure remains rather tedious as for each sludge concentration, the settling velocity versus sludge concentration curve needs to be consulted to estimate the settling velocity. Perhaps the biggest drawback is that the differences between the settling characteristics of different sludges can only be described graphically.

6. ANALYTICAL APPLICATION OF FLUX THEORY

6.1 Empirical equation for settling velocity versus sludge concentration

In order to overcome the difficulties associated with the graphical approach, research has been undertaken to find an empirical relationship between the settling velocity and sludge concentration. Such a relationship would allow direct analytical solutions to the design equations.

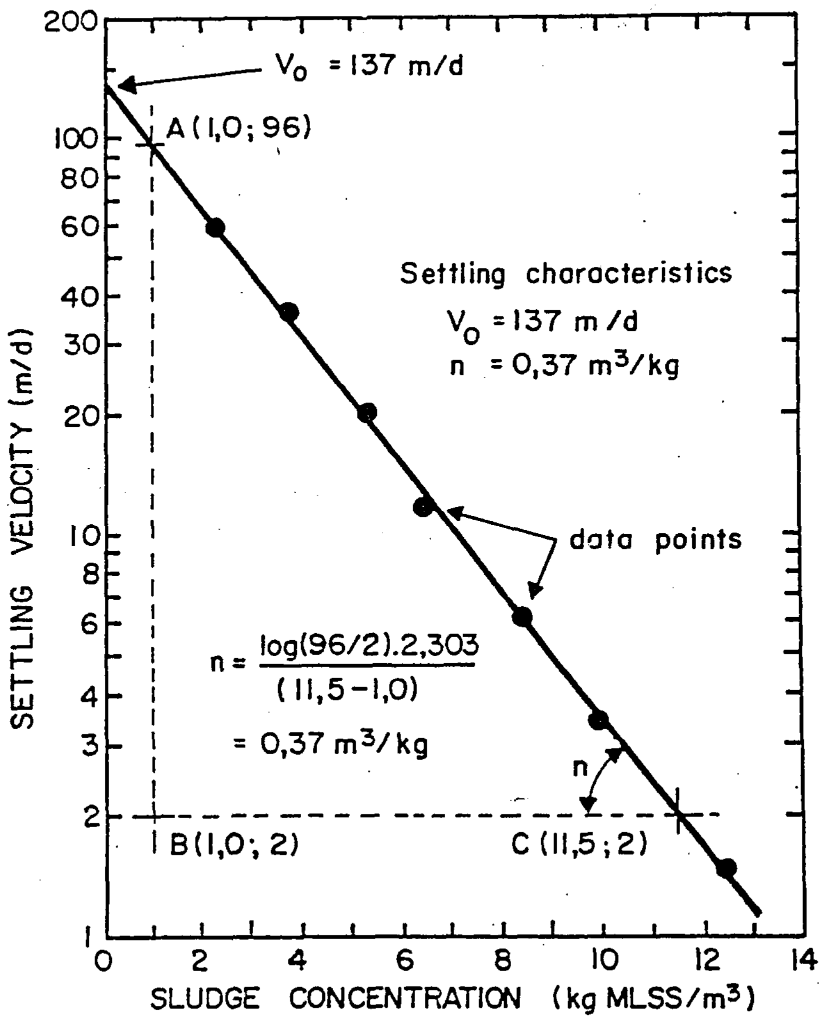

FIGURE 8.11: Determination of settling characteristics V_o and n from multiple batch test settling data.

Many mathematical expressions linking the settling velocity and sludge concentration have been proposed but the two most widely accepted are the following

$$V_s = V_o X^{-n} \qquad (8.13)$$

$$V_s = V_o e^{-nX} \qquad (8.14)$$

where

V_s = settling velocity (m/d)

X = MLSS concentration (kg/m^3)

V_o,n = constants describing settling characteristics of sludge.

The first expression can be represented by a straight line on a log V_s versus log X plot (i.e. log – log), the second by a straight line on a log V_s versus natural X plot (i.e. semi-log). Considerable research effort has been expended to establish the applicability of each expression to flux theory – Dick and Young, 1972 and Pitman, 1978 using Eq. 8.13 and Vesilind, 1968a, 1968b and White, 1975 using Eq. 8.14. A comprehensive comparison is presented by Smollen (1981), who concluded that Eq. 8.14 is superior to Eq. 8.13 because it (1) gives a theoretically consistent description of the observed gravity flux curve with defined turning and inflexion points and (2) gives a closer correlation with data measured on laboratory, pilot and full scale plants over a number of years ($r^2 > 0{,}95$). Consequently in this chapter, Eq. 8.14 is accepted for describing the interface zone settling velocity versus sludge concentration in the batch settling test.

The settling characteristics V_o and n are readily obtained by linear least squares regression analysis of multiple batch test settling data over a range of sludge concentrations from 1,5 kg/m^3 to 15 kg/m^3 on the expression (see Figure 8.11)

$$\log V_s = \log V_o - nX/2{,}303 \qquad (8.15)$$

Substituting Eq. 8.14 for V_s into Eq. 8.1 yields the gravity flux in terms of the concentration and settling characteristics i.e.

$$G_s = X V_o e^{-nX} \qquad (8.16)$$

Equation 8.16 describes the gravity flux curve shown in Figure 8.4.

6.2 Mathematical properties of theoretical flux equation

The slope of the flux curve is given by the derivative of Eq. 8.16 i.e.

$$dG_s/dX = V_o e^{-nX}(1 - nX) \qquad (8.17)$$

which, when set to zero, gives a turning point at $X = +1/n$ and a maximum flux of V_o/ne. Differentiating Eq. 8.17 with respect to X and substituting 1/n for X shows that the turning point with coordinates ($X = 1/n$; $G = V_o/ne$) is a maximum. Setting the second derivative to zero shows that there is an inflexion point in the flux curve at coordinates ($X = 2/n$; $G = 2V_o/ne^2$).

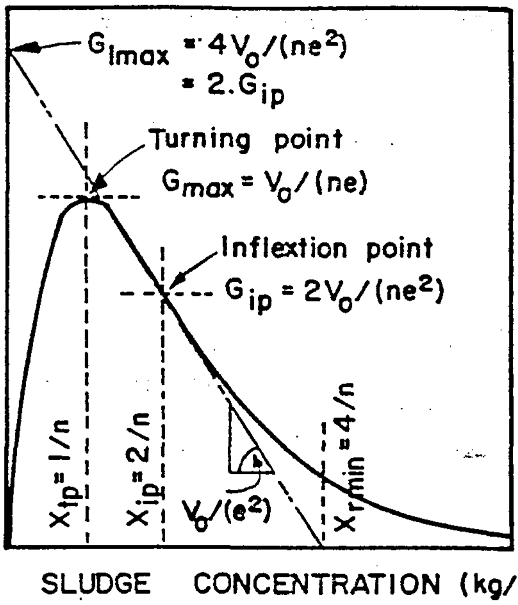

FIGURE 8.12: Mathematical properties of the flux curve based on the settling velocity-sludge concentration equation $V_s = V_o \exp(-nX)$.

From an inspection of the flux curve, no tangential underflow line can be constructed on the flux curve (see Section 5.1 above) with a slope greater than that of the flux curve at the inflexion point. The intercepts of the tangent at the inflexion point with the vertical and horizontal axes therefore give respectively the maximum limiting flux (G_{1max}) and the minimum underflow concentration (X_{rmin}) for thickening to govern the design of the settling tank. The slope of this tangent is $-V_o/e^2$ (from Eq. 8.17), and hence

$$G_{1max} = 4V_o/ne^2 = 2.G_{IP} = 2/e.G_{max} \qquad (8.18)$$

$$X_{rmin} = 4/n \quad = 2.X_{IP} = 4.X_{TP} \qquad (8.19)$$

where subscripts IP and TP refer respectively to the inflexion and turning points.

The above features of the flux curve are shown in Figure 8.12.

6.3 Application of flux equation to settling tanks

Defining s as the underflow recycle ratio, i.e.

$$s = Q_r/Q_i \qquad (8.20)$$

the maximum overflow rate to satisfy the thickening criterion is found from Eqs. 8.3 and 8.5 to 8.8 and is given by

$$Q_i/A = G_1/\{X_o(1 + s)\} = G_1/(X_r s) \qquad (8.21)$$

where

$$X_r = (1 + s)X_o/s \qquad (8.22)$$

The equation of the underflow line with a point of tangency to the flux curve is

$$G = G_1(1 - X/X_r) \qquad (8.23)$$

and its slope is

$$dG/dX = -G_1/X_r \qquad (8.24)$$

At the point of tangency (see Figure 8.8) $X = X_1$, and both the fluxes and the slopes of the flux curve and the underflow tangent are equal i.e. Eq. 8.16 = Eq. 8.23 and Eq. 8.17 = Eq. 8.24. From these equations taking X_r as known, solving for X_1 and, ignoring the impractical solution, yields

$$X_1 = (X_r/2)\{1 + \sqrt{1 - 4/(nX_r)}\} \tag{8.25}$$

Substituting Eq. 8.25 into Eq. 8.16 gives the limiting flux i.e.

$$G_1 = V_o X_r \frac{(1 + \alpha)}{(1 - \alpha)} \exp\{-nX_r(1 + \alpha)/2\} \tag{8.26a}$$

where

$$\alpha = \sqrt{1 - 4/(nX_r)} \tag{8.26b}$$

If $\alpha < 0$, there is no solution for G_1. Setting $\alpha = 0$ yields $X_r = 4/n$ and hence $G_1 = 4V_o/ne^2$ which corresponds to X_{rmin} and G_{1max} respectively as found above (see Figure 8.11). No solution for G_1 is possible for $X_r < 4/n$ because the curvature of the flux curve is such that no valid tangential underflow line can be constructed within the envelope of the flux curve. This result can be interpreted as the thickening criterion not being important in a design when $X_r < 4/n$, leaving only the clarification criterion to be met (see condition 4 in Section 5.1 above).

Substituting Eq. 8.22 for X_r into Eq. 8.26 and substituting Eq. 8.26 into Eq. 8.21 yields

$$\frac{Q_i}{A} = \frac{V_o}{s}\frac{(1+\alpha)}{(1-\alpha)} \exp\{-n(1+s)X_o(1+\alpha)/(2s)\} \tag{8.27a}$$

where

$$\alpha = \sqrt{1 - 4s/\{n(1+s)X_o\}} \tag{8.27b}$$

Equation 8.27 relates the overflow rate to the recycle ratio, s, for a selected value of the sludge concentration in the reactor and settling characteristics of the sludge to meet the thickening criterion. If $\alpha = 0$, $X_o = 4s/(1+s)n$. Hence from Eq. 8.27, if $\alpha = 0$

$$\frac{Q_i}{A} = V_o/(e^2 s) \tag{8.28}$$

Consequently, if Q_i/A is less than $V_o/(e^2 s)$ then both the thickening and clarification criteria have to be met. If Q_i/A is greater than $V_o/(e^2 s)$, then thickening apparently is not a governing criterion and the clarification criterion only has to be met. This criterion is met if the overflow rate is less than the settling velocity of the sludge at the feed concentration i.e. from Eq. 8.9

$$Q_i/A = V_o e^{-nX_o} \tag{8.29}$$

6.4 Settling tank design and operating chart

By plotting Q_i/A versus s from Eqs. 8.27 to 8.29 for different values of X_o, a steady state design and operating chart for the settling tank is obtained (Figure 8.13). The hyperbola on the diagram described by Eq. 8.28 distinguishes between the domain in which both the thickening and clarification criteria have to be met (area between axes and hyperbola) and in which the clarification only has to be met (area above hyperbola). The

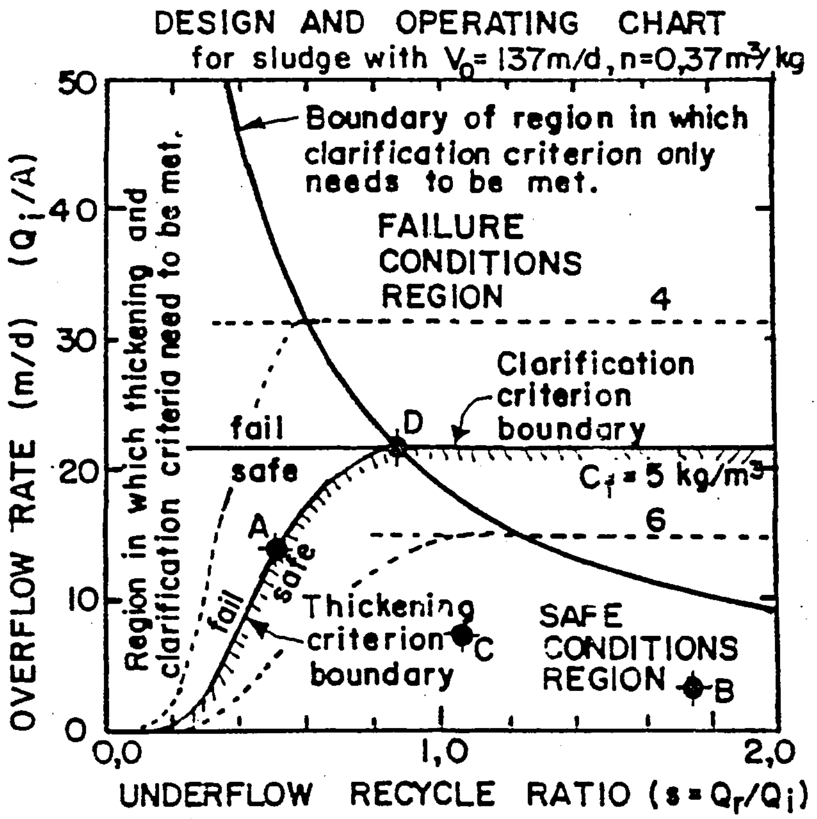

FIGURE 8.13: Design and operating chart for secondary settling tanks showing the different types of failure that can occur for different combinations of overflow rate and underflow recycle ratio.

thickening criterion is given by Eq. 8.27 which, for different X_0 values, is a family of curves from the hyperbola to the origin. These curves show that as underflow recycle ratio s decreases the allowable overflow rate decreases. The clarification criterion is given by Eq. 8.29 and being independent of s, is a family of horizontal lines. In the reqion below the hyperbola, the thickening and clarification criteria have to be met simultaneously. In some instances in this region, the allowable overflow rate is higher for thickening than for clarification so that clarification is the governing criterion for the settling tank. Consequently for a specified reactor concentration X_0, *failure* of the settling tank is represented by overflow rate-underflow recycle ratio data pairs falling above the thickening and the clarification lines.

It should be noted that the design and operation chart is also valid for daily cyclic conditions. For successful operation of the settling tank, the conditions prescribed by the design and operation chart need to be met at any instant of the day. For example, for a fixed settling tank area (A) and underflow rate (Q_r), as the influent flow to the plant (Q_i) increases, so the overflow rate (Q_i/A) increases and the recycle ratio (s) decreases, which moves the intersection point of the overflow rate and recycle ratio upwards to the left in Figure 8.13. At each instant of the day, the overflow rate and recycle ratio intersection point must subscribe to the thickening and clarification criterion in the diagram i.e. must at all times be below the thickening and clarification lines.

6.5 Design example

Assume that the MLSS concentration of the nutrient removal processes designed in Chapters 4, 5, 6 and 7 is 5 kg/m^3. For the nutrient removal process, the settling characteristics are selected to be rather poor i.e. V_0 = 137 m/d or 5,7 m/h and n = 0,37 m^3/kg giving V_0/n = 15,4 kg/m^2/h (see Table 8.1). The average dry weather flow (ADWF) is 13,3 Mℓ/d and assume that over the day the flow varies in a cyclic pattern with a minimum (MDWF) of 0,6.ADWF and a peak (PDWF) of 2,0.ADWF, and the peak wet weather flow (PWWF) is estimated to be 3,0.ADWF.

For the settling characteristics of the sludge the design and operating chart is constructed as set out above (see Figure 8.13). Assume that at PWWF, clarification must be achieved and that the thickening criterion will be met by adjusting the underflow recycle. The PWWF is 3.13,3 = 40 Mℓ/d. From Figure 8.13 the maximum overflow rate at X_0 = 5 kg/m^3 is 21 m/d (given by horizontal clarification criterion line). Hence the settling tank area is 40.10^3/21 = 1905 m^2. In order that thickening failure does not occur under PWWF conditions, the recycle ratio at PWWF must be greater than 0,82 (given by the s value of the intersection point of the clarification line and the hyperbola, see point D on Figure 8.13). Hence at PWWF of 40 Mℓ/d the underflow Q_r should be 0,82.40 = 32,8 Mℓ/d or 2,46 times ADWF otherwise thickening failure will occur (i.e. for s < 0,82, the Q_i/A and s intersection point will fall above the thickening criterion line in Figure 8.13).

Under dry weather conditions, the overflow rate at PDWF is 2,0.13,3.10^3/1905 = 14 m/d. In order that thickening failure does not take place at PDWF, the underflow recycle ratio s must be at least 0,52 (given by the s value of the intersection point of the horizontal Q_i/A = 14 m/d line and the thickening criterion line for X_0 = 5 kg/m^3 see point A on Figure 8.13). Hence the underflow should be at least 0,52.26,6 = 13,9 Mℓ/d or approximately 1:1 based on ADWF. Accepting the settling tank area of 1905 m^2 and fixing the recycle flow at a constant 14 Mℓ/d, the overflow rate and recycle ratio at ADWF are 7 m/d and 1,05:1, at MDWF are 4,2 m/d and 1,75:1 and the PWWF are 14 m/d and 0,52:1 respectively. Hence in the design and operating chart, the locus of the points defining the operating conditions over the day under cyclic influent flow moves between points of coordinates Q_i/A = 14 m/d and s = 0,52 at PDWF (see point A in Figure 8.13) and Q_i/A = 4,2 m/d and s = 1,75 at MDWF (see point B in Figure 8.13) through Q_i/A = 7,0 m/d and s = 1,05 at ADWF (see point C in Figure 8.13). From the chart, all these points represent safe operating conditions with respect to thickening and clarification and hence under dry weather conditions, the settling tank should operate satisfactorily. Provision for wet weather flow was made up to a PWWF of 3.ADWF but the underflow recycle needs to be increased to 32,8 Mℓ/d to accommodate this. For lower PWWF, the underflow need not be increased so high. The required underflow can be determined from the design and operating chart.

The above design is based on the accepted settling characteristics V_0 = 137 m/d and n = 0,37 m^3/kg and should the actual characteristics be better than these the settling tanks will still operate satisfactorily. However, if the settling characteristics are poorer than the accepted ones, difficulty in operating the settling tank may be encountered. To a degree, a deterioration in settling characteristics can be accommodated under dry weather conditions by increasing the underflow recycle but this strategy depends on how poor the characteristics are below the accepted ones; however, to increase the underflow recycle beyond the value where the thickening criterion no longer needs to be met will not improve the situation and will be wasted on pumping costs. Under peak wet weather conditions, sludge will probably be lost depending on the duration of the peak wet weather flow.

High peak wet weather flows can be accommodated by the settling tank by allowing them to bypass the main biological process and discharging them to the effluent from the last aerobic reactor. Although the overflow rate in the settling tank still increases, the applied flux to the settling tank does not increase due to the dilution of the mixed liquor concentration by the influent. The effect of this is to discharge to the settling tank a significantly reduced operating sludge concentration. In contrast, if the storm flow is discharged to the head of the biological process, both the overflow rate and applied flux would increase and a large mass of sludge would be displaced from the biological process to the settling tank. The emergency strategy of bypassing storm flows is permissible because usually storm flows are very low in COD, TKN and P concentrations and a short contact period with the mixed liquor in the settling tank will significantly reduce the COD and TKN; however, a higher P concentration may have to be accepted for the short duration of the storm flow.

7. PRACTICAL ASPECTS OF FINAL CLARIFIER DESIGN

The basic purposes of acivated sludge final clarifiers are to produce a clear effluent free of suspended solids and a thickened sludge for recycle to the inlet of the process. However, in nutrient removal versions, two additional requirements are desired. The quality of the underflow sludge should be such that a large mass of nitrate is not recycled to the anaerobic zone. Also, if the anaerobic conditioning of the activated sludge can be initiated in the bottom of the clarifier it will benefit phosphate removal.

Full-scale experience on sloping bottom scraped clarifiers has shown that the solids in the sludge layer move in semi-plug flow and despite their endogenous nature, can denitrify quite well. For example, up to 5 mgN/l nitrate can be lost on passing through this layer and if denitrification is complete, anaerobic conditions are initiated. The phosphate released under these anaerobic conditions will not cause problems as it is usually trapped in the sludge layer and moves out with the return sludge before diffusing into the effluent.

If the sludge can be encouraged to thicken well in the clarifier, denitrification will be favoured even more. For example, if the sludge is thickened up to give a recycle concentration of 20 000 mg/l instead of 10 000 mg/l, half the mass of nitrate will be returned even if no denitrification occurred at all. However, at 20 000 mg/l solids the ratio of biomass to nitrate is very large and despite the endogenous nature of the bacteria present, very good denitrification will occur in most cases using up all the available nitrate. Under this situation there will be no nitrate feedback to the anaerobic zone. Of course, the ability to thicken up to 10 000 mg/l solids will depend on the sludge properties and will usually only happen with low SVI sludges (i.e. less than 100 ml/g). In this respect operators should be encouraged to ensure that such sludges are produced in this process.

The denitrification and anaerobic conditions that can be encouraged in final clarifiers can only be of benefit to the process and designers should bear this in mind in clarifier design.

As mentioned earlier, good denitrification has been noticed on full-scale sloping bottom scraped clarifiers as well as in conical bottom (60° cone) settling tanks. However, these effects are less pronounced on flat-bottomed suction scraped clarifiers, as these are designed to lift settled sludge off the floor before it has the opportunity to denitrify. It has also been found that suction type clarifiers tend to return a relatively weak sludge and if attempts are made to thicken the sludge and so obtain better denitrification, the sludge will refuse to pass up the withdrawal tubes due to its increased viscosity. Also, the rotating suction arm tends to cause hydraulic instability in the tank at high recycle rates.

Therefore with the view to encouraging denitrification in final clarifiers and the initiation of anaerobic conditions in the return sludge, these clarifiers should incorporate the following features:

(i) they should preferably be circular and a sloping bottom with mechanised scrapers to move sludge to a central bottom withdrawal point,

(ii) mixed liquor to be introduced near the surface at the centre with the inlet system being well baffled to destroy kinetic energy of the influent,

(iii) the side water depth should be at least 4 m,

(iv) the sludge withdrawal system should be designed to handle sludges with a higher viscosity than water.

The clarifier surface area has a very important influence on performance. It is recommended that the solids flux theory (described in Sections 5 and 6 above) be used to determine this. The Vesilind equation:

$$V = V_0 e^{-nx}$$

which describes the settling of activated sludge is an important aspect of this theory with the two factors V_0 and n varying with sludge settling properties. Experimentally derived values of these factors can show a large variation depending on the degree of filamentous growth in the sludge.

Typical results obtained are shown in Table 8.1 in order of deteriorating settling properties.

It was found that the ratio V_0/n is a very good indicator of settling properties which correlates with more wellknown parameters such as SVI. Thus, it can be seen in Table 8.1 that Plant D has a poor settling sludge; the sludge in Plant A settles fairly well. High values of this ratio (i.e. 30 kg/m²/h) indicate good settling properties, while low values (i.e. 10 kg/m²/h) are found with poor settling sludges. Designers must decide what types of sludges are likely to occur in their plant during its lifetime. If they are sure that poor settling sludges will never be encountered, a V_0/n ratio of 30 or above can be used but if poor settling sludges are encountered, values of less than 15 must be chosen. However, this decision must be taken carefully as a value of 15 will lead to a clarifier design with twice the surface area as would be obtained using $V_0/n = 30$. It should also be noted that experience to date has shown that biological nutrient removal plants can be prone to the encouragement of filamentous growths which give rise to poor settling sludges so that a conservative approach to clarifier design might be prudent.

Plant operation can also have a dramatic effect on activated sludge settling properties and hence V_0/n values. An important variable is the dissolved oxygen concentration in aerobic zones. It has been found both in South Africa and overseas that plants run at low dissolved oxygen concentrations (below 1,0 mg/l and in particular, below 0,5 mg/l) are prone to the growth of filamentous bacteria which are the cause of poor settling sludges. Designers should therefore bear in mind the quality of operation that will be available, in particular, skills for the control of dissolved oxygen input and process overloads.

Another factor to be remembered is that the solids-

TABLE 8.1 SETTLING CHARACTERISTICS OBSERVED AT FOUR DIFFERENT ACTIVATED SLUDGE PLANTS IN JOHANNESBURG

Plant	V_o (m/h)*	n (m^3/kg)	V_o/n (kg/m^2/h)*	SVI (mℓ/g)
A. NORTHERN WORKS MODULE 2				
mean value	7,90	0,29	27,24	100
best result	9,49	0,28	35,89	65
worst result	6,07	0,35	17,34	180
B. NORTHERN WORKS MODULE 1				
mean value	7,94	0,36	22,06	120
best result	9,49	0,27	35,16	70
worst result	7,02	0,40	17,55	170
C. GOUDKOPPIES MODULE 2				
mean value	6,68	0,40	16,70	210
best result	9,84	0,31	31,7	70
worst result	3,84	0,37	10,4	300
D. GOUDKOPPIES MODULE 1				
mean value	5,95	0,43	13,8	250
best result	8,96	0,41	21,9	130
worst result	5,20	0,55	9,5	300

*To convert to units per day, multiply by 24.

flux theory covered in this chapter assumes an idealised settling tank which will not occur in practice and although it has been found that some full-scale clarifiers obey the flux model within 20%, designers should make use of safety factors either in the initial choice of V_0 and n or applied to the final clarifier area calculated from this model.

8. REFERENCES

ALKEMA, K.L. (1971) The effect of settler dynamics on the activated sludge process. MSc thesis, Dept. of Chemical Engineering, Univ. of Colorado.

BRATBY, J.R. and MARAIS, G.v.R. (1976) A guide for the design of dissolved-air (pressure) flotation systems for activated sludge processes. *Water SA*, 2 (2) 87-100.

BRATBY, J.R. and MARAIS, G.v.R. (1977) Aspects of sludge thickening by dissolved-air flotation, *Wat. Pollut. Control*, *77*, (3) 421-432.

COE, H.S. and CLEVENGER, G.H. (1916) Methods for determining the capacities of slime settling tanks, *Trans. Am. Inst. Mining Eng.*, **55**, 356.

DICK, R.I. (1967) Evaluation of activated sludge thickening theories, *J. Sanit. Eng. Div., ASCE*, **93**, SA4.

DICK, R.I. (1970) Role of activated sludge final settling tanks, *J. Sanit, Eng. Div., ASCE*, **96**, 423.

DICK, R.I. and YOUNG K.W. (1972) Analysis of thickening performance of final settling tanks, Prod. 27th Ind. Waste Conf., Purdue Univ., Series No. 161, Lafayette, Indiana.

KYNCH, G.J. (1952) A theory of sedimentation, *Trans. Faraday Soc.*, **48**, 166.

PITMAN, A.R. (1980) Settling properties of extended aeration sludges, *JWFCF*, **52**, 3.

SMOLLEN, M. (1981) Behaviour of secondary settling tanks in activated sludge processes, MSc thesis, Dept. of Civil Eng., Univ. of Cape Town.

TRACY, K.D. and KEINATH, T.M. (1973) Dynamic model for thickening of activated sludge, Water 1973, **70**, 136, Chem. Eng. Progress, Symp. Series AICHE.

VESILIND, P.A. (1968a) Design of prototype thickeners from batch settling tests, *Wat. and Sew. Works*, **115**, 302.

VESILIND, P.A. (1968b) Discussion of evaluation of activated sludge thickening theories, *J. Sanit. Eng. Div., ASCE*, **94**, SA1, 185-190.

WHITE, M.J.D. (1975) Settling of activated sludge. Technical Report TR11, Water Research Centre, Medenham, England.

YOSHIOKA, N, HOTTA, Y., TANAKA, S., NAITO, S. and TSUGAMI, S. (1957). Continuous thickening of homogenous flocculated suspensions, *Chem. Eng., Tokyo*, **21**, 66.

CHAPTER NINE

MICROBIOLOGICAL ASPECTS

by

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1. EUKARYOTIC AND PROKARYOTIC CELLS

The basic units of living organisms are cells. Organisms are divided into two major sub-groups on the basis of the internal structure of their cells. Two distinct types of cells exist, i.e. eukaryotic and prokaryotic cells. Eukaryotic cells are the more complex and although their complexity may vary in different organisms, it is the basic structural unit of plants, animals, rotifera, protozoa, fungi and algae. Protozoa, algae and fungi are relatively simple organisms and have been placed in a sub-group of the eukaryotes, that is, the protists. This is done to distinguish them from the more complex plants and animals. The prokaryotic cell is a less complex structure and is the structural unit of bacteria.

2. PROTISTS IMPORTANT IN ACTIVATED SLUDGE

Fungi and algae are not considered to be important in activated sludge.

2.1 Protozoa

Protozoa exhibit a wide range of form and way of life. Three classes are important in activated sludge. The *Rhizopoda* engulf food particles within the *pseudopodia* by which they move, i.e. phagotrophic ingestion; soluble foods may, however, also be absorbed through the cell membrane, i.e. osmotrophic ingestion. The *Mastigophora* move by means of whiplike flagella and these organisms may also be phagotrophic and/or osmotrophic. The *ciliophora* are motile by means of numerous short hairlike projections which are termed cilia. The *Ciliophora* in activated sludge are mostly phagotrophic. The *Ciliophora* are divided into free-swimming types, for example, *Paramecium*, crawling types, e.g. *Aspidisca* and stalked types, e.g. *Vorticella*.

2.1.1 Metabolism and growth

Metabolism is the process whereby an organism obtains the necessary energy and cell building blocks to survive and proliferate. Protozoa respire aerobically, i.e. require oxygen for their metabolism. Organisms that ingest soluble matter are the primary feeders and agents of purification; they thus form what is termed the basic trophic level. Phagotrophic animals occupy higher levels in the system in that they prey on the primary feeders, i.e. they form the second trophic level.

A definite succession of protozoa occurs when activated sludge forms and as it becomes more efficient. The food of the primary feeders is the organic matter in the sewage, thus bacteria (osmotrophic feeders) and osmotrophic protozoa, mostly *rhizopods* and *flagellates*, are in direct competition for their food supply. The osmotrophic protozoa cannot compete with the more efficient bacteria and as the latter increase in number, the osmotrophic protozoa are eliminated. Generally then, the osmotrophic *Rhizopoda* and flagellates occur in systems just starting up, in inefficient systems or when organic concentrations are high. The phagotrophic flagellates then appear before the ciliates because they have relatively lower energy requirements than the ciliates. However, because of the less efficient feeding mechanism of the flagellates the ciliates start predominating. Within the ciliates there is also a succession of species. Free-swimming ciliates predominate when there are a large number of free-swimming bacteria and their presence thus points to poor flocculation. The attached and crawling ciliates have lower energy requirements and thus replace the free-swimming forms as their food becomes limited. The attached and crawling forms, together with higher animals such as rotifers, indicate a well-flocculated sludge and a low COD effluent, i.e. they indicate that the number of free-swimming bacteria have been reduced below the demands of the free-swimming ciliates.

3. PROKARYOTES

The prokaryotes (bacteria) are small osmotrophic unicellular organisms. Some of these organisms can form multicellular filamentous forms of growth (e.g. *Sphaerotilus*) and some can form a mycelial (fungal-like) structure (*Actinomycetes*). Although a certain amount of morphological variation exists among these organisms, their differences are primarily expressed in terms of metabolic behaviour, i.e. the organisms are classified in terms of what they do rather than on their appearance.

The average size of the prokaryotes is considerably smaller than that of the protists. Whereas the volume of a structural unit of the eukaryotes is 5 to 150 000 000 μm^3, that of the prokaryotes is 0,01 – 5 000 μm^3. Because the rate of metabolism is inversely proportional to the size of the cell, the small size of bacteria confers a selective advantage where bacteria must compete with protists for available nutrients.

3.1 Nutritional requirements

Bacteria are extremely diverse in their requirements for nutrients and this property forms a basis for their classification.

3.1.1 Carbon

Most organisms obtain their carbon from organic nutrients for satisfying their biosynthetic and energetic requirements. Such organisms are termed heterotrophic. Some organisms can obtain their energy from the oxidation of inorganic compounds such as ammonia (*Nitrosomonas* and *Nitrobacter*). These organisms obtain their carbon for cell synthesis from CO_2 and these latter organisms are termed autotrophic.

Organisms such as *Pseudomonas* and *Acinetobacter* can use many different compounds as carbon and energy sources, whereas others are extremely limited.

3.1.2 Nitrogen and sulphur

Nitrogen and sulphur occur in the building blocks of cells. Many cells can obtain these elements from nitrates and sulphates, others must obtain the elements in a reduced form, e.g. as ammonium salts. These elements may also be present in the organic material metabolised by the bacteria and thus become available to them.

3.1.3 Growth factors

Organic compounds which cells require, but cannot synthesize, are known as growth factors, i.e. these compounds must be supplied to the organisms. Organisms vary greatly in their growth factor requirements and this property exerts a considerable selectivity effect in any given system. This is one of the reasons why organisms such as *Pseudomonas* and *Acinetobacter*, which have very low or no growth factor requirements, proliferate in activated sludge, whereas fastidious organisms like the enteric bacteria, rapidly die off.

3.2 Bacterial energetics

The basis for the diversity of chemical activities among bacteria and for the extensive array of extreme environments in which they are to be found, lies in the methods they have available for obtaining the energy to drive metabolism. These organisms are thus capable of using various electron acceptors as alternatives to oxygen, viz photosynthesis, anaerobic respiration and fermentation.

3.2.1 Aerobic respiration

Organisms that require molecular oxygen (O_2) for respiration, i.e. molecular oxygen functions as the terminal electron acceptor (oxidising agent), are termed obligate aerobes.

3.2.2 Anaerobiosis

Two general mechanisms exist which permit bacteria to grow in the absence of oxygen, viz. fermentation and anaerobic respiration. While certain species are classified as facultative anaerobes in that they are able to adapt their metabolism to grow either in the presence of oxygen or its absence, others are obligate anaerobes and can only grow in an oxygen-free environment.

(i) Anaerobic respiration

The principal electron acceptors which can serve as an alternative to oxygen are nitrate, nitrite, sulphate and carbon dioxide.

(ii) Nitrate and nitrite as terminal electron acceptors

Some facultative anaerobes, in the absence of oxygen can make use of nitrate as the terminal electron acceptor where the oxygen in the nitrate ion is used instead of molecular oxygen. Nitrate respiration is the term used for the reduction of nitrate to nitrite. It is a widespread reaction and can be carried out by many genera including *Pseudomonas*, *Bacillus*, *Clostridium* and various *Enterobacteriaceae*. Denitrification is the term used for reduction through nitrite to nitrogen. This process is confined to a limited number of species, e.g. *Paracoccus denitrificans* and *Thiobacillus denitrificans*. Both these organisms can use nitrate as the terminal electron acceptor and CO_2 as carbon source.

(iii) Sulphate as terminal electron acceptor

These suphate reducing organisms are strict anaerobes and their respiratory mechanism can only be coupled with sulphate or CO_2. Their nutrition is restricted in that only a limited range of substrates can serve as electron donors. There are basically only two genera, i.e. *Desulfovibrio* and *Desulfotomaculum*. These organisms thus couple their substrate oxidation to the reduction of sulphate to sulphide. Their electron donors are limited to the oxidation of hydrogen, pyruvate, lactate, ethanol, formate or malate, which are converted to acetate and CO_2. These organisms are thus only found in sulphate-rich anaerobic environments in association with other organisms which excrete products such as lactate.

(iv) Fermentation

This is the process where oxidation of one substrate is coupled to the reduction of another. Fermentation occurs in a number of bacterial groups and genera. i.e. the lactic acid and propionic acid bacteria, the *Enterobacteriaceae*, *Staphylococcus*, *Bacillus* and *Clostridium*. All *Clostridia* are obligate anaerobes and extremely sensitive to even small quantities of oxygen. The propionic acid bacteria are also strict anaerobes but do have some oxygen tolerance. The other organisms are facultative anaerobes and are thus capable of aerobic respiration in the presence of oxygen. Fermentation processes have a low yield of energy and hence growth for the amount of nutrient consumed; these processes lead to the excretion into the surrounding medium of fermentation products such as short chain fatty acids and alcohols.

3.3 Summary of nutritional categories among bacteria of importance in activated sludge

The most simple classification is that based on the nature of the energy and principle carbon source. Organisms that are dependent on a chemical energy source (this is applicable to all organisms in activated sludge) are termed chemotrophs. Phototrophs are organisms that can use light as the energy source. Organisms able to use CO_2 as principle carbon source are termed autotrophs and organisms dependent on an organic carbon source are termed heterotrophs.

3.3.1 Chemoautotrophs

These organisms use a chemical energy source and CO_2 as the principal carbon source. Energy is obtained by the oxidation of reduced inorganic compounds such as NH_3, NO_2, reduced forms of sulphur (H_2S, S) and ferrous iron. Ammonia is oxidised to nitrite by organisms belonging to the genera *Nitrosomonas*, *Nitrosospira*, *Nitrosococcus* and *Nitrosolobus*. Nitrite is then oxidised to nitrate by *Nitrobacter*, *Nitrospira* and *Nitrococcus spp*.

Reduced sulphur compounds are oxidised by strictly aerobic organisms and also by the organism *Thiobacillus denitrificans* which can use nitrate as the terminal electron acceptor. The *Thiobacillus spp* are the most widespread sulphur oxidising organisms. The filamentous *Beggiatoa-Thiothrix* group also oxidise sulphide. Ferrous iron is oxidised to ferric iron by organisms such as *Thiobacillus ferro-oxidans*, *Ferrobacillus ferro-oxidans* and the stalked bacteria of the genus *Gallionella*. The filamentous *Sphaerotilus* and *Leptothrix spp* can also oxidise Fe^{2+} but these two latter *spp* are heterotrophs and their importance in Fe^{2+} oxidation is unclear.

3.3.2 Chemoheterotrophs

The chemoheterotrophs include rotifers, protozoa, fungi and most bacteria. These organisms use a chemical energy source and an organic substrate as the carbon source. Because these organisms obtain their energy from the oxidation of organic compounds they are often termed chemo-organotrophs. The bulk of the bacteria in activated sludge are considered to be gram-negative, chemoheterotrophic (chemo-organotrophic) rod-shaped bacteria.

4. SEWAGE AS A SUBSTRATE FOR MICRO-ORGANISMS

All micro-organisms require certain basic nutrient elements. Apart from the carbon required for energy and synthesis, adequate supplies of various other elements must be present in assimilable form. All micro-organisms require nitrogen, phosphorus, sulphur, potassium, calcium and magnesium. Trace elements such as iron, copper, zinc and cobalt are also required by many species and more fastiduous organisms also require growth factors such as vitamins. Carbon, nitrogen and phosphorus are required by micro-organisms in balanced amounts, the ratios can differ but a reasonable estimate for the ratio is COD:N:P of 100:6:1,5. In the activated sludge process however, this ratio can be mostly different. Experience with full scale nutrient removal plants has shown up ratios more in the region of 100 total COD:3 total N to 0,5 total P. This is because only a fraction of the incoming constituents are involved in bacterial synthesis. Large quantities of inert or non-biodegradable material can be present which are removed by processes other than metabolism (i.e. adsorption onto floc particles). Nitrogen is only considered fully available when present as ammonia and phosphorus as soluble phosphate, although other forms of these elements can be converted to ammonia and phosphate and thus become available.

Domestic sewage usually provides a nutritionally balanced food with the necessary elements and vitamins for bacterial activity. However, for the processes required for denitrification and enhanced phosphorus removal, the organic carbon source is often insufficient.

5. ORGANISMS PRESENT IN SEWAGE AND THEIR FATE IN ACTIVATED SLUDGE

The predominant bacteria in sewage are those found in human intestines. However, most intestinal bacteria are strict anaerobes and die rapidly on leaving their host. The coli-aerogenes group is probably the principal group of organisms entering a plant; these intestinal bacteria function optimally at 37 °C, and as activated sludge plants operate at 15 to 15 °C, these organisms do not function effectively. Furthermore, the dilute nutrient concentration in an activated sludge plant also does not favour the proliferation of the intestinal bacteria. Thus, most of the organisms entering the plant with the sewage, die rather than grow. The organisms that do proliferate in a plant enter by wind or infiltration into sewers. Such organisms include species of *Pseudomonas*, *Alcaligenes* and *Acinetobacter*. The temperatures and low nutrient concentrations in activated sludge plants are natural conditions for the saprophytic organisms.

6. SELECTIVE PRESSURES AND THE SUCCESSION OF ORGANISMS IN THE ACTIVATED SLUDGE ENVIRONMENT

6.1 General

In activated sludge the relative number of each species will be determined by their growth rates, availability of suitable food and predation. Furthermore, physical conditions prevailing in the plant will have a different effect on the rate of proliferation of various species. Such factors include temperature, availability of oxygen, pH and mode of agitation. Of the many species entering the plant only a relatively small number will find the environment suitable and these species will dominate the population to differing degrees. However, because of successive different conditions in the plant and influent, it is impossible for specific species to totally predominate. The organic waste to be treated forms the food of the primary feeders, i.e. the bacteria and osmotrophic protozoa. Phagotrophic protozoa and rotifers prey pn these osmotrophic forms. The autotrophic bacteria require less complex energy sources and are thus not in competition with the heterotrophic organisms. The numbers of predators are influenced by the number of prey and this is also influenced by the extent to which the prey can seek refuge from the predator. The degree of flocculation is significant in this respect, as also is the extent to which the bacteria can clump or form clusters too large for the predators to ingest. Obviously organisms which cluster will also be retained in the system whereas free-swimming forms will be washed out.

The dominant organisms in terms of activity are not necessarily those occurring in the largest numbers but rather those which constitute the bulk of the biomass and/or those that make the largest contribution to the total metabolic activity.

The dominant bacteria will be those capable of most effectively utilising the organic waste and of forming flocs to ensure their retention in the system. The nature of the bacteria in the plant is thus chiefly determined by the nature of the organic waste and secondly by conditions in the plant, chiefly the degree of aeration and particularly anoxic, anaerobic and aerobic fractions of the total solids retention time and the magnitude of the latter in nutrient removing activated sludge processes.

6.2 Selection of organisms in an aerobic environment

The dominant aerobic bacteria present in such systems appear to be gram-negative rods belonging to the genera *Pseudomonas, Achromobacter, Alcaligenes* and *Flavobacterium*. *Pseudomonas* and *Acinetobacter* species are versatile chemoheterotrophs and can use a wide range of common substrates as sole energy and carbon sources. *Acinetobacter spp* can generally only utilise short chain carbon compounds as carbon and energy sources. If these compounds are present in a sewage *Pseudomonas spp* will generally predominate with low agitation but in a well-agitated medium *Acinetobacter spp* will often predominate. The reason for this phenomenon is that whereas the *Pseudomonas spp* are motile organisms and thus capable of migrating towards food and oxygen, the *Acinetobacter spp* are non-motile. The latter organisms, although metabolically very active and competitive, thus need to be brought into contact with their substrates and oxygen.

6.3 Selection of organisms in a nitrogen removing anoxic/aerobic system

In a totally aerobic system certain aerobic heterotrophs and autotrophs will predominate, depending principally on the nature of the sewage.

The inclusion of a zone devoid of oxygen but containing nitrates as electron acceptor will cause organisms capable of nitrate respiration and denitrification to proliferate. These organisms have a preference for short chain organics as substrates and will thus deplete these compounds present in the influent sewage. This in turn will affect the nature of the populations established in the aerobic zone, as aerobic organisms which can only utilise short chain organics will not be able to grow unless short chain organics can be produced extracellularly from adsorbed longer chain organics.

6.4 Selection of organisms in an anaerobic/anoxic/aerobic system

As substrate and mode of aeration are the principal factors in the establishment of microbial populations, the combination of an anaerobic/anoxic/aerobic sequence in an activated sludge process will result in a restructuring of populations in comparison to an aerobic or anoxic/aerobic system.

In the present context, anaerobic refers to the absence of oxygen and nitrates, and anoxic refers to the presence of nitrates and the absence of dissolved oxygen. The anaerobic zone will lead to the enrichment of organisms capable of fermentation, i.e. organic compounds serve as electron donors and electron acceptors. Organisms normally living in soil and water and capable of fermentation, are *inter alia*, species of *Enterobacter*, *Serratia*, *Proteus* and *Aeromonas*. These organisms are also capable of aerobic respiration and would thus be capable of surviving the alternating anaerobic/aerobic conditions prevailing in the plant. As fermentation is an inefficient mode of energy production, the cell yield per unit mass of substrate utilised is considerably less than under aerobic conditions and the population increase of the fermentating organisms will thus be limited.

The fermentative organisms mentioned are also capable of growing under aerobic conditions. However, the transition from a fermentative metabolism to a respiratory metabolism does not occur rapidly and consequently, by the time these organisms have switched their metabolism to a respiratory mode in the aerobic zone, available substrate will probably have been utilised by the obligate aerobes. This will thus limit the population increase of these organisms under aerobic conditions. The conversion from a respiratory mode to a fermentative mode of metabolism is rapid and the organisms have the necessary pathways operational for fermentation before the DO has reached zero, i.e. fermentation can commence virtually immediately when the organisms enter the anaerobic zone.

During fermentation compounds such as lactic, succinic, propionic, butyric, acetic acids and ethanol are produced, and as these compounds cannot be utilised further under anaerobic conditions, they are excreted into the medium. These highly biodegradable compounds can then be used in the anoxic zone for denitrification and the balance thereof will pass onto the aerobic zone where aerobic organisms, particularly *Acinetobacter sp*, which can store PHB in the anaerobic zone, can then proliferate. It has also been suggested that certain aerobic organisms can accumulate the organics excreted in the anaerobic zone by the fermentative organisms and store these compounds in the form of internal carbon and energy reserves, probably as poly-β-hydroxybutyrate (PHB). As *Acinetobacter spp* are metabolically very active the availability of their preferrred and prestored intracellular substrates in the presence of oxygen could thus lead to their proliferation in the aerobic zone. It has been shown at the Contra Costa Phostrip plant in California that up to 10 hours of anaerobiosis has no effect on respiratory rates when aerobic activated sludge organisms are once again exposed to oxygen.

Anaerobic conditions thus probably do not adversely affect the aerobic populations established in the aerobic zone.

It has been shown that certain *Acinetobacter spp* predominate in activated sludges removing excess amounts of phosphorus and these organisms contain large inclusions of polyphosphates. These organisms have also been observed to release some of this accumulated phosphorus under anaerobic conditions, thus providing a probable explanation for the increase in ortho-phosphate often observed in the anaerobic zone. Chemical adsorption/desorption effects could also be relevant to the nett release and uptake observed in anaerobic zones. In enhanced-phosphorus-removing sludges the cells are mainly present in large clusters containing hundreds of cells. Intracellular polyphosphate inclusions can be observed in activated sludge by staining dried smears of mixed liquor on microscope slides for 1 min with an aqueous solution of 1 % methylene blue plus 1 % borax. The slide is rinsed in water, allowed to dry and observed by light microscopy. Polyphosphate inclusions stain a purplish colour, whereas the cells and sludge matrix stain a blue colour. The property of a dye to change its colour when it reacts with certain compounds, is termed metachromacy, i.e. polyphosphates stain metachromatically with methylene blue. That these purplish inclusions are rich in phosphorus has been confirmed by electron microscopy combined with microprobe analysis. By the latter method is has been estimated that the polyphosphate granules contain in excess of 25 % P (m/m) and in excess of 8 % Ca (m/m). It has also been observed that these inclusions can occupy the bulk of the cell volume.

Although a controversy still exists as to whether *Acinetobacter spp* are the principal organisms responsible for enhanced phosphorus uptake, the dominance of these organisms has been shown in many enhanced-phosphorus-removing plants. The assay of these organisms has been performed by plating, biochemical and immunofluorescent techniques. The abundance of these organisms has also been indicated by the observation that protozoa in such sludges contain mainly the characteristic *Acinetobacter* cells.

Acinetobacter possesses a relatively characteristic morphology, i.e. short, plump rods 1,0 – 1,5 µm by 1,5 – 2,5 µm in logarithmic phase approaching coccus shape in stationary phase and predominantly in pairs and short chains. The cells are very large compared to most other bacteria. *Acinetobacters* are non-motile, gram-negative and oxidase-negative.

An *Acinetobacter* species isolated from enhanced-phosphorus-removing sludge has been acknowledged as a distinct species, the organism is designated *A phosphodevorans* (i.e. devourer of phosphorus).

If these organisms are the principal agents of enhanced-phosphorus uptake, the question still to be answered is how these organisms can be enriched and maintained within a system, i.e. what causes these organisms to have a selective advantage within the activated sludge environment.

The fact that many plants performing enhanced phosphorus removal receive septic sewage, points to the substrate nature as being important in enriching these organisms. The augmentation of the feed to the 50 Mℓ/d plant with the supernatant of an acid digester has been performed with great success, whereas the augmentation of the same plant with a similar additional COD but not of acidic nature, did not succeed in inducing enhanced P-uptake. Whether other factors such as the exposure of the organisms to an anaerobic condition is necessary still needs to be elucidated.

It thus appears as if the anaerobic zone provides suitable substrates for the proliferation of the aerobic phosphorus-accumulating bacteria. This is supported by the observation that the mixed liquor of phosphorus-removing sludges is often very black in colour. The sulphate-reducing bacteria can only use a restricted range of compounds, e.g. lactate or malate as their source of carbon and energy. Such compounds are rare in fresh sewage and the sulphate reducers must therefore utilise the metabolic products of fermentation processes which in turn utilise more complex organic compounds, for example, cellulose. The sulphide formed by these bacteria reacts with ferrous iron to form black ferrous sulphides. The production of sulphide thus indicates that fermentative processes do occur in plants with black MLSS. The presence of sulphides also encourages the growth of *Beggiotoa* and *Thiothrix spp* of filamentous organisms often observed in phosphorus-removing sludges. The filamentous *Sphaerotilus Leptothrix* group are also encouraged to grow in such anaerobic systems as their preferred substrates are also formed during fermentation.

Nitrates are inhibitory to fermentation processes and it may be for this reason that their presence in the anaerobic basin inhibits overall phosphorus removal by the plant.

For Photomicrographs see Appendix 5, page A5.1.

CHAPTER TEN

PRACTICAL DESIGN CONSIDERATIONS

by

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1. INTRODUCTION

This section is not intended as a comprehensive and detailed guide to the design of biological treatment plants but as a guide to emphasize those aspects of the design of biological nutrient removal plants which require particular attention. As such only the broad design principles which should be applied in the design of these plants are outlined. As each particular plant is in many respects unique, the detailed implementation of the principles is left to the Design Engineer.

2. PLANT CONTROL

There are a number of design parameters which can be determined and fixed at the design stage and others which can be selected subsequently during plant operation. In the case of some parameters, selection is completely controlled by the designer whereas in the case of other parameters, the designer exercises limited control only. The parameters chosen relate to process input, process output and factors internal to process control.

2.1 Process input parameters

Inputs to the process over which the designer may exercise full control are:

(a) Influent flows including peak flows and recycle flows.

The extent to which the process must be designed to cater for peak flows and diurnal flow variations can be modified by the use of flow balancing or equalization. Where excess flows are likely to have a detrimental effect on the process, these flows may be diverted around sensitive stages in the process scheme. Thus provision could be made in the design for peak storm flows and certain recycle flows to be diverted to the aerobic zone, bypassing the anaerobic and anoxic zones and avoiding the detrimental effect resulting from dilution and reduced retention times in these zones. Similarly, flows resulting from severe storm peaks could be bypassed to the end of the aeration zone. The need for bypass facilities depends on the magnitude of the peak flows in the system and the assessed sensitivity of the system to these flows.

(b) Influent load.

Variations in influent load (as distinct from influent flow) can be reduced by the introduction of a load balancing step to the process scheme. Methods by which this can be achieved will be considered at a later stage.

(c) Dissolved oxygen levels.

The desired residual dissolved oxygen level is selected at the design stage and the design of the aeration system oxygenation capacity is based on this level. In selecting the residual dissolved oxygen level, the minimum desirable level necessary for reliable nitrification and the production of a sludge of good settleability should be provided. This is generally accepted as a dissolved oxygen concentration of not less than 2 mg/*l*.

There are some parameters over which the designer can exercise only limited control. These include:

(a) Influent sewage characteristics which can to some extent be predetermined by an effective industrial effluent control policy, and

(b) by the incorporation of selected additional stages in the process scheme such as:

- (i) Primary sedimentation to reduce the organic load on the biological treatment stage in the case of high strength wastewater with a favourable TKN:COD ratio; or
- (ii) high rate primary sedimentation to partially reduce the organic load on the biological treatment stage while maintaining a favourable TKN:COD ratio; or
- (iii) load balancing achieved by subjecting a proportion of the influent to primary sedimentation and using the primary sludge so derived to increase the organic load of the influent during periods when lower than average loads are experienced.

The parameters that are dependent on the process are sludge age and effluent quality.

2.2 Process output parameters

The designer can select the design sludge age or a design range for the sludge age within normally accepted limits.

The standard of effluent to be produced by the treatment process will, in most cases, be prescribed by the regulatory authority.

2.3 Other process considerations

Based on the influent sewage characteristics and the required effluent quality, sludge and mixed liquor recycle rates can be selected, and a suitable biological reactor configuration chosen. In selecting the biological reactor configuration the possibility of increasing or decreasing zone detention times to cater for the initial flow conditions and possible future changes in the character of the sewage should be considered and provided for if necessary. This can be achieved by compartmentalization of the anaerobic and anoxic zones, providing for step feed of sewage to the various zones of the reactor and the installation of appropriate mechanical equipment, such as a combination of mechanical mixers and associated air spargers in all zones.

3. INSTRUMENTATION

The successful monitoring and control of the biological nutrient removal process depends on robust, reliable and easily maintained instrumentation. In the selection of instrumentation, the initial cost, the availability of local agents and their capability to service and repair the equipment must be taken into account.

In areas with a high incidence of lightning, the possibility of lightning damage to electronic instruments must be considered. Lightning protection devices should be incorporated and special consideration must be given to the siting of instruments and the location of signal cables to minimize the likelihood of lightning damage.

3.1 Flow metering

To facilitate process control, the following process streams must be metered. The flow meters should in all cases incorporate instantaneous rate of flow indication and flow totalization.

(1) Influent to the plant.
(2) The flow to each treatment module.
(3) The flow of waste sludge.
(4) The sludge recycle flows.
(5) The mixed liquor recycle flows.

In small plants where the installation of flow meters on the waste sludge, sludge recycle and mixed liquor recycle flows cannot be economically justified, Venturi flumes should be constructed for these flows and the flow rates checked at regular intervals using a dip stick and flume calibration chart.

3.2 Oxygenation control

Oxygenation control is important from two aspects:

(1) it enables energy usage to be optimized, and
(2) it enables the dissolved oxygen concentration to be maintained at levels necessary for nitrification and good sludge settleability.

Where the character of the sewage is such that biological nutrient removal becomes marginal, the ability to closely control the oxygen levels in the biological reactor becomes extremely important. Dissolved oxygen levels should be controlled so that the levels are high enough to ensure good nitrification and sludge settleability but not so high that denitrification or phosphate removal are adversely affected.

Oxygen control methods include:

(1) Dissolved oxygen monitoring equipment:

The most common system in use today operates electronically by means of a dissolved oxygen sensor which is submerged in the mixed liquor of the biological reactor and which provides a direct reading of dissolved oxygen level at the point where it is situated. The output signal can be used to automatically control the oxygen input to the process. The major advantage of the system is that the equipment is readily available from a variety of manufacturers. The major disadvantage is that the sensors are sensitive to membrane fouling and regular cleaning of the membrane (at least once weekly) is required to ensure satisfactory operation.

Portable dissolved oxygen meters with recorders can be used to determine the temporal variation of oxygen demand. This demand can then be matched by setting up time controls to switch the aerators on and off according to the predetermined demand pattern.

(2) Methods other than the direct monitoring of dissolved oxygen concentration which are presently in the developmental stage are:

(a) On-line monitoring of the ammonia concentration of the aerobic zone. A low ammonia concentration indicates that sufficient oxygen is being provided to ensure full nitrification. This method depends on the presence of a small ammonia residual in the mixed liquor since if zero ammonia is detected, the possibility exists that excess oxygenation is occurring.

(b) By measuring respiration rate of the activated sludge, the oxygen demand of the microorganisms can be determined and the oxygen input and oxygen demand matched.

(c) Total oxygen demand measurements. By measuring the total oxygen demand of the influent, sufficient oxygen can be provided to meet the measured demand.

3.3 Temperatures

Thermometers must be provided to measure both ambient and liquor temperatures. As biological reaction rates are temperature dependent, knowledge of the liquor temperature is important for setting process control parameters such as sludge age and recycle rates.

3.4 Energy consumption

To check the continuing efficiency of the process and equipment, the total energy consumed in the process and in particular the energy used by the aeration system should be recorded on kilowatt-hour meters.

3.5 Sludge blanket sensors

Process operation can be further optimized by monitoring the level of sludge in the clarifiers. In the case of primary sedimentation tanks, monitoring the sludge level in the tanks will provide data on the inventory of sludge available for addition to the process or which must be treated by anaerobic digestion.

Where poorly settling sludges could occur the monitoring of sludge levels in the final clarifiers will provide an early warning of imminent failure of the clarifiers. Appropriate counter-measures such as reducing the influent to the plant, increasing the sludge wasting rate or reducing the sludge recycle rate can be taken timeously to prevent carry-over of sludge into the final effluent.

A system in which the sludge recycle rate is varied in direct proportion to the influent flow rate can also assist in greater process stability.

3.6 Sampling

Regular sampling of the process streams is essential for monitoring and control of the process. Sampling points should be identified at the design stage and necessary facilities provided. Sampling points must be located where conditions are such that a representative sample is obtained. The sampling point must also be readily accessible.

When the sampling frequency at each point has been determined, the method of sampling must be decided on. Where sampling at frequent intervals is required, on-line monitoring should be considered. Such intensive sampling should only be carried out where it is essential for hour to hour process control. Sampling at less regular intervals can be carried out using automatic sampling machines.

In both cases a source of electric power should be provided at the sampling point to power the sampling equipment.

3.7 Running hour meters

The implementation of a routine and preventative maintenance programme for mechanical plant and machinery on a works is facilitated if running hour meters are provided on all plant and machinery.

4. DISSOLVED OXYGEN CONTROL

One key aspect in the design of biological nutrient removal plants is the control of dissolved oxygen levels in order to ensure true anaerobic and anoxic states in the reactors. These states are essential for conditioning of the sludge for phosphate removal and for denitrification respectively.

Direct or indirect transfer of oxygen to the liquid can occur through the aeration system or as a result of turbulent hydraulic conditions.

4.1 Indirect oxygen transfer

Indirect transfer of oxygen to the liquid can occur as a result of:

- Steep site topography which gives rise to hydraulic drops in the distribution channels, pipework and into the biological reactor. (The same aeration effect can also occur in the sewer reticulation system).
- Hydraulic turbulence at venturi flumes and sharp bends in channels and pipework
- The lack of adequate internal baffling in the reactor which allows backmixing of aerated liquor into the anoxic and anaerobic zones.
- The use of weirs for hydraulic control; especially on primary sedimentation tanks, dividing tanks and adjustable weirs for controlling the level of dissolved oxygen in the reactor.
- Hydraulic turbulence caused by vortexing, induced by mechanical mixers in the anaerobic and anoxic zone.
- Hydraulic turbulence resulting from the poor selection of recycle pumps.

It has been shown that the dissolved oxygen concentration in mixed liquor can be substantially increased by its passage through Archimedian screw pumps. If these pumps are used they should be located remote from the anaerobic zone in order to allow bacterial metabolism to reduce the DO before the mixed liquor enters this zone. Air lift type pumps should not be used for obvious reasons. Mixed liquor recycle pumps should be selected whose operation will result in minimum turbulence to the mixed liquor so that excess entrainment of air is avoided.

Where the readily biodegradable fraction of the sewage is low, the use of an aerated grit chamber for the removal of grit from the sewage should be avoided since the aeration may further deplete the readily biodegrable fraction.

The following precautions can assist in eliminating the indirect entrainment of oxygen:

- Pipes should be used in preference to channels and where channels are used, the chosen channel cross-section should minimize the liquid surface area.
- By suitable positioning of baffle walls and direction of rotation of surface aerators, backmixing in the biological reactor between the aerobic and anoxic zones can be utilized to eliminate the need for mixed liquor recycle pumps.
- Where possible, weirs should be designed as submerged weirs.

- The plant site should be selected and the plant layout arranged so that the drop in elevation (the slope of the hydraulic grade line) through the plant is minimized.

4.2 Aeration systems

It is important that adequate aeration capacity is provided in the aeration zone where phosphorus uptake occurs. In selecting an aeration system, initial capital cost, oxygen transfer efficiency, operational flexibility, system reliability, maintenance costs, operating costs and ease of maintenance must be considered. Systems in use include mechanical surface aeration, diffused aeration, turbine aeration and pure oxygen systems. When determining the peak aeration demand, the possible load balancing effect within the biological reactor and reduced peak oxygen demand should be taken into account. The load balancing effect within the reactor will depend on the type of aeration system, the configuration of the aerobic zone, denitrification effects and the magnitude of the internal recycle flows.

4.2.1 Mechanical surface aerators

A design incorporating numerous small aerators favours operational flexibility, system reliability, efficiency and uniformity of dissolved oxygen concentration in the aeration zone. Against this must be weighed the increased capital and maintenance costs of such a design. As surface aerators operate most efficiently at maximum immersion and electric motors operate most efficiently at maximum load, optimum efficiency is obtained with a dissolved oxygen control strategy that matches oxygen input and oxygen demand by stopping and starting aerators or by using aerators with two-speed motors rather than a system in which the depth of immersion of the aerator is varied.

The ability to effectively control dissolved oxygen concentrations in the biological reactor depends on the uniformity of the dissolved oxygen concentration levels throughout the aeration zone. Surface aerators with good mixing characteristics should promote uniform dissolved oxygen concentration levels. As a rule, dissolved oxygen levels in mechanically aerated reactors are not uniform. The measured dissolved oxygen level is affected by the proximity of the sensor to the aerator, whether nearby aerators are operating or not, and the proximity to the basin inlet. This non-uniformity of dissolved oxygen levels must be considered when deciding on the number and position of dissolved oxygen sensors to be provided in each aeration zone. A dissolved oxygen sensor located where it is likely to be affected by any of the abovementioned phenomena is unsuitable for control purposes. In general a number of dissolved oxygen sensors or sockets for installing dissolved oxygen sensors should be provided in the aeration zone. A suitable control sensor can be selected from these depending on the operating conditions and experience with the system. Because of the non-uniform dissolved oxygen levels in the reactor only actual experience will enable the selection of a suitable control sensor and suitable control limits for that sensor.

4.2.2 Fine bubble diffused aeration

A number of different systems utilizing this principle are commercially available and each should be individually evaluated to assess its suitability for a particular application. Important in a fine bubble diffusion system is its "turn down" capability or the range of air flows over which the system is capable of operating. This capability may be of importance for example if the plant is to be very underloaded at start up or if it is to be subjected to large variations in load. Systems using ceramic domes to produce fine air bubbles require a minimum air flow through the domes to prevent both fouling of the domes and the ingress of water into the air pipes. This limits their "turn down" capability and thus their operational flexibility. Systems in which the air flow can be stopped completely provide greater flexibility. With such systems air flow to sections of the aerobic zone can be stopped, if necessary, to provide a larger anoxic volume in the reactor. Dissolved oxygen control is normally accomplished by controlling the volume of air delivered to the reactor by means of inlet vane control on centrifugal type blowers. On smaller plants positive displacements blowers with variable speed drives are common. Where there is more than one reactor, control of dissolved oxygen levels in each reactor should be independent of the other reactors.

4.2.3 Turbine aeration system

In this system mechanical mixers are combined with air spargers to produce an aeration system which has an efficiency somewhere between that of a fine bubble and a coarse bubble aeration system. Although less efficient in energy usage than a fine bubble system or mechanical aeration system, it is attractive because of its operational flexibility. By stopping the air flow to the sparger the aerator can be converted to a mixer. Depending on the reactor design this system would allow virtually unlimited operational flexibility in determining the position and volume fractions of the anaerobic, anoxic and aerobic zones.

4.2.4 Pure oxygen systems

There is little experience with the use of pure oxygen in biological nutrient removal processes although one such system has been patented. The use of oxygen for aeration is most likely to find an application in hybrid systems designed either to correct existing process deficiencies or to upgrade existing works. The system also has applications where strong wastes are encountered which cannot be handled with a conventional aeration system.

5. HYDRAULICS

In the case of the biological nutrient removal processes, particular attention should be given to the following aspects of hydraulic design:

(1) To facilitate process control the inlet must be designed to ensure equal distribution of flow to the various treatment modules.

(2) Where flow balancing is included in the process scheme, the design of the balancing tank should attempt to eliminate or at least minimize the deposition of solids in the tank. A means of removing the deposited solids which does not cause high strength sludge loads to be scoured from the equilization tank should be provided.

(3) In general the hydraulic design should attempt to eliminate the unintentional entrainment of oxygen. This can be done by eliminating cascades and other unnecessary hydraulic turbulence. See section 4.1.

6. "BACK UP" FACILITIES

In situations where conditions for reliable biological nutrient removal are marginal or where, from time to time, the process is not able to produce an effluent complying with the prescribed standard, "back-up" facilities to bring the effluent into compliance with the standard should be considered.

Possible methods of accomplishing this include:

(1) Chemical addition to the main flow stream or to a sidestream of the process to chemically precipitate residual phosphorus not removed by the biological nutrient removal process. The object is to utilize the more expensive chemical precipitation process only when conditions are unfavourable for biological phosphorus removal. The use of chemicals should be avoided unless absolutely necessary to achieve the prescribed standard.

(2) When sub-standard effluents arise, arrangements other than discharge to streams can be made to deal with them. Examples of strategies to prevent the sub-standard effluent from entering the receiving water are:

  (a) Storage of the sub-standard effluents which can then be returned to the process for retreatment when conditions are more favourable.

  (b) Using sub-standard effluents as low grade industrial water or, where sufficient land is available, for irrigation.

One of the main reasons why the biological nutrient removal process fails to produce effluents complying with the phosphorus standard is the occurrence from time to time, of an unacceptably high TKN:COD ratio in the influent sewage. This problem can be overcome by supplementing the influent COD during periods of COD deficiency with COD removed from the influent sewage during periods of excess organic load and stored for use in such a contingency.

7. SCUM HANDLING AND CONTROL

Under certain conditions the biological nutrient removal process is characterised by the growth of filamentous organisms which results in the formation of filamentous scums on both the biological reactor and the final clarifiers. This can give rise to bad odours and high effluent suspended solids concentrations. These scums are known to occur in plants both with and without primary sedimentation. The cause of these filamentous scums is not known but they are believed to result from the occurrence of one or more of the following:

(a) suppressed dissolved oxygen levels and excessively long anoxic retention times;

(b) the occurrence of temperatures and/or substrates which favour the growth of these organisms in preference to other micro-organisms.

As the cause of these scums is unknown and there is no known method of preventing or eliminating the scums, both the biological reactor and the final clarifiers must be provided with adequate descumming facilities. Descumming troughs on the final clarifiers should be positioned taking into account the prevailing wind direction.

Methods which have proved successful in ameliorating the problem are:

(1) spraying the scum layer with water to break it up and entrain the scum in the liquid;

(2) spraying the sum with chlorine solutions or the periodic dosing of chlorine to the aeration basin.

Both methods appear to control the scum by suppressing the growth of the filamentous organisms.

8. TREATMENT OF WASTE ACTIVATED SLUDGE

The principles applied in the treatment of waste activated sludge from conventional biological processes can be applied in the case of sludge from biological nutrient removal processes but, in the case of biological nutrient removal sludges, special consideration must be given to the excess phosphorus associated with the sludge which will be released into solution under anaerobic conditions. The sludge must be either maintained in an aerobic condition or, if the sludge is allowed to go anaerobic, provision must be made for the treatment and disposal of phosphorus enriched liquors arising in the treatment process.

The following principles should be applied:

(1) Mixed liquor should be wasted from the aerobic zone of the biological reactor where the dissolved oxygen concentration is high in preference to wasting from the clarifier underflow where dissolved oxygen concentrations are lower and release of phosphorus from the sludge could already be occurring.

(2) The sludge treatment facilities should be located in close proximity to the biological reactor so that long conveyance of the sludge is avoided. When pumping the waste sludge is unavoidable, the retention periods in the pump station sump and in the rising main should be kept as short as possible.

(3) The retention time of the sludge in the thickening and dewatering processes should be as short as possible.

(4) Preference should be given to thickening and dewatering processes which will maintain the sludge in an aerobic condition.

In the case of sludge thickening, dissolved air flotation (DAF) and centrifugation are the most suitable processes currently available for rapid dewatering. Thickening of waste activated sludge can be carried out by centrifugation or by means of filter belt presses. In the case of both gravity thickeners and filter presses, the sludge is retained in the process for a considerable length of time during which it goes anaerobic and the phosphorus is released into solution.

Wherever possible the underflows, centrates, filtrates and wash waters from the thickening and dewatering process should be segregated so that any sidestream containing a high phosphorus concentration as a result of phosphorus release from the sludge can receive further treatment. Such treatment can take the form of chemical addition or land irrigation to immobilize the phosphate ion. Where the phosphorus-rich waste activated sludge is stabilized by anaerobic or aerobic digestion, the excess phosphorus in the sludge will be released into solution and the supernatant liquors from these processes must receive suitable treatment.

Dissolved air flotation thickening of waste activated sludge produces a sludge which is saturated with air. To avoid the possibility of air binding in the pumps, positive displacement pumps should be used with this sludge in preference to centrifugal pumps.

The sludge also has the rheological properties of a non-Newtonian fluid. The pumping system and pipelines must be designed accordingly.

9. SECONDARY CLARIFIERS

The secondary clarifiers are expected to produce a final effluent containing a minimum of suspended solids as well as a highly concentrated underflow for recycling to the biological reactor. In the light of the known poor settling properties of these sludges, clarifier design should be based on the "solids flux theory" approach. The total clarifier area must be sufficient to pass the design sludge mass at the predetermined flux and the clarifier must be deep enough to permit the required thickening of the underflow. Laboratory studies should be carried out on a representative sludge to determine these parameters.

To avoid anaerobic conditions developing in the clarifier with the subsequent release of phosphorus into solution, the clarifier should be designed so as to minimize the length of time the recycled sludge is retained in the clarifier. In this regard the activity of the sludge as measured by the respiration rate is important. Very active sludges will quickly utilize the available dissolved oxygen and nitrates in the sludge liquid and anaerobic conditions will develop rapidly. Provided the released phosphorus is retained within the underflow liquid, the phosphorus will not be released into the effluent. The likelihood that the phosphorus will escape into the effluent will depend on the depth of the clarifier, the height of the sludge blanket and the method by which sludge is withdrawn from the clarifier.

Circular clarifiers with conical bottom and centre draw off of sludge under hydrostatic head, provided with mechanical scraper mechanism and a depth not less than three metres but preferably four metres are considered to be most suitable. Tanks with these characteristics are most likely to have hydraulic regimes approaching the ideal "plug flow" condition.

Circular clarifiers with suction-type sludge removal mechanisms enable sludge to be removed rapidly from the clarifier but this type of clarifier has a number of disadvantages:

(1) They require much attention from the operator to ensure that the syphoning action is maintained.

(2) The design should enable each suction pipe to be monitored visually and controlled independently to ensure that each is removing sludge with high solids concentration.

(3) In essence the syphon method is not a "fail safe" design and if failure of the syphoning action is not detected, the tank can fill with sludge and sludge overflows into the effluent can result.

The suction type mechanism produces an inherently unstable hydraulic regime in the clarifier which is likely to cause mixing and greater deviation from ideal "plug flow" conditions. It has also been observed that excessive rotational speed of the scraper mechanism can cause turbulence in the sludge blanket. If it is proposed to use this type of clarifier, the design parameters should be checked against known successful installations.

The number and size of clarifiers to be provided will depend on considerations of overall cost, operational factors and the possibility that under initial underloaded conditions there will be increased retention time in the tanks. When the clarifiers are sized allowance must be made for any flow variations which may occur including possible surges in flow resulting from rapid lowering of the adjustable weirs used to automatically control dissolved oxygen input to the reactor. Because of the prevalence of filamentous scums it is essential that adequate means of descumming is provided.

10. BIOLOGICAL REACTOR

The biological reactor consists of a number of different zones. Each has a specific function to perform and the design of each zone must provide the optimum conditions for the performance of these functions.

10.1 Anaerobic zone

The retention time of the influent sewage in this zone plus the development of authentic anaerobic conditions are of extreme importance. Conditions which are likely to affect either, such as return flows with high dissolved oxygen or nitrate concentrations, must be avoided. To ensure an adequate retention time and no dilution of the readily biodegrable COD concentration in this zone, provision should be made to divert peak storm flows past this zone directly into the aeration zone.

Enclosure of this zone to exclude atmospheric oxygen is not necessary provided excessive surface turbulence as a result of mixing does not occur. To limit turbulence, the energy imparted to the liquid by the mixers should be kept to a minimum. It has been found that an energy input of about 4 W/m^3 provides adequate mixing without causing excessive turbulence. Actual power requirements for mixing will depend on the shape and size of the reactor and the type of mixer installed. Mixer manufacturers should be consulted regarding the most suitable configuration. Where feasible, the use of the kinetic energy in the influent flow or recycle streams to achieve hydraulic mixing in both the anoxic and/or anaerobic zone should be considered.

The concentration of readily biodegradable COD in the anaerobic zone should be as high as possible to promote biological phosphorus removal. This can best be achieved in a plug flow reactor or in a series of small completely mixed reactors rather than in a single large completely mixed reactor. Baffles between the zones should be designed to prevent backmixing but need not be completely watertight.

10.2 Anoxic zone

The introduction of extraneous dissolved oxygen into these zones must be prevented. The size of the primary anoxic zone must be based on denitrification rates determined by experiment using the sewage to be treated. The second anoxic zone is not always needed. It is most effective when treating high strength sewage particularly when the sewage has a high biodegradable solids content. Under these conditions particulate organic matter is absorbed onto the sludge and is carried through to the second anoxic zone on the sludge. Metabolism of the absorbed particulate COD in the second anoxic zone enables higher denitrification rates to be achieved.

With low strength sewages, for example settled sewages, and with sewages containing a large proportion of soluble and colloidal biodegradable matter which are readily removed in the primary anoxic zone, there is likely to be a deficiency of organic matter in the second anoxic zone and only minimal quantities of nitrate may be removed.

10.3 Aerobic zone

The main aerobic zone must be sized for complete nitrification at the minimum sewage temperature, at the selected residual dissolved oxygen concentration and sludge age. The choice of aeration system and the measurement and control of dissolved oxygen levels has been discussed in Section 4.2.

The final aerobic zone is required:

(1) to boost the dissolved oxygen concentration in the mixed liquor before it enters the clarifiers; and
(2) to polish the final effluent by removal of additional phosphorus and the oxidization of residual ammonia.

Excess aeration in this zone must be avoided as it will result in the conversion of organically bound nitrogen to nitrate and the slow aerobic release of phosphorus.

11. ECONOMICS

The main objective of wastewater treatment is to produce an effluent meeting the required standard at the lowest possible cost. To accomplish this the trade-offs between the cost components of the various wastewater treatment processes must be determined in order to establish the optimum scheme. For example, when comparing the cost of biological filtration with chemical phosphorus removal and the cost of biological phosphorus removal in the activated sludge process, the major trade-off is cost of chemicals as opposed to the cost of electricity. Thus for an existing biological filter plant where little additional capital expenditure need be incurred, the most cost effective solution is probably chemical precipitation of phosphorus. The alternative is some modification of the activated sludge process which will incur additional capital expenditure and will involve the use of electricity.

11.1 Typical treatment costs for a biological nutrient removal plant

The treatment costs of a first generation (Phoredox) 150 Mℓ/d biological nutrient removal plant are given below. The major cost components are:

(i) Interest on and redemption of capital.
(ii) The cost of electrical energy.

Treatment Costs – 150 Mℓ/d 'Phoredox' Plant:
Year Ending June 1981

	Actual Flow	Design Flow
Flow (m^3 per annum)	31 587 x 10^3	54 750 x 10^3
Costs (cents/m^3)		
Operating	0,43	0,25
Maintenance	0,38	0,22
General overheads	0,77	0,44
Electricity	0,77	1,71
Laboratory Services	0,58	0,34
Interest and Redemption	4,67	2,69
Total Costs	7,60	5,65

NOTES:

1. Cost of electricity taken at 2,03 cents per kWh, inclusive of service charge, demand charge and energy charge in accordance with Johannesburg Electricity Department tariffs.
2. In addition to the abovementioned costs, the cost of treatment and disposal of waste activated and primary sludge must be added.
3. Based on the latest models of the biological nutrient removal process, for process configuration which optimizes the operating costs by minimizing electricity consumption, the electricity cost will be of the order of 0,5 cents/m^3.

11.2 Selection of process configuration

If energy usage is to be minimized, a process configuration should be selected which can operate at a high TKN:COD ratio and will minimize load to be oxidized. In addition, a smaller biological reactor, smaller final clarifiers, reduced oxygenation capacity with resulting energy saving will be required. These savings will be offset to some extent by the additional cost of primary sedimentation and the cost of the treatment and disposal of raw sludge.

Provided that the TKN:COD ratio after primary sedimentation is such that the process remains viable, the benefits of incorporating primary sedimentation into the process scheme are:

(i) lower loadings on the biological reactor;
(ii) reduced energy consumption;
(iii) smaller civil and correspondingly reduced land requirements;
(iv) smaller mass of waste sludge to be treated.

The disadvantages are:

(i) the need for anaerobic digestion to treat primary sludge;
(ii) the additional costs associated with the primary sedimentation stage in the process scheme, including the cost of anaerobic digestion.

11.3 Effect of equalization (flow balancing) on the sizing of the units and associated cost implications.

The most notable benefits of flow equalization are:

(i) Reduced clarifier size as a result of reduced peak flow rates downstream of the equalization tank.
(ii) A levelling of aeration power demand resulting in reduced installed aeration capacity and lower expenditure on energy.
(iii) The biological treatment process benefits from reduced fluctuation in loading as well as more stable retention periods and biomass concentrations in the aeration basin.
(iv) Reduced reactor volume.
(v) Simplified manual and automated control of flow dependent operations such as chlorination and filtration.
(vi) Savings realised on both capital and operating cost resulting from smaller clarifiers and aeration basin, fewer or smaller aerators and smaller pipes and channels.

In the following analysis an attempt is made to evaluate the cost savings resulting from smaller clarifiers, smaller aeration basins and fewer or smaller aerators for an existing works with a nominal treatment capacity of 150 Mℓ/d.

The additional cost of the balancing tank must be offset against these savings. In the analysis it is assumed that the additional ground required for the balancing tank is offset by the additional land which would be required for secondary clarifiers if flow equalization is not practised.

The parameters used in the original design of the works have been used to estimate the requirements without flow equalization.

11.3.1 Aeration

The original design was based on:

Maximum 2 hour load without balancing	4 286 kg BOD
Maximum 2 hour load with balancing	3 000 kg BOD
Minimum 2 hour load without balancing	570 kg BOD
Minimum 2 hour load with balancing	900 kg BOD
Maximum: Minimum 2 hour load ratio	
without balancing	7,5
with balancing	3,3
2 hour peak BOD: Average daily BOD	
without balancing	2,3
with balancing	1,6

The total installed oxygenation capacity (main aerators only) with flow equalization had been calculated to be 3 244 kg O_2 based on an oxygen input per aerator of 90,1 kg/h (see Annexure 1).

The number of additional aerators required without flow equalisation:

$$\frac{4\,296 - 3\,244}{90,1} = 11,58 \quad \text{(say 12 aerators)},$$

or four additional aerators per module, which could be accommodated adequately in the existing aeration volume with little increase in energy density.

The cost of the additional aerators (excluding provision of civil support structures) is estimated at R413 000.

11.3.2 Biological reactor

In a conventional plant the volume of the aeration basin is designed for the average flow/load and a safety factor is then applied to account for variations in flow and load when there is no flow equalisation.

Factors ranging from 1,3 – 2,5 have been proposed to account for influent variability where treatment capacity is estimated using a steady state model. Furthermore, reactions in the anaerobic and anoxic zones are rate-dependent and a minimum detention time must be provided. Where no flow equalization is provided these zones should be designed to provide the minimum detention time at peak flow.

Under conditions of no equalisation the volume of these zones would have to be increased to maintain the existing detention times.

ZONE	INCREASED VOLUME (per module) m^3	ADDITIONAL COST Rand
Anaerobic Zone	915	41 000
Primary Anoxic Zone	2 110	93 000
Main Aeration Zone	4 410	195 000
		329 000

In this analysis the secondary anoxic zone has been omitted as experience has shown that it is ineffective and should possible be excluded in future designs.

11.3.3 Secondary sedimentation

Based on the original design parameters six clarifiers would be required for each module for non-equalized flow instead of the four clarifiers provided with equalisation (see Annexure 3).

The cost of these additional clarifiers would amount to R299 000 for mechanical equipment and R263 000 for civil structures, a total of R492 000. The cost could be reduced to some extent by using a smaller number of larger clarifiers but it is felt that the size of the clarifier in use is approaching the limit for the design.

11.3.4 Pipework, channels, etc.

Flow conveying elements of the treatment works such as pipes, channels etc. must be designed to convey the balanced flow. Sizing of these elements is very site specific; greater costs are likely to be incurred on flat sites than on sites with adequate falls. No attempt therefore has been made to quantify the expected cost differences.

11.3.5 Downstream processes

Other downstream processes also benefit from flow equalisation since these processes can be sized to treat the average flow rather than the expected peak flow. Chlorination equipment, chlorine control tanks and filtration units will be smaller in size. Because of the constant flow, these processes will also be easier to operate.

11.3.6 Cost implications

Details of the estimated increase in capital and electricity cost where no equalisation is provided are given in Annexures 2 and 4.

The estimated increase in capital cost for a plant with a nominal treatment capacity of 150 Mℓ/d if no provisions is made for flow balancing is approximately R1,0 million.

The estimated annual savings resulting from the incorporation of flow balancing, based on savings on interest and redemption of capital and electricity costs, is R180 000 per annum or a saving of 0,33 c/m^3 of sewage treated.

Both primary sedimentation and flow balancing can adversely affect the TKN:COD ratio of sewage. If certain minimum TKN:COD ratios are necessary for the operation of a particular biological nutrient removal process, the use of primary sedimentation and possible flow balancing might be counter-productive and lead to unstable performance.

The implications of the TKN:COD ratio in process design and selection of a reactor configuration are dealt with fully in Chapter 7.

11.4 Effect of load balancing

A similar exercise to the above can be carried out when it is intended to incorporate load instead of flow balancing. Assuming that the total influent load eventually all passes to the biological reactor, the advantages of load balancing are:

(i) A more stable process in that when periods of adverse TKN:COD ratios occur in the sewage these can be balanced out.

(ii) A reduction in maximum/minimum power demand commensurate with the degree of load balancing applied to the reactor. The installed aeration capacity can be reduced because of the reduction in peak loads. A corresponding reduction in peak demand electricity charges also results. As it is assumed that the total load is not reduced, the total aeration demand and hence energy consumption will remain unchanged.

(iii) If all the primary sludge produced in the load balancing stage is returned to the process stream, no additional costs will be incurred in the treatment and disposal of primary sludge, but an increased mass of waste activated sludge will have to be treated and disposed of.

The disadvantages are:

(i) As peak flows will not be attenuated by the load balancing operation, the final clarifiers, biological reactor, pipework and channels must be designed to cater for these peak flows.

(ii) Additional capital expenditure will be incurred for the provision of a sludge storage tank and primary sedimentation tanks; although the latter could be smaller and less efficient than the normal design.

11.5 Full and partial nitrification and denitrification

Depending on the effluent standards with respect to ammonia and nitrates, the plant can be designed and/or operated to achieve full and/or partial nitrification and denitrification.

Full nitrification without full denitrification is less efficient in respect of energy usage as the energy stored as bound oxygen in the nitrates is only partially recovered. In a process configuration suitable for sewage with a high TKN:COD ratio, less energy is consumed by the oxidation of the carbonaceous substrate but there would be a lower energy recovery as a result of only partial denitrification and higher nitrate concentrations would be discharged in the effluent.

Low dissolved oxygen concentrations in the biological reactor can achieve reduced energy usage by suppressing nitrification. This procedure can however result in operational problems as a poorly settling filamentous sludge may be produced.

12. STAFFING

There is no doubt that the operation of a biological nutrient removal plant requires greater understanding of the theory behind the process than is normally required from sewage treatment plant operators in South Africa. This is particularly true when the conditions for biological phosphorus removal are marginal. The effect on the process of variability in the influent may necessitate regular process adjustments and may require the full-time attendance of a competent certificated operator who is fully conversant with the biological nutrient removal activated sludge process.

This has the following implications:

(i) Higher salaries will be necessary to attract a better quality, properly qualified operator who must be capable of understanding the operational requirements and making the correct process adjustments when necessary.

(ii) Provision must be made for adequate staff training to ensure that the operator is competent and capable of operating the plant.

(iii) Full-time attendance of a trained operator is required. It is most unlikely that such a plant will produce results consistently within standard if it is provided with only limited operator attendance.

The biological nutrient removal process places greater reliance on instrumentation for success than other biological processes. The operator should be capable of minor routine maintenance on the instruments such as checking zero adjustment and renewing DO probe membranes. There is an extreme shortage of qualified instrument technicians in South Africa, and provision should be made for maintenance contracts for instrumentation.

ANNEXURE 1

CALCULATED AERATOR PERFORMANCE

Size of electric motors = 110 kW
Number of electric motors = 12 per module
= 36 total
Full load efficiency of motors = 94%
Efficiency of gearboxes = 93%
Guaranteed oxygen transfer efficiency = 1,81 kgO_2/kWh

Oxygenation capacity of the aerators must be adjusted to allow for:

(a) Altitude (1 500 m above sea level).
(b) Sewage as opposed to clean water ($\alpha = 0{,}85$).
(c) Temperature (13°C).

$$\text{Oxygenation capacity} = 1{,}81 . \frac{(7.6-1.0)}{9.02} . 0{,}85 . 0{,}94 . 0{,}93 = 0{,}98 \text{ kg/kWh}$$

Power at aerator drive output shaft at maximum immersion = 92 kW

Oxygen input per aerator = 92.0,98 = 90,10 kg/h

TOTAL OXYGEN INPUT (3 MODULES = 36 AERATORS) = 3 244 kgO_2/h

ANNEXURE 2

CALCULATION OF ELECTRICITY COSTS

Aeration is the major consumer of energy and thus has a predominating effect on the costs of power. Mixing, pumping and secondary clarifiers constitute only a small proportion of the total power demand and can be neglected in a comparison of power costs.

The analysis is based on the following electricity tariff:

Service charge	22c per day
Demand charge	11,66c per kVA per day
Consumption charge	1,39c per kWh

The power required depends on the oxygen demand of the incoming load (COD and ammonia). The average daily load to be treated is the same whether flow equalization is provided or not. Only the peak load will vary. This must be catered for by the provision of additional aerators where no flow equalization is provided. This will result in higher peak electricity demands and a higher demand charge only.

With flow equalization		Without flow equalization	
Number of aerators per module	12	Number of aerators per module	16
Size of aerator motors (kW)	110	Size of aerator motors (kW)	110
Power factor (with PF correction)	0,96	Power factor (with PF correction)	0,96
kVA per aerator	114,6	kVA per aerator	114,6
Maximum kVA per module	1 375	Maximum kVA per module	1 834
Demand charge per module per day	R159,52	Demand charge per module per day	R212,70
Demand charges per annum	R174 678,00	Demand charges per annum	R232 904,00

ADDITIONAL COST OF ELECTRICITY WITHOUT FLOW EQUALIZATION: R58 226,00 per annum.

ANNEXURE 3

SECONDARY SEDIMENTATION REQUIREMENTS WITHOUT FLOW EQUALIZATION

Peak (2 hour) dry weather flow without balancing = 8 967 m^3/h
per module = 2 989 m^2/h

Solids loading (flux): In winter = 2,5 kg/m^2h
In summer = 3,5 kg/m^2h

Mixed liquor suspended solids: In winter = 5,1 kg/m^3
In summer = 3,5 kg/m^3

Winter Conditions

Solids load on clarifiers = 2 989 m^3/h × 5,1 kg/m^3 = 15 224 kg/h

$$\text{Area of clarifiers required} = \frac{15\,224}{2{,}5} = 6\,097 \text{ m}^2$$

Summer Conditions

Peak wet weather flow without balancing = 13 078 m^3/h
per module = 4 359 m^2/h

Solids loads on clarifiers = 4 359 × 3,5 = 15 258 kg/h

$$\text{Area of clarifiers required} = \frac{15\,258}{3{,}5} = 4\,359 \text{ m}^2$$

Winter conditions are critical and 6 – 36,2 m diameter clarifiers would be required for each module.

ANNEXURE 4

COST IMPLICATIONS OF FLOW EQUALIZATION

	COST (R)	
	WITH EQUALIZATION*	WITHOUT EQUALIZATION
UNIT		
Primary sedimentation tanks	221 000	221 000**
Balancing tank	236 000	—
Aeration basins	1 720 000	2 049 000
Secondary clarifiers	525 000	788 000
MECHANICALS		
Primary sedimentation tanks	152 000	152 000
Balancing tank	—	—
Aerators only	1 240 000	1 653 000
Secondary clarifiers	458 000	687 000
	R4 552 000	R5 550 000

*Actual final cost(1976) rounded off to nearest R1 000.

**Primary sedimentation retained as exclusion would require larger aeration basin volume and greater aeration capacity.

Based on a redemption period of 15 years at 10% p.a. interest for mechanical equipment and 30 years at 10% p.a. for civil construction work, the annual cost of the alternatives is:

SCHEME		CAPITAL (R)	INTEREST AND REDEMPTION (R p.a.)	ADDITIONAL COST OF ELECTRICITY (R p.a.)
With equalization				
Civils		2 702 000	287 000	–
Mechanicals		1 850 000	243 000	–
	TOTAL	R4 552 000	R530 000	
Without equalization:				
Civils		3 058 000	324 000	–
Mechanicals		2 492 000	328 000	58 000
	TOTAL	R5 550 000	R652 000	R58 000

ADDITIONAL ANNUAL COST OF SCHEME WITHOUT EQUALIZATION AMOUNTS TO R180 000.

CHAPTER ELEVEN

OPERATION OF BIOLOGICAL NUTRIENT REMOVAL PLANTS

by

A.R. Pitman

1. BACKGROUND

This chapter will highlight operating experiences on full-scale biological nutrient removal activated sludge plants. Most of the discussion will reflect information gained on the five-stage Phoredox process layout but should easily be adapted to other processes such as the three-stage Phoredox or UCT systems. The basic principles involved should be the same for all biological nutrient removal plants.

Even when research provides sufficient understanding of phosphorus removal mechanisms to permit designers to provide more optimum plants, the successful operation of any wastewater treatment plant will depend on the quality of its operators. In addition, the extra complexity of nutrient removal processes requires that they be controlled by a better calibre of operator than one who operates a trickling filter plant. In this respect, suitably trained and experienced operators are most important. This is not a peculiarity of biological plants as good quality operation would also be required on chemical nutrient removal plants.

Suitable laboratory facilities manned by properly trained, experienced staff would also be required on nutrient removal plants in order to ensure that regular analyses are carried out. Owner authorities must therefore ensure that adequate staff will be available before embarking on the planning and design of these plants.

2. INFLUENT SEWAGE

2.1 Characteristics

In general, biological nutrient removal is more suited to strong sewages than weak ones. Experience has shown that for successful operation the feed COD should be at least 300 mg/*l* and preferably above 400 mg/*l*. Also, the TKN/COD ratio of the feed is important; this should preferably be below 0,10. The sewage should also have a relatively high readily biodegradable COD content. The above requirements will ensure that denitrification is virtually complete and that phosphate release in the anaerobic zone will readily occur.

Where readily biodegradable COD is available as a waste product from industry, its discharge to the sewer should be encouraged. Primary sedimentation tanks can have a marked detrimental effect on the desired sewage characteristics and thus lead to inferior phosphorus removal. The COD is reduced in primary sedimentation tanks and the TKN/COD ratio can increase adversely as COD is usually removed to a greater extent than TKN. The reduction in COD load and concentration over weekends, holiday periods and after heavy rainfalls, can also be a problem. This can cause phosphorus removal to deteriorate for a period and only recover once the normal load has been restored.

The combination of primary sedimentation and low strength weekend flows could result in the complete loss of phosphorus removal with phosphate peaks occurring in the effluent.

As with many other continuous processes, biological nutrient removal can benefit from a constant quality of feed. In this respect the use of flow balancing tanks can be an advantage. They must however, be controlled to ensure that effluent flowrates are constant and that solids deposition on the balancing tank floors are minimised.

2.2 Storm flows

During periods of rain, sewage flows tend to be of greater magnitude and dilute strength due to the infiltration of rainwater into the sewer system. These factors can produce some unique problems to the biological nutrient removal process.

The weaker sewage could result in inadequate COD being available to:

(i) achieve the usual degree of denitrification; and

(ii) ensure that the anaerobic zone is properly anaerobic. Also, if adequate DO control is not practised, DO levels will rise as the respiration rate drops and excessive nitrification of organic nitrogen can occur.

The higher flow would result in the retention times in the zones being reduced to the extent that the desired action does not take place in them. This is particularly so in the case of the anaerobic zone.

For the latter reason the possibility of bypassing peak storm flows should be considered. These could flow to holding tanks to be pumped back during low flows or go to oxidation ponds for stabilisation. If none of these facilities are available, peak flows could be bypassed to the final clarifiers or preferably to the end of the biological reactors. This will increase the upward velocity in the clarifiers but not increase the solid flux load to the same extent because of the dilution effect on the MLSS. The advantage of flowing through the last zone or the end of the reactor is that some contact with biomass is achieved, which will absorb COD and not result in a severe deterioration in effluent quality.

In the case of weaker sewages the dilution of the strength caused by storm water can be a serious problem leading to the failure of phosphorus removal.

2.3 Improving feed characteristics

If the quality of the sewage fed to the process is inadequate, various options can be tried to improve the situation. Possibilities are:

- Overloading of the primary clarifiers to increase solids carry-over and hence the feed COD.
- Continuously feeding back of primary sludge into feed to biological reactors.
- The addition of volatile fatty acids produced by high rate anaerobic digestion of primary sludge. This would be the ideal substrate for *Acinetobacter* and would also be a readily biodegradable substrate producing rapid denitrification. These acids would also increase the degree of release in the anaerobic zone and produce a much more (biologically) active mixed liquor. Either an acid supernatant liquor can be added or the whole acid sludge containing sludge solids can be used. The latter will however, increase the MLSS and power demands far more than the former. The acid material should be bled back into the sewage feed to the anaerobic zone.

3. BASIC CONCEPTS IN BIOLOGICAL PHOSPHORUS REMOVAL

In the conventional activated sludge process which is not designed or operated to achieve biological excess phosphorus removal, the bacteria only use phosphorus in quantities which satisfy their basic metabolic requirements. Because of the usual nutrient imbalance in sewage only a limited quantity of the feed phosphorus will be removed in such plants. For example, from a feed sewage with 10 mg P/*l* total phosphorus, an effluent containing 7 – 8 mg/*l* total phosphorus will be produced. The waste activated sludge from such a process will contain relatively little phosphorus in a typical case, 2 % on a dry solids mass basis.

Bacteria which accumulate excess quantities of phosphorus will be absent or only present in low numbers in such plants and the conditions which induce excess phosphorus uptake by the biomass will not occur.

In plants designed for biological phosphorus removal however, the environment is created for the proliferation of a biomass which will accumulate phosphorus in excess of normal metabolic requirements. Certain bacteria growing in such plants can accumulate and store within their cell walls large quantities of phosphate as a relatively insoluble storage product. Furthermore, the possibility exists that certain chemical precipitates containing phosphorus can be adsorbed to the surfaces of organic matter in activated sludge. Such plants can therefore produce an effluent containing less than 1 mg/*l* phosphorus from a feed sewage containing 10 mg P/*l*. The waste activated sludge will contain enhanced quantities of phosphorus, typically 4 – 5 % (as P) in the case of most local plants.

The sludge from a plant that is removing phosphorus well, will have the following properties:

3.1 Ability to release phosphate

Under anaerobic conditions, that is, in any environment free of nitrate and dissolved oxygen an activated sludge that has accumulated excess phosphorus will tend to release phosphate from storage areas out into surrounding liquid. The rate of release will depend on the activity of the sludge, but over a few weeks of anaerobiosis a large proportion of the accumulated phosphate will be released. However, in the short retention period in the anaerobic zone of a plant only 10 – 20 % of the total phosphorus in the sludge will be released. When necessary, anaerobic release can be prevented by keeping the sludge under aeration or by maintaining a certain concentration of nitrate in the sludge.

Anaerobic release of phosphate can be stimulated by the presence of biodegradable COD. When a phosphate rich activated sludge is fed with readily biodegradable soluble COD, the rate of release under anaerobic conditions is speeded up. In fact, volatile fatty acids such as acetic acid can produce very rapid phosphate releases.

Anaerobic release is an important prerequisite for biological phosphorus removal. For a plant to remove phosphate well, release must always occur, either in the anaerobic zone or elsewhere in the process. Phosphate release indicates that the sludge has been conditioned to remove further phosphorus. Anaerobic release conditions are usually achieved by passing activated sludge continually through a zone which is free of nitrates and molecular oxygen so that at any one time, a certain fraction of the total activated sludge inventory is being subjected to the anaerobic environment. In certain plants, on/off operation can be used whereby all the sludge in the process is anaerobically conditioned at once, usually by switching off all the aerators for a period of time. This 'off' period is repeated daily or after a number of days and the plant is operated in its normal aerobic state for the rest of the period. If release in the anaerobic zone is lost, for example, due to the presence of nitrates, phosphate removal will fail. In most plants this usually manifests itself as an increase in effluent phosphate concentration a day or two after release was lost, i.e. there is a time delay between the loss of release and the failure of phosphorus removal. Where all the sludge has been anaerobically conditioned on a batch or on/off basis, the conditioned sludge will continue to take up phosphate under aerobic conditions for a number of days or even weeks before the effect of the conditioning is lost. Sufficient aeration capacity should be available to ensure rapid uptake of released phosphate. If this is not on hand, temporary chemical precipitation should be considered. Plants in Johannesburg, Pinetown and Brits have been successfully operated in this mode without addition of chemicals.

A phosphate rich sludge can also release soluble phosphate under aerobic conditions. This is a far slower process than anaerobic release and takes place when the sludge is aerated at relatively high DO levels in the absence of substrate, i.e. endogenous conditions. Examples of where such release can occur are, at the end of a plug flow aeration basin and during aerobic sludge digestion. However, it must be emphasized that this type of release is not the same as the anaerobic release required to induce phosphorus removal.

Phosphate release under aerobic conditions will also occur if the pH of the sludge is decreased (say to 5 – 6) by the additional acids or by excess nitrification in a poorly buffered water. The latter can typically occur during aerobic digestion of the sludge.

Although anaerobic release must occur before biological phosphorus removal is possible, release of phosphate can cause problems in certain circumstances. For example, *excessive* release under anaerobic conditions can result in inadequate phosphate uptake in the subsequent aerobic zone. Also, if a phosphate rich sludge is anaerobically digested, phosphate release will occur and liquors and filtrates from the dewatering of such a sludge will have a high phosphate content. Furthermore, when waste activated sludge is dewatered on drying beds, the bed underflow can have an elevated phosphate content. The recycle of these liquors to the nutrient removal activated sludge reactors can impair phosphorus removal. If this is a problem artificial precipitation of phosphate from these liquors by chemical addition should be considered.

3.2 Phosphorus uptake ability

Under aerobic conditions the sludge will remove phosphate from the liquid phase and accumulate it in quantities which are in excess of normal metabolic requirements. However, the sludge will only do this if it has been previously subjected to anaerobic release conditions, i.e., the sludge will only take up phosphate if it has previously released phosphate. During aeration such a conditioned sludge will accumulate phosphate fairly rapidly, often leaving only trace quantities in solution. Sludges which do not perform this way have either not been anaerobically conditioned properly or the rate of aeration is inadequate. In particular, sludges which have released excessive quantities of phosphate might need a longer aeration time to reduce solution phosphate concentrations to low levels.

4. OPERATING CONDITIONS IN VARIOUS ZONES AND CLARIFIERS

This section will describe typical operating conditions in the zones of a five-zone plant and point out the conditions which have been found to be necessary for efficient nutrient removal.

4.1 Anaerobic zone

This zone is essential for phosphorus removal as the bacteria in the activated sludge passing through it are preconditioned to take in excess phosphate. The release of a certain quantity of phosphate from the biomass into solution indicates that the bacteria have been suitably conditioned. To achieve this the feed sewage and activated sludge are retained in the absence of nitrate and dissolved oxygen, for a period of anything between 30 minutes and 4 hours. The actual retention time is dependent on the volume of this reactor and the flow rates of the feed sewage and recycled activated sludge. The strength of the sewage can also affect the retention time needed for successful preconditioning. For example, a high COD feed, particularly one containing a large readily biodegradable fraction will cause a rapid phosphate release and hence conditioning of the sludge, while a low COD feed will produce a far slower rate of release with a concomitantly lower excess P removal.

Bacteria in the mixed liquor are stressed or conditioned in this zone to the extent that they will take up excess phosphate under aeration in the aerobic zones. The level of phosphate release in the anaerobic zone is a measure of the degree of anaerobic stress, i.e., the greater the release the greater the stress achieved.

Conditions in the anaerobic zone should be such that phosphate is released into solution to levels above the feed sewage phosphate concentration. For example, for good phosphorus removal to occur in a plant treating a sewage containing 8 mg/ℓ total P with a MLSS concentration of 3 000 mg/ℓ, the *orthophosphate concentration in solution in the anaerobic* zone should average about 20 mg P/ℓ and never drop below 10 mg/ℓ. Although release is necessary for phosphorus removal, excessive release should be avoided as it might prove difficult to take up all of the released phosphate in the aeration time available in the aerobic zone. In the example above, a release of 30 mg P/ℓ may be counter-productive as 20 mg P/ℓ might be perfectly adequate.

Nitrate and dissolved oxygen discharged to this zone should be zero or as near as possible to zero at all times. If nitrates or DO are detected anaerobic conditions will not be possible, phosphate release will not occur and biological excess phosphorus removal will not operate, that is, even a nitrate concentration of 0,5 mg N/ℓ will block phosphate release and hence uptake. The situation might even occur where nitrates and DO are not detected in the anaerobic zone and adequate phosphate release will still not occur in this case and nitrate feedback via the return sludge might just be sufficient to prevent anaerobic conditions occurring but the amount returned would be denitrified to the extent that no measurable quantities of nitrate were present. The loss of release in the anaerobic zone indicates that phosphate removal is about to fail. This usually occurs either the following day or the day thereafter, for example, low phosphorus release patterns over weekends can result in effluent phosphate concentrations increasing on Tuesdays.

4.1.1 Nitrate feedback

In plants which have an anaerobic zone at the inlet end, the problem of nitrate feedback in this zone via the return sludge cannot be over-emphasized. When nitrate concentrations in the return sludge are high, an excessive denitrification load can be placed on this zone and the anaerobic conditions required for phosphate removal are lost. As soon as nitrates are detected in the anaerobic zone, even at low concentrations (0,5 – 1,0 mg N/ℓ) impending failure of phosphorus removal is indicated. The ideal condition is that the return sludge should contain no nitrates but this is not always possible in practice. The degree of nitrate feedback that can be tolerated depends on the strength of the sewage feed to the anaerobic zone and in particular, its readily biodegradable COD concentration. For a high strength sewage (COD = 600 mg/ℓ) with a good readily biodegradable content, a return sludge nitrate level of 7 mg N/ℓ might be tolerable. Alternatively, for a low strength feed of COD = 200 mg/ℓ, a return sludge nitrate level of 1,5 mg N/ℓ might be undesirable. However, it should be borne in mind that the greater the recycle of nitrate, the lower the degree of P removal obtained in the process (see Figure 7.13). If nitrates cannot be detected in the anaerobic zone and a good phosphate release is occurring, then phosphate uptake is possible. If nitrates cannot be detected in the anaerobic zone and phosphate release is still at low levels, then the nitrate feedback in the return sludge might be just sufficient to prevent proper release. Although the key to obtaining low return sludge nitrates is efficient denitrification in the biological reactor, any additional denitrification which can be achieved in the sludge layer in the final clarifier can be a great help. Although the design of the clarifier is important in this regard, adaptations in clarifer operation can also help. For example, reducing the clarifier underflow rate (recycle ratio) will mean that the sludge will stay a longer time in the clarifier, thicken up a bit more and thus consume a larger amount of residual nitrates. Care should be taken however, in reducing clarifier underflows to not reduce the solids handling capacity of the clarifier (see Chapter 8) and risk solids carryover in the effluent. Also, if solids are retained too long in the clarifier, all the residual nitrates can be consumed, the sludge will go anaerobic and a phosphate leach-out into the effluent will occur. In particular, this could take place in a very active sludge having a relatively low nitrate content or where the sludge surface level is standing fairly high in the clarifier.

The easiest way of ensuring that the return sludge nitrates are low is to ensure that the effluent from the biological reactor has a low nitrate concentration. This depends on the denitrification efficiency which depends on many other factors described elsewhere in this document.

Nitrate feedback to the anaerobic zone can also occur via backmixing from adjacent anoxic zones. For example, when the baffle separating the anaerobic and anoxic zones has ports in it to allow transmission of flow, the mixers in the anoxic zone can easily backmix mixed liquor containing nitrate into the anaerobic zone. In multi-mixer plants, the mixer nearest the baffle could be turned off to minimise this effect. (This should only be done if the remaining mixer/s can still keep the solids in suspension).

4.1.2 Oxygen input to anaerobic zone

Experience has shown that molecular oxygen can enter the anaerobic zone from a number of sources and if difficulties are experienced in obtaining suitable phosphate release, the operator should investigate and try and eliminate these unwanted oxygen inputs.

Excessive turbulence in the streams feeding into the anaerobic zone and in the zone itself can introduce unnecessary oxygen. For instance, cascades, screw pumps and aerated grit channels can oxygenate the feed sewage while screw pumps in the sludge recycle line can have the same effect. Excessive mixing in the anaerobic zone could also be detrimental.

Where this occurs and more than one mixer is available, consideration should be given to operating fewer mixers. In addition, by alternating the operating mixers, better utilisation of mechanical plant will be obtained.

Although many of the sources of oxygen input in the anaerobic zones of existing plants will be due to inherent design faults which the operator will not be able to correct easily, knowledge of the extent of this problem can be of great use in plant operation.

4.1.3 Retention time

The basic actual retention time of the liquid in the anaerobic zone will have been fixed by the designer and the operator will have little leeway to vary it. In cases where suitable anaerobic conditions are difficult to achieve consideration should be given to increasing the retention time in this zone by reducing the sludge recycle rate. This will also reduce the mass of nitrate recycled to the anaerobic zone. Before reducing the sludge recycle however, the operation of the final clarifiers should be considered. When the underflow recycle from the clarifiers is decreased, the solids handling capacity of the clarifiers is reduced as well, so that if the underflow is reduced too much, a solids flux failure will occur in the clarifiers and activated sludge solids will be carried into the effluent. Thus, there will be a limit to how much the sludge recycle can be reduced, and operators should be aware of this limit before considering ways of increasing the anaerobic retention time. In general, plants should be operated with as low an underflow recycle as possible while maintaining a safety factor in the clarifier solids handling capacity for peak load conditions. Large feed flowrates above the design level can also drastically reduce the anaerobic retention time. These can occur during storm-flow conditions so that consideration should also be given to bypassing storm peaks to the anoxic or aerobic zones.

The anaerobic retention time can also be increased by making the first anoxic zone anaerobic as well. This can be done by cutting the recycle of mixed liquor to this anoxic zone. If this is done however, the operator must be sure that good denitrification can be achieved elsewhere in the plant, that is, by creating anoxic zones in the aerobic zone. This option is usually only available with surface aerators when certain aerators (usually at the inlet end of the aerobic zone) can be switched off to create an anoxic zone. If this cannot be done, i.e., in diffused air plants, and plants with limited aeration capacity, then the effluent nitrate levels will increase when the mixed liquor recycle is cut, nitrate feedback to the anaerobic zone will increase and any beneficial effect of the bigger anaerobic zone will be negated. Also, it should be remembered that aerators should not be switched off at the expense of achieving adequate dissolved oxygen levels in aerobic zones. On certain plants treating a high strength sewage having a large readily biodegradable content, phosphate release in the anaerobic zone could be excessive and the retention time in the aerobic zone would then be inadequate for good phosphate uptake. In this case the retention time in the anaerobic zone can be *decreased* by increasing the sludge recycle. The sludge recycle should be increased only until the level of phosphate release drops to the optimum level (say 20 mg P/ℓ in the case of the earlier example.

4.2 Anoxic zone or primary anoxic zone

This zone is the main denitrification reactor in the process. It is fed by the effluent from the anaerobic zone and the mixed liquor recycle from the aerobic zone. Nitrates can be present in solution but oxygen inputs should be restricted.

The size of this zone will have been fixed by the designer using denitrification kinetics and the mass of nitrate recycled to it can be varied by changing the mixed liquor recycle rate. (see Chapter 6).

If desired, the size of this zone could be increased on surface aerator plants by switching off the first few aerators in the adjacent aerobic zone. The first section of the aerobic zone will be anoxic and the operating aerators will backmix nitrates into the zone, further assisting denitrification. This practice can only be done in surface aerator plants with excess aeration capacity. This advantage afforded by surface aerator plants can be further exploited in multi-aerator zones where anoxic areas can be created in the aerobic zone by the judicious switching off of certain aerators.

The introduction of oxygen into the anoxic zone can be detrimental so that knowledge of its points of ingress is important. High mixing intensities can change the liquid surface rapidly and bring oxygen from the atmosphere. In plants where this might occur, consideration should be given to not operating some of the mixers in this zone. Also, oxygen can enter via the mixed liquor recycle from the aerobic zone, particularly when high intensities of aeration occur near the mixed liquor abstraction point. In this case, aeration can be tapered off near the end of the aerobic zone (provided it is not completely mixed), to give lower dissolved oxygen levels at the abstraction point. Oxygen can also enter the anoxic zone via backmixing through gaps in the baffle separating it from the aerobic zone. Surface aerators can easily backmix oxygen through holes in this baffle and a possible remedy is to modify the baffle to create one-way flow only, or to switch off the aerators nearest the baffle.

The mixed liquor recycle rate from the aerobic to the anoxic reactor is a variable that can be controlled. An upper operating limit for the flowrate would be six times the feed flowrate, i.e. 6:1 and this can be reduced as low zero, as mentioned earlier. A good operating figure in most cases would be 4:1. The optimum would be to control the mixed liquor recycle so that nitrate concentrations are not very high in this anoxic zone. Due to variations in the strength and volume of the feed sewage and the usual lack of on-line nitrate monitoring, this ideal is not attainable and usually a fixed ratio of say, 4:1 is set and is not changed very often. Even with the mixed liquor pumps not operating, denitrification can still occur in certain plants. For example, surface aerators on some plants backmixed nitrates to the anoxic zone through holes in the baffles and in highly loaded plug flow diffused air plants the oxygen demand at the inlet end of the aerobic zone could be so great that the aerators cannot cope, oxygen levels drop to zero and an anoxic zone is created. If the mixed liquor recycle to the anoxic zone is zero and other backmixing does not occur, then this zone becomes a further anaerobic zone and additional release of phosphate can occur. This can be beneficial to phosphate removal when the main anaerobic zone is not functioning properly. When good release of phosphate occurs in the anaerobic zone however, a further release in the anoxic zone could be detrimental.

In the case of a plant treating a high strength feed with a low TKN/COD ratio and having a high readily biodegradable fraction, a good release will occur in the anaerobic zone and nitrates will be readily consumed in the anoxic zone causing it to go anaerobic despite a mixed liquor recycle ratio of say 4:1 being used. A further phosphate release could then occur in the anoxic zone. The degree of release could then be so high that the phosphate will not be taken up by the biomass again in the time allowed in the aerobic zone. Thus, despite there being a good phosphate release, phosphorus removal will be bad due to poor uptake in the aerobic zone. The remedy for this would be to reduce the retention time in the anoxic zone by increasing the mixed liquor recycle to 6:1 or even 8:1 provided pumping capacity for this is available.

At the opposite end of the scale, plants which treat a low COD waste with a small readily biodegradable fraction, will have relatively high nitrate concentrations in the primary anoxic zone (e.g. 5 mg N/ℓ) when the mixed liquor recycle is 4:1. In this case the mixed liquor recycle rate is excessive and could be reduced to prevent overloading the zone with nitrates. If however, nitrates cannot be reduced by lower recycles then inadequate COD is available in the feed sewage or dissolved oxygen levels in the returned mixed liquor are too high.

In the typical nutrient removal plant phosphate and ammonia concentrations in mixed liquor solution will drop between the anaerobic and anoxic zones. This is due to the dilution effect of the recycled mixed liquor. If however, they decrease to a greater extent than can be accounted for by dilution, particularly in the case of ammonia, then the oxygen input to this zone is too great, nitrification is occurring in the anoxic zone and the process is probably heading for a failure. Ways should then be found to restrict this oxygen input as described earlier.

4.3 Aerobic zone

The function of this zone is to oxidise organic matter in the sewage, to oxidize ammonia to nitrate and provide and environment in which the biomass can take up all the phosphate released in the anaerobic zone, plus all the phosphate which enters the process in the feed sewage. The reactor size and dissolved oxygen levels should be such that all these processes can operate effectively.

Although the size of this zone will have been fixed in the design so that nitrification will occur in the process, the DO levels in this zone and the aerobic sludge age or overall sludge age should be such that nitrification will occur at all operating temperatures. This should be borne in mind particularly when aerators are switched off or aeration is trimmed to create anoxic zones within the aerobic zone.

As mentioned earlier, the primary anoxic zone size can be increased by switching off aerators at the inlet of the aerobic zone, but aerators can also be cut at the end of this zone to increase the size of the second anoxic zone and/or to reduce DO levels in the mixed liquor recycle to the primary anoxic zone. However, it should be remembered that increasing the size of the first anoxic zone will be far more profitable than increasing the size of the second as the denitrification rate in the latter is very slow. (See 4.4). If possible therefore, the operating aerobic sludge age should be 1,5 times the minimum required by nitrification kinetics at the operating temperature. For example, if the minimum required in winter is 8 days then an aerobic sludge age of 12 days should be used. This would probably give in many plants a total sludge age of about 20-25 days.

In most aerobic zones where completely mixed conditions do not occur the oxygen demand of the mixed liquor is higher at the inlet end. Therefore, if possible, more air should be introduced here than at the outlet end. That is, if possible, tapered aeration should be used. However, the latter is not easily achieved with equally spaced surface aerators. Where it is possible to vary aerator depths, then the aerators near the inlet could be set lower in order to introduce more oxygen. With diffused air plants, diffuser spacing might be such that tapered aeration has been built into the plant or the operator can vary air supplies to various points in the zone to give the correct aeration pattern.

Phosphate is taken from solution into storage by bacteria in the aerobic zone. Various factors can affect the rate of phosphate uptake by the bacteria and ideally uptake should be complete at the end of the aerobic zone so that only trace quantities (i.e. $<$0,2 mg P/ℓ) are detected in solution at this point. Aeration rate or aeration intensity can affect phosphate uptake so that if DO levels are too low, phosphate uptake will be suppressed and relatively high phosphate concentrations will be in solution at the end of aeration (i.e. 2,0 mg P/ℓ). Therefore, if effluent phosphate concentrations are high despite there being a good release of phosphate in the anaerobic zone, lack of aeration in the aerobic zone is often the cause. (Sometimes however, an increase in oxygen levels can cause a slight increase in effluent phosphate concentration instead of a decrease as described in 3.1 above).

Aeration should therefore be adequate to

(i) promote rapid uptake of released plus feed phosphorus,

(ii) ensure oxidation of carbon compounds and ammonia, and

(iii) suppress the growth of filamentous microorganisms which produce poor settling sludges.

The aeration rate should be well matched to the incoming load variations and be just adequate to ensure the above, for if excess aeration is used, then:

(i) Further nitrification can occur when some of the organic nitrogen bound up with the sludge is converted to nitrate. This increases the quantity of nitrates to be removed and can result in higher effluent nitrate levels, a possible greater nitrate load on the anaerobic zone and the eventual loss of phosphorus removal. This is however, not a critical factor but emphasizes that good dissolved oxygen control is important in these plants.

(ii) The slow aerobic release of absorbed phosphate will be encouraged resulting in unnecessarily high effluent phosphate levels. This is much slower than anaerobic release.

(iii) Wastage of energy will occur.

(iv) High DO feedback to anoxic zones will occur thus reducing their efficiency.

(v) Floc break-up can occur leading to pinpoint floc in effluent.

It is difficult to give a recommended DO concentration which will suffice in all operating conditions. Experience has shown however, that for the above criteria to be met, the average DO in the aerobic zone should be about 2 – 3 mg/ℓ and should not be allowed to drop much below 1 mg/ℓ at any one point. A DO profile starting at say 1 mg/ℓ near inlet and ending at 3 mg/ℓ, would be very suitable. DO levels in aerobic zones should never be allowed to drop below 0,5 mg/ℓ as this will encourage filamentous growths and suppress nitrification. Excessively high DO levels should be avoided, i.e. 4 – 5 mg/ℓ, as these can cause floc break-up to pinpoint floc and wastage of power. By trial and error an optimum DO level for each plant can be found. But, it should be remembered that if filamentous growths are allowed to proliferate when DO levels are dropped, it can take many months of prolonged aeration to restore good settling sludge. Often, the monitoring of ammonia and nitrate levels in filtered mixed liquor samples from the end of the aerobic zone can give a good indication of aeration levels. For example, high ammonia levels and low nitrate levels could indicate inadequate aeration or *vice versa*.

On some plants which experience low strength sewages or variable strength with say, low COD levels over weekends, inadequate COD might be available for good denitrification. This would mean that when the feed COD was low, effluent nitrate would be high and phosphorus removal could be lost. At such times, it might be advantageous to suppress nitrification by reducing DO levels and allowing some ammonia out with the effluent. The General Standard would still be adhered to but the effluent ammonia levels would be allowed to increase from say, 1 mg N/ℓ to 7 mg N/ℓ.

This could however, encourage the growth of filamentous organisms which produce very poor settling sludges and thus be counter-productive in the long-term. For these reasons the suppression of nitrification should only be considered as a short-term emergency procedure.

Operation under these conditions is difficult as one tends to have an 'all or nothing at all' situation, i.e. if the DO levels are up, nitrification will be complete. If they are too low, nitrification will tend to stop completely and effluent ammonia levels will exceed Standard requirements. However, it is possible by judicious aeration control, to allow some ammonia to leave with the effluent and thus keep nitrate concentrations low. DO levels for this operation will usually be below 1 mg/ℓ and possibly below 0,5 mg/ℓ so that the risk of encouraging filamentous growths should be kept in mind.

During periods of low load it is also desirable to reduce aeration in order to match it to the load and obviate over-aeration and consequent over-nitrification. To do this successfully can create problems, particularly in the early life of new plants when aeration capacity installed to handle future loads remains unused.

On diffused air* plants, aeration might not be able to be reduced enough without going below the minimum airflow requirements for diffusers, i.e. there might not be sufficient turn-down capacity.

In the case of surface aerators, particularly those having small numbers of aerators, so many aerators might have to be turned off that mixing efficiency is lost and solids settle on the floor of the aeration tank and go anaerobic. This will cause a release of phosphate into solution so that when this sludge is resuspended, the released phosphate could be washed out with the effluent before it can be taken up by solids again. The anaerobic conditioning of the sludge on the floor can be beneficial to the process as it will stimulate the sludge to accumulate excess phosphate but the problem of phosphate washout cannot always be obviated.

With both types of aeration care must be taken not to reduce oxygen input to levels where phosphate uptake in the aerobic zone is prejudiced. It should also be borne in mind that filamentous organisms which can cause the activated sludge to settle poorly are encouraged by large unaerated zones and low dissolved oxygen levels in the aerobic zone. Operators should therefore take care when reducing aeration, not to produce an environment that will encourage these nuisance growths.

For good control of oxygen input the number and position of surface aerators is important as is the distribution of air in diffused air plants. In surface aeration it is better to have a number of smaller aerators than a few large aerators. This gives a greater flexibility.

The placement and maintenance of dissolved oxygen probes is an important aspect of aeration control. If possible three or four probes should be placed at strategic points in the aerobic zone. In the case of surface aerator plants, the probe should not be placed too near the operating aerators or they will read non-representative high values. Also they should not be placed in areas where aerators are to be switched off for long periods. The variation in dissolved oxygen levels with tank depth should also be borne in mind and the measurement of dissolved oxygen profiles with a portable meter can give a good idea of the best location for the probes. In fact, a good, reliable, portable dissolved oxygen meter should always be available for calibrating the *in situ* probes and checking dissolved oxygen levels elsewhere in the process. Continuous recording of the measured dissolved oxygen values can also be very useful in plant operation. Dissolved oxygen probes should also be cleaned out and calibrated weekly and regularly serviced.

Although the control of aeration via the reading from dissolved oxygen probes will suffice in most cases, it is possible to control oxygen input by making use of ammonia and ortho-phosphate concentrations in the mixed liquor at the end of the aerobic zone. Experience has shown that *provided an adequate release occurs in the anaerobic zone*, when aeration is decreased in the aerobic zone, phosphate and ammonia levels will be increased and when aeration is increased they will decrease. When both parameters reach approximately zero concentration, the optimum aeration level has been reached. Thus, by monitoring trends and changes in ammonia and phosphate levels in the effluent or if possible, at the end of the aerobic zone and phosphates in the anaerobic zone, it is possible to control the aeration to optimum levels. Such a system however, will depend on the use of on-line analysers and is thus only suitable for large plants. Small plant operators could also approximate the above system if quick on-site laboratory analyses were available.

4.4 Second anoxic zone

Further denitrification takes place in this zone. Experience on many plants has shown that because of the very slow denitrification rate in this zone, the quantity of nitrate removed by it is small, i.e. 1 – 2 mg N/ℓ. Therefore, this zone is only really needed when very low effluent nitrate levels are desired.

In the case of low strength sewages, particularly those containing small amounts of biodegradable solids (settled sewage), this zone will remove minimal amounts of nitrate. This is because most of the substrate will have to be metabolised in the aerobic zone, resulting in a sludge with a very low activity entering the second anoxic zone. For high strength sewages containing larger amounts of biodegradable solids, these solids might not all be consumed in the aerobic zone and thus create a higher activity and hence denitrification rate in this zone.

Sometimes, when treating a high strength sewage containing a fair proportion of biodegradable solids, activity in this zone might be so great that denitrification is complete; the zone goes anaerobic and phosphate is released again. This could be detrimental as this phosphate may not be picked up quickly enough in the final aerobic zone and effluent phosphate levels will be increased. A method of avoiding phosphate release in this zone is to allow DO or some nitrate to bleed into it from the main aerobic zone. Quantities should be just

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*Another disadvantage of diffused air plants is that one is not able to create anoxic zones in the aerobic zone as is possible with surface aerator plants.

sufficient to keep this zone from going anaerobic and not in excess, particularly in the case of DO. Sometimes, phosphates will continue to be picked up by the solids in this zone. This will usually take place when DO and nitrate enter it in sufficient quantities and can be useful when uptake tends to be slow in the main aerobic zone.

Although the feeding in of untreated feed sewage into this zone would be beneficial in speeding up denitrification, it is not really advisable as it adds phosphate which might not be removed in time, and ammonia which will tend to be nitrified in the final aerobic zone thus increasing effluent and return sludge nitrates. The latter would increase the nitrate load on the anaerobic zone unnecessarily.

4.5 Final aerobic zone

This zone has a short retention time and is used to increase the dissolved oxygen in the mixed liquor before it enters the final clarifier. Residual ammonia in the mixed liquor will continue to be nitrified in this zone and if phosphate had not been completely taken up in the previous zones it will also continue to be taken up in this zone.

Aeration should be sufficient to promote phosphate uptake and maintain good aerobic conditions but excess aeration should be avoided as it will encourage the conversion of organically bound nitrogen to nitrate and also cause the slow aerobic release of phosphate from the solids.

Mixing in this zone should always ensure homogeneous conditions. In the case of multiple final clarifiers this will ensure the equal distribution of solids to the clarifiers and when waste sludge is removed from this zone, will ensure that the correct quantity of waste sludge is withdrawn. For example, if such a zone has a number of surface aerators and those near the waste sludge take off are switched off to prevent over-aeration then a liquor lower in suspended solids than the mixed liquor could be wasted if settlement of solids occurs near the take-off point.

4.6 Final clarifiers

These should be operated to give the minimum suspended solids in the final effluent and a thick sludge for recycle to the inlet of the biological reactors. The use of the 'solids flux' approach will give the best procedure for optimising the design and operation of these clarifiers.

Effluents from biological nutrient removal plants usually contain quantities of nitrate so that the greater the suspended solids content in the return sludge, the less the mass of nitrate returned to the anaerobic zone. High return sludge solids concentrations are achieved by lowering sludge recycle rates or ratios, but in lowering these, care should be taken not to exceed the solids handling capacity of the clarifier. For a fixed feed-sewage flowrate, the lower the sludge recycle rate, the greater the retention time in the anaerobic zone so that in general, low recycle rates should encourage phosphorus removal. These recycle rates should however, still be above the minimum allowed by the solids flux theory in addition to allowing a safety factor and the handling of peak sewage flows.

Phosphate release into solution can taken place in final clarifiers particularly in the case of high activity sludges (i.e. sludges having a high respiration rate). With these sludges, DO and residual nitrate are quickly consumed and anaerobic conditions ensue in the sludge layer. Anaerobic phosphate release occurs. If this released phosphate remains in the liquid interstices in the sludge blanket, problems will not occur. In fact, the initiation of anaerobic conditions in the sludge in the clarifier might benefit the overall process. However, if this released phosphate diffuses into or is mixed into the upper layers of the clarifier liquid, it can get into the effluent and negate some of the phosphorus removal achieved in the biological reactor.

In this respect, the depth of the clarifier, the height of the sludge blanket and the distribution and retention time of solids in the clarifier, are of great importance. Therefore, in design, sloping bottom clarifiers which encourage a 'plug-flow' of sludge solids in the sludge layer are recommended. Mechanical scrapers would then compliment this system.

Although the operation of the final clarifier can greatly affect its performance, sludge settling properties have an overriding effect with the clarifier operations having to be adapted to the sludge settling characteristics. The latter are largely influenced by the operating conditions in the biological reactor, in particular, the DO levels in the aerobic zones and the fraction of the plant that is unaerated.

If phosphate release in the clarifier becomes a problem and contamination of the effluent occurs, this can be removed by bleeding a little more nitrate or DO from the biological reactor. This should not be overdone, as it will increase effluent nitrate levels and run a greater risk of nitrate feedback to the anaerobic zone.

Because this process appears to run the risk of encouraging the proliferation of organisms which cause a frothy growth on the biological reactor, final clarifiers should be equipped with good scum baffles and scum removal facilities. Scum withdrawal points at the end of the biological reactor would also be useful. Once this scum has been removed, it should not be recycled back to the plant inlet. Its disposal with the waste activated sludge is thus recommended.

The growth of algae in effluent launders is also a problem and facilities should be available to minimise these growths or to clean them off on a regular basis.

5. SLUDGE AGE CONTROL

As was shown in Chapter 4, the parameter sludge age is useful in the design of activated sludge plants. It can also be of great benefit in the operation of such plants.

It is desirable therefore to use the concept of sludge age control in the operation of nutrient removal plants in preference to the control of MLSS concentration. In sludge age control the aim is to operate at fixed predetermined sludge age (or solids retention time). The MLSS concentration will then vary freely in response to incoming COD load changes. Procedures for controlling sludge age are adequately covered in Chapter 4 and will thus not be repeated here.

6. REMEDIAL MEASURES WHEN NUTRIENT REMOVAL FAILS

Occasionally certain biological nutrient removal plants will stop removing phosphorus while others might be badly planned and designed, giving no excess phosphorus removal at all. When the usual operational adjustments (mentioned elsewhere in the text) do not bring results, the following options can be considered:

6.1 Supplementing COD

In many cases the failure of phosphorus removal will be due to the lack of COD, particularly readily biodegradable COD in the sewage feed to the reactors. This situation can sometimes be remedied by the addition of extra COD to the feed from sources within the works or from external supplies. It can be added continually or only during low load periods, i.e. weekends and holidays.

Primary Sludge. On works having primary sedimentation tanks, and where primary sludge is not needed for other purposes, (i.e. gas production). The sludge should only be added in quantities, which boost the feed COD into the range where phosphorus removal occurs as any extra might overtax the aeration system and waste energy. Addition should preferably be continuous but often batch dosing of the plant can be successful. One problem which might occur is the concurrent addition of rags with the sludge. On some plants these can tangle and foul up mixers and clarifier sludge withdrawal systems. In this case, precautions should be taken to remove these rags, i.e. additional screens on the sludge supply line.

Acid Digested Primary Sludge. When sludge is subjected to high rate fermentation some of the solid material is converted by bacteria to volatile fatty acids and other soluble organic compounds. The use of this material for the increase in COD load is very beneficial as the soluble organics will be a readily biodegradable carbon source to intensify denitrification and stimulate anaerobic zone release while the fatty acids would be the ideal substrate for certain phosphate accumulating bacteria. Either the whole fermented sludge could be added or the sludge could be partially settled and an acid supernatant liquor added. Acid sludge can be produced by the high rate digestion of primary sludges in a separate reactor or by allowing primary sludge to accumulate in sedimentation tanks or in any other holding tank.

Anaerobically Digested Sludge or Supernatant Liquor. Sludge from a conventional methane producing anaerobic digester can also be used as a supplementary COD source, although it is not as suitable as the above two sludges. Such a sludge will contain greater quantities of inert solids which do not benefit the process but increase the MLSS concentration. The relatively high concentrations of ammonia and phosphate could cause problems. The ammonia would have to be oxidized to nitrate which might not be denitrified completely. Also, the process might not be able to remove the additional phosphorus. The latter problem could be overcome by pre-treatment of say, the supernatant liquor to precipitate phosphorus and remove ammonia.

Waste Materials from Industry. Some industries produce waste biodegradable organic materials. These are often used in boosting the COD, particularly if they are soluble. Examples are certain food and chemical industries. Such materials could be discharged to the sewer feeding the works or tankered out and stored on site. The discharge of these wastes could be timed so that they arrive at the plant at the desired time.

Purchased Organic Compounds. When the above options are not available, the purchase of organic carbon sources such as methanol or acetic acid might be necessary, but in most instances this would prove prohibitively expensive.

6.2 Chemical addition

This practice provides a useful alternative to COD boosting with various carbon sources. Chemicals can be added to the activated sludge during periods when effluent phosphate peaks are likely to occur or continuously to remove any residual phosphate which the biological mechanism cannot remove, i.e. in the case of sewages having high phosphorus contents. Lime or ferric chloride are useful chemicals for this application with ferric chloride addition to the end of the aeration basin being a well-tried process. Any residual unused iron will then be absorbed onto the floc and remain in the process to be available for further phosphate removal. Iron and aluminium salts are usually added to the mainstream, i.e. in to the biological reactor, while lime is used to treat phosphate with side-streams such as the liquid phase separated from anaerobic zones. Either industrial chemicals can be bought in bulk or suitable chemical by-products can be obtained from metal or mining industries. In general, the addition of chemicals should only be considered as a last resort and the reader is referred to the WRC publication 'Guidelines for the Chemical Removal of Phosphates from Municipal Wastewaters' for details of methods that can be used.

7. UNWANTED MICROBIOLOGICAL GROWTHS

Biological removal plants are susceptible to two types of unwanted biological growths which can adversely affect their performance.

7.1 Nuisance scums

Experience has shown that biological nutrient removal plants tend to favour the growth of scums which accumulate on the surface of the biological reactor and final clarifiers. They are usually caused by the excessive growth of filamentous micro-organisms such as *Nocardia* and *Microthrix* which produce a surface active agent enabling them to be frothed up and float as a surface mat. These scums are an unsightly nuisance, can decompose, producing offensive odours and when they escape into the effluent, can significantly increase its COD and suspended solids content.

The exact cause of these growths is not known at present but it is possible that they are favoured by certain substrates in the feed sewage, or the environmental conditions in the plant (DO levels; temperatures). On certain plants these growths are seasonal occurring only in the winter months but on others they can occur at any time of the year.

Successful methods of controlling these scums are not readily available at present but possible remedies to be tried are:

- Surface spraying with heavily chlorinated water (i.e. surface chlorination).
- Complete physical withdrawal of all the scum on the plant and the scrubbing down of all surfaces to remove residual contamination which might reseed the process. In this respect, very good scum removal facilities are needed which are not available on many plants.
- Surface spraying with effluent or pumped mixed liquor.
- Changes in aeration.
- Drastic decrease in sludge age. A combination of any of the above could be more effective. Once the scum solids are removed from the plant they should be kept out and not recycled to the inlet of the works. Recycling can mean that the process is reinnoculated with a seed of these organisms.

7.2 Filamentous growths

Filamentous growths which result in poor settling activated sludges can also occur. Large quantities of these organisms can cause problems with final clarifier operation and result in the washout of activated sludge solids into the effluent.

Experience has shown that these growths appear to be favoured by long anoxic retention times. If the actual anoxic retention time in a plant is over 1,5 h, filamentous growths could be encouraged. Also if the dissolved oxygen values in the aerobic zones are allowed to drop too low (i.e. below 0,5 mg/ℓ), these unwanted growths will be encouraged.

Therefore, in order to suppress filamentous growths, actual anoxic times should be reduced as much as possible (without adversely affecting denitrification) and dissolved oxygen levels at any point in the aerobic zones should not be allowed to drop below 0,5 mg/ℓ). Further research is, however, needed to find better long-term solutions to these problems. An effective short-term remedy for filamentous growths is the wellknown practice of chlorinating the return activated sludge line. This can produce a dramatic improvement but is costly and requires careful control.

8. MONITORING

In general, because of their greater sensitivity, biological nutrient removal plants need more intensive monitoring than the usual sewage treatment processes. Although some authorities will have better laboratory facilities than others it is essential that these plants be monitored regularly to ensure proper operation.

For the typical 5-stage plant the following would be a recommended daily monitoring schedule with all samples being 24 h composites:

Feed sewage (24 h composite): COD, TKN, NH_4, Total P, o-PO_4
Anaerobic zone* (24 h composite of filtered samples): NO_3, o-PO_4
First anoxic zone* (24 h composite of filtered samples): NO_3 o-PO_4
Main aerobic zone* (24 h composite of filtered samples): NO_3, o-PO_4, NH_4
Second anoxic zone* (24 h composite of filtered samples): NO_3, o-PO_4
Final aerobic zone* (24 h composite of filtered samples): NO_3, o-PO_4, NH_4, SS, SVI (or equivalent)
Return sludge* (24 h composite of filtered samples): NO_3, o-PO_4, SS
Final effluent (24 h composite sample unfiltered): NO_3, o-PO_4, NH_4, SS

Feed, return sludge and waste activated sludge flowrate should be monitored continuously as should the DO levels in the unaerated zones. Feed and return DO concentrations should be checked occasionally with a portable DO meter to detect unnecessary oxygen input to these zones. A portable DO meter is also useful to check and calibrate permanently installed DO meters.

The analysis of filtered mixed liquor samples from the various zones can provide very useful information about the operation of a plant. These samples should be taken and filtered *immediately*. Either a snap sample can be taken or a 24 h composite whereby the zone is sampled every hour, the sample is filtered *immediately* and the filtrates composited into one sample. It is very important to filter the mixed liquor samples quickly as the sample will change if stored in the presence of solids (i.e. nitrates will be consumed and phosphates will be released into solution).

The feed and effluent samples may be analysed on a weekly basis for other parameters such as BOD, pH, alkalinity, OA, etc., and weekly microbiological examination of the mixed liquor is also recommended.

Occasionally a 24 hour run can be done on the plant in which the feed, effluent and possibly some of the zones would be sampled hourly and each sample tested to determine the variation in plant performance.

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*If a 24 h composite of filtered samples is not possible to obtain, as many snap samples as possible per day should be taken and filtered immediately after being obtained.

This procedure is very useful in picking up operational problems such as the appearance of unnecessary phosphate or ammonia concentration peaks in the effluent. If such peaks can be detected, operational changes could be made to eliminate them and improve overall performance.

In addition, the feed sewage could occasionally be monitored to determine the concentration of certain metals such as Ca, Mg, Fe, Al, Zn and Mn.

On-line chemical monitoring systems using automatic analysers or specific ion electrodes can be very useful in detecting plant performance variations but their use will probably only be economically justified on larger installations. These units are also quite expensive and require trained personnel to maintain them.

Although the above sampling and analysis programme may seem ambitious to certain smaller authorities, it is a model which should be aimed at. The absolute minimum requirement to achieve reasonably good operation would be the once weekly analysis of 24 h composite samples for o-PO_4, NH_4, NO_3 and suspended solids, with continuous monitoring of flows and DO levels. Also, snap samples of filtered mixed liquor and effluent should be analysed for NH_4, NO_3 and o-PO_4 as often as possible.

8.1 Typical results of chemical monitoring

Table 11.1 shows examples of typical results which could be expected from a five-stage plant under various modes of operation.

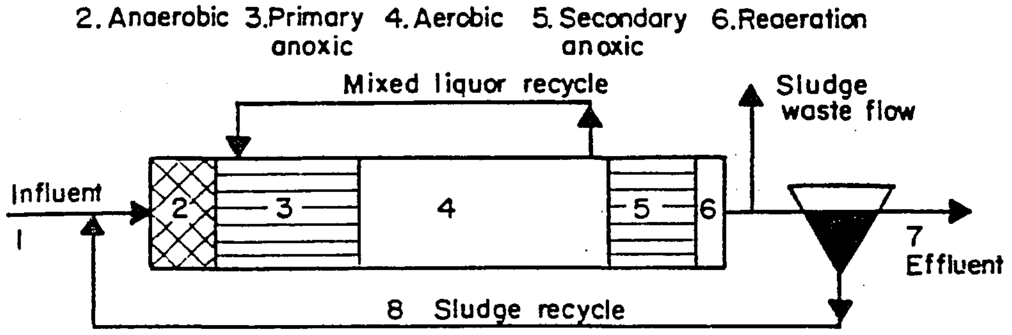
Example A: Shows probably the ideal situation with good nitrogen and phosphorus removal. A substantial phosphate release has occurred in the anaerobic zone with food uptake in the aerobic zone leading to very low effluent phosphate levels. Nitrification and denitrification are good while the additional denitrification in the final clarifier has resulted in very low levels of nitrate in the return sludge. The nitrate load on the anaerobic zone was thus minimal, contributing greatly to the achievement of suitable anaerobic conditions and release in the first zone.

Example B: Shows a plant where no excess phosphorus removal occurred. Feed and effluent ortho-phosphate concentrations are much the same with no release in Zone 1. Virtual complete nitrification has occurred but denitrification is only partial probably due to the lack of COD in the feed sewage. Effluent nitrate levels are high, reflecting in high return sludge nitrates and a large nitrate load on the anaerobic zone manifested by the 1 mg N/*ℓ* nitrate in this zone and the lack of release.

Example C: In this case, denitrification is adequate, the nitrate returned to the first zone is low and anaerobic phosphate release is good. However, effluent phosphate levels are relatively high due to a poor uptake in the aerobic zones. This slow rate of uptake is probably due to a lack of aeration in Zone 3 which is also shown by the poor nitrification in this zone. In this case the aeration could have been decreased too much. In cer-

TABLE 11.1: TYPICAL CONCENTRATION PROFILES THAT CAN BE FOUND IN 5 – ZONE PHOREDOX PLANTS

		(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
A	o-PO_4/P	6	20	6	0,4	0,2	0,2	0,2	—
	NH_4/N	30	14	4	0,6	0,5	0,5	0,5	—
	NO_3/N	0	0	0,3	2	2	2	2	0,5
B	o-PO_4/P	6	6	6	6	6	6	6	—
	NH_4/N	30	15	4	0,6	0,5	0,4	0,4	0,5
	NO_3/N	—	1,0	9	15	15	16	16	15
C	o-PO_4/P	6	20	8	2,5	2,8	2,0	2,0	—
	NH_4/N	30	25	13	8	8	7	7	—
	NO_3/N	—	0	2	3	3	3	3	1
D	o-PO_4/P	6	30	15	3,5	2,2	1,8	1,8	—
	NH_4/N	30	15	5	0,2	0,2	0,2	0,2	—
	NO_3/N	—	0	0,1	0,6	0,6	0,6	0,6	0,2



tain cases the ammonia levels could be much lower than indicated in this example and the aeration could stil be inadequate for complete phosphate uptake.

Example D: Here the phosphate release in the anaerobic zone is very high and is followed by a further release in the primary anoxic zone (a mass balance using a mixed liquor recycle of 4:1 and a return sludge recycle of 1:1 will confirm this). This excessive release has meant that phosphate uptake in the aerobic zone is incomplete, resulting in phosphate washout into the effluent. It can be seen that phosphate uptake under aeration was poor, despite there being enough oxygen to achieve complete nitrification.

9. SCHEME FOR PROBLEM SOLVING IN NUTRIENT REMOVAL PLANTS

Figure 11.1 shows an algorithm for the identification and solving of problems in Phoredox plants and incorporates many of the concepts expounded in earlier sections. It should be studied by operating personnel and used as a means of identifying problem areas and suggesting solutions.

In this algorithm ▭ represents an action. For example, the first action at the start is "measure effluent phosphate concentration". The symbol ⬭ represents a question and decision, i.e. one must decide 'yes' or 'no' whether the effluent phosphate concentration is below 1 mg P/ℓ or not. The symbol ⊖ means that one ceases action as the process should be performing as required. One then merely follows the arrows through the algorithm until a solution to the problem has been found or the effluent phosphate concentration is below 1 mg P/ℓ.

10. TREATMENT OF WASTE ACTIVATED SLUDGE

When the sludge from a biological nutrient removal plant is allowed to go anaerobic, it releases accumulated phosphate. This fact should be borne in mind when considering methods for the treatment of waste activated sludges.

- If phosphate release into the liquid phase is to be minimised, treatment should take place as soon as possible after removal from the process, while the sludge is still in a fresh 'aerobic' state. The time available before significant release of sludge occurs, depends on the activity or respiration rate of the sludge.

- In thickening and dewatering, liquid low in phosphate concentration should be removed quickly from the sludge, particularly if these liquids are to be recycled back to the process.

- When large quantities of phosphate have been released as in anaerobic digestion, supernatant liquors and filtrates from sludge dewatering should not be recycled to the main process. These will contain high concentrations of phosphate and will result in a phosphate build-up and a consequent loss of phosphorus removal.

- Phosphate rich liquors should either be chemically treated to remove phosphate or disposed of on land.

Gravity thickening of waste activated sludge should be approached with care. Although it is a relatively cheap operation, the risk of anaerobic phosphate release is real. If gravity thickening is done quickly, release can be minimised. Therefore, sludges with good thickening properties (low SVI), which are well stabilised (low respiration rate), could be gravity thickened without large releases of phosphate. As with final clarifiers, phosphate will be released mostly in the sludge layer at the bottom of the thickener where it will do no harm provided it is not allowed to diffuse or be mixed into the upper layers. If this thickened sludge is subsequently dewatered, the filtrate or centrate will then contain the released phosphate. Gravity thickening will usually give thickened sludges of 2 – 3 % solids.

Dissolved air flotation thickening is an alternative operation of phosphate rich sludges. It thickens far quicker than gravity thickening and beause air is injected, the sludge remains aerobic for longer periods. This operation will therefore give a liquid phase of low phosphate content and a thicker sludge than gravity thickening. Float solids concentrations vary from 3,0 to 6 %, depending on sludge properties. It is obviously more costly than gravity thickening.

Dewatering. Continuous dewatering equipment such as belt-presses or centrifuges provide a rapid means of activated sludge dewatering, thus minimising phosphate release during this operation. Experience has shown that belt-presses give drier cakes than centrifuges. Even drier sludge is produced by the plate filter press but because of the relatively longer pressing time, phosphate release will be greater in this type of equipment. When dewatering on drying beds, the recycle of bed underflow to the biological reactor should be avoided as it will usually be rich in phosphate due to anaerobic conditions in the sludge layer.

Digestion. anaerobic digestion provides a convenient way of stabilising waste activated sludge, particularly if it is carried out in conjunction with primary sludge. However, anaerobic digestion does cause the solubilisation of accumulated phosphate so that the disposal of digested sludge or digester liquors must be carefully considered. If anaerobic liquors are to be recycled to the liquid phase they must be chemically treated to remove phosphates. Also, the presence of activated sludge in an anaerobic digester can deteriorate the thickening properties of the digested sludge.

- Aerobic digestion will also stabilise the sludge and cause the release of phosphates into solution. Aerobic phosphate release is generally slower than anaerobic release and is via a different mechanism, that of cell lysis. Also, if the pH in the aerobic digester is reduced by nitrification of organic

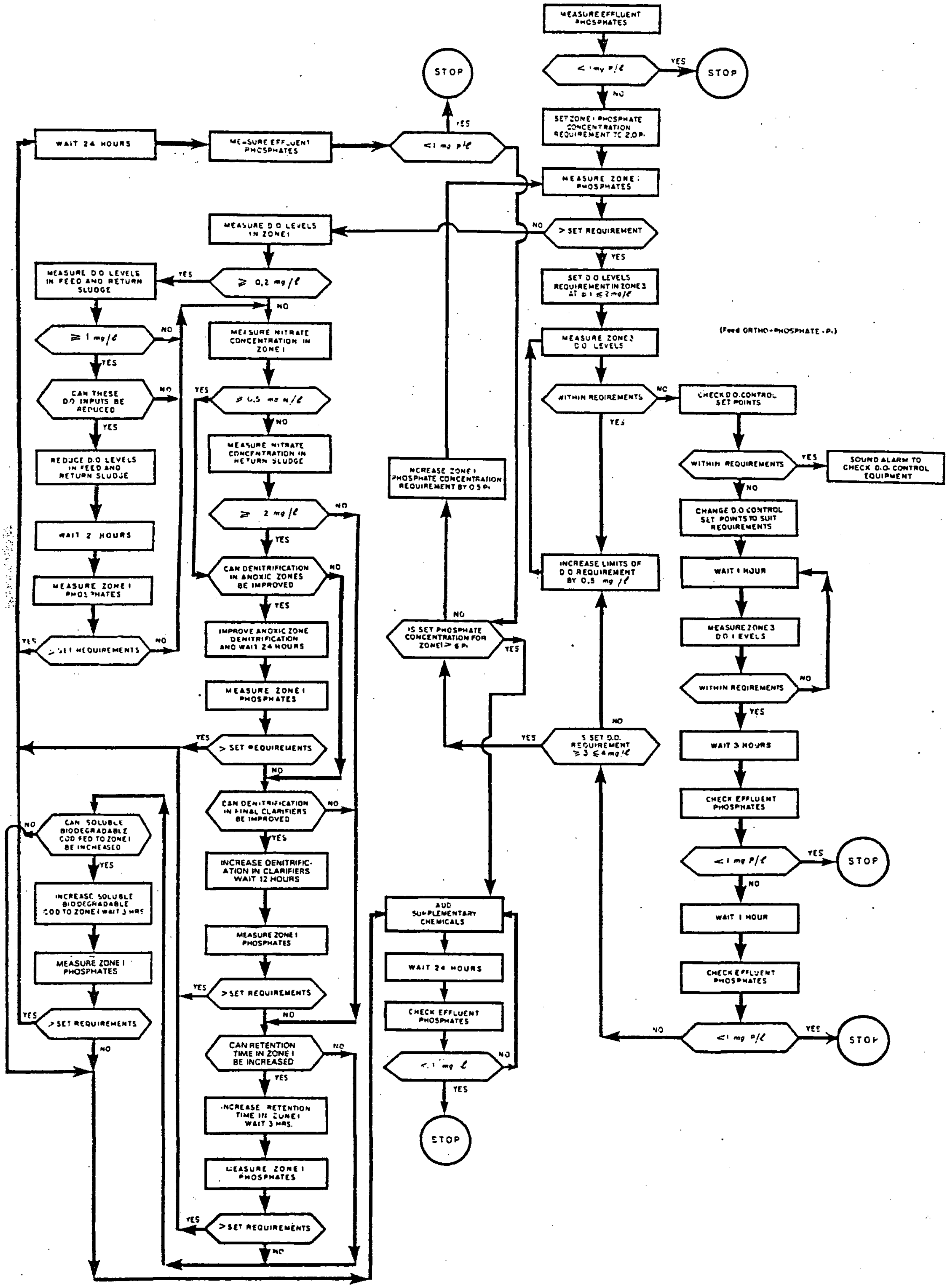


FIGURE 11.1: Scheme for problem solving in biological phosphorus removal plants

nitrogen in the sludge, phosphate release can be dramatically increased. This could be partially avoided by incorporating a denitrification step in the digestion process. The anoxic step will preserve the alkalinity of the system and mean that chemicals such as lime will not have to be added to maintain the pH.

11. SLUGE DISPOSAL

Because the sludge from this type of process will be higher in phosphorus content than the other sludges, agricultural disposal could be beneficial. However, care should be taken to ensure that the sludge is properly stabilised either anaerobically or aerobically before it is allowed out into the environment.

- Activated sludge cannot be really stabilised in the aeration tank so that once away from an aeration source it will putrefy readily, causing significant odour problems. Thickened and dewatered activated sludges are even more prone to causing odour problems and should be stabilised or disposed of as soon as possible.
- In dry climatic areas sludge may be disposed of on land without problems, provided it is spread thinly and not allowed to accumulate in pools (in the case of liquid sludge) or heaps (in the case of dewatered or dried sludge).

12. DISPOSAL OF LIQUORS

Liquids separated during the thickening and dewatering of sludges can be recycled to the liquid phase, providing the returned nitrogen and phosphorus can be removed in the main process. If this is not possible, intermediate chemical treatment for their removal will be necessary or the liquors can be disposed of elsewhere. For example, the irrigation of liquor onto agricultural land could be possible. Agricultural disposal has attractive possibilities as chemical precipitation can often be costly.

13. COMPLIANCE WITH STANDARDS

In designing nutrient removal plants, the designer must be aware of the effluent standards requirements and the ability of the proposed plant to meet desired effluent quality limits. In this respect, the reliability of the process and the effluent quality variability must be considered.

In most of the "sensitive" areas in South Africa, the modified General Standards will have to be complied with. These are summarised in Table 11.2

TABLE 11.2: SUMMARY OF GENERAL STANDARDS

Parameter		Limit mg/ℓ
Chemical Oxygen Demand	COD	Less than 75
Ammonia	N	Less than 10
Suspended Solids		Less than 25
o-Phosphate	P	Less than 1,0

Regulations in terms of the 1956 Water Act require that these limits be adhered to at all times. However, because of the variability of any real process, 100 % compliance will not be possible. There will always be occasions when the limit is exceeded even in the most optimum designs. Therefore, in this chapter we will consider the case of say, 95 % compliance with the Standards with the aim being to design reliable, stable plants which will achieve this goal.

Statistical analyses of the data from sewage treatment plants has shown that the performance of these plants on an annual basis is quite variable. Also, the data is not normally distributed, with a log-normal distribution giving the best fit of observed data. This means that when plotted on log-probability graph paper, the data gives a straight line with deviations from linearity at either end.

Figure 11.2 shows a log-probability plot of data from the Goudkoppies plant in Johannesburg. When this plant was underloaded (top graph), it could only comply with the phosphate limit for just over 60 % of the time. However, when a high strength sewage was treated (lower graph), a 94 % compliance was possible.

Figures 11.3; 11.4 and 11.5 show similar graphs for ammonia, COD and nitrate. Although there is at present no nitrate standard it is interesting to note how improved phosphate removal coincides with better denitrification.

As can be seen the ammonia limit of 10 mg N/ℓ was very easy to meet, in fact, better than 99 % compliance was possible for both sets of data shown. In the case of the COD, better than 95% compliance with the 75 mg/ℓ limit was possible with the weak sewage but only 45 % when treating the strong sewage due to the latter's industrial waste component.

Tables 11.3 and 11.4 show further results of the statistical handling of these two sets of data. As can be seen, the geometric means (obtained by averaging the logarithms of the results), were usually much lower than the arithmetic means. This indicates that the data is "skewed", i.e., that there are a few results much higher than the bulk of the results. It is these results which prevent compliance with the standard 100% of the time.

Also shown on the bottom lines are the mean design values which should be used (bearing in mind the variability of the data) to achieve the General Standards for 95 % of the time. A nitrate limit of 10 mg N/ℓ has also been included for interest sake. It is interesting to note that to *achieve the phosphate limit of 1 mg P/ℓ for 95 % of the time, one must use a design value of ≅0,4 mg P/ℓ. Similarly, design values of ≅3 mg N/ℓ to achieve an ammonia of 10 mg/ℓ and ≅45 mg O/ℓ for COD of 75 mg/ℓ* must be used. (It should be remembered however, that the effluent COD is a function of the effluent suspended solids and the influent soluble

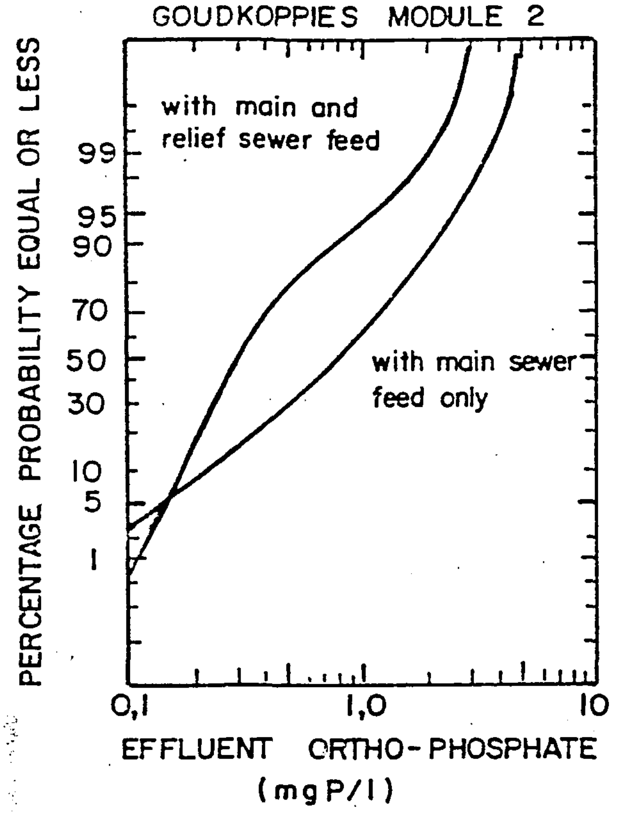


FIGURE 11.2: Log-probability plot of effluent phosphate concentrations from 4-hourly monitoring of Module 2, Goudkoppies

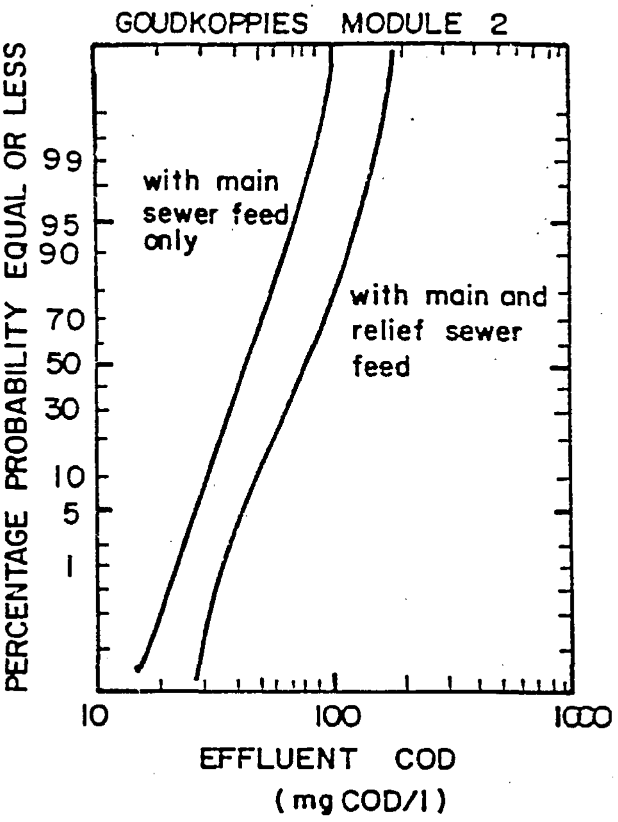


FIGURE 11.3: Log-probability plot of effluent ammonia concentrations from 4-hourly monitoring of Module 2, Goudkoppies

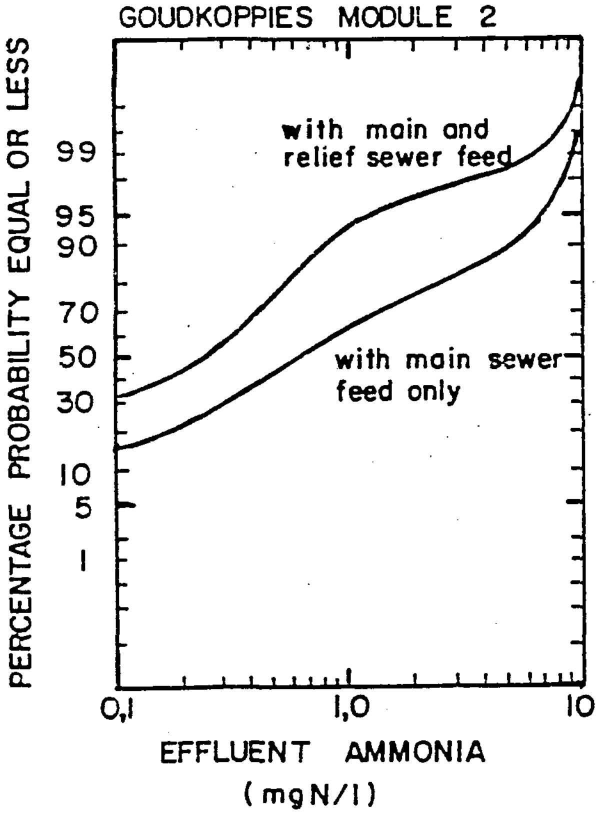


FIGURE 11.4: Log-probability plot of effluent COD concentrations from 4-hourly monitoring of Module 2, Goudkoppies

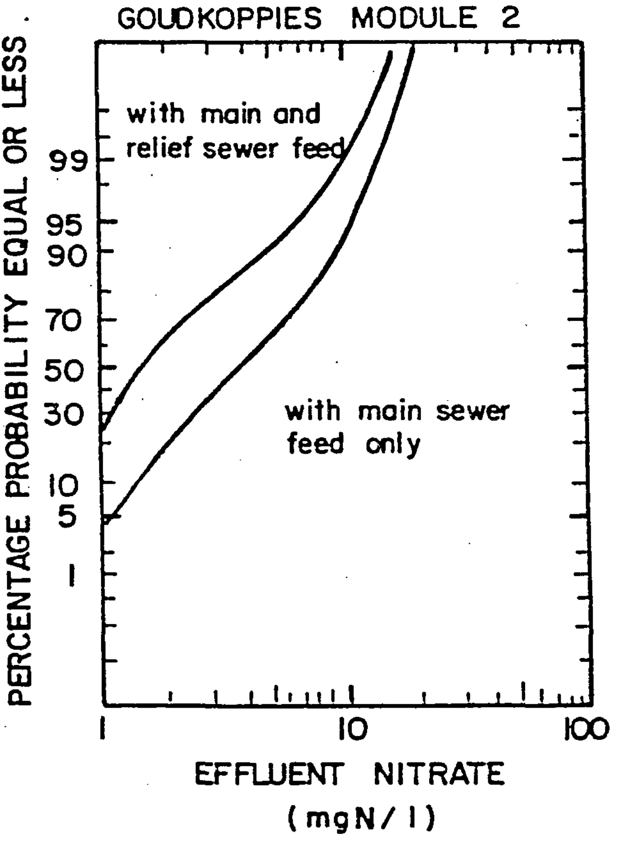


FIGURE 11.5: Log-probability plot of effluent nitrate based on 4-hourly monitoring of Module 2, Goudkoppies

TABLE 11.3: RESULTS OF A STATISTICAL ANALYSIS OF THE PERFORMANCE OF GOUDKOPPIES MODULE 2 WHEN TREATING THE *MAIN* SEWER FLOW
(4 hourly samples taken between 14 June and 31 December 1981)

	Total Phosphorus	Ortho-Phosphate	TKN	Ammonia	COD	Nitrate
Number of results	961	1 123	988	1 050	1 123	1 109
Arithmetic mean	1,11	0,89	2,71	1,33	43,43	4,51
Standard deviation	0,72	0,74	1,84	2,10	12,59	2,94
Coefficient of variation	0,65	0,82	0,68	1,58	0,29	0,65
Geometric mean	0,93	0,48	2,23	0,13	41,41	2,91
Mean design value for 95 % reliability	—	0,40	—	3,0	49	4,5

TABLE 11.4: RESULTS OF A STATISTICAL ANALYSIS OF THE PERFORMANCE OF GOUDKOPPIES MODULE 2 WHEN TREATING THE *MAIN* AND *RELIEF* SEWER FLOW
(4 hourly samples taken between 10 January and 11 July 1982)

	Total Phosphorus	Ortho-Phosphate	TKN	Ammonia	COD	Nitrate
Number of results	1 064	1 077	906	1 071	1 070	1 065
Arithmetic mean	0,66	0,36	2,78	0,39	77,18	1,60
Standard deviation	0,42	0,35	0,93	0,97	25,68	2,16
Coefficient of variation	0,65	0,91	0,33	2,52	0,33	0,35
Geometric mean	0,58	0,27	2,63	0,01	72,97	0,54
Mean design value for 95 % reliability*	—	0,45	—	2,7	46	3,1

*Based on an effluent Standard of COD <7/ mg/ℓ : Ammonia <10 mg N/ℓ o-PO_4 <1 mg P/ℓ and NO_3 <10 mg N/ℓ

non-biodegradable COD fraction and this cannot really be designed for).

Thus designers should use these values in their designs rather than the values in the General Standards in order to conceive plants which will comply with the Standard in a more reliable manner.

14. REFERENCE

NIKU, S., SCHROEDER, E.D., TCHOBANOGLOUS, G., AND SAMANEIGO, F.J. (1981) Performance of Activated Sludge Process: Reliability, Stability and Variability. USEPA Publication, December 1981.

CHAPTER TWELVE

SUMMARY OF CASE STUDIES ON BIOLOGICAL NUTRIENT REMOVAL AT

I. JOHANNESBURG'S SEWAGE PURIFICATION WORKS by A.R. Pitman

II. THE NIWR DASPOORT PILOT PLANT by A. Gerber

1. INTRODUCTION

Biological nutrient removal is today practised widely in South Africa, as well as in isolated instances in South West Africa, Zimbabwe, the United States, Canada and Europe. At the IAWPR post conference seminar on phosphate removal in biological treatment processes, held in Pretoria during April 1982, a number of case studies on biological nutrient removal were presented, amongst others, by Paepcke (1982), Pitman *et al* (1982), Kerdachi and Roberts (1982) and Barnard (1982).

In this chapter two case studies carried out under contract to the Water Research Commission are briefly dealt with. In the first case study operating experience with nutrient removal activated sludge plants in Johannesburg is described. For more detailed information the reader is referred to the joint publication by City Council of Johannesburg and the Water Research Commission entitled "Optimisation of the Modified Activated Sludge Process for Nutrient Removal" which can be obtained from the Water Research Commission, P O Box 824, Pretoria, Republic of South Africa. In the second case study information is presented of pilot-scale application of the Phoredox and UCT processes at the Daspoort Sewage Treatment Works in Pretoria. Greater details of this study are given in the joint National Institute for Water Research and Water Research Commission publications entitled "Technical Guide on Biological Nutrient Removal: The Phoredox Process" and "Biological Nutrient Removal from Wastewater Effluents: Performance Evaluation of the Phoredox and UCT Processes". Both these publications are obtainable from the Water Research Commission at the abovementioned address.

2. CASE STUDY NO. I

OPERATING EXPERIENCE WITH NUTRIENT REMOVAL ACTIVATED SLUDGE PLANTS – GOUDKOPPIES AND NORTHERN PURIFICATION WORKS, JOHANNESBURG

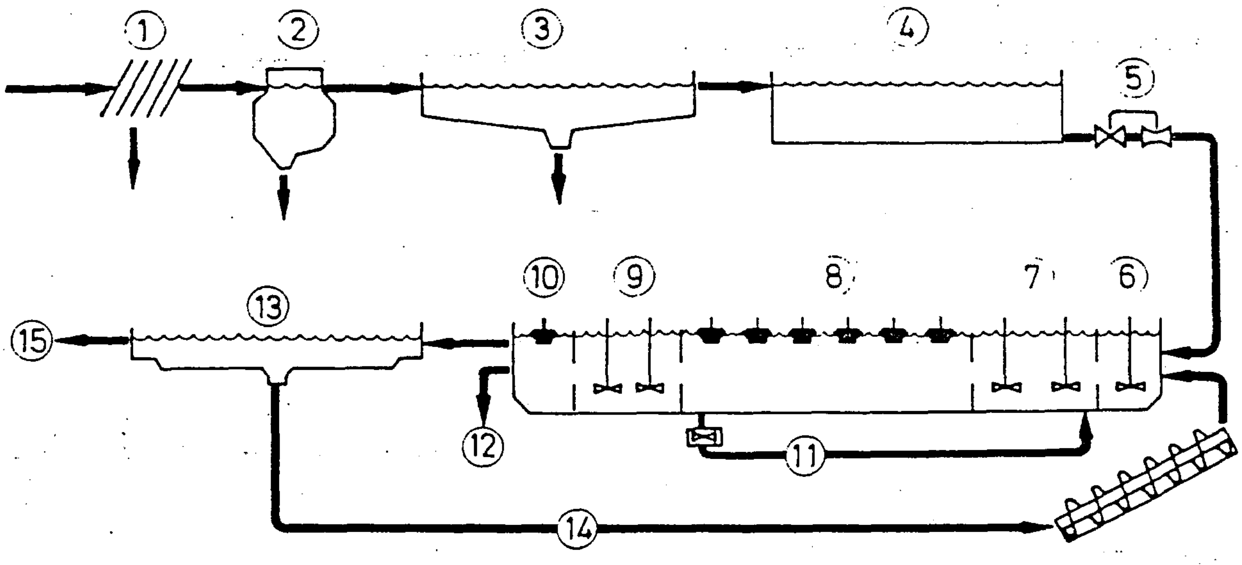
The Johannesburg City Council operates two large biological nutrient removal activated sludge plants at Goudkoppies and the Northern Works. Each has a design capacity of 150 Mℓ/d which is split into three 50 Mℓ/d modules incorporating the five-stage Phoredox layout (See Figure 12.1). Both plants have primary sedimentation tanks and settled sewage is fed to the biological reactors via in-line flow balancing tanks. These plants which were designed in 1973 and 1974 respectively, represent the initial phase of the process design evolution of biological nutrient removal plants. Each has anaerobic zones of approximately 1 hour nominal retention time which by 1983 technology is rather short. The performance of these plants should thus be viewed with the knowledge that their design does not represent the present state of the art of process design.

Experience at Goudkoppies covers two distinct periods; first when the plant was underloaded, receiving 50 per cent of design load and second, when the waste flow to the plant included a high strength flow containing considerable quantities of yeast industry effluent and the plant operated at approximately its design organic loading.

When the plant was underloaded, and the influent wastewater total phosphate was between 7 to 8 mg/ℓ (as P), an effluent with a mean ortho-phosphate concentration of just under 1 mg P/ℓ could be produced (see Table 12.1), but performance was erratic with frequent effluent phosphate peaks usually occurring after

TABLE 12.1: AVERAGE PERFORMANCE

Wastewater flow and characteristics		July 1980 – June 1981 Influent	Effluent
Mean daily flow	($m^3 \times 10^3$/d)	88	
Suspended solids	(mg/ℓ)	210	16
COD	(mg/ℓ)	640	48
OA (SABS)	(mg/ℓ)		7
TDS	(mg/ℓ)		550
NH_4 as N	(mg/ℓ)	20	4,7
NO_3 as N	(mg/ℓ)		3,5
Total P as P	(mg/ℓ)	7,5	1,4
PO_4 as P	(mg/ℓ)	2,7	0,7
pH		7,3	7,6
Construction Cost		R11 000 000	
Commissioning Date		1978	



1 SCREENING
2 GRIT REMOVAL
3 PRIMARY SEDIMENTATION
4 FLOW BALANCING
5 FLOW CONTROL
6 ANAEROBIC ZONE
7 FIRST ANOXIC ZONE
8 MAIN AERATION ZONE
9 SECOND ANOXIC ZONE
10 REAERATION ZONE
11 MIXED LIQUOR RECYCLE
12 WASTE ACTIVATED SLUDGE
13 CLARIFICATION AND SLUDGE THICKENING
14 ACTIVATED SLUDGE RETURN
15 EFFLUENT

FIGURE 12.1: Goudkoppies Purification Works – Plant layout

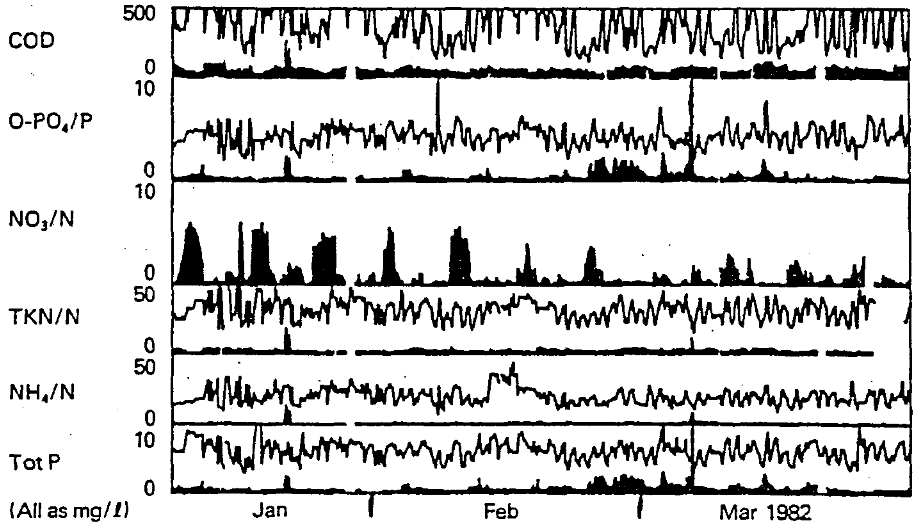


FIGURE 12.2 Performance (Based on 4-hourly sampling)
= Feed = Effluent

weekends and holiday periods. Phosphate removal was very dependent on denitrification. Feed sewages having high TKN/COD ratios (0,15) and/or low COD concentrations (300 mg/ℓ) produced effluents containing relatively high nitrate levels (i.e. 5-10 mg N/ℓ). This was reflected in high return sludge nitrates (4-7 mg N/ℓ) from the clarifiers and excessively high nitrate loads on the anaerobic zone with the consequent loss of the desired anaerobic conditions. Phosphate release into solution in the anaerobic zone was found to be necessary for good phosphate removal with orthophosphate concentrations in the range 15-25 mg P/ℓ being ideal. The recycle of nitrate to the anaerobic zone tended to inhibit this phosphate release. For this reason it was essential to maintain an effluent with a low nitrate concentration (less than 2 mg N/ℓ). When adequate denitrification could not be achieved the operation was changed to limit the oxygen input to the main aerobic zone. This action suppressed nitrification and ensured that the nitrate in the recycle remained below 2 mg N/ℓ but the ammonia concentration in the effluent increased to approximately 5 mg/ℓ. However, this mode of operation proved difficult to optimise due to the varying quality of the feed sewage.

When the higher strength sewage containing the yeast waste was treated the performance of the Goudkoppies Works improved dramatically. Denitrification was more efficient and effluent total phosphate concentrations were decreased from 7-8 mg/ℓ (as P) to below 1 mg/ℓ (as orthophosphate P) for 95 per cent of the time and often below 0,2 mg P/ℓ. Nutrient removal was far more stable and less erratic (see Figure 12.2), while the plant was easier to control. A mean effluent orthophosphate concentration of 0,36 mg P/ℓ and a total nitrogen (TKN + NO_3) level of 4,3 mg N/ℓ was obtained during this period.

Underloading and unfavourable characteristics* of the settled sewage also led to poor nutrient removal at the Northern Works. Both the effluent nitrate and phosphate concentrations exceeded 5 mg/ℓ and there was no phosphate release in the anaerobic zone due mainly to the large recycle of nitrates via the clarifier underflow and the low concentration of readily biodegradable COD. In an experiment however, when acid digested primary sludge was fed to the anaerobic zone of one module, to increase the readily biodegradable COD concentration effluent phosphate concentrations dropped to below 0,5 mg P/ℓ and nitrates to below 2 mg N/ℓ. The higher COD and more favourable characteristics of the feed to the plant imbued by the acid sludge, greatly improved nutrient removal.

Studies also showed that these plants (particularly Goudkoppies) were prone to the growth of filamentous micro-organisms which resulted in poor settling sludges and nuisance scums. Poor sludge settling created frequent operational problems, particularly when SVI values in excess of 300 mℓ/g were encountered. Sludge settling properties could change very rapidly, often for no obviously apparent reason, but once filamentous organisms predominated in the activated sludge they were extremely difficult to eradicate. Factors which appeared to encourage these nuisance growths were operation of modules with large unaerated mass fractions and the suppression of aerobic zone dissolved oxygen levels to below 1 mg/ℓ.

Microscopic observations during periods of good phosphate removal showed activated sludges containing large quantities of clustered bacteria which contained significant amounts of polyphosphate storage material. These bacteria had the appearance of *Acinetobacter spp* and seemed to be a dominant species in the sludges. Poor settling of activated sludges was found to be due to the prolific growth of filamentous organisms.

Application of the Marais-Ekama activated sludge model (described in previous chapters) to data from Johannesburg's nutrient removal works demonstrated that the model accurately predicated the COD and nitrogen removal performance of both the Goudkoppies and Northern Works as well as their oxygen consumption rates and mixed liquor suspended solids concentrations. The model also gave a good indication of why phosphate removal was not optimum and what measures would be necessary to improve performance. These predictions were verified when the addition of acid sludge at Northern Works and yeast waste at Goudkoppies greatly improved phosphate removal.

Another interesting experiment at Goudkoppies was the investigation of an on/off operating mode. This study was undertaken during the initial period when the plant was underloaded. Because of inadequate sewage flow and COD concentration, Module 3 was operated on a reduced feed of approximately 20 Mℓ/day, which represented about 25 per cent of its design COD load. Under these circumstances, phosphate removal was impossible with the normal continuous operating mode. However, when an on/off operating mode was adopted, phosphate removal improved. In this mode sewage flow to Module 3 was stopped completely between 07h00 Saturdays and 08h00 on Mondays and all mechanical plant was switched off for this period. Activated sludge settled to the floor of the reactor and went anaerobic releasing phosphate into solution. On Monday mornings feed to the module was gradually reinstated and when aerators became sufficiently immersed again, mechanical plant was switched on and the module run at its usual setting until the next weekend. Performance was characterised by a high effluent phosphate peak soon after start up on Mondays which dropped rapidly on Tuesdays to become well below 1 mg P/ℓ for the rest of the week. On Thursdays and Fridays only trace quantities of phosphate were found in the effluent and overall there was a net 50 per cent improvement in phosphate removal over previous operating modes.

An important finding during the operation of these plants was the inefficient performance of the second anoxic zone. On an average this zone removed approximately 1 mg N/ℓ of nitrate and contributed very little to the overall denitrification achieved in these plants.

The detailed study of the performance and optimisation of these plants was the topic of a Water Research Commission contract with the Johannesburg City Council. The work was undertaken in collaboration with the University of Cape Town and the National In-

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*High TKN/COD ratio of influent; low readily biodegradable COD.

stitute for Water Research. For further details of the Council's contribution to this project, the reader is referred to the detailed final report entitled "Optimisation of the Modified Activated Sludge Process for Nutrient Removal" which is available from the Water Research Commission, P O Box 824, Pretoria, 0001, Republic of South Africa. References to other publications produced during this study can also be found in this report.

3. CASE STUDY NO. II

BIOLOGICAL NUTRIENT REMOVAL AT THE DASPOORT SEWAGE WORKS PILOT PLANT, PRETORIA

The National Institute for Water Research of the Council for Scientific and Industrial Research investigated biological nutrient removal under simulated practical conditions over a period of five years.

For this purpose a 100 m³/d pilot plant, situated at the Daspoort Sewage Treatment Works in Pretoria, was operated in the Phoredox and UCT modes at sludge ages varying between 15 and 20 days. A multiple basin design was used, with the number of reactors in the primary anaerobic and anoxic stages varying from 2 to 3 and in the aerated zone between 3 and 5. When included, the secondary anoxic and aerated stages consisted of 4 and 1 basins respectively. Each basin had a nominal hydraulic retention time of approximately 1,2 h. The mixed liquor (a) and sludge (s) recycle ratios usually were maintained at values of 2:1 and 1:1 respectively for the Phoredox configurations but a number of mixed liquor return rates were used when studying the UCT mode. In the Phoredox configuration the mixed liquor recycle ratio of 2:1 was selected after several pilot plant tests indicated this ratio was near optimal for the experimental conditions and plant layout used.

Settled sewage was fed during the study, with the exception of a six month period during which screened raw sewage was used in an effort to determine the effect of increased influent COD concentrations on phosphate removal efficiency. Median values for feed COD (as mg/*ℓ*), total phosphate (as mg P/*ℓ*) and the TKN/COD ratio (mg N/mg O) during the five-year investigation were respectively 320, 6,2 and 0,10 while the central 80 per cent of the individual values lay in the respective ranges 210 to 410, 4,2 to 8,6 and 0,14 to 0,08.

Several additional studies were performed during this period apart from evaluating the long term behaviour of the Phoredox and UCT configurations *per se*. These included the effect of variable a-recycle rates on process performance, the elimination of peak effluent phosphate concentration by the addition of relatively high strength sewage collected beforehand, the influence of elevated DO concentrations in the s-recycle and the influence of low DO concentrations in the aerated stage.

The median effluent concentrations observed for the five-stage Phoredox process during a 14-month investigation between March 1977 to May 1978 were 22; 0,8; 0,02 and 5,3 mg/*ℓ* for COD, orthophosphate (as P), ammonia (as N) and nitrate (as N) respectively. The central 80 per cent of all observations lay in the respective ranges 14 to 37; 0,1 to 3,1; 0,0 to 0,2 and 2,3 to 8,0 mg/*ℓ*. These results were achieved at mean influent COD and total phosphate (as P) values of 277 and 5,5 mg/*ℓ* respectively, and a TKN/COD ratio of 0,11. During the period October 1978 to February 1979, when the mean influent total phosphate concentration was 6,0 mg/*ℓ* and the COD and TKN/COD values slightly more favourable for enhanced phosphate accumulation, namely 307 mg/*ℓ* and 0,10 respectively, the process produced effluents of which the median COD, orthophosphate, ammonia and nitate concentrations were 20; 0,4; 0,01 and 3,8 mg/*ℓ* respectively. The central 80 per cent of all observations during this phase of the study lay in the respective ranges 18 to 23; 0,1 to 2,5; 0,0 to 0,15 and 1,7 to 6,7 mg/*ℓ*. When the mean influent COD value was artifically increased to 330 mg/*ℓ*, by collecting peak strength sewage on Fridays and blending this product in a 1:1 ratio with the relatively low strength feed normal for weekends, the plant produced median effluent COD, orthophosphate, ammonia and nitrate concentrations of 22; 0,4; 0,1 and 5,9 mg/*ℓ*. This was achieved despite a relatively high mean influent total phosphate concentration of 7,2 mg/*ℓ* and an unfavourable TKN/COD ratio of 0,11. The corresponding ranges in which the central 80% of the aforementioned parameters lay were 18 to 25; 0,1 to 1,5; 0,0 to 0,1 and 3,5 to 9,5 mg/*ℓ*. While not necessarily considered to be a practical procedure the blend resulted in a sharp improvement in effluent and an almost complete elimination of the peak values characteristically occurring on Sundays or Mondays.

The phosphate removal capability of the pilot plant deteriorated upon changing the feed from settled to screened raw sewage, despite relatively favourable mean influent COD and total phosphate values of 369 and 5,3 mg/*ℓ* respectively and a favourable TKN/COD ratio of 0,09. During the period May to October 1978, when this mode of operation was evaluated, the median effluent concentrations were 23; 1,3; 0,03 and 4,0 mg/*ℓ* for COD, orthophosphate, ammonia and nitrate respectively, with the central 80 per cent of all values in the corresponding ranges 16 to 30; 0,3 to 3,2; 0,0 to 0,4; and 0,4 to 6,9 mg/*ℓ*. The relatively poor performance of the plant with respect to phosphate removal during this period led to the conclusion that primary sedimentation had a beneficial effect on the subsequent phosphate uptake capabilities of Phoredox plants under similar experimental conditions to those of this study.

The UCT process was evaluated during three distinct operational periods. The best of these was that from November 1980 to December 1981, during which the mean influent COD and total phosphate values were

324 and 7,5 mg/ℓ respectively and the mean TKN/COD ratio 0,10. The median effluent concentrations for COD, orthophosphate, ammonia and nitrate were 26; 0,8; 0,14 and 8,0 mg/ℓ respectively, with the central 80 per cent of all observations in the corresponding ranges 19 to 35; 0,1 to 3,1; 0,0 to 0,7; and 4,0 to 12,0 mg/ℓ. This best performance of the UCT process was achieved only after the anaerobic stage was extended from two to three basins.

The study showed that the processes of carbon oxidation, nitrification and denitrification should not present the designer with significant problems in establishing an appropriate nutrient removal plant. Such a facility should prove capable of producing a virtually completely nitrified effluent with average COD and total nitrogen values in the respective ranges 20 to 30 and 4 to 7 mg/ℓ for situations with wastewater and plant characteristics similar to or more favourable than those pertaining to the Daspoort plant (refer to Chapter 7). Effluent phosphate concentrations from the Phoredox process can be expected to exhibit peak values in sympathy with reduced COD input over weekends but otherwise should be sufficiently low to achieve overall average levels in the range 0,5 to 1,0 mg/ℓ (as P) when settled sewage is fed to the plant. It is noteworthy that this degree of phosphate removal can be achieved in the Phoredox process despite average nitrate concentrations around 5 mg/ℓ (as N) being returned to the anaerobic stage. Nonetheless, it should be remembered that effluent concentration is a function of influent phosphate concentration and achievable removal, and this should be considered for each sewage and process. The UCT process is highly successful in suppressing effluent peak phosphate values but overall is not expected to prove advantageous over the Phoredox process for conditions similar to those under which this case study was performed. Detailed results pertaining to the performance of the pilot plant under various modes of operation may be found in the reports by Simpkins and Gerber (1981) and Gerber and co-workers (1982).

4. REFERENCES

BARNARD, J.L. (1982) Design consideration regarding phosphate removal in activated sludge plants. IAWPR Post Conference Seminar on Phosphate Removal in Biological Treatment Process, CSIR Conference Centre, Pretoria, 5-6 April 1982.

GERBER, A., SIMPKINS, M.J., WINTER, C.T. and SCHEEPERS, J.A. (1982) Biological nutrient removal from wastewater effluents: Performance evaluation of the Phoredox and UCT processes. Contract report C WAT 52 prepared for the Water Research Commission by the National Institute for Water Research (Obtainable from the Water Research Commission, P O Box 824, Pretoria, Republic of South Africa 0001)

KERDACHI, D.A. and ROBERTS, M.R. (1982) Full scale phosphate removal experiences in the Umhlatuzana works at different sludge ages. IAWPR Post Conference Seminar on Phosphate Removal in Biological Treatment Processes, CSIR Conference Centre, Pretoria, 5-6 April 1982.

PAEPCKE, B.H. (1982) Performance and operational aspects of biological phosphate removal plants in South Africa. IAWPR Post Conference Seminar on Phosphate Removal in Biological Treatment processes, CSIR Conference Centre, Pretoria, 5-6 April 1982.

PITMAN, A.R., VENTER, S.L.V. and NICHOLLS, H.A. (1982) Practical experience with biological phosphorus removal plants in Johannesburg. IAWPR Post Conference Seminar on Phosphate Removal in Biological Treatment Processes, CSIR Conference Centre, Pretoria, 5-6 April 1982.

PITMAN A.R. (1983) Optimisation of the Modified Activated Sludge Process for Nutrient Removal. Joint publication by the City Council of Johannesburg and the Water Research Commission. (Obtainable from th Water Research Commission, P O Box 824, Pretoria, Republic of South Africa 0001).

SIMPKINS, M.J. and GERBER, A. (1981) Technical guide on biological nutrient removal: The Phoredox process. Confidential report prepared for the Water Research Commission by the National Institute for Water Research, 1981. (Obtainable by arrangement with Messrs SAIDCOR, P O Box 395, Pretoria, Republic of South Africa 0001).

APPENDIX 1

LIST OF SYMBOLS

a = mixed liquor recycle ratio from the aerobic to the anoxic reactors.
a_o Subscript o denotes optimum.

A = surface area of secondary settling tank

b_{hT} = endogenous mass loss rate for heterotrophic organisms at T°C (/d)
= $b_{h20}(1,029)^{(T-20)}$

b_{nT} = endogenous respiration rate for nitrifying bacteria at T°C (/d)
= $b_{n20}(1,029)^{(T-20)}$

b_{h20} = the rate at 20°C = 0,24/d

b_{n20} = the rate at 20°C = 0,04/d

D_p = denitrification potential (mgN/ℓ influent)

D_{p1}, D_{p3} Subscripts 1 and 3 refer respectively to the primary and secondary anoxic reactors

D_{pd1}, D_{pd2} Subscripts d1 and d2 refer respectively to the first and second anoxic reactor of a Modified UCT process

D_{pp} = denitrification potential of the process when the maximum anoxic sludge mass fraction is all in the form of a primary anoxic reactor

f = unbiodegradable fraction of active mass = 0,20 mgVSS/mgVSS

f_{av}, f_{at} = active fraction of the sludge mass with respect to the volatile and total solids concentrations respectively

f_{bs} = readily biodegradable COD fraction of the influent with respect to the biodegradable COD concentration

f_{cv} = COD to VSS ratio of the volatile sludge mass = 1,48 mgCOD/mgVSS

f_i = MLVSS to MLSS concentration ratio of the mixed liquor

f_n = nitrogen fraction of the MLVSS (mgN/mgVSS) = 0,10 mgN/mgVSS

f_{na} = amonia fraction of the influent TKN

f_{nu} = soluble unbiodegradable organic nitrogen fraction of the influent TKN

f_{ns} = TKN/COD concentration ratio of the influent

f_p = phosphorus fraction of the inert MLVSS and endogenous residue MLVSS = 0,015 mgP/mgVSS

f_{rps} = fraction of raw total COD removed by primary sedimentation

f_u = unbiodegradable COD fractions in the influent (mgCOD/mgCOD). Additional subscripts p and s refer respectively to particulate and soluble fractions

f_v* = general parameter for volume fractions

f_x* = general parameter for sludge mass fractions
*Additional subscripts a, b, d, 1, 3 and m refer respectively to anaerobic, total aerobic, total anoxic, primary anoxic, secondary anoxic and maximum unaerated allowable sludge mass fraction

f_{xdt}, f_{xdm} Additional subscripts t and m following the d subscript refer respectively to the total and maximum allowable sludge mass fractions

f_{x1min}, f_{x3min} Subscript min following subscripts 1 and 3 refer to the minimum primary and secondary anoxic sludge mass fractions

f_{xd1}, f_{xd2} Subscripts 1 and 2 following the d subscript refers respectively to the first and second anoxic reactors of a Modified UCT process

F/M = Food to Microorganism ratio

G = General parameter denoting flux (kg/m^3/d). Subscripts ap, b, l and s refer respectively to applied flux, bulk flux, limiting flux and gravity settling flux

K, K_1, K_2, K_3 = general parameter for denitrification rate (mgNO$_3$/mgVASS/d). Subscripts 1 and 2 refer respectively to the 1st and 2nd rates in the primary anoxic and 3 to the rate in the secondary anoxic. Additional subscripts T and 20 refer to T°C and 20°C respectively

K_r = conversion rate of biodegradable organic nitrogen to free and saline ammonia by active mass = 0,015 ℓ/mgVASS/d. Subscripts T and 20 refer to rates at T°C and 20°C

K_n = half saturation coefficient in Monod equation for nitrification = mgN/ℓ.
Subscripts T and 20 refer respectively to T°C and 20°C.
Subscripts pH and 7,2 refer respectively to K_n at different pH values and that at pH 7,2

LF = load factor

M = prefix denoting mass as opposed to concentration of a variable

N = general parameter denoting nitrogen concentration (mgN/ℓ)

N_a, N_n, N_o, N_t, N_u = Subscripts a, n, o, t and u refer respectively to ammonia, nitrate, biodegradable organic nitrogen, total TKN and soluble unbiodegradable organic nitrogen concentrations. Additional subscripts e, i, r, s and a refer respectively to the concentrations in the effluent, influent, r-, s- and a-recycle flows

N_c = nitrification capacity (mgN/ℓ influent)

N_{pi} = unbiodegradable particulate organic nitrogen concentration in influent (mgN/ℓ)

N_s = nitrogen required for sludge production (mgN/ℓ influent)

N'_{ti} = nitrogen available for nitrification
= $N_{ti}-N_s$ (mgN/ℓ influent)

O = general parameter for oxygen

O_c, O_n, O_d, O_t Subscripts c, n, d and t refer respectively to the oxygen demands for carbonaceous material degradation, nitrification, that recovered by denitrification and total oxygen demand

O_i, O_a, O_s Subscripts i, a and s refer respectively to the dissolved oxygen concentrations in the influent a- and s-recycles

OUR = oxygen utilization rate (mgO/ℓ/h)

P_f = excess P removal propensity factor (mgCOD/ℓ)

P_s = phosphorus in daily sludge wastage per ℓ influent flow (mgP/ℓ) i.e. the phosphorus removal from the wastewater

P_t = total phosphorus concentration (mgP/ℓ).
Additional subscripts i and e refer respectively to influent and effluent

q = flow rate for waste sludge from the process reactor (ℓ/d)

Q = daily mean influent flow rate (ℓ/d)

r = mixed liquor recycle ratio from the anoxic to the anaerobic reactor in the UCT type processes

R_{hn} = process total average nominal hydraulic retention time (d)

R_a, R_n = general parameters for actual and nominal hydraulic retention times. Subscripts p and s refer to primary and secondary anoxic reactors

R_s = sludge age (d)

R_{sm} = minimum sludge age for nitrification

s = underflow recycle ratio

S_f = factor of safety with respect to nitrification

S_{fd} = factor of safety with respect to conversion of a UCT process to a Modified UCT process

S = general parameter denoting COD concentration

S_b, S_u, S_t Subscripts b, u and t refer respectively to biodegradable, unbiodegradable and total COD concentrations

S_{bs}, S_{up}
S_{bi}, S_{bsi} Additional subscripts i, e, s and p refer respectively to concentrations in the influent and effluent, and readily biodegradable and particulate COD

S_{bsa} = readily biodegradable COD concentration in the anaerobic reactor

S_{COD} = substrate concentration with respect to COD

S_{BOD} = substrate concentration with respect to BOD

SUR = substrate utilization rate
Subscripts a, v and t refer respectively to the rate with respect to active, volatile and total sludge concentrations
Additional subscripts COD and BOD refer to the rate with respect to the substrate measurement in terms of COD or BOD

t = time

t_1 = duration of first phase of denitrification in the primary anoxic reactor

U_o, U_u = overflow and underflow velocity in the secondary settling tank (m/d)

v = volume of waste sludge abstracted from process reactor per day

V = general parameter denoting volume
Subscripts p and r refer respectively to the total process and reactor

V_s = gravity settling velocity of sludge (m/d)
Additional subscripts denote velocity at specified sludge concentrations

X = general parameter denoting sludge mass concentration
Subscripts a, e, i, v, t and n refer respectively to active, endogenous, inert, volatile, total and nitrifier sludge concentrations
Additional subscripts f and i, and a, d and b refer respectively to concentrations in effluent and influent and those in the anaerobic, anoxic and aerobic reactors

$\overline{X}_t$ = process average total sludge concentration (mgTSS/ℓ)

X = for settling tank theory denotes total MLSS sludge concentration (kgMLSS/m^3)
Subscripts af, dz, ef, l, o and r refer respectively to the sludge concentration of the layer above the feed point, the dilute zone below the feed point, the effluent, the limiting zone, in the biological reactor and the underflow sludge return

Y_h = heterotrophic organism yield coefficient = 0,45 mgVSS/mgCOD

Y_n = nitrifier organism yield coefficient = 0,10 mgVSS/mgN

α = denitrification attributable to the readily biodegradable COD (mgNO_3-N/mg biodegradable influent COD)

γ = coefficient of excess phosphorus removal (mgP/mgVASS) i.e. the proportion of phosphorus in the active mass

Δ = prefix denoting the change in the parameter following

ϕ_{ns} = pH sensitivity parameter for nitrification in the simplified Arrhenius equation = 2,35

μ_n = specific growth rate of the nitrifiers (/d)
Subscript m denotes the maximum rate
Additional subscripts T and 20 refer respectively to the rate at T°C and 20°C
Additional subscripts pH and 7,2 refer respectively to the rate at different pH values and at pH 7,2

2,86 = oxygen equivalent of nitrate i.e. 2,86 mg oxygen can accept as many electrons as 1 mgNO_3-N nitrate

8,6 = mg mass of COD utilized per mgNO_3-N nitrate denitrified

4,57 = mg mass of oxygen required for nitrifying 1 mgN ammonia to 1 mgN nitrate

= decimal place

= multiplication symbol

APPENDIX 2

BRIEF DESCRIPTION OF THE DETERMINATION OF THE READILY BIODEGRADABLE COD FRACTION OF A MUNICIPAL WASTEWATER

1. INTRODUCTION

In the experimental investigation into the dynamic behaviour of the activated sludge process (Dold, Ekama and Marais, 1980), it was found that in the aerobic completely mixed process at short sludge ages (1,5 to 3 days) under daily cyclic square wave loading conditions (feed 12 hours on, 12 hours off), there is a precipitous decrease in oxygen consumption rate immediately after the termination of the feed period (see Figure A2.1). In modelling the activated sludge process, it was found that this behaviour was due to the readily biodegradable COD in the influent; at feed termination, the addition of readily biodegradable COD stops so that the utilization of the readily biodegradable COD stops, causing a step change in the oxygen consumption rate. Hence a method for determining the readily biodegradable COD in the influent is the measurement of the step change in oxygen consumption rate at feed termination in a daily cyclic square wave loaded completely mixed process operated at a very short sludge age.

2. APPARATUS

The test is performed using a single reactor completely mixed aerobic activated sludge unit operated at a sludge age of about 1,5 to 3 days depending on the temperature; the higher the temperature the shorter the sludge age. The data shown in Figure A2.1 are from a 2,5 day sludge age unit at 20°C. A schematic representation of the process is shown in Figure A2.2.

The following points warrant notice regarding the setting up and operation of the process:

(1) *Aeration.* The most convenient method of supplying aeration is to use an ordinary fish tank air pump with a fine bubble sparger stone attached to a perspex tube. By adjusting the depth of immersion of the sparger the DO level can be controlled to the desired value of about 2 mgO/*ℓ*.

(2) *Inlet and outlet from aeration reactor.* All streams to and from the reactor should enter the reactor below the liquid surface and not be dripped into the reactor; dripping feed into the reactor may result in incorrect readings by exposing the feed to air and allowing DO entrainment via the liquid/air interface.

(3) *Mixing.* The contents of the feed tank should be stirred slowly and gently in order to keep particulate matter in suspension, but to avoid excessive DO entrainment into the reactor contents. Covering the reactor contents by floating on it a Sagex disc is a very effective method for stopping DO entrainment into the reactor caused by mixing.

(4) *Sludge wastage.* In order to maintain the correct sludge age, the correct volume of mixed liquor must be withdrawn from the reactor. Say the reactor volume is 10 *ℓ* and the sludge age is to be 2,5 days, then 10/2,5 = 4,0 *ℓ* of mixed liquor needs to be wasted daily. Because this volume constitutes a large proportion of the volume of the reactor, (this would be the case up to about 10 days sludge age) the sludge is best wasted with the aid of a pump which is activated by a timer. The wastage arrangement should be such that about 0,3 to 0,5 *ℓ* is abstracted over a few minutes and repeated every few hours. The high rate of abstraction is necessary to avoid settlement of the sludge from the abstraction tube back into the reactor; if this occurs the correct sludge age would not be established. In setting up the sludge wastage system, it is recommended that about 0,5 *ℓ* less than the required volume is withdrawn so that the exact volume can be made up by hand. Also note that if sludge is abstracted when the influent feed is stopped, the reactor volume will be reduced. Consequently it is best not to waste sludge during the non-feed period.

3. METHOD

3.1 Unit start-up

From start up, operate the unit under constant flow and load conditions for at least 3 sludge ages allowing the unit to reach steady state – steady state is achieved when the daily oxygen consumption rate and reactor MLVSS concentration show approximately steady values. When steady state is achieved switch the unit over to daily cyclic square wave loading conditions by discharging the daily volume of feed over say 12 hours instead of 24. (Keep the feed volume and COD concentration constant; simply increase the flow rate). Operate the unit under square wave conditions for about 2 sludge ages. Note that now the oxygen consumption rate will vary over the day but the MLVSS concentration should not change significantly.

3.2 Readily biodegradable COD measurement

(1) At some time at least 6 hours after commencing the feed, measure the oxygen utilization rate (OUR)

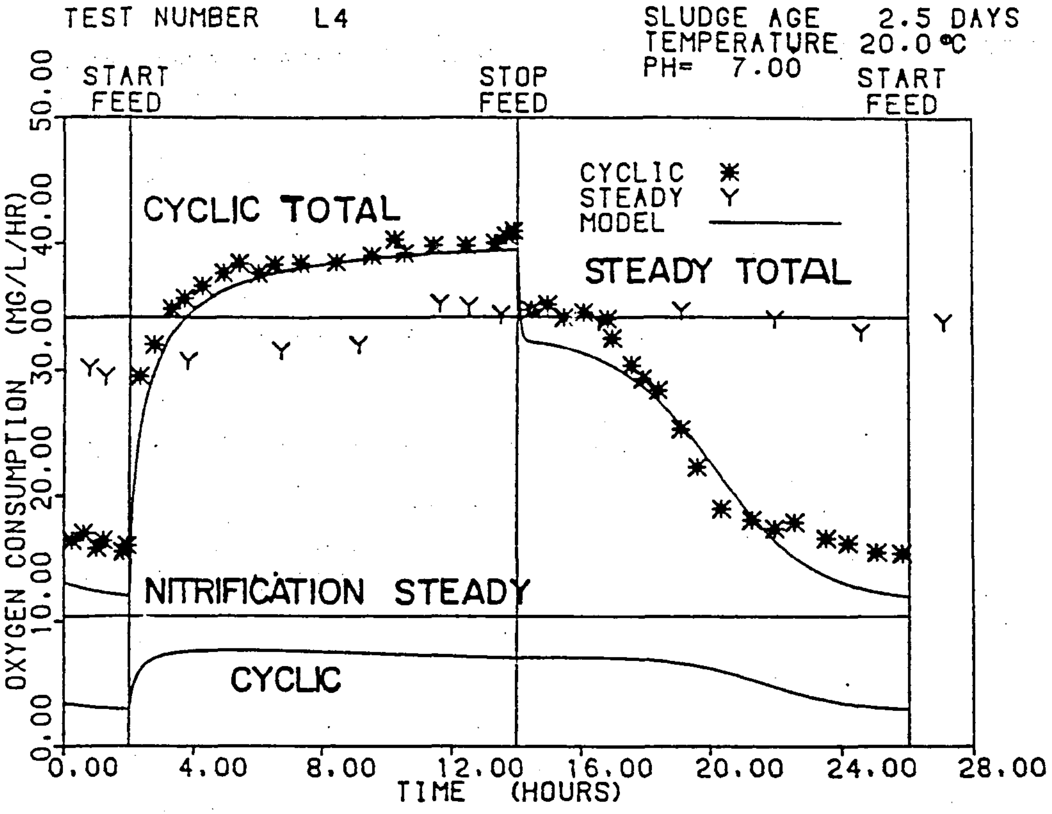

FIGURE A2.1: Experimental (and theoretical) oxygen consumption rate over a 26 hour period in a single completely mixed aerobic activated sludge process at 2,5 days sludge age and 20°C under daily cyclic square wave loading conditions (12 hours on, 2 to 14 hours and 12 hours off, 14 to 26 hours). Note precipitous decrease in the measured rate at feed termination (14h00) but that the oxygen consumption rate for nitrification does not change (theoretical line). Experimental (and theoretical) data for the equivalent steady state process (i.e. receiving the same feed volume of 24 hours) are also shown (from Dold, Ekama and Marais 1980).

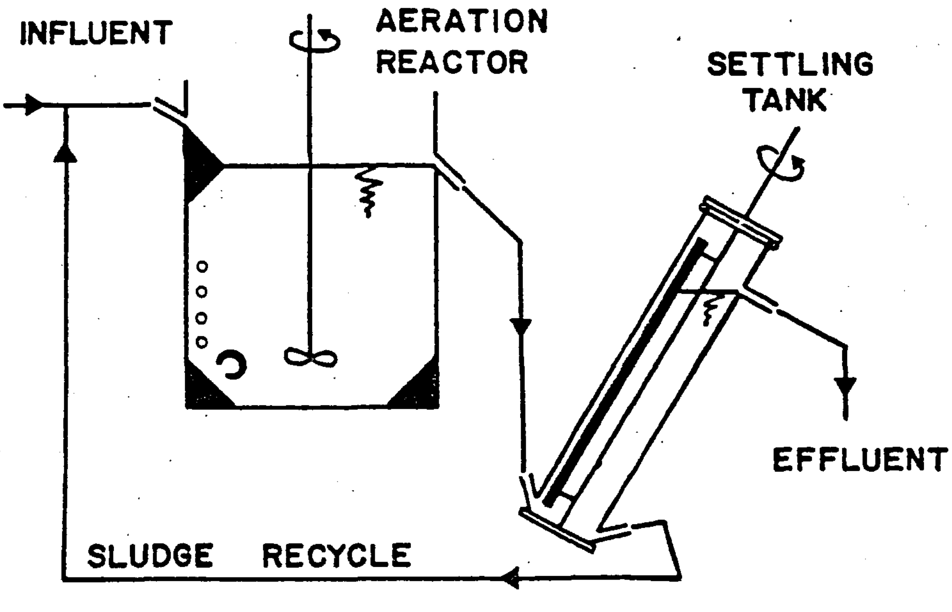

FIGURE A2.2: Schematic layout of process configuration.

by raising the DO to about 6 mgO/ℓ, switching off the air supply, and monitoring the rate of decrease of DO on a recorder. Measure the OUR every 20 to 30 minutes for about 2 hours before the feed is terminated.

(2) Switch off the *feed* to the reactor and again raise the DO to about 6 mgO/ℓ. About 1 to 2 minutes after the feed has been stopped again measure the OUR. Measure the OUR every 20 to 30 minutes for about 2 hours after the feed has been terminated.

(3) Restart the feed and run plant for about 4 hours before repeating the measurement to allow the OUR to stabilize to the feeding condition again.

(4) Plot OUR for each period of about 2 hours before and after the feed was terminated. Estimate the average OUR before and after the feed termination.

The following information is required to calculate the concentration of readily biodegradable COD in the influent feed:

Q = feed flow rate (ℓ/d) over the 3 hour period before the feed is terminated

V_p = reactor volume (ℓ)

OUR_b = average OUR before feed termination (mgO/ℓ/h)

OUR_a = average OUR after feed termination (mgO/ℓ/h)

S_{ti} = total influent COD concentration of feed

$\Delta OUR = (OUR_b - OUR_a)$ (mgO/ℓ/h)

Then the influent readily biodegradable COD concentration

$S_{bsi} = (\Delta OUR . V_p . 24)/(Q . 0{,}334)$

where

24 = number of hours per day

0,334 = general conversion factor for COD to oxygen. This factor states that for every unit of COD utilized by the micro-organisms, 0,334 units of COD are oxygen consumption and the remaining 0,666 units of COD are converted to new organism mass (see Chapter 1).

The influent readily biodegradable COD fraction with respect to the total COD (f_{ts}) is given by

$$f_{ts} = S_{bsi}/S_{ti}$$

and the influent readily biodegradable COD fraction with respect to the biodegradable COD (f_{bs}) is given by

$$f_{bs} = S_{bsi}/S_{bi}$$

$$= S_{bsi}/S_{ti}(1 - f_{us} - f_{up}).$$

3.3 Example calculation

The following data were obtained from an experimental unit operated on primary settling tank effluent at the Goudkoppies plant

Q = 50 ℓ/d for period prior to feed termination

V_p = 7,5 ℓ

OUR_b = 24,1 mgO/ℓ/h

OUR_a = 20,1 mgO/ℓ/h

S_{ti} = 370 mgCOD/ℓ

$\Delta OUR = 24{,}1 - 20{,}1 = 4{,}0$ mgO/ℓ/h

$S_{bsi} = (4{,}0 . 7{,}5 . 24)/(50 . 0{,}334)$

= 47,5 mgCOD/ℓ

S_{bs}/S_{ti} = 47,5/370 = 0,13 mgCOD/mgCOD

$f_{bs} = 47{,}5/[370(1 - 0{,}09 - 0{,}04)]$

= 0,15 mgCOD/mgCOD.

4. NOTES

(1) *Suggested loading:* The COD load on the reactor should be such that the OUR (excluding nitrification which may or may not occur at the short sludge age depending on μ_{nm}) is greater than 25 mgO/ℓ/h. This can be achieved if the feed rate, COD feed concentration and reactor volume are such that the load is greater than 2 400 mgCOD/ℓ reactor/day. Thus, if the feed concentration is about 350 mgCOD/ℓ and a reactor volume of 10ℓ is used the feed flow rate Q prior to feed termination should be greater than 2400.10/350 = 68,5 ℓ/d.

(2) *Nitrification:* The degree of nitrification which takes place in the unit does not affect the readily biodegradable COD determination (see Figure A2.1). This is because at the short sludge ages at which the determination is done, the oxygen utilization for nitrification is the same before and after feed termination due to the high ammonia concentration in the reactor and effluent. If the ammonia concentration in the reactor and effluent before and after feed termination is low (<5 mgN/ℓ), the change in the oxygen utilization rate for nitrification may lead to errors in the readily biodegradable COD determination.

5. REFERENCES AND SUGGESTED READING

DOLD, P.L., EKAMA, G.A. and MARAIS, G.v.R. (1980) A general model for the activated sludge process. *Prog. Wat. Tech.* 12 47-77. An understanding of the paper is strongly recommended as background information.

EKAMA, G.A. and MARAIS, G.v.R. (1979). Letter to the Editor. *Water SA* 5(1) 57-60. A detailed discussion on the method and possible errors that can be incurred in measuring the oxygen consumption rate.

MARAIS, G.v.R. and EKAMA, G.A. (1976) The activated sludge process, Part I – Steady state behaviour. *Water SA* 2(4) 163-200. This contains further detail on the operation of the experimental units.

APPENDIX 3

BRIEF DESCRIPTION OF THE DETERMINATION OF THE MAXIMUM SPECIFIC GROWTH RATE OF THE NITRIFIERS

1. INTRODUCTION

The maximum specific growth rate constant of the nitrifiers (μ_{nm}) defines the maximum rate at which the nitrifying organisms can grow in the activated sludge process. Although μ_{nm} is a kinetic "constant" in the nitrification theory, its magnitude can differ appreciably between different municipal wastewaters, and can vary even between batches of the same wastewater. This behaviour is so marked that μ_{nm} should be classified as a wastewater characteristic.

The magnitudes of μ_{nm} that have been observed in municipal wastewaters at 20°C range from 0,30 to 0,65 per day. The low values appear to be due to inhibition by some substance(s) in the wastewater. These substances are more likely to be present in wastewaters having significant industrial* contributions; in general, the greater the contribution, the lower the μ_{nm} value. The reduction in μ_{nm} is not the result of toxicity because a high efficiency of nitrification can be achieved provided the sludge age is sufficiently long – toxicity would tend to terminate nitrification irrespective of sludge age.

The magnitude of μ_{nm} has a significant effect on the minimum sludge age for nitrification R_{sm}; the lower μ_{nm}, the longer R_{sm}. Clearly, due to the link between the wasteflow composition and μ_{nm}, an experimental estimate of the μ_{nm} value is important for efficient design of nitrifying plants. A method for estimating μ_{nm} is set out below.

2. METHOD AND APPARATUS

2.1 General

The test is performed in a single completely mixed reactor at about 6 to 10 days sludge age receiving a constant flow and load but with alternating cycles of anoxic and aerobic periods of 2 to 3 hours each. The selection of the sludge age depends on two factors, (i) the temperature of the wastewater and (ii) whether or not the μ_{nm} value is expected to be high (0,5/d at 20°C) or low (0,2/d at 20°C). The higher the temperature the shorter the sludge age, and the lower the expected μ_{nm} value the longer the sludge age.

The minimum sludge age for nitrification R_{sm} at T°C is given by (see Chapter 5, Section 4.4)

$$R_{sm} = S_f/\{\mu_{nmT}(1-f_t) - b_{nT}\} \qquad \text{(A3.1)}$$

where

μ_{nmT} = maximum specific growth rate at T°C

$$= \mu_{nm20}(1,123)^{(T-20)} \qquad \text{(A3.2)}$$

μ_{nm20} = rate at 20°C

b_{nT} = endogenous respiration rate of the nitrifiers at T°C

$$= 0,04(1,029)^{(T-20)} \qquad \text{(A3.3)}$$

T = temperature in deg C.

S_f = factor of safety to ensure complete nitrification, usually 1,25.

f_t = fraction of the total time that the process is under anoxic conditions.

With the aid of Eqs. A3.1 to A3.3, the sludge age required for nitrification in the process can be estimated for an expected μ_{nm20} value, and generally for μ_{nm20} around 0,40/d will be found to be between 6 to 10 days at 20°C for a 2-hourly cycle of anoxic and aerobic periods (f_t = 0,50).

The length of the anoxic and aerobic cycles is chosen such that the concentration of ammonia during the aerobic period *does not decrease below 2 mgN/ℓ.* The reason for this is that if the ammonia concentration is greater than 2 mgN/ℓ, the nitrification rate proceeds at the maximum possible and production rate of nitrate can be directly linked to μ_{nm}; if the ammonia concentration is below 2 mgN/ℓ, the nitrate production rate no longer can be directly linked to μ_{nm} because now the limited supply of ammonia reduces the nitrification rate even though the μ_{nm} value remains unchanged (see Figure 5.1 in Chapter 5). If it is found that the ammonia concentration reduces to below 2 mgN/ℓ in too short a time interval, the sludge age should be reduced. This has the effect of reducing the concentration of nitrifying organisms, with the result that less ammonia can be nitrified per unit time.

The experimental data shown in Figure A3.1 were measured at 6 days sludge age and 20°C with an anoxic and aerobic cycle of 2 hours each. The data are tabulated in Table A3.1

A schematic representation of the experimental process and some points regarding its operation are given in Appendix 2.

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*The type of industry referred to here are those which produce effluents which have high metal ion concentrations e.g. metal finishing industry, plating works, production, tannery wastes, etc.

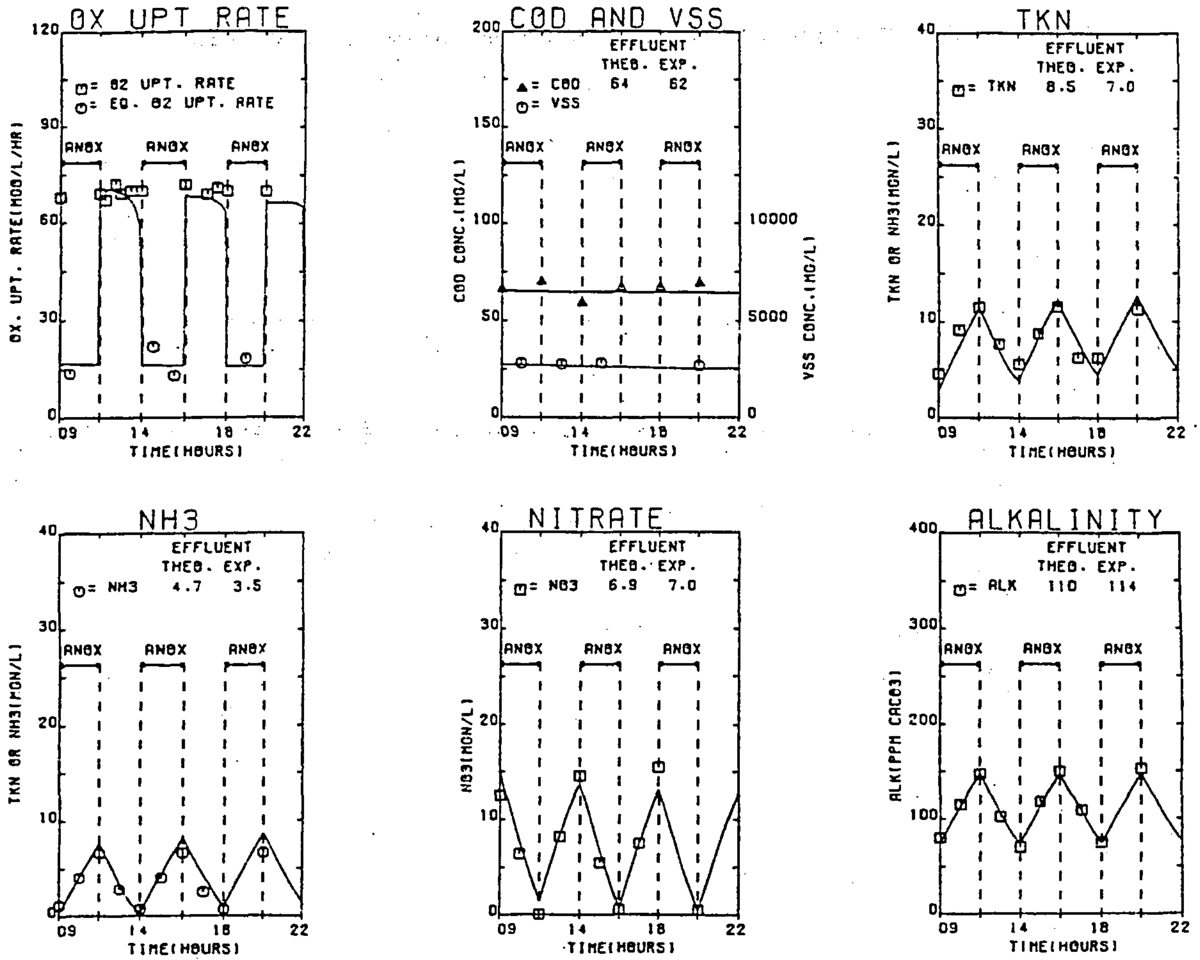

FIGURE A3.1: Experimental (and theoretical) data of (a) the oxygen consumption rate (b) reactor MLVSS concentration and filtered COD concentration and filtered reactor, (c) TKN, (d) ammonia, (e) nitrate and (f) Alkalinity concentrations versus time in a 2 hourly alternating anoxic and aerobic process under constant flow and load conditions at 20°C and 6 days sludge age. (From van Haandel and Marais 1981).

2.2 Experimental unit start-up

From start-up, operate the unit under constant flow and load conditions at the selected sludge age under *purely aerobic conditions* for about 2 to 3 sludge ages by which time nitrification should be taking place.* If necessary operate for 1 or 2 sludge ages longer to ensure nitrification is complete and the nitrifiers are well established in the process. Only when nitrification is complete should the process be switched over to the alternating aerobic-anoxic phase, say two hours aerobic, two hours anoxic. Even for very low values of μ_{nm} (0,20 at 20°C), it is unlikely that nitrification will not take place at sludge ages around 10 days at 20°C with 50% anoxic times. Operate the process under alternating conditions for not less than one sludge age to allow the system to stabilize to the alternating conditions.

During the stabilization period, measure on a daily basis the following –

(i) influent COD and TKN concentration
(ii) reactor MLVSS concentration and pH
(iii) effluent** TKN, ammonia, nitrate concentrations
(iv) effluent** COD concentration.

When these data show approximate stability over a period of about one sludge age, the system can be considered to be stable and the analyses for the μ_{nm} determination can be commenced.

2.3 Analysis requirement for μ_{nm} measurement

Once the process is operating satisfactorily, an intensive testing programme of about 14 to 16 hours is performed on it. The data measured during such a programme are the folowing:

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*A simple qualitative test to check the presence of nitrate is the Merckoquant® paper test, or to leave a 100 mℓ (or larger) measuring cylinder with mixed liquor standing overnight; if upon tapping the cylinder, bubbles are released from the sludge layer, or some sludge has risen to the surface, nitrification is taking place – evidenced by the nitrogen bubbles released by denitrification of the nitrate in the sludge layer.

**Grab samples collected from the effluent bucket after a full 24 hour period.

TABLE A3.1 EXPERIMENTAL DATA MEASURED IN A 2-HOURLY ALTERNATING ANOXIC-AEROBIC PROCESS AT 6 DAYS SLUDGE AGE AND 20°C

Process Parameters
Reactor Volume = 6 ℓ
Influent flow = 20 ℓ/d
Sludge age = 6 days
Temperature = 20°C
Recycle ratio = 1:1

Day's average	COD	TKN	NH_3	Alk	NO_3	pH
Influent (unfilt.)	492	49,0	33,5	245	–	7,8
Effluent (unfilt.)	62	7,1	3,5	114	7,1	7,5
Effluent (filt.)	54	6,9	3,5	114	5,8	–

TIME	OCR	VSS	COD	TKN	NH_3	Alk	NO_3	pH
9h30·	69	–	66	4,6	1,1	80	12,5	7,1
10h30	–	2822	–	9,2	4,0	115	6,4	7,2
11h30*	69	–	70	11,5	6,6	147	0,0	7,4
11h45	67	–	–	–	–	–	–	–
12h15	72	–	–	–	–	–	–	–
12h30	69	2773	–	7,7	2,8	102	8,2	7,2
13h00	70	–	–	–	–	–	–	–
13h30·	70	–	59	5,6	0,7	70	14,6	7,1
14h30	–	2832	–	8,8	4,0	118	5,5	7,2
15h30*	72	–	67	8,6	6,6	150	0,5	7,4
16h30	69	–	–	6,3	2,5	109	7,5	–
17h00	71	–	–	–	–	–	–	–
17h30·	70	–	67	6,3	0,7	75	15,5	–
19h30*	70	2692	69	11,3	6,6	153	0,5	–

*start of aerobic period
·start of anoxic period

(a) grab samples from influent and effluent buckets over 24 hours period during which testing period occurs;
  (i) unfiltered influent: COD, TKN, NH_3, Alk and pH
  (ii) unfiltered effluent: COD, TKN, NH_3, Alk and pH
  (iii) filtered effluent: COD, TKN, NH_3, Alk, NO_3 and pH.

(b) on grab samples from the reactor:
  (i) samples at hourly intervals, filtered TKN, NH_3, NO_3, Alk and pH
  (ii) samples at 2- to 3-hourly intervals, MLVSS and filtered COD.

(c) on reactor every ½ hour during aerobic periods:
  (i) oxygen consumption rate.

As an example of the data measured during an intensive testing programme, the data plotted in Figure A3.1 are listed in Table A3.1.

3. THEORY AND CALCULATION

The μ_{nm} value is calculated from the increase in the nitrate concentration during the aerobic stages of the alternating cycle. From increase in the nitrate concentration the nitrification rate of the nitrifying organisms r_{Nn} can be calculated, and provided the ammonia concentration does not decrease below about 2 mgN/ℓ, this rate is directly linked to μ_{nm} i.e.

$$\mu_{nm} = r_{Nn} Y_n / X_n \quad (/d) \qquad (A3.4)$$

where

r_{Nn} = nitrification rate by the nitrifying organisms in mg NO_3-N/ℓ reactor volume/day
Y_n = yield coefficient for the nitrifiers
= 0,10 mgVSS/mgNO_3-N nitrified
X_n = reactor nitrifier VSS concentration (mgVSS/ℓ).

The nitrifier sludge concentration is calculated from the nitrate concentration generated per litre influent. Note that this concentration is not equal to the nitrate concentration in the effluent because during the anoxic periods nitrate is removed from the system by denitrification. An estimate of the nitrate generated per litre influent, N_{ng}, is found from the following equation:

$$N_{ng} = N_{ti} - N_{te} - N_s \quad (\text{mgN}/\ell) \qquad (A3.5)$$

where

N_{ng} = nitrate generated per litre influent (mgN/ℓ)
N_{ti} = influent TKN concentration (mgN/ℓ)
N_{te} = effluent TKN concentration (mgN/ℓ)
N_s = nitrogen required for sludge production (mgN/ℓ influent)

$$= f_n \frac{V_p X_v}{Q . R_s} \quad (\text{mgN}/\ell) \qquad (A3.6)$$

f_n = nitrogen content of the VSS
= 0,10 mgN/mgVSS
V_p = volume of the reactor (ℓ)
X_v = average VSS concentration over testing period (mgVSS/ℓ)
Q = influent flow rate (ℓ/d)
R_s = sludge age (d).

Knowing N_{ng}, the concentration of nitrifier sludge mass in the reactor is given by

$$X_n = \frac{Y_n (N_{ng}) R_s}{(1 + b_{nT} R_s) R_{hn}} \quad (\text{mgVSS}/\ell) \qquad (A3.7)$$

where

R_{hn} = nominal retention time of the process
= V_p/Q (d).

The increase in the nitrate concentration in the reactor during aerobic period r_{Nr} is the net effect of two reactions (i) the increase in nitrate by nitrifying organisms which is the r_{Nn} rate required for the μ_{nm} determination and (ii) the change in nitrate concentration in the reactor caused by nitrate leaving the system via the effluent, i.e. hydraulic effects, r_{Nh}, i.e.

$$r_{Nn} = r_{Nr} + r_{Nh} \qquad (A3.8)$$

Taking a time interval of $\Delta t = t_2 - t_1$ during the aerobic period where t_1 and t_2 are the beginning and end respectively of the time interval, and the nitrate concentrations in the reactor at times t_1 and t_2 are N_{n1} and N_{n2} mgN/ℓ respectively, then the net rate of nitrate increase per litre reactor volume

$$r_{Nr} = (N_{n2} - N_{n1})/(t_2 - t_1) \quad (\text{mgN}/\ell/\text{d}) \qquad (A3.9)$$

The rate at which nitrate leaves the reactor per litre reactor volume at a time instant is simply the product of effluent flow and the concentration of nitrate at the time instant divided by the reactor volume i.e. at time t_1: $r_{Nh1} = Q.N_{n1}/V_p = N_{n1}/R_{hn}$. Similarly at t_2: $r_{Nh2} = N_{n2}/R_{hn}$. Now the rate of nitrate leaving the reactor *during* the time interval bounded by t_1 and t_2 can be approximated by the average rates at times t_1 and t_2. Hence the average nitrate loss over the interval Δt due to hydraulic effects is given by

$$r_{Nh} = (N_{n1} + N_{n2})/(2\,R_{hn}) \quad (mgN/\ell/d) \qquad (A3.10)$$

The hydraulic effect causes the rate of nitrate generation to decrease so that this rate must be added to the rate of nitrate generation measured from the nitrate concentrations in the reactor as given by Eq. A3.8.

It should be noted that although only the nitrate concentrations during the aerobic cycles are used for the determination of μ_{nm}, the other analyses are useful to check that the experiment has progressed satisfactorily. These data can be used to determine and/or check constants other than μ_{nm}; the nitrate concentrations during the anoxic cycles can be used to determine the 2nd rate of denitrification in the primary anoxic reactor (K_2), but this determination requires a knowledge of the readily biodegradable COD concentration of the influent (f_{bs}). Also the Alkalinity values during the aerobic and anoxic periods can be used to check the μ_{nm} and K_2 determinations. For details see van Haandel and Marais (1981).

When all the data are measured, the constants can be determined and/or checked with the aid of computer simulation (see Figure A3.1). However, computer simulation is not essential because generally it has been found that when the manually calculated values of μ_{nm} (and K_2) are used in the simulation, very good correspondence with the experimental data is obtained.

4. EXAMPLE

To demonstrate the calculations, the μ_{nm} value will be calculated from the data listed in Table A3.1.

The nitrate concentrations measured at hourly intervals during the aerobic periods are listed in Table A3.2. From these data, the reactor nitrate generation rate r_{Nr} and the rate of nitrate loss by hydraulic effects r_{Nh} are calculated from Eqs. A3.9 and A3.10 respectively. The sum of the two rates gives the rate of nitrate generation by the nitrifiers r_{Nn} i.e. Eq. A3.8. From the 4 periods of one hour, a mean value of r_{Nn} is 8,45 mgNO$_3$-N/litre reactor volume/hour is found.

The average MLVSS concentration during the experiment is 2780 mgVSS/ℓ. From this, the nitrogen required for sludge production N_s is determined from Eq. A3.6 i.e.

TABLE A3.2: CALCULATION OF NITRIFICATION RATES FROM NITRATE CONCENTRATION PROFILES DURING THE AEROBIC PERIOD OF AN ALTERNATING ANOXIC-AEROBIC COMPLETELY MIXED PROCESS UNDER CONSTANT FLOW AND LOAD

Time	NO_3	r_{Nr}*	r_{Nh}**	r_{Nn}*
11h30	0,0			
		8,2	0,57	8,77
12h30	8,2			
		6,4	1,58	7,98
12h30	14,6			
15h30	0,5			
		7,0	0,56	7,56
16h30	7,5			
		8,0	1,60	9,60
17h30	15,5			
Mean				8,45

*units mgNO$_3$-N/ℓ reactor volume/h.
**nominal hydraulic retention time = 6/20 = 0,30 d ≅ 7,2 h.

$$N_s = 0{,}10\,\frac{6.2780}{20.6} = 13{,}9 \text{ mgN}/\ell \text{ influent flow.}$$

The average effluent TKN concentration (unfiltered) is 7,1 mgN/ℓ. Hence from Eq. A3.5, the nitrate generated per litre influent, N_{ng}, is

$$N_{ng} = 49{,}0 - 7{,}1 - 13{,}9 = 28{,}0 \text{ mgN}/\ell.$$

At 6 days sludge age and measured temperature 20°C, the nitrifier concentration in the reactor is found from Eq. A3.7

$$X_n = \frac{0{,}10.28.6}{(1 + 0{,}04.6)\;6/20} = 45{,}2 \text{ mgVSS}/\ell$$

Hence from Eq. A3.4, the maximum specific growth rate of the nitrifiers μ_{nm} at 20°C is

$$\begin{aligned}\mu_{nm} &= (8{,}45.24).0{,}10/45{,}2\\ &= 0{,}45\text{ /d. at } 20°C\end{aligned}$$

Hence $\mu_{nm20} = 0{,}45$/d.

If in this example, the temperature was not 20°C, the endogenous respiration rate in Eq. A3.7 for X_n needs to be adjusted with Eq. A3.3 for the given temperature. The value of μ_{nm} found from Eq. A3.4 is rate at the given temperature and to calculate μ_{nm20}, this value needs to be adjusted to 20°C with Eq. A3.2.

5. REFERENCES

VAN HAANDEL, A.C. and MARAIS, G.v.R. (1981) Nitrification and denitrification kinetics in the activated sludge process. Research Report W 39, Dept. of Civil Eng., Univ. of Cape Town.

SEHAYEK, E. and MARAIS, G.v.R. (1981) Kinetics of biological nitrogen removal in the activated sludge process. Research Report W 41, Dept. of Civil Eng., Univ. of Cape Town.

MARAIS, G.v.R. and EKAMA, G.G. (1976) The activated sludge process: Part I – Steady state behaviour, *Water SA*, 2(4) 163-200.

APPENDIX 4

LIST OF RSA BIOLOGICAL NUTRIENT REMOVAL PLANTS

1. Driefontein Water Pollution Control Works, City of Roodepoort

City of Roodepoort
Private Bag X30
ROODEPOORT
1725

2. Units 53 and 2/53, Sasol II, Secunda

SASOL II
P O Box 1
SECUNDA
2302

3. Meyerton Sewage Purification Works

Municipal Chemist
Meyerton Town Council
P O Box 9
MEYERTON
1960

4. Goudkoppies Sewage Purification Works, City Council of Johannesburg

Assistant Chief Sewerage Engineer
Wastewater Reclamation
City Engineer's Department
P O Box 4323
JOHANNESBURG
2000

5. Unit 3 of Northern Sewage Purification Works, City Council of Johannesburg

City Engineer's Department
Sewerage Branch
P O Box 4323
JOHANNESBURG
2000

6. Baviaanspoort Sewage Purification Works City Council of Pretoria

City Council of Pretoria
P O Box 1409
PRETORIA
0001

7. Rynfield Purification Works, Town Council of Benoni

Town Council of Benoni
Private Bag X014
BENONI
1500

8. Randfontein

Consulting Chemist
City of Krugersdorp
P O Box 94
KRUGERSDORP
1740

9. Standerton Sewage Purification Works

Municipality of Standerton
P O Box 66
STANDERTON
2430

10. Klerksdorp Sewage Purification Works

Town Council of Klerksdorp
P O Box 9
KLERKSDORP
2570

11. Stilfontein Sewage Purification Works

Town Council of Stilfontein
P O Box 20
STILFONTEIN
2550

12. Mitchell's Plain

and

13. Cape Flats Sewage Purification Works

Cape Town City Council
P O Box 1694
CAPE TOWN
8000

14. Gammans Sewage Works

City Engineer's Department
P O Box 59
WINDHOEK
9000
SOUTH WEST AFRICA

15. Rietspruit Sewage Works, Town of Vanderbijlpark

Town Council of Vanderbijlpark
P O Box 3
VANDERBIJLPARK
1900

16. Hartebeesfontein Sewage Purification Works, City of Kempton Park

Chief Chemist
City of Kempton Park
P O Box 13
KEMPTON PARK
1620

17. MONDI TIMBERS, Sabie

Chief Engineer
MONDI TIMBERS
P O Box 69
SABIE
1260

18. Delmas Sewage Purification Works

Municipality of Delmas
P O Box 6
DELMAS
2210

19. Pietersburg Sewage Purification Works

Municipality of Pietersburg
P O Box 111
PIETERSBURG
0700

20. Middelburg Sewage Purification Works

Municipality of Middelburg, Tvl
P O Box 14
MIDDELBURG, Tvl
1050

21. Potchefstroom Sewage Purification Works

Municipality of Potchefstroom
P O Box 113
POTCHEFSTROOM
2520

22. Bethal Sewage Purification Works

Municipality of Bethal
P O Box 3
BETHAL
2310

23. Umhlatuzana Wastewater Treatment Plant Borough of Pinetown

Borough of Pinetown
P O Box 49
PINETOWN
3600

24. Bethlehem Sewage Purification Works

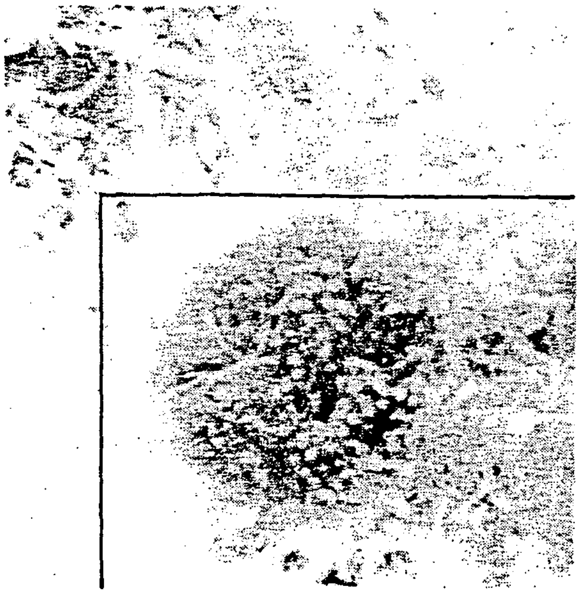
City Engineer's Department
Municipality of Bethlehem
BETHLEHEM
9700

25. Bellville Sewage Purification Works

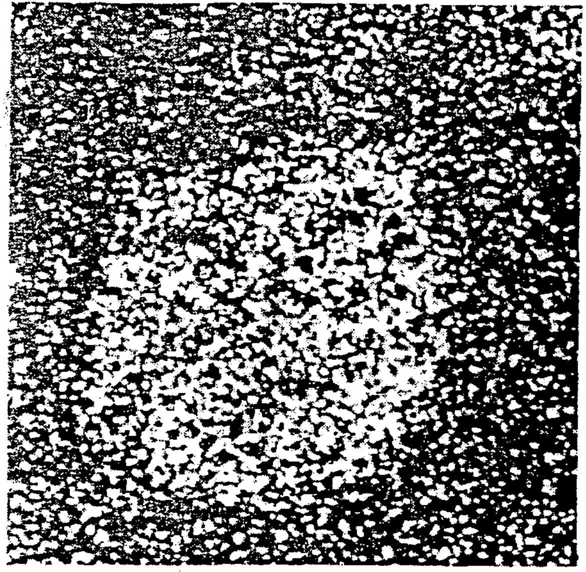
Municipality of Bellville
P O Box 2
BELLVILLE
7530

26. Brits Sewage Purification Works

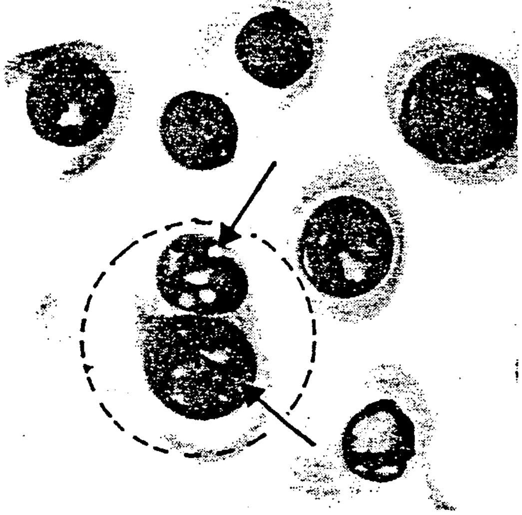
Municipality of Brits
P O Box 16
BRITS
0250



(a) The appearance of a sample of enhanced P-removing sludge in a scanning electron microscope. Framed area indicates a cluster of cells in the sludge matrix.



(b) Energy dispersive analysis of X-rays area map of phosphorus for the same area as (a) above. Concentration of dots indicated that phosphorus was associated with the cell cluster.

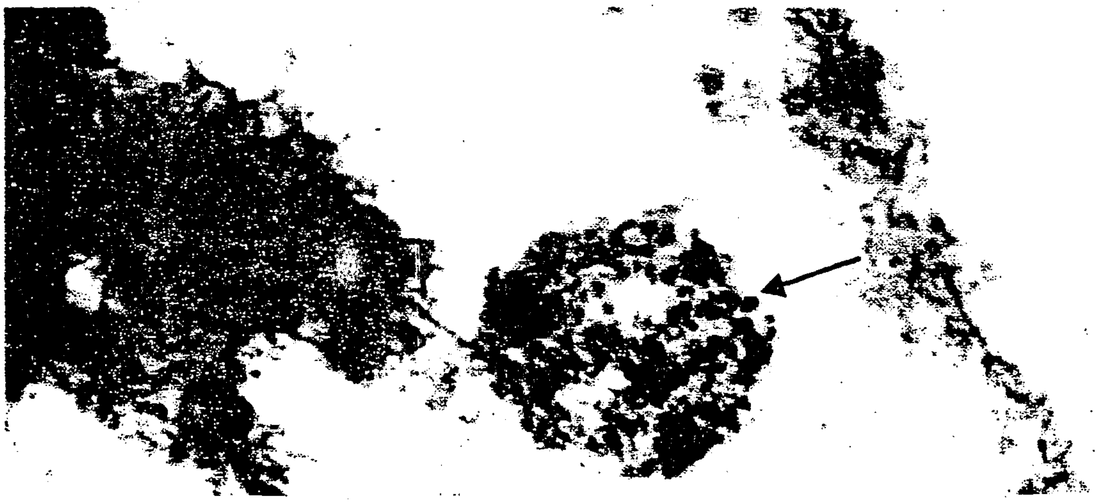


(c) Polyphosphate-containing cells as observed in a transmission electron microscope. Dotted outlines indicate single cells; arrows indicate polyphosphate inclusions.

FIGURE A5.2: Photomicrographs relating to *Acinetobacter spp*

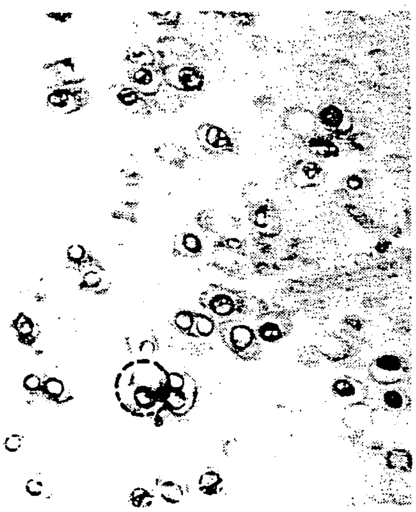
APPENDIX 5

PHOTOMICROGRAPHS OF BACTERIA FOUND IN NUTRIENT REMOVAL ACTIVATED SLUDGE PROCESSES



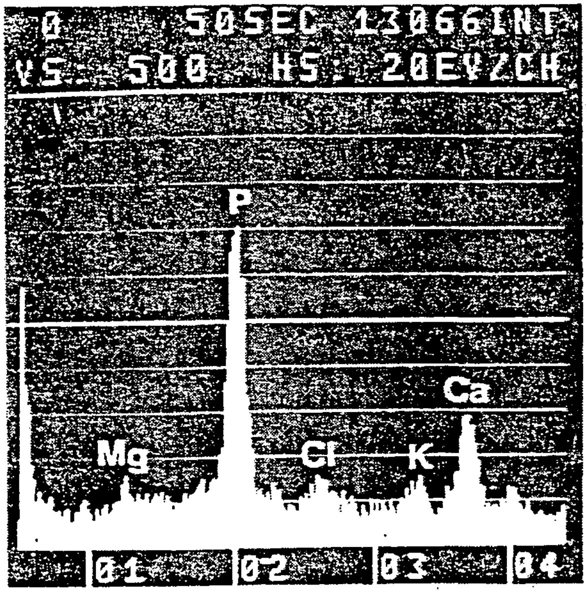
(a) A methylene-blue-stained specimen of enhanced P-removing sludge. Note the purple colour of the cluster of cells. The arrow indicates a single cell within the cluster.

(b) A cluster of phosphorus-rich cells in an enhanced P-removing sludge as observed in a scanning electron microscope. Dotted outline indicates a single cell.



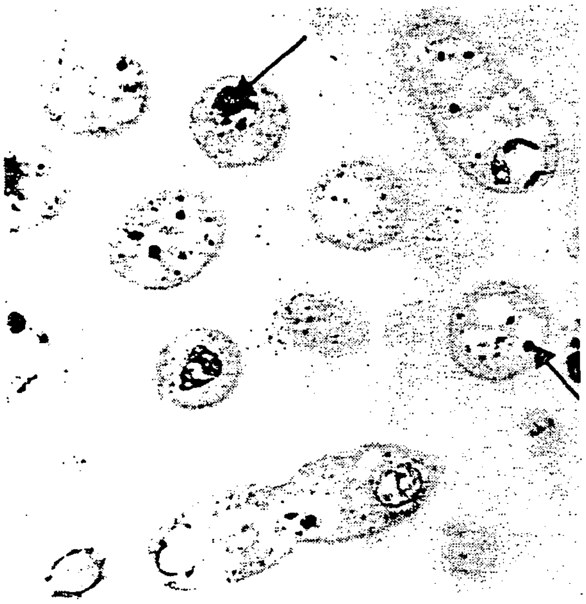
(c) A section of a cluster of phosphorus-rich cells as observed in a transmission electron microscope. The black spheres were polyphosphate granules and the white areas were where the polyphosphate granules had been torn from the cells during specimen preparation. Dotted outline indicates single cells.

FIGURE A5.1: Photomicrographs relating to *Acinetobacter spp*



(d) Energy dispersive X-ray spectrum for a polyphosphate granule indicating elements associated with the granule. Quantitative analysis indicated that the granules contained >25 % P and >8 % Ca.

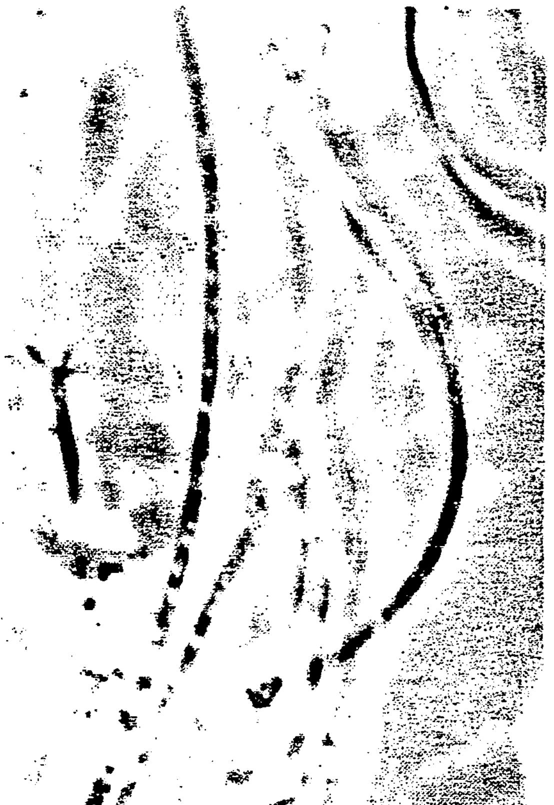
(e) The appearance of a specimen of sludge taken from the anaerobic zone of an enhanced-phosphorus-removing activated sludge plant as observed by light microscopy. The specimen was stained with Sudan-Black, a stain used for indicating PHB. The black granules indicated PHB inclusions.



(f) A transmission electron microscope picture of a sludge specimen taken from the anaerobic zone of an enhanced-phosphorus-removing activated sludge plant. The polyphosphate granules had disintegrated (black arrows) and PHB had appeared (white arrows).

FIGURE A5.2 (Continued).

(a) Light microscopic pictures of bulking activated sludge indicating the filamentous nature of organisms within the sludge.



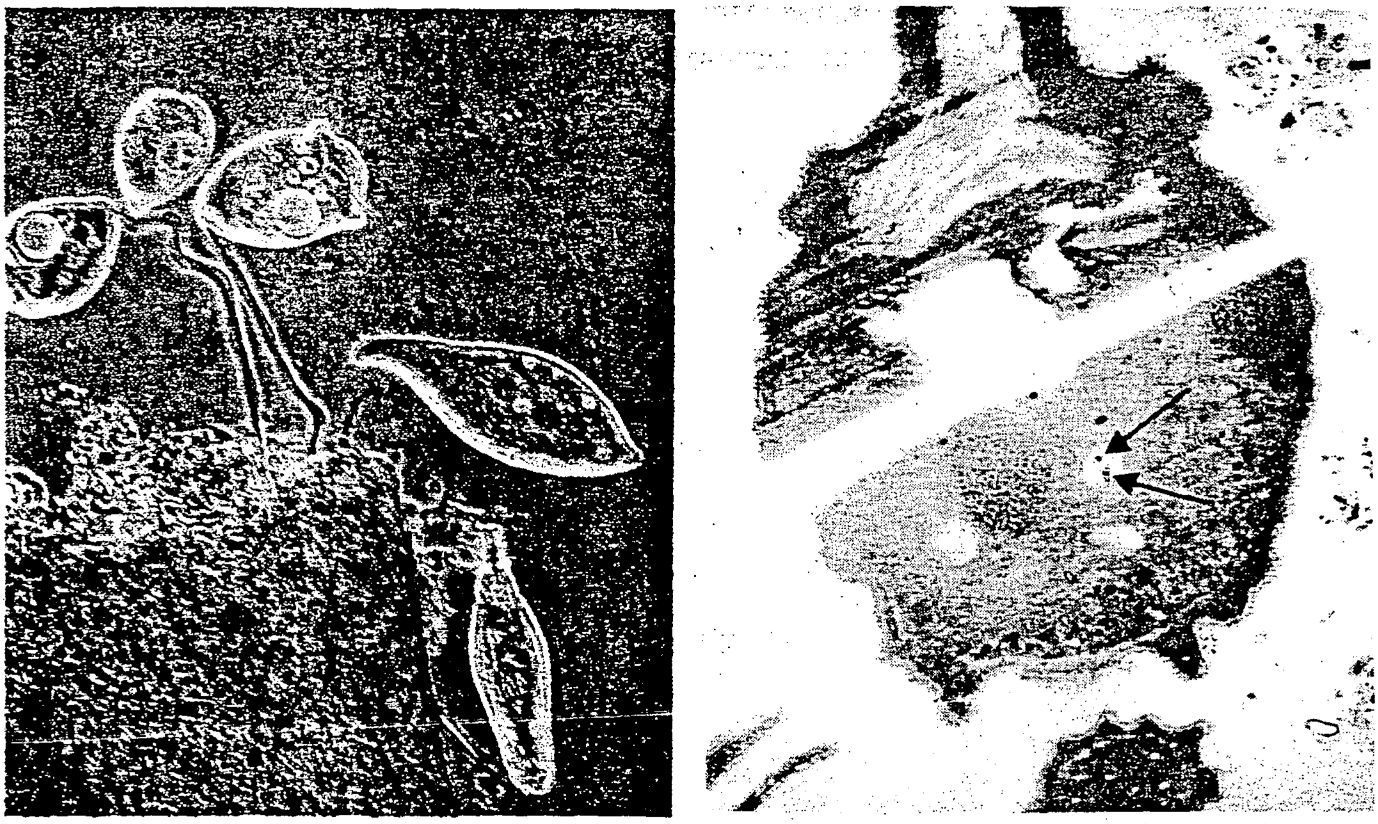
(b) As in (a) above

FIGURE A5.3: Photomicrographs of filamentous activated sludge

(c) A scanning electron microscope picture of a bulking sludge.

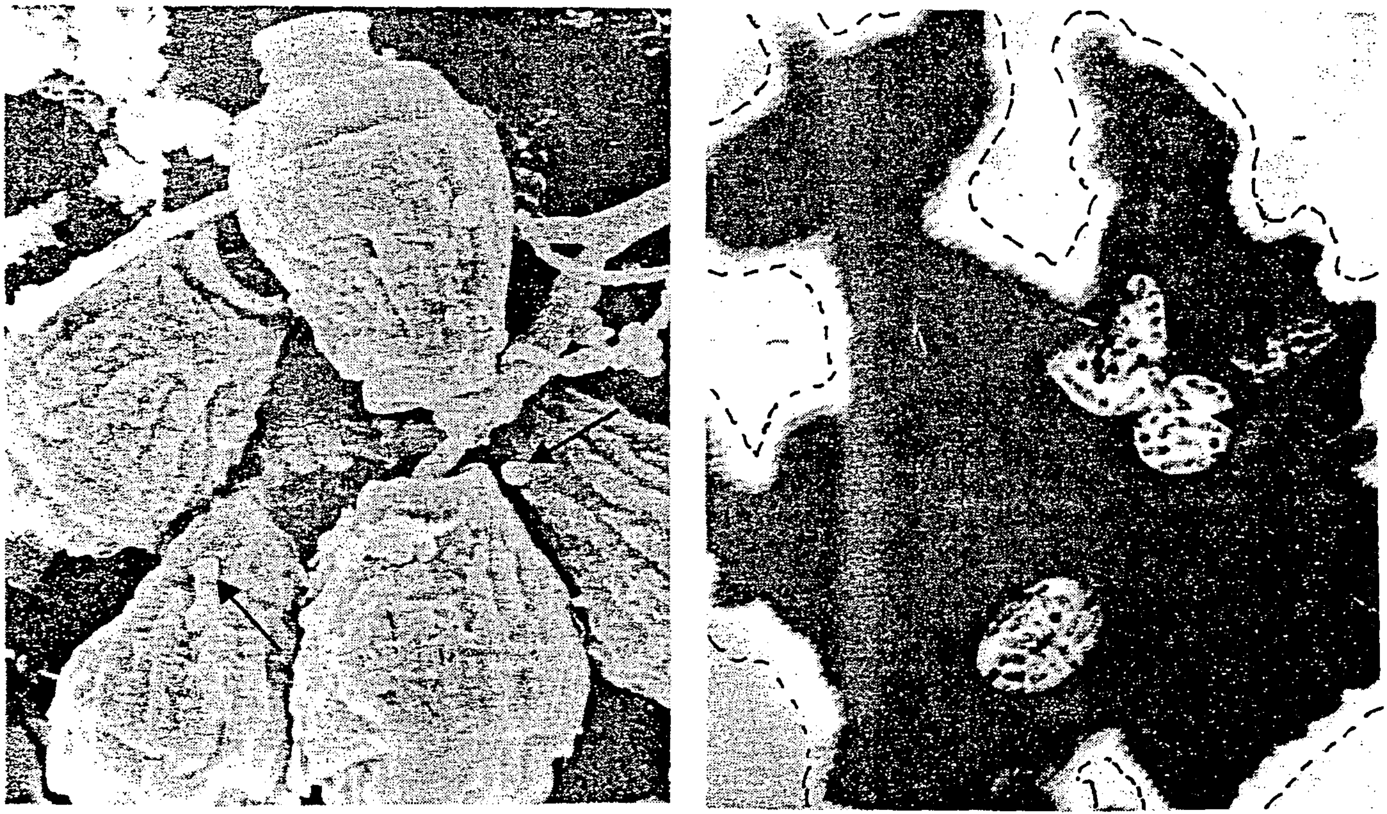
(d) A transmission electron microscope picture of a filamentous organism in bulking activated sludge. The single cells (black arrow) are enclosed within a sheath (white arrow).

FIGURE A5.3 (Continued)



(a) A light microscopic picture of the protozoan, *Vorticella*, in activated sludge.

(c) A section of a *Vorticella* as observed in a transmission electron microscope. The arrows indicate two phosphorus-rich cells within the *Vorticella*.



(b) A scanning electron microscopic picture of *Vorticella* in activated sludge. Note the bacteria (arrows) on the *Vorticella*.

(d) A section of an amoeba as observed in a transmission electron microscope. The amoeba had ingested a large number of phosphorus-rich cells. Dotted outline indicates amoeba. Note the three clumps of cells within the amoeba – the arrow indicates a single cell.

FIGURE A5.4: Photomicrographs of various protozoa and metazoa