EVALUATION AND OPTIMIZATION OF DUAL DIGESTION OF SEWAGE SLUDGE

PART 1 : OVERALL SYSTEM PERFORMANCE

by

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GLOSSARY OF SYMBOLS AND ABBREVIATIONS

ATAD	Autothermal thermophilic aerobic digestion
COD	Chemical oxygen demand (g/L or kg/m^3)
CST	Capillary suction time test (seconds)
DD	Dual digestion
H_{bi}	Rate of biological heat generation (MJ/h)
H_{mi}	Rate of heat generation in the sludge due to mechanical action of pumps or stirrers (MJ/h)
H_{net}	the rate of aerobic reactor sludge enthalpy increase during the specific heat yield test (MJ/h)
H_{se}	the rate of heat loss from the aerobic reactor due to the sensible heat of the reactor effluent sludge (MJ/h)
H_{si}	the rate of heat gain to the aerobic reactor due to the sensible heat of the reactor feed sludge (MJ/h)
H'_{se}	the net rate of heat loss from the aerobic reactor due to the sensible heat of the reactor feed and effluent sludges (MJ/h). This is the rate of heating required to change the temperature of the sludge from the feed to the effluent value (MJ/h)
H_{ve}	the rate of heat loss from the aerobic reactor in the latent heat of the reactor vent gas (MJ/h)
H_{we}	the rate of heat loss from the surfaces of the aerobic reactor and the Vitox oxygenation system (MJ/h)
IA	Intermediate Alkalinity - determined by titrating from pH 5.75 to pH 4.3. This approximates the Volatile Acid alkalinity
M_{wc}	the mass flow rate of condensate accumulation in the water traps of the aerobic reactor vent gas metering system during the specific heat yield test (kg/h)
O_c	Oxygen consumption rate (kgO/h)

O_{eff}	the fraction of oxygen supplied to the aerobic reactor which is consumed by the reactor sludge. This symbol is used in the specific heat yield test and is equivalent to OTE, the symbol preferred in design procedures.
O_s	the mass rate at which oxygen is supplied to the aerobic reactor sludge biomass (kgO/h)
OSR	Oxygen supply rate to the aerobic reactor [kgO/(m^3.h)]
OSS	Ordinance on Sewage Sludge (Germany)
OTE	Oxygen transfer efficiency i.e. the fraction of the oxygen supply which the oxygenation system is able to transfer to the mixed liquor. This symbol is used in design procedures and is equivalent to O_{eff}, the symbol preferred in the specific heat yield test.
OTR	the oxygen transfer rate i.e. the rate at which oxygen is transferred to the mixed liquor [kgO/(m_3.h)]
OUR	the biological oxygen utilization rate, i.e. the volume specific rate at which sludge consumes oxygen under oxygen sufficient conditions y[kgO/(m_3.h)]
PA	Partial Alkalinity - determined by titrating from original sample pH to pH 5.75. This corresponds roughly to the bicarbonate alkalinity.
PFRP	Process to Further Reduce Pathogens
PREST	Aerated Thermophilic Digestion Prestage before Anaerobic Digestion
$_s$	the sludge density (ton/m^3)
PSA	Pressure swing absorption
PSRP	Process to Significantly Reduce Pathogens
R^H	the hydraulic retention time of the sludge in the aerobic reactor
SOUR	Specific oxygen uptake rate (mg/h.gTSS)
SRF	Specific resistance to filtration (m/kg)
SRT	biological solids retention time (h)
SRT^c	Critical value for SRT, after which process failure occurs (h)

SRT^d	design solids retention time (h)
STP	Standard Temperature and Pressure
T^{ge}	the temperature of the vent gas at the point of exit from the aerobic reactor (°C)
T^{gm}	the temperature of the aerobic reactor vent gas passing through the vent gas meter during a specific heat yield test (°C)
t^{min}	Undisturbed reaction time (hours or minutes)
T^{se}	the temperature of the sludge in the aerobic reactor or the temperature of the effluent sludge from the aerobic reactor (°C)
T^{si}	the temperature of the feed sludge to the aerobic reactor (°C)
TS	Total solids concentration (g/L or kg/m^3)
TSS	Total suspended solids concentration (g/L or kg/m^3)
U	heat transfer coefficient [$MJ/(m^2.°C.h)$]
Vitox	Patented oxygenation system
V^{gm}	the volumetric flow rate of vent gas measured by the vent gas meter during the specific heat yield test (m^3/h)
VKS	Vereinigung Kommunaler Städtereinigungsbetriebe
V_p	the operating volume of the aerobic reactor (m^3)
VS	Volatile solids concentration (g/L or kg/m^3)
VSS	Volatile Suspended solids concentration (g/L or kg/m^3
WRC	Water Research Commission
Yco_2	Respiration quotient, i.e. mole CO_2 generated per mole O_2 consumed.
Y_H	Specific heat yield i.e. the quantity of biological heat generated per unit of oxygen consumed (MJ/kgO)
Zimpro	Patented high temperature, high pressure wet air oxidation process used to treat sewage sludge

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CHAPTER 1

INTRODUCTION

1.1 SEWAGE SLUDGE TREATMENT AND DISPOSAL

Sludge is produced by all biological waste water treatment processes and the treatment and disposal of the sludge generated has become a problem of increasing magnitude. In most cases, nearly half of the cost of waste water treatment plant operation is incurred by the handling and disposal of sludge. Final disposal of sludge is usually by one of the four most common options, viz: landfill, ocean dumping, incineration, or by land application. All of these options could create health, environmental and economic problems if not judiciously managed, and consequently prescribed codes of practice for sludge disposal exist in most countries.

Land application for agricultural purposes is the oldest and most common means of using sludge beneficially. Sludge used for this purpose should, however, be stabilized and disinfected to such a degree so as not to create nuisance conditions or be hazardous to health. It should also contain only limited amounts of heavy metals and chemical compounds which would accumulate i the soil if the sludge were to be applied over long periods.

Sludge stabilization by aerobic or anaerobic digestion is practised worldwide. In the United Kingdom and South Africa the anaerobic process is favoured, while the aerobic process is more common in most European countries and in the United States. The type of waste water treatment plant employed also influences the choice of sludge stabilization process. Aerobic digestion is most commonly used at activated sludge plants, while anaerobic digestion is historically associated with trickling filter plants. Other factors such as energy considerations, plant location, the type of community served, operator skill, and sometimes personal preferences may affect the choice of stabilization process employed.

All the digestion processes used for stabilization effect a degree of disinfection as well. Disinfection or pasteurization of sewage sludge before disposal or being applied to land for agricultural purposes is now mandatory in some countries. The degree of disinfection prescribed varies, but in general the sludge has to undergo treatment aimed specifically at improving its hygienic quality. To achieve this objective, a number of options are available. These include chemical treatment (lime, methyl bromide, ammonia, etc.), irradiation, and various forms of heat treatment, including pasteurization.

The heat treatment option is the one that is the most commonly used. High temperature processes such as the Porteous, Zimpro, flash drying or rotary kiln drying are very effective in inactivating pathogens and parasites, but are usually

very costly. Pasteurization, both pre- and post-stabilization, has frequently been used, but the latter was plagued by rapid reinfection and regrowth of pathogenic organisms. Pre-pasteurization is now commonly used, but it is an expensive process if waste energy is not available.

The heat treatment processes mentioned all require external energy input, which, with rapidly increasing energy prices, make them less attractive.

Alternative processes have been developed which utilize the heat liberated by the biological degradation of the organic fraction of the sludge to effect both stabilization and disinfection under the right conditions. These processes require a much reduced external energy input, often as mechanical energy, which makes them attractive in the energy conscious world of today. The processes in this category include forced aeration composting, autothermal thermophilic aerobic digestion (ATAD), and dual digestion (DD). The evaluation of latter process is the subject of this report.

The patented dual digestion process combines autothermal thermophilic aerobic digestion and anaerobic digestion. The process is claimed to give a high degree of disinfection and stabilization together with the added benefits of a low energy requirement, reduced capital cost and excellent process stability.

Recognizing the potential of the dual digestion process while it was still under development in the USA, the Water Research Commission (WRC) sponsored research by the Johannesburg Municipality into autothermal thermophilic aerobic digestion, i.e. the first stage of dual digestion, on a pilot scale The promising results (Trim, 1984) prompted further research and accordingly, a tripartite agreement was finalized between the WRC, the Division of Water Technology of the CSIR, and Milnerton Municipality to evaluate a full-scale dual digestion process at Milnerton's Potsdam waste water treatment works.

This report covers the research carried out under the joint agreement and also incorporates detailed research into the ATAD process, specifically as it is applied in the first stage of dual digestion. The latter research was submitted as a PhD (Eng.) thesis at the University of Cape Town (Messenger, 1991).

1.2 DUAL DIGESTION

The dual digestion process was developed to overcome one of the inherent disadvantages associated with autothermal aerobic digestion using oxygen. In the USA several oxygen aerobic sludge digestion plants have been installed at works using the "Unox" oxygen activated sludge process, since this combination can utilize oxygen generated on site fairly efficiently. The oxygen is usually produced

by means of the Pressure Swing Absorption (PSA) process. The high energy cost of producing the oxygen and then dissolving it efficiently during aerobic digestion to achieve full stabilization has limited the application of this process, hence the advent of the dual digestion process.

As the title suggests, dual digestion is a two stage process which was developed in 1975 by the Union Carbide Corporation. The first stage consists of an auto-thermal aerobic reactor which operates at a retention of approximately one day and is capable of maintaining thermophilic temperatures. The second is an anaerobic digester operating at eight to ten days retention, and usually operates at temperatures in the mesophilic range.

Sludge is fed to the first stage where it is oxygenated. Heat released during biological oxidation of some of the organic matter maintains the temperature in the reactor at thermophilic levels. Disinfection and partial stabilization is therefore achieved in the first stage. The pre-treated sludge passes to the anaerobic stage where stabilization is completed and biogas is produced.

Dual digestion claims the following main advantages over its component processes:

1. Disinfection and stabilization of the sludge occurs in one process. The thermophilic temperatures in the first stage disinfect the sludge while full stabilization occurs in the second stage.

2. The use of energy for the supply and dissolution of oxygen is low since the degree of oxidation of the sludge limited, the only requirement being the maintenance of the chosen temperature to effect disinfection. Stabilization is completed in the anaerobic stage, which uses no oxygen.

3. No external heating of the anaerobic digester is required as all the heat required is provided by the hot sludge fed from the first stage.

4. The process is an overall net producer of energy since all the biogas, which contains approximately 65% methane, is available for use external to the digestion plant. By way of comparison, a conventional anaerobic digestion process consumes 40 to 100% of the methane produced to maintain the optimum operating temperature.

5. The process may be started up much more quickly than an anaerobic plant, eight to ten days being required instead of thirty to forty days to reach full operating capacity.

6. The stability of the anaerobic stage is considerably improved by the conditioning effect of aerobic stage, allowing more rapid digestion (8 to 10 days retention), by the following processes:

* The effect of organic toxic contaminants of the feed sludge is reduced;
* Oxidation of the most rapidly degradable components of the feed sludge which helps to maintain the balance of the growth rates of the acid forming and methane producing bacteria;
* Solubilization of part of the solid organic material at the thermophilic temperature of the first stage;
* Production of ammonium ions by the deamination of organic nitrogen compounds. These increase the alkalinity of the sludge which is very beneficial for maintaining the buffer capacity of the anaerobic process. It is unlikely that any nitrification will occur in the aerobic stage as the temperature is too high;
* Temperature fluctuations are minimized by controlling heat input from the aerobic stage.

7. Although the system requires two separate digestion tanks, the overall capital cost of the process is lower than that of conventional anaerobic digestion. This is a consequence of the reduced tank volumes required (60% less volume for the anaerobic stage due to increased digestion rate) and the fact that no heating equipment is necessary.

8. The dual digestion process is particularly suited to cases where existing anaerobic digestion plant has to be upgraded. By installing an aerobic reactor ahead of the existing anaerobic digester, the capacity of the plant may be increased by a factor of 2 to 3.

From a practical point of view the last claim is of particular importance and if it can be substantiated then dual digestion would be a viable system for upgrading existing fully loaded anaerobic digesters. Its application would be particularly appropriate in South Africa where mesophilic anaerobic digestion is the preferred method for sludge stabilization. Where disinfection of the sludge is required or where existing digesters are overloaded, dual digestion could be implemented to effect both disinfection and increased digestion capacity. At many treatment plants anaerobic digesters have been rendered ineffective due to ageing or inoperative heating equipment. In these cases, converting the sludge treatment to dual digestion by the addition of an aerobic stage would be doubly beneficial.

1.3 RESEARCH AND DEVELOPMENT PROGRAMME

The research and development programme carried out at Milnerton to evaluate the performance of the dual digestion process comprised the following main tasks:

1. Establish a large-scale dual digestion facility capable of treating about 50 m^3 sludge per day, using existing tanks for the sludge thickener, aerobic reactor and for the anaerobic digester. The necessary modifications to the existing equipment is to be carried out and the required pumps, pipework, oxygenation equipment, controls, monitoring apparatus, etc. installed.

2. Commission and evaluate the aerobic stage and when operational, commission the anaerobic digester and evaluate. Specific items to be addressed are:

* Oxygen requirements and oxygen utilization efficiency, with pure oxygen and air oxygenation - the latter to also include control of foam formation when using air;

* Minimum retention time in the aerobic reactor while maintaining thermophilic temperatures and effective disinfection;

* Temperature control of the aerobic reactor and the anaerobic digester;

* Evaluate disinfection potential of the process and establish mode of operation to prevent recontamination;

* Evaluate anaerobic digester performance with respect to minimum retention time for effective stabilization, gas production, sludge stability, operating stability, and dewater ability of the final sludge;

* Establish process operating and control procedures, design criteria for future plants;

* Economic and technological assessment of the process. Its performance is to be compared to that of the Zimpro process at Milnerton;

* Final report, including a literature review.

CHAPTER 2

LITERATURE REVIEW

2.1 INTRODUCTION

It has long been recognized that the most beneficial means of sludge disposal to land is for agricultural use. Ideally the sludge used for this purpose must be stabilized and disinfected to a degree where it will not create nuisance conditions or pose a health threat to humans or animals. The classical sludge treatment processes such as aerobic or anaerobic digestion, usually at mesophilic temperatures, are capable of stabilizing the sludge but generally provide only a low degree of disinfection. Disinfection is usually effected by an additional process either before or after stabilization. Heat treatment, such as pasteurization is commonly used, but this always entails a large energy requirement.

The principal purpose of digestion is to convert, at minimum cost, the raw sludge to a stabilized product. The degree of stabilization required or achieved is often ill-defined or entirely subjective. Sewage plant operators usually judge a sludge to be stable when it does not produce unpleasant odours or attract flies. Scientists or engineers require quantification of the degree of stability, which is generally based on the reduction in the available biodegradable energy. The decrease in the volatile solids (VS) content has traditionally been used to gauge stability, but other parameters such as COD reduction or specific oxygen uptake rate (SOUR) is now also finding favour.

In search of a process that would both stabilize and disinfect sludge, autothermal thermophilic aerobic digestion (ATAD) was developed. This process is capable of disinfecting sludge during a very short retention period, typically 2 to 4 hours, but stabilization is only attained after 5 to 10 days. As a relatively high energy input is required because of the necessity of supplying oxygen to the process, full treatment by ATAD would therefore be uneconomical. In contrast anaerobic digestion can achieve good stabilization with a low energy demand, but over considerably longer retention times, typically 15 to 30 days, which requires large digesters and a high capital cost.

The dual digestion (DD) process was developed to incorporate the advantages of both ATAD and anaerobic digestion in terms of disinfection and stabilization. The anaerobic digestion process is fairly well understood and is commonly applied for sewage sludge digestion. It is, however, important to examine aerobic digestion, and specifically thermophilic aerobic digestion, in greater detail, particularly when heat production is a key process operating criterion to ensure disinfection.

A detailed analysis of the stoichiometry and kinetics of biological heat generation in the ATAD process, as applied in the first stage of the dual digestion process, is given in the PhD thesis by Messenger, of which a summarized version forms

Part 2 of this report. A more general overview of the theory and of recent studies is presented in this review to serve as a background to the research carried out for this project.

2.2 SEWAGE SLUDGE AS A FEED SUBSTRATE

Waste sludge produced by primary and secondary sewage treatment processes is a complex and highly variable feed substrate consisting of biodegradable and non-biodegradable matter, both solid and liquid. Non-biodegradable matter, irrespective of its physical state, is by definition not biodegradable. Some of this matter may, however, be modified during digestion processes so that its behaviour is significantly changed which may, for example, enhance its dewatering properties. Most of the non-biodegradable solids will pass through the stabilization process essentially unchanged and will ultimately comprise the non-putrefactive fraction of the stabilized sludge.

Biodegradable matter may comprise solids of microbial or non-microbial origin as well as soluble, water immiscible and sorbed compounds (Hamer *et al.*, 1986).

The biodegradable microbial fraction comprises viable, non-viable and dead microbes which forms a significant part of waste secondary sludge, although a small fraction may consist of potentially pathogenic micro-organisms which were present in the original sewage. Microbes generated as a result of the utilization of biodegradable substrate during the stabilization process also represent additional feed for subsequent biodegradation.

A wide selection of other biodegradable solids of non-microbial origin are also present in waste sludge, with cellulose in its various forms being the predominant constituent (Hamer *et al.*, 1986). Degradation models for cellulose indicate that only partial breakdown will occur during the relatively short period of treatment and that there will always be some non-degraded cellulose present in the treated sludge. Cellulose, being a natural compound, will decay slowly without significant adverse effects when the treated sludge is spread on agricultural land.

Relatively low amounts of soluble biodegradable matter usually occur in waste sludges. When feed sludges are subjected to extended holding times in storage tanks prior to treatment, significant concentrations of soluble partial biodegradation products can accumulate together with the facultative anaerobic bacteria that mediate degradation during storage (Hamer and Bryers, 1985).

Water immiscible biodegradable compounds, mainly oils, occur in sludges as oil is frequently discharged into sewers, usually against the regulations imposed in most countries. During sewage treatment, oil will be removed either by skimming or, if sorbed onto solids, by sedimentation. Both the skimmed material and the sedimented solids end up as primary sludge, which comprises a significant fraction of the total waste sludge produced.

2.3 AEROBIC DIGESTION

2.3.1 Microbiological Aspects

When biodegradable matter (substrate) is aerobically degraded by microbes, assuming that all other nutrients essential for the aerobic growth of the population are present, a fraction of the substrate carbon is converted into carbon dioxide and energy is generated (catabolism). Part of this energy is used by the microbes performing the oxidation to fix another fraction of the substrate carbon to form additional microbial biomass (anabolism). A small part of the remaining energy is used for maintenance requirements, the balance is released as heat (Hamer and Zwiefelhofer, 1986). Thus the dominant activity is the production of biomass at the expense of substrate. When an active microbial culture is deprived of its feed substrate, it reverts to using stored intracellular matter and the lysed contents of dead cells as its carbon substrate, the processes being termed endogenous respiration and "cryptic" growth (Mason, 1986).

As microbes can utilize only soluble and/or miscible carbon energy substrates, the solubilization of particulate matter is a vital step in treatment of wastewater sludges. Despite the fact that the rate of solubilization is the limiting step in the biological degradation of sludge, very little is known about this aspect of the treatment processes and not much research is being undertaken in this field (Mason *et al.*, 1987; Eastman and Ferguson, 1981). A recent study by Bomio *et al.*, (1989), indicated that only 32 % of the thermophilic activity measured during an aerobic digestion experiment was attributable to growth on the soluble fraction of sewage sludge, while the main activity (68 %) was bound to growth on particulate matter. They also found that the thermophilic populations preferentially utilize the proteinaceous material in the sludge as carbon source and that the proteolytic activity was the main enzymatic activity in aerobic thermophilic sewage sludge digestion. Other extracellular activities (lipase, amylase, pectinase and cellulase) were not detected in the sludge investigated, which would suggest that carbohydrates, lipids and other polymers were either not present in significant amounts or were passed through the aerobic thermophilic process almost unchanged.

Aerobic thermophilic digestion has been proposed as a means of improving the effectiveness of waste sludge treatment, possibly in conjunction with existing anaerobic mesophilic digestion. This would supposedly give improved removal of toxic substances and pathogenic organisms. Improved solubilization of particulates in the aerobic process should definitely enhance the performance of the anaerobic stage, as the hydrolysis of particulates to soluble substrates is recognized as being the rate limiting step in the acid phase of anaerobic digestion. (Eastman *et al.*, 1981). It has been suggested that the aerobic stage should be an "acidification stage" where volatile fatty acid production should be stimulated. The methanogenic bacteria in the anaerobic stage could then utilize the acetate produced directly. Bomio *et al*, (1989) found that only minute amounts of fatty acids could be produced under severe oxygen limitation and that these were only isomers of butyrate and valeriate.

They deduced that although aerobic thermophilic populations were able to produce acetate in the presence of starch and mono- or disaccharides, these would only occur infrequently in sewage, so that acetate production would not occur under normal circumstances.

2.3.2 Process Aspects

There is a wealth of information and research findings available on conventional aerobic digestion at both mesophilic and thermophilic temperatures. The main focus of this information is, however, directed at the reduction of the amount of biodegradable material in the sludge, i.e. to maximize stabilization, and the disinfection aspect is rarely considered.

Aerobic digestion of municipal wastewater sludge was started in the early 1950's and indications were that results at least as good as those produced by anaerobic digestion could be achieved (USEPA, 1979; Eckenfelder, 1956; Lawton *et al.*, 1964). Aerobic digestion has become popular since then because it has some significant advantages i.e. lower capital costs, ease of operation and it produced a less problematical supernatant. The disadvantages are: high power cost to oxygenate the sludge and production of a digested sludge which has poor dewatering characteristics.

In conventional aerobic digestion the aim is to stabilize the sludge in the shortest time at the lowest cost. The hydraulic retention time and the oxygen demand required to achieve this goal are the primary factors governing the design of the process. The rate of biodegradable volatile solids destruction (dBVS/dt) dictates the hydraulic retention time and this rate must be as high as possible to minimize the retention time and therefore the volume requirements of the plant. The oxygen demand is obtained from the stoichiometric ratio of the mass of oxygen required per unit mass of volatile solids destroyed.

First order kinetics are generally used to describe the aerobic digestion process:

$$dBVS/dt = -k \times (BVS)t \quad (kg.h^{-1})$$

Where:

BVS	=	biodegradable volatile solids	$(kg.m^{-3})$
k	=	VS reduction rate coefficient	(h^{-1})
t	=	time	(h)

From the above the hydraulic retention time R_h can be deduced, assuming that the system is continuous flow and fully mixed:

$$R_h = \left(\frac{1}{k} \frac{(BVS)_i}{(BVS)_f} - 1 \right) \qquad (h)$$

Where:

R_h = hydraulic retention time (h)

and subscripts *i* and *f* denote initial and final states respectively.

The oxygen consumed during aerobic digestion at ambient temperatures is used not only for oxidation of the carbon fraction but also for nitrification. The oxygen used for nitrification does not contribute to volatile solids reduction, so omitting this fraction, the carbonaceous oxygen utilization rate (OUR) may be calculated from the product of the rate of VS reduction and the COD/VS ratio of the sludge:

$$OUR = k \times [BVS]_f \times \frac{COD}{VS} \qquad (kgO.m^{-3}.h^{-1})$$

From the above equations it is clear that the value of the rate coefficient "*k*" has a considerable effect on the entire process. Its value increases with temperature and is generally believed to follow an Arrhenius relationship over a narrow temperature range, approximately in the meso- and lower thermophilic range and can be approximated by:

$$k_T = k_{20}\, \theta(T-20) \qquad (h^{-1})$$

It is also a function of the sludge type, waste activated sludge requiring less time to digest, therefore having a higher *k* value. (The above from: Messenger, 1991; Matsch *et al.*, 1977; Gould *et al.*, 1978).

2.4 THERMOPHILIC AEROBIC DIGESTION

At thermophilic temperatures the rate of VS breakdown is considerably higher than at ambient temperatures. For example, an aerobic digester operating at ca. 50 °C can attain the same degree of degradation as a system operating at 20 °C in half the retention time (Matsch *et al.*, 1977; Fuggle *et al.*, 1985). The total oxygen requirement is the same, because although the OUR increases, the increased rate is required for a shorter period. Note that at operating temperatures above 45 °C nitrification will be totally inhibited (Matsch *et al.*, 1977; Brock *et al.*, 1970). No oxygen is therefore required for nitrification.

The theoretical oxygen requirement for aerobic digestion at thermophilic temperatures was deduced by Matsch and Drnevich (1977) by assuming that the degradable fraction of sludge consists of microbial cell mass which may be represented by the formula $C_5H_7NO_2$. The oxidation of this material can then be expressed as:

$$C_5H_7NO_2 + 5O_2 = 5CO_2 + 2H_2O + NH_3 + \text{energy}$$

Note that nitrification is inhibited at thermophilic temperatures.

From this equation the theoretical oxygen demand can be calculated as being 1,42 kg O_2/kg organic material oxidized. Also, 1 mole of CO_2 is produced from each mole of oxygen consumed. In practice, however, figures of around 2,0 kg O_2 per kg organic matter have been reported (Gould and Drnevich, 1978; Booth and Tramontini, 1983; Trim,1984).

Pure oxygen or air may be used for oxidation. Heat balance calculations (Matsch *et al*, 1977) indicated that high operating temperatures could not be reached in systems using air at total solids levels of 5% or less. It was calculated that the large volume of air required would carry away too much of the heat liberated by the digestion process. Subsequent calculations by Fuggle and Spensley (1985) showed that in treating a thick sludge (6% TSS) the maximum temperature rise was 83°C using oxygen, and only 44°C using air (at 20% O_2 utilization). With a "thin" sludge (3% TSS) temperature increases of 44°C and 24°C were possible for oxygen and air systems respectively.

Another drawback of using air is the additional energy required to transfer the oxygen to the sludge. The 20% efficiency assumed was considered to be high for conventional transfer devices, especially at the high temperatures employed. Lower transfer efficiencies would result in higher gas flows with higher heat losses due to vaporization and lower digester temperatures. When using oxygen, poor oxygen transfer efficiencies have a much smaller effect on the heat loss and digester operating temperature.

Jewell and Kabrick (1980) showed that it was possible to reach temperatures in excess of 50°C employing air, using a special aerating device which gave transfer efficiencies in excess of 20% in pilot digesters fed with a mixture of waste activated sludge and primary sludge. Maximum temperatures and organic removal rates were attained at loading rates of between 8 and 10 kg of biodegradable COD/m^3.d. It was determined that if 70% of the biological heat generated could be conserved, a simple equation could be used to estimate the temperature increase in the digester:

$$\Delta T = 2.4 \Delta COD$$

Where:

ΔT = increase in temperature in digester, °C
ΔCOD = COD in g/L oxidized at <8 days hydraulic retention.

Wolinsky (1985), reporting on the operation of a 60m^3 ATAD plant at Palmesford using air injected into a venturi in an external loop through which the sludge in the digester was continuously circulated, found that temperatures of 60°C or higher could be maintained. This was ascribed mainly to the enhanced oxygen transfer efficiency obtained in the foam blanket which formed on top of the sludge in the digester. This foam formation apparently could result in 100% oxygen transfer in some cases. The foam layer did, however, occupy about half the digester volume.

The capital benefits of ATAD due to smaller digestion vessels etc. are considerable. Retention times as low as 0,53 d have been reported while maintaining a digester temperature of 55°C (Farrell, 1984).

An important additional benefit of thermophilic operation is the disinfection achieved at the higher digestion temperature. This aspect is covered in greater detail in Section 2.6 on disinfection.

To maintain a digester at thermophilic temperatures requires suitable precautions to limit heat losses, and sufficient heat must be supplied to the process at a defined rate. Some of the heat supplied may be derived from direct process sources e.g. mixing energy, heat exchange or pumping. Other external sources may also contribute e.g. steam injection. Usually the major contribution is derived from the biological activity which makes the process self sustaining, hence the term autothermal thermophilic aerobic digestion (ATAD).

In the literature, three parameters have been used to predict biological heat generation. These are:

* Heat of combustion of the sludge;
* Chemical oxygen demand (COD) of the sludge; and
* The volatile solids (VS) of the sludge.

The first of these paramaters (Linton *et al.*, 1978; Hamer *et al.*, 1985), although theoretically correct, is not practically applicable in a normal wastewater laboratory. The determination of COD and VS reductions are within the normal scope of the average laboratory.

As sludge treatment and disposal regulations frequently define sludge stability in terms of a specified % VS reduction, VS has been most often used to define biological heat generation in ATAD (Andrews *et al.*, 1969; Jewell *et al.*, 1980;

Vismara, 1985; Booth and Tramontini, 1984; Wolinski, 1985). The use of COD utilization can be justified on theoretical and empirical bases (Servizi and Bogan, 1963; McCarty, 1972) and has also been frequently applied to predict heat release in ATAD (Andrews *et al.*, 1971; Gould *et al.*, 1978).

For the aerobic stage of the dual digestion process the requirements are different - the rapid generation of heat is the primary goal to achieve disinfection and pre-conditioning of the sludge, while destruction of biodegradable material is left to the second anaerobic stage. Also, in dual digestion the aerobic stage is likely to operate under oxygen limited conditions, which may mean that the ATAD heat generation/VS removal kinetics are not directly applicable.

To get a proper perspective of the requirements of the aerobic stage of dual digestion it is necessary to examine the basic concepts of aerobic digestion, as well as those of heat generation in the process. These aspects are covered in detail in Part 2, of this Final Report.

2.5 ANAEROBIC DIGESTION

Anaerobic digestion, which forms the second stage of the dual digestion process, is one of the oldest forms of wastewater sludge treatment. It is the biological degradation of complex organic substances in the absence of free oxygen with the release of energy and the conversion of much of the organic matter to methane, carbon dioxide and water. Since little carbon and energy remain to sustain further biological activity, the remaining solids are rendered stable.

The process is used extensively all over the world for treating wastewater sludges which have a volatile solids content of more than 50%, as well as for high strength organic wastes. The process has developed a reputation for being unstable and failures often occur. This was usually the result of the poor understanding of the process operating conditions and, till recently, poor design. Extensive research and development of the anaerobic digestion process has led to significant improvements in both reliability and performance, but good control and regular monitoring of the critical parameters such as load, gas production, volatile acids, alkalinity and pH, is essential for successful operation (USEPA Report, 1979).

2.5.1 Advantages and Disadvantages

Anaerobic digestion has several advantages over other sludge stabilization methods:

* *Methane Production*:

In most cases the process is a net energy producer, since the energy content of the digester gas exceeds the energy demand for heating and mixing. Surplus gas may be used for other purposes such as space heating or for fuelling gas engines (Ward, 1974; USEPA Reports, 1977, 1978).

* *Sludge Mass Reduction:*

Commonly from 25 to 45% of the raw sludge solids are destroyed and converted primarily to methane, carbon dioxide and water. This significantly reduces the problem and the cost of sludge disposal.

* *Stable Residual Solids:*

The residual solids in the treated sludge are stable, do not produce unpleasant odours or support fly breeding and can be used as a valuable soil conditioner.

* *Pathogen Reduction:*

Pathogenic organisms tend to die off during the relatively long retention times required to stabilize the sludge. Although significant reductions in the numbers of most pathogens occur, from one to four log reductions in various pathogenic bacteria and viruses (Moore *et al.*, 1976), helminth ova are considerably more resistant and *Ascaris lumbricoides* ova are reported to have survived thermophilic anaerobic digestion at 50°C (Stern *et al.*, 1977).

The principal disadvantages of anaerobic digestion are:

* *High Capital Cost*:

Owing to the rate of growth of the methane producing bacteria, it takes from 15 to 30 days to stabilize sludge. This slow growth rate limits the speed at which the process can adjust to changes in feed load, temperature and other environmental conditions (McCarty, 1964). Very large, fully enclosed digestion tanks are required, with mixing and heating facilities in the case of high rate digestion.

* *Process is susceptible to upsets*:

Anaerobic digestion is a complex fermentation process, but in simplified form it may be regarded as two distinct degradation phases viz. acid formation and methane production. During the first phase facultative bacteria convert the complex organic substrate to volatile short chain organic acids. In the second phase, strictly anaerobic bacteria (methanogens) convert the volatile acids to methane, carbon dioxide and water.

When an anaerobic digester is working properly, the two phases are in equilibrium, i.e. the volatile acids are converted to methane at the same rate that they are produced from the feed substrate. Volatile acid levels are low in a digester operating normally. The inherent problem in the process is that the methanogens are extremely slow-growing organisms, with a doubling time measured in days. They are also very sensitive and are adversely affected by fluctuations in pH, substrate concentration and temperature. On the other hand, the acid formers are extremely robust and can function effectively over a wide range of environmental conditions. In addition, they grow very rapidly, with a doubling time measured in

hours. Consequently, when a digester is subjected to overloading, temperature fluctuation or other stresses, the methanogenic activity decreases first and volatile acids cannot be converted to methane as rapidly as they are formed. The acids accumulate and the pH drops, further inhibiting the methanogenic activity, with the result that the digester fails rapidly unless corrective action is taken (USEPA, 1979).

* *Poor Quality Supernatant*

The supernatant produced after settling anaerobically digested sludge usually has a high COD, nitrogen and suspended solids concentration which can cause considerable extra loading on the liquid waste treatment processes. In nitrogen removal wastewater plants this could cause serious problems or lead to increased costs.

* *Poor Settling and Dewatering*

Anaerobically digested sludges do not exhibit good settling or dewatering characteristics (Mignone, 1977; Karr *et al.*, 1978; Smollen *et al.*, 1986). The sludge is often supersaturated with digester gas which, when it comes out of solution as small bubbles, causes flotation of the sludge with settling only occurring when these bubbles are released. The sludge contains a high proportion of very fine particles which are produced by both mixing and by the natural break-down of particle size during the biological decomposition of the substrate. These "fines" interfere with settling and dewatering. Particle size is considered to be the single most important factor influencing sludge dewaterability (Karr *et al.*, 1978). As the average particle size decreases, primarily caused by mixing or shear, the surface/volume ratio increases exponentially (Heukelekian *et al.*, 1958). Increased surface area means greater hydration and increased resistance to loss of water which results in poor dewaterability.

2.5.2 Operational Considerations and Treatment Capacity of Mesophilic Anaerobic Sludge Digesters

Anaerobic sewage sludge digesters are generally classified as low-rate or high-rate depending on their efficiency. High-rate digesters usually include features such as heating, auxiliary mixing, raw sludge thickening and uniform feeding. These features act together to create optimum conditions for the biological processes to ensure maximum stability with minimum retention time and consequently minimum volume requirement.

The second stage of the dual digestion process is usually a high-rate anaerobic digester, operating in the mesophilic temperature range (30 to 38°C). The high-rate process will therefore be discussed in this review.

2.5.2.1 Digester Load Rate

The digester volume required to treat sludge from a community has in the past been based on criteria such as per capita volume allowance, or later, on volatile solids loading rate. The former criterion was usually based on assumptions

regarding per capita waste load, solids removal efficiency in the treatment works and the digestibility of the sludge. The latter criterion, which specifies a reactor volume requirement for each unit of volatile solids (dry) in the feed per unit of time, has recently been commonly used. It, however, does not relate to the fundamental component of anaerobic digestion, the actual micro-organisms which break down and stabilize the sludge (USEPA 1979).

2.5.2.2 Solids Retention Time

In a completely mixed high-rate anaerobic digester, the active organisms are theoretically evenly distributed throughout the vessel. To maintain steady state operation, the organism growth must equal or exceed the loss of active organisms when digested sludge is removed, otherwise the population of bacteria in the digester declines and the process eventually fails. As it is not practical to monitor organism growth directly, their growth rate is determined by monitoring the rate at which substrate is reduced and end products are formed, i.e. the kinetics of the process. Extensive studies of the kinetics of anaerobic digestion have been carried out, and these have resulted in a better understanding of how to control the process to prevent the frequent failures that used to occur in the past (Pretorius, 1969; Lawrence, 1971; USEPA,1979).

The most important design and operating parameter in anaerobic digestion is the biological solids retention time (SRT), defined as the average time a unit of microbial mass is retained in the system (Lawrence *et al.*, 1970). In digesters without recycle, the SRT equals the hydraulic detention time. To keep the size of the digester vessel to the minimum or to utilize an existing digester to its maximum, the lowest feasible SRT is used. As the SRT is reduced and approaches a critical value (SRT_c) where process failure occurs, substrate breakdown deteriorates and methane production decreases. An increase in volatile acids, reduction in bicarbonate alkalinity, and a fall in pH, together with changes in gas production and composition, are signs of impending failure. It is therefore recommended that the design solids retention time (SRT_d), which provides a margin of safety, should be used in practice. McCarty (1964), recommends that SRT_d should be 2,5 times SRT_c. Process kinetics calculations predict a SRT_c of 4.2 days for sewage sludge at 35°C (O'Rourke, 1968). A pilot-scale study by Torpey (1955) confirms this value as he found that the performance of anaerobic sludge digesters operating at 37°C deteriorated rapidly below 5 days retention, and failed completely at an SRT of 2,6 days.

Several researchers have recommended ten days as a minimum solids retention time for high rate anaerobic digesters operating at about 35°C (Sawyer *et al.*, 1960; Malina *et al.*, 1968; Dague, 1968; Zablatzky *et al.*, 1971). There is, however, not much evidence available of continuous stable operation of full-scale mesophilic anaerobic sludge digesters at SRT's of 10 days. This is thought to be primarily due to the more inefficient mixing, with concomitant poor temperature distribution, in full-scale digesters (Thompson *et al.*, 1984).

2.5.2.3 Solids Reduction

Anaerobic digestion not only stabilizes the sludge but effects a large reduction in the amount of solids for ultimate disposal. It is usually assumed that a breakdown of the volatile portion of the sludge solids occurs and consequently a common measure of digester performance is the percentage of volatile solids destroyed. This may range from 40 to 65%, depending on the character of the sludge and the operating parameters of the digestion process (USEPA, 1979).

The maximum solids reduction depends largely on the characteristics of the raw sludge. The bacteria can only break down part of the volatile feed solids. Research to date suggests that 60 to 80% of volatile solids in municipal waste water sludge is readily biodegradable (Kountz *et al.*, 1959; McKinney, 1974). The remaining fraction comprises relatively inert organics such as lignins and tannins, which can be considered indigestible within the retention times normally associated with anaerobic digestion.

Research has indicated that during anaerobic digestion primary sludge degrades faster than a mixture of primary and waste activated sludge, which in turn degrades faster than straight waste activated sludge (USEPA 625/4-78-012, 1978).

The most important operating parameters affecting volatile solids reduction are the solids retention time and the temperature of digestion. VS reduction increases rapidly to 50 or 60% as the SRT is increased to 20 or 25 days, but beyond this point the rate falls off rapidly and further reduction is minimal (Pfeffer *et al.*, 1967; Hindin *et al.*, 1960). Raising the operating temperature to its optimum range (35 to 38°C) significantly increases the proportion of volatile solids destroyed. At higher SRTs this effect is less pronounced, but still significant. The response to temperature change is not instantaneous as the organisms require time to acclimatize to the new conditions.

Solids reduction and gas production are considered indicative parameters of the degree of stabilization achieved. However, Loll (1981), used the ratio of BOD_5 to COD as stabilization coefficient. He found that this ratio decreased most rapidly between 10 and 20 days, i.e. the most effective stabilization occurred during this period. The dewaterability of the sludge, as determined by the Capillary Suction Time (CST) test, also improved most rapidly between 10 and 20 days. Further improvement in both parameters occurred up to SRTs of 30 days.

Typical normal operating parameters for a well-functioning high-rate unit are given below (Ross, 1988):

Digester Load Rate	1,6 to 6,4 kg $VS.m^{-3}.d^{-1}$
Total Solids (TS) in feed sludge	3 to 8%
Volatile Solids (VS) in	
: feed sludge	70%
: digested sludge	45%

VS reduction in process	40 to 65%
Solids Retention Time (SRT)	15 to 25 days
Biogas production	0,4 to 0,7 m^3/kg VS destroyed
CO_2 concentration in gas	25 to 33%
Digestion Temperature	32 to 38°C
pH of settled digester liquor	6,8 to 7,2
Volatile Acids of digester liquor	<400 mg/L CH_3COOH
Total Alkalinity of digester liquor	>2000 mg/L as $CaCO_3$
Volatile Acids/Alkalinity Ratio	0,1 to 0,25

2.5.2.4 Buffer Capacity and pH

For stable operation it is important to maintain the digester pH between 6,4 and 7,5, as the methanogens are pH sensitive and operate best in this pH range (Heukelekian *et al.*, 1939; Barker, 1943; Mylroie *et al.*, 1954). In this pH range, the carbon dioxide-bicarbonate relationship is the most important controlling factor. The system is controlled by the CO_2 concentration of the gas phase and the bicarbonate alkalinity of the liquid phase. A digester with a given gas phase CO_2 concentration and liquid phase bicarbonate alkalinity can exist only at one specific pH. If the bicarbonate alkalinity is increased and the CO_2 remains the same, the pH must increase. Increasing the bicarbonate alkalinity raises the system buffering capacity, which stabilizes the pH and makes the system less prone to upset (McCarty, 1964; Brovko *et al.*, 1977; Pohland, 1968).

The decomposition of nitrogenous compounds, e.g. proteins, has an important bearing on buffer capacity. These compounds are reduced to ammonia, part of which is metabolized by the organisms, while the remainder combines with carbon dioxide to produce stable inorganic compounds which act as buffers, e.g. ammonium carbonate (Ross *et al.*, 1987).

2.5.2.5 Volatile Acids

Up to the early 1960's volatile acid levels in excess of 2000 mg/L were believed to be toxic to the anaerobic digestion process. There was also considerable controversy about the addition of alkaline substances to maintain buffer capacity. Carefully controlled studies in the early 1960's (McCarty and McKinney, 1961; McCarty and Brosseau, 1963) showed that volatile acid levels up to 6000 to 8000 mg/L were not toxic as such to the methanogens provided that adequate buffer capacity was available to maintain the system pH between 6,6 and 7,4.

They also showed that alkali addition to control pH was a valid procedure provided that the cation associated with the alkali did not cause toxicity. In general, it was safer to use magnesium or calcium compounds than those of ammonium, sodium or potassium.

2.6 DISINFECTION

The fundamental concept on which sludge disinfection is based is one of subjecting pathogenic organisms present in the sludge to hostile environmental conditions for a sufficient period to kill them. Exposure to elevated temperatures, such as is encountered during thermophilic digestion, enhances the death rates of most pathogenic organisms, but there is relatively little data available on the kinetics of organism death at the relatively low temperatures (i.e. typically below 70 °C) employed in most sludge treatment processes (Hamer, 1985; Hamer *et al.*, 1985).

2.6.1 Occurrence of pathogens in sewage

Pathogens are organisms or biological agents which cause disease. Most of the important pathogens found in human wastes will also be present in sewage. Many of the pathogens in sewage have little significance in the spreading of disease because their survival outside the body is limited, or the routes involving sewage, sludge and water do not enable the cycle of infection to be completed back to man. Several studies have demonstrated that conventional waste water treatment processes will remove many of these enteric micro-organisms from the liquid fraction, especially if the final effluent is disinfected prior to discharge. However, many organisms become associated with the separated solids, i.e. the sludge, resulting in high concentrations of these organisms in both the raw (primary) and secondary treatment sludges (Kabrick *et al.*, 1982; Pike *et al.*, 1988).

Table 2.1 gives the major pathogens found in sewage waters and the diseases associated with these pathogens (Burge and Marsh, 1978). In South Africa the most important helminths that occur commonly in sewage sludge are *Ascaris lumbricoides, Trichuris trichiura, Enterobius vermicularis, Hymenolepis nana* and *Taenia* spp.

Adult forms of helminths and protozoa do not persist in the environment outside their hosts. Their ova and cysts are, however, capable of surviving in the environment and are the forms which continue the infection cycle via sewage wastes (Burge, 1983).

Helminth ova become concentrated in sewage sludge owing to their higher specific gravity (Foster *et al.*, 1973; Reimers *et al.*, 1983). Reimers *et al.* (1983) showed that 91 to 98 % of *Ascaris* ova are removed from the liquid phase by secondary clarification and are concentrated in the sludge. Under favourable conditions these ova can survive for a number of years (Smith, 1977).

Enteric bacteria and viruses do not produce resistant forms. The micro-organisms are themselves found in sewage and sludges. *Salmonellae* appear to be the only pathogenic bacteria which are capable of sustained growth outside their hosts. Numerous cases of regrowth after disinfection treatment have been reported (Burge, 1983; Yeager and Ward 1981; Rus and Yanko 1981). In sludge and compost that has been sterilised the regrowth is rapid and extensive due to the elimination of competition from other organisms (Brandon *et al.*, 1977).

TABLE 2.1: Major pathogens found in sewage and diseases associated with these pathogens *(Burge and Marsh, 1978)*

Organism	Disease
Helminthic parasites (Intestinal worms)	
Ascaris lumbricoides	Ascariasis
Ancylostoma duodenale	Hookworm infection
Necator americanus	Hookworm infection
Taenia saginata (Beef tapeworm)	Taeniasis
Trichuris trichiura (Whipworm)	Trichuriasis
Protozoa	
Entamoeba histolytica	Amoebic dysentery
Bacteria	
Salmonella spp	Salmonellosis
Shigella spp	Shigellosis
Mycobacterium tuberculosis	Pulmonary tuberculosis
Viruses	
Poliovirus	Poliomyelitis
Coxsackievirus	Aseptic meningitis, gastroenteritis*
Echovirus	Aseptic meningitis**
Reovirus	Mild respiratory infection, gasteroenteritis
Rotavirus	Gastroenteritis
Norwalk	Gastroenteritis
	Acute respiratory infection***
Adenovirus	Infectious hepatitis
Hepatitis A virus	

* Two of the diseases caused by several serotypes of this virus. Diseases caused range from trivial to lethal.

** Diseases caused are similar to those of coxsackievirus.

*** Other diseases include pharyngitis and infant pneumonia.

Because of their affinity to bind to particulate matter in sewage (Lund, 1970; Lund *et al.*, 1973; Wellings *et al.*, 1976; Englande, 1983) most enteroviruses become a component of the sludge. Since human viruses are obligative intracellular parasites and cannot reproduce outside the host cell, the sludge does not provide a suitable environment for reproduction. Enteric viruses do not therefore regrow and will be inactivated at a rate that increases with temperature. Viruses encapsulated in particle aggregates are given protection against physical and chemical adversities in the environment and can therefore persist for extended periods outside the host cell.

Laboratory studies have shown that virus survival can range from 32 to 170 days in soil, depending on factors such as soil type, pH, organic content, moisture, temperature, etc. (Englande, 1983).

2.6.2 Criteria for Disinfection

The ongoing search for suitable criteria for establishing whether a sludge is "disinfected" has been fraught with problems. The most logical first approach is to monitor the degree of destruction of the pathogenic organisms themselves. Unfortunately this approach is impractical as the procedures involved are too difficult and costly for routine application.

Monitoring the inactivation of a specific pathogen has frequently been advocated and tried, but this method also has disadvantages. To be effective, the concentration of the chosen organism in the untreated sludge must be high and the organism must be resistant enough so that its destruction (i.e. its absence) in the disinfected sludge will guarantee the inactivation of other pathogens that were present in the untreated sludge.

The use of indicator organisms, typically faecal coliforms, is also not suitable. Although these organisms are destroyed during the disinfection process, reinfection and regrowth in the treated compost or sludge almost inevitably occurs (Burge, 1983). Doubts have also been expressed as to the suitability of coliforms as indicators, as experiments have shown that *Salmonellae* could still be isolated from "disinfected" samples in which coliforms were absent (Strauch *et al.*, 1984).

The detection of viruses also presents problems. The number of viruses isolated from waste water and sludge samples may be one or two logs (90 to 99%) lower than the actual concentrations due to the limitations of the virus recovery procedures (Englande, 1983). The use of a temperature-by-time monitoring system was proposed by Burge *et al.*, (1978), using the relatively heat resistant bacteriophage f2 as yardstick. The D-values of common enteric pathogens as well as that of f2 are shown in Table 2.2. Note that a D-value is the time required to effect a ten-fold, or one log, reduction in the population numbers. For aerated composting Burge (1983) decided to use a 15-log reduction of f2 as the disinfection standard, which translates to 2,5 days at 55°C or 0,9 days at 60°C. Using these criteria at the Dickerson sewage sludge composting site in Montgomery County, Maryland, has consistently yielded good results and no samples positive for *salmonellae* have been found (Burge, 1983).

Although it has been suggested that the resistance of *Salmonellae* to increased temperature varies with serotype (Kabrick and Jewell, 1982), the range of variability is not great and the variation may even be attributable to extraneous or experimental factors.

TABLE 2.2: Time required for a 10-fold reduction of enteric pathogens by heat as compared with that for f2 bacteriophage *(Burge et al., 1978)*

Organism	D Value (Minutes) 55 °C	60 °C
Adenovirus, f2 NIAID	11	0,17
Poliovirus, type 1	32	19
Ascaris ova	-	1,3
Histolytica cysts	44	25
Salmonella senftenberg (775W)	80	7,5
Bacteriophage, f2	267	47

One of the most environmentally resistant pathogenic organisms is the ovum of the roundworm *Ascaris* sp. In South Africa *Ascaris lumbricoides* occurs quite commonly and as its ova are highly resistant, their presence may be a useful indicator of the hygienic quality of the sludge (Trim, 1984). The ova may persist for up to seven years in soil, and up to 35 days on the surface of vegetables (Reimers *et al.*, 1983). Infection occurs when viable ova are ingested. A real danger therefore exists when eating raw vegetables which may have been contaminated by sewage sludge.

Various studies have shown that *Ascaris* ova are more temperature resistant than ova from other common helminths e.g. *Taenia saginata* (Pike *et al.*, 1983) and bacteria such as *Salmonellae* (Strauch *et al.*, 1984). Temperature/time studies on the inactivation of *Ascaris* ova have frequently been reported in the literature. Reyes *et al.*, (1963) reported complete inactivation by heating to 55 °C for 60 minutes, or to 45 °C for 20 days. Strauch *et al.*, (1984) found that 55 °C for 3 days would completely degenerate *Ascaris suum* ova, while Pike *et al.*, (1983) reported at least 99 % inactivation of the same organism in anaerobic digestion for 10 to 20 days at 49 °C. In a review paper, Hays (1977) reported that it was generally agreed that a temperature of 60 °C for 30 minutes or more was necessary in anaerobic digestion to destroy *Ascaris* ova. Reyes *et al.*, (1963), reported zero survival after 20 minutes at 55 to 60 °C in anaerobic digestion. Carrington (1984) showed that *Ascaris* ova added to sludge were not viable after exposure to temperatures in excess of 55 °C for periods of about 2 hours. Pike *et. al.* (1988) during laboratory studies found that the viability of *Ascaris* suum ova was destroyed during thermophilic digestion at 49 °C for 10 to 20 days, or by heating to 55 °C for 15 minutes. Trim (1984), during tests on a pilot scale ATAD plant, reported complete destruction of *Ascaris lumbricoides* ova at 60 °C and 3,5 day retention. At 1.8 day retention only 98 to 99 % destruction was achieved, while at 45 to 55 °C and one day retention the performance was considerably poorer owing to short-circuiting in the reactor.

Summarizing the criteria used to determine the disinfection achieved, it appears that:

(1) Monitoring the survival of the usual indicator organisms is not effective;

(2) Monitoring the fate of specific pathogens is in most cases difficult, costly and prone to error;

(3) Specifying the process conditions for the required degree of disinfection holds more promise. These minimum conditions can be determined experimentally, e.g. on a pilot scale process.

Effective disinfection can, however, only be assured if the operating procedures are strictly followed.

2.6.3 Standards or Guidelines for Sludge Disinfection

Many countries regulate the use of sewage-derived sludge for agricultural applications by enforced standards, others promulgate guidelines or recommendations for such use. Some of the regulations are very vague on what constitutes "hygienically safe" or "disinfected" sludge, while others specify acceptable processes, degree of stabilization, minimum microbiological standards, maximum limits for heavy metals and certain chemical agents as well as restrictions regarding the use of the treated sludge. In general, most of these regulations are being looked at more critically and many are in the process of being revised to be less ambiguous and more appropriate and enforceable.

To place the new South African "Guidelines for the Use of Sewage Sludge" (Appendix 1) in the right perspective, it is probably wise to look at existing and proposed new regulations in other countries which also control the use of sludge on agricultural land. The situation in the USA and the Federal Republic of Germany are discussed briefly (Heidman 1990).

2.6.3.1 Existing U.S. Regulations

Land application of municipal wastewater sludge is subject to the U.S. Federal Regulations which are defined in 40 CFR Part 257 "Criteria for Classification of Solid Waste Disposal Facilities and Practices" of 1979. These technology based regulations require the sludge to be treated by appropriate technologies prior to land application to prevent the creation of nuisance conditions or health hazards. The regulations define two categories of technologies in terms of their ability to reduce the pathogen content of sludge (USEPA, 1989):

(1) Processes to Significantly Reduce Pathogens (PSRPs);

(2) Processes to Further Reduce Pathogens (PFRPs).

Processes to Significantly Reduce Pathogens (PSRPs) must be capable of reducing wastewater sludge pathogens to levels which compare with a well run anaerobic digester. The following are classified as PSRPs:

* Aerobic digestion (detention time 60 days at 15°C to 40 days at 20°C, achieving minimum volatile solids reduction of 38%);
* Air drying (sludge depth 23 cm maximum, 3 month drying period with 2 months at temperatures greater than 0°C);
* Anaerobic digestion (detention time 60 days at 20°C to 15 days at 35 to 55°C, minimum volatile solids reduction 38%);
* Composting (minimum operating conditions 40°C for 5 days, temperature must exceed 55°C for at least 4 hours);
* Lime stabilization (pH of 12 after 2 hours contact);
* Other methods which reduce pathogens and volatile solids to an extent equivalent to the reduction achieved by any of the above methods.

Aerobic sludge treatment processes must achieve the following performance levels to be classified as PSRPs:

* A 2 $\log_{10}$ reduction in either:

  (i) Faecal coliform and faecal streptococci; or
  (ii) Faecal coliform and enterococci;

* A 38% reduction in volatile suspended solids (VSS):

* A Specific Oxygen Uptake Rate (SOUR) of 1mg O_2 per hour per g TSS.

Processes to Further Reduce Pathogens (PFRPs) are assumed to be capable of reducing pathogens below detectable levels. The following are classified as PFRPs:

Composting: Within-vessel and static aerated pile methods: Temperature of 55°C or more for 3 days;

Windrow method: Temperature of 55°C or more for at least 15 days during composting period, with a minimum of 5 turnings of the windrow;

Heat Drying: Dewatered sludge dried to reduce moisture to 10% or lower. Temperature to exceed 80°C;

Heat Treatment: Liquid sludge heated to 180°C for 30 minutes;

Thermophilic Aerobic Digestion: Aerobic conditions to be maintained for 10 days at 55°C to 60°C. A volatile solids reduction of at least 38% must be achieved;

Other Methods: Acceptable if equivalent pathogen and volatile solids reductions to the above treatment methods are achieved.

The processes listed below, if added to a PSRP:

Beta Ray Irradiation: Sludge irradiated at dosages of at least 1,0 megarad at room temperature;

Gamma Ray Irradiation: Sludge irradiated at dosages of at least 1,0 megarad at room temperature;

Pasteurization: Sludge maintained at a minimum temperature of 70°C for at least 30 minutes;

Other Methods: Acceptable if equivalent pathogen reduction to the above add-on methods is achieved.

To be classified as PFRPs, aerobic sludge treatment processes must attain the following performance levels:

* Pathogen criteria - maximum in a 100 mL sludge sample at 5% solids:

  3 MPN Salmonella spp.;
  1 PFU total enteroviruses;
  1 viable *Ascaris* spp. ovum;
  1 viable *Toxocara* spp. ovum;
  1 viable *Trichurus trichiura* ovum.

* A 38% reduction in volatile suspended solids (VSS).

* A Specific Oxygen Uptake Rate (SOUR) of less than 1 mg O_2 per hour per g TSS.

The use of sludge produced by PSRP processes is restricted owing to their reduced ability to remove pathogens, whereas PFRP processed sludge is at present unrestricted with regard to pathogen criteria. There are no performance monitoring criteria relating to pathogen reduction specified in the present Federal regulations, so it is not necessary to test the treated sludge for pathogen content to confirm that the theoretical performance level is being achieved.

2.6.3.2 Proposed New U.S. Regulations

The proposed new standards of February, 1989, which will most likely replace the existing Part 257 regulations, include performance based standards for sludge quality. Routine sampling and analysis of the sludge produced will be required.

The proposed standards will allow for three classifications, viz. A, B and C. Class A standards correspond to PFRP performance levels, while Class B standards correspond to PSRP levels. The Class C standards correspond to PSRP levels for sewage treatment systems without primary clarification and with an extended sludge residence time.

CLASS A

One of the following two groups of standards must be met to attain Class A:

Group 1

≤ 3 *Salmonella* spp. per g VSS;
≤ 1 PFU virus per g VSS;
≤ 1 Protozoan organism per g VSS;
≤ 1 Helminth ovum per g VSS.

Group 2

The following time and temperature conditions must be met:

t_{min} ≤ 120 hours at 53°C;
t_{min} ≤ 72 hours at 55°C;
t_{min} ≤ 0,5 hour at 70°C.

Concentrations of indicator organisms in treated sludge must be:

≤ 2 $\log_{10}$ faecal coliform per g VSS;
≤ 2 $\log_{10}$ faecal streptococci (enterococci) per g VSS.

CLASS B

Similarly, one of the following groups of standards must be met to achieve Class B:

Group 1

2 $\log_{10}$ reduction of *Salmonella* spp. per g of wastewater plant influent VSS;
2 $\log_{10}$ reduction of viruses per g waste water plant influent VSS.

Group 2

Concentrations of indicator organisms in treated sludge must be:

≤ 2 $\log_{10}$ for *Salmonella* spp. per g VSS;
≤ 2 $\log_{10}$ for viruses per g VSS.

CLASS C

One of the following groups of standards must be met:

Group 1

1,5 $\log_{10}$ reduction of *Salmonella* spp.waste water plant influent VSS;
1,5 $\log_{10}$ reduction of viruses per g of waste water plant influent VSS.

Group 2

Concentration of indicator organisms in the treated sludge must be:

$\leq$ 6,3 $\log_{10}$ faecal coliform per g VSS;
$\leq$ 6,7 $\log_{10}$ faecal streptococci (enterococci) per g VSS.

In addition all three classes of treated sludge must reduce the potential for vector (i.e. disease vectors such as flies, mosquitos, rodents, etc.) attraction by meeting at least one of several other criteria. For aerobic processes, these include:

* a 38% VSS reduction;
* a Specific Oxygen Uptake Rate (SOUR) $\leq$1 mg O_2/h/g TS;
* TSS >75%;
* treated sewage sludge is subsurface injected.

The use of Class B and Class C sludges is subject to application restrictions with regard to crop types, foraging, public access and time, these being more severe for Class C owing to its higher potential pathogen content.

2.6.3.3 Existing German Federal and State Regulations

The Ordinance on Sewage Sludge (OSS) of 25 June, 1982, enacted on 1 April, 1983, which controls the use of sewage sludge for agricultural application, only permits the utilization of "hygienically safe" sludge for such purposes. This Ordinance defines a "hygienically safe" sludge as one from which all pathogenic organisms have been eliminated by appropriate technologies. The types of pathogens are not specified, neither are the technologies which are considered to be appropriate. Although the details are vague, the intent was clear, i.e. to prevent the introduction of pathogens into the food chain. Only "disinfected" sludge was deemed fit for utilization in vegetable and fruit cultivation, and on pasture and forage land. Other applications, not related to the production of food or forage, such as ploughed fields and forests, can use sewage sludge independent of its hygienic condition.

Owing to the vague terms "disinfected" and "hygienically safe" the regulations have been criticized by the scientific and engineering community, and modifications to the OSS is likely to be introduced in the near future.

2.6.3.4 Recommendations for Future Changes to the OSS

A joint working group comprising engineers, human and veterinary experts of the Abwassertechnische Vereinigung (ATV) and the Vereinigung Kommunaler Städtereinigungsbetriebe (VKS) has proposed a definition of "disinfected" and has

made recommendations regarding the "appropriate technologies" to achieve the required disinfection. These proposals will most likely be incorporated into an administrative regulation for the "Disinfection of Sludge" and become legally binding in the near future.

The proposals state that a sewage sludge can be defined as "disinfected" if it conforms to the following three-stage control concept:

1. **Treatment Process Qualification**

The sewage sludge must have been treated with an appropriate technology. Processes which have pre-qualified through approved studies are deemed acceptable. They must have been shown to be capable of:

* reducing the number of indigenous or added *Salmonellae* by at least 10^4;
* rendering indigenous or added eggs of *Ascaris* non-infectious.

2. **Treated Sludge Performance Criteria**

Directly after treatment the sludge shall contain:

* no *Salmonellae* per g of treated sludge;
* not more than 1000 *Enterobacteriaceae* per g of treated sludge.

3. **Process Specific Operational Control Criteria**

Operational control criteria must be established specifically for each process, defining the boundary conditions which will ensure disinfection as required under item 2 above.

The operational control criteria which may be applicable to the dual digestion process are:

(i) Autothermal Thermophilic Aerobic Digestion (ATAD).

* At least two stages must be used to avoid short-circuiting. The undisturbed reaction time, t_{min}, for a given temperature T must be as follows:

  $t_{min} \geq 23$ h at T = 50°C
  $t_{min} \geq 10$ h at T = 55°C
  $t_{min} \geq 4$ h at T = 60°C

  These temperature requirements take into consideration the temperature drop caused by batch feeding occurring once per day.

* The total detention time must be at least 5 days, assuming the use of two equal size vessels.

Operational controls include:

* continuous temperature monitoring and recording of each vessel in at least two locations;
* measurement of raw and final sludge pH values;
* continuous on-line monitoring of sludge feed volumes to enable accurate calculation and control of detention time.

(ii) Aerated Thermophilic Digestion Prestage before anaerobic digestion (PREST).

This process is essentially similar to the dual digestion (DD) process, except that it is specified that the temperature in the aerobic stage of the PREST process has to be maintained with additional heating. The temperature/time requirements for the aerobic stage are the same as for pre-pasteurization:

$t_{min} \geq 10$ minutes at T = 80°C
$t_{min} \geq 20$ minutes at T = 75°C
$t_{min} \geq 25$ minutes at T = 70°C
$t_{min} \geq 30$ minutes at T = 65°C

The following parameters have to be monitored to ensure proper control of the process:

* The undisturbed reaction time (contact time) t_{min};
* The temperature in at least two locations, in both the aerobic and anaerobic stages;
* The temperature of the anaerobic stage has to be kept at more than 30°C to maintain disinfection.

2.6.3.5 South African Guidelines for the use of sewage sludge

According to section 20 of the Health Act (Act 63 of 1977), every local authority shall take all steps necessary to prevent nuisance, unhygienic or offensive conditions from occurring in their district, which could affect the population detrimentally, and must correct such conditions if they occur. Clearly this Act also includes sewage and sewage sludge treatment and disposal.

The large local authorities are generally well equipped and capable of complying with the conditions of the Act, but the smaller local authorities often need proper guidance to cope with problems such as, for example, sewage sludge management.

Although the Minister of Health is empowered to pass regulations to enforce certain conditions in the Act, this has not been done, but Guidelines have been drafted to help local authorities to comply with the Act.

The first set of Guidelines for the Utilization of Sludge in South Africa, presented by Oberholster (1983) does not permit the use of raw sludge for any agricultural applications. They provided clear guidance as to the degree of treatment required before the sludge could be applied for agriculture or horticulture. The guidelines relate principally to the hygienic quality of the sludge and do not take into account the effects of heavy metals, toxic chemicals etc. on the environment. Great emphasis was placed on the control of *Ascaris lumbricoides*.

Secondary (digested) sludge may be used for agriculture, but not on bulb-type, tuberous or low growing vegetables or fruits which could be directly contaminated. The use on lawns, sports fields or parks is prohibited, except during initial development or planting. Forage crops for animals may not use secondary sludge, but crops not eaten raw by humans, for example sugar cane, may use digested sludge.

Tertiary sludge, defined as secondary sludge matured on drying beds for more than 90 days; raw, primary or secondary sludge composted according to accepted criteria at 50 to 65 °C, or sludge pasteurized at less than 80 °C, may be used on vegetable crops if pathogen free. The term "pathogen free" specifies zero viable *E.coli, Ascaris lumbricoides* ova or pathogenic viruses per 100 g sample of sludge. If well mixed with soil, it may be used unrestricted on other crops, but not as lawn top dressing.

Sludge which has undergone advanced treatment, for example by irradiation or high temperature treatment (150 - 230 °C), may be used without any restriction.

New guidelines are being prepared and have been presented for comment (Vivier *et al.*, 1988). These Guidelines attempt to limit the effects of the use of sludge on human health and the environment to acceptable levels, imposing only essential requirements and conditions, with limited control determinations, but at the same time to allow maximum beneficial use of all types of sewage sludge. Heavy metal restrictions are included, but at the stage of initial publication no definite restrictions for hazardous organic chemicals were incorporated, but it was mentioned that sludge producers should be prepared to undertake the necessary measures when waste water from chemical industry is discharged into their sewers.

The guidelines consist of the classification of various sewage sludges and applications of such sludges. For more details, see Appendix 1. In brief, sewage sludge is classified into three basic types A, B and C in a decreasing order of its potential to cause odour or fly-breeding nuisance as well as to transmit pathogenic organisms to man and his environment, as follows:

* Type A: Unstable with high odour and fly nuisance potential, having high content of pathogens.

* Type B: Stable with low odour and fly nuisance potential, reduced content of pathogens.

* Type C: Stable with no odour or fly nuisance potential, containing insignificant numbers of pathogens.

A further category, Type D, is of similar hygienic quality as Type C but since it may be used without restriction on land at a maximum application rate of 8 dry tonne per hectare per year, it has a metal content restriction.

The guidelines recognize the fact that it is impractical to analyze for a wide range of pathogens, so it concentrates on the numbers of faecal coliforms, *Salmonellae* and *Ascaris* ova as additional hygienic quality requirements for sludge Types C and D.

As the origins/acceptable treatment and the hygienic requirements of Types C and D are mainly of interest for the purpose of this investigation, they are listed below:

ORIGIN - TREATMENT (EXAMPLES)

- Pasteurized sludge
- Heat treated sludge
- Lime stabilized sludge
- Composted sludge
- Irradiated sludge
- Fumigated sludge

CHARACTERISTICS / QUALITY STANDARD

- Stabilized - should not cause odour nuisances or fly breeding.
- Contain no viable *Ascaris* ova per 10 g of dry sludge.
- Maximum of 1 *Salmonella* organism per 100 g dry sludge.
- Maximum of 10 faecal coliforms per 100 g dry sludge immediately after treatment (disinfection/sterilization).

The sludge is required to be certified to comply with the above requirements, otherwise it is considered a Type B sludge.

Note that the treatment process parameters are not specified, nor are the sampling procedures or frequency. It is assumed that the authority will implement suitable control measures. The success of implementation of these guidelines relies heavily on the goodwill and conscientiousness of the authority, who may at their discretion enforce these guidelines in, for example, local bye-laws.

2.7 DUAL DIGESTION

2.7.1 Historical Development

The dual digestion process was developed in 1975 at the Tonawanda Laboratories of the Union Carbide Corporation. Bench and pilot scale testing of the dual digestion system showed that autothermal heat generation in the aerobic stage was feasible and that the combined aerobic-anaerobic system could attain a degree of stabilization comparable with conventional processes (Drnevich *et al.*, 1978;

Castillo *et al.*, 1979). This led to a full-scale EPA supported demonstration of the process at the Hagerstown Wastewater Treatment Plant, Maryland, during 1980 (Haas, 1980). In the USA further dual digestion systems have since been installed at the Nutbush Creek Wastewater Treatment Plant, Henderson, North Carolina, and at the Lackawanna Sewage Treatment Plant, Lackawanna, New York. A second aerobic reactor has also been installed at the Hagerstown plant to double its capacity as part of an overall plant expansion. Three further plants were also under construction by 1986 (Appleton and Venosa, 1986).

In Europe there was also considerable interest in developing processes that would stabilize and disinfect sewage sludge. Zwiefelhofer (1983) recognized the benefits of combining the autothermal aerobic thermophilic hygienization with subsequent anaerobic mesophilic digestion to achieve this goal. A full-scale operating process using this concept was described by him. Pilot plant investigations on a similar system was carried out by Keller and Berninger (1984).

As a consequence of Zwiefelhofer's work the UTB Aerotherm system was developed to aerobically pre-treat and disinfect the sludge, using air, before passing it to a mesophilic anaerobic digester. More than 70 UTB Aerotherm plants with aerobic thermophilic pretreatment first stages are in operation or under construction in Europe. Performance data for periods of up to 6 years is available from more than 30 plants (Baier and Zwiefelhofer, 1991).

It is noteworthy that the American dual digestion plants all use pure oxygen from pressure swing absorption (PSA) oxygen generation systems for the aerobic stage. The oxygen is introduced via coarse bubble diffusers and a mechanical shear arm mixer is used to distribute and promote oxygen dissolution. The Swiss plants are more sophisticated and probably more efficient, employing an external recirculation loop with a venturi for dissolving the air and distributing the aerated sludge throughout the reactor. The plants are fully automated and conserve heat effectively by using batch heat exchangers and good all-round insulation.

2.7.2 Initial Bench and Pilot Scale Evaluation

The initial bench and pilot scale tests on the new concept of dual digestion at the Tonawanda Laboratories of Union Carbide Corporation is described by Drnevich and Matsch (1978). Bench tests were performed in 14-litre fermentors using various combinations of retention times for the aerobic and anaerobic units. The results are summarized in Table 2.3.

The results obtained were very encouraging with stable operation at anaerobic retention times as low as three days. It also showed that the aerobic stage was a net producer of alkalinity, which was a consequence of the deamination of amino acids which led to the production of ammonium bicarbonate. The additional alkalinity enhanced the stability of the anaerobic digester.

Pilot scale tests were carried out using a 190 litre insulated aerobic reactor followed by a 380 litre anaerobic digester. Pure oxygen was used in the aerobic stage and the units were fully mixed by using draft tube mixers. No external heat was added. A summary of the results obtained appears in Table 2.4.

TABLE 2.3: Aerobic/Anaerobic Digestion Data (Bench Scale)

Phase			I	II	III
Retention Time	Aerobic	Days	1,0	1,0	1,0
	Anaerobic	Days	3,0	4,0	7,0
TSS Feed		mg/L	34700	31700	27800
VSS Feed		mg/L	27800	25900	21200
Temperature	Feed	°C	20	20	20
	Aerobic	°C	50	50	50
	Anaerobic	°C	35	35	35
pH	Feed		6,4	7,2	6,9
	Aerobic		8,2	8,5	8,0
	Anaerobic		7,0	7,2	7,0
VSS Reduction	Aerobic	%	14	14	15
	Anaerobic	%	16	23	31
	Overall	%	28	34	41
Loading	Aerobic	VSS/d.ft^3	1,7	1,6	1,3
	Anaerobic	VSS/d.ft^3	0,49	0,35	0,16
	Overall	VSS/d.ft^3	0,43	0,32	0,16

TABLE 2.4: Aerobic/Anaerobic Digestion Data (Pilot Scale)

Phase			Start-up	Steady-state
Retention Time	Aerobic	Days	1,6	1,0
	Anaerobic	Days	10,4	8,7
TSS Feed		%	4,39	4,89
VSS Feed		%	2,85	2,98
Temperature	Feed	°C	22	18,2
	Aerobic	°C	49	50,2
	Anaerobic	°C	31	30,5
pH	Feed		6,4	6,9
	Aerobic		7,5	7,4
	Anaerobic		7,0	7,0
VSS Reduction	Aerobic	%	16	9,5
	Overall	%	45	44
Loading	Aerobic	VSS/d.ft^3	1,1	1,7
	Anaerobic	VSS/d.ft^3	0,14	0,17
	Overall	VSS/d.ft^3	0,15	0,18
Anaerobic gas production per overall VSS reduction		ft^3/Lb	13,6	13,9

Further pilot studies using a 380 litre aerobic reactor and a 1500 litre anaerobic digester, run in parallel with a 1500 litre standard anaerobic digester to act as a control, produced further encouraging results (Castillo *et al.*, 1979).

2.7.3 Full-scale Demonstration of Dual Digestion

The dual digestion technology was demonstrated at full-scale at the Hagerstown Wastewater Treatment Plant in 1980 (Haas, 1981) where primary and secondary sludge from the oxygen activated sludge plant was treated. The design detention times for the aerobic and anaerobic stages were 0,9 and 12 days respectively. Dissolved air flotation was used for pre-thickening the raw sludge, and plate-and-frame press to dewater the product.

The layout of the Hagerstown plant is shown in Figure 2.1. Note that the anaerobic digester was initially operated half-full to obtain the design retention time. During 1984 the aerobic stage was doubled, and the full digester operating volume utilized.

During the four month demonstration period the aerobic reactor achieved an average temperature of 50°C. After the demonstration period higher temperatures were reached in spite of the colder weather. This was attributed to the increase in feed volatile solids. Heat losses were low, owing to the low flow of vent gas from the reactor with subsequent low evaporative cooling effect. Oxygen utilization efficiency was reported as being good at 65% at a mixer speed of 32 r.p.m.

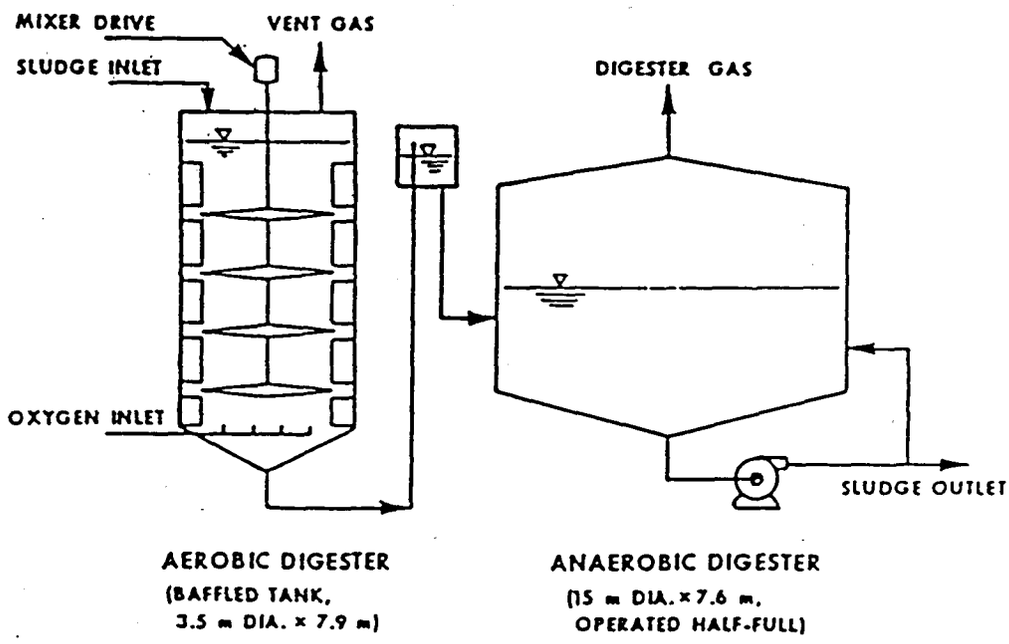

FIGURE 2.1: Dual Digestion at Hagerstown, Maryland

The temperature of the aerobic reactor was accurately controlled at temperatures up to 55°C by adjusting the oxygen flow rate. Problems were encountered in controlling the anaerobic digester temperature, which depended entirely on the aerobic sludge temperature and the heat losses in the digester. The residence time in the digester could also not be properly controlled and accurate data was not obtained for this unit.

One of the main features claimed for the process was its ability to produce a disinfected sludge. The initial results showed that *Salmonella* sp. and viruses was eliminated in the aerobically treated sludge, but that *Salmonella* sp. were present in the anaerobic sludge. This was attributed to either slow wash-out of these organisms from the digester which had previously operated on raw sludge, or to regrowth in the digester. Total coliform, Faecal coliform and Faecal streptococcus reductions of 3,6; 2,6 and 1,2 log were reported for the overall process (Farrell, 1984).

2.7.4 Evaluation of the Process (USA)

A technology evaluation of the dual digestion system as well as a study on the pathogen and indicator organism destruction by the process, which formed part of the Innovative/Alternative Technology evaluation programme of the EPA, was undertaken and the results published during 1986 (Appleton and Venosa, 1986; Appleton, Leong and Venosa, 1986).

Performance data from the Hagerstown demonstration are shown in Table 2.5 (Farrell, 1984).

TABLE 2.5: Summarized Performance of Dual Digestion System at Hagerstown, Maryland (1980)

	Demonstration period	Post Demonstration Period (weekly averages)			
	23/6 - 2/11	24/11/80	8/12/80	22/12/80	5/1/81
Temperature (°C)					
Feed	23,8	14,5	15,3	10,9	8,2
Aerobic effluent	50,5	55,3	53,6	55,6	55,4
Anaerobic effluent	41,7				
Ambient air	21,5	3,8	6,7	-3,2	-7,0
Residence time (d)					
Aerobic stage	1,38	0,97	0,73	0,68	0,53
Anaerobic stage	19,9				
Input solids (%)					
Total solids	3,02	3,79			
Volatile solids	2,41	3,16*			
pH					
Feed	6,0	5,7			
Aerobic effluent	7,1	7,2			
Anaerobic effluent	7,3	7,3			
Vol solids removal (%)					
Aerobic stage	27,2	25,5			
Overall	41,9				

* estimated

The project comprised two parts:

(i) post-construction evaluation at all the municipal wastewater treatment facilities using dual digestion; and
(ii) a performance study at one of the facilities with special emphasis on the degree of pathogen destruction.

The latter aspect was important because the options for the ultimate disposal of the stabilized sludge depend on the degree of disinfection achieved. The post-construction evaluation was carried out at the Hagerstown, Henderson and Lackawanna plants, while the performance study was undertaken at the Lackawanna plant. The dual digestion process was funded at an increased level at these plants through the EPA Innovative/Alternative Technology programme and its classification as a process to significantly reduce pathogens (PSRP) or as a process to further reduce pathogens (PFRP) depended to a large extent on the results of this evaluation. (These classifications are detailed in Section 2.6 Disinfection).

The most significant conclusions reached at the completion of the study were:

* The dual digestion technology is a promising process alternative for sludge stabilization and is especially attractive to those plants using pure oxygen activated sludge.
* Adequate volatile solids reduction could be achieved relative to equivalent performance PSRP's and PFRP's (i.e. >38%).
* The pH and alkalinity increase in the aerobic stage conditions the sludge for subsequent anaerobic digestion. The anaerobic digester maintains optimal pH without any chemical addition.
* Although the temperature in the aerobic stage often reached or exceeded the thermophilic operating goal of 55°C, the temperature varied sufficiently daily so that the mean reactor temperatures were less than desired. Reductions of the three indicator organisms viz. total coliform, faecal coliform and faecal *streptococci* were greater than those achieved by PSRP's and greater than or equal to those achieved by thermophilic aerobic digestion which is classified as a PFRP. *Salmonella* sp. data were found to be difficult to compare, but it appeared that better reductions could be achieved with PSRP's.
* Treatment plant staff reported the dual digestion system to be easy to operate and did not require significant labour beyond that for anaerobic digestion.
* Temperatures in the aerobic stage were effectively controlled by varying the pure oxygen feed rate. Temperature control in the anaerobic digester was more erratic and large variations occurred and the optimum mesophilic temperature was not consistently maintained.
* Mechanical reliability problems occurred. The mass transfer device (i.e. the mixing device) in the aerobic reactor was prone to various failures, which entailed shutting down the plant.

* Variable degrees of performance were noted at one of the plants assessed. Although temperatures reached the target thermophilic levels and solids reductions were adequate, consistent long-term performance was not demonstrated.

Other noteworthy points mentioned in the evaluation report are:

* The pre-thickening of the raw sludge is important. A total solids concentration of at least 5% is recommended.
* Accurate liquid level control in the aerobic reactor is important.
* Off-gas odours from the aerobic stage may cause problems. A simple granular activated carbon filter worked well on the plants evaluated.
* At the three plants evaluated the aerobic stage was a retrofit into the existing sludge treatment system, which used two-stage anaerobic digestion. The former second-stage anaerobic digesters were used to provide additional anaerobic digestion and solids/solution separation. A floating cover was fitted to this digester to provide storage for the biogas generated in the anaerobic digester.
* Sludge dewaterability following dual digestion treatment could not be quantitatively evaluated, but the minimum sludge concentration requirement for landfill (20% total solids) was met at the Lackawanna plant.

2.7.5 European Dual Digestion Plants

2.7.5.1 Process Description

Although the basic dual digestion process developed and marketed by UTB in Switzerland is the same as the process in the USA, there are some significant differences. The UTB process comprises the following treatment steps (Baier and Zwiefelhofer, 1991):

* Pre-thickening of primary and waste activated sludge, either combined or separately. Small thickening tanks are used, residence times being 10 to 20 hours. No mixing or additives are used.
* Aerobic thermophilic pre-treatment in insulated cylindrical steel tanks with a height to diameter ratio of 2:1. A pump recirculates the sludge through an external loop fitted with a venturi where air and recycled vent gas is introduced. The recycled vent gas from the reactor is partially recycled to increase the oxygen utilization and to minimize vent gas heat losses. The thermophilic reactor operates as a fully mixed unit and is run in a batch-fed mode. From 5 to 20% of the reactor volume is withdrawn periodically and replaced by an equal volume of raw thickened sludge. Residence time varies from 18 to 24 hours and the reactor temperature is controlled in the range of 60 to 68°C. To pre-heat the raw feed sludge a sludge to sludge heat exchanger is used which recovers heat from the hot sludge before introducing it into the anaerobic digester. The digester temperature is controlled by the sludge temperature leaving the exchanger.

* Anaerobic mesophilic digestion in cylindrical vessels with a height to diameter ratio of 1:2. The digestion temperature is between 34 and 40°C. Mixing is by external pumps, internal draft tubes, internal propellers or by biogas recirculation. The biogas produced is used for external applications, often to generate electricity by driving a gas engine equipped with exhaust gas heat recovery, with the recovered heat being utilized in the treatment process.
* Storage and consolidation of the treated sludge. Thickening of the digested sludge occurs in large consolidation and storage tanks equipped with mixing facilities. The mean hydraulic residence time is from 1 week to 6 months, the latter value applying during winter. Consolidation and storage tanks are run on a fill-and-draw mode or with continuous supernatant discharge.

The most significant differences from the original American dual digestion plants are:

* Use of air for oxygenation of the first stage;
* Provision of sufficient head-space in the reactor for foam. Also the provision of equipment (foam cutters) to cope with foam generation associated with using air;
* Conservation of heat throughout the plant. The use of high quality insulation and heat exchangers limit losses. Owing to the cold climate the aerobic stage is usually indoors.

  Vent gas from the aerobic reactor, which is a major source of heat loss, is partially recirculated. Cooling water from stationary gas engines powered by biogas from the anaerobic digester is in some cases used to heat the recirculation loop on the aerobic reactor.

* Extensive automation throughout to limit labour requirements.

2.7.5.2 Plant performance

Performance data for periods of operation for up to six years are available for 30 or more plants in Europe (Baier *et al.*, 1991). Details of results obtained at a small plant (2500 inhabitants) at Unterterzen, Switzerland, was reported by Hamer and Zwiefelhofer, (1986), while Mechsner (1986), produced a report on the performance of the plants at Thun, Wartau, Schöftland and Münsterlingen.

Most of the results reported show that the degradation of volatile solids was about 25% better with the dual digestion process than with the original mesophilic anaerobic digestion. The final product contained less organic matter and was thus more stable. It was also reported that in spite of the destruction of 5 to 14% of the raw sludge volatile solids in the aerobic stage, the biogas production was as high as that obtained by straight anaerobic digestion.

The survey by Baier *et al.*, (1991) showed that at treatment plants where sludge thickening procedures, using static thickeners, had not changed after introducing an aerobic stage ahead of the anaerobic digester, an increase of from 15 to 40% (as dry solids) was achieved. There was also a considerable improvement reported in the dewaterability of the treated sludge on belt filter presses. A direct comparative study at one plant showed an increase in dry solids from 27% to 37% after the addition of an aerobic thermophilic stage ahead of the digester. In other plants an increase of 8 to 10% (DS) was found, using identical dosages of polymers before and after implementing dual digestion.

Data on the disinfection of the treated sludge shows that the process is very effective. Enterobacteria are reduced by a factor of 10^7 to 10^8. In completely mixed digesters following aerobic thermophilic treatment the level of Enterobacteria detected was found to be from 1 to 50/g of wet sludge. These organisms have a very low regrowth potential in dual digested sludge so the concentration stays low even if the sludge remains stored for up to a year. *Salmonellae* have never been found in the dual digested sludge. Parasitological investigations on dual digested sludge samples have failed to detect any viable worm eggs (Baier *et al.*, 1991).

2.7.6 Concluding Comments

In general, the European dual digestion plants, most of which are upgraded mesophilic anaerobic plants, appear to work very well. No serious maintenance problems have been reported in the literature and the operators comment favourably on the ease of operation. The results reported indicate that the process effectively stabilizes and disinfects sludge to comply with the Swiss and German regulations for treatment of sludge to be applied to land.

CHAPTER 3

PLANT DETAILS

3.1 LOCATION OF PLANT

The dual digestion plant is located at Milnerton's Potsdam Waste Water Treatment Works, which serves a population of 80 000 and treats an average dry weather flow of approximately 18 ML/d. The process train comprises screening and grit removal, 7 primary sedimentation tanks, 8 biofilters followed by 8 humus sedimentation tanks and finally a series of maturation ponds. The humus tank sludges are returned to the primary sedimentation inlet channels and removed with the primary sludge. The primary sedimentation tanks are desludged at about 5 hourly intervals, the sludge being discharged to a central collection sump. The total daily sludge production amounts to approximately 150 m^3, but it is extremely difficult to estimate the relative proportions of humus and primary sludge. All the sludge was treated by a Zimpro plant until the dual digestion plant was commissioned. The latter now treats from 36 to 45 m^3 of the total sludge flow daily.

A layout of the treatment works is shown in Figure 3.1, indicating the sludge flows and the position of the dual digestion vessels in relation to the other units.

3.2 DUAL DIGESTION PLANT EQUIPMENT

The dual digestion plant utilised available equipment and structures where possible. The plant comprised three tanks in series. All three vessels, i.e. the thickening tank, the aerobic reactor and the anaerobic digester were available on site. A large capacity macerator pump was also available and was employed as the aerobic reactor feed pump.

A schematic layout of the dual digestion plant is shown in Fig 3.2.

The details of the equipment are as follows:

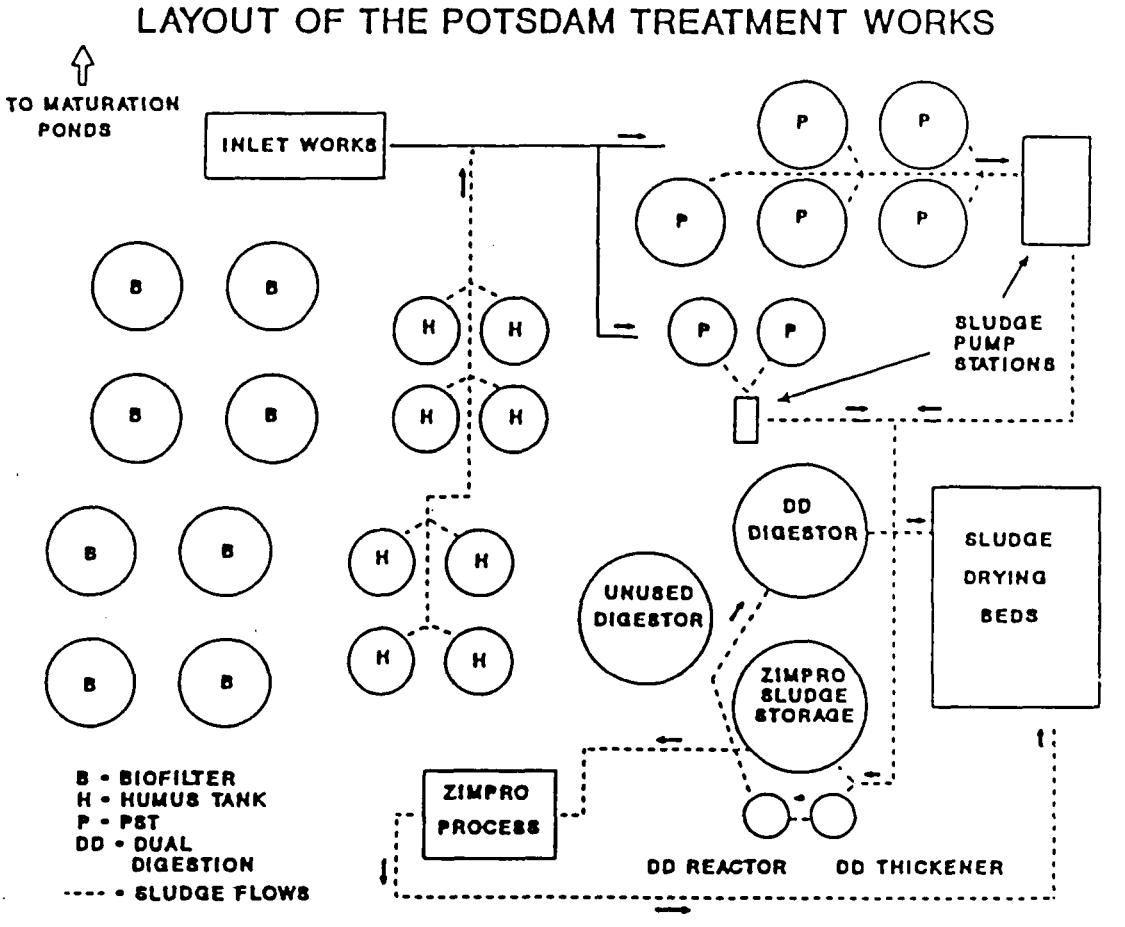

FIGURE 3.1: A schematic layout of the Potsdam waste water treatment works showing the integration of the dual digestion research plant with the treatment facility. Note that only sludge flows are depicted in the illustration - flows of sewage are not shown.

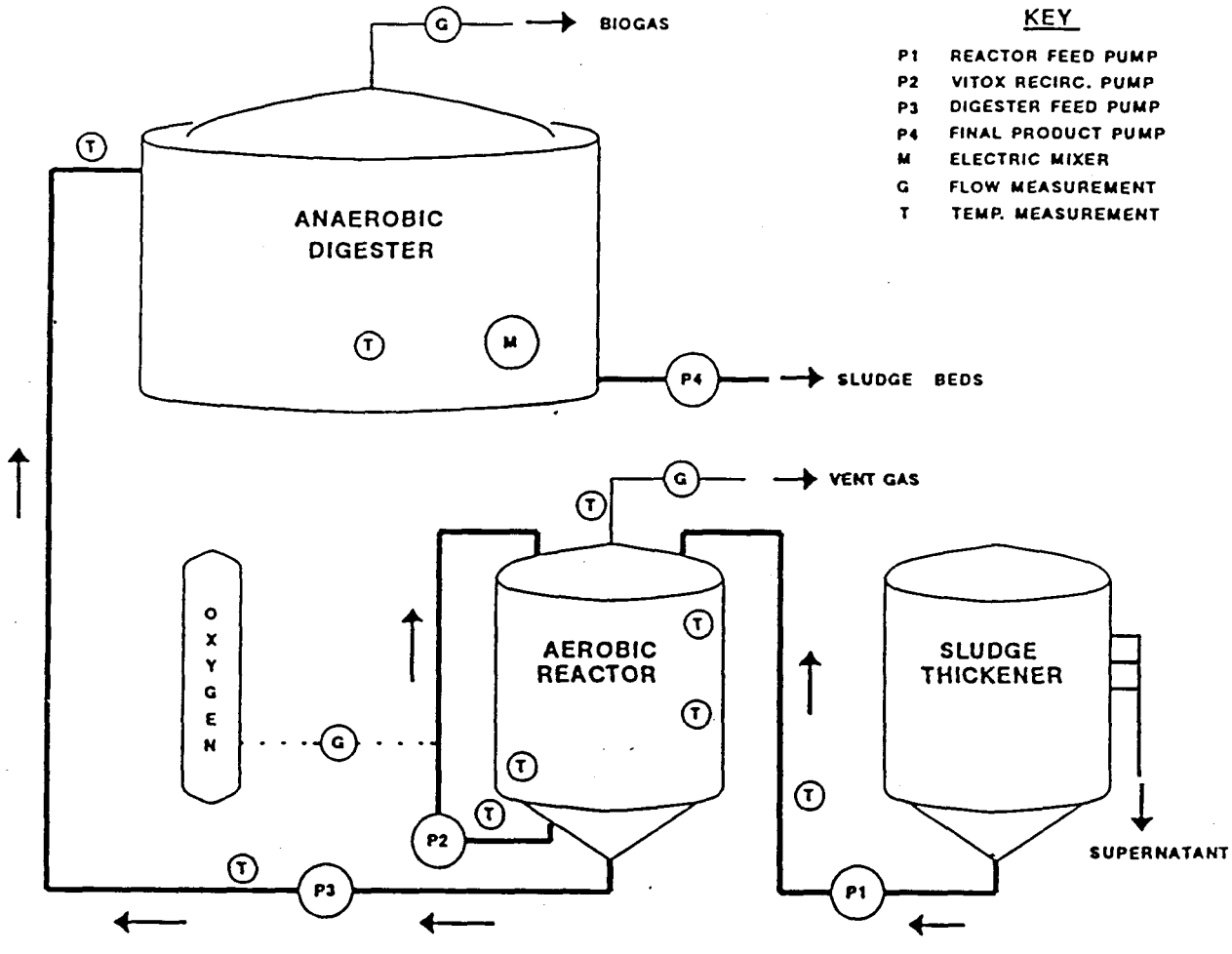

FIGURE 3.2: A schematic layout of the dual digestion plant at Potsdam showing the sludge thickener, the aerobic reactor and the anaerobic digester, pumps and positions of temperature and gas flow measurements.

3.2.1 Sludge Thickener

A 50 m^3 closed cylindrical glass fibre tank with a 45° conical bottom, 3,2 m in diameter and 5,7 m deep (excluding the conical bottom) was used as sludge thickening tank. Three draw-off points were provided at one metre levels from the upper edge of the cylindrical section to enable supernatant decanting. The supernatant from the thickener was returned to the head of works. The thickener was filled as required from the sludge collection sump to which the sludge from the primary settling tanks was fed. This sludge also included humus sludge as explained earlier. The thickened sludge was withdrawn from the apex of the settling cone to a macerator/feed pump. This pump, controlled by the electronic timer/level control system, feeds the thickened sludge into the top of the aerobic reactor.

3.2.2 Aerobic Reactor

The aerobic reactor is a fully lagged 50 m^3 glass fibre tank with the same dimensions as the sludge thickener. The lagging comprises a 50 mm polyurethane foam layer with an outer covering of epoxy resin and glass matting for mechanical strength and waterproofing. The normal operating sludge level is 0,6 m below the upper rim of the tank, giving an effective operating volume of 45 m^3.

A float operated level control system allows a fixed volume of hot disinfected sludge to be withdrawn down to a preset level by a centrifugal pump which transfers the aerobically treated sludge to the anaerobic digester. Immediately afterwards a macerator/feed pump refills the reactor to its operating level with sludge from the thickener. This draw/fill mode of operation prevents untreated sludge from short-circuiting to the anaerobic digester, thus ensuring effective disinfection. It also allowed the sludge feed rate to the process to be monitored very accurately, as each cycle passed a batch of 3,75 m^3 (1/12 of the reactor operating volume) through the plant. The draw/fill operation normally took 7 to 8 minutes.

The resin used in the construction of the thickener and the reactor tanks was sensitive to heat and the manufacturer's guarantee would be invalidated if the sludge temperature in the reactor rose above 70°C. Accordingly 68°C was set as the maximum allowable reactor sludge temperature.

3.2.3 Sludge Oxygenation System

The sludge in the aerobic reactor is oxygenated and mixed by a "VITOX"[1] oxygenation system supplied by Afrox. The system forms an external recirculation loop around the aerobic reactor. It comprises a large torque-flow centrifugal pump (Allis Chalmers Model NSR 4 X 4 X 12, 150 m^3/h nominal delivery) which draws sludge from near the apex of the conical bottom of the reactor and passes it through a venturi on its delivery side where pure oxygen is introduced under pressure. The sludge/oxygen mixture is passed through a fixed length of pipe to facilitate oxygen dissolution and is reintroduced into the reactor where it is

dispersed below the liquid surface by a special nozzle[2]. The 22 kW of energy provided by the system ensures good mixing in the reactor.

The pure oxygen is supplied from a liquid oxygen storage vessel. It passes via an evaporator and the necessary pressure regulators and flow meters (Rotameters) to the venturi. The oxygen flow rate is adjusted manually as required.

3.2.4 Anaerobic Digester

The anaerobic digester is an existing 1000 m^3 concrete unit with a domed roof. The unit has a diameter of 16,7 m and a sidewall depth of 6,1 m.

The sludge depth is controlled between predetermined levels by regularly discharging treated sludge to the drying beds or to the Zimpro storage tank by means of an external pump. A large bore 6 m high sight glass which is suitably calibrated and fitted with a fresh water flushing system is used to gauge the level of the sludge in the digester. By varying the level in the digester it is possible to operate the unit at any retention period from 8 to 40 days, assuming a daily input of aerobically treated sludge of between 25 to 45 m^3.

Mixing is effected by means of a directionally adjustable 7 kW Flygt submersible propeller mixer. An adjustable timer is used to activate the unit as required. It is also possible to use the discharge pump for mixing, this unit withdrawing sludge from the bottom centre of the digester and discharging it on the surface near the sidewall. The latter mixing mode is used when the sludge level is too low for the submersible mixer to operate effectively, for example, during start-up of the anaerobic digester.

The hot sludge transferred from the aerobic reactor is introduced from the top of the digester, near the sidewall, via a downpipe and siphon breaker vent. The latter is required to avoid siphoning sludge from the aerobic reactor, which is at a higher level. Digester gas is discharged from a gas dome at the top of the digester. The dome is equipped with a safety water seal. Normally the discharge pressure is 0,6 kPa but if the pressure exceeds 3 kPa the water seal is blown out, allowing the pressure to drop. The gas is passed via a long condensing pipe followed by a water trap and a re-heat section to a positive displacement totalizing gas meter. The meter, manufactured by Parkinson Cowan, was designed to operate on town gas and had to be considerably modified to cope with the corrosive digester gas.

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[1] Patented by British Oxygen Company (BOC)
[2] Patented by African Oxygen Ltd (AFROX)

3.2.5 Temperature Monitoring and Recording

A programmable ten channel Honeywell Model 1312-10-A-3 recorder was used to monitor temperatures continuously, using Pt 100 resistance thermometers, at the following locations in the process:

* 1 point in the macerated sludge feed pipe to the aerobic reactor, indicating the feed sludge temperature;
* 3 points in the aerobic reactor;
* 1 point in the "VITOX" recirculation loop;
* 1 point in the vent gas line from the aerobic reactor;
* 2 points in the transfer line to the anaerobic digester;
* 1 point in the anaerobic digester;
* 1 point to monitor ambient temperature.

Outputs from the recorder are used to trigger a warning siren when over-temperature conditions occur. When this happens the oxygen flow rate is automatically reduced by re-routing the oxygen feed through a smaller preset flowmeter, thereby reducing the heat generation in the reactor.

3.3 MODE OF OPERATION

The primary settling tanks at the sewage plant are manually desludged about every 5 hours, the sludge gravitating to a collecting sump from where it is pumped to either the Zimpro or dual digestion thickener. The shift operator drains supernatant from and tops up the dual digestion thickener at regular intervals. The success of this part of the operation depends entirely on the skill of the operator, as it is important to thicken the sludge to at least 3,5% total solids.

The aerobic reactor operates on a draw/fill batch cycle. An accurate fully adjustable quartz timer controls the cycle frequency, while a level control system consisting of a large steel float which operates external magnetic reed switches, ensures that a very accurate "batch" is treated during each cycle.

The following sequence occurs:

1. The timer initiates the draw/fill cycle, typically every 2 hours for a one day average detention time in the aerobic reactor. The transfer pump starts, transferring hot aerobically treated sludge to the anaerobic digester.

2. When the level control float reaches its set lower level, the transfer pump stops and the feed pump starts, transferring sludge from the thickener to the aerobic reactor.

3. When the level control float reaches its set upper level, the feed pump stops, the reactor having been filled to its normal operating level. Throughout the test period, the levels were set so that a batch constituted 1/12 of the reactor volume (3,75 m^3).

4. The recirculation pump operates continuously and is not affected by the draw/fill cycle. Oxygen is introduced continuously while the pump is operating. This effectively mixes and aerates the aerobic reactor contents continuously.

5. The anaerobic digester is desludged daily down to a level marked on the sight glass in order to maintain a predetermined retention time. This operation is carried out by the shift operator at a fixed time daily. The anaerobically treated sludge is pumped to the sludge drying beds if space is available, or otherwise to the Zimpro storage tank. The digester receives a batch (3,75 m^3) of aerobically treated hot sludge at the preset intervals. The sludge volume increases until the desludging operation occurs, when the predetermined level is restored.

6. The temperature of the anaerobic digester depends entirely on the volume and the temperature of the sludge transferred from the aerobic reactor. Some control is effected by bypassing batches of heated aerobic sludge to the Zimpro storage tank, but this in turn changes the retention time. Towards the end of the test period a batch heat exchanger was installed to cool the aerobic sludge, but it was never tried owing to commissioning problems and failure of the aerobic reactor vessel.

7. The propeller mixer in the anaerobic digester operates on a preset cycle, typically it would run for 20 minutes in every hour.

CHAPTER 4

OPERATION OF PLANT

4.1 COMMISSIONING AND OPERATING HISTORY

The aerobic stage of the dual digestion plant was completed and hydraulically tested by March 1987, when tests on sewage sludge commenced. The initial operation was plagued with numerous equipment problems and it was only from mid-November 1987 that continuous stable operation could be achieved. In spite of the intermittent operation during this early period, it was evident that the plant would easily reach and maintain the temperature levels required for disinfection and that stable operation would be possible once the equipment problems had been solved. In the event, the experience gained in correcting all the problems proved invaluable in understanding the aerobic process better and in improving its performance.

The anaerobic digester was started up early in February 1988, but this unit was not operating properly till May 1988. A problem with the aerobic stage recirculation pump caused an interruption in operation of about one month at the end of 1988. A period of continuous operation, with only very short interruptions for minor repairs and maintenance, followed till the plant was shut down on 11 January 1990 for installing a heat exchanger and an improved computerized monitoring system.

During 1990 the operation of the plant was essentially aimed at investigating the start-up and operation of the anaerobic digester and to establish the effect the aerobic pretreatment had on its performance. The investigation would be performed by a M.Sc student and it was hoped that the computerized temperature monitoring system and operation with a heat exchanger could be evaluated. Problems with the new apparatus caused long delays and the heat exchanger was seriously damaged during commissioning. Tests with the heat exchanger were shelved until it could be repaired or replaced, which did not happen before the research was precipitously terminated by the structural failure of the aerobic reactor tank on the 23 November, 1990.

4.2 PLANT OPERATION AND MONITORING

4.2.1 Sludge Thickening

The efficiency and stability of the aerobic stage is directly affected by the feed sludge density. It was therefore considered important to attempt to maintain a sludge density of at least 3,5 g/L to ensure efficient operation.

Owing to the relatively poor performance of some of the settling tanks, a thin sludge arrives at the central sludge collection sump, which was pumped to the dual digestion sludge thickener as required. At regular intervals, the plant operators decanted the clear supernatant through one of the three drain ports on the thickener. At times dilute sludge was wasted in the same manner in an attempt to ensure the most concentrated feed sludge possible for the aerobic reactor.

When the aerobic reactor is operating at high retention (1,5 days or more) and the ambient temperature is low, the thickener produces a sludge having a total solids (TS) content of 3,5 to 4,0%. At low aerobic reactor retention time (1,0 to 1,25 d) with concomitant high feed volume requirements, the TS could drop below 3%. This condition was alleviated by installing a submersible pump in the sludge collection sump to decant supernatant before pumping to the thickener. Judicious management of the overall thickening process improved the density of the feed sludge by up to 1% on average.

Long retention times in the Milnerton sewerage reticulation system resulted in some sludge septicity, and the degree of consolidation achieved in the thickener was often compromised by the generation of gas by the sludge. Occasionally during very hot weather the sludge was buoyed to the surface resulting in complete sludge/supernatant inversion which necessitated the withdrawal of supernatant from the base of the thickener to achieve thickening.

The efficiency of the thickening process relies entirely on the skill and diligence of the shift operators. That they have performed their tasks creditably is testified by the good results achieved.

The sludge density was regularly monitored by analysis of composite samples taken by the operators.

4.2.2 Aerobic Reactor

4.2.2.1 Sludge feeding and level control

Accurate sludge level monitoring inside the reactor was essential to control sludge levels during the batch draw and fill operation. The first system used conductivity probes to sense sludge level. The system was known to be extremely accurate when sensing water levels, but proved to be useless in this application owing to the foam layer causing spurious switching with consequent inaccurate filling. A capacitance probe level control system from a reputable manufacturer which was tried next produced even worse results.

A "home-made" level control system was eventually built which has proved to be very accurate and reliable. It comprised a heavy stainless steel cylindrical float, which rested on the sludge surface, to which was attached a rigid rod with a plastic bottle containing a magnet fastened to its top end. The magnet would activate reed switches fixed to a rule at predetermined points. The reed switch at the high level corresponded to the 45 m^3 operating volume and the low level to

41,25 m^3, the difference between the two levels being the 3,75 m^3 batch volume. The level control system is illustrated in Figure 4,1.

At predetermined intervals of not less than 2 hours, a quartz pulse timer activated the transfer pump and aerobically treated sludge was pumped from the reactor to the anaerobic digester, causing the float to drop until the low level switch was activated. This terminated the transfer cycle with exactly 3,75 m^3 having been transferred. Immediately after termination of the sludge transfer to the digester, sludge feeding from the thickener would commence. The feed cycle was terminated by the high level reed switch when the operating volume once again reached 45 m^3. The pulse timer merely initiated the transfer cycle and after initiation, the level control system controlled the transfer and feed operations.

Additional reed switches were attached to the rule above and below the high and low level control switches. These were alarm switches which would be activated in the event of a failure of one of the level control switches.

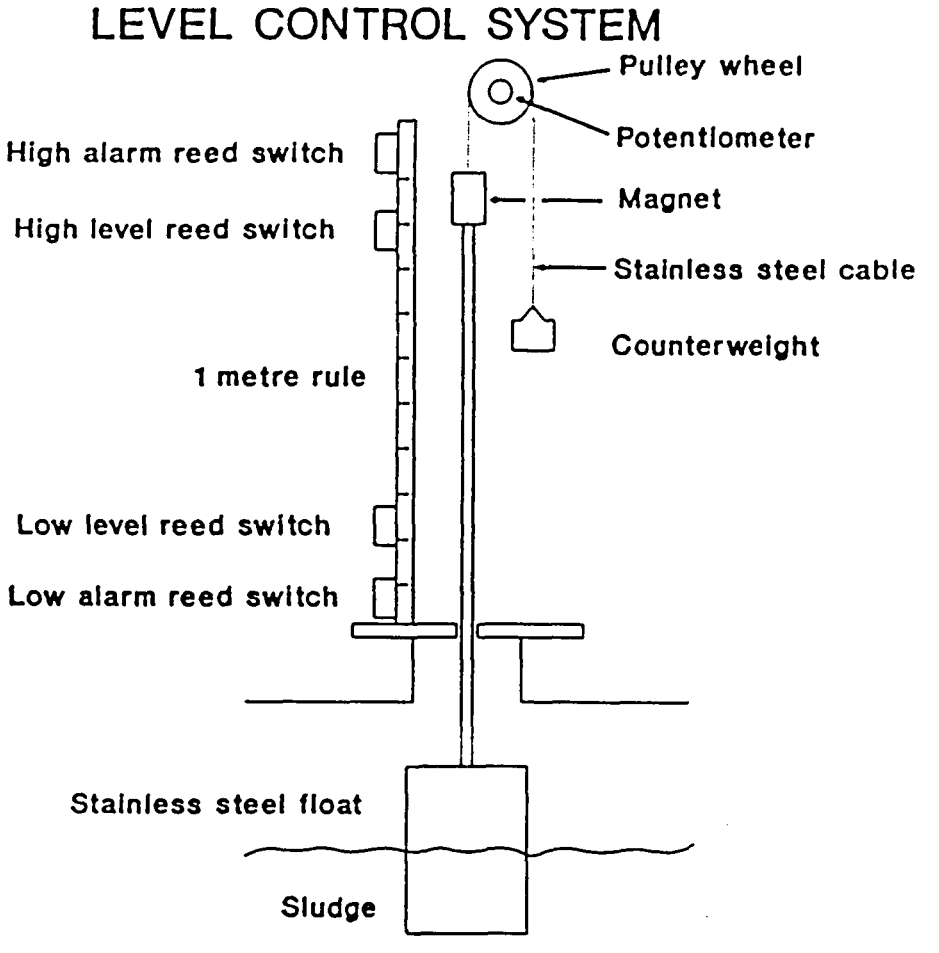


FIGURE 4.1: The aerobic reactor level control system which maintained a constant sludge level of 45 m_3, and which controlled the batch draw and fill system to ensure that 3,75 m^3 batches of sludge were transferred from and fed to the reactor.

A 1,5 mm stainless steel cable was connected to the top of the float rod. The cable was looped over a pulley wheel fitted on the shaft of a potentiometer and a counterweight attached to its free end. Vertical movement of the float thus rotated the pulley wheel and the potentio-meter. This changed the resistance in a constant voltage circuit and the change in current was displayed on an ammeter located in the control hut. This arrangement provided a visual indication of the sludge level in the reactor.

4.2.2.2 Oxygenation system

A Vitox system was used to oxygenate and mix the aerobic reactor. This system comprised a recirculation loop through which the aerobic reactor sludge was pumped continuously. Oxygen was injected via a venturi located in the loop. Sludge was drawn from a point on the conical base of the reactor, passed through the recirculation loop by a torque flow centrifugal pump and discharged into the reactor below the surface through a single nozzle. The nozzle was offset axially and radially so as to provide effective mixing of the entire contents of the reactor. The flow rate was approximately 150 m^3/h, giving a reactor contents turnover time of about 20 minutes.

The torque flow pump used for recirculation was not prone to blockages by rags and other debris present in the sludge. Its efficiency was only 40-50%, but the heat balance tests showed that the relatively large amount of power consumed ($\pm$ 20 kW) was not wasted as about 90% of it was converted to heat and contributed directly to heating the sludge.

During the first eight months of the project numerous problems were experienced with the pump which included:

(1) Poor standard of manufacture and assembly;

(2) Cavitation in the absence of oxygen injection;

(3) Shaft sleeve wear and gland leakage;

(4) Rapid erosion of the impeller and pump casing.

The latter two problems were solved to the extent that the pump would operate satisfactorily for approximately six months with only routine maintenance. The pump casing wear made it necessary to rebuild the casing every six to eight months, but this was accepted by the operating staff as routine maintenance and not inconsistent with over 5000 hours of pumping an abrasive sludge. A spare identical pump was purchased and suitably modified to act as a standby unit, thus the downtime was reduced to a minimum when the pump in service developed problems.

Details of the problems encountered with the recirculation pump and the modifications made to solve these problems are fully detailed in Appendix 5B.1.6 of Part 2 of this Final Report.

The oxygen supply was stored as a liquid in a 6000 litre vacuum insulated vessel. The liquid oxygen was vaporized to ambient temperature before being piped to the regulating system in the control hut. This comprised a pressure regulator followed by two accurately calibrated rotameters in parallel, each fitted with a needle valve and a solenoid valve. Either the low flow rotameter (6 to 20 kg/h) or the high flow rotameter (10 to 55 kg/h), was used depending on the requirements. When the maximum set temperature of the reactor was exceeded, the temperature recorder activated the control unit which would shut off the preset high flow rate and turn on the low flow rate, preventing any further rise in temperature but still keeping the process alive. The oxygen was injected at 400 kPa into the venturi located in the recirculation loop downstream of the pump. The oxygen partially dissolved in a tortuous path (four 180° bends) recirculation loop prior to discharge into the reactor.

Both the recirculation pump and the oxygen supply operated continuously. The oxygen flow rate was fixed for any specific "steady state" condition. Control of the oxygen supply on the basis of dissolved oxygen concentration in the reactor was not used, as previous research (Trim, 1984) had found that continuous oxygen feed at a fixed rate was more reliable and eliminated the problems associated with measuring dissolved oxygen in the sludge.

4.2.2.3 Temperature monitoring

Temperatures were measured at various points around the dual digestion process using Pt 100 resistance thermometers. The location of these measuring points is described in Chapter 3, 3.2.5, and the approximate positions indicated in Figure 3.2.

The Honeywell recorder used for monitoring and recording the data provided an immediate digital readout of the temperatures in a fixed sequence as well as producing traces of the individual temperatures on a continuous chart in different colours and printing styles.

The chart display of the four temperature monitoring points directly associated with the aerobic reactor i.e. one each in the upper, middle and lower regions of the reactor, and one in the recirculation loop, was of particular significance. The draw and fill batch feed operation produced a distinctive saw tooth profile for each of these temperatures. A typical profile for a single temperature probe is shown in Figure 4.2. There are 3 phases in a batch cycle viz: (1) transfer phase; (2) feeding phase; and (3) the heating phase. From the temperature measurements, sludge transfer from the reactor (phase 1) had no discernible effect on the temperature profile and the heating rate during this phase appeared to be unchanged from the previous phase (the heating phase 3).

Note that in order to illustrate the three phases more clearly in Figure 4.2, the durations of phases 1 and 2 are shown longer than they actually were during the test. In reality these phases were very short (about 3 minutes each) compared with phase 3 (2 to 6 hours).

The feeding of cold sludge (phase 2) caused a sharp drop of about 3°C in temperature as it rapidly mixed with the hot reactor contents. Immediately after cessation of feeding, phase 3 commenced and the temperature increased linearly at constant oxygen supply rate until the next batch cycle commenced.

4.2.2.4 Mixing in the reactor

Continuous and thorough mixing of the aerobic reactor contents is essential for maximum oxygen dissolution and distribution of feed sludge. The mixing action imparted by the recirculation system, which turned over the reactor volume about once every 20 minutes, proved to be more than adequate. The calculated mechanical heat input (H_{mi}) for mixing was 20 MJ/h, which was probably considerably more than required, but the excess energy was not wasted as it contributed to the reactor heating.

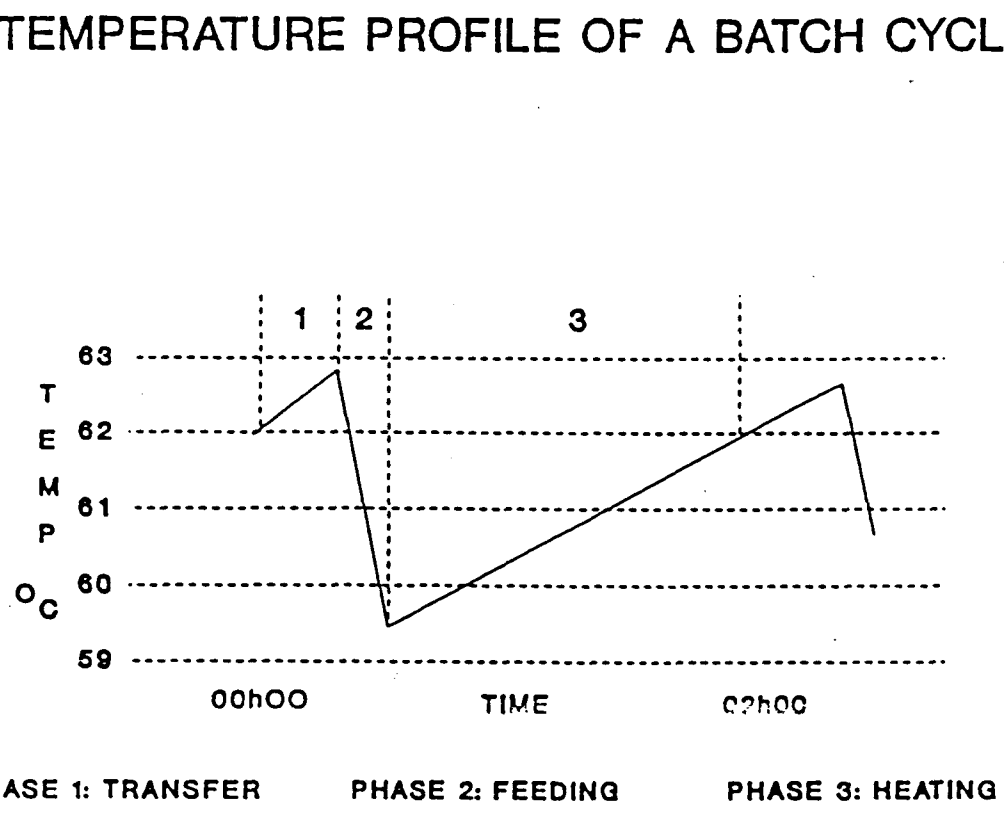


FIGURE 4.2: A typical profile of the reactor sludge temperature ($T_{..}$) against time for the 3 phases of a complete batch cycle.

By examining the temperature traces of the four probes in the aerobic reactor and the recirculation loop it was possible to gauge the effectiveness of mixing in the reactor. The rapidity and similarity of the four individual thermometer responses to the sudden temperature changes that occurred during feeding were used as indicators of the quality of reactor mixing. When the cold feed sludge was rapidly distributed into the hot reactor contents and the mixing was good, the temperatures sensed by the four probes in the reactor showed an equally rapid and similar response to the sudden temperature decrease. If for some reason the mixing was poor and the cold influent sludge was not distributed rapidly and uniformly, the temperature probes showed differing rates of response and produced dissimilar readings. Figures 4.3 and 4.4 illustrate the effect of poor mixing on the response of the temperatures in the reactor, the latter figure showing a section of an actual recorder chart during a period when poor mixing occurred while evaluating a two nozzle manifold and the change to good mixing when one of the nozzles got blocked.

The conditions for good mixing in the reactor was found by trial and error, various manifold and nozzles designs being tested before achieving success with the single angled nozzle described earlier. Reactor mixing is described in more detail in Part 2, Appendix 5B.

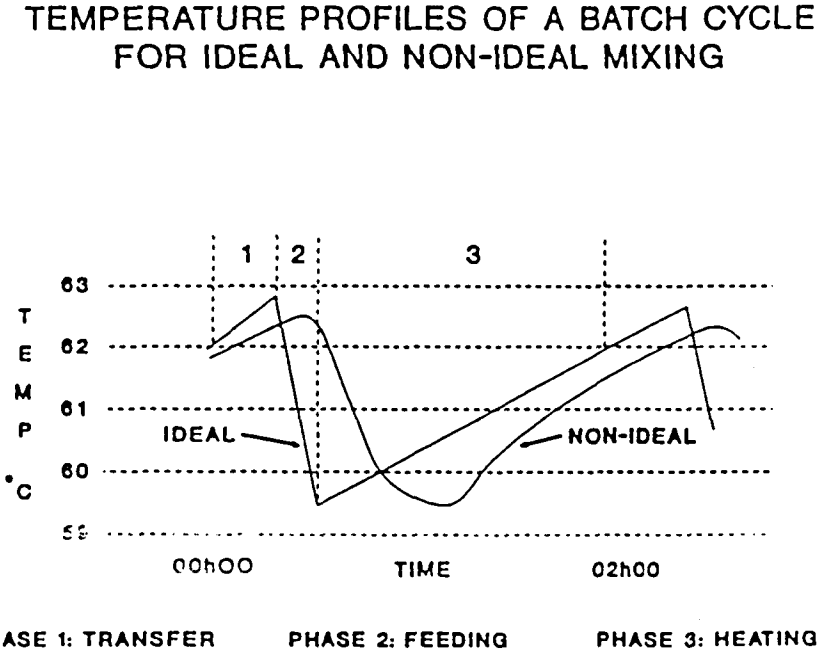


FIGURE 4.3: The profiles of reactor sludge temperature (T_{sc}) against time, illustrating ideal and non-ideal mixing via the responses of 2 temperature probes. For ideal mixing the response of the probe to change in sludge temperature is rapid and the points of change are clearly defined. For non-ideal mixing the probe response lags behind the changes in sludge temperature with poorly defined points of change.

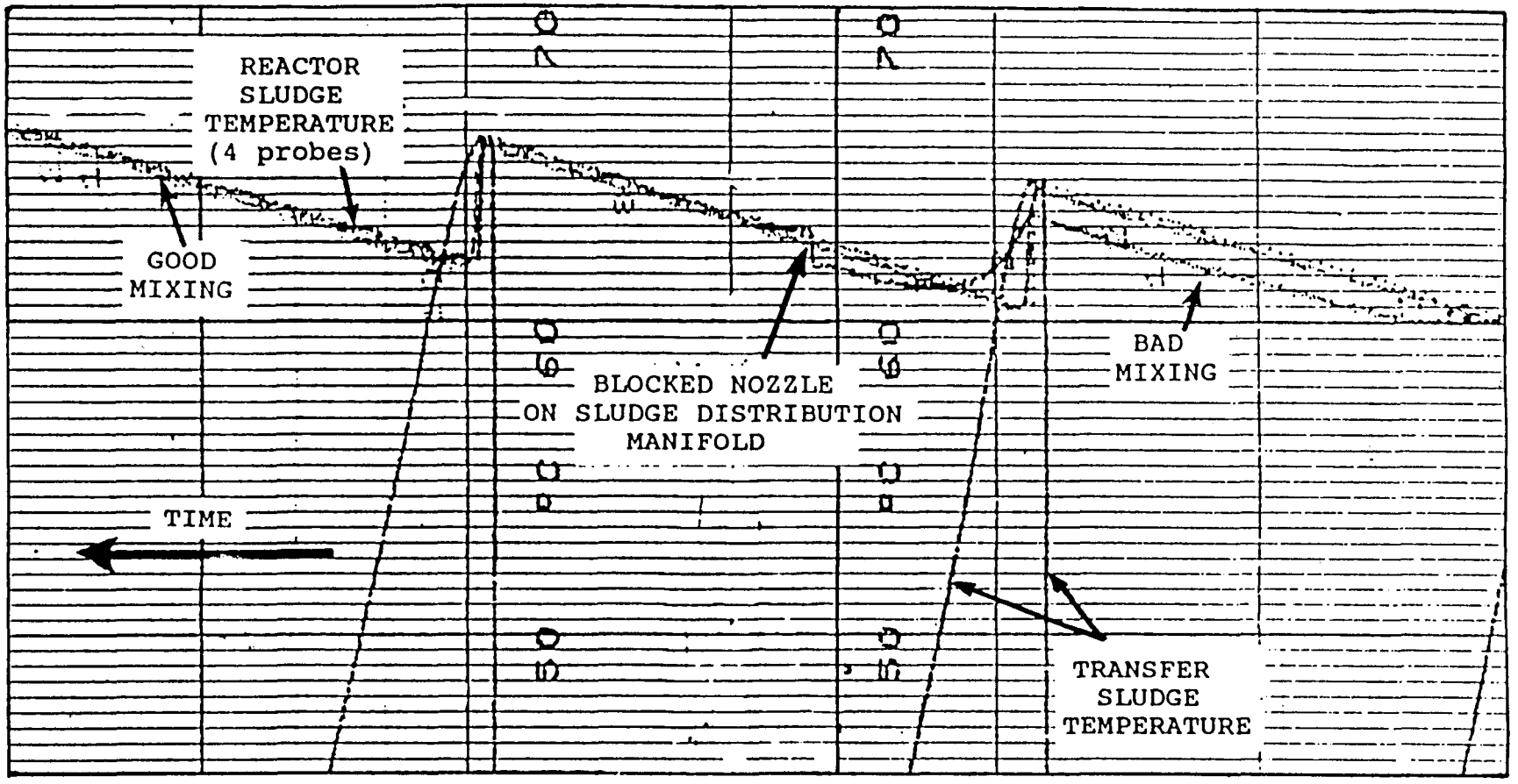


FIGURE 4.4: A chart recorder trace of the temperatures in the dual digestion system. The saw tooth temperature-time profile on the RHS of the illustration is typical of the bed mixing achieved with the sludge distribution manifold with two nozzles. One of the two nozzles in this manifold blocked at a point shown in the illustration and the saw tooth temperature-time profile to the left of this point indicates that good mixing was achieved with the remaining nozzle.

4.2.2.5 Gas measurement

Oxygen was fed continuously to the oxygenation loop of the reactor. The flow rate was controlled at a preset fixed rate by the two rotameters mentioned earlier i.e. a high flow rotameter (10 to 55 kg O_2/h) and a low flow rotameter (6 to 20 kg O_2/h). Apart from the automatic alarm control switch from the high to the low flow rotameter in the event of the reactor sludge temperature exceeding 68°C, the oxygen supply rate was manually set with the aid of the rotameters, which had been very accurately calibrated as described in Part 2, Appendix 3A.

In order to carry out the heat balance and oxygen mass balance, it was necessary to accurately measure the volume of vent gas from the reactor. The high humidity of the vent gas complicates the flow rate measurement as condensation in the positive displacement gas meter interferes with its operation and accuracy. To overcome this difficulty, the vent gas was passed through a water cooled condenser and water trap followed by a re-heating section to ensure that no further condensation would occur in the gas meter. This system functioned faultlessly and is shown in Figure 4.5. Not only did this arrangement protect the gas meter, but it also provided a means for measuring the water vapour condensate volume, which enabled the vent gas water vapour heat loss to be accurately determined. Further details are given in Part 2, section 3.4.1.2, and Appendices 3A and 3B.

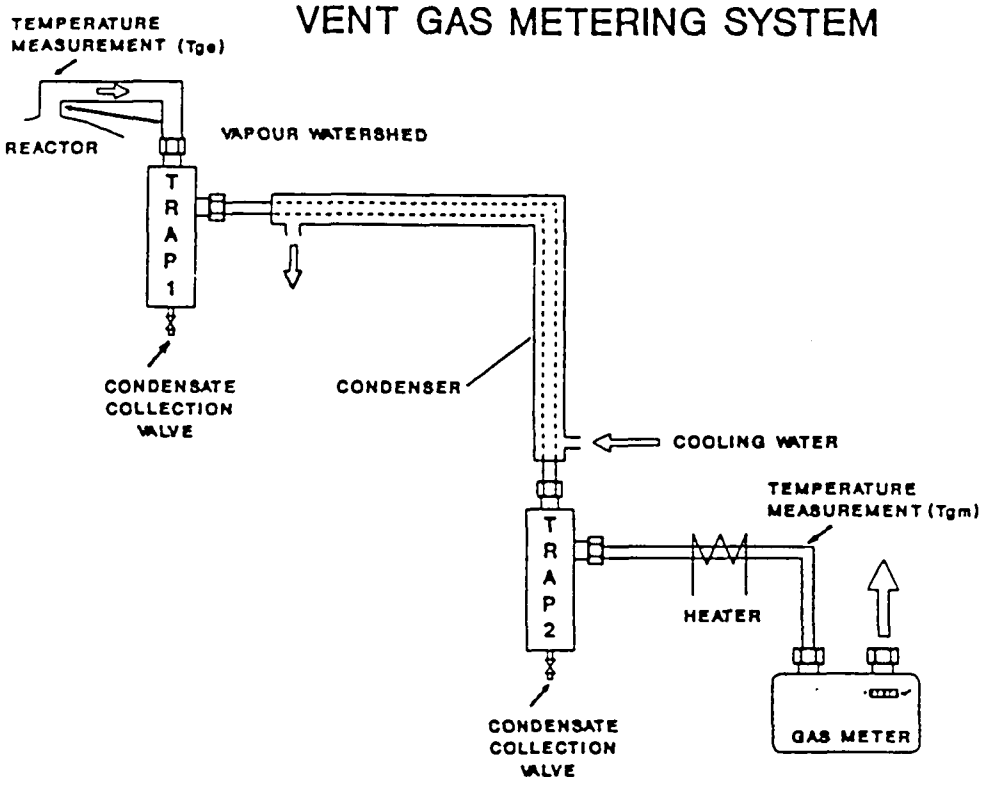


FIGURE 4.5: The layout of the vent gas metering system. The vent gas metering system (1) cooled the hot, moisture laden vent gas to 20°C, (2) enabled the condensate from the cooled and partially dried vent gas to be collected and (3) reheated this cooled and partially dried vent gas prior to the measurement of its volume in the gas meter.

4.2.2.6 Sludge flow and analyses

The daily volume of sludge treated by the process was determined by the product of the number of batches and the batch volume. As described earlier, the level control system ensured that each batch of sludge fed to the reactor had a volume of 3,75 m^3. From the temperature traces on the chart recorder, the saw-tooth profile clearly indicated when feeding occurred and the number of feed cycles per dy could easily be counted.

The nature of the sludge, especially the raw thickened sludge, made accurate sampling difficult. A system of composite sampling was devised to ensure that representative samples were obtained for analysis. Samples were taken for each of those batch feeds that occurred between 08h00 and 24h00, i.e. during the period when the shift operators were on duty. Samples were drawn from the discharge pipes from the transfer and feed pumps. To minimize the effect of stratification in the feed sludge thickener, the feed sludge sample was tapped from the discharge line in short, regular bursts over the entire duration of the feed cycle. The same procedure was followed for taking the transfer sludge sample, although, given the high rate ofmixing (400 W/m^3) in the aerobic reactor, this was not really necessary as the sludge consistency was much more uniform. The pumps each operated for about 3 to 4 minutes, during which time the sampling occurred. A fixed volume (1,5 L) of each collected sample was then transferred to the appropriate composite sample bucket. The following morning the composite feed and transfer samples were thoroughly stirred and a 1,5 L sample of each taken for analysis.

During heat balance and other intensive tests, samples were taken daily. Otherwise weekly samples were taken and analyzed. Daily samples were analyzed at the laboratory on site at the treatment works, while weekly samples were analyzed at the laboratories of the DWT of the CSIR at Bellville. When daily samples were analyzed at Milnerton, control samples were also done at least once weekly at Bellville to check the accuracy of the results (see Part 2, Appendix 4D).

4.2.2.7 Aerobic reactor start up

The aerobic reactor was started up with fresh sludge on at least 8 occasions. Initially, when little was known about the response of the system to the various factors that influenced it and the control equipment had not been sorted out (e.g. the level control in the reactor), severe initial foaming occurred. Once the cause of the excessive foam generation had been established and the process was better understood, successful start up could be achieved in most cases in about 4 days. The start up phase was considered to be completed once an active aerobic culture had been established at a temperature of 60°C or more. A typical start up temperature profile is shown in Figure 4.6.

To achieve aerobic reactor start up, the unit was filled with sludge from the thickener to its operating level and continuous oxygenation started. During the first year of operation (1987) from 15 to 20 kg oxygen was fed per hour, but in

subsequent start ups this rate was reduced to 6 kg O_2/h in an attempt to control foam generation. Foam was generated in the reactor throughout the start up phase, but ceased once an active aerobic culture had been established. During the start up, the reactor was periodically fed with raw sludge from the thickener to ensure that sufficient substrate was available to establish an active aerobic culture.

In Figure 4.6 the times at which the reactor was fed can be seen as the sharp temperature decreases.

The point at which an active culture had been established was easily identified. Besides the cessation of foaming, another indicator of incipient biological activity was the rate of change of the reactor sludge temperature. Over the initial period after start up, approximately the first 24 hours, the temperature increase is nearly linear and is due to the heat input from the recirculation pump. When biological activity commenced, oxygen consumption began and biological heating started.

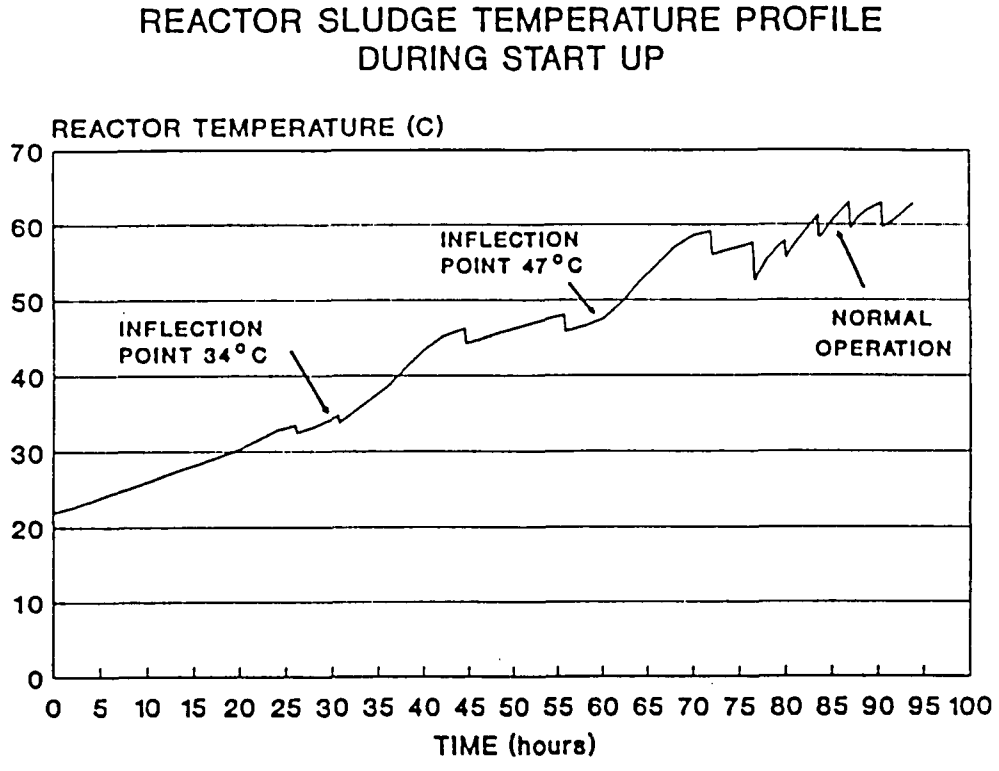


FIGURE 4.6: The change in reactor sludge temperature with time during a typical start-up of the aerobic reactor from ambient temperature. While the temperature change over the first 24 hours is approximately linear and is due to heat input from the Vitox recirculation pump, inflection points on the temperature trace indicate heat generation associated with incipient biological activity. The introduction of sludge feed to the reactor is shown by sharp decreases in reactor sludge temperature. In the above illustration, about 90 hours (3,75 days) elapsed between starting up with raw sludge and the development of a fully active biological culture.

This was indicated as a sudden increase in the temperature trace. In Figure 4.6 such an inflection point is shown when the sludge temperature had reached 34°C. A further one occurred at about 47°C. These points probably indicate the points of transition for mesophilic and thermophilic cultures.

Usually biological activity occurred within 24 to 36 hours after oxygenation commenced, and the operating temperature of ± 60°C was attained after about 96 hours.

On one occasion the start up phase took two weeks instead of the usual four days. Usually restarting the process was a straight forward operation, but on this occasion (21/12/88) the only temperature increase that was noted was due to the energy input from the pump. The sludge that had been put into the reactor had been relatively old, so it was assumed that this was the cause of the problem. The reactor was emptied and refilled with fresher sludge, but once again the temperature rise was unusually slow. It was noticed that the pH also remained low, so 75 kg sodium carbonate was added to the reactor, increasing the pH from 5,15 to 7,35. Within 12 hours a normal temperature rise was evident and the rest of the start up sequence occurred as expected.

The feed sludge usually had a very low pH (± 5,1) but previously no pH correction had been necessary and the pH had increased normally to 7 or more within about 4 days. The above experience indicated that apparently a minimum alkalinity level is required for biological activity to start, thereafter the process generated sufficient alkalinity on its own. Nevertheless, subsequent to this occasion, the precaution was taken of correcting the feed sludge to pH 7 with sodium carbonate or lime before starting up the reactor.

4.2.2.8 Foam production in the aerobic reactor

At Milnerton it was found that foaming occurred in the aerobic reactor when oxygen was fed to the unit and the sludge was biologically inactive. Foam production was never observed during normal operation when oxygen was being consumed and biological heat was being generated.

Foaming always occurred during start up with fresh sludge and continued until normal biological activity had been established and most of the oxygen fed to the reactor was being consumed. Start up with very low oxygen flow rates (6 kg O_2/h) reduced the volume of foam produced but its formation could not be eliminated. It was assumed that the foam was caused by large amounts of undissolved oxygen passing through the sludge mass and as the sludge became active and consumed more oxygen, more dissolved, and foam generation tapered off. This explanation would possibly account for the continuous foaming associated with air oxygenated aerobic reactors. The large volumes of undissolved nitrogen passing through the sludge produces continuous heavy foaming. If this theory is correct it should be possible to start up an oxygen fed reactor without foaming by accurately controlling the oxygen input so that all the oxygen is used.

Foaming also occurred when the sludge became inactive or "died" after having been deprived of oxygen for more than an hour. The foaming ceased as soon as the sludge became active.

Only on one occasion did some foaming occur while the sludge was active. This occurred during a test when the retention time was suddenly decreased from 3 to 1,25 days, when slight foaming was observed which dissipated after about 30 minutes. It appeared to be the result of an acclimatization phase during which, after its activity had been constrained by the low oxygen supply (6 kg O_2/h) for about 14 days, which is a typical supply rate for maintaining 60+ °C at 3 days retention, the sludge culture adapted to the higher oxygen supply rate and shorter retention time.

Foaming did not cause undue problems at Milnerton - it seldom occurred and was highly predictable. When foam was produced, the operating staff took the necessary steps to control its effects and to prevent it from becoming a nuisance. Excess foam was diverted to a drain after closing off the vent gas pipe to prevent the foam from getting into the gas measuring equipment or being blown out to the surroundings. When the vent was isolated, care had to be taken against creating a vacuum in the reactor headspace, as the only connection to the atmosphere was now the overflow pipe to the drain. In the event of a sudden vacuum in the head-space e.g. when sludge was transferred from the reactor, normally air would be drawn in through the vent pipe or through the overflow pipe. When the vent pipe was closed and the overflow obstructed by an accumulation of sludge and foam, as often occurred during foaming, there was a real danger that a vacuum created by transferring sludge could have caused the glass fibre reactor to implode. Care was therefore taken not to transfer sludge during periods of foaming.

4.2.2.9 Aerobic vent gas odour

The vent gas emitted from the aerobic reactor had a pungent, rather unpleasant odour. Fortunately the volume produced was small and the odour dispersed very rapidly. Initially the gas was vented to the atmosphere and caused no problems, but owing to the relatively close proximity of residential areas to the treatment works it was decided to discharge the gas into the vent line from the Zimpro storage tank. This line passes through a scrubber before discharging to the atmosphere. The condensate which collects in the water trap of the vent gas line contains an oily substance which has an extremely bad odour. Whether the removal of the condensate had any effect on the gas finally discharged was not determined.

4.2.2.10 Aerobic reactor vessel

Notwithstanding the initial doubts regarding the suitability of the glass fibre tank used for the aerobic reactor, it performed well for nearly four years during which time it showed no sign of undue wear. The upper temperature limit of 70°C specified by the manufacturers did not prove to be a serious handicap as the temperature could be accurately regulated by controlling the oxygen input. The

alarm system that was devised for ensuring the safety of the reactor gave warning as soon as the reactor temperature exceeded 68°C and at the same time it cut down the oxygen flow rate to a much lower preset level. The upper temperature limit was rarely reached although the process was capable of producing much higher temperatures.

The reactor, with its 50 mm polyurethane/epoxy cladding, proved to have very good heat insulating properties. Heat loss tests were performed to obtain accurate data for the heat balance determinations (see Part 2, Appendix 3D). The overall heat transfer coefficient U was calculated to be 0,0073 MJ/(m^2.°C.h). As an example with the reactor sludge temperature at 63,7°C and the ambient temperature at 12,5°C, the wall heat loss rate amounted to 28,8 MJ/h.

The structural integrity of the tank was initially considered to be more than adequate. The unit survived a partial collapse of the dome (top) when a vacuum was inadvertently created in the reactor (see Part 2, Appendix 5B.1.7). This mishap caused only very minor damage to the insulation. A vacuum relief valve was subsequently fitted.

During a plant inspection shut down on 22 September, 1988, the reactor was drained, cleaned and given a full inspection by the manufacturers. No visible wear was found in spite of the abrasive nature of the sludge. The vessel was declared to be in excellent condition.

On 23 November, 1990, almost four years after the project was started, the reactor suffered terminal structural failure. The bottom cone sheared completely and very suddenly from the cylindrical tank body, flooding the surrounding area, including the pump house, with hot sludge. This accident put paid to further research but fortunately by this time most of the important aspects had been addressed. In the light of this catastrophic failure of the reactor the possible future use of glass fibre vessels for the high temperature aerobic stage of dual digestion cannot be recommended.

4.2.3 Aerobic Reactor - Heat generation, temperature control and oxygen requirements

As part of the overall research project, a detailed investigation into the aerobic reactor performance was undertaken. Detailed aspects of this investigation are reported in Part 2 of this Final Report.

A brief summary of some of the research into the performance and behaviour of the aerobic reactor, as applied in the dual digestion process, is presented below.

The objectives were to examine:

(1) Oxygen requirements and oxygen utilization efficiency, with pure oxygen and air oxygenation;

(2) Minimum retention time that can be achieved in the aerobic reactor while still maintaining the required thermophilic temperatures for disinfection, and effecting a degree of pretreatment with regard to solubilization of particulate matter;

(3) Temperature control of the aerobic reactor.

In order to meet these objectives, it was necessary to examine the kinetics of biological heat generation in the dual digestion aerobic reactor. It was realized that there was a distinct difference between the objectives of aerobic thermophilic digestion as a stand alone process and those of the aerobic thermophilic reactor in the dual digestion process. In the case of the former, the principal objective is sludge stabilization, i.e. to reduce the sludge energy as measured by volatile solids (VS) removal, and accordingly the kinetics are defined in terms of VS degradation rate. Consequently the retention time to achieve a specific VS removal, the rate of biological heat generation and the oxygen utilization rate are all related to the rate of VS degradation via separate constants if proportionality. For the kinetics to be controlled by the VS degradation rate, the process needs to be operated under conditions of oxygen sufficiency (Andrews and Kambhu, 1971). In the case of the dual digestion aerobic reactor biological heat generation was the principal objective and not sludge stabilization, which would be accommodated in the second anaerobic digestion stage of the process. This investigation found that the reactor should be operated under oxygen limiting conditions, which resulted in limited VS removal but produced large quantities of biological heat. Also, Mason (1986) found that the solubilization of particulate matter was promoted by operation under oxygen limitation and particulate solubilization was found to be the rate limiting step in the acid phase of anaerobic digestion (Eastman and Ferguson, 1981).

Accordingly, as biological heat generation can theoretically be shown to be stoichiometrically related to oxygen consumption rate and the operation occurs under oxygen limitation, it was proposed that the oxygen supply rate (OSR) and the sludge oxygen consumption rate (O_c) would be the most useful parameters for operation, control, design and simulation of the aerobic reactor. Consequently the primary objectives of the aerobic reactor research was to substantiate the above proposal by:

(1) Verifying the direct proportionality between the rate of biological heat generation and oxygen consumption via the specific heat yield (Y_H, MJ/kgO); and

(2) Demonstrating that by means of the oxygen supply rate (OSR), instantaneous and complete control of the reactor sludge temperature could be obtained.

In order to determine the specific heat yield, Y_H, in terms of the oxygen utilized, accurate heat and oxygen mass balances were required across the aerobic reactor.

4.2.3.1 Heat balance across the reactor

Two types of heat balance can be made across the aerobic reactor, a steady state balance and an unsteady state balance. The steady state heat balance is generally applied to a continuously fed reactor. The reactor is considered to be at steady state if the heat balance parameters, such as reactor temperature, do not change with time. If this condition applies, the sum of the heat sources to the reactor equals the sum of the heat sinks from the reactor.

A typical steady state heat balance for the reactor would be:

$$H_{bi} + H_{mi} + H_{si} = H_{se} + H_{we} + H_{ve} \qquad (4.1)$$

(net sources) = (net sinks)

Where:

H_{bi}	=	rate of biological heat generation	(MJ/h)
H_{mi}	=	rate of heat generation in the reactor due to the mechanical action of mixing	(MJ/h)
H_{si}	=	rate of heat generation due to sensible heat of the influent sludge	(MJ/h)
H_{se}	=	rate of heat loss due to sensible heat of the effluent sludge	(MJ/h)
H_{we}	=	rate of heat loss from the surfaces of the reactor and Vitox oxygenation system	(MJ/h)
H_{ve}	=	rate of heat loss in the vent gas water vapour	(MJ/h)

Note that as the oxygenation is accomplished with pure oxygen, the contributions by the influent and effluent gas sensible heats are negligible, typically <0,3% of the sources and <0,4% of the sinks.

However, because the aerobic reactor is batch fed, it is never at a steady state. The heat sources are either greater than or smaller than the heat sinks, causing the typical saw tooth fluctuation in the reactor temperature shown in Figure 4.2. A "steady state" condition in such a batch fed reactor can only be identified over a complete batch cycle, i.e. between corresponding times encompassing all three phases (Figure 4.2) together, and should comply with the following criteria:

(1) The reactor sludge temperature at corresponding points of successive batch cycles must be equal;

(2) Reactor operating parameters must not have changed for 3 retention times (sludge ages).

From the above, if time intervals other than those between corresponding times in successive batch cycles are taken, an unsteady state heat balance must be applied to a batch fed reactor.

At Milnerton, an unsteady state heat balance was applied over the linear heating phase (Phase 3 in Figure 4.2) to form the basis of determining the rate of biological heat generation (H_{bi}). The heating phase was chosen because:

(i) It was the longest period and allowed sufficient time for carrying out the measurements for the heat balance and oxygen mass balances;

(ii) During the heating phase the sludge volume in the reactor remains constant and does not disturb the vent gas flow rate or composition. Also, the heat gained by the sludge was directly proportional to the sludge temperature increase.

As the heat gained by the sludge is directly proportional to the temperature increase, the unsteady state heat balance over the linear heating phase may be expressed as:

Change in reactor enthalpy = Σ heat sources - Σ heat sinks

Accepting gas sensible heats as negligible for the pure oxygen oxygenated reactor, the unsteady heat balance may be expressed as:

$$H_{net} = H_{bi} + H_{mi} - H_{ve} - H_{we} \quad \text{(MJ/h)} \qquad (4.2)$$

and

$$H_{net} = V_p \times \rho_s \times C_p \times \frac{dT_{se}}{dt} \quad \text{(MJ/h)} \qquad (4.3)$$

Where:

H_{net}	=	the rate of reactor sludge enthalpy increase during specific heat yield test	(MJ/h)
Vp	=	the operating volume of the reactor	(m^3)
Cp	=	the specific heat of the sludge	[MJ/(ton.°C)]
ρ_s	=	the density of the sludge	(ton/m^3)
$\frac{dT_{se}}{dt}$	=	rate of reactor sludge temperature change	(°C/h).

For sewage sludge the specific heat and the density were accepted to be equal to those of water, i.e:

C_p	=	4,184	MJ/(ton.°C)
ρ_s	=	1,000	ton/m^3

The presence of up to 4% dry solids was shown to have an insignificant effect on the calculation.

With the exception of H_{bi} all the terms in equations 4,2 and 4,3 can be calculated from measurements. H_{bi} can therefore be determined.

The methods employed for determining dT_{se}/dt, H_{mi}, H_{we} and H_{ve} are all described in detail in Part 2, Chapter 3, sections 3.4.3.1 to 3.4.3.4 of this Final Report.

4.2.3.2 Oxygen mass balance

An oxygen mass balance was carried out over the same time interval in which the heat balance was conducted, enabling the oxygen consumption by the sludge during that interval to be determined (Refer to Part 2, section 3.4 for details).

The influent oxygen mass flow was measured by means of the accurately calibrated rotameters of the oxygen control system (described in section 4.2.2.5 above). The effluent oxygen mass flow was determined by measuring the volumetric flow rate of the vent gas as well as the oxygen concentration in the vent gas (as described in section 4.2.2.5 above). The water vapour removed from the vent gas as condensate ahead of the gas volume meter provided a means of accurately determining the water vapour heat loss in the vent gas.

The following data were measured during a specific heat yield test:

(1) the temperature of the vent gas at the point of exit from the reactor T_{ge};

(2) the water vapour condensate accumulation rate M_{wc};

(3) the flow rate of the partially dried vent gas V_{gm} at the temperature of T_{gm};

(4) the fraction, by volume (f_{O2}) of oxygen in the vent gas.

With the data obtained, it was possible to calculate:

(1) the mass of oxygen consumed by the reactor, O_c (kg O/h), at a given oxygen feed supply rate, O_s (kg O/h);
(2) the oxygenation efficiency of the Vitox system, O_{eff};
(3) the respiration quotient, Y_{CO2} (i.e. mole CO_2 generated per mole O_2 consumed);
(4) the degree of water saturation of the vent gas.

Items (1) to (3) are calculated thus:

$$O_c = O_s - \frac{380.7 \times f_{o2} \times V_{gm}}{273 + T_{gm}} \quad (kg/h) \quad (4.4)$$

$$O_{eff} = \frac{O_c}{O_s} \times (100as\%) \quad (4.5)$$

$$Y_{co_2} = \frac{380.7 \times (1 - f_{o_2}) \times V_{gm}}{O_c \times (273 + T_{gm})} \quad (4.6)$$

Where:

O_c	=	The mass oxygen consumption rate of the reactor (kgO/h)	
O_s	=	Mass oxygen supply rate to the reactor	(kgO/h)
f_{O2}	=	The volume fraction of oxygen measured in the vent gas passing through the Parkinson Cowan gas meter	
O_{eff}	=	The fractional oxygen transfer/utilization efficiency	
V_{gm}	=	The volumetric flow rate f vent gas measured by the Parkinson Cowan gas meter	(m^3/h)
T_{gm}	=	The temperature of the vent gas passing through the Parkinson Cowan gas meter	(°C)

The derivation of these equations, as well as the more complex treatment of (4), the degree of water saturation of the vent gas, is described in detail in Part 2, Appendix 3B.

4.2.4 Anaerobic digester

4.2.4.1 Digester structure

An old decommissioned 1000 m^3 domed concrete cold anaerobic digester was refurbished and used as the second stage of the dual digestion plant. The digester was in fact far too large for the second stage of the dual digestion plant, but as it was available the decision was made to use it, but to control the operating volume by limiting the sludge depth in the unit. The effect of different retention times could then be investigated by simply varying the operating sludge depth.

Numerous cracks and fissures in the structure were sealed using a commercial product (Barralatex 100) and when this had been satisfactorily accomplished, the interior of the vessel was epoxy coated (ABE 356 Epoxy Tar). The gas dome and its equipment was also completely refurbished.

The digester proved to be gas tight but further structural repairs had to be undertaken during early 1990 after the unit had been emptied and cleaned.

4.2.4.2 Sludge feeding and discharge

The hot aerobically treated sludge was fed to the digester in batches of 3,75 m^3 at intervals which were determined by the aerobic stage cycle settings. During the test period the time between batches varied from 1 to 3 hours. The sludge was pumped from the aerobic reactor via about 75 m of 90 mm HDPE pipe, including a 14 m section of metal pipe which acted as an anti-siphon loop, to the top of the anaerobic digester. This arrangement caused severe problems during initial trials when the hot sludge ($\pm$ 60°C) falling from the top of the digester to the sludge surface caused rapid expansion of the gas volume in the digester, thereby repeatedly blowing out the 150 mm water seal on the gas dome. To overcome this problem a down pipe was fitted to feed the sludge to below the surface of the sludge in the digester. To avoid any back siphoning, a T-piece was fitted to the top of the down pipe.

Sludge was pumped daily from the digester (except Sundays) to restore its operating level. In effect, the retention time was never fixed, but varied as the level in the digester increased and decreased daily.

The sludge level in the digester was gauged by means of a large bore external sight glass, suitably calibrated and equipped with a clean water flushing system. To carry out the daily desludging operation, the operator simply pumped sludge to the drying beds, or the Zimpro storage tank, until the correct level was restored.

4.2.4.3 Mixing

Owing to the large diameter (16,7 m) and the very shallow operating depth, which depended on the retention time required e.g. 1,7 to 3,4 m for retentions between 10 and 20 days with the aerobic reactor operating at 1,25 day retention, thorough mixing was seen as a problem. Two mixing systems were provided. A 7 kW Flygt submersible propeller mixer was installed on a mounting bracket of which the direction could be externally adjusted. Although it was installed as near to the digester floor as feasible, it could only be used when the sludge level exceeded 1,6 m, otherwise severe cavitation would occur. When the sludge level was too low to use this unit, mixing was effected by means of the external sludge pump which, apart from being employed to pump sludge to the drying beds or to storage, could also recirculate the digester contents. This pump drew sludge from the centre of the floor of the digester and discharged it near the side wall.

Initially the propeller mixer was run continuously, but later it was operated by means of an adjustable timer. Running the mixer for 20 minutes in every hour provided adequate mixing under most circumstances.

It was found that, in spite of its good performance, using a mixer located inside the digester had serious drawbacks. When the unit failed (19 April, 1990) the digester

had to be emptied out and flushed before repairs could be effected. The down time incurred, including a lengthy acclimatization period, must necessarily be a costly exercise.

4.2.4.4 Heating and temperature monitoring

All the heat required to maintain the anaerobic digester at mesophilic temperatures was derived from the hot (± 60°C) sludge transferred batchwise from the aerobic reactor. Temperature probes located at the ends of the transfer line were used to monitor the heat loss during transfer. They also gave an excellent indication of the exact time of feeding the digester as well as indicating whether a batch or batches of hot sludge had been bypassed to the Zimpro tank. A single temperature probe located on the side wall about one metre above the digester floor was used to monitor the digester temperature.

Generally it was attempted to maintain the digester temperature at about 38°C. During the latter half of summer, when ambient temperatures frequently exceeded 30°C, the rate of heat loss from the digester to the ambient surroundings was too low to cool the digester contents sufficiently and as a result the sludge temperature reached 44°C on occasions. In the absence of any means to cool the feed sludge and without the use of a reactor feed to effluent sludge heat exchanger, the only way to limit the heat input to the digester was to reduce the flow of hot feed sludge from the reactor. This was done by bypassing a number of batches of hot sludge to the Zimpro storage tank. In summer it was thus not possible to maintain short stable retention times in the digester. In an attempt to maintain as short as possible anaerobic retention times, the digester volume was reduced as far as possible. It could, however, not be reduced below 380 m^3 without adversely affecting the operation of the submersible mixer.

A simple jacketed batch heat exchanger was manufactured and installed during early 1990. This would have been used to transfer heat from the aerobically treated sludge to the cold incoming feed sludge from the thickener. Unfortunately the unit was seriously damaged while undergoing hydraulic tests. By the time it had been repaired the aerobic reactor had collapsed so it was not used during this investigation. The heat exchanger would have enabled a proper evaluation of the minimum retention required for stabilization to be carried out under long term, stable conditions.

4.2.4.5 Biogas monitoring

The volume of biogas produced in the digester was continuously monitored using a large capacity positive displacement totalizing town's gas meter. Initially the gas was led directly from the gas dome to the gas meter which was located in a hut at ground level. The meter soon failed as a moisture trap installed ahead of the meter proved ineffective and the unit was clogged and suffered severe corrosion.

After being overhauled the meter was reinstalled, but this time a white painted metal pipe was used from the gas dome to the meter (approximately 16 m long)

to promote condensation before entering the water trap. The fairly dry gas was led to the gas meter via a metal pipe around which a heating element was wound. This raised the temperature of the gas to well above its dew-point and effectively prevented any further condensation in the meter.

The meter was read daily and the daily gas production calculated by difference after applying a correction for temperature and pressure.

The composition of the dry biogas was checked only on two occasions, using an ORSAT apparatus, by absorbing the carbon dioxide and assuming that the balance who is virtually all methane. The composition was found to be 33% CO_2 and 67% CH_4, as expected from a properly operating anaerobic digester.

4.2.4.6 Digester start up

Starting up the anaerobic digester without seeding it with sludge from an active anaerobic digester proved to be a problem. It was not possible to seed it as this would compromise its disinfected state and it would take a considerable time for organisms introduced during seeding to wash out of the system.

The start up procedure during summer and winter will now be described in some detail to show how problematical it was without some form of supplementary heating or cooling.

The first time start up was attempted the empty digester was filled with nitrogen before hot sludge was introduced from the aerobic reactor. Initially the water seal was repeatedly blown out of the gas dome due to the rapid expansion of the cool gas in the digester when hot batches of sludge were introduced. This problem resolved itself after some time, but a down pipe was later fitted, as described earlier, to prevent recurrence of this difficulty. In spite of the sludge being "pre-treated" by the aerobic process and having a high initial alkalinity and pH, no gas production was evident after one month, and feeding had to be stopped as the sludge temperature was becoming too high, volatile acid levels were increasing and the alkalinity had dropped alarmingly. Over the next three weeks lime, sodium bicarbonate and sodium carbonate were added daily to the digester. In total it required the following to restore the pH and alkalinity to acceptable levels (pH 7,4 and total alkalinity of 4463 mg/L):

3250 kg $Ca(OH)_2$
200 kg $NaHCO_3$
800 kg Na_2CO_3

After re-establishing suitable conditions for methanogenesis, gas production started and feeding was gradually increased and the digester was functioning normally 47 days later.

The entire start up procedure had taken 96 days. It was expected that starting up a "sterile" digester with thermophilically pretreated sludge would not be a straight forward operation, but the difficulties were compounded by the fact that the sludge temperature could not be maintained without feeding the digester. The

converse was also true - the digester could not accept all the aerobically treated sludge from the reactor without overheating during summer. The need for means of controlling the temperature of the effluent sludge from the reactor was evident.

The next time the digester was started (25 May 1990) it was nearly winter and difficulties were again experienced in getting the anaerobic process underway. Before sludge was introduced, 500 kg lime was spread in the digester in an attempt to prevent the initial pH drop. After 26 days of feeding 30 to 36 m^3 sludge daily, the pH had dropped and the temperature had only risen to 25°C due to the increased heat loss at that time of year. During the next 21 days cold sludge was drained from, and hot sludge and lime added to the digester, a total of 2150 kg lime and 1000 kg sodium carbonate being added. In spite of these efforts, the temperature was down to 22°C after 47 days with no sign of gas production.

During the next 56 days, periodic feeding and addition of another 600 kg sodium carbonate maintained the pH to between 6,6 and 7,0 and very slight gas production (<50 m^3/d) was evident. On day 103 after start up, it was decided to attempt to increase the sludge temperature, so all the batches of hot sludge produced by the reactor were fed to the digester. On day 108 the sludge level was dropped to reduce the volume from 640 to 340 m^3 to get a greater effect from the hot sludge being added. By day 113 the temperature had risen from 19°C to 30°C. Another 825 kg lime and 175 kg sodium carbonate was added, but the pH still dropped to 6,69. Feeding was stopped till day 118, and a drop in VFA's from 4400 to 3600 mg/L was noted, while the pH rose to 6,79. More lime was added to raise the pH to 7,1 and feeding resumed on day 118.

From day 120 a steady improvement was noticed as the VFA level was slowly decreasing and gas production increasing. The feed rate was cut by half from day 125 and during the next 20 days the VFA's decreased from 2005 to 1329 mg/L with gas production never below 80 m^3/d at 25°C and occasionally as high as 220 m^3/d at 20°C. By day 148 the VFA level was down to 55 mg/L and the pH had reached 7,32, so the feed rate was doubled to 30 m^3/d. On day 153 the feed rate was increased to the full output of the aerobic reactor (36 m^3/d).

From then on until the structural failure of the aerobic reactor (on 23 November, 1990, 183 days after restarting) full capacity operation was maintained, with the VFA level between 70 and 230 mg/L and gas production between 200 and 400 m^3/d.

From the experiences described above it is clear that starting up the anaerobic digester from scratch, without seeding, is a long and frustrating process. If some sludge could have been retained in the digester from the previous run to act as inoculum, even if it had been cold and inactive, the start up would probably have been more rapid and much easier to manage.

4.2.4.7 Final sludge dewatering

The structural failure of both the heat exchanger and later the aerobic reactor put paid to the proposed final sludge dewatering study which was to be performed once stable retention conditions had been established in the anaerobic digester. Practical experience with drying the sludge on the drying beds and the results of a laboratory investigation on the characterization of the final sludge can, however, be reported.

The final effluent sludge was extremely fine and thin, with the consistency of black ink. During dewatering on the sludge beds, the sludge neither attracted flies nor generated offensive odours. Its odour was characteristic of a well stabilized anaerobically digested sludge. Although it settled well, it did not exhibit good dewatering characteristics on the drying beds as the very fine sludge particles tended to blind the bed surface.

In spite of the high degree of volatile solids reduction of about 55% by the dual digestion process, and its apparent stability, the sludge continued to generate gas after being discharged to the drying beds. This gas generation floated some of the sludge to the surface, forming a "crust", while the remainder settled and blinded the drying bed media. Sandwiched between the upper and lower sludge layers was a relatively clear supernatant layer. A system was devised, consisting of a series of small open containers placed in the drying bed media, projecting to just above the normal lower sludge level, with drain pipes to decant the middle clear layer to the main drain channel. This clear layer could be decanted within a day of the sludge being discharged to the beds. During hot, dry weather the remaining sludge would dry reasonably well so that it could be manually removed from the beds 6 to 8 days after being discharged onto the beds.

Laboratory characterization tests, which gave some indication of the response to different dewatering methods, were performed on samples of the aerobically treated final sludge. The results are reported in the next chapter, but these confirmed the impression that the sludge was very difficult to dewater using filtration based technology.

CHAPTER 5

RESULTS AND DISCUSSION

5.1 INTRODUCTION

During the period 1987 to 1990 a vast amount of data was gathered on the operation of the dual digestion plant at Milnerton which will not be presented in detail, but instead the general trends and significant findings will be discussed. A large part of the research effort was directed at attempting to get a better insight into the operation of the aerobic reactor, more specifically on all the aspects of biological heat generation and means of predicting and controlling it. A detailed exposition on the stoichiometry and kinetics of biological heat generation in the aerobic stage of the dual digestion process is presented in Part 2 of this Final Report, but some of the more general aspects are also presented in this Chapter.

In Appendices 2, 3 and 4 some of the operating, analytical and microbiological data are given in more detail. Detailed results are also presented in the Appendices of Part 2.

5.2 AEROBIC REACTOR

Tests on the aerobic reactor started on 6 March 1987, but development work on the reactor and oxygenation system took almost 9 months. It was only by 18 November 1987 that the aerobic reactor was functioning reliably. At that stage most of the teething problems had been solved and continuous testing was possible, with only relatively short interruptions for maintenance and repair of malfunctioning equipment.

The modifications and repairs to the anaerobic digester was completed by the beginning of February 1988. On 10 February the effluent sludge from the aerobic reactor was transferred to the digester for the first time. Prior to that, the aerobic sludge had been fed into the sludge storage tank of the Zimpro plant for further treatment.

The initial operation of the aerobic reactor during the developmental stage had been very erratic, with frequent stoppages for modifications and repairs. The results of tests during this period are not reported on, as they did not represent steady operation of the aerobic process.

A summary of the first period of continuous operation from 18 November 1987 until 9 January 1988 is presented in Appendix 2A. The salient details of this first continuous run are given in Table 5.1.

5.2.1 Process Reliability

The general reliability of the aerobic stage is demonstrated by the fact that during the period covered in Table 5.1 the plant operated for 715 days out of a possible 783 days (91,3% on line). This was achieved in spite of stoppages for further modifications, inspections and for experimental reasons, which would not normally

occur on a production plant. The one item that did cause frequent problems, especially during the early stage of the project, was the Vitox recirculation pump. But after having been suitably modified it would run quite reliably for five to six months before requiring an overhaul.

5.2.2 Reactor Temperature

Table 5.1 and Appendix 2A show that the average reactor sludge temperature was 61°C. The apparently large temperature fluctuations evident in Appendix 2A and illustrated graphically in Appendices 2D to 2G were not the result of instability of the aerobic process, but were in most cases due to the requirements of the experimental procedures.

The reactor operating temperature was easily controlled by the oxygen supply rate and would remain at a "constant" level, allowing for the approximately 3°C saw tooth fluctuation, for lengthy periods. After some operating experience had been gained in operating the aerobic reactor, a mathematical model was developed to simulate temperature changes with time (Refer Part 2, Chapter 6 of Final Report). Using this model, the temperature changes in response to a wide range of operating conditions could be accurately predicted. Actual operating temperatures versus temperatures predicted by the model are given in Table 5.2.

TABLE 5.1: Aerobic Reactor - Operating Data (18/11/87 to 9/1/90)

	Average Reactor Temp. (°C)	Average Retention (d)	Total Sludge Treated (m^3)	Oxygen Supplied (kg)	Oxygen Supplied to sludge (kg/m^3)	Power Supplied (kWh)	Power Supplied (kWh/m^3)
715	61	1,4	22396	285951	12,8	351500	15,7

TABLE 5.2: Predicted vs. Actual Reactor Temperatures

Retention R_H (days)	O_2 Supplied (kg/h)	$\%O_2$ Consumed	Feed Temp. T_{si} (°C)	Actual T_{se} (°C)	Predicted T_{se} (°C)
1,25	24,3	69,2	17,9	61,0	60,1
1,25	22,0	77,9	19,0	60,5	60,8
1,25	18,5	83,5	18,3	57,1	56,8
1,50	18,5	84,4	19,7	65,8	66,6
1,50	16,5	85,4	20,8	62,4	62,0
1,25	14,5	92,6	18,8	53,0	54,0
3,00	6,3	100,0	20,8	67,8	67,0

The temperature response to changes in the oxygen supply rate are discussed in Section 5.2.7.

5.2.3 Retention time in the reactor (R_H)

The retention time in the reactor was varied between 1,0 and 3,0 days. Most of these variations were due to research requirements, but during 1989/90 the retention period was often changed to try to accommodate the heat requirements of the anaerobic digester.

Steady operation at retention periods of from 1,25 to 3,0 d has been demonstrated under all conditions. Continuous operation at a one day retention has also been shown to be possible during summer, but during the winter when both the feed sludge and ambient temperatures are lower, operation becomes marginal. If some heat could be recovered from the effluent reactor sludge to heat the feed sludge, or if supplementary heat sources (e.g. the combustion of methane from the anaerobic digester) could be exploited, operation at one day retention or less, using pure oxygen, appears to be feasible all year round (Refer Part 2, Section 5.5.4).

A heat exchanger would have effected a considerable improvement in the flexibility and performance of the entire dual digestion plant.

By employing a relatively simple batch heat exchanger, which would transfer some of the heat from the batch of sludge leaving the reactor to the batch of cold influent sludge fed to the reactor, a number of benefits could be achieved, viz.

(1) Shorter retention times could be maintained.
(2) Oxygen limitation could be achieved more easily at lower oxygen transfer rates.
(3) Reactor operating costs could be reduced by lowering the oxygenation power requirements as well as reducing the oxygen consumption.

The heating requirements of the anaerobic digester must be considered if heat exchange is used. If the anaerobic digester is not well insulated, it may require all the available heat generated by the aerobic reactor to maintain mesophilic operation. At Milnerton, during the winter, all the heat of the reactor effluent sludge, with the reactor operating at 1,25 day retention and at 60°C, was required to maintain the digester at 38°C at 21 day retention time. During summer, a heat exchanger would have prevented the serious overheating problems that plagued the anaerobic digester while at the same time allowing shorter retention in both stages of the dual digestion plant.

The benefits to be derived from a batch heat exchanger were recognized and a unit was manufactured and installed (Refer Section 4.2.4.4). Unfortunately the unit was damaged during hydraulic testing and could not be evaluated during the contract period.

5.2.4 Oxygen Supplied and Consumed

From Table 5.1, the average mass of oxygen supplied to treat 1 m^3 of sludge was 12,8 kg. However, depending on the operating conditions,this value varied from 6,0 to 24,3 kg O/m^3.

The oxygen utilization efficiency, which is defined as:

$$O_{eff} = O_c/O_s \text{ (x 100 as \%)} \qquad (5,1)$$

Where:

O_c = Oxygen consumption rate (kgO/h)
O_s = Oxygen supply rate (kgO/h)

varied between 66 and 100% (Detailed results in Part 2, Appendix 4A). It was found that as O_s increased, O_{eff} decreased. This was most probably due to the Vitox oxygenation system transferring oxygen less efficiently as the oxygen supply rate was increased. At the higher supply rates it is likely that the oxygen utilization efficiency was also limited by the biological oxygen utilization rate (OUR) of the sludge.

Figure 5.1 illustrates the relationship between oxygen supplied, oxygen consumed and the oxygen utilization efficiency. The results used for this illustration were obtained during the heat yield tests undertaken as part of this investigation (Part 2, Section 4.5).

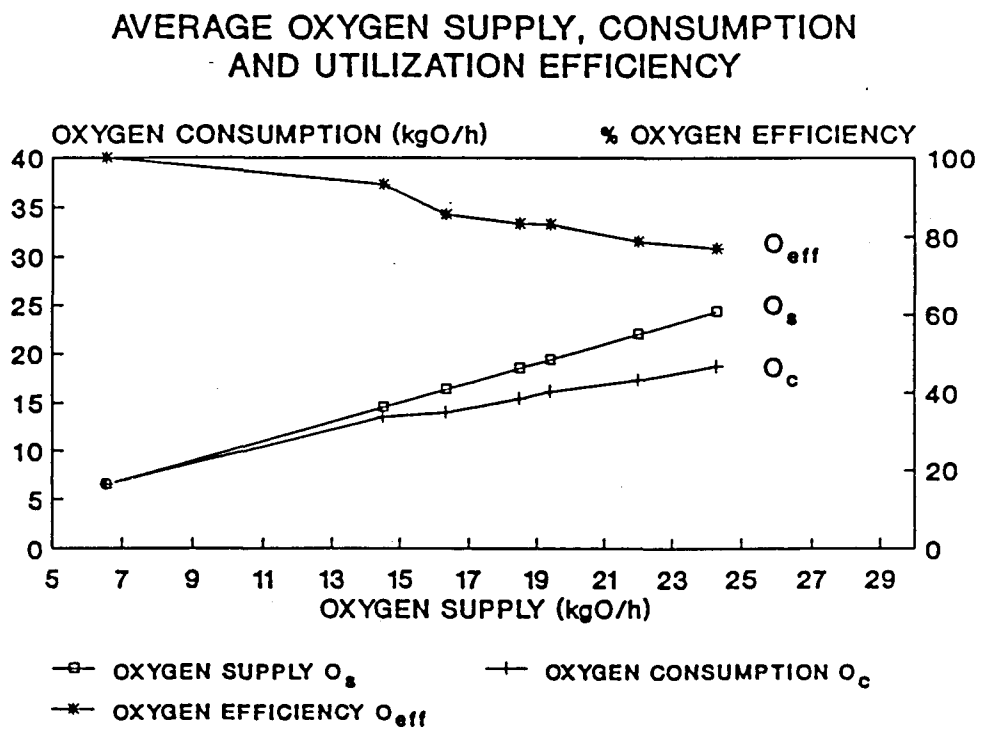


FIGURE 5.1: **The average oxygen consumption rate O_c, and average oxygen efficiency, O_{eff}, calculated for the 7 ranges of oxygen supply, O_s, which were employed during the 116 specific heat yield tests.**

5.2.5 Power Consumption

The amount of electrical power consumed by the three pumps of the aerobic reactor (i.e. feed, transfer and recirculation pumps) was accurately monitored over a period of 49 days while the reactor was operating at retention times of 1,75; 1,25 and 1,0 day. Details of the results are shown in Table 5.3.

The Vitox recirculation pump consumes by far the greater part of the total power used. The nominal power drawn by this pump is 20,5 kW, but some 17,4 kW of this goes to heating the sludge. As this pump operates continuously, it has a constant power consumption of about 492 kWh/d or 1772 MJ/d. The power consumed by the feed and transfer pumps vary according to the retention time in the reactor, as, for example, three times more sludge must be fed to and transferred from the reactor if the retention time is decreased from 3 days to 1 day. The actual variation in the overall power consumption is negligible as at one day retention, which would be the worst case, the recirculation pump consumes 98,07%, the feed pump 1,17% and the transfer pump 0,76% of the total power.

Over the 49 day period covered in Table 5.3, the average power requirement for the reactor was almost exactly 500 kWh/d or 1800 MJ/d.

TABLE 5.3: Aerobic Reactor - Power Consumption of Pumps

Period 1987/1988	Retention (h)	Hours run			kWh		
		Transfer	Feed	Recirc	Transfer	Feed	Recirc.
23/11 - 27/11	1,75	2,15	1,35	120,0	11,0	16,6	2460
28/11 - 4/12	1,75	2,88	2,83	165,8	14,7	34,8	3399
5/12 - 11/12	1,75 - 1,25	3,37	1,19	166,1	17,2	14,6	3406
12/12 - 18/12	1,25	4,74	2,75	167,4	24,2	33,8	3432
19/12 - 25/12	1,00	4,99	3,39	168,1	25,5	41,7	3446
26/12 - 1/01	1,00	5,32	3,42	167,9	27,1	42,1	3442
2/01 - 8/01	1,00	5,15	3,13	162,4	26,3	38,5	3330
9/01 - 11/01	1,00	2,17	1,54	70,0	11,1	18,9	1435
	TOTAL	30,77	19,6	1187,7	157,1	241,0	24350

5.2.6 Specific Heat Yield (Y_H) Determinations

During 1988 a total of 116 heat and oxygen mass balances were performed on the aerobic reactor. The operating conditions under which these balances were performed were varied widely, the reactor sludge temperature 54 to 69°C, retention times from 1,2 to 3,0 days, ambient temperatures varied between 8 and 30°C, oxygen supply rates from 0,13 to 0,44 kgO/(m^3.h), which produced oxygen transfer rates (OTR) or, equivalently sludge oxygen utilization rates (O_c) from 0,13 to 0,39 kgO/(m^3.h). Oxygen transfer efficiency varied from 0,80 to 1,00. Full results and details of these tests are given in Part 2, Chapter 4.

Table 5.4 gives the ranges of aerobic reactor conditions under which these tests were carried out. A broad outline of the heat and oxygen mass balance is also given in this report in Chapter 4, Section 4.2.3.

From each unsteady state heat balance (Equation 4.2) the biological heating rate was calculated from the experimentally determined values of H_{net}, H_{mi}, H_{ve} and H_{we}. In Part 2, Section 3.4.3 details are given for measuring these terms, while a sample calculation of the specific heat yield is given in Part 2, Section 3.5. The results obtained for these terms, as well as the calculated biological heating rate (H_{bi}) for all 116 tests are illustrated in Figure 5.2, the data being ranked in order of increasing H_{net} (rate of sludge enthalpy increase).

In Figure 5.3, the biological heating rate H_{bi} is plotted against the oxygen consumption rate O_c (see Equation 4.4) which was determined at the same time. The very well defined direct proportionality between H_{bi} and O_c is clear, with very little deviation from the best fit line drawn through the plotted points.

The constant of proportionality between H_{bi} and $O_c$is the specific heat yield Y_H. From the 116 specific heat yield tests, Y_H was found to vary between 11,4 and 14,6 MJ/kgO, with an average value of 13,1 MJ/kgO. This value compares closely with thermodynamically and bioenergetically calculated values (McCarty, 1972; Kambhu, 1971; Wright, 1975). On pilot or full scale ATAD plants values for Y_H from 5,3 to 21 MJ/kgO have been reported (Trim, 1984; Wolinsky, 1985), but in these investigations the research was not specifically directed at determining the specific heat yield. In the absence of reliable experimental data on Y_H with which to compare the value of 13,1 MJ/kgO determined at Milnerton, the data was carefully checked by:

(1) Statistical evaluation of the Y_H data.

(2) Determining the effect of the reactor operating conditions such as reactor temperature, retention time and oxygen consumption rate on the Y_H data.

(3) The sensitivity of the Y_H data to experimental measurement errors.

This evaluation is described in Part 2, Sections 4.4.2 to 4.4.4.

From this evaluation it was concluded that the values determined for Y_H were reliable and accurate. A linear increase in Y_H with increasing oxygen limitation which showed up during the evaluation was therefore assumed to be real and not due to experimental error (Part 2, Section 4.4.3.3).

TABLE 5.4: Ranges of Aerobic Reactor Operating Conditions

T_{se} (°C)	54	55	56	57	58	59	60	61
TESTS	4	2	9	6	7	17	9	13
T_{se} (°C)	62	63	64	65	66	67	68	69
TESTS	12	9	7	6	1	5	5	4
O_s (kg/h)	5,9 - 8,1		14,5		16,0 - 16,5		18,5	
TESTS	13		7		29		15	
O_s (kg/h)	19,4		22		24,3			
TESTS	13		28		11			
R_H (days)	1,2		1,25		1,5		3	
TESTS	25		47		31		13	

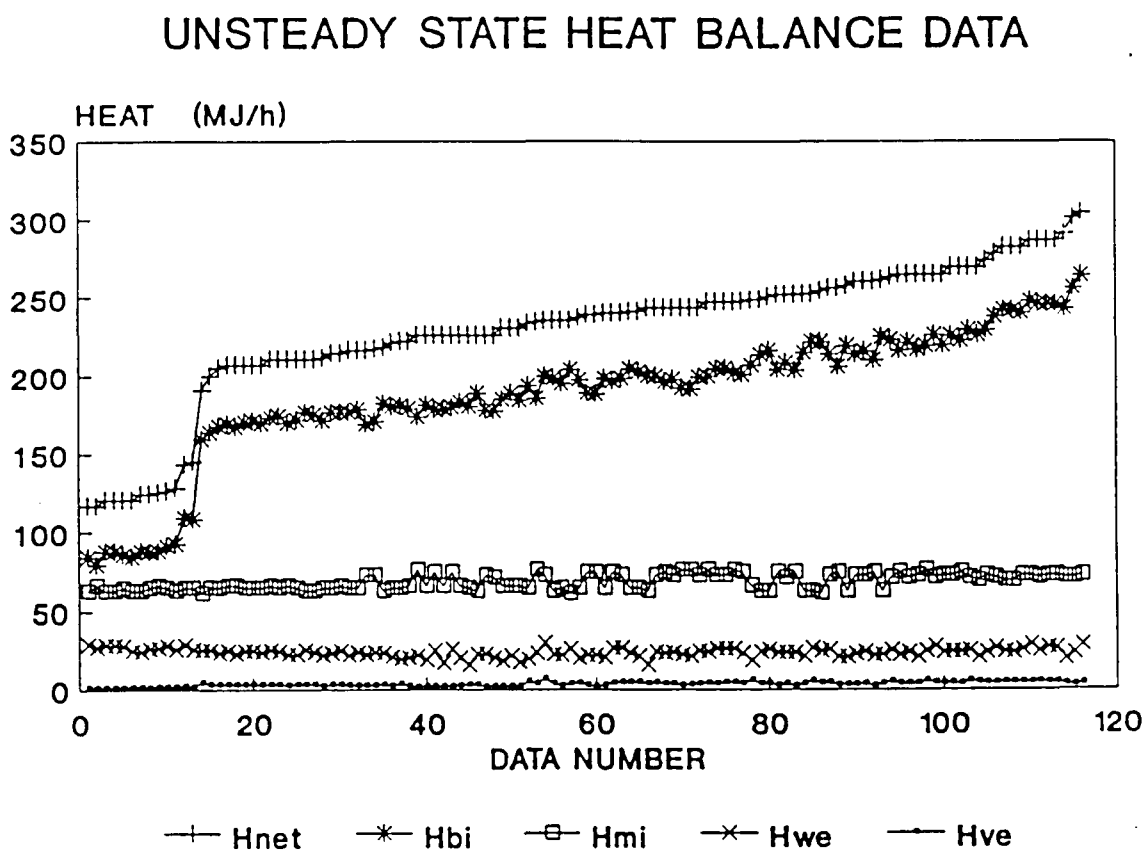


FIGURE 5.2: The data for the 5 components of the unsteady state heat balance, H_{net}, H_{be}, H_{mi}, H_{we} and H_{ve}, measured during the 116 specific heat yield tests and ordered in terms of increasing H_{net}.

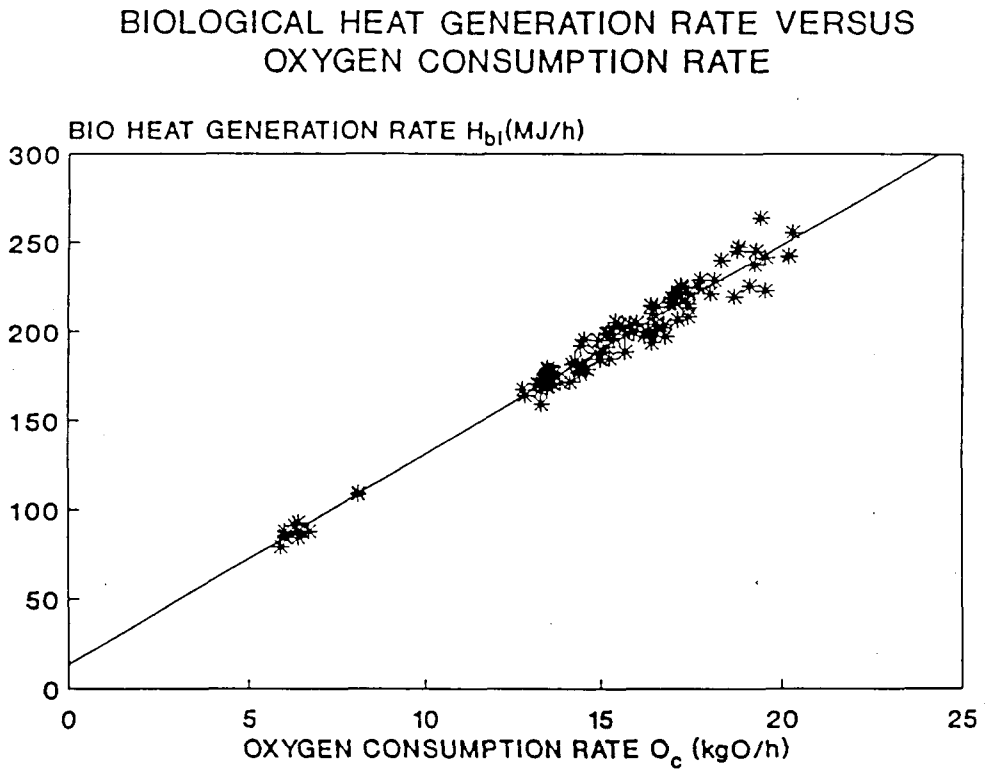


FIGURE 5.3: Plot of the biological heat generation rate (H_{bi}) versus the oxygen consumption rate (O_c) from those values measured during the 116 specific heat yield tests. Direct proportionality between H_{bi} and O_c is shown with the constant of proportionality being the specific heat yield Y_h (MJ/kgO).

5.2.7 Temperature Control - Aerobic Reactor (Part 2, Section 4.8)

The temperature of the sludge in the aerobic reactor could be completely and almost instantly controlled by means of the oxygen supply rate (OSR), for as long as the reactor was operated under oxygen limiting conditions, i.e. while the oxygen transfer rate (OTR) was less than the maximum oxygen utilization rate (OUR). For the Milnerton sludge, which is a blend of primary and humus tank sludge from a biofilter process, the maximum biological oxygen consumption rate for the 45 m^3 was approximately 17 kgO/h, which gives an OUR of about 0,38 kgO/(m^3.h) at an average volatile solids concentration of 30 kgVS/m^3. Step increases in the OSR as high as 330% (from 0,13 to 0,43 kgO/(m^3.h)), which increased the oxygen transfer rate and hence the biological oxygen consumption rate by 300% (from 0,13 to 0,39 kgO/(m^3.h)), caused an immediate increase in the biological heat generation rate (H_{bi}), which increased the reactor sludge temperature. Similarly, decreasing the OSR caused an immediate decrease in H_{bi}, resulting in a lower steady reactor sludge temperature.

Note that the increase in OSR did not produce an equivalent increase in the OTR, as the oxygen transfer efficiency (OTE = OTR/OSR) is decreased with increasing OSR (refer to Figure 5.1).

The fact that the reactor temperature can be accurately controlled by means of the oxygen supply rate on a reactor using a Vitox oxygenation system is considered to be of major significance in the design of future dual digestion plants.

5.2.8 Steady State Heat Balance (Part 2, Section 4.9)

For a steady state heat balance to be made across a reactor, all of the heat balance parameters must remain constant over the defined time interval of the test. As explained in Section 4.2.3.1 the aerobic reactor was judged to be at steady state when:

(1) the reactor sludge temperature at the start of a number of consecutive batch cycles remained the same; and

(2) the operating parameters of the plant had not been changed for at least 3 retention times.

By assuming that the batch fed reactor operating under the conditions defined above is continuously fed and that the steady state temperature of the reactor is equal to the temperature of the effluent sludge T_{se}, which is the highest reactor sludge temperature attained during a batch cycle, a steady state heat balance can be carried out.

The steady state heat balance is given by equating the heat sources with the heat sinks, i.e.

Heat sources = Heat sinks

$$H_{bi} + H_{mi} + H_{si} = H_{se} + H_{ve} + H_{we} \quad \text{(MJ/d)} \quad (5,2)$$

Where:

H_{bi}	=	biological heating rate	(MJ/d)
H_{mi}	=	mechanical heat input rate	(MJ/d)
H_{si}	=	feed sludge heat input rate	(MJ/d)
H_{se}	=	effluent reactor sludge heat loss rate	(MJ/d)
H_{ve}	=	vent gas water vapour heat loss rate	(MJ/d)
H_{we}	=	wall heat loss rate	(MJ/d)

For a reactor oxygenated with pure oxygen it is reasonable to accept that the influent and effluent sludge specific heats are the same. This is because the total or volatile solids suspended in the water medium make a negligible contribution to the specific heat (C_p) of the mixed liquor, so that any change in the solids does not affect the specific heat of the liquor. Also, by using oxygen, a negligibly small proportion of the water in the reactor is vaporized (<0,007%).

Accepting the influent and effluent sludge specific heats (C_p) and the volumetric flow rates to be the same, the net effluent reactor sludge heat loss rate is related to the temperature difference between the influent (T_{si}) and the effluent (T_{se}) sludges as follows:

$$H'_{se} = H_{se} - H_{si} = F_s \rho_s C_p(T_{se} - T_{si}) \quad \text{(MJ/d)} \quad (5,3)$$

Where:

H'_{se}	=	Net rate of heat loss in the reactor effluent sludge in respect of the influent sludge temperature T_{si}	(MJ/d)
ρ_s	=	density of the sludge liquor	(ton/m^3)
C_p	=	specific heat	(4,18 MJ/ton.°C)
F_s	=	sludge volumetric flow rate through reactor	(m^3/d)
F_s	=	V_p/R_H	
V_p	=	reactor operational volume	(m^3)
R_H	=	retention time	(d)

H_{se}' is not only the rate of heat loss in the reactor effluent sludge, but when compared with the unsteady state heat balance (Eqn 4.2) is also equal to the rate of increase of the sludge heat content, H_{net}, from feed sludge temperature T_{si} to reactor sludge temperature T_{se}. Another way of looking at it is that in the unsteady heat balance the 45 m^3 sludge volume, which was cooled about 3°C when the cold feed sludge was added, is reheated to the effluent sludge temperature, where as in the steady state heat balance the reactor sludge temperature is accepted as remaining constant at the effluent sludge temperature, and only the influent sludge volume is heated from the feed temperature (T_{si}) to the effluent sludge temperature (T_{se}).

By substituting Eqn (5.3) in Eqn (5.2), the steady state heat balance becomes:

$$H_{bi} + H_{mi} = \frac{V_p}{R_h} \rho_s C_p(T_{se} - T_{si}) + H_{we} + H_{ve} \qquad (5.4)$$

Taking one day as the unit of time, examples of the steady state heat balance calculated by equation (5.4) is illustrated in Figure 5.4 in the form of pie charts for 1,25 and 3,0 days retention. The conditions applied were based on the actual conditions of two of the heat yield tests at these retention times.

In Figure 5.4, the bottom half of each of the pie charts constitute the heat sources and the top half the heat sinks. The biological heat generation rate H_{bi} and the effluent sludge heat loss rate H_{se} constitute the largest source and sink respectively, and both decrease with increasing retention time. The values of H_{bi}, H_{se} and H_{ve} more than doubled with the reduction in the retention time from 3,0 to 1,25 days. In the case of H_{bi} and H_{ve} this was due to the increased oxygen supply

rate, and in the case of H_{se} due to the increased sludge flow rate F_s. In contrast, H_{we} decreased when the retention time was reduced from 3,0 to 1,25 days. This reduction (16%) was caused by the higher reactor temperature at the 3,0 day retention. Although it was attempted to maintain the reactor temperature at about 60°C for both tests, the higher (67°C) temperature at 3 days retention was a consequence of the inability of the smaller (low flow rate) rotameter to dispense less than 6,0 kgO/h accurately. More heat than required to maintain the reactor temperature at 60°C was therefore generated.

The calculated values of H_{bi}, H_{mi}, H_{we} and H_{ve} are plotted versus increasing H_{se}' in Figure 5.5, for the 65 specific heat yield tests that were done while the reactor was at steady state. Because the temperature difference between the transfer and feed sludges (T_{se} - T_{si}) did not change very much (35,6 to 48,6°C) during these tests, the change in retention time R_H represents the largest influence on H_{se}', which according to equation 5.3 decreased as R_H increased. Consequently in Figure 5.5, increasing $H_{se}'$is approximately related to decreasing R_H.

The interrelationships between $H_{se}'$and H_{bi}; H_{mi}; M_{we} and H_{ve} is important in the steady state design of the aerobic reactor and this aspect is discussed in Part 2, Sections 4.9.1 to 4.9.3.

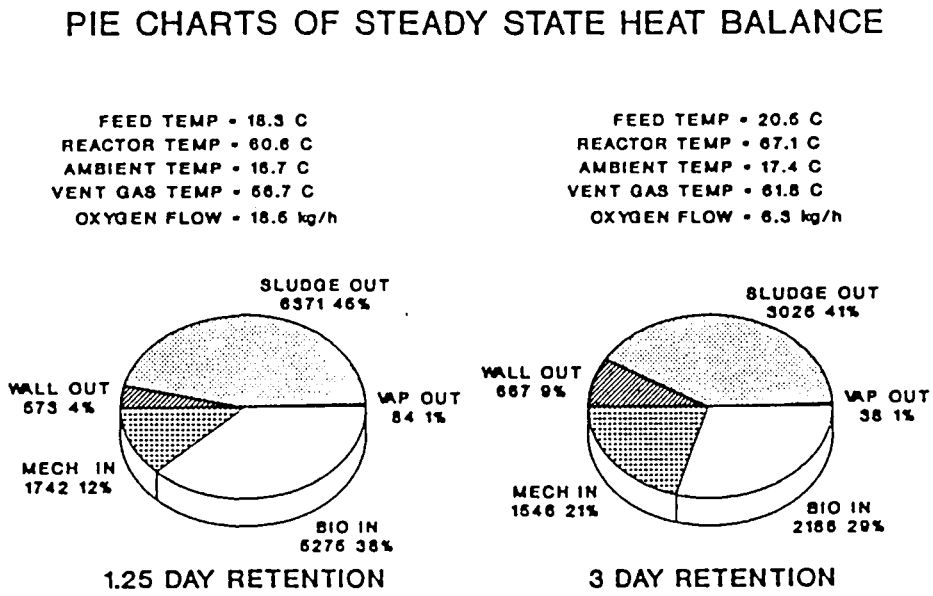


FIGURE 5.4: Two pie charts illustrating typical daily heat balance data for the aerobic reactor measured under "steady state" conditions at a 1,25 and a 3 day hydraulic retention time.

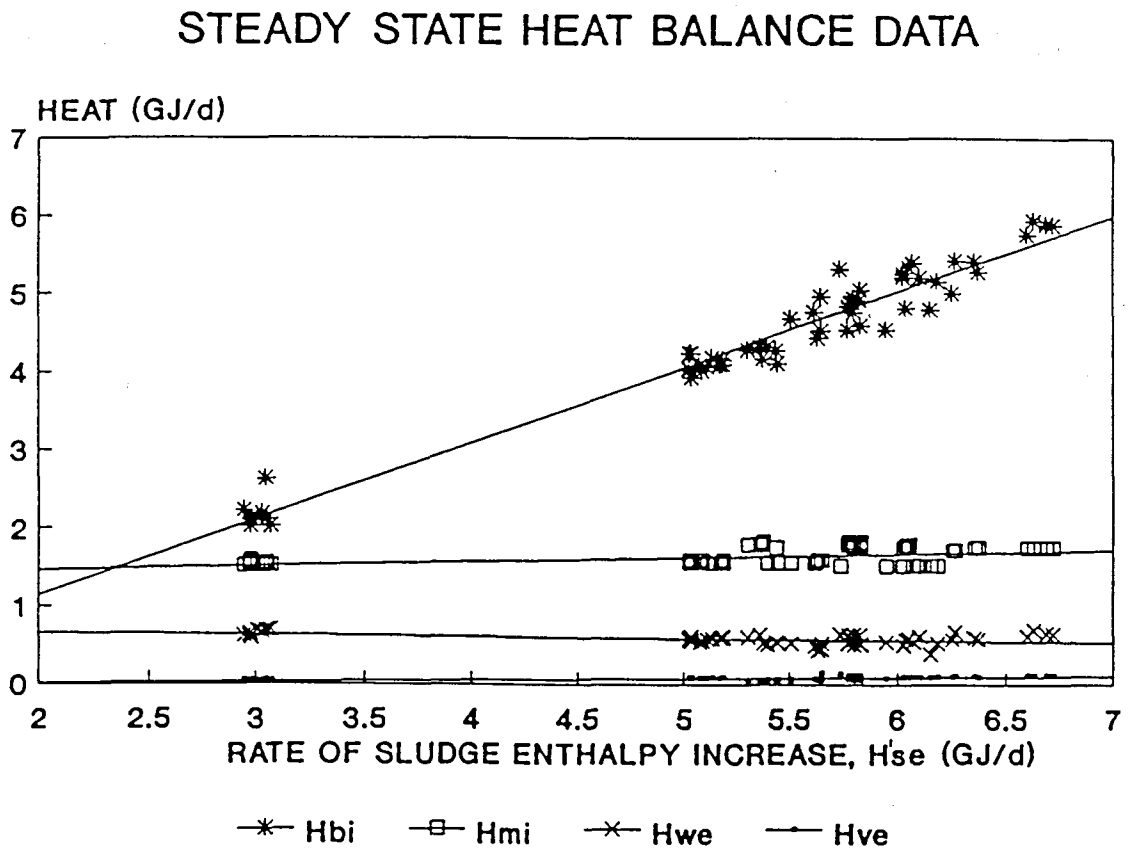


FIGURE 5.5: The daily rates of heat gain or heat loss for 4 of the 5 terms of the steady state heat balance across the aerobic reactor viz. H_{bi}, H_{mi}, H_{we} and H_{ve}. These 4 rates are plotted against the 5th term of the heat balance $H'_{s}e$, the net rate f sensible heat loss in the reactor effluent sludge. H'_{se} also denotes the net rate of heating required to maintain the reactor at the desired thermophilic operating temperature T_{se}. This data in this plot were derived from 61 of the 116 unsteady state heat balances which were conducted while the reactor was at "steady state".

5.2.9 Respiration Quotient (Y_{co2})

The respiration quotient, Y_{co2}, is the number of moles of carbon dioxide produced per mole of oxygen consumed by the reactor sludge. It is calculated from the oxygen mass balance data by applying Equation 4.6, as given in Chapter 4, Section 4.2.3.2. The Y_{co2} values obtained from the 116 heat yield tests are given in Part 2, Appendix 4A. These ranged from 0,53 to 0,85, with an average of 0,66 $\pm$ 0,04. The theoretical values of Y_{co2} found in the literature, based on the complete oxidation of "typical sludge" formulations such as, for example $C_5H_7O_2N$, to CO_2, H_2O and NH_3 at higher temperatures which inhibit nitrification, yield values of 1,0 or very close to 1,0. The data was statistically examined but this did not show up any reason for not accepting the value 0,66 as valid.

Possible causes for the low Y_{co2} value was considered, i.e.:

(1) The actual substrate is different from the accepted formulations for a typical sludge.

(2) Some of the CO_2 produced remains in solution in dissolved state as a result of the increase in bicarbonate alkalinity in the reactor by ammonification of organic nitrogen in the sludge.

The second reason was dismissed because if it were true, it would tend to exercise a buffering effect on any sudden changes in CO_2 production and a relatively stable value of Yco_2 would be measured.

It was, however, found that in 6 of the 9 times that the oxygen supply was either increased or decreased, Y_{CO2} decreased immediately after the change and rose again over the next 24 hours.

Details of evaluation of the respiration quotient are given in Part 2, Section 4.6.

5.2.10 Volatile Solids and COD Removal

Although the purpose of the aerobic stage was not to remove VS or COD beyond the requirement for producing biological heat, this aspect was routinely monitored throughout the duration of the evaluation programme. An intensive study was also undertaken to determine whether the biological heat generated could be consistently and reliably related to VS or COD removal in the reactor.

Table 5.5 gives the average values of analyses performed from 30 March 1987 until 3 January 1990 on feed and transfer samples, while detailed analyses are shown in Appendices 3A to 3C for this period.

The results indicate that a relatively low reduction in both the TS and VS occurred in the reactor. The average reduction in TS was 4,3 g/L or about 11,3%, while the reduction in VS was 3,9 g/L or 12,8%.

TABLE 5.5: Aerobic Reactor - Average Analytical Results (30/3/87 - 3/1/90)

	Average of Analyses		Removal or Increase	
	Feed	Transfer	Change	Change %
No. of Analyses	59	60		
Total Solids g/L	37,65	33,38	-4,27	-11,3
Volatile Solids g/L	30,47	26,58	-3,90	-12,8
Volatile Solids %	80,93	79,63		
COD g/L	50,26	41,25	-9,01	-17,9
COD/VS Ratio	1,649	1,552	-0,097	-
Ammonia-N mg/L	174	278	+104	-
Alkalinity as $CaCO_3$				
(pH 5,75) mg/L	11	575	+564	-
(pH 4,3) mg/L	701	891	+190	-
pH	5,60	7,09	+1,49	

The removal of COD in the aerobic reactor was somewhat higher - the average reduction being 9 g/L or about 18% of the feed COD.

The observed average COD/VS ratio in the feed sludge was 1,649, while after aerobic treatment it was lower at 1,552.

When comparing these results with those obtained during the intensive specific heat yield evaluations reported in Part 2, Section 4.10, there are only minor differences. However, in the latter results the a considerably lower VS removal of only 1,3% of the sludge mass treated was found (Refer to Part 2, Section 4.10.1). The reason for this discrepancy is not clear, but after careful evaluation of the methods of sampling and analysis, no errors could be detected. This matter had not been resolved by the end of the evaluation, but owing to the small difference between the influent and effluent sludges it was in any case not possible to derive a meaningful relationship between the rate of biological heat generation and the VS removal in the reactor.

The average COD removal was about 18% for the routine sample results in Table 5.5. The heat yield results indicate a lower average COD removal of 11,9% (Part 2, Section 4.10.2). After careful statistical evaluation of the VS and COD removal data it was concluded that both the VS and COD removal were not suitable parameters on which to base biological heat generation for temperature control of the aerobic reactor in a dual digestion process.

In both the routine VS and COD analyses and the analyses performed during the heat yield tests, the aerobic reactor produced on average a larger reduction in sludge COD than in the sludge VS. The routine results (Table 5.5) indicate a VS removal of 12,8% and a COD removal of 17,9%. The heat yield tests (Part 2, Section 4.10.3) produced an average VS removal of 1,3% and a COD removal of 11,9%. The removal of more COD than VS causes the sludge COD/VS ratio to decrease across the reactor in both cases. For the routine analyses the mean COD/VS ratios for the feed and transfer sludges were 1,649 and 1,552 respectively, which amounts to a 7,7% decrease. For the heat yield results the COD/VS ratios were 1,721 and 1,548 for the feed and transfer sludges respectively, giving a reduction of 10%.

In models of autothermal thermophilic aerobic digestion which describe the rate of biological heat generation in terms of VS removal kinetics and the stoichiometry in terms of oxygen consumption, the COD/VS ratio forms the link between the kinetics and the oxygen consumption. This approach is inappropriate in the case of the aerobic reactor in dual digestion, as the COD/VS ratio changes through the reactor. To estimate the oxygen consumption on the basis of VS removal kinetics by assuming a fixed COD/VS ratio would produce incorrect results for aerobic reactor design.

5.3 ANAEROBIC DIGESTER

Aerobically treated sludge was fed to the anaerobic digester for the first time on 10 February 1988, but after a protracted start-up period described in Section 4.2.4.6 it was acclimatized and operational by 17 May 1988. The operating results for the period from 30 April 1988 until the process was shut down on 12 January 1990 are reported on a weekly basis in Appendices 2B and 2C. The averages of these operating results are shown in Table 5.6. Detailed analytical results for the dual digestion process appear in Appendices 3A, 3B and 3C. The latter two include the analytical data for the anaerobic stage. The average values for these analyses are presented in Table 5.7.

The temperature of the digester is also shown graphically, together with the other aerobic reactor temperatures, in Appendices 2D to 2G.

5.3.1 Operation and Control

During the first period of operation from the time that the digester was considered to be fully acclimatized until the end of November 1988, when the failure of the aerobic reactor recirculation pump resulted in a temporary shut-down of 5 weeks, operation was erratic. This was caused mainly by the inability to control the temperature and the retention period adequately. Failure of the gas meter, requiring the unit to be completely overhauled, and extensive modifications made to dry the gas before entering the gas meter, resulted in a 10 week period during which no gas measurement was possible. In spite of the considerable fluctuations in digester sludge temperature and retention time, no instability problems were evident and the performance was generally very good (See Table 5.7).

During the warm summer months it was found that the anaerobic digester temperature could be maintained by feeding about 20 m^3/d of aerobic sludge at 60°C, if the ambient temperature was 26°C or higher. This corresponds to a heat input of about 2850 MJ/d. However, the sludge heat output of the aerobic reactor when operating at, for example, 1,25 d retention and 60°C, was about 6022 MJ/d. The digester would therefore overheat rapidly if all the aerobically treated sludge was fed daily. In the absence of a heat exchanger or any other means of cooling the aerobic effluent sludge, a number of batches of hot sludge had to be diverted to the Zimpro storage tank daily. This made it almost impossible to maintain a constant sludge retention time and also resulted in significant temperature fluctuations in the digester.

TABLE 5.6: Anaerobic Digester - Averages of Operational Data (30/4/1988 - 12/1/90)

Period	Digester Temp. (°C)	Feed Volume (m^3/d)	Average Digester Vol. (m^3)	Gas. Prod. (m^3/d)	Average Gas/Feed Ratio	Average Retention (d)	Total Sludge Treated (m^3)
30/4/88 - 6/1/89	33,2	21,22	559	259	13,2	21,1	5347
7/1/89 - 12/1/90	37,7	26,67	539	359	13,93	21,7	9896

Another factor that influenced the operation of the digester was the effects of changes in the operation of the aerobic reactor during experimental procedures, e.g. during the heat yield tests when the temperature had to be changed quite often.

During 1989 (Appendix 2C) the general operation was much steadier, but the digester was still very prone to overheating during warm periods. The period from 7 January 1989 till 12 January 1990 is therefore the one which best represents typical operation of the anaerobic digester, and is the one to which will generally be referred when discussing digester performance.

5.3.2 Digester Temperature

As explained in the preceding section, the temperature of the digester sludge varied considerably during the initial period of operation. During 1989/90 it was more constant when the dual digestion process was operated more as a production process rather than an experimental one. During the colder months, it was possible to treat almost all the sludge from the reactor, i.e. about 250 m^3/week or 36 m^3/d without exceeding a sludge temperature of 40°C. Fairly large fluctuations of up to 2°C/d occurred regularly, but this appeared to have no effect on the digestion process other than causing fluctuations in the gas production.

The average temperature during the 1989/90 period was 37,7°C and 40°C was reached only on a few occasions before it could be controlled by diverting some of the aerobic sludge to the Zimpro storage tank (Refer to Appendices 2C, 2F and 2G).

TABLE 5.7: Anaerobic Digester - Average Analytical Results (17/5/88 - 24/11/88)

		Transfer	Anaerobic Digestion	Change	Change %
No of Analysis		17	17		
Total Solids	g/L	35,00	20,1	-14,9	42,6
Volatile Solids	g/L	27,90	14,1	-13,8	49,5
Volatile Solids	%	79,8	69,8	-	-
COD	g/L	43,5	23,8	-19,7	45,3
COD/VS Ratio		1,559	1,688	+0,129	-
Ammonia-N	mg/L	250	656	+406	-
Alkalinity as $CaCO_3$					
(pH 5,75)	mg/L	529	1942	+1413	-
(pH 4,3)	mg/L	860	2575	+1715	-
IA/PA Ratio		-	0,30	-	-
Volatile Acids	mg/L	-	208	-	-
pH		7,00	7,50	+0,5	-

5.3.3 Retention Period in the Digester

The operating level of the sludge in the digester, which had a direct influence on the retention period, fluctuated considerably, both on a daily and on a weekly basis. As the hot sludge was fed to the digester in batches of 3,75 m^3 approximately ten times daily and an approximate equivalent total amount was withdrawn as a single batch once daily, this mode of operation resulted in a 5 to 7% fluctuation in the operating volume. An average retention time, taking into account the volume fluctuation, was used for all calculations. On most Sundays, no sludge was withdrawn so a 10 to 15% occurred over weekends. Clearly this was not an ideal mode of operating the digester, but it did not appear to have an adverse effect on the process other than complicating the retention time calculations.

To have operated at retention times of less than 12 days with an aerobic sludge feed of 36 m^3/d would have been marginal at best due to the operating level in the digester being very close to the lower limit at which the mixer could operate.

For the greater part of the year the retention time was governed by the temperature restriction and retention times of around 21 days was the norm. During a period of very cold weather from August to September 1989 it was attempted to operate the digester at a retention time of 12 days. The average operating volume was reduced to about 450 m^3 and with a daily feed from the aerobic reactor of 37,5 m^3 (1,2 day retention in the reactor) an approximate retention time of 12 days was obtained. Owing to an increase in volatile acids (VA) and a continuing fall in the gas production, which are signs of incipient digester failure, the operating volume was increased and the feed reduced to allow the process to recover after 5 weeks at 12 days retention.

Except for this short period, the average retention was 21,7 days (Table 5.6). At 15 days retention, which was obtained during July, 1989, operation appeared to be quite stable, with normal gas production and low VA's (Refer Appendices 2C and 3C, from 8/7/89 to 4/8/89).

It appears that retention times of 15 days in the anaerobic digester produces stable operation, but that a 12 day retention results in instability and all the typical signs of overloading are evident, even under the favourable mesophilic conditions maintained in the digester.

5.3.4 Gas Production

The quantity of gas produced by the digester varied considerably, but was usually between 13 and 15 m^3 at 25°C per m^3 of feed sludge under stable operating conditions. At an average VS removal of 12,6 kgVS/m^3 (Table 5.8), this translates to a specific gas production of between 0,94 and 1,09 m^3/kgVS removed (STP corrected), which is as good or better than the specific gas production of normal mesophilic (heated) anaerobic digesters which produce from 0,75 to 1,1 m^3/kgVS destroyed (USEPA, 1979).

The biogas produced was accurately measured by means of a positive displacement Parkinson Cowan town gas meter. Prior to being measured, the gas was cooled to condense the moisture and reheated to about the same temperature as the digester sludge. A temperature correction was applied in all calculations, but it was assumed that the actual measurement was at atmospheric pressure.

The gas composition when measured appeared to be very normal, with 33% CO_2 and 67% CH_4, which, according to the literature (USEPA, 1979) is typical of healthy anaerobic digestion.

During the period of incipient failure while operating at 12 days retention (Refer 5.3.3) a significant decrease in gas production occurred, the gas/feed ratio falling from a value of more than 12 to around 8 m^3/m^3 (Refer to Appendix 2C,

weeks 31 to 38). At the same time a downward trend in VS and COD removal was noted. The specific gas production under these conditions was approximately 0,67 m^3/kgVS destroyed.

Although the biogas produced by the digester was not used at Milnerton, it was calculated that combustion of the gas having a composition of 67% methane and 33% carbon dioxide to give combustion products CO_2 as gas and H_2O as liquid, both at 65°C, would produce 26,2 MJ/m^3 of energy. In the literature (USEPA, 1979) the heat value for digester gas is given as a "high" heat value of 26,2 MJ/m^3 for combustion in a calorimeter and as a more realistic "low" heat value of 23,8 MJ/m^3, which is the heat value obtained when none of the water vapour formed is condensed. The latter value is used for the determination of gas engine efficiencies.

Using the low heat value and assuming a daily gas production of 350 m^3, the digester generates about 8330 MJ/d of energy which could be applied for supplementary heating of the feed sludge to the aerobic reactor or for other purposes at the treatment works.

5.3.5 Chemical/Physical Analyses

From the average values presented in Tables 5.7 and 5.8, it is clear that the anaerobic digester effects a large reduction in VS (49%) and COD (44%). This performance was achieved at more than double the retention time claimed possible for an anaerobic digester operating in the dual digestion process. Plotting the COD values of the digested sludge against the retention time (Figure 5.6) produced a wide scatter of points, but the best fit line confirmed the trend observed from the results in Appendix 3C, i.e. that the COD of the treated sludge would increase to between 30 and 35 g/L at retentions of 12 days or lower. Similarly, the gas/feed (m^3/m^3) ratio, depicted in Figure 5.7 shows that the quantity of gas produced per m^3 of sludge treated dropped considerably as the retention decreased. Both these trends indicate that a considerably lower COD removal could be expected at the lower retention times claimed (10 days or less) for the process, if the stability of the digestion process could be maintained at these times.

At the average retention time of 21 to 22 days (Table 5.6) the average COD removed was 17,7 g/L or 44%. From Figure 5.6, by extrapolation it appears if the COD removed could drop to about 10g/L (from 40g/L in the feed to 30 g/L in the treated sludge) which is a 25% reduction at a retention of 10 days.

This appears to be much too low for effective stabilization, as a minimum reduction of 38 to 40% in VS or COD is required in most of the new regulations (refer to Chapter 2, Section 2.6.3).

TABLE 5.8: Anaerobic Digester - Average Analytical Results (31/1/89 - 17/1/90)

	Transfer	Anaerobic Digestion	Change	Change %
No of Analysis	22	26		
Total Solids g/L	32,3	18,4	-13,9	43,0
Volatile Solids g/L	25,7	13,1	-12,6	49,0
Volatile Solids %	79,8	70,8	-	-
COD g/L	40,2	22,5	-17,7	44,0
COD/VS Ratio	1,564	1,718	+0,154	-
Ammonia-N mg/L	318	651	+333	-
Alkalinity as $CaCO_3$				
(pH 5,75) mg/L	637	1733	+1096	-
(pH 4,3) mg/L	950	2260	+1310	-
IA/PA Ratio	-	0,38	-	-
Volatile Acids mg/L	-	298	-	-
pH	7,1	7,6	+0,5	-

The stability of the digestion process was very good while operating at retention times of 15 days or more. The pH and the bicarbonate alkalinity (approximated by titration to pH 5,75) remained high while the volatile acid levels remained low. The exceptions were during the start-up phase and when operation at 12 days retention was attempted. At 12 days retention, the VA's increased progressively from about 150 mg/L to 1686 mg/L after 5 weeks (Appendix 3C). On reducing the feed it rapidly decreased again to its normal level.

One of the best indicators of instability was the IA/PA ratio, defined as the ratio of the Intermediate Alkalinity divided by the Partial Alkalinity (Ripley et al., 1985). These terms are:

PA = Partial alkalinity = Determined by titration from sample pH to pH 5,75. This corresponds approximately to the bicarbonate alkalinity.

IA = Intermediate alkalinity = Determined by titration from pH 5,75 to pH 4,3. This approximates the volatile acid alkalinity.

An IA/PA ratio of lower than 0,4 is supposed to indicate good buffering and stable digestion.

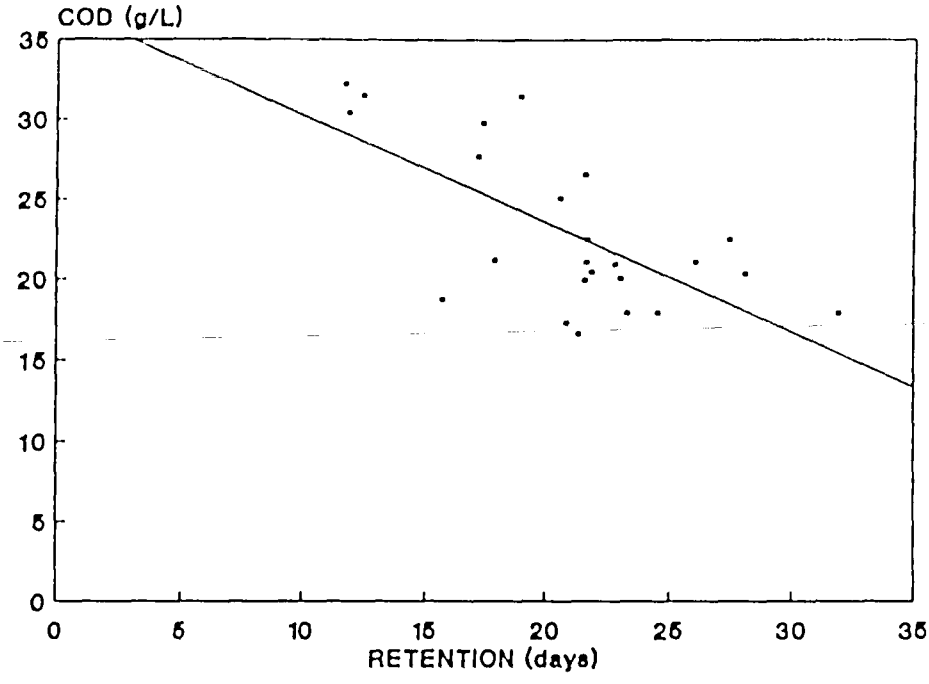


FIGURE 5.6: Treated sludge COD vs Retention.

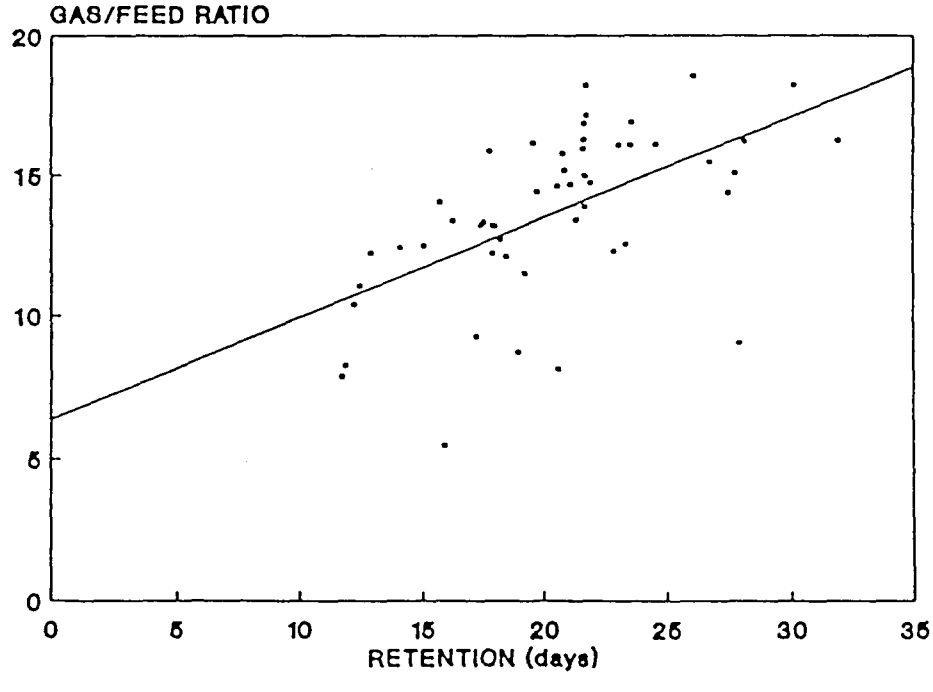
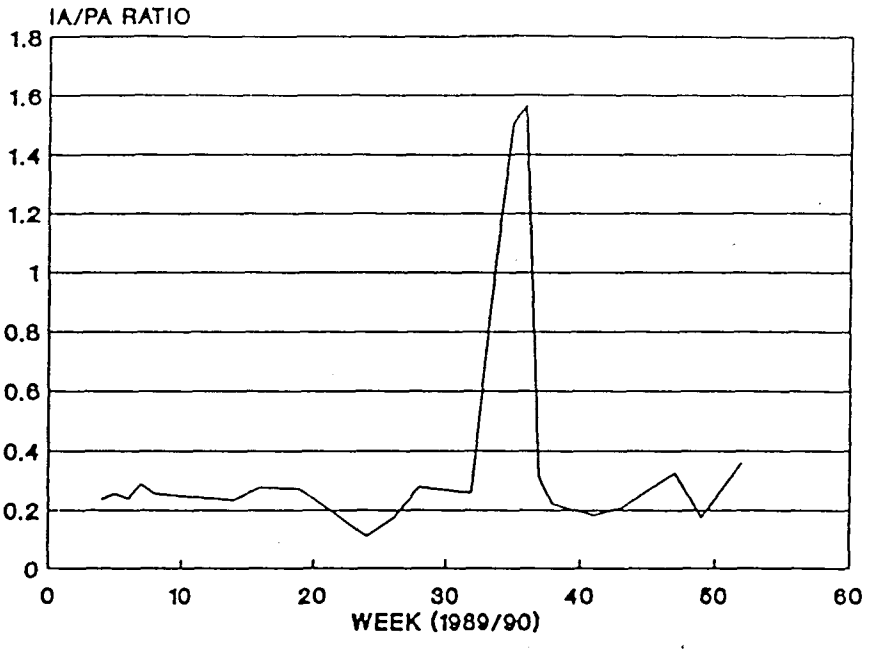


FIGURE 5.7: Gas/Feed Ratio vs Retention.



IA/PA Ratio

PA = Partial Alkalinity - Determined by titrating from original sample pH to pH 5.75. This corresponds roughly to the bicarbonate alkalinity

IA = Intermediate Alkalinity - Determined by titrating from pH 5.75 to pH 4.3. This approximates the Volatile Acid alkalinity.

FIGURE 5.8: Anaerobic Digester - IA/PA Ratio

In Figure 5.8 the IA/PA ratio for the 1989/90 period (Refer data, Appendix 3C) is plotted against time. The sudden onset of instability is clearly shown where the IA/PA ratio exceeded 1,5 at that point. This control test is easier to perform than the VA determination and appears to give a very good indication of the buffering capacity of the digester sludge.

The physical appearance of the sludge clearly reflected the degree of solids breakdown in the digester. The digested sludge had the appearance of watery black ink, with only very small solid particles suspended in it. The solids settled out relatively slowly once the evolution of gas had ceased. The supernatant had a murky, colloidal appearance, but no further settlement occurred.

The results obtained, particularly over the period 1989/90, indicate that although the anaerobic process operates very stably and is immune to fluctuations in load and temperature, it reaches a point somewhere between 12 and 15 days retention when instability sets in. Typical signs of overloading appear, i.e. reduced gas production, increased VA levels, reduction in bicarbonate alkalinity, increase in the IA/PA ratio and a significant reduction in COD breakdown. It was unfortunate that further tests to determine the lower limit of the retention period could not have

been completed. If the heat exchanger could have been used it would have been relatively easy to establish the limits of the anaerobic digestion process as applied in the dual digestion process. The further tests during 1990 were largely confined to restarting the anaerobic process after the digester had been emptied and repaired. The aerobic reactor collapsed soon after the digester had been acclimatized, so no further tests were possible.

5.4 FINAL SLUDGE DEWATERING

After attempting to dry the dual digestion final sludge on the standard drying beds at Milnerton and achieving reasonable success during warm weather (Chapter 4, Section 4.2.4.7), the decision was made to carry out an extensive investigation into the dewatering and drying characteristics of the sludge. It had been planned to subject the treated sludge to standard laboratory sludge characterization tests once the plant was fully operational with the heat exchanger installed and the anaerobic digester fully acclimatized. These tests would have been undertaken during 1990. This goal could not be attained owing to difficulties in restarting the anaerobic digester, damage to the heat exchanger during commissioning and finally, the structural failure of the aerobic reactor vessel.

5.4.1 Laboratory Studies - Sludge Characterization

After the problems with the heat exchanger and restarting the anaerobic digester, a decision was made to start with the tests even though the process was not operating at its proper potential.

The anaerobic digester had undergone a protracted start-up phase of 153 days before it appeared to be reasonably well acclimatized and was put on full load (36 m^3/d). Two weeks later the first of a series of three sets of tests were performed, on final sludge samples taken as follows:

Run 1 5 November 1990
Run 2 15 November 1990
Run 3 27 November 1990

The aerobic reactor had failed on 23 November 1990, so the last samples were taken after digester feeding had ceased and its temperature was declining. The full testing procedure and results obtained on these samples is described by Smollen and Kafaar, 1991. A brief outline of their report follows.

5.4.1.1 Sludge conditioning and characterization tests

To determine the effect of conditioning agents on the sludge, jar tests were carried out and it was established that, on the basis of visual floc formation, a dose of 5 kg/ton DS (dry solids) Zetag 32 polyelectrolyte was the most appropriate conditioning treatment for the sludge. Another non-conventional treatment was also tried, i.e. conditioning the sludge by the addition of 400 kg/ton DS char (pyrolyzed domestic refuse) plus 1 kg/ton DS polyelectrolyte.

The sludge samples, with and without conditioning additives as described, were subjected to the following tests:

* dewaterability tests;
* drying tests;
* electro-osmotic dewatering tests.

The dewaterability tests comprised:

(1) The commonly used specific resistance to filtration (SRF);

(2) The filterability parameter test. This is obtained by dividing the total solids concentration of the sample by the capillary suction time, i.e. c/CST. Good filtration is indicated by a value of 2 or higher;

(3) Cake solids concentration after filtration, centrifugation or gravity dewatering;

(4) Settling properties, using the dilute sludge volume index (DSVI).

The drying tests were performed at low temperature (35°C) in a specialized computer controlled apparatus which closely regulated the drying conditions and continuously recorded the rate of drying.

The electro-osmotic tests were carried out in a special filtering cell equipped with electrodes to subject the sludge to an electric field to facilitate the movement of water in a downwards direction and the sludge particles in an upward direction.

5.4.1.2 Results and conclusions of characterization tests

The results of the sludge characterization tests are given in Tables 5.9 to 5.12. These results show that the dual digestion sludge was very difficult to dewater using any filtration based technology. It was found that the very high content of fine particles in the sludge caused blinding of the filter medium and blocked the drainage passages through the sludge cake that forms on the filter medium.

Polyelectrolyte addition agglomerated the fine particles to larger particles which increased the filtration rate. This was demonstrated by the improvement in the SRF and CST tests. The poly addition did not effect a significant increase in cake solids content after a pressure filtration test. In the drying test it was found that poly addition decreased the immobilized moisture and increased the bound moisture which cannot be removed by mechanical means. In the gravity settling test (24 hour) the poly addition had a beneficial effect, increasing the ratio of final to initial solids from 1,6 (without poly) to a value of 3,4.

The addition of both polyelectrolyte and char produced the highest centrifuged cake solids content of 19%, with a ratio of final to initial solids of 6,7.

The application of electro-osmotic dewatering techniques did not prove to be effective with the dual digestion sludge.

It was concluded that the most effective dewatering technique for the dual digestion sludge was by means of centrifugation, after addition of char and a small quantity of polyelectrolyte.

The overall laboratory evaluation of the dewatering properties of the dual digestion sludge indicated that its dewatering characteristics would be a major limitation in practice. It was, however, that the results of laboratory tests should only be considered to be a trend indicator as there are considerable differences between the laboratory tests and full scale sludge dewatering practices.

TABLE 5.9: Results of Sludge Characterisation Tests (No addition of polyelectrolyte)

Dewaterability Parameters		Runs 1	2	3
Total solids (%)		2,0	2,2	2,0
Vol. Solids (%)		1,4	-	-
Cake solids after filtration (%)		2,7	2,6	2,3
Cake solids after centrifugation (%)		13,1	14,1	13,0
Cake solids after 24 hrs gravity settling (%)		3,3	3,3	3,7
Electro-osmotic dewatering (%)		5,1	5,4	3,9
CST(s)		405	365	425
c/CST		0,05	0,06	0,05
SRF (x 10 ^{12}m/kg)			unable to determine	
DSVI (mL/g)			unable to determine	
Moisture content (g/gDs)	[Total]	7,1	-	7,0
	[Immobilised]	3,8	-	3,7
	[Bound]	2,1	-	1,8
	[Chemical]	1,7	-	1,5
Drying time (h)	[Critical]	1,6	-	2,2
	[Total]	3,5	-	5,0

TABLE 5.10: Results of Sludge Characterisation Tests (With addition of polyelectrolyte)

Dewaterability Parameters		Runs 1	2	3
Total solids (%)		2,9	2,9	2,6
Vol. Solids (%)		-	-	-
Cake solids after filtration (%)		5,9	7,0	5,8
Cake solids after centrifugation (%)		12,8	13,2	12,1
Cake solids after 24 hrs gravity settling (%)		8,4	9,6	7,5
Electro-osmotic dewatering (%)		5,2	5,3	6,2
CST(*)		29	28	47
c/CST		0,98	1,04	0,55
SRF (x 10 ^{12}m/kg)		127,1	192,8	920,0
DSVI (mL/g)		18	20	29
Moisture content (g/gDs)	[Total]	6,8	-	7,5
	[Immobilised]	2,9	-	2,8
	[Bound]	3,3	-	2,8
	[Chemical]	0,6	-	1,9
Drying time (h)	[Critical]	1,0	-	1,9
	[Total]	3,4	-	5,0

TABLE 5.11: Results of Sludge Characterisation Tests
(With addition of char and polyelectrolyte)

Dewaterability Parameters		Runs 1	2	3
Total solids (%)		3,0	2,9	2,6
Vol. Solids (%)		-	-	-
Cake solids after filtration (%)		3,1	2,9	3,3
Cake solids after centrifugation (%)		19,0	18,8	19,2
Cake solids after 24 hrs gravity settling (%)		4,9	4,7	4,2
Electro-osmotic dewatering (%)		12,9	12,0	10,6
CST(s)		189	169	200
c/CST		0,16	0,17	0,13
SRF (x 10 ^{12}m/kg)		582	432	1140
DSVI (mL/g)		22	22	25
Moisture content (g/gDs)	[Total]	6,0	-	4,2
	[Immobilised]	2,3	-	2,0
	[Bound]	2,7	-	1,6
	[Chemical]	1,0	-	0,6
Drying time (h)	[Critical]	1,3	-	1,8
	[Total]	4,5	-	4,5

TABLE 5.12: Comparison of the Results from Laboratory Centrifugation Tests

No Additives				Additives					
				Polyelectrolyte			Char and Polyelectrolyte		
	Solids concentration %		Final initial ratio	Solids concentration %		Final initial ratio	Solids concentration %		Final initial ratio
	Initial	Final		Initial	Final		Initial	Final	
Run 1	2.0	13.1	6.5	2.9	12.8	4.4	3.0	19.0	6.3
Run 2	2.2	14.1	6.4	2.9	13.2	4.5	2.9	18.8	6.5
Run 3	2.0	13.0	6.5	2.6	12.1	4.6	2.6	19.2	7.4
4 hours gravity settling									
Run 1	2.0	3.3	1.6	2.9	8.4	2.9	3.0	4.9	1.6
Run 2	2.2	3.3	1.5	2.9	9.6	3.3	2.9	4.6	1.6
Run 3	2.0	3.7	1.8	2.6	7.5	3.8	2.6	4.2	1.6

5.5 DISINFECTION

The results of microbiological analyses performed on samples from the dual digestion process are presented in Appendices 4A to 4C.

The results obtained during the initial development period are presented in Appendix 4A. At that stage only the aerobic reactor was operating. The samples were examined for total coliforms, faecal coliforms, *Salmonellae*, Coliphage and for *Ascaris lumbricoides* ova. The aerobic process demonstrated that it was capable of achieving a 7 log reduction in total coliforms as well as in faecal coliforms. Coliphage was reduced only marginally by about 2 log on average. No *Salmonellae*

were found in either the feed sludge from the thickener or in the aerobically treated sludge. It was thought that this was due to the low pH of the raw sludge caused by the septic conditions in the long sewers that serve the Milnerton works, as well as the considerable retention time in the sludge sump and in the dual digestion thickener.

Only on one occasion did viable *Ascaris* ova occur in the aerobically treated sludge. In that particular case the reactor was operating at a fairly low temperature (58°C maximum) and the cycle time was two hours i.e. the reactor was operating at one day retention. It was, however, thought to be more likely that the positive count was a consequence of poor sampling technique rather than inadequate disinfection by the aerobic process.

After examining the results achieved during this initial period, it was decided to limit the microbiological analyses to faecal coli and *Ascaris* ova as it was too expensive to continue with the full microbiological test programme. During 1988, 21 samples of thickened feed sludge and aerobically treated sludge were examined as well as 19 samples of the final sludge from the anaerobic digester. These results are presented in Appendices 4B and 4C. Also included are the tests on a sample taken during 1990, as no microbiological tests were performed during 1989 and the process was not fully operational during most of 1990.

The results (Appendices 4B and 4C) indicate that on average, there was a 9 log reduction in faecal coli, as well as a number of zero counts for the aerobic sludge. Viable *Ascaris* ova were effectively zero after the sludge had passed through the reactor. One positive count occurred, but it was again suspected that the sampling technique was to blame as it was very difficult to collect a sample from the fairly large sampling ports without any risk of outside contamination. It was unlikely that *Ascaris* ova could have survived a sludge temperature of 58 to 61°C in the reactor over the 2,5 hour time cycle. Note that this was the absolute minimum time that any sludge particle or organism would spend in the reactor while it was operating at 1,25 day retention. According to the literature, *Ascaris* ova are completely inactivated after being subjected to 55°C for 2 hours (Carrington, 1984).

The anaerobic digester (Appendix 4C) appears to effect little or no further reduction in the faecal coliform counts. On a number of occasions actual regrowth was indicated. Three positive viable *Ascaris* counts occurred in the digested sludge and quite often the non-viable counts were found to be higher on the digested sludge than in the aerobic feed sludge. Why this occurred is not clear, as at no time was any sludge other than the aerobically processed sludge added to the digester. The sampling procedure could have been to blame as it is very strange that positive viable ova counts occurred in all the samples taken on 5 July 1988. Other unknown factors may have caused this anomaly, but it is most unlikely that there were viable *Ascaris* ova present in the digester and that they only appeared occasionally in the samples.

An intensive microbiological evaluation programme over a one month period while the plant was operating with its heat exchanger and while optimum conditions

prevailed had been planned for 1990. Unfortunately this programme could not be started owing to the failure of the heat exchanger and finally the total collapse of the aerobic reactor. In view of the lack of further microbiological tests, the results presented indicate an acceptable degree of disinfection. If the information cited in the literature with regard to the destruction of *Ascaris* ova is correct, it is impossible for any viable *Ascaris* ova to survive the conditions in the aerobic reactor while it is operating at a maximum temperature of 60°C. It is also just as unlikely that viable ova will be present in the final treated sludge from the anaerobic digester if they were not introduced there in the first instance.

5.6 OVERALL PERFORMANCE OF THE DUAL DIGESTION PROCESS

Apart from being easy to operate and control, the dual digestion process as applied at Milnerton performed well, both in adequately stabilizing the sludge and in providing a high degree of disinfection.

The overall performance of the process with regard to stabilization is presented in Table 5.13, which shows the changes effected through the plant (1989/1990). The data shows that very little stabilization occurred in the aerobic reactor, with the total solids being reduced by 13,6%, the volatile solids by 14,9% and the COD by 18,1%. Very substantial increases in ammonia N and alkalinity occurred, and the pH of the septic feed sludge was increased to a value of 7,1, which was probably ideal for the next anaerobic stage.

The claim that the aerobic process "conditions" the sludge for further stabilization in the anaerobic stage is therefore correct.

The aerobic reactor appeared to solubilize particulate material present in the sludge as substantial amounts of colloidal material were observed in the effluent sludge from the reactor. It was almost impossible to filter the aerobic sludge as these colloids caused severe blinding of the filter medium and was also resistant to centrifugation, making the determination of suspended solids impossible.

Whether the aerobic treatment had made the sludge more amenable to anaerobic digestion could not be determined from the results of the evaluation. A control process would have to be operated in parallel to determine this factor and would best be done on a laboratory scale.

The anaerobic digestion process effected a very large removal of solids and COD. On average 43% total solids, 49% volatile solids and 44% COD was removed from the aerobic sludge by the anaerobic digester.

Overall the dual digestion process reduced the total solids by 50,8%, the volatile solids by 56,6% and the COD by 54,4%. In effect this means that more than half of the solid material is removed and the final disposal problem is significantly reduced. The problem of dewatering the sludge produced is difficult, but it did dry on the sludge beds once the supernatant had been drained off. On the sludge beds it did not attract flies or produce unacceptable odours. The sludge odour was

characteristic of a well stabilized anaerobically digested sludge. However, the laboratory characterization tests showed that dewatering the sludge using any means of filtration was extremely difficult and that centrifugation produced better results.

The dual digestion process as operated at Milnerton produces as much biogas as a normally operated anaerobic digester, and all of it is available for heating or other purposes. The aerobic process generated enough heat to maintain adequate mesophilic temperatures in the anaerobic digester during winter, and too much for this purpose during summer. A heat exchanger is considered to be an essential part of

a dual digestion installation, as it is needed for controlling the temperature of the digester and would improve the operation and reduce the running cost of the whole process.

It must, however, be borne in mind that the retention period in the anaerobic digester was on average 21 days. Considerable reductions in the gas production and in the solids breakdown will occur if the retention time is reduced to the more realistic 15 days, which proved to still give stable operation. It seems most unlikely that the claims of 8 to 10 days retention in the anaerobic digester could be achieved without instability and total digester failure. The Milnerton digester, in spite of being operated at the optimum temperature and being provided with good mixing, showed definite signs of instability at 12 days retention.

Operation using air instead of, or together with pure oxygen was not tried, but the design criteria developed which were based on the thorough evaluation of the aerobic reactor (Part 2) cover this aspect.

The overall degree of disinfection achieved by the dual digestion process was good. *Ascaris lumbricoides* ova were inactivated in all but a few instances by the pasteurizing action of the aerobic reactor. Considerable reductions of faecal coli occurred in the aerobic reactor, but it appeared as if some regrowth occurred in the anaerobic digester. Since regrowth of coliform organisms have been reported in the literature on, for example, sludge drying beds, this is likely to occur in any case, but as an indicator organism it does show that the aerobic stage exerts a considerable pasteurizing effect on the sludge. If required, it would be entirely feasible to operate the aerobic reactor at a higher maximum temperature, for example 70°C to obtain full pasteurization of the sludge.

The draw and fill mode of operating the aerobic reactor completely eliminates the possibility of any pathogenic organism short-circuiting the aerobic reactor and entering the anaerobic digester. If the aerobic reactor is operated correctly, which is very easy to do, the dual digestion process must be regarded as an excellent process for disinfecting sewage sludge.

TABLE 5.13: Average Values of Analyses and Changes through Dual Digestion Plant

Parameter		Average Value of Analyses			% Removal(-) or Increase (+)		
		Feed	Aerobic	Anaerobic	Aerobic	Anaerobic	Dual Digest
Total solids	g/L	37,4	32,3	18,4	-13,6	-43,0	-50,8
Volatile solids	g/L	30,2	25,7	13,1	-14,9	-49,0	- 56,6
Volatile solids	%	81,0	79,8	70,9	-	-	-
COD	g/L	49,1	40,2	22,5	-18,1	-44,0	-54,2
Ammonia-N	mg/L	175	318	651	+81,7	+104,7	+272
Alkalinity as $CaCO_3$	mg/L						
To pH 5,75		15	637	1733	+(4147)	+(172)	+(11453)
To pH 4,3		679	950	2260	+39,9	+137,9	+233
pH		5,6	7,1	7,6	-	-	-
Volatile Acids	mg/L	-	-	298	-	-	-

CHAPTER 6

CONCLUSIONS AND RECOMMENDATIONS

6.1 COMPLIANCE WITH PROCESS CLAIMS

The conclusions to be drawn from this evaluation of the dual digestion process must, first of all, be considered in the light of the claims made for the process as described in Chapter 1.

(1) Claim: Disinfection and stabilization of the sludge occurs in a single process.

This has been adequately established at Milnerton.

(2) Claim: The use of energy for the supply and dissolution of oxygen is low since the degree of oxidation of the sludge is limited, the only requirement being the maintenance of the chosen temperature to effect disinfection. Stabilization is completed in the anaerobic digester, which does not require oxygen.

This claim is essentially correct, but it was found that even if an oversized pump is used in the Vitox system, the excess energy is mostly converted to heat, increasing the reactor sludge temperature. Less oxygen is therefore required and as the cost of electrical energy and that of the equivalent amount of oxygen to generate a given amount of heat does not differ significantly, nothing is lost by using a larger recirculation pump for mixing and oxygen dissolution.

(3) Claim: No external heating of the anaerobic digester is required as all necessary heat is provided by the hot sludge fed to the digester from the aerobic reactor.

This claim is valid. It does not, however, mention that during hot summer conditions far more heat than required is generated, resulting in overheating of the digester. If a well insulated digester were used, this would be a permanent problem. A means of cooling the aerobically treated sludge before feeding it to the anaerobic digester is essential. A batch heat exchanger which would cool the effluent sludge from, and heat the influent sludge to the aerobic reactor would solve this problem and reduce the oxygen consumption as well.

(4) Claim: The process is an overall net producer of energy, since all the biogas, which contains approximately 65% methane, is available for use external to the dual digestion plant.

Depending on the approach taken, the aerobic reactor could be considered to be a net producer of energy. The maximum biological power generated by the aerobic stage at Milnerton was 73kW as low grade heat. The oxidation reaction consumed 20 kg/h pure oxygen from the 24 kg/h supplied. The electrical power to dissolve

the oxygen at 83% efficiency was 19 kW, but 17 kW was recovered as heat by the sludge. Thus a total of 90 kW of power as heat was obtained from 19 kW electrical power. But the oxygen also represents a power input. To produce liquid oxygen requires 1100 kWh per ton, therefore each kg pure oxygen supplied is equivalent to 1,1 kWh electrical energy. The 24 kg/h therefore represents an additional 26 kW power input which is not recoverable.

In total, therefore, 45 kW electrical power (= 19 + 26) was converted into 90 kW heat, which gives an effective gearing of 2,0.

If the oxygen had been produced on site as gaseous oxygen of 95% purity using a pressure swing absorption (PSA) plant of 10 tons/d capacity, this would consume 300 kWh energy per ton. Therefore 1 kg O_2 would be equivalent to 0,3 kWh electrical energy. The oxygen supplied (24 kg/h) thus represents 7 kW power which is not recoverable. The total power supplied is thus 26 kW (= 19 + 7), for an output of 90 kW.

In real (cost associated) terms, the aerobic reactor can therefore be considered a net energy producer (For details, refer to Part 2, Appendix 5B.1.8).

The anaerobic digester produces about 350 m^3/d of biogas. This is approximately equivalent to 8 330 MJ/d, or 96 kW of power which is available for heating or for fuelling a gas engine which may, for example, be used to generate electrical power.

The dual digestion process can therefore be considered to be a net producer of energy.

(5) Claim: The process may be started up much more quickly than an anaerobic plant, eight to ten days being required instead of thirty to forty days.

When starting the process from scratch, this statement is not even remotely true. The aerobic reactor can be started in approximately four days. To start the anaerobic digester without seeding it from another pathogen free anaerobic digester may take a month or more. At minimum it will require as much time as starting up a normal anaerobic digester.

(6) Claim: Although the system requires two separate digestion tanks, the overall capital cost is reduced compared with conventional anaerobic digestion. This is a consequence of the reduced tank volumes required (60% less volume is needed for the anaerobic stage) and the fact that no heating equipment is necessary.

It is not possible to comment on this statement based on the Milnerton experience. There it was found that 15 days was about the lowest retention time that could be employed without endangering the stability of the process. This was approximately half the retention required with relatively inefficient older type mesophilic digesters, but not very much better than is achieved with modern high

rate digesters which reputedly give effective stabilization in 15 to 20 days. A new dual digestion process may therefore not necessarily be less costly than a new high rate anaerobic digester, but the latter does not provide disinfection of the sludge. The capital cost of an anaerobic digester plus a pasteurizer will in all likelihood be more than a dual digestion plant.

(7) Claim: The stability of the anaerobic stage is considerably enhanced by the conditioning effect of the aerobic stage by:

* Reducing the effects of toxic organic contaminants in the feed sludge.

This was not evaluated at Milnerton.

* Oxidizing the rapidly degradable components of the sludge which helps to maintain the balance of growth rates of acid formers and methane producers in the digester.

It was not attempted to prove this theory. It may have been a factor in the good stability achieved.

* Solubilizing part of the solid organic material at the thermophilic temperatures reached in the reactor.

This was not specifically investigated. The aerobically treated sludge did contain a significant quantity of extremely fine particles which would not settle on standing. This colloidal nature made the aerobic sludge virtually unfilterable. The larger particles present in the feed sludge had definitely been broken down, but to what degree solubilization had occurred was not established.

* Ammonium ions are produced by the deamination of organic nitrogen compounds, increasing the alkalinity.

Ammonia N increased on average by 80%, while the bicarbonate alkalinity (as approximately determined by titration to pH 5,75) increased by over 4000%. The pH on average increased from 5,6 to 7,1. This claim is therefore correct.

* Temperature fluctuations are minimized by controlling the heat input from the aerobic stage.

This was not possible at Milnerton. The temperature of the digester can be controlled as stated, but not without either changing the reactor throughput or by changing its temperature or by bypassing some of the aerobic sludge to storage.

None of these provide a satisfactory means of controlling the digester temperature as either the capacity or the disinfection of the system is affected.

(8) Claim: The dual digestion process is particularly suited to cases where an existing anaerobic plant has to be upgraded, by installing an aerobic reactor ahead of the existing digester. The capacity of the plant may be increased by a factor of 2 or 3.

This claim is essentially correct. A number of anaerobic plants have been upgraded by addition of an aerobic reactor in Europe and America. Information on the improved treatment capacity achieved has not been available. Judging by the performance of the Milnerton plant (which was also an "upgrade"), where the capacity of the anaerobic digester was in effect, doubled, similar performances were probably achieved elsewhere. But as a 15 day retention was found to be necessary in the anaerobic digester, a plants' capacity would be trebled only if the original anaerobic plant had operated at a retention of 45 days or more.

6.2 AEROBIC REACTOR CONCLUSIONS

The comprehensive investigation into the operation of the aerobic reactor of the dual digestion process has yielded accurate information on the basics of the process of biological heat generation as well as on the control of the reactor temperature. It was demonstrated that under oxygen limiting conditions the biological heat generation is directly proportional to the oxygen consumption, the constant of proportionality being the specific heat yield Y_H, which on average was found to be 13,1 MJ/kgO. The maximum heat generation rate occurs when the sludge consumes oxygen at its maximum rate, i.e. at its biological oxygen utilization rate [OUR, kgO/(m^3.h)]. At this point, the sludge is no longer oxygen limited and any increase in the oxygen supply rate will not increase the oxygen consumption of the sludge or the biological heat generation rate. The Milnerton sludge, which is a blend of primary and humus tank sludge, had a steady state OUR of 0,38 kgO/(m^3.h) at a concentration of 30 kg VS/m^3.

Under oxygen limiting conditions it was shown that the temperature could be completely and instantaneously controlled by the oxygen supply rate (OSR). Provided that the oxygen transfer rate was less than the oxygen utilization rate of the sludge, step increases as high as 330% in the oxygen supply rate which effectively increased the transfer rate and the biological consumption rate by 300%, caused an immediate increase in the biological heating rate which increased the reactor temperature. Decreasing the oxygen supply rate in the same way lowered the reactor sludge temperature.

The maximum oxygen transfer rate of the oxygenation system is very important in the design and operation of the aerobic reactor. To reach adequate disinfection temperatures at the design retention time, the oxygenation system must be capable of transferring sufficient oxygen to the sludge. The maximum biological heating rate is controlled by the maximum oxygen transfer rate, provided that the latter is less than the oxygen utilization rate of the sludge. In the design of a system, the oxygen transfer rate should be slightly greater than the oxygen utilization rate of the sludge. If so, the biological kinetics, i.e. the oxygen utilization rate, will govern the minimum retention time in the reactor rather than the maximum oxygen transfer rate of the oxygenation system.

The respiration quotient (Y_{CO2}), i.e. the mol CO_2 generated per mol O_2 utilized, was found to be 0,66 instead of the usual 1.0 assumed from stoichiometric considerations. This appears to indicate that the heat generating oxidation reactions are

probably not simply those of volatile solids degradation. The value of Y_{CO2} enables an estimate to be made of the volumetric flow rate of the reactor vent gas and its associated heat loss. It is not of great significance, as with pure oxygen systems the rate of heat loss in the vent gas is negligible. In air oxygenated systems the large volume of nitrogen gas passing through the reactor masks the effect of the losses associated with the CO_2 that was generated.

A significant heat loss occurs in the water vapour in the vent gas. It was shown that the vent gas was saturated at all vent gas flow rates.

The COD and/or VS removal rates were demonstrated to be poor parameters for quantifying the biological heat generation rate or for controlling the reactor sludge temperature. The COD and VS tests are prone to sampling and analysis errors and to obtain an accurate difference between the influent and the effluent values is virtually impossible.

Aerobic reactor design and simulation procedures were derived from the results of the Milnerton reactor evaluation (Detailed in Part 2). These were founded on basic heat and mass balance principles and are suitable for general application to reactors oxygenated by oxygen, air or by oxygen enriched air.

The Milnerton aerobic reactor, oxygenated with pure oxygen, could be operated at disinfection temperatures of 60°C or higher at a 1 day retention time during hot weather when the feed sludge (26°C) and ambient (30°C) temperatures were high. During winter, when the feed sludge (17°C) and ambient (10°C) temperatures were low, a retention time of 1,25 days was required to maintain a reactor temperature of 60°C.

In operation, the aerobic reactor was stable, reliable and easy to operate. For 8 hours per day (24h00 to 08h00) it was left unattended by the operating staff. During the rest of the day they had to see that the thickened feed sludge was available and take the composite samples required during the evaluation period. Equipment failure rarely occurred, but the Vitox recirculation pump required periodic attention due to abrasive wear of the casing. The pump had to be rebuilt at about 5 to 6 monthly intervals.

6.3 ANAEROBIC DIGESTER CONCLUSIONS

The anaerobic digester of the dual digestion process behaves like any high rate anaerobic digester, but it appears that it is more stable and less prone to failure when temporarily overstressed.

Problem areas experienced at Milnerton was the difficulty in starting the digestion process and the control of the digester temperature and retention period.

The digester was started twice from a completely empty state. The first time it took 96 days to complete the starting phase to the point where conditions were normal. The second time it took more than five months to reach a stable operating

state. Seeding is clearly required, but this must be done by introducing sludge from another pathogen free digester if the disinfection capability of the process is not to be compromised.

At Milnerton the digester temperature and retention period were interdependent because it was not possible to control the temperature of the aerobic feed sludge to the digester. The aerobic sludge provided ample heat to maintain optimum mesophilic temperature in the digester during winter, but in summer the heat input was far too high if the total flow were accommodated and overheating occurred. Reducing the heat input was only possible by reducing the feed volume from the reactor, but this increased the retention time. A simple batch heat exchanger to cool the sludge transferred to the anaerobic digester and warm the feed sludge to the aerobic reactor is an essential requirement of the dual digestion process.

The production of biogas by the anaerobic digester was found to be normal in all respects. At the average retention of 21 days with a VS removal of 12,6 kg VS/m^3, from 13 to 15 m^3 of biogas was produced (at 25°C). The specific gas production of 0,94 to 1,09 m^3/kg VS removed was as good as that of "stand alone" heated and mixed mesophilic anaerobic digesters, which produce between 0,75 and 1,1 m^3/kg VS removed. The biogas comprised 67% methane and 33% carbon dioxide. At an average daily gas production of 350 m^3 the digester produces about 8300 MJ/d energy which is available for other purposes.

Both VS and COD removal was high at 49% and 44% respectively. Both parameters decrease as the retention time reduced. At 15 days retention, which was the lowest at which adequate stability could be maintained, VS and COD removals would be down to 40 and 35% respectively. The gas production would also be lower.

The anaerobic process stability was generally excellent. Regular tests indicated that a high pH, high bicarbonate alkalinity and low volatile acid levels were maintained at retention times down to 15 days. At 12 days retention all the signs of instability were present,so it was concluded that 15 days was the minimum safe operating retention time for this particular digester.

The treated sludge produced by the digester contained only very fine particulate and colloidal matter. It proved to be difficult to dewater and sludge characterization tests showed that any filtration based dewatering process would give poor results, but that with limited preconditioning with polyelectrolytes, centrifugation would provide a reasonable degree of dewatering and thickening.

6.4 OVERALL DUAL DIGESTION PROCESS CONCLUSIONS

The dual digestion process was shown to give effective disinfection by subjecting the sludge to the chosen disinfection temperatures. The batch draw and fill mode of operation of the aerobic reactor ensures that no short circuiting occurs and that each batch of sludge is subjected to the predetermined temperature for at least the batch cycle time (not less than 2 hours).

Sludge stabilization at typically 1,25 days retention in the aerobic reactor and 21 days in the anaerobic digester was excellent, with VS and COD reductions of 56% and 54% being obtained respectively. At 15 days retention the overall reductions would be lower but would still be from 40 to 50%.

Gas production by the anaerobic digester was normal and was not reduced by the aerobic pretreatment which reduced the available substrate slightly by about 14% (as VS). All the biogas was available for other uses as none was required for heating in the dual digestion process.

The heat generated by the aerobic process was sufficient for maintaining the anaerobic digester at optimum mesophilic temperature levels during the colder winter months, but proved to be excessive during summer. Some of this heat should be beneficially exploited to heat the feed sludge, thereby reducing the oxygen requirements and enabling the anaerobic digester temperature to be controlled.

The comprehensive evaluation of the aerobic process has produced design data which may be confidently applied. The response of the anaerobic digester to operation on the aerobically pretreated sludge was not fully evaluated owing to the problem of controlling the temperature without affecting the retention time. Stable operation at 15 days retention and incipient digestion failure at 12 days would suggest that 15 days is the lower safe limit for design purposes.

Dewatering the final sludge could be difficult. On the sludge beds it dries at about the same rate as well stabilized anaerobically digested sludge, provided that the supernatant is first decanted. Most dual digestion plants elsewhere employ thickening and storage tanks following the anaerobic digester stage. The treated sludge is allowed to settle in these tanks and supernatant is withdrawn so that a fair degree of thickening occurs. The sludge is then used in liquid form for agricultural purposes. These methods should be considered if the dual digestion process is to be applied in South Africa.

The process is simple to operate. At Milnerton it was routinely operated by relatively unskilled operators who had received a minimum of instruction in running the process.

A dual digestion plant may be designed as a "high tech", high capital investment operation, with a high degree of insulation, heat exchange, employing supplementary heating by utilizing part of the digester gas, computer controlled, requiring a minimum of operator attention and having low operating costs. Alternatively, it could be designed as a "low tech", low capital cost process, with manual control requiring more operator attention, but still incorporating some form of heat exchange or external cooling to control the anaerobic digester temperature. In this form it would have higher operating costs.

Both systems would work well and produce a disinfected, well stabilized final sludge. The choice depends on the specific requirements and location of the plant.

CHAPTER 7

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APPENDIX 1

TABLE 1 : CLASSIFICATION OF SEWAGE SLUDGE TO BE USED OR DISPOSED OF ON LAND

TYPE OF SEWAGE SLUDGE	ORIGIN/TREATMENT (EXAMPLES)	CHARACTERISTICS/QUALITY STANDARD
TYPE A SLUDGE	RAW SLUDGE COLD DIGESTED SLUDGE SEPTIC TANK SLUDGE OXIDATION POND SLUDGE (NIGHT SOIL)	* USUALLY UNSTABILISED AND CAN CAUSE ODOUR NUISANCES AND FLY-BREEDING * CONTAIN PATHOGENIC ORGANISMS * VARIABLE METAL CONTENT
TYPE B SLUDGE	ANAEROBIC DIGESTED SLUDGE (HEATED DIGESTER) SURPLUS ACTIVATED SLUDGE HUMUS TANK SLUDGE	* FULLY OR PARTIALLY STABILISED - SHOULD NOT CAUSE SIGNIFICANT ODOUR NUISANCES OR FLY-BREEDING * CONTAIN PATHOGENIC ORGANISMS * VARIABLE METAL CONTENT
TYPE C SLUDGE	PASTURISED SLUDGE HEAT TREATED SLUDGE LIME STABILISED SLUDGE COMPOSTED SLUDGE IRRADIATED SLUDGE FUMIGATED SLUDGE	* CERTIFIED TO COMPLY WITH THE FOLLOWING QUALITY REQUIREMENT: / STABILISED - SHOULD NOT CAUSE ODOUR NUISANCES OR FLY-BREEDING / CONTAIN NO VIABLE ASCARIS OVA PER 10g DRY SLUDGE / MAXIMUM 1 SALMONELLA ORGANISM PER 100g DRY SLUDGE / MAXIMUM 10 FAECAL COLIFORMS PER 100g DRY SLUDGE IMMEDIATELY AFTER TREATMENT (DISINFECTION/STERILISATION) NB IF NOT CERTIFIED THIS SLUDGE IS CONSIDERED A TYPE B SLUDGE * VARIABLE METAL CONTENT
TYPE D SLUDGE PRODUCED FOR UNLIMITED USE ON LAND WITH OR WITHOUT ADDITION OF PLANT NUTRIENTS OR OTHER MATERIALS	PASTURISED SLUDGE HEAT TREATED SLUDGE LIME STABILISED SLUDGE COMPOSTED SLUDGE IRRADIATED SLUDGE FUMIGATED SLUDGE	* CERTIFIED TO COMPLY WITH THE FOLLOWING QUALITY REQUIREMENT: / STABILISED - SHOULD NOT CAUSE ODOUR NUISANCES OR FLY-BREEDING / CONTAIN NO VIABLE ASCARIS OVA PER 10g DRY SLUDGE / MAXIMUM 1 SALMONELLA ORGANISM PER 100g DRY SLUDGE / MAXIMUM 10 FAECAL COLIFORMS PER 100g DRY SLUDGE IMMEDIATELY AFTER TREATMENT (DISINFECTION/STERILISATION) / MAXIMUM METAL CONTENT IN mg/kg DRY SLUDGE Cadmium 20 Cobalt 100 Chromium 2750 Copper 750 Mercury 10 Molibdenium 25 Nickel 200 Lead 250 Zinc 2750 * USER MUST BE INFORMED ABOUT THE MOISTURE AND N P K CONTENT * USER MUST BE WARNED THAT NOT MORE THAN 8 t/ha/yr (OR kg/10 sq.m) (DRY SLUDGE) MAY BE APPLIED TO SOIL AND THAT THE pH OF THE SOIL SHOULD PREFERABLY BE HIGHER THAN 6

TABLE 2 : CROP TYPES THAT MAY BE FERTILISED WITH SEWAGE SLUDGE AND OTHER METHODS OF UTILISATION OR DISPOSAL

CROP TYPES/OTHER OPTIONS	TYPE A SLUDGE	TYPE B SLUDGE	TYPE C SLUDGE	TYPE D SLUDGE
1 SELLING OR ALIENATION	PERMISSIBLE - RESTRICTION 1	PERMISSIBLE - RESTRICTION 1	PERMISSIBLE - RESTRICTION 1	PERMISSIBLE - NO RESTRICTION
2 HOUSEHOULD VEGETABLES CONSUMED RAW OR COOKED	NOT PERMISSIBLE	NOT PERMISSIBLE	NOT PERMISSIBLE	PERMISSIBLE - RECOMMENDATION 2, 6
3 VINEYARDS AND FRUIT TREES	NOT PERMISSIBLE	PERMISSIBLE - RESTRICTION 3, 6	PERMISSIBLE - RESTRICTION 6	PERMISSIBLE - RECOMMENDATION 6
4 CEREAL CULTURE AND SUGAR-CANE	NOT PERMISSIBLE	PERMISSIBLE - RESTRICTION 3, 6	PERMISSIBLE - RESTRICTION 6	PERMISSIBLE - RECOMMENDATION 6
5 PUBLIC GARDENS AND TRAFFIC ISLANDS ONLY FOR BEAUTIFYING WITH MINIMUM HUMAN CONTACT	NOT PERMISSIBLE	PERMISSIBLE - RESTRICTION 6	PERMISSIBLE - RESTRICTION 6	PERMISSIBLE - RECOMMENDATION 6
6 PUBLIC PARKS, RECREATION AREAS, LAWNS AT SCHOOLS, SWIMMING POOLS, SPORTS FIELDS	NOT PERMISSIBLE	PERMISSIBLE - RESTRICTION 2, 6	PERMISSIBLE - RESTRICTION 2, 6	PERMISSIBLE - RECOMMENDATION 6
7 PRIVATE GARDENS - LAWNS, SHRUBS, TREES, VEGETABLES	NOT PERMISSIBLE	NOT PERMISSIBLE	NOT PERMISSIBLE	PERMISSIBLE - RECOMMENDATION 6
8 NURSERIES - SHRUBS, TREES AND OTHER PLANTS	NOT PERMISSIBLE	PERMISSIBLE - RESTRICTION 3, 5	PERMISSIBLE - RESTRICTION 5	PERMISSIBLE
9 INSTANT LAWN CULTIVATION	NOT PERMISSIBLE	PERMISSIBLE - RESTRICTION 3, 5	PERMISSIBLE - RESTRICTION 3, 5	PERMISSIBLE
10 GRAZING FOR MILK, MEAT AND EGG PRODUCING ANIMALS	NOT PERMISSIBLE	PERMISSIBLE - RESTRICTION 2, 6	PERMISSIBLE - RESTRICTION 6	PERMISSIBLE - RECOMMENDATION 6
11 CROPS NOT FOR GRAZING, BUT UTILISED AS DRY FODDER	NOT PERMISSIBLE	PERMISSIBLE - RESTRICTION 3, 4, 6	PERMISSIBLE - RESTRICTION 6	PERMISSIBLE - RECOMMENDATION 6
12 STABILISING OF MINE DUMPS - GRASS OR OTHER PLANTS	NOT PERMISSIBLE	PERMISSIBLE - RESTRICTION 6, 7	PERMISSIBLE - RESTRICTION 6, 7	PERMISSIBLE - RECOMMENDATION 6, 7
13 COMPOSTING WITH OTHER ORGANIC MATERIAL	PERMISSIBLE (SEE TYPE C SLUDGE)	PERMISSIBLE - (SEE TYPE C SLUDGE)	PERMISSIBLE	PERMISSIBLE
14 NATURAL VELD AND TREE PLANTATIONS	PERMISSIBLE - RESTRICTION 4,5,6,8	PERMISSIBLE - RESTRICTION 3, 4, 6	PERMISSIBLE - RESTRICTION 6	PERMISSIBLE
15 LAND APPLICATION - PLOUGHED IN REPEATEDLY, LANDFILL	PERMISSIBLE - RESTRICTION 4,5,6,8	PERMISSIBLE - RESTRICTION 4,5,6,8	PERMISSIBLE - RESTRICTION 6, 8	PERMISSIBLE
GENERAL REQUIREMENTS AND PRECAUTIONARY MEASURES ACCORDING TO TYPE OF SLUDGE	- MAXIMUM SLOPE OF SITE 1:25 (4%) - DEPTH OF AQUIFER > 5m - > 500m FROM DWELLING - > 200m FROM RIVER, DAM, BOREHOLE - SOIL pH > 6	- MAXIMUM SLOPE OF SITE 1:17 (6%) - DEPTH OF AQUIFER > 2m - > 200m FROM DWELLING - > 200m FROM RIVER, DAM, BOREHOLE - SOIL pH > 6	- SOIL pH > 6	- SOIL pH PREFERABLY > 6

SEE TABLE 3 FOR RESTRICTION/RECOMMENDATION 1 TO 8

TABLE 3 : RESTRICTIONS/RECOMMENDATIONS REFERRED TO IN TABLE 2

1 ONLY AS PER CONTRACT.....SEE TABLE 7.

2 APPLICATION ONLY DURING PLANTING.

3 APPLICATION ONLY WITH PLANTING AND DURING THE PERIOD SUBSEQUENT TO HARVESTING AND PRIOR TO THE NEXT GROWING SEASON IN ORDER TO MINIMISE SEWAGE SLUDGE COMING INTO CONTACT WITH CROPS TO BE HARVESTED.

4 APPLICATION PERMISSIBLE, ONLY IF THE AREA IS EFFECTIVELY FENCED TO KEEP OUT UNAUTHORISED PERSONS AS WELL AS MILK, MEAT AND EGG PRODUCING ANIMALS.

5 NO SUBSEQUENT SELLING OR ALIENATING OF SLUDGE OR ANY MIXTURE CONTAINING SUCH SLUDGE IS ALLOWED BY THE USER.

6 ALL SLUDGE MUST BE MIXED OR COVERED WITH SOIL WHENEVER POSSIBLE.

7 SOIL pH AND SLOPE REQUIREMENTS COULD BE RELAXED ON CONDITION THAT NO CONTAMINATED RUNOFF AND SEEPAGE WATER WILL POLLUTE ANY SURFACE OR UNDERGROUND WATER.

8 APPLICATION OF EXCESSIVE QUANTITIES OF SEWAGE SLUDGE TO LAND CAUSE THAT SITE TO BE UNFIT FOR ANY OTHER PURPOSE DURING SUCH OPERATION AND FOR A MINIMUM PERIOD OF TWO YEARS AFTER TERMINATION THEREOF.

NO NUISANCE OR ANY OTHER CONDITION POSING A POTENTIAL HEALTH HAZARD OR WHICH MAY CAUSE POLLUTION OF ANY WATER SOURCE WILL BE TOLERATED ON SUCH SITE.

UTILISATION OF THIS SITE FOR ANY OTHER PURPOSE WILL ONLY BE PERMITTED AFTER THE NECESSARY INVESTIGATION HAS PROVED THIS TO BE SAFE.

TABLE 4 : MAXIMUM SOIL METAL CONTENT AND PERMISSIBLE METAL APPLICATION

METAL		MAXIMUM PERMISSIBLE METAL CONTENT IN SOIL mg/kg	MAXIMUM AMOUNT OF METAL THAT CAN BE APPLIED TO SOIL kg/ha/25yr	g/ha/yr	MAIN REASON WHY METALS ARE LIMITED PHYTOTOXICITY	ZOOTOXICITY
CADMIUM	Cd	2	4	160		X
COBALT	Co	20	20	800		X
CHROMIUM	Cr	80	550	22000	X	
COPPER	Cu	100	150	6000	X	
MERCURY	Hg	0.5	2	80		X
MOLIBDENIUM	Mo	2.3	5	200		X
NICKEL	Ni	15	40	1600	X	
LEAD	Pb	56	50	2000		X
ZINC	Zn	185	550	22000	X	

TABLE 5 : EXAMPLES OF APPLICATION RATES OF SEWAGE SLUDGE CONTAINING VARIOUS CONCENTRATIONS OF METALS

		MAXIMUM AMOUNT OF DRY SEWAGE SLUDGE IN t/ha/yr (OR kg/10 sq.m/yr) PERMISSIBLE ON SOIL FOR DIFFERENT METAL CONCENTRATIONS IN SLUDGE					
		1	2	5	10	20	50
METAL		METAL CONCENTRATION OF DRY SLUDGE - mg/kg (OR g/t)					
CADMIUM	Cd	160	80	32	16	8	3
COBALT	Co	800	400	160	80	40	16
CHROMIUM	Cr	22000	11000	4400	2200	1100	440
COPPER	Cu	6000	3000	1200	600	300	120
MERCURY	Hg	80	40	16	8	4	2
MOLIBDENIUM	Mo	200	100	40	20	10	4
NICKEL	Ni	1600	800	320	160	80	32
LEAD	Pb	2000	1000	400	200	100	40
ZINC	Zn	22000	11000	4400	2200	1100	440

TABLE 6 : GENERAL INDICATION OF NITROGEN DEMAND

CROP OR USE	NITROGEN DEMAND kg N/ha/yr
HORTICULTURE AND AGRICULTURE	50 - 200
SHRUBS AND TREES	50 - 100
CEREAL CROP	75 - 175
LUCERNE	300 - 500

TABLE 7 : ESSENTIAL ITEMS TO BE INCLUDED IN CONTRACTUAL AGREEMENT (Type A,B or C Sludge)

SUPPLIER/PRODUCER

1 NAME AND ADDRESS
2 TYPE OF SEWAGE SLUDGE - A, B OR C
3 QUALITY: HYGIENIC - STABILITY AND MICRO-ORGANISMS
MOISTURE CONTENT
NITROGEN, PHOSPHATE AND POTASSIUM CONTENT
MAXIMUM METAL CONTENT (Cd, Co, Cr, Cu, Hg, Mo, Ni, Pb, Zn)
4 LIMITING METAL AND MAXIMUM APPLICATION RATE IN DRY t/ha/yr
5 RECOMMENDED MAXIMUM APPLICATION RATE IN t/ha/yr IN TERMS OF NITROGEN DEMAND OF CROP
6 NOTIFICATION OF LOCAL AUTHORITIES INVOLVED

RECEIVER/USER

1 NAME AND ADDRESS
2 NAME OF TRANSPORTER OF SLUDGE
3 NAME OF FARM OR SITE WHERE SLUDGE WILL BE STORED AND USED
4 SIZE OF SLUDGE APPLICATION AREA
5 CROPS TO BE FERTILISED OR ALTERNATIVE USE
6 PREVIOUS SEWAGE SLUDGE APPLICATION - ANNUAL RATE AND FREQUENCY
7 METAL CONTENT OF SOIL (Cd, Co, Cr, Cu, Hg, Mo, Ni, Pb, Zn)-
SOIL TO BE ANALYSED AFTER 10 YEARS OF SLUDGE APPLICATION
8 DETAILS OF SEWAGE SLUDGE PROCESSING, ADDITION OF OTHER MATERIALS
OR CHEMICALS AND QUALITY OF FINAL PRODUCT, IF PRODUCED FOR SELLING

AGREEMENT

1 SEWAGE SLUDGE TO BE USED IN TERMS OF GUIDE
2 INSPECTION OF USER'S ACTIVITIES BY LOCAL AUTHORITY
3 BREACH OF CONTRACT - TERMINATION OF SEWAGE SLUDGE SUPPLY
AND PUNITIVE MEASURES

APPENDIX 2A: OPERATING SUMMARY - AEROBIC STAGE - 18/11/87 TO 09/01/90

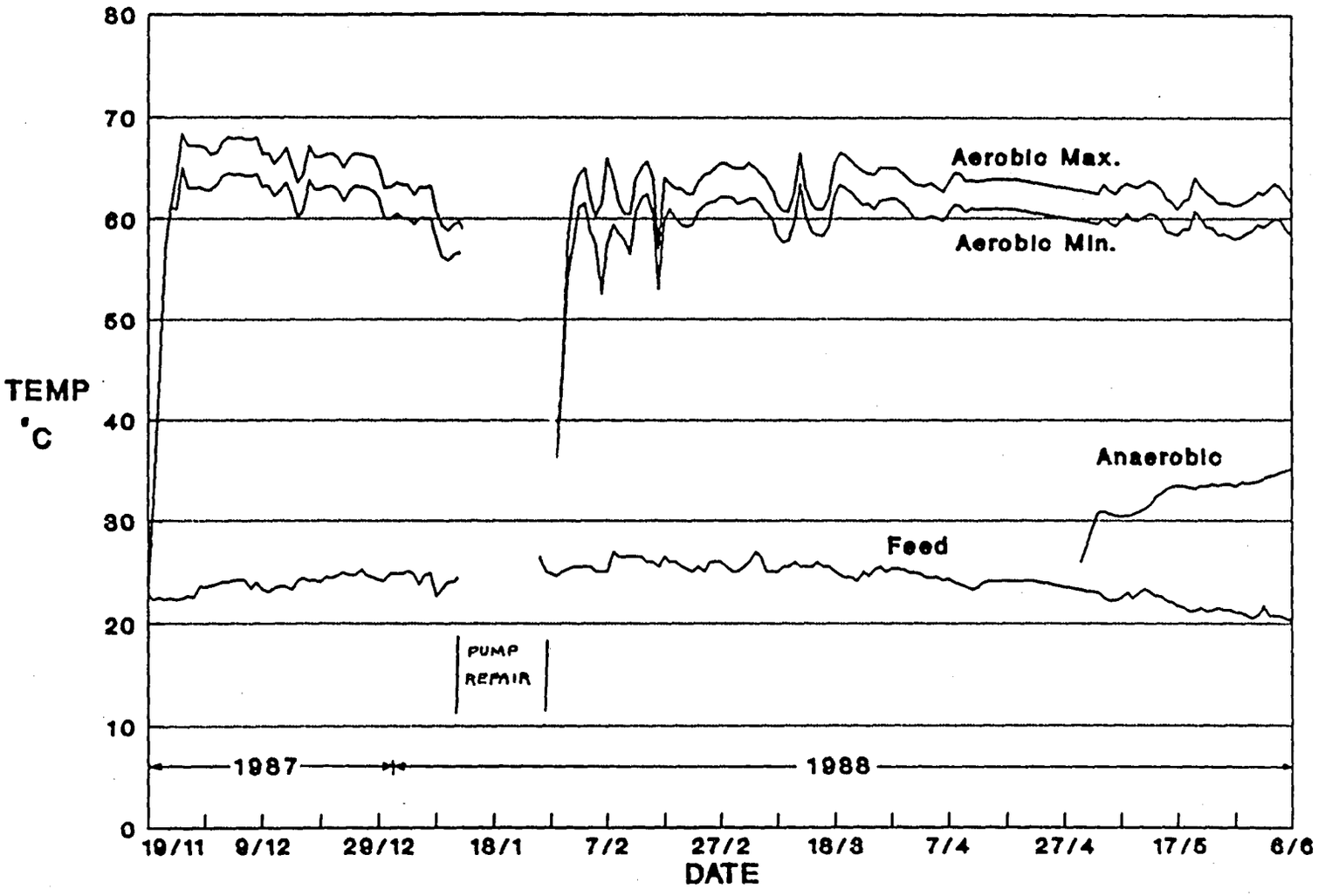
Period	Days	Reactor Temp (Mean) °C	Retention (Days)	Sludge Volume Treated m3	Oxygen Supplied (kg)	Oxygen Supplied kg/m3 Sludge	Power Supplied (kWh)	Power Supplied kWh/m3 Sludge
18/11-23/11/87	6		Startup					
24/11-7/12	13	67	1.75	335	4094	12.2	6500	19.4
08/12-17/12	10	66	1.25	380	4540	12.8	5000	13.9
18/12-10/01/88	24	63	1.00	1060	14450	13.4	12000	11.1
26/01-01/02	6		Startup					
02/02	1	61	1.50	30	583	19.4	500	16.7
03/02-05/02	3	63	1.00	135	1749	13.0	1500	11.1
06/02-08/02	3	62	1.00	90	864	9.6	1500	16.7
09/02-12/02	4	60	1.00	162	2332	14.4	2000	12.3
13/02-02/03	19	63	1.25	653	9120	14.0	9500	14.5
03/03-06/03	4	63	1.25	143	1564	11.1	2000	14.0
07/03-10/03	4	60	1.25	143	1392	9.7	2000	14.0
11/03	1	61	1.25	34	583	17.1	500	14.7
12/03-17/3	6	61	1.25	218	2068	9.6	3000	13.8
18/03	1	64	1.25	34	583	17.1	500	14.7
19/03-19/05	62	62	1.25	2205	29760	13.5	31000	14.1
20/05-23/06	35	61	1.25	1206	17634	14.6	17500	14.5
24/06-29/06	6	54	1.25	218	2124	9.7	3000	13.8
30/06-31/06	63	59	1.25	2138	33264	15.6	31500	14.7
01/09-13/09	13	56	1.25	465	5772	12.4	6500	14.0
14/09-22/09	9	63	1.50	255	3996	15.7	4500	17.6
23/09-03/10	Equipment inspection. Stop and restart on completion.							
04/10-18/10	15	66	3.00	244	2584	10.6	7500	30.7
19/10-26/10	8	58	1.20	293	3728	12.7	4000	13.7
27/10-31/10	5	57	1.50	148	2330	16.0	2500	17.1
01/11-09/11	9	60	1.50	270	3564	13.2	4500	16.7
10/11-13/11	4	63	1.50	120	2112	17.6	2000	16.7
14/11-17/11	4	64	1.50	124	1548	12.5	2000	16.1
18/11-28/11	11	57	1.20	398	4250	10.7	5500	13.8
29/11-19/12	Plant stopped - Pump failure.							
20/12-05/01/89	Restarting attempts.							
06/01-09/01	4	62	2.00	71	910	12.8	2000	28.2
10/01-12/01	3	59	1.20	116	1044	9.0	1500	12.9
13/01-16/01	4	58	1.50	120	947	7.9	2000	16.7
17/01-19/02	34	58	2.00	776	5389	6.9	17000	21.9
20/02-29/03	38	59	1.00	1586	16688	10.4	19000	11.9
30/03-03/04	5	58	1.20	184	1882	10.2	2500	13.6
04/04-01/05	28	57	2.00	641	6087	9.5	14000	21.8
02/05-02/08	93	58	1.50	2723	32880	12.1	46500	17.1
03/08-26/09	55	57	1.20	2048	36433	17.8	27500	13.4
27/09-16/11	51	59	1.50	1519	17667	11.6	25500	16.8
17/11-25/12	39	61	2.00	866	7720	8.9	19500	22.5
26/12-28/12	Electrical fault - Plant off and restart.							
29/12-31/12	3	58	3.00	45	454	10.1	1500	33.3
01/01-09/01/90	9	67	2.00	188	1244	6.6	4500	23.9
Average/Total	715	61	1.40	22396	285851	12.8	351500	15.7

APPENDIX 2B: ANAEROBIC DIGESTER - WEEKLY TOTALS / AVERAGES 1988/89

WEEK No.	WEEK	AVG TEMP °C	FEED VOL m3	AVG DIG VOL	GAS PROD m3	GAS/ FEED RATIO	AVG RTN d	MIXER RUN h/d
1	30/4-6/5/88	29.4	180	569	1593	8.9	22	24
2	7/5-13/5	31.1	139	662	2514	18.1	33	24
3	14/5-20/5	33.2	244	865	3435	14.1	25	24
4	21/5-27/5	33.5	251	897	3295	13.1	25	24
5	28/5-3/6	34.0	251	630	3187	12.7	18	10
6	4/6-10/6	34.7	244	488	2439	10.0	14	10
7	11/6-17/6	33.8	191	582	2234	11.7	21	12
8	18/6-24/6	35.0	251	533	2346	9.3	15	10
9	25/6-1/7	31.7	41	531			26	10
10	2/7-8/7	32.7	251	659			18	9
11	9/7-15/7	34.6	146	637			17	9
12	16/7-22/7	34.9	252	706			20	13
13	23/7-29/7	34.1	96	731			23	6
14	30/7-5/8	32.6	184	621			24	9
15	6/8-12/8	34.7	255	740			20	9
16	13/8-19/8	36.1	251	711			20	9
17	20/8-26/8	37.2	251	708			20	9
18	27/8-2/9	35.3	49	685			28	8
19	3/9-9/9	30.9	131	666	1740	12.6	20	8
20	10/9-16/9	30.5	109	694	1448	13.3	25	9
21	17/9-23/9	33.1	228	564	2000	8.8	17	11
22	24/9-30/9	31.8	0	420	263			8
23	1/10-7/10	28.8	56	423	575	10.2	30	8
24	8/10-14/10	31.0	105	442	1603	15.3	29	8
25	15/10-21/10	34.1	173	428	2674	15.5	17	13
26	22/10-28/10	37.8	248	436	3016	12.2	12	9
27	29/10-4/11	36.2	150	410	2566	17.1	19	8
28	5/11-11/11	37.2	180	413	2877	16.0	16	10
29	12/11-18/11	39.6	188	588	2730	14.6	22	8
30	19/11-25/11	37.8	128	432	1756	13.8	24	8
31	26/11-2/12	38.8	124	393	1961	15.9	13	8
32	3/12-9/12	30.3	0	370	460			0
33	10/12-16/12	27.6	0	370	202			0
34	17/12-23/12	27.5	0	370	158			0
35	24/12-30/12	26.7	0	370	96			0
36	31/12-6/1/89	26.4	0	370	0			0
TOTAL	36 WEEKS	1194.7	5347	20114	47168			345
AVG		33.2	149	559	1814	13.1	21.1	9.6

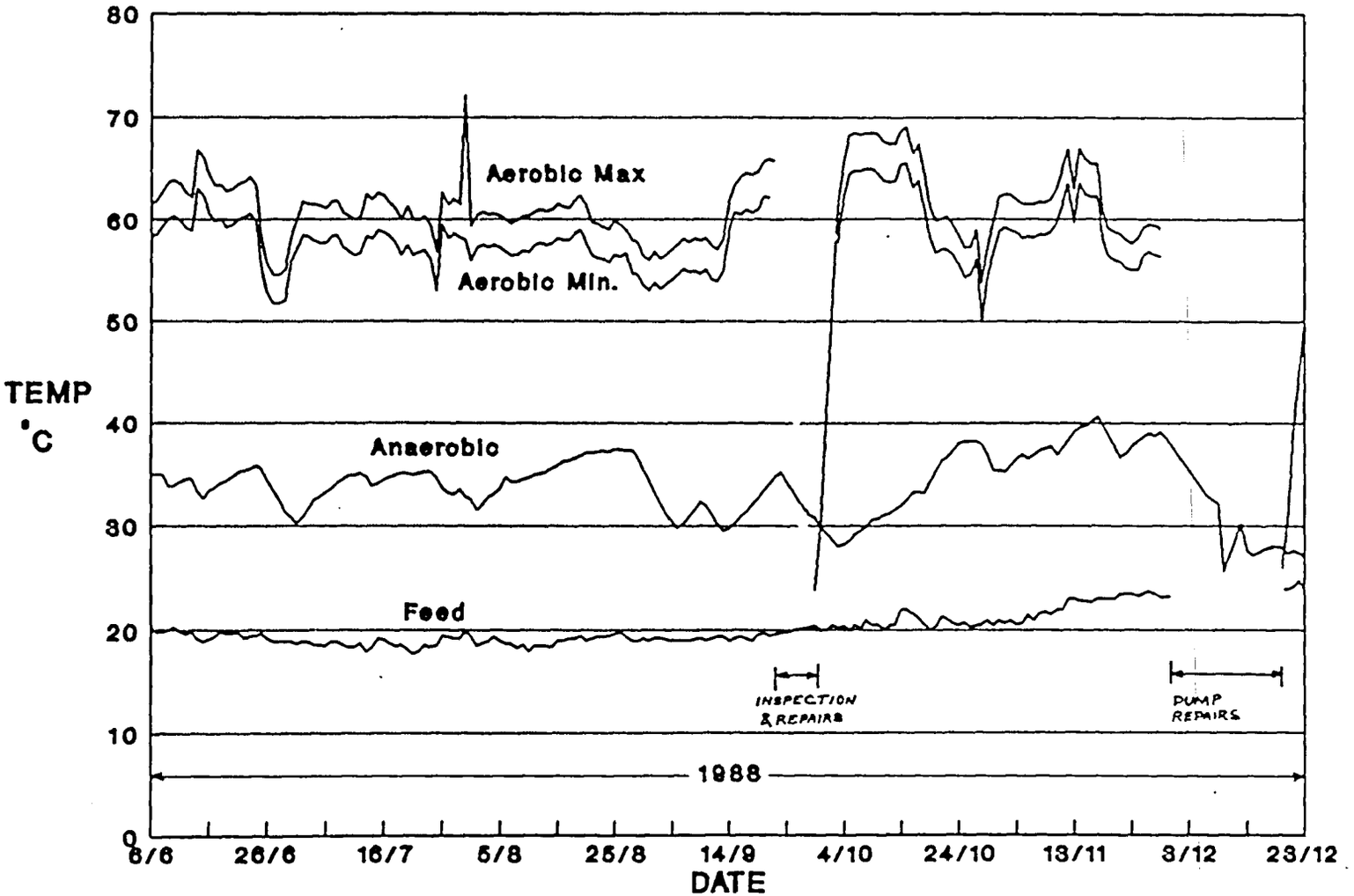
APPENDIX 2C: ANAEROBIC DIGESTER - WEEKLY TOTALS / AVERAGES 1999/00

WEEK No.	WEEK	AVG TEMP °C	FEED VOL m3	AVG DIG VOL	GAS PROD. m3	GAS/ FEED RATIO	AVG RTN d	MIXER RUN h/d
1	7/1-13/1	30.4	187.5	425	1025	5.46	15.9	6.0
2	14/1-20/1	37.6	176.3	463	2029	11.51	19.2	10.3
3	21/1-27/1	38.3	157.5	462	2302	14.61	20.5	8.5
4	28/1-3/2	38.7	120.0	447	2227	18.56	26.1	8.5
5	4/2-10/2	38.8	131.3	400	2112	16.09	24.5	8.9
6	11/2-17/2	38.6	157.5	390	2061	13.21	17.3	9.3
7	18/2-24/2	38.1	127.5	363	2076	16.28	21.6	8.6
8	25/2-3/3	38.7	127.5	425	1601	12.56	23.3	8.2
9	4/3-10/3	38.7	176.3	463	2136	12.13	18.4	9.0
10	11/3-17/3	38.6	150.0	505	2534	16.89	23.8	8.4
11	18/3-24/3	37.5	127.5	506	1155	9.06	27.9	11.5
12	25/3-31/3	37.9	123.8	404	1525	12.32	22.8	8.8
13	1/4-7/4	38.9	191.3	477	2550	13.33	17.5	8.5
14	8/4-14/4	38.8	157.5	487	2361	14.99	21.7	9.5
15	15/4-21/4	38.8	172.5	436	2737	15.87	17.7	10.5
16	22/4-28/4	38.3	157.5	466	2871	18.23	21.7	7.2
17	29/4-5/5	37.5	187.5	579	3160	16.86	21.6	8.8
18	6/5-12/5	37.8	176.3	547	3024	17.16	21.7	7.9
19	13/5-19/5	38.4	191.3	629	3072	16.06	23.0	8.2
20	20/5-26/5	37.0	210.0	501	3026	14.41	19.7	8.4
21	27/5-2/6	37.3	210.0	545	2671	12.72	18.2	8.8
22	3/6-9/6	37.5	210.0	565	3387	16.13	19.5	8.0
23	10/6-16/6	37.9	210.0	538	2769	13.19	17.9	7.9
24	17/6-23/6	38.3	210.0	656	3092	14.73	21.9	8.1
25	24/6-30/6	37.9	210.0	535	2570	12.24	17.8	7.9
26	1/7-7/7	37.3	210.0	535	2779	13.23	17.8	7.9
27	8/7-14/7	36.8	202.5	433	2532	12.51	15.0	7.9
28	15/7-21/7	36.9	196.8	444	2769	14.03	15.8	7.9
29	22/7-28/7	36.4	210.0	485	2809	13.37	16.2	7.9
30	29/7-4/8	36.0	228.8	459	2839	12.41	14.0	7.9
31	5/8-11/8	37.6	262.5	462	3209	12.22	12.9	7.8
32	12/8-18/8	36.0	258.8	456	2859	11.05	12.4	8.4
33	19/8-25/8	37.9	262.5	456	2726	10.38	12.2	7.6
34	26/8-1/9	37.7	262.5	444	2169	8.26	11.8	8.8
35	2/9-8/9	37.8	258.8	432	2035	7.86	11.7	8.4
36	9/9-15/9	36.5	262.5	643	2430	9.26	17.1	10.6
37	16/9-22/9	39.0	255.0	686	2224	8.72	18.9	8.2
38	23/9-29/9	39.1	243.8	716	1983	8.13	20.5	8.8
39	30/9-6/10	39.2	210.0	649	2915	13.88	21.8	8.7
40	7/10-13/10	38.9	196.8	667	3197	16.09	23.5	9.0
41	14/10-20/10	38.8	206.3	635	3287	15.94	21.6	9.0
42	21/10-27/10	38.6	210.0	622	3312	15.77	20.7	8.5
43	28/10-3/11	38.7	210.0	625	3185	15.17	20.8	8.8
44	4/11-10/11	38.3	210.0	632	3083	14.68	21.1	9.0
45	11/11-17/11	39.1	202.5	617	2714	13.40	21.3	8.9
46	18/11-24/11	38.9	157.5	624	2379	15.10	27.7	9.2
47	25/11-1/12	38.5	157.5	618	2266	14.39	27.5	8.9
48	2/12-8/12	38.2	157.5	601	2436	15.47	26.7	8.5
49	9/12-15/12	38.2	157.5	632	2554	16.22	28.1	8.5
50	16/12-22/12	38.2	146.3	629	2667	18.23	30.1	8.8
51	23/12-29/12	35.8	82.5	613	1746	21.16	52.0	5.0
52	30/12-5/1/00	31.9	142.5	650	2313	16.23	31.9	3.6
53	6/1-12/1	33.0	75.0	631	1533	20.44	58.9	6.4
TOTAL	53 WEEKS	1996	9696	28578	133062	738	1151	444
AVG		37.7	186.7	539.2	2510.6	13.9	21.7	8.4

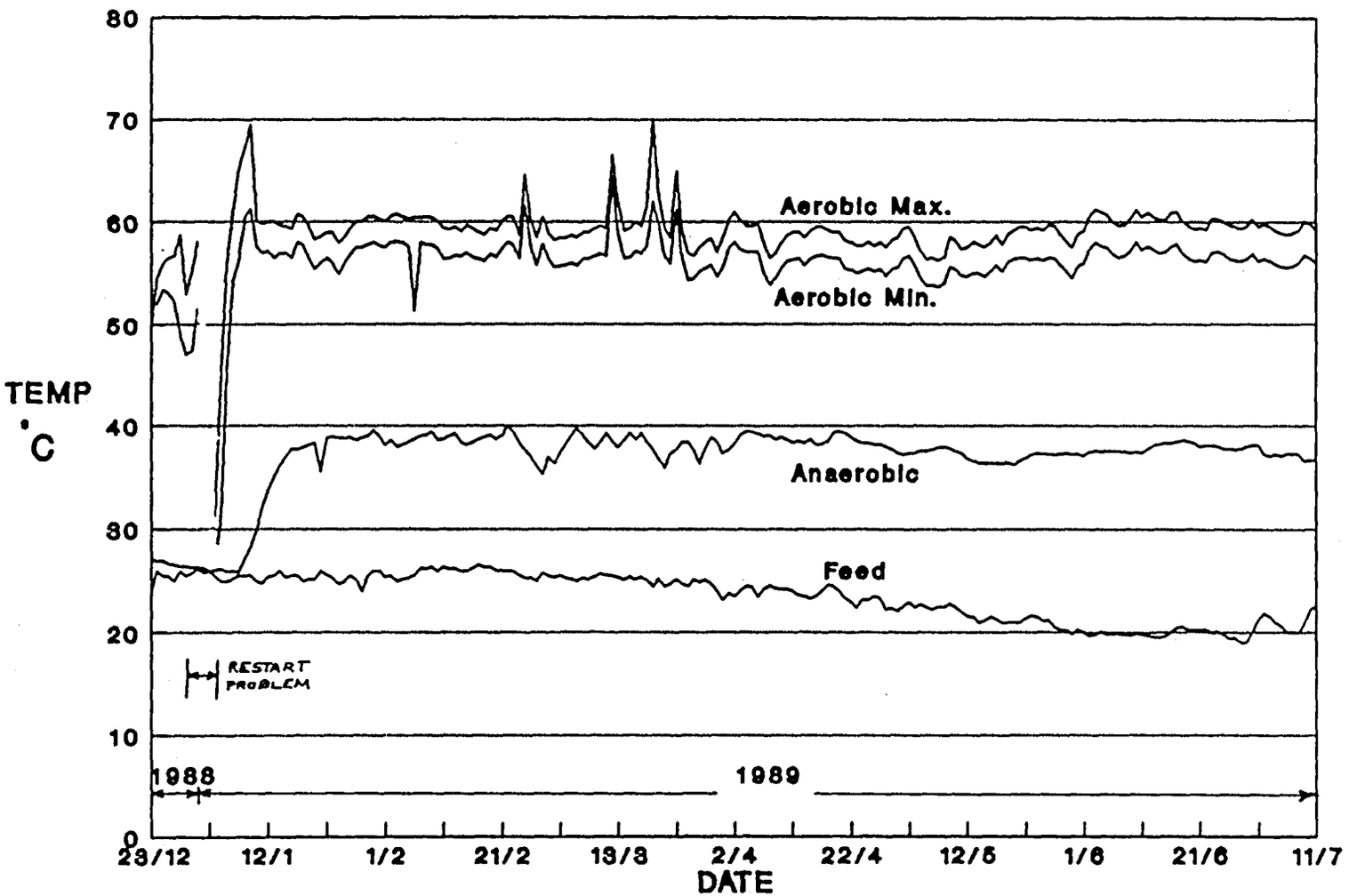


APPENDIX 2D: DUAL DIGESTION TEMPERATURES

(1987 - 1988)

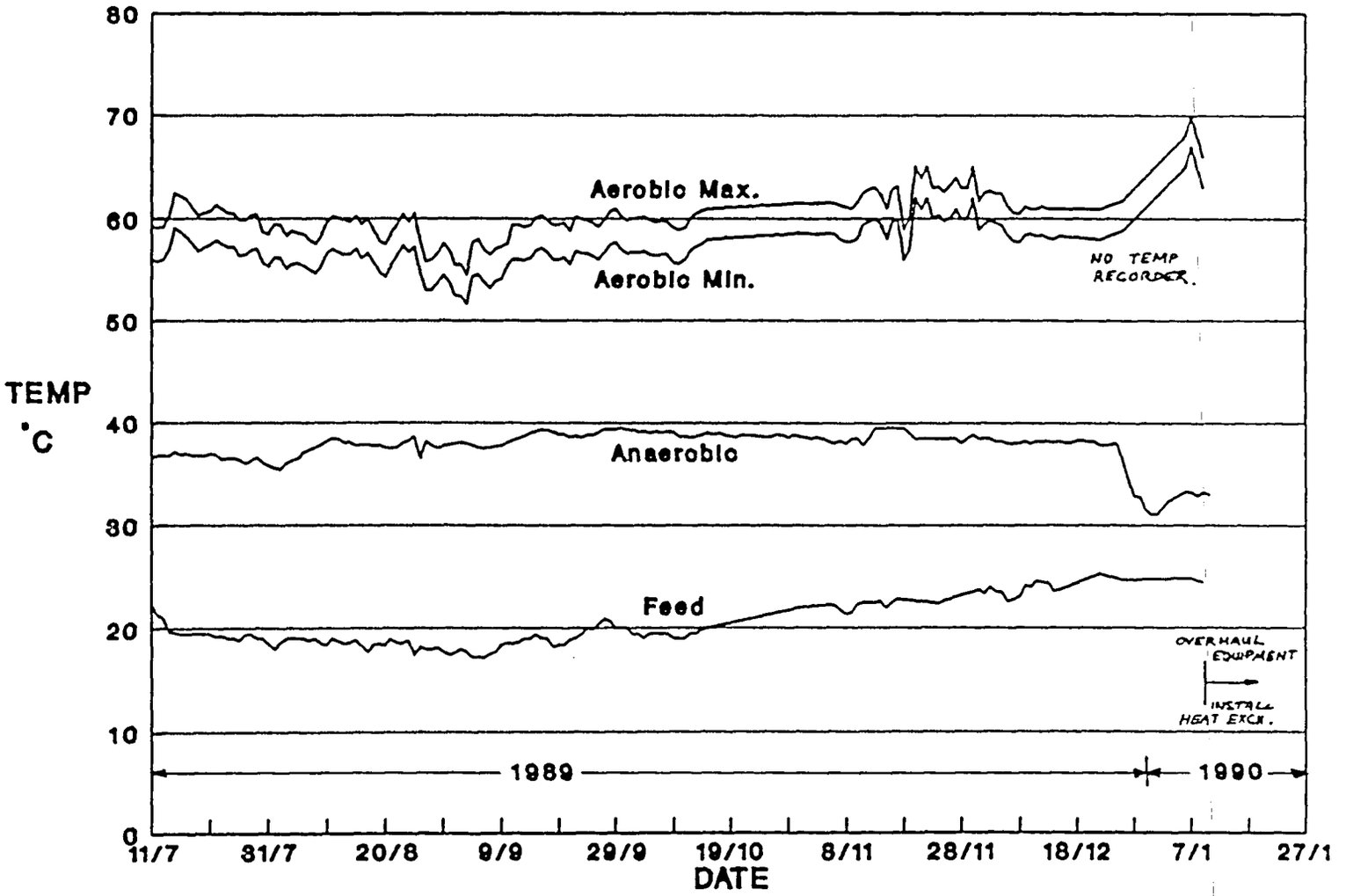


APPENDIX 2E: DUAL DIGESTION TEMPERATURES
(1988)



APPENDIX 2F: DUAL DIGESTION TEMPERATURES

(1988 - 1989)



APPENDIX 2G: DUAL DIGESTION TEMPERATURES

(1989 -1990)

APPENDIX 3A: ANALYTICAL RESULTS - DUAL DIGESTION AT MILNERTON 1987/88.

Date	FEED										TRANSFER									
	TS g/l	VS g/l	%VS %	NH4-N mg/l	TKN mg/l	TOT P mg/l	COD g/l	pH	Alk 5.75 mg/l	Alk 4.3 mg/l	TS g/l	VS g/l	%VS %	NH4-N mg/l	TKN mg/l	TOT P mg/l	COD mg/l	pH	Alk 5.75 mg/l	Alk 4.3 mg/l
30/3/87	35.9	29.7	82.7	112	1670	437	84.2	5.5			31.1	25.0	80.4	294	1560	375	43.7	6.6		
20/7	46.6	37.5	77.2	126	1615	463	87.6	5.6	0	600	28.1	22.4	78.7	320	1378	344	29.6	7.5	480	625
28/7	45.2	35.9	79.4	178	1770	484	84.5	5.3	0	725	34.9	26.4	75.6	336	1566	430	51.8	7.7	640	825
3/8	37.0	29.8	80.5	140	1372	362	48.1	5.5	0	675	28.3	22.5	79.5	238	1504	332	33.7	7.5	540	800
10/8	40.6	33.1	81.5	154	1652	480	53.1	5.9	20	640	27.8	22.1	79.2	180	1333	404	31.8	7.0	340	560
17/8	37.1	30.5	82.2	112	1366	332	44.5	5.8	0	755	34.5	26.5	76.8	248	1512	388	37.0	7.4	710	1015
26/8	40.0	33.2	83.0	115	1033	325	53.4	6.0	61	825	32.7	25.7	78.6	324	1470	325	40.6	7.7	813	1144
8/12	30.9	26.0	84.1	196	1230	373	47.2	5.3	0	702	33.3	26.8	80.5	368	1360	326	36.9	7.7	782	1067
11/1/88	52.7	44.1	83.7	133	1262	245	50.8	5.7	0	800	27.7	22.7	81.9	101	1327	240	22.9	6.7	300	575
N	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9	6	6
MAX	52.7	44.1	84.1	196	1815	484	84.2	6.0	61	825	34.9	26.8	81.9	368	1560	430	51.8	7.7	813	1144
MIN	30.9	26.0	77.2	112	1033	245	44.5	5.3	0	640	27.7	22.1	75.6	101	1327	240	22.9	6.6	300	560
SUM	368.0	298.8	734.4	1263	13210	3501	473.4	50.4	101	5922	276.5	220.1	712.3	2411	13082	3184	331.0	65.8	4565	6731
AVG	40.8	33.3	81.6	140	1468	388	52.6	5.6	13	740	30.9	24.5	79.1	268	1451	352	36.8	7.3	573	841

APPENDIX 3B: ANALYTICAL RESULTS - DUAL DIGESTION AT MILNERTON 1996

Date	FEED								TRANSFER								ANAEROBIC DIGESTER									
	TS	VS	%VS	NH4-N	COD	pH	Alk 5.75	Alk 4.3	TS	VS	%VS	NH4-N	COD	pH	Alk 5.75	Alk 4.3	TS	VS	%VS	NH4-N	COD	pH	Alk 5.75	Alk 4.3	IA/PA Ratio	Vol Acid
	g/L	g/L	%	mg/L	g/L		mg/L	mg/L	g/L	g/L	%	mg/L	mg/L		mg/L	mg/L	g/L	g/L	%	mg/L	g/L		mg/L	mg/L		mg/L
15/2/96	30.9	25.2	81.6	65	39.0	8.5	0	653	32.2	26.0	80.8	127	36.0	7.1	375	686										
29/2	35.7	29.7	83.4	90	46.5	8.6	0	725	30.5	24.8	81.3	92	38.8	6.8	388	710	25.8	21.2	81.8	230	36.2	5.2	0	1336	-	3180
7/3	28.7	23.4	81.5	113	38.8	8.8	0	808	31.4	25.2	80.3	167	36.1	6.7	313	756	20.8	16.6	79.4	220	33.5	5.8	0	1382	-	-
15/3	42.1	33.3	79.1	76	57.1	8.3	0	675	31.8	25.8	81.4	86	40.8	6.9	475	713	24.8	18.8	76.7	187	32.9	5.4	0	2081	-	-
22/3	36.8	29.8	80.8	118	48.8	8.8	0	880	31.8	25.7	80.8	225	36.8	6.8	950	813	27.8	18.8	66.3	344	34.6	6.7	513	3265	5.404	4200
29/3	43.8	34.8	79.3	145	51.8	8.3	0	738	33.7	26.8	79.8	211	38.8	6.7	475	813	26.8	18.8	66.4	364	34.8	7.4	1413	4463	2.156	5250
11/4	33.8	28.3	83.5	147	48.8	8.8	0	650	31.4	26.8	82.8	261	36.8	7.8	878	875	30.2	20.1	66.8	467	35.8	7.6	2713	4850	0.798	2150
18/4	30.4	25.7	84.5	156	41.8	8.8	10	888	31.8	26.3	84.8	208	40.8	6.8	812	1012	30.8	21.1	68.5	443	36.8	7.4	2500	4850	0.824	2850
25/4	30.8	25.3	82.1	141	34.1	8.8	0	800	31.5	25.8	82.2	175	37.2	7.3	531	819	30.7	20.8	67.8	421	37.7	7.7	2788	5013	0.798	3000
2/5	30.4	24.8	81.8	128	40.0	8.7	0	133	32.7	26.8	81.3	140	36.3	7.0	450	884	27.3	18.3	70.7	256	33.8	7.8	2250	3763	0.636	2670
9/5	33.2	28.1	84.6	148	54.8	8.7	0	819	32.3	26.8	83.0	208	47.0	6.8	850	888	28.1	18.1	66.3	486	33.3	7.7	2875	3752	0.433	675
17/5	34.3	28.4	82.8	188	46.1	8.7	0	575	34.8	28.8	83.5	198	46.8	6.7	444	713	23.4	16.8	70.8	512	26.5	7.8	2650	3350	0.264	317
7/6	38.2	31.8	81.2	256	48.5	8.8	0	863	33.5	26.4	78.8	238	38.7	7.1	475	675	22.3	15.8	70.8	588	25.8	7.5	2113	2563	0.213	265
15/6	38.4	31.8	80.2	188	56.4	8.8	25	700	36.8	28.8	78.3	237	46.8	6.8	731	813	22.8	16.2	73.8	813	25.8	7.4	1988	2219	0.330	182
20/6	38.8	32.1	83.4	248	51.4	8.8	63	800	32.1	26.1	81.3	375	36.8	7.8	888	888	20.8	15.2	73.1	822	24.1	7.8	1863	2375	0.275	182
29/6	36.8	30.4	84.4	288	53.8	8.8	19	888	34.8	28.1	82.8	348	42.8	7.1	563	875	20.1	14.9	74.1	880	24.2	7.7	2038	2588	0.281	294
5/7	36.1	29.8	82.5	253	46.8	8.5	0	800	31.8	25.8	81.8	304	36.8	7.1	850	875	18.8	13.4	71.3	838	22.8	7.5	1875	2400	0.280	380
12/7	43.8	31.7	72.8	276	48.4	8.8	25	850	43.8	29.7	68.1	283	42.3	6.8	425	763										
19/7	43.8	34.8	78.5	253	56.8	8.8	25	813	42.8	33.8	78.8	352	46.8	6.8	536	1075	18.8	13.4	67.7	581	21.3	7.3	1813	2513	0.366	240
3/8	35.8	28.2	82.0	234	47.7	8.7	0	731	36.4	31.2	81.3	277	45.8	7.8	300	463	18.8	13.5	67.8	721	22.8	7.5	1513	1988	0.314	262
17/8	36.1	33.8	84.8	194	65.5	8.8	38	584	31.7	26.2	82.8	220	45.7	6.9	450	825	20.8	14.3	71.5	856	25.2	7.3	1963	2575	0.312	72
25/8	38.2	32.3	82.4	211	52.2	8.4	0	856	35.8	29.8	82.5	208	50.2	6.7	325	738	16.5	13.2	71.4	838	20.8	7.3	1800	2488	0.308	125
8/9	34.8	28.5	76.8	225	48.2	8.7	0	744	35.2	27.8	78.4	258	38.5	7.3	844	1031	17.8	12.3	68.1	721	22.1	7.4	1918	2831	0.475	250
13/9	35.7	27.9	78.2	222	45.8	8.8	25	863	33.8	26.2	77.5	262	42.0	6.8	528	850	16.3	12.1	66.1	686	21.8	7.6	1913	2725	0.424	80
12/10	41.4	31.8	76.8	234	56.0	8.2	0	950	43.8	32.7	74.8	431	46.2	7.5	663	1019	18.1	13.3	68.8	722	21.8	7.5	2113	2963	0.290	180
18/10	41.2	33.1	80.3	220	72.7	8.3	0	825	40.3	31.2	77.4	451	46.8	7.1	750	1238	20.3	13.8	68.0	732	23.7	7.5	2088	2725	0.306	125
1/11	42.8	33.8	78.3	144	57.8	8.7	0	831	42.8	33.8	78.3	256	66.8	7.0	650	650	18.5	13.0	68.7	653	23.8	7.8	2250	2619	0.184	180
10/11	36.4	28.1	78.4	200	50.4	8.2	0	763	36.8	28.8	77.2	352	44.8	7.4	825	1263	21.3	13.9	65.3	662	25.8	7.4	1850	2475	0.336	135
24/11	36.1	30.8	80.1	197	48.8	8.4	0	744	36.8	29.5	79.8	270	48.4	7.0	429	900	20.8	14.0	70.0	783	25.8	7.6	1488	2700	0.815	185
N	28	28	28	28	28	28	28	28	25	25	25	25	25	25	25	25	17	17	17	17	17	17	17	17	17	17
MAX	43.8	34.8	84.8	276.0	72.7	8.8	63	863	43.8	33.8	84.8	451.8	66.8	7.8	825	1263	23.4	16.8	74.1	783	26.5	7.8	2650	3350	0.8	380
MIN	28.7	23.4	72.8	76.8	34.1	8.2	0	133	30.5	24.8	68.1	86.0	36.8	6.8	300	463	17.8	12.1	65.3	512	20.8	7.3	1488	1988	0.2	72
SUM	1088.8	863.0	2346.8	5317.8	1460.7	182.8	236	20472	1014.5	808.1	2320.7	7252.8	1260.3	202.8	15331	24846	341.8	236.8	1187.0	11150	403.8	127.8	33018	43778	5.7	3534
AVG	38.8	29.8	80.9	183.3	50.4	8.8	8	706	35.0	27.9	80.0	250.1	43.8	7.0	529	860	20.1	14.1	69.8	656	23.8	7.5	1942	2575	0.3	208

APPENDIX 3C: ANALYTICAL RESULTS - DUAL DIGESTION AT MILMERTON 1988/89.

Date	FEED								TRANSFER								ANAEROBIC DIGESTER								
	TS	VS	%VS	NH4-N	COD	pH	Alk 5.75	Alk 4.3	TS	VS	%VS	NH4-N	COD	pH	Alk 5.75	Alk 4.3	TS	VS	%VS	NH4-N	COD	pH	Alk 5.75	Alk 4.3	IA/PA Ratio	Vol Acid
	g/L	g/L	%	mg/L	g/L		mg/L	mg/L	g/L	g/L	%	mg/L	mg/L		mg/L	mg/L	g/L	g/L	%	mg/L	g/L		mg/L	mg/L		mg/L
31/1/89	29.9	24.2	80.9	155	47.6	6.0	50	975	29.8	24.1	81.7	380	46.5	7.3	700	1013	17.0	11.2	65.9	614	21.0	7.6	1337	1875	0.253	144
7/2	43.4	35.3	81.3	180	46.5	5.5	0	736	31.3	25.5	81.5	181	37.0	6.6	266	556	15.1	10.7	70.9	561	17.9	7.5	1413	1775	0.256	150
7/2																	15.7	11.2	71.3	616	18.7	7.7	2063	2675	0.297	182
16/2	36.2	31.4	82.2	231	42.7	5.3	0	750	27.7	22.6	81.6	403	31.6	7.3	975	1363	25.2	16.3	72.6	850	29.7	7.7	1966	2436	0.236	372
21/2	37.2	30.6	82.3	177	46.2	5.1	0	763	26.6	23.6	81.8	421	35.0	7.0	1013	1484	18.4	14.0	72.2	626	26.5	7.5	1863	2400	0.266	240
2/3	43.4	35.7	82.3	180	46.5	5.5	0	736	31.3	25.5	81.5	181	37.0	6.6	266	556	15.1	10.7	70.9	561	17.9	7.5	1413	1775	0.256	150
9/3	29.3	24.2	82.9	186	36.2	5.3	0	700	23.2	19.4	83.6	445	27.0	7.0	450	988	19.0	13.6	72.6	643	20.9	7.7	1966	2466	0.242	264
14/4	46.1	41.0	85.2	186	36.1	5.2	0	713	26.6	24.3	84.4	466	46.2	7.1	1125	1566	12.9	9.6	74.4	676	21.0	7.6	2036	2513	0.233	136
28/4	35.3	28.4	80.5	182	46.0	5.5	0	436	26.3	20.7	77.5	566	46.6	7.4	900	1266	16.2	12.2	75.3	667	22.4	7.6	1536	1863	0.276	96
17/5	36.5	30.9	83.3	222	56.4	5.6	25	650	33.6	27.0	79.9	607	46.0	7.1	1223	1715	16.6	11.5	68.5	736	20.0	7.5	2327	2960	0.272	112
30/5									34.7	28.4	81.8	383	36.1	7.2	836	1106	16.9	12.0	71.0	703	20.4	7.5	2221	2463	0.109	132
4/7	36.1	26.3	80.6	196	46.1	5.6	0	754	32.4	25.6	79.8	253	35.0	7.1	533	786	22.6	16.6	73.7	660	21.1	7.7	1716	2015	0.173	108
21/7	36.6	29.4	80.3	181	47.0	5.6	0	901	36.3	28.3	80.2	262	43.4	7.2	415	654	14.9	10.3	68.1	625	18.7	7.5	1854	2366	0.276	120
15/8																	24.7	18.6	80.2	556	31.6	7.6	1718	2184	0.256	60
29/8	41.0	30.1	73.4	225	57.6	5.6	0	734	36.6	27.7	75.7	136	44.0	6.6	316	513	22.6	15.7	69.5	611	20.4	7.3	819	1733	1.116	824
5/9	40.5	30.6	75.6	186	54.6	5.1	78	966	36.1	26.2	74.0	176	42.5	7.0	276	460	27.9	19.5	69.9	606	32.2	7.2	896	1750	1.504	1374
12/9	32.7	26.6	81.3	156	46.1	6.0	64	876	33.6	27.6	82.2	187	43.6	7.0	370	621	20.6	14.6	71.2	560	27.6	7.0	671	1720	1.563	1666
19/9																	20.1	14.3	71.1	629	31.4	7.6	1553	2036	0.312	450
26/9	36.6	30.0	81.5	186	46.2	5.6	0	794	34.7	27.0	77.6	183	36.1	7.2	300	500	18.9	14.2	71.4	679	25.0	7.5	1900	2319	0.221	132
10/10	36.9	29.6	80.5	186	50.4	5.9	50	705	33.9	26.6	78.1	328	33.1	7.4	636	815	17.2	11.8	68.6	712	18.9	7.8	2018	2365	0.182	150
2/11	36.5	31.4	81.6	196	47.3	5.9	46	725	32.5	26.1	80.3	345	35.7	7.5	861	1196	17.6	12.6	71.6	666	17.3	7.6	2131	2570	0.206	180
16/11	36.0	29.0	80.6	136	43.7	5.5	0	636	33.2	26.3	79.2	186	36.6	7.0	431	864	17.2	12.0	69.8	646	16.6	7.7	1966	2466	0.266	106
23/11	36.0	28.6	79.4	91	72.6	5.2	0	650	30.0	21.6	72.7	290	37.2	7.2	615	966	17.9	12.1	67.6	625	22.4	7.5	1866	2501	0.324	132
14/12	36.6	30.1	82.2	172	43.4	5.7	0	793	31.6	25.3	80.1	348	34.0	7.3	756	1044	18.2	12.3	67.6	715	20.3	7.8	2136	2513	0.175	66
3/1/90	33.6	27.6	82.1	156	56.6	5.1	0	666	30.3	24.1	79.5	360	40.6	6.9	675	1215	14.5	10.2	70.3	707	17.6	7.6	1650	2513	0.356	132
17/1																	12.2	8.1	66.4	732	15.5	7.7	1955	2567	0.313	66
N	31	21	21	21	21	21	21	21	22	22	22	22	22	22	22	22	26	26	26	26	26	26	26	26	26	26
MAX	46.1	41.0	85.2	231	72.6	6.1	78	794	36.3	28.7	84.4	566	97.2	7.5	1223	1715	27.9	19.5	80.2	736	32.2	7.8	2327	2960	1.56	1666
MIN	29.3	24.2	73.4	91	36.2	5.1	0	436	23.2	19.4	72.7	136	27.0	6.6	276	460	12.2	8.1	65.9	556	15.5	7.0	671	1675	0.11	60
SUM	784.6	636.2	1700.6	3877	1051.2	117.2	313	14264	708.8	566.4	1736.2	7004	863.7	156.6	14011	20896	477.8	338.7	1843.5	16813	564.2	197.8	45050	56752	9.87	7736
AVG	37.4	30.2	81.0	175	48.1	5.6	15	679	32.3	25.7	79.6	318	40.2	7.1	637	950	18.4	13.1	70.9	651	22.5	7.6	1733	2260	0.38	298

APPENDIX 4a

MICROBIOLOGICAL ANALYSES - AEROBIC PROCESS

(Preliminary Trials of Aerobic Reactor)

Reactor Conditions				Feed Sludge from Thickener						Transfer to Anaerobic Digester					
Date	React Temp. (°C)	Time Cycle (h)	Reten Time (d)	Total Coliform per g dry	Faecal Coliform per g dry	Coliphage per g dry	*Salmonellae* per g dry	*Ascaris lumbricoides* ova (per g dry) viable	Non-viable	Total Coliform per g dry	Faecal Coliform per g dry	Coliphage per g dry	*Salmonellae* per g dry	*Ascaris lumbricoides* ova (per 10g dry) Viable	Non-viable
30/3/87	65	2	1	$2,86x10^{13}$	$1,58x10^{13}$	$5,1x10^{5}$	0	223	368	$4,4x10^{5}$	$2,38x10^{5}$	$9,0x10^{3}$	0	0	424
20/7/87	58	1,7	0,9	$1,6x10^{13}$	$3,0x10^{12}$	$9,0x10^{5}$	0	0	1317	$1,6x10^{6}$	$7,6x10^{5}$	$1,4x10^{5}$	0	0	925
28/7/87	67	2	1	$1,2x10^{14}$	$5,8x10^{12}$	$1,3x10^{5}$	0	1725	88	$7,7x10^{4}$	$7,7x10^{4}$	$1,7x10^{3}$	0	0	0
3/8/87	58	2	1	$6,4x10^{12}$	$3,1x10^{12}$	$5,7x10^{4}$	0	486	1135	$1,3x10^{4}$	$1,3x10^{4}$	$1,9x10^{4}$	0	141	2403
10/8/87	57	3	1,5	$5,5x10^{12}$	$4,7x10^{11}$	$9,9x10^{5}$	0	542	0	$3,6x10^{5}$	$1,8x10^{5}$	$5,0x10^{4}$	0	0	143
17/8/87	67	3	1,5	$1,3x10^{13}$	$4,9x10^{12}$	$1,3x10^{6}$	0	485	54	$1,5x10^{4}$	$1,4x10^{3}$	$7,2x10^{3}$	0	0	0
28/9/87	65	2,5	1,25	$4,7x10^{11}$	$1,2x10^{10}$	$1,2x10^{6}$	0	175	0	$5,7x10^{5}$	$1,1x10^{5}$	$9,4x10^{4}$	0	0	0
8/12/87	68	3,5	1,75	$2,25x1^{011}$	$6,8x10^{8}$	$2,2x10^{6}$	0	388	518	$9,0x10^{3}$	0	$6,0x10^{3}$	0	0	0
11/1/88	59	2	1	$3,2x10^{10}$	$9,8x10^{7}$	$9,2x10^{5}$	0	-	-	$1,05x10^{7}$	$6,5x10^{6}$	$1,9x10^{5}$	0	-	-

APPENDIX 4B (1)

MICROBIOLOGICAL ANALYSES - AEROBIC PROCESS

Date	Reactor Temp max. (°C)	Time Cycle (d)	Retention Time (d)	Feed sludge			Transfer to A.D.		
				Faecal coli (per g dry)	Ascaris ova/10g dry		Faecal coli (per g dry)	Ascaris ova/10g dry	
					Viable	N-viable		Viable	N-viable
15/2/88	64	2,5	1,25	$7,77x10^{8}$	712	6194	0	0	0
29/2/88	66	2,5	1,25	$7,26x10^{8}$	611	266	0	0	68
07/3/88	63	2,5	1,25	$8,01x10^{7}$	0	321	0	0	0
15/3/88	61	2,5	1,25	$1,19x10^{8}$	0	428	$7,86x10^{1}$	0	63
22/3/88	65	2,5	1,25	$5,69x10^{7}$	0	108	*	0	31
28/3/88	65	2,5	1,25	$5,15x10^{7}$	0	368	*	0	158
11/4/88	64	2,5	1,25	$1,87x10^{8}$	246	197	$1,26x10^{4}$	0	100
19/4/88	64	2,5	1,25	$5,66x10^{8}$	122	244	$6,55x10^{3}$	0	98
26/4/88	63	2,5	1,25	$4,97x10^{10}$	403	161	0	0	427
02/5/88	63	2,5	1,25	$3,59x10^{9}$	301	0	0	0	311
09/5/88	63	2,5	1,25	$2,02x10^{11}$	31	217	$1,09x10^{4}$	0	146
17/5/88	61	2,5	1,25	$2,16x10^{12}$	102	266	$2,56x10^{5}$	0	81
07/6/88	62	2,5	1,25	$1,08x10^{12}$	196	66	0	0	676
15/6/88	66	2,5	1,25	$1,27x10^{12}$	0	53	$1,08x10^{3}$	0	163
20/6/88	63	2,5	1,25	$1,28x10^{13}$	40	80	$1,96x10^{3}$	0	116
28/6/88	55	2,5	1,25	$4,11x10^{12}$	164	817	$5,09x10^{3}$	0	735
06/7/88	61	2,5	1,25	$2,38x10^{12}$	203	264	$2,53x10^{3}$	36	348
12/7/88	60	2,5	1,25	$1,08x10^{12}$	406	180	$5,28x10^{4}$	0	610
19/7/88	60	2,5	1,25	$2,68x10^{12}$	228	913	$4,52x10^{4}$	0	730
03/8/88	61	2,5	1,25	$1,89x10^{12}$	102	153	$6,82x10^{3}$	0	210
17/8/88	61	2,5	1,25	$4,45x10^{12}$	210	105	$3,01x10^{4}$	0	588
14/11/90*	60	2,5	1,25	$5,18x10^{10}$	733	33	$3,67x10^{4}$	0	33

* Sample contained zero *Salmonellae*

APPENDIX 4B(2)

MICROBIOLOGICAL ANALYSES - ANAEROBIC PROCESS

Date	Digester Time	Retention Time	Transfer sludge to A.D.			Anaerobic digester sludge		
			Faecal coli	Ascaris ova/10g dry		Faecal coli	Ascaris ova/10g dry	
	(°C)	(d)	(per g dry)	Viable	N-viable	(per g dry)	Viable	N-viable
29/2/88	41	-	0	0	68	0	26	77
07/3/88	35	-	0	0	0	0	0	33
15/3/88	37	-	$7,8x10^{1}$	0	63	0	82	1224
22/3/88	35	-	0	0	31	$1,79x10^{2}$	0	143
28/3/88	33	-	0	0	158	$1,37x10^{4}$	0	24
11/4/88	26	-	$1,25x10^{4}$	0	100	$7,65x10^{4}$	0	167
19/4/88	25	-	$6,55x10^{3}$	0	98	$1,53x10^{4}$	0	175
25/4/88	22	-	0	0	427	$5,31x10^{3}$	0	170
02/5/88	29	22	0	0	311	$8,11x10^{4}$	0	349
09/5/88	31	33	$1,09x10^{4}$	0	146	$2,17x10^{4}$	0	359
17/5/88	34	25	$2,56x10^{5}$	0	81	$3,16x10^{3}$	0	101
07/6/88	35	14	0	0	675	$2,51x10^{3}$	0	806
15/6/88	33	21	$1,08x10^{3}$	0	163	$2,77x10^{3}$	0	350
20/6/88	35	15	$1,96x10^{3}$	0	116	$1,59x10^{4}$	0	267
28/6/88	33	26	$5,09x10^{3}$	0	735	$4,98x10^{2}$	0	697
05/7/88	33	18	$2,53x10^{2}$	35	348	$1,53x10^{5}$	57	848
19/7/88	35	20	$4,52x10^{4}$	0	730	$1,75x10^{4}$	0	947
03/8/88	33	24	$6,82x10^{3}$	0	210	$3,30x10^{4}$	0	251
17/8/88	36	20	$3,01x10^{4}$	0	588	$1,05x10^{4}$	0	600
14/11/90	36	22	$3,57x10^{4}$	0	33	$1,50x10^{2}$	0	0