The Performance and Kinetics of Biological Nitrogen and Phosphorus Removal with Ultra-Filtration Membranes for Solid-Liquid Separation

Report to the
Water Research Commission

by

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EXECUTIVE SUMMARY

E.1 MOTIVATION

Effective solids-liquid separation in suspended medium biological wastewater treatment systems is an essential step in the process, because it has a major influence on effluent quality. Traditionally in biological wastewater treatment systems this step has been accomplished in secondary settling tanks (SSTs). However, membranes in place of SSTs for solids-liquid separation would appear to offer several potential advantages, including:

(1) Insensitivity to sludge settleability and filamentous bulking; this is a significant advantage as biological nutrient removal (BNR) systems notoriously produce rather poor settling sludges (DSVI~150 mL/g) when aerobic mass fractions are low (<60%)

(2) Insensitivity to activated sludge flocculation characteristics and hydraulic shear in the reactor; membrane retain all solids > 0.1 μm, which includes free swimming bacteria.

(3) SSTs are not required – a wastewater treatment plant footprint reduction.

(4) Very high reactor concentrations 15 to 20 g TSS/ℓ (1.5 to 2%) resulting in reduced biological reactor volumes compared with conventional BNR systems with SSTs (a further footprint reduction).

(5) Production of possibly disinfected effluent for industrial or horticultural use; membranes appear to be an effective barrier against bacteria and viruses (Churchouse and Brindle, 2003), indicating effective pore sizes considerably smaller than the nominal pore size.

(6) Possibly obviate waste activated sludge thickening when reactor concentrations are at the high end of the range (2% TSS).

Membrane activated sludge systems have been operated successfully both technically and economically in the UK, Europe and Japan. However, application has been largely restricted to COD (BOD₅) and free and saline ammonia (FSA) removal (nitrification), i.e. mostly as fully aerobic systems. Membrane bioreactor (MBR) technology application to BNR activated sludge systems is limited and uncertainty exist as to the impact of the conditions induced by the MBRs on the biologically mediated processes of nutrient removal.

Accordingly, in April 2003, a one year consultancy (K8/514) was set up between the University of Cape Town (UCT), the Water Research Commission (WRC) and Aquator, to investigate BNR behaviour in MBR systems. The preliminary investigation under the consultancy indicated that a more in-depth investigation was warranted. Therefore, in April 2004 a follow-on two-year research contract (K5/1537) was set up between UCT and the WRC to continue the research.

E.2 OBJECTIVES AND AIMS OF RESEARCH

The principle objectives of both the research consultancy and contract were to:

Aim 1: Evaluate biological nutrient removal (BNR) performance at typical membrane bioreactor (MBR) total suspended solids (TSS) concentrations (14-18 g/ℓ).

Aim 2: To compare the performance and kinetics of biological N and P removal under MBR conditions (high reactor TSS concentration (16 g/ℓ)) with those in conventional BNR systems (low reactor TSS concentration (4 g/ℓ)).
Aim 3: To identify a BNR plant where MBR technology can be applied at full scale and negotiate as a project team with owners of BNR plants to implement BNR at a demonstration WWTP site.

During the course of the research it became evident that installing membranes for solids/liquid separation into BNR activated sludge systems makes a profound difference not only to the design of the BNR system itself, but also to the design approach for the whole wastewater treatment plant. Accordingly, a further objective for the project was identified which was not explicitly included in either project proposal:

Aim 4: Evaluate the impact of membrane solid liquid separation on the design of biological nutrient removal (BNR) activated sludge (AS) systems.

E.3 PROJECT RESULTS

E.3.1 Aim 4: The impact of membrane solid liquid separation on the design of BNRAS systems.

Installing membranes for solid-liquid separation into biological nutrient removal (BNR) activated sludge (AS) systems makes a profound difference not only to the design of the BNR system itself, but also to the design approach for the whole wastewater treatment plant (WWTP). In multi-zone BNR systems with membranes in the aerobic reactor and fixed volumes for the anaerobic, anoxic and aerobic zones (i.e. fixed volume fractions), the mass fractions can be controlled (within a range) with the inter-reactor recycle ratios. This zone mass fraction flexibility is a significant advantage of membrane BNR systems over conventional BNR systems with SSTs, because it allows changing the mass fractions to optimise biological N and P removal in conformity with influent wastewater characteristics and the effluent N and P concentrations required.

In contrast to conventional BNRAS systems with secondary settling tanks, the size of which is governed by organic (COD) load and system sludge age, the size of MBR BNR systems is governed by hydraulic load and oxygen transfer rate. For peak wet weather flow (PWWF) to Average dry weather flow (ADWF) ratios ($f_{q}$) in the upper range ($f_{q} \approx 2.0$), aerobic mass fractions in the lower range ($f_{\text{aer}} < 0.60$) and high (usually raw) wastewater strengths, the indicated mode of operation of MBR BNR systems is as extended aeration WWTPs. However, the volume reduction compared with equivalent conventional BNR systems with secondary settling tanks will not be large (40-60%), but the cost of the membranes can be offset against sludge thickening and stabilisation costs. Moving from a flow unbalanced raw wastewater system to a flow balanced ($f_{q}=1$) low (usually settled) wastewater strength system can double the ADWF capacity of the biological reactor, but the design approach of the WWTP changes away from extended aeration to include primary sludge stabilisation. The cost of primary sludge treatment then has to be paid from the savings from the increased WWTP capacity.

E.3.2 Aim 1: Evaluate MBR BNR performance at high total suspended solids (TSS) concentrations (14-18 g/l).

To assess the impact of high VSS concentration in membrane bioreactor biological nutrient removal (BNR) activated sludge (AS) systems on biological N and P removal performance, two identical (except for the hydraulic retention time) parallel laboratory scale University of Cape Town (UCT) nitrification denitrification (ND) biological excess phosphorus removal (BEPR) systems were operated. One system was a low VSS concentration (3 g VSS/l) and solid liquid separation with a
secondary settling tank (CAS system), the other at a high VSS concentration (13 g VSS/ℓ) and solid liquid separation with submerged panel membranes (MBR system). Both were fed the same real wastewater were operated over two successive periods (Phases 1 and 2) in excess of 450 days each from 2003 to 2006. From the BNR performance of these two systems, the following conclusions were drawn (Table E.1).

Table E.1: Summary of the influent and effluent qualities, and the resultant removals, of both UCT systems.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Units</th>
<th>Influent</th>
<th>MBR System</th>
<th>CAS System</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Effluent</td>
<td>Efficiency %</td>
<td>Effluent</td>
</tr>
<tr>
<td>COD</td>
<td>mg COD/ℓ</td>
<td>951</td>
<td>41, 115&lt;sup&gt;a&lt;/sup&gt; 69&lt;sup&gt;b&lt;/sup&gt;</td>
<td>95.7, 88.0&lt;sup&gt;a&lt;/sup&gt;, 92.7&lt;sup&gt;b&lt;/sup&gt;</td>
</tr>
<tr>
<td>TKN</td>
<td>mg N/l</td>
<td>106.5</td>
<td>1.52</td>
<td>98.6</td>
</tr>
<tr>
<td>FSA</td>
<td>mg N/l</td>
<td>81.7</td>
<td>0.74</td>
<td>99.1</td>
</tr>
<tr>
<td>NO&lt;sub&gt;3&lt;/sub&gt;</td>
<td>mg -NO&lt;sub&gt;3&lt;/sub&gt;-N/l</td>
<td>0</td>
<td>16.53</td>
<td>-</td>
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<tr>
<td>TN</td>
<td>mg N/l</td>
<td>106.5</td>
<td>18.05</td>
<td>83.1</td>
</tr>
<tr>
<td>TP</td>
<td>mg P/l</td>
<td>30.3</td>
<td>8.4</td>
<td>22.5&lt;sup&gt;c&lt;/sup&gt;</td>
</tr>
<tr>
<td>TSS</td>
<td>mg TSS/ℓ</td>
<td>N/A</td>
<td>0</td>
<td>-</td>
</tr>
<tr>
<td>E. coli</td>
<td>CFU/100 ml</td>
<td>N/A</td>
<td>0</td>
<td>-</td>
</tr>
</tbody>
</table>

<sup>a</sup> unfiltered sample; <sup>b</sup> 0.45 filtered sample; <sup>c</sup> P removal mg P/ℓ; N/A = value not available

E.3.2.1 Organics (COD) removal

The COD removal efficiency of the MBR system (96%) was superior to that of the CAS system (92% unfiltered, 95% 0.45 μm filtered). However, the MBR system produced an “effluent” 0.45 μm filtered COD concentration (measured on the supernatant of the MBR aerobic reactor mixed liquor) that was consistently higher than the effluent filtered (0.45 μm) COD concentration from the CAS system (Table 3). Similarly the MBR unfiltered “effluent” COD values (measured on the unfiltered supernatant at the 800 mℓ mark in the 1 ℓ DSVI measuring cylinder test on MBR aerobic reactor sludge after 30 minutes settling) were consistently higher (139 mg COD/ℓ) than those from the CAS system (73 mg COD/ℓ).

From these observations (Table 1), the following aspects can be highlighted:

1. The membranes, because of smaller pore size, retain organics that would be considered “soluble” (<0.45 μm) in a CAS system with SSTs;
2. Even though the total COD removal of the MBR appeared to show a better performance (lower effluent concentration) due to the filtration effect of the membranes, in terms of biological removal the CAS system seems to have better performance. The FISH analysis results confirmed this observation because the samples of MBR system on average indicated that only 50% of total DAPI (total DNA) were eubacteria (EUB probe) compared with 80% in the CAS system (Maharaj et al., 2007);
3. Although the nominal pore size of the Kubota® membranes used in this study were 0.4 μm, the considerably lower MBR effluent COD (41 mg COD/ℓ) than the 0.45 μm filtered COD (69 mg COD/ℓ) indicates that a dynamic gel layer forms on the membrane which reduces their effective pore size.

The differences between the MBR and CAS effluent COD can be accommodated in the steady-state and dynamic models as differences in the soluble unbiodegradable COD fraction (f<sub>S'us</sub>) which were 0.045 and 0.066 respectively for the two systems. As expected, the MBR system produced a solids free effluent, whereas from the CAS system there was a continual loss of solids with the effluent (19.3 mg TSS/ℓ), quantified as the difference between filtered and unfiltered effluent COD concentrations (Table 1). This confirms that the MBR system effluent quality is independent of the
floc characteristics of the mixed liquor. The MBR system’s sludge settleability did not exceed 120 mℓ/g but that of the CAS system varied between 80 and 180 mℓ/g over the investigation.

**E.3.2.2 Pathogen (faecal coliform) removal**

The microbiological quality (faecal coliforms) of the MBR system effluent was superior. The results indicated pathogen counts were not detectable for the MBR system whereas for the CAS system, the average pathogen count was 2250 CFU/100 ml (Table 1). While these absolute CFU concentrations are not reflective of effluents from full-scale WWTPs, from the relative removals by the MBR and CAS systems, the membrane effluent has a higher quality for reuse purposes.

**E.3.2.3 Trans-membrane pressure (TMP)**

Due to the constant flow and load conditions, the membrane flux was constant and the TMP was controlled to maintain the flux [0.24 m³/(m².h)]. Because the volume of the MBR system aerobic reactor needed to be constant to not change its VSS concentration, the external U-bend over which the effluent exited the aerobic reactor was lowered below the reactor ML level to increase the TMP as required. From day 50 to 350, the TMP increased from around 20 to 80 mm (0.2 mm/d). While the membranes were rinsed under running water ever month, no chemical cleans were applied. From day 350 to 430 the TMP increased to 200 mm (1.5 mm/d) caused by inorganic colloidal solids from construction work at the WWTP, which also increased the influent and reactor ISS concentrations. A chemical clean on day 433 restored the TMP to its earlier low value of 55 mm.

**E.3.2.4 N and COD mass balances**

Nitrogen mass balances were calculated from the wastewater batch average results by comparing the exiting N via the effluent, waste sludge stream and nitrate denitrified (from a nitrate balance over the anoxic and anaerobic reactors) with the N entering the systems via the influent TKN. N balances for the MBR system ranged between 75 and 120% with an overall average of 96%. Similarly, the COD balance was calculated by comparing the exiting COD via the effluent, waste sludge stream and oxygen utilised in the aerobic reactor (corrected for nitrification) with the COD entering the systems via the influent COD. COD balances for the MBR system ranged between 89 and 107% with an overall average of 103%. These mass balances are good and validate that the measured results are reliable. In fact, the COD balances over the MBR system are among the best achieved in NDBEPR systems at UCT (Ekama and Wentzel, 1999), possibly as a result of the membranes.

**E.3.2.5 Biological nitrogen removal**

The TKN removal efficiency of the MBR system was marginally better than that of the CAS system (Table E.1). This is again attributed to the retention of solids by the membranes compared with that lost with the effluent from the CAS system. In both systems, nitrification was virtually complete, as indicated by the low residual free and saline ammonia (FSA) concentrations. The two systems achieved similar total N removals indicated by the similar effluent nitrate concentrations (Table 1).

**E.3.2.6 Biological phosphorus removal**

In both systems, TP was dosed in excess of the system removal capacity in order to observe the maximum BEPR possible. Thus P removal performance is represented by total P (TP) removals. System average P removals of 22.5 mg P/l and 17.4 mg P/l were achieved in the MBR and CAS systems showing that the P removal performance of the CAS was inferior to that of the MBR system. Reasons for this are: (1) Anoxic P uptake was higher in the CAS system at 22% of total P uptake in the anoxic reactor compared with the MBR system at 8%. This was also evident in the P removal anoxic batch tests results. With anoxic P uptake by denitrifying PAOs, significantly
reduced BEPR has been reported (Ekama and Wentzel, 1999; Hu et al., 2002) probably due to less efficient utilisation of stored polyhydroxyalkanoates (PHA) by PAOs when nitrate serves as electron acceptor (Hu et al., 2002, 2007a,b). (2) The nitrate load on the anoxic reactor of the MBR system was lower than the reactor’s denitrification potential identified by zero nitrate concentration in this reactor. In contrast, denitrification was not complete in the anoxic reactor of the CAS system, identified by >1 mg NO₃-N/l nitrate concentration in this reactor. This is one of the main factors that stimulates anoxic P uptake (Ekama and Wentzel, 1999; Hu et al., 2002). For the CAS system, the recycle of nitrate to the anaerobic reactor from the anoxic reactor reduced the RBO available for PAOs and additionally reduced the P removal. The main reason for the difference in denitrification capacity of the MBR and CAS anoxic reactors was the dissolved oxygen (DO) in the recycles, which had a negligible effect on denitrification at the very high VSS concentration in the MBR system.

The observations above indicate that the nitrate load in the CAS should have been reduced by reducing the aerobic to anoxic a-recycle ratio to maximise BEPR. While reducing the a-recycle ratio does not change the anoxic mass fraction and denitrification potential, the N removal performance will decrease, i.e. more nitrate out with the effluent rather than recycled to the anaerobic reactor via the anoxic reactor. In contrast, in the MBR system, from the design procedures (see 3.1 above), reducing the a-recycle ratio causes a corresponding reduction in the anoxic mass fraction and nitrate load and hence produces a greater decrease in N removal performance. This highlights the conflicting requirements between N and P removal in BNRAS systems at elevated influent TKN/COD ratios, which is by no means new but operates differently in the MBR BNR systems than in the CAS BNR systems. The balance between optimising N and P removal will be influenced by a number of factors, such as influent TKN/COD ratio, N and P removal requirements and influent RBO fraction.

At present anoxic P uptake BEPR is not explicitly incorporated in the steady-state design procedures for NDBEPR systems (Wentzel et al., 1990, Henze et al. 2008), as quantitative relationships linking the extent of anoxic P uptake to the system design and operating parameters have not been definitively established. Several dynamic models attempt to include it (ASM2d, Henze et al., 1999; UCTPHO+, Hu et al., 2007a,b) with varied success because the stimuli that promote it and the changes in kinetic behaviour it causes are not well defined yet.

E.3.2.7 System stability

With regard to COD, N and P removal over the whole investigation, the MBR system showed a greater stability than the CAS system – the MBR system performance recovered more quickly after power failures or pump stoppages and was generally less sensitive to disruptions such as sludge withdrawal for batch tests and monthly system cleaning (draining all the sludge from the reactors for about 2-3 h).

E.3.2.8 Sludge production

The MBR UCT system had a greater sludge production per COD load (0.31 kg VSS wasted/kg COD load) than the CAS system (0.20 kg VSS wasted /kg COD load). This increased sludge production can be explained only in part by (1) the retention of solids that would normally escape via the effluent from CAS systems with SSTs as TSS and unbiodegradable organics that would normally would be considered “soluble” (<0.45 µm), and (2) a higher P removal resulting in a larger PAO population which produces more VSS and ISS per unit influent COD than OHOs due to their lower endogenous respiration rate (Wentzel et al., 1990) and higher ISS content (Ekama and Wentzel, 2004). In the literature, previous studies comparing CAS and MBR BNR systems run under the same operating conditions (sludge age, influent wastewater) have indicated similar results.
as those outlined above (Cicek et al., 1999; Smith et al., 2002; Holbrook et al., 2005). More data at different sludge ages are required to determine if this increase is consistent (predictable) and to identify the underlying causes.

E.3.3 Aim 2: Compare the kinetics of biological N and P removal under MBR conditions (high reactor VSS concentration (13 g VSS/ℓ)) with those in conventional BNR systems (low reactor VSS concentration (3 g VSS/ℓ)).

The influence of high VSS concentration (up to 13 g VSS/ℓ) in the MBR system on the nitrification, denitrification and phosphorus release and uptake kinetic rates was measured with aerobic, anoxic and anaerobic batch tests on mixed liquor (ML) harvested from the MBR system, diluted to different VSS concentrations, and from the CAS system. Also, the limitation of ammonia, oxygen, nitrate and acetate on the kinetic rates was investigated with batch tests.

E.3.3.1 Calculation of kinetics rates from batch tests results.

In order to assign the observed bioprocess kinetic rates to the organism group mediating it, the measured VSS concentration in the batch tests was fractionated into ordinary heterotrophic (OHO), phosphorus accumulating (PAO) and autotrophic nitrifier (ANO) organism active concentrations by applying the steady state BNR AS model to the measured performances of the MBR and CAS systems. To account for the higher sludge production in the MBR system, the unbiodegradable particulate (f_{S'up}) and soluble (f_{S'us}) COD fractions are higher and lower respectively for the MBR system (f_{S'up}=0.241, f_{S'us}=0.045) compared with those of the CAS system (f_{S'up}= 0.084, f_{S'us} = 0.066). The higher f_{S'up} fraction affected the OHOVSS and PAOVSS active fractions and hence also the biomass specific kinetic rates. However, this affect was unavoidable because VSS specific kinetic rates are not comparable between different BNR systems and steady state models aligned with and based on the same but simplified principles as kinetic models are the interface between experimental systems and the kinetic models. This made the kinetic rates associated with BEPR measured in the batch tests incomparable between the two systems and so the rates were compared with literature values.

E.3.3.2 Nitrification

From the aerobic nitrification batch tests: (1) At the same low VSS concentration, the MBR system exhibited lower VSS specific ammonia utilisation rate (SAUR) and autotrophic nitrifier organism (ANO) maximum specific growth rates (μ_A) than the parallel CAS system, apparently due to different selection pressures imposed by membranes and SSTs. (2) For the MBR system, as the VSS concentration increased, the SAUR and μ_A decreased, apparently due to ammonia and/or oxygen transfer limitations. (3) For the MBR system at the VSS concentration, as the initial ammonia concentration increased, the SAUR and μ_A increased, indicating possible ammonia transport limitation at increasing VSS concentration.

From the above, it was evident that the ANOs in the MBR and CAS systems exhibited different behaviour, apparently induced by different environments under which the ANOs develop. The reasons for this possibly are: (1) In CAS systems with SSTs, organism loss via the effluent occurs including ANOs. Therefore CAS system may select ANOs with higher maximum specific growth rates (μ_A) than MBR systems. In the MBR system all the ANOs are retained, including slow growing ones. (2) At the high VSS concentrations in the MBR system, oxygen and ammonia transport limitations decrease the observed SAUR and μ_A.
E.3.3.3 Denitrification

From the anoxic-aerobic batch tests, the OHOVSS specific denitrification rate by OHOs (K_{2OHO}) utilizing slowly biodegradable organics (SBO) obtained at different MBR system VSS concentrations (2.5-12 g VSS/ℓ) and different initial nitrate concentrations ranging from 30 to 90 mg N/ℓ showed no effect to initial nitrate concentration, in agreement with past work (Van Haandel et al., 1981; Clayton et al., 1991; Ekama and Wentzel, 1999) and no effect to VSS concentration. From all the anoxic batch tests, the average K_{2OHO} was 0.264 mg NO_3-N/(mg OHOVSS.d), which is very close to the average K_{2OHO} rate reported in the literature for conventional (low VSS) BNR systems with SSTs, i.e. 0.255 from Ekama and Wentzel (1999).

E.3.3.4 Biological P removal

From the anaerobic-anoxic-aerobic batch tests, the specific VSS and specific PAOVSS anaerobic acetate (as COD) uptake and P release rates showed no effect of VSS or initial acetate concentration. Also, the results obtained with different concentrations of acetate added showed the acetate uptake rate to be zero order with respect to acetate concentration, which is in agreement with literature studies (Wentzel et al., 1985, 1989). The P release to acetate uptake ratio also showed no effect with acetate dose and VSS concentration. The specific VSS and specific PAOVSS aerobic and anoxic P uptake rates also showed no effect of VSS concentration. The average PAOVSS specific anaerobic acetate uptake and P release rates and the aerobic P uptake rate obtained over the VSS concentration range were within the range of literature rates observed on enhanced PAO culture systems, confirming that within experimental variation, high VSS concentration does not affect the rates.

In the anaerobic-anoxic/aerobic batch tests with acetate uptake, the PAOs showed significantly higher anoxic P uptake and denitrification rates than in the MBR system itself, where high acetate and excess nitrate did not occur. In the former the PAOs denitrified 22% of the nitrate whereas in the MBR system only 11%. The OHOVSS specific denitrification rates were within the same 0.2 to 0.3 mg NO_3-N/(mg OHOVSS.d) range in all the batch with an anoxic phase. While the PAOVSS specific denitrification rate in the anaerobic-anoxic/aerobic batch tests was about half of the OHOVSS specific denitrification rate, in the MBR system, the PAOVSS specific denitrification rate was only 1/14^{th} of the OHOVSS specific denitrification rate because the conditions in the anaerobic-anoxic/aerobic batch tests (high acetate and nitrate) were not prevalent in continuous flow BNR systems fed real wastewater. The large reduction in P removal resulting from significant anoxic P uptake BEPR seems counter-productive for the very small PAO contribution to denitrification.

The results from this investigation show that the BNRAS steady state and kinetic models developed for low VSS concentration BNRAS systems with secondary settling tanks can be applied with reasonable confidence to predict the performance of high VSS concentration BNRAS systems with membranes, except for the maximum specific growth rate of the nitrifiers, which was observed to be significantly lower in the MBR system.

E.3.4 Aim 3: Identify a BNR plant where MBR technology can be applied at full scale and negotiate as a project team with owners of BNR plants to implement BNR at a demonstration WWTP site.

During this project there were several possibilities for pilot or full-scale MBR BNR system implementation by the City of Cape Town, viz. (1) retrofit/ refurbishing the pilot-scale plants at Mitchell’s Plain WWTP, (2) an upgrade and extension of the Melkbos WWTP and (3) a
“greenfields” implementation of a new 20 MP/d MBR plant at Zandvliet WWTP. At the completion of this project none of these possibilities were realised.

E.4 CONCLUSIONS

The results from this investigation show that biological N and P removal are not negatively affected by high suspended solids concentrations (up to 13 g VSS/l) typical of MBR activated sludge plants, except the maximum specific growth rate of nitrifiers, which decreases as VSS concentration increases. However, for BNR system design and operation, because the maximum specific growth rate of nitrifiers varies considerably from one wastewater to another, it is regarded a wastewater characteristics rather than a kinetic constant. Hence the BNRAS steady state and kinetic models developed for low VSS concentration BNRAS systems with secondary settling tanks can be applied with reasonable confidence to predict the performance of high VSS concentration BNRAS systems with membranes, except for the maximum specific growth rate of the nitrifiers, which was observed to be significantly lower in the MBR system. Provided the models are given the appropriate input information, they will predict the performance of MBR BNR systems at high VSS concentration equally reliably (or unreliably) as that of conventional BNR systems at low VSS concentration with secondary settling tanks for solid liquid separation.
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LIST OF SYMBOLS AND ABBREVIATIONS

ADWF  Average dry weather flow
AS  Activated sludge
ASM  Activated sludge model
BEPR  Biological excess phosphorus removal
BNR  Biological nutrient removal
CASS  Conventional activated sludge system
COD  Chemical oxygen demand
Dpp  Total system denitrification potential (mg N/l)
DSVI  Diluted sludge volume index
f_{bs's}  RBCOD fraction with respect to total influent wastewater
f_{CV}  Mixed liquor COD/VSS ratio
f_{i}  VSS to TSS ratio of the mixed liquor
f_{N}  Mixed liquor TKN/VSS ratio
f_{q}  Peak wet weather to average dry weather flow
f_{S'up}  Influent unbiodegradable particulate COD fraction
f_{S'us}  Influent unbiodegradable soluble COD fraction
FSA  Free and saline ammonia
HRT  Hydraulic retention time
IAWPRC  International Association for Water Pollution Research and Control
IAWQ  International Association on Water Quality (formerly IAWPRC)
ISS  Inorganic settleable solids
IWA  International Water Association (formerly IAWQ, IAWPRC)
JHB  Johannesburg
ℓ  Litres
MBR  Membrane bioreactor
mg  Milligram
MF  Micro-filtration
MLSS  Mixed liquor suspended solids
mm  Millimetre
N  Nitrogen
ND  Nitrogen-denitrification
NDBEPR  Nitrogen-denitrification-biological excess phosphorus removal
NO_{2}  Nitrite
NO_{3}  Nitrate
NO_{x}  Nitrite and nitrate
OHO  Ordinary heterotroph organism
OSR  Oxygen supply rate
OTR  Oxygen transfer rate
OUR  Oxygen utilisation rate
P  Phosphorus
PAO  Polyphosphate accumulating organism
PST  Primary settling tank
PWWF  Peak wet weather flow
Q_{i}  Influent flow rate
RBCOD  Readily biodegradable COD
SCFA  Short-chain fatty acids
SQW  Square-wave (fed activated sludge system)
SRT  Solids retention time
SS  Sewage sludge
<table>
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<tr>
<th>Symbol</th>
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<tr>
<td>SSD</td>
<td>Sample standard deviation</td>
<td></td>
</tr>
<tr>
<td>SST</td>
<td>Secondary settling tank</td>
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<tr>
<td>$S_{\text{use}}$</td>
<td>Unbiodegradable soluble COD concentration of the system effluent</td>
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<tr>
<td>$S_{\text{ini}}$</td>
<td>Unbiodegradable soluble COD concentration in the influent wastewater</td>
<td></td>
</tr>
<tr>
<td>TKN</td>
<td>Total Kjeldahl Nitrogen</td>
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<tr>
<td>TMP</td>
<td>Trans-membrane pressure</td>
<td></td>
</tr>
<tr>
<td>TSS</td>
<td>Total suspended solids</td>
<td></td>
</tr>
<tr>
<td>UCT</td>
<td>University of Cape Town</td>
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<tr>
<td>UF</td>
<td>Ultra-filtration</td>
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</tr>
<tr>
<td>VSS</td>
<td>Volatile suspended solids</td>
<td></td>
</tr>
<tr>
<td>WAS</td>
<td>Waste activated sludge</td>
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</tr>
<tr>
<td>WRC</td>
<td>Water Research Commission</td>
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**Note:** Only symbols and abbreviations used in the text are included, those in equations are defined below in the appropriate equations.
PUBLICATIONS GENERATED DURING WRC PROJECT K5/1537

Refereed journal papers


International conference papers


Local conference papers


Reports/Theses


CAPACITY BUILDING

The following BSc (equivalently Hons), MSc and PhD students completed thesis research projects within this WRC project.

Janson, Christina (BSc, 2003) Biological kinetics in membrane bioreactor nutrient removal systems.
Du Toit, Geoff JG (MSc, 2007) Design and performance of BNR activated sludge systems with flat sheet membranes for solid liquid separation (W129).
CHAPTER 1
INTRODUCTION

1.1 MOTIVATION

Effective solids-liquid separation in suspended medium biological wastewater treatment systems is an essential step, because it has a major influence on effluent quality. Traditionally in biological wastewater treatment systems this step has been accomplished in secondary settling tanks (SSTs). However, membranes appear to be an attractive alternative: “Due to plummeting costs and dramatically improving performance, water-treatment applications based membranes are blossoming. Today, membrane processes are robust, simple to operate, and ever more affordable. They take up little space, need modest technical support, and can remove many contaminants in one step. These advantages make it practical, for the first time, to protect public health and safely reuse water for non-potable uses.” (Bellagio Statement, 2003).

Membrane solid-liquid separation in place of sedimentation in secondary settling tanks (SSTs) offers several advantages for the activated sludge (AS) system:

1. Insensitivity to sludge settleability and filamentous bulking; this is a significant advantage as biological nutrient removal (BNR) systems notoriously produce rather poor settling sludges (DSVI~150 mℓ/g) when aerobic mass fractions are low (<60%)
2. Insensitivity to activated sludge flocculation characteristics and hydraulic shear in the reactor; membranes retain all solids > 0.1 μm, which includes free swimming bacteria.
3. SSTs are not required, resulting in a wastewater treatment plant footprint reduction.
4. Very high reactor solids concentrations 15 to 20 g TSS/ℓ (1.5 to 2%) resulting in reduced biological reactor volumes compared with conventional BNR systems with SSTs (a further footprint reduction).
5. Production of possibly disinfected effluent for industrial or horticultural use; membranes appear to be an effective barrier against bacteria and viruses (Churchouse and Brindle, 2003), indicating effective pore sizes considerably smaller than the membrane nominal pore size.
6. Possibly obviate waste activated sludge thickening when reactor concentrations are at the high end of the range (2% TSS).

Membrane activated sludge systems have been operated successfully both technically and economically in the UK, Europe and Japan. However, application has been largely restricted to COD (BOD₅) and free and saline ammonia (FSA) removal (nitrification), i.e. mostly as fully aerobic systems. Membrane bioreactor (MBR) technology application to BNR activated sludge systems is limited, and uncertainty exists as to the impact of the conditions induced by the MBRS on the biologically mediated processes of nutrient removal. Two obvious conditions associated with MBRS are:

1. High total suspended solids (TSS) concentrations (12 to 19 g/ℓ).
2. Different selective pressures; the membranes retain organisms that would normally escape with the effluent in conventional systems.

Accordingly, in April 2003 a one year consultancy (K8/514) was set up between the University of Cape Town (UCT), the Water Research Commission (WRC) and Aquator, to investigate BNR behaviour in MBR systems. The primary purpose of this project was to assess whether or not a longer term more in-depth investigation into the kinetics of biological excess phosphorus removal (BEPR) and nitrification denitrification (ND) is required when BNR is operated within an MBR.

The preliminary investigation under the consultancy indicated that a more in depth investigation was warranted. Accordingly, in April 2004 a two year research contract (K5/1537) to March 2006
was set up between UCT and the WRC to continue the research. This report presents the consolidated results from both the consultancy and the contract.

1.2 OBJECTIVES AND AIMS OF RESEARCH

The main objective of both the research consultancy and contract was to (1) evaluate biological nutrient removal (BNR) performance at typical membrane bioreactor (MBR) total suspended solids (TSS) concentrations (14-18 g/l). For the research contract additional objectives were identified, viz. (2) Compare the performance and kinetics of biological N and P removal under MBR conditions (high reactor TSS concentration (16 g/l) with those in conventional BNR systems (low reactor TSS concentration, 4 g/l), (3) Identify a BNR plant where MBR technology can be applied at full scale and negotiate as a project team with owners of BNR plants to implement BNR at a demonstration WWTP site.

During the course of the preliminary research consultancy it became evident that installing membranes for solids/liquid separation into BNR activated sludge systems makes a profound difference not only to the design of the BNR system itself, but also to the design approach for the whole wastewater treatment plant. Accordingly, a second main objective for the project was identified which was not explicitly included in either project proposal, viz. (4) evaluate the impact of membrane solid liquid separation on the design of biological nutrient removal activated sludge systems.

1.3 RESEARCH APPROACH

The research approach adopted was to operate two parallel laboratory-scale BNR activated sludge systems in the constant temperature Water Research Laboratory (20°C), one MBR system and one conventional system. Both BNR systems were 3 reactor anaerobic – anoxic – aerobic UCT configurations, with 3 recycle flows, called “a” – mixed liquor from aerobic to anoxic, “s” – sludge return from SST to anoxic and “r” – mixed liquor from anoxic to aerobic. The “a” and “s” recycles for the MBR system were combined as one recycle “as”. The UCT configuration was selected because it allows the biological excess phosphate removal (BEPR) to operate independently of the nitrogen removal, i.e. zero nitrate recycle to the anaerobic reactor, provided the recycle(s) to anoxic reactor do not overload this reactor with nitrate.

The two systems were set up to have identical design parameters such as anaerobic, anoxic and aerobic mass fractions, recycles and sludge age. The only differences would be the influent flow and total reactor volume (i.e. hydraulic retention time, HRT), due to the physical constraints imposed by the dimensions of the membranes, and the much higher reactor concentrations in the MBR system.

The performance of the two systems was to be monitored, evaluated and compared to identify and quantify the influence of the membranes on system response. The kinetics of biological N and P removal processes would be determined in anaerobic, anoxic and aerobic batch tests on sludge harvested from the different reactors of the two systems, to measure the nitrification, denitrification, P release and P uptake rates and compare these between the two systems.

1.4 SPECIFIC PROJECT RESEARCH TASKS

To address the four objectives above, a number of specific research tasks were identified:

Task 1: Evaluate the impact of the membranes on the system design. A theoretical study will be undertaken to quantify the impact of membrane solids liquid separation on the design of BNR
activated sludge systems, to evaluate the advantages and disadvantages of membrane BNR systems. From this study design procedures and criteria for membrane BNR activated sludge systems will be developed. The outcomes of this research task is reported in detail by Ramphao et al. (2004) and summarised in Chapter 3.

**Task 2:** Operate under controlled laboratory conditions two NDBEPR UCT configuration BNR activated sludge systems, one a conventional system with SST and the other with MBR. All the engineering parameters will be measured on the systems and with these the usual consistency checks for data reliability such as COD and N mass balances will be made. The outcomes of this research task are reported in detail by Ramphao et al. (2004) and du Toit et al. (2006) and are summarised in Chapters 4 and 5.

**Task 3:** Evaluate and compare the performance of the laboratory-scale systems. From the measurements on the two parallel systems, the COD, N and P removal behaviour and performance of the two systems will be evaluated directly and by comparison with predictions of the models for BNR activated sludge systems. Evaluation of system performances for the first year of operation (June 2003 to July 2004) is reported in detail by Ramphao et al. (2004) and summarised in Chapter 4. System performance evaluation for the second year operation (July 2004-Aug 2006) is summarised in Chapter 5.

**Task 4:** Impact of membranes and high TSS (15 g TSS/ℓ) concentration on the N and P removal bio-process kinetics of nitrification, denitrification, P release and P uptake. This was done with aerobic, anoxic and anaerobic batch tests on sludge harvested from the two BNR systems. The outcomes of this research task are reported in detail by Parco (2006) and are summarised in Chapter 6.

**Task 5:** Identify a plant for full scale implementation. During this project there were several possibilities for pilot of full scale MBR BNR system implementation by the City of Cape Town, viz. (1) retrofit/ refurbishing the pilot-scale plants at Mitchell Plain WWTP, (2) an upgrade and extension of the Melkbos WWTP and (3) a “greenfields” implementation of a new 20 MP/d MBR plant at Zandvliet WWTP. At the completion of this project none of these possibilities were realised.
1.5 PROJECT REPORTS

This final report summarises both the consultancy (K8/514) and contract (K5/1537) research outcomes. During the course of the project, the WRG produced three reports which describe the research investigation in more detail, viz.


The first two of these reports are available from the Dept of Civil Engineering, University of Cape Town, Private Bag, Rondebosch, 7701, Cape, South Africa.

1.6 REFERENCES

BELLAGIO STATEMENT (2003) A statement emerging from the IWA specialised conference on membranes, April 2003, Bellagio, Italy.


CHAPTER 2 LITERATURE REVIEW

2.1 INTRODUCTION

Membrane technology – typically thought of in terms of water treatment – has begun to emerge as one of the most significant advances in wastewater treatment in the past 20 years, and its application is expected to become widespread in the future. Early use of membranes for treatment of wastewater appeared more than 25 years ago, but over the past 5 years there has been a rapid increase in the volume of wastewater that is treated with membranes, typically for reuse purposes. In fact, in 2000 there were about 300 full-scale municipal wastewater treatment facilities using membrane technologies, about 150 of them in the North America (Thompson et al., 2000), and this number is increasing rapidly.

The potential to use membranes exists wherever they provide the ability to remove contaminants that cannot be removed by other technologies, remove contaminants at lower cost than other alternatives, or require less land area than competing technologies. For wastewater treatment applications, membranes are currently being used for the tertiary removal of dissolved salts, organic compounds, phosphorus, colloidal and suspended solids, and human pathogens, including bacteria, protozoan cysts, and viruses, or as the solid-liquid separation step in activated sludge type systems. Thus, membrane technologies for wastewater treatment include:

- High-pressure membranes – nano-filtration or reverse osmosis pressure systems for treatment of wastewater plant effluents for the production of high-quality product water suitable for indirect potable re-use and high-purity industrial process water.

- Low-pressure membranes – usually microfiltration (MF) or ultrafiltration (UF) membranes, either as a pressure system or an immersed system, providing a higher degree of solids removal following secondary clarification (Fig. 2.1b).

- Membrane bioreactors – usually MF or UF membranes immersed in aeration tanks, or implemented in external pressure-driven membrane units, as a replacement for secondary clarifiers (Fig. 2.1c).

This research project focuses on membrane bioreactors (MBR). In this chapter, applications of MBR for wastewater treatment will be reviewed, to highlight potential advantages or disadvantages associated with these systems.
2.2 MEMBRANE BIOREACTOR

The membrane bioreactor (MBR) technology derives from coupling of a conventional suspended growth process with micro-filtration or ultra-filtration, i.e. the MBR process is a variation of the conventional activated sludge process (Cicek et al., 1998a). In an MBR process, micro-filtration or ultra-filtration membranes are used instead of secondary settling tanks to separate mixed liquor suspended solids (MLSS) from the treated effluent. Only the treated effluent passes through the membrane, and the sludge is returned to or retained in the biological reactor. The size of the membrane pores (typically 0.1-0.5 µm) essentially results in complete removal of the total suspended solids (TSS) and significantly traps a high proportion of the pathogenic organisms (Cicek et al., 1998b).

2.2.1 MBR System Configurations

MBRs are composed of two primary parts; the biological suspended growth (activated sludge) unit which is responsible for the biological degradation of the waste compounds, and the membrane
module for the physical separation of the treated water from the mixed liquor. MBR systems can be classified into two major groups according to their configuration. The first group, which is commonly known as integrated or submerged MBR, involves the use of outer skin membranes that are submerged into the biological reactor (Fig. 2.2). The driving force across the membranes is achieved by pressurizing the membranes on the bioreactor side or by creating a negative pressure on the permeate side of the membranes (Buisson et al., 1998; Cote et al., 1997; Rosenberger et al., 2002).

Cleaning of the membranes is achieved through frequent permeate back-pulsing and/or occasional chemical backwashing. Diffusers are usually placed directly beneath the membrane modules to facilitate air and liquid scouring across the membrane surface to reduce fouling. Aeration and mixing are also achieved with the same diffusers.

The second configuration is the recirculated (external) MBR, which involves the recirculation of the mixed liquor through a membrane module that is located outside the bioreactor. Both inner-skin and outer-skin membranes can be used in this application. The driving force is the pressure from the pump that circulates mixed liquor through the membrane module. This pressure also acts perpendicular to the axis of flow and forces the water through the membrane surface. A schematic of the external MBR is shown in Fig. 2.3.
The choice between operating options is dependent upon the specific implementation, as both systems have advantages and disadvantages as shown in Table 2.1.

<table>
<thead>
<tr>
<th>Submerged MBR</th>
<th>External (Side-stream) MBR</th>
</tr>
</thead>
<tbody>
<tr>
<td>Aeration cost is high (~90%)</td>
<td>Aeration cost is low (~20%)</td>
</tr>
<tr>
<td>Very low pumping costs (higher if suction pump is used ~28%)</td>
<td>High pumping costs (60-80%)</td>
</tr>
<tr>
<td>Lower flux (larger footprint)</td>
<td>Higher flux (smaller footprint)</td>
</tr>
<tr>
<td>Less frequent cleaning is required</td>
<td>More frequent cleaning is required</td>
</tr>
<tr>
<td>Lower operating costs</td>
<td>Higher operating costs</td>
</tr>
<tr>
<td>Higher capital costs</td>
<td>Lower capital costs</td>
</tr>
</tbody>
</table>

### 2.2.2 Membrane Types

Various types of membranes have been used for MBR applications. These include tubular, hollow fibre, sheet (plate) and inorganic (ceramic) micro-filtration or ultra-filtration membranes. The pore sizes of the membranes (especially micro-filtration or ultra-filtration membranes) typical range from 0.1-0.5 μm. The flux obtained in these types of membranes ranges from 0.5 to 1.2 m³/(m².d) and 1 to 4 m³/(m².d) for the submerged and external membranes respectively, at 20°C. The applied trans-membrane pressure ranges from 100 mm to 3000 mm for the submerged membranes (depending on the type), whilst in the case of external membranes, the pressure ranges from –75 mm to –3000 mm.

The membranes used in MBR systems must satisfy several criteria, i.e. they must be inert and unbiodegradable. The membranes should be easy to clean and should be resistant to the cleaning agents, high temperatures and pressures. Uniform pore distribution and high porosity are desired characteristics for membranes.
2.2.3 MBR Operational Considerations

One of the biggest technical challenges with the use of membranes for wastewater treatment is the high fouling that occurs universally. Membrane fouling appears to be mainly due to colloids, soluble organic compounds, and bacteria. Membrane fouling results in an increase in pressure to maintain the required flux and necessitates frequent cleaning of membranes. This leads to a reduction in overall facility efficiency and a shorter membrane life.

To reduce the fouling, in the MBR system aeration within the bioreactor not only provides the required (part or full) oxygen transfer for the growth of the biomass and turbulence for mixing of the reactor, but in the case of the submerged MBR configuration (Fig. 2.2) the rising bubbles also provide a turbulent cross flow velocity (approximately 0.5 m/s) over the surface of the membranes. This provides scour across the surface of the membranes which helps to maintain the flux through the membranes, by reducing the build up of material at the membrane surface. This scours increases the operational cycle time of the system. In this type of system, aeration usually is via coarse bubble aeration to promote scour efficiency. With scour induced by aeration, less frequent and less rigorous cleaning of the membranes is required to restore the operational flux compared with the side-stream system.

In the side-stream system, the aeration is usually through fine bubble diffusers, which offer much more efficient oxygen transfer than the coarse bubble diffusers. The cross flow velocity utilised in these systems usually is higher (2-4 m/s) and is provided by circulation of the mixed liquor across the membrane surface through pumping. As a higher differential head usually drives this system, the operational flux of the system is higher. The disadvantage of this is that fouling is more pronounced and much more rigorous cleaning regimes are required to restore the operational flux. The useful life span of the membranes may be reduced by such an operating regime.

In both types of systems, to increase the scour effectiveness of the air, the solids (MLSS) concentration has to be very high, that is, 8 000-20 000 mg TSS/ℓ, which, on the one hand is advantageous for the activated sludge system because it reduces the bioreactor volume, but on the other hand is disadvantageous because it reduces the oxygen transfer rate (kg O/kWh).

2.3 TYPES AND SUPPLIERS OF MBR SYSTEMS

The two main suppliers of MBR systems for wastewater treatment are Kubota® (Japan) and Zenon® (USA). Other suppliers are Degremont® (France), US Filter® (USA) and MembraneTek® (RSA). In the rapidly changing market, by the time this report is published, these trade names may have changed.

2.3.1 Kubota® Membranes

Kubota® uses flat sheet membranes made of chlorinated polyethylene with a non-woven cloth base giving a nominal pore size of 0.1-0.5 μm. Each membrane cartridge consists of solid acrylonitrile butadiene styrene (ABS) support plate with a spacer layer between it and an ultrasonically welded flat sheet membrane on both sides. The typical membrane cartridge (Type 510) has dimensions of 1 000 mm (Height) x 490 mm (Width) x 6 mm (Thick). Filtered water (effluent) passes through the membrane into the interior of each panel to an outlet nipple cast into the top of the support plate. Each cartridge provides an effective filtration area of 0.8 m². Fig. 2.4 shows the Type 510 Kubota® membrane.
The Kubota® MBR operates with membrane panel units submerged in the bioreactor in which the MLSS is maintained within the range of 12 000 to 20 000 mg TSS/l. The Kubota® submerged membrane units are manufactured from stainless steel and can be installed in either single or double deck configurations. The single deck can house up to 200 membrane cartridges, while the double deck units can house up to 400 cartridges. The gap between each cartridge is approximately 7 mm. Each panel in the cartridge is connected to a filtered effluent manifold.

The single deck submerged membrane unit (Figure 2.5a) comprises a diffuser case and a membrane case. The double deck submerged membrane unit comprises a diffuser case, a lower membrane case, an intermediate case and an upper membrane case. In a double deck membrane unit (Figure 2.5b), the lower membrane case and the intermediate case are bolted together. The diffuser case contains the coarse bubble diffuser. This section supports the membrane case and directs the mixture of air bubbles and mixed liquor between the membrane panels. The air-water mixture maintains an upward cross-flow velocity over the membrane surface greater than 0.5 m/s, minimising fouling of the membranes. The minimum air requirement is 7.5Nm³/(m² membrane surface.h).

The Kubota® system operates by gravity, with a mixed liquor (water) head (TMP) of 100-200 mm above the membranes sufficient to drive permeate (effluent) through the membranes. Wastewater fine screening (3 mm 2 dimensional screens) is a pre-requisite and grit removal recommended prior to the MBR treatment. The average membrane flux for the Kubota® system is approximately 0.5 m³/(m².d) (submerged system at a TMP of ~ 150 mm) and a temporary maximum of 1.0 m³/(m².d) can be accommodated for a few hours.

In municipal applications, chemical cleaning of the membranes is typically required every 6 months using sodium hypochlorite (0.5% w/w). Cleaning requires 3 ℓ of chemical solution per panel and the cleaning cycle takes up to 2 hours.

Kubota® has a reference list of over 1 000 plants (September 2003) treating domestic and industrial wastewater, with most of the sites located in Japan. The Kubota® plants range in size from systems to treat the equivalent of individual households to the 5 800 m³/d ADWF plant at Swanage in the South of England. Recommended biological design parameters are listed in Table 2.2.
Table 2.2: Recommended design parameters for MBR with Kubota® membranes (Judd, 2002).

<table>
<thead>
<tr>
<th>PARAMETER</th>
<th>VALUE</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow through membrane</td>
<td>In-to-out</td>
</tr>
<tr>
<td>Pore size</td>
<td>0.1 -0.5 microns</td>
</tr>
<tr>
<td>MLSS</td>
<td>12 000-20 000 mg/L</td>
</tr>
<tr>
<td>Trans-membrane pressure (TMP)</td>
<td>100 mm to 200 mm</td>
</tr>
<tr>
<td>Operational flux</td>
<td>between 20 (average)-27 (peak) L/m².h</td>
</tr>
<tr>
<td>Mean aeration demand</td>
<td>0.75 Nm³/hr/m² membrane</td>
</tr>
<tr>
<td>Mean aeration demand</td>
<td>32 Nm³ air/hr/m³ permeate product</td>
</tr>
<tr>
<td>Mean permeability</td>
<td>190 l/m²/hr/TMP</td>
</tr>
<tr>
<td>Clean in place, frequency</td>
<td>8 mths for 2 hrs with 1.0% hypochlorite</td>
</tr>
<tr>
<td>HRT</td>
<td>greater than 2 hours</td>
</tr>
<tr>
<td>SRT</td>
<td>greater than 15 days</td>
</tr>
</tbody>
</table>

**Figure 2.5a: Kubota® single deck membrane unit.**
Figure 2.5b: Kubota® double deck membrane unit.

Figure 2.6 illustrates a laboratory-scale immersed MBR, with A4 size Kubota® membranes inserted directly in the aerobic bioreactor. A typical full-scale plant will have as main equipment, the membranes installed within the membrane cases within the reactor, a fine screen, a bioreactor, blowers for the aeration process and equipment for easy membrane cleaning and washing.

Figure 2.6: Typical aerobic reactor of a laboratory-scale Kubota® MBR.
2.3.2 Zenon® Membranes

Zenon® markets the ZenoGem® system, based on the ZeeWeed® membrane. This is a hollow fibre tube with an external diameter of 1.9 mm and consists of an inner support layer covered by the outer membrane filtration layer. The membrane has a nominal pore size of 0.1-0.5 μm. The fibres are mounted on vertical frames into modules with filtered effluent passing into the centre of the fibre and extracted from both ends. The ZW-500 module is 2000 mm (H) x 700 mm (W) x 200 mm thick with 46 m² of filtration surface area. The ZeeWeed® hollow fibre membranes are contained in bundles called modules, which are assembled into cassettes of 8 to 48 modules. The membrane modules are immersed directly in the aeration reactor, in direct contact with the mixed liquor.

The ZeeWeed® MBR process is a proprietary ZENON technology that consists of a suspended-growth biological reactor integrated with ZeeWeed® ultra-filtration membranes. Membranes replace the solids separation function of conventional SSTs and sand filters. For wastewater applications, the ultra-filtration membrane's pore size ensures that no particulate matter greater than 0.1 μm is discharged with the effluent. The ZeeWeed® membranes are typically immersed directly in the aeration reactor in direct contact with the mixed liquor.

Through the use of a centrifugal pump, a vacuum pressure varying between -15 kPa and -65 kPa is applied to a header connecting the membrane modules. The vacuum draws the treated water through the hollow fibre membranes. All the particulate matter and the mixed liquor solids are rejected at the surface of the membrane and retained in the bioreactor. The filtration capacity is in the range of 1.0-1.7 m³/m²/d.

The ZeeWeed® membranes are automatically back pulsed on a regular basis using collected permeate (effluent). Air is supplied to the system by a combination of coarse bubble aerators integrated into the bottom header of the modules, to gently agitate the membrane fibres and to keep the reactor contents mixed, and by the fine bubble aeration to supply the balance of the total biological oxygen demand. The airflow provided by the diffuser scours the external surface of the membranes transferring the rejected solids away from the membrane surface. Sludge is wasted directly from the aeration tank in which the operating MLSS concentration is maintained between 8 000 mg TSS/ℓ to 12 000 mg TSS/ℓ. The recommended biological design parameters are listed in Table 2.3.

Table 2.3: Recommended design parameters for MBR with Zenon® membranes (Judd, 2002).

<table>
<thead>
<tr>
<th>PARAMETER</th>
<th>VALUE</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow through membrane</td>
<td>Out-to-in</td>
</tr>
<tr>
<td>Pore size</td>
<td>0.2 -0.5 microns</td>
</tr>
<tr>
<td>MLSS</td>
<td>8 000-12 000 mg/L</td>
</tr>
<tr>
<td>Trans-membrane pressure (TMP)</td>
<td>150 mm to 375 mm</td>
</tr>
<tr>
<td>Operational flux</td>
<td>between 27 (average)-40 (peak) L/m².h</td>
</tr>
<tr>
<td>Mean aeration demand</td>
<td>0.91 Nm³/hr/m² membrane</td>
</tr>
<tr>
<td>Mean aeration demand</td>
<td>27 Nm³ air /hr/m³ permeate product</td>
</tr>
<tr>
<td>Mean permeability</td>
<td>140 l/m²/hr/TMP</td>
</tr>
<tr>
<td>Back flush duration</td>
<td>~ 25% of permeate product in minutes</td>
</tr>
<tr>
<td>Clean in place, frequency</td>
<td>1-2 times per week with hypochlorite</td>
</tr>
<tr>
<td>HRT</td>
<td>greater than 2 hours</td>
</tr>
<tr>
<td>SRT</td>
<td>greater than 15 days</td>
</tr>
</tbody>
</table>
In addition to the scouring action of the coarse bubble aeration, cleaning of the membranes to control fouling is provided by automatic pulses of backwashing with stored permeate described above, and periodic in-situ membrane cleaning with a hypochlorite solution or other chemicals approved by the manufacturer. Zenon has a reference list of over 150 plants treating domestic and industrial wastewater (Cicek et al., 1998a).

2.3.3 Selection of Membrane Type for the Investigation

Kubota® and Zenon® membranes have their advantages and disadvantages. For this investigation the Kubota® membranes were selected due to their ease of operation in a laboratory system (no pumping). Their selection in this project is not an endorsement for their selection at full scale.

2.4 POTENTIAL ADVANTAGES OF MBRs OVER CONVENTIONAL ACTIVATED SLUDGE SYSTEMS

The membrane bioreactor (MBR Fig. 2.5c) process offers a number of advantages over the conventional system with SSTs, such as the absolute control of solids and hydraulic retention times, high treated water quality, retention of microorganisms and viruses, maintenance of high biomass concentration and compactness of system (Cicek et al., 1999a,b,c). These are described in more detail in this section.

In the MBR system, the traditional secondary settling tanks (SSTs) for the separation of treated water from the mixed solution in the bioreactor are replaced by a membrane unit (MF or UF). Hence, activated sludge and solids-liquid (membrane) separation are integrated into a single treatment step. This configuration of MBR is known as a submerged membrane bioreactor (sMBR). The sMBR technology would appear to offer a number of advantages over a conventional activated sludge system (CASS) with SSTs:

- The absence of SSTs means that the overall sMBR plant size (footprint) can be significantly reduced in comparison with that of a CASS. Also, the cost of the membranes can be offset by absence of costs for SSTs.

- In MBRs, the membranes replace the SSTs, providing solids separation by means of filtration instead of gravity settling. This means that the system is insensitive to settling characteristics of the sludge; even poorly settling sludge can be readily retained. This is a significant advantage, as some aerobic and biological nutrient removal (BNR) systems produce poor settling sludges.

- The membranes with pore sizes < 0.5 μm (Tables 2.1 and 2.2) would retain virtually all solids. This would cause the effluent quality to be insensitive to activated sludge flocculation characteristics and hydraulic shear in the reactor.

- Elimination of SSTs would eliminate problems with foaming or rising sludges, and hence the impact of these on effluent quality.

- Since in conventional systems solids-liquid separation is in SSTs, the settling velocity of the mixed liquor in the SST must exceed the upflow velocity. The settling velocity decreases as the solids concentration increases (WRC, 1984). Thus, as the solids concentration increases, the area of the SST must increase to reduce the upflow velocity, which results in increased cost for the SST. However, an increase in solids concentration causes a decrease in bioreactor volume, reducing the cost for this. The optimum cost in
wastewater treatment plants with SSTs usually is at a solids concentrations of about 3 to 5 g TSS/ℓ. Hence, activated sludge systems with SSTs usually are operated in this concentration range. In MBR systems, the SSTs no longer are required and hence this concentration range restriction no longer applies, but the concentration is limited by the effect of the increased concentration on liquid viscosity which influences oxygen transfer efficiency. In MBR systems, reactor concentrations usually are in the range 8-20 000 mg TSS/ℓ. Such high concentrations also improve the scour effectiveness across the membrane surface. These very high reactor concentrations can result in significantly reduced reactor volumes compared with the equivalent conventional activated sludge systems, a further footprint reduction.

- A possible consequence of the high concentrations above, is that waste activated sludge thickening prior to further treatment/disposal may not be required.

- The membranes, with a nominal pore size < 0.5 μm (Tables 2.1 and 2.2) and an effective pore size that is considerably lower, provide an effective barrier against bacteria and perhaps viruses (Cicek et al., 2001, 2002; Churchouse and Brindle, 2003). This results in the production of a high quality effluent by the MBR system, with reduced tertiary treatment and disinfection requirements for industrial or horticultural re-use. Also, subsequent treatment steps for water reclamation for potable purposes will be reduced.

- In a conventional biological system, performance and efficiency are limited by the ability of the SSTs to settle the solids from the mixed liquor stream. This is a function of operator skill, sludge settleability, basic SSTs design, solids management, the extent and the rate of variability in hydraulic or organic loading. In the case of an upset occurring in the system, the solids can be lost and the plant performance can be compromised. With complete retention of solids, MBR systems effluent are less prone to these system upsets.

2.5 APPLICATIONS OF MBRs

2.5.1 Applications in Municipal Wastewater Treatment

MBRs were initially used for municipal wastewater treatment, primarily in the area of recycling and water re-use. Compactness, production of re-usable water, and water that is free from biomass solids and contaminants made the MBR an ideal process for recycling municipal wastewater in space limited environments. Legislation in parts of the world, encouraging water reuse in large commercial buildings, triggered the development and application of alternative technologies. Thus, several types of MBR systems were implemented on a large scale basis and were made available commercially (Kimura, 1991; Yokomizo 1994).

Plate and frame ultra-filtration membranes, connected to an aerobic bioreactor, were used to treat wastewater originating from kitchens and bathrooms in Tokyo, Japan (Kimura, 1991; Minami, 1994). The treated water was then recycled and used for flushing of toilets. An effluent free of suspended solids and very low organic content (below 15 mg COD/ℓ) was obtained consistently. To maintain the flux of the membrane, the reactor was aerated continuously to provide the cross-flow velocity (scour) over the surface of the membranes to minimise fouling.

In Michigan, USA, an anoxic and aerobic biological treatment system coupled to an external tubular organic membrane was developed for the treatment of municipal wastewater and re-use of the effluent produced (Irwin, 1990). Activated carbon and ozone addition to the effluent (permeate) were carried out to remove odour and prevent biological activity by disinfection. Through
facilitating water re-use, this system was capable of reducing “new” water use by 70 to 90% and has since been applied in more than 30 locations in the United States (Irwin, 1990). A similar system, using ceramic tubular membranes was developed by Lyonnaise des Eaux in France. A pilot-scale aerobic MBR was used to treat municipal wastewater at the Aubergenville Wastewater Treatment Plant, nearby Paris. Steady operation was achieved and complete nitrification and over 93% of COD and suspended solids removal were accomplished (Fan et al., 1996).

The MBR system was also used in the treatment of human excreta in domestic wastewater. These applications, also known as night soil treatment systems, were typified by the high strength of the waste and the need for on-site treatment. These properties promoted the applications of MBR processes which became highly feasible under such conditions. The MBR system successfully replaced a rather complex set of treatment systems which incorporated denitrification, coagulation, filtration, and activated carbon treatment (Magara and Itoh, 1991).

Another application of the MBR has been in the area of sludge treatment. Conventionally, sludge stabilisation in wastewater treatment plants is achieved by a single pass, anaerobic digester. Since the HRT and the SRT are identical in these systems, the capacity is limited and long sludge ages are required for effective solids destruction. It has been proven that the addition of a micro-filtration unit will enhance the performance of the digester by uncoupling the HRT and the SRT and, thereby, allowing higher volumetric throughput (Ross et al., 1990). An economic evaluation of such an MBR process was performed at a wastewater treatment plant in Durban, South Africa. It has been shown theoretically that the MBR system could reduce both the capital and operational cost of a conventional anaerobic digester (Pillay et al., 1994). However, avoiding blockages in the membrane system may make the system impractical.

The applications above essentially coupled the implementation of MBRs with the re-use of the produced effluent or with site specific problems and conditions (e.g. limited space), probably due to economic conditions. As described above, implementation of the coupled schemes had been successful at a number of sites, and these will increase significantly in response to sustainable development pressures. More recently, the increasing affordability of modern membranes and their advantages, have resulted in implementation in the treatment step alone.

Wen et al. (2003) operated a fully aerobic system with MBR for treatment of municipal wastewater for 216 days. They observed a stable and excellent effluent quality in terms of COD (95% removal) and complete nitrification, but since the system was aerobic total nitrogen removal was poor (67%).

Gao et al. (2003) compared an aerobic reactor submerged MBR with a conventional aerobic activated sludge system operated in parallel over a period of 210 days, treating synthetic ammonia-bearing inorganic wastewater under similar conditions. In the MBR system, almost complete conversion of ammonia to nitrate was constantly achieved over an influent ammonia concentration range from 180 to 1 300 mg N/l at a hydraulic retention time of 24 h, compared with an average conversion of only 90% in the parallel conventional system.

2.5.2 Applications in Industrial Wastewater Treatment

High organic loadings and very explicit and difficult to treat compounds are two major characteristics of industrial waste streams that render alternative treatment techniques such as the MBR desirable. Traditionally, high COD content wastewaters were treated under anaerobic conditions, and hence initial MBR applications for industrial wastewater were in the field of anaerobic treatment which proved to be successful (Cicek et al., 1998a).
Another application of MBR systems in “industry” is in the area of landfill leachate treatment. Landfill leachates usually contain high concentrations of organic and inorganic compounds. Conventionally, the treatment of leachates involves a physical, biological or membrane filtration process (or a combination of them). MBR systems have been successfully utilised with an additional treatment step for inorganics and heavy metal removal, such as reverse osmosis. Several industrial scale plants, combining a MBR and reverse osmosis system, are presently being explored in France (Manem and Sanderson, 1996), to generate water of high quality for re-use.

2.6 APPLICATIONS OF MBR FOR BIOLOGICAL NUTRIENT REMOVAL (BNR)

2.6.1 System Configuration

Most of the MBR systems implemented to date have incorporated fully aerobic (COD removal and nitrification) reactors with very few incorporating biological nitrogen (N) removal in anoxic-aerobic (additional denitrification) sequences and phosphorus (P) removal in an anaerobic-anoxic-aerobic sequence of reactors. However, effluent quality requirements are becoming more stringent or are already stringent (e.g. USA and South Africa) in terms of the nutrients nitrogen and phosphorus (N and P), to limit receiving water eutrophication. This, together with the increasing affordability of membranes, has stimulated interest in combining membranes with biological nutrient (N and P) removal (BNR) activated sludge systems. Such BNR systems require an anaerobic-anoxic-aerobic sequence of reactors (Wentzel et al., 1990). Hence, the membranes would have to be submerged in the aerobic reactor, which forms only part of the system volume. This in itself would introduce restrictions and limitations that would have to be taken into account in the design of such a system (see Chapter 3 for details).

2.6.2 Reports on BNR Performance

Some information is available in the literature on biological N removal in systems incorporating membranes. Hasar and Kinaci (2002) operated an N removal (anoxic-aerobic) sequencing batch reactor (SBR) with submerged membranes (details on the system are not clear), and compared this with a conventional SBR with SSTs (Table 2.4)

In the sMBR system, the average removal of COD was 97.1%. In the conventional system (CASP), the removal of COD was lower, at approximately 86%. While the average COD concentration in the effluent of the sMBR was 15 mg/P, the average COD in the conventional system was 75 mg/ℓ. It was concluded that the removal of organic pollutants by the sMBR was high in terms of COD, and therefore a good quality effluent could be achieved during the long term operation.

Table 2.4: Performances of sMBR and conventional activated sludge system (CASP) operated for 95 days (Hasar and Kinaci, 2002)

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Average influent</th>
<th>CASP Effluent</th>
<th>Effluent Efficiency (%)</th>
<th>sMBR Effluent</th>
<th>Efficiency (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>COD (mg/ℓ)</td>
<td>520</td>
<td>75</td>
<td>85.6</td>
<td>15</td>
<td>97.1</td>
</tr>
<tr>
<td>TP (mg/ℓ)</td>
<td>15.0</td>
<td>7.9</td>
<td>47.3</td>
<td>2.25</td>
<td>85.0</td>
</tr>
<tr>
<td>PO₄³⁻-P (mg/ℓ)</td>
<td>10.5</td>
<td>7.1</td>
<td>32.4</td>
<td>1.90</td>
<td>81.9</td>
</tr>
<tr>
<td>SS (mg/ℓ)</td>
<td>110</td>
<td>40</td>
<td>63.6</td>
<td>ND</td>
<td>100</td>
</tr>
<tr>
<td>Turbidity (NTU)</td>
<td>38</td>
<td>15</td>
<td>60.5</td>
<td>0.44</td>
<td>98.8</td>
</tr>
<tr>
<td>TKN (mg/ℓ)</td>
<td>48.3</td>
<td>30.2</td>
<td>37.5</td>
<td>3.4</td>
<td>93.0</td>
</tr>
<tr>
<td>NH₄⁺-N (mg/ℓ)</td>
<td>35.0</td>
<td>20</td>
<td>42.9</td>
<td>1.90</td>
<td>94.6</td>
</tr>
<tr>
<td>NO₃⁻-N (mg/ℓ)</td>
<td>0.94</td>
<td>3.0</td>
<td>-</td>
<td>13.5</td>
<td>-</td>
</tr>
</tbody>
</table>
In the sMBR the average removals of TKN and NH$_4^+$-N were 93% and 94.6% respectively, indicating successful nitrification. In contrast, in the CASP the nitrification efficiency was only 42.9%, for some unidentifiable reason. The P concentration in the effluent from the sMBR system was significantly lower than that in the effluent from the CASP. However, from the operating data it is not possible to determine if this was due to biological excess P removal (BEPR) or chemical supplementation.

The SS concentration in the filtrate of the sMBR was not detected because of the pore size of the UF membrane module. However, the effluent SS in the CASP was 40 mg/L.

Clearly, the sMBR system performance was superior to that of the CASP. However, comparison of the BNR achieved by the two systems is not clear: It is difficult to assess N removal performance due to the partial nitrification in the CASP system. Further, BEPR behaviour is not clearly defined.

Although, the suppliers of membranes do include BNR systems for N and P removal in their lists of full-scale applications, no information in the literature on such applications could be found.

2.7 CLOSURE

From this review it is evident that the inclusion of membranes for solid/liquid separation in the biological nutrient removal (BNR) activated sludge system offers several significant potential advantages, such as reduced bioreactor volume, elimination of secondary settling tanks (SSTs), effluent quality independent of mixed liquor settling and flocculation characteristics, improved effluent quality in terms of microbiological characteristics, higher waste activated sludge concentrations possibly obviating sludge thickening. These advantages, coupled with the robustness, simplicity to operate and increasing affordability of modern membranes, are making membranes increasingly attractive as the solid/liquid separation process in the BNR activated sludge system. However, the membranes impose a set of unique selection properties on the BNR activated sludge bioconoenosis; most if not all bacteria and higher organisms are retained in the system, whereas in the conventional activated sludge systems (CASS) with SSTs only those organisms that flocculate and settle are retained; the membrane systems are operated at considerably higher mixed liquor concentrations than the CASS, 8 000-20 000 mg TSS/l versus 3 000-5 000 mg TSS/l. Although the effects of these differences on the design and performance of completely aerobic biologically mediated COD removal and nitrification activated sludge systems are reasonably well established (Brindle et al., 1996 and Stephenson et al., 2000), their possible impacts on the design, operation and performance of the biologically mediated nutrient (nitrogen and/or phosphorus) removal activated sludge system remains to be quantitatively established. Very little information exists in the scientific literature on this aspect. This research project aims to address this deficiency.

2.8 REFERENCES


CHAPTER 3
THE IMPACT OF MEMBRANES ON THE DESIGN OF BNR ACTIVATED SLUDGE SYSTEMS

3.1 INTRODUCTION

This chapter summarises the process design requirements for integrated BNR and membrane solid-liquid separation systems, and compares these with the design of a conventional BNR system with SST solid-liquid separation, to evaluate the advantages and disadvantages of membrane BNR systems (detailed procedures are given by Ramphao et al., 2004, 2005). The intention of this evaluation is to illustrate the principles and procedures in the design of membrane BNR activated sludge systems and not to provide numerical values that can be directly applied in design. The focus of the evaluation is on flat panel membrane systems, since the Kubota® membranes implemented in this research are of this type. However, the principles developed in this chapter are general (but not the quantitative results), and can be applied to other membrane types.

![Figure 3.1: Membrane unit showing membrane panels with effluent manifold. (Photo: Aquator)](image)

3.2 EVALUATION METHODOLOGY

In principle, the design procedure for MBR BNR systems is as follows: With selected zone (anaerobic, anoxic and aerobic) mass fractions and associated recycle ratios (for desired N&P removal), the peak wet weather flow (PWWF), i.e. hydraulic considerations, determines the surface area of membranes required. This required surface area has associated with it a required aerobic zone volume and oxygen transfer rate (OTR). If the OTR of the membranes can meet the peak biological oxygen demand (i.e. low wastewater strength), then the aerobic zone (and hence biological reactor) volume is governed by the volume required to accommodate the membranes. However, if the membranes cannot meet the peak biological oxygen demand, then the aerobic (and hence biological reactor) volume is increased to accommodate additional fine bubble aeration capacity. Under these circumstances, (i.e. high wastewater strength), the OTR of the membrane and additional aeration systems govern the volume of the aerobic zone and hence the biological reactor. Accurate aeration information is therefore essential to correctly size the reactor. With the
volume of the biological reactor fixed by these two criteria, the sludge age is calculated such that the applied organic load generates sufficient sludge mass to meet the aerobic zone MLSS concentration required by the membranes (12 to 18 g/l for the Kubota® membranes considered here). The impact of these two criteria on the sludge age and volume requirements of membrane BNR systems in comparison with conventional BNR systems is evaluated for three cases, i.e. diurnal flow and load variation with a PWWF to average dry weather flow (ADWF) ratio (fq) of 2:1 with single (Case 1) and double (Case 2) storey membrane layouts in the aerobic reactor and an influent flow balanced case with a single storey membrane layout (Case 3). These three cases are evaluated treating both normal raw and settled municipal wastewater. The steady state BNR design model of Wentzel et al. (1990) formed the basis for the MBR BNR system design equations, which are detailed by Ramphao et al. (2004, 2005) and are not repeated here.

3.3 SOME PRELIMINARY COMMENTS

3.3.1 Adjustable sludge mass fractions

In multi-zone BNR systems with membranes in the aerobic reactor and fixed volumes for the anaerobic, anoxic and aerobic zones (i.e. fixed volume fractions as in an existing system), the mass fractions can be varied (within a range) by varying the inter-reactor recycle ratios (Ramphao et al., 2004, 2005). This zone mass fraction flexibility is a significant advantage of MBR BNR systems over BNR system with SSTs, because it allows changing the mass fractions to optimise biological N and P removal in conformity with influent wastewater characteristics and the effluent N and P concentrations required.

3.3.2 Requirement for nitrification

Because for a selected aerobic mass fraction (f_{maer}), the membrane or aerobic volume requirements fix the sludge age of the system, the lowest maximum specific growth rate of the nitrifiers at 20°C (\(\mu_{nm20}\)) to ensure nitrification can be calculated (for details see Ramphao et al., 2004, 2005). If this lowest \(\mu_{nm20}\) is less than the anticipated \(\mu_{nm20}\) design value (~0.45 to 0.50 /d), then the system can be expected to nitrify. Generally, because sludge ages in MBR BNR systems are long (see below), nitrification usually will not be a problem.

3.3.3 MBR BNR system simulation

If required, the performance of MBR BNR systems can be simulated with current BNR activated sludge models by returning the SST underflow into the aerobic zone from which the SST feed flow exits. However, such simulations require *a priori* information on the reactor and zone volumes and recycle flows, which would need to be determined with the steady state procedures set out by Ramphao et al. (2004). Experimental MBR BNR systems were operated in this research project to evaluate the applicability of these models to MBR BNR systems, i.e. do the kinetics developed for low TSS concentration CASS BNR systems apply also to high TSS concentration MBR BNR systems?

3.3.4 Installing membranes into BNR systems with secondary settling tanks

Combining membranes and SSTs for solid-liquid separation in the same BNR system is possible in a side stream membrane reactor with a low sludge return flow (20%) to the reactor to harvest a high grade effluent from the BNR reactor for re-use. Also, wasting activated sludge directly from such a side stream reactor obviates waste sludge thickening but the costs of this would have to be compared with other thickening systems such as dissolved air flotation.
3.4 EVALUATION

For this evaluation, a MLE ND (Figure 3.2) and a UCT NDBEPR (Figure 3.3) system are selected and the biological reactor volumes required for MBR systems are compared with those required for the equivalent conventional systems with SSTs treating the same wastewater and having the same design parameters (Table 3.1), except the sludge age and aerobic zone MLSS concentration ($X_{\text{taer}}$) which were fixed at 15d and 4.0 g TSS/l for the conventional systems with SSTs.

![Figure 3.2: The MBR modified Ludzack-Ettinger (MLE) nitrification-denitrification system with membrane solid-liquid separation.](image1)

![Figure 3.3: The MBR University of Cape Town (UCT) ND biological excess P removal (BEPR) system with membrane solid-liquid separation.](image2)

Table 3.1: Wastewater characteristics, organism kinetic and stoichiometric constants and system design parameters used in the evaluation.

<table>
<thead>
<tr>
<th>Wastewater characteristics</th>
<th>Organism constants</th>
<th>System design parameters</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Species</td>
<td>OHOs</td>
</tr>
<tr>
<td>WW</td>
<td>Y</td>
<td>0.45</td>
</tr>
<tr>
<td>Raw Settled</td>
<td>fS'u</td>
<td>0.13</td>
</tr>
<tr>
<td>p</td>
<td>fS'us</td>
<td>0.04</td>
</tr>
<tr>
<td>X_{\text{SS}}/S_{\text{ri}}</td>
<td>0.25</td>
<td>0.35</td>
</tr>
<tr>
<td>f_{\text{am}}</td>
<td>0.08</td>
<td>0.11</td>
</tr>
<tr>
<td>%f_{\text{SS},s}</td>
<td>90</td>
<td>80</td>
</tr>
<tr>
<td>Temp °C</td>
<td>14</td>
<td>14</td>
</tr>
</tbody>
</table>

Selecting for Case 1, a maximum membrane flux ($q_{\text{max}}$) of 1.0 m$^3$/m$^2$.d, a single storey membrane layout and a PWWF/ADWF ratio ($f_q$) of 2:1, the required sludge age ($R_s$) of the membrane system, the % peak total biological oxygen demand supplied by the membrane units and the % volume of the membrane system with respect to the equivalent SST BNR system versus the aerobic zone MLSS to influent COD concentration ratio ($X_{\text{taer}}/S_{\text{ri}}$) are shown in Figs 3.3a, b and c for the MLE ND system treating raw wastewater at 14°C with different aerobic mass fractions ranging from 1 (fully aerobic) though 0.65, 0.55, 0.45 and 0.35. Also shown in Figure 3.4a is the lowest $\mu_{\text{nm}20}$ to ensure nitrification for a factor of safety on $\mu_{\text{nm}20}$ ($S_f$) of 1.25.

In Figure 3.4a, for a raw wastewater COD ($S_{\text{ri}}$) of say 750 mg COD/l and an aerobic zone MLSS concentration ($X_{\text{taer}}$) of 15.0 g TSS/P, the $X_{\text{taer}}/S_{\text{ri}}$ ratio is 15000/750 = 20. With an aerobic mass fraction of say 0.55 in a MLE ND system, the sludge age needs to be 28d (Figure 3.4a). At this sludge age and aerobic mass fraction, the lowest $\mu_{\text{nm}20}$ to ensure nitrification is about 0.32 /d. Because this is well below an expected value of around 0.45 to 0.50 /d for normal municipal wastewater, the system will nitrify. From Figure 3.4b at $X_{\text{taer}}/S_{\text{ri}} = 20$, the membrane units supply...
90% of the peak biological oxygen demand, which indicates that the aerobic zone volume, and hence the sludge age, is governed by OTR. The aerobic zone therefore will comprise two compartments, a first with fine bubble aeration, which supplies 10% of the peak biological oxygen demand and a second containing the membranes supplying 90% of the peak biological oxygen demand. From Figure 3.4c, the reactor volume is 48% of the equivalent SST MLE ND system at the selected 15d sludge age and aerobic zone MLSS of 4 g/ℓ.

In Figure 3.4a, as X_{taer}/S_{ti} increases, so effectively S_{ti} decreases, because X_{taer} remains “fixed” at the selected aerobic zone MLSS concentration. A high S_{ti} (low X_{taer}/S_{ti}) sets a high organic load on the biological reactor with the result that a short sludge age generates sufficient sludge mass to achieve the selected aerobic zone MLSS concentration (Figure 3a). However, the high organic load stimulates a high peak biological oxygen demand with the result that the membranes supply a small percentage of the biological oxygen demand (Figure 3b). Additional aerobic zone volume therefore has to be provided to accommodate fine bubble aeration to supply the biological oxygen demand deficit. This increases the sludge age (and also the oxygen demand, but marginally so), because the additional volume requires a greater sludge mass to be carried in the reactor to maintain the selected aerobic zone MLSS concentration (X_{taer}) – hence the flatter slope of the sludge age line in Figure 3a for X_{taer}/S_{ti} <23.

At high S_{ti} (low X_{taer}/S_{ti}), the reactor volume is therefore governed by the OTR (as indicated on Figure 3.4a), which establishes a longer sludge age in the system than if accommodation of the membranes only governs the volume of the aerobic zone (at low S_{ti} or high X_{taer}/S_{ti}).

Figures 3.3a, b and c show also the effect of the aerobic mass fraction (f_{maer}) from 1.00 (fully aerobic) through 0.65, 0.55, 0.45 and 0.35. The f_{maer} is a very important design parameter for BNR systems as generally the lower the f_{maer}, the higher the N and P removal, especially because membranes eliminate the concern about poor settling sludges. From Figure 3.4a it can be seen that the smaller the f_{maer}, the longer the sludge age for the same X_{taer}/S_{ti}. This is because essentially the aerobic zone volume, and hence the whole reactor volume, is governed by the biological oxygen demand or membrane volume requirements. The smaller f_{maer}, the larger the reactor volume and concomitantly, the longer the sludge age to generate sufficient sludge mass to operate at the selected X_{taer}. From Figure 3.4c, it can be seen that at low aerobic mass fractions (0.45 and 0.35), the volume of the membrane system increases above 60% of the equivalent SST BNR system, which begins to erode the footprint advantages of using membranes.
Figure 3.4: MBR MLE ND system sludge age (bold lines) and lowest maximum specific growth rate of nitrifiers at 20°C ($\mu_{\text{min20}}$, thin lines) (Figure 3.4a, top left), % peak total biological oxygen demand supplied by the membranes (bio OD, Figure 3.4b, top right) and % volume with respect to equivalent conventional BNR system with SSTs (% Conv, Figure 3.4c, bottom left) versus aerobic zone TSS to influent COD concentration ratio (mg TSS/mg COD) for aerobic mass fractions of 1.00 FA (fully aerobic), 0.65, 0.55, 0.45 and 0.35 treating raw wastewater (Table 3.1) at 14°C for Case 1 (PWWF/ADWF ratio $f_q = 2.0$, single storey membrane units).

For the raw wastewater and MLE ND system (Figure 3.4a), when $X_{\text{aer}}/S_{ti} \sim 23$, the membrane units supply 100% of the peak biological oxygen demand. The biological oxygen demand therefore no longer needs to be supplemented with fine bubble aeration with the result that the reactor volume is governed by the aerobic zone volume to accommodate the membranes. At $X_{\text{aer}}/S_{ti} = 23$ and $X_{\text{aer}} = 15000 \text{ mg/ℓ}$, $S_{ti} = 650 \text{ mg COD/ℓ}$. This influent COD concentration can therefore be regarded as the "balanced design" value, above which the aerobic zone volume, and hence the whole reactor volume, is governed by the OTR and below which, the reactor volume is governed by accommodation of the membranes with the membrane OTR supplying oxygen in excess of the peak biological oxygen demand. Treating the raw wastewater in a UCT NDBEPR system with membranes leads to similar trends as for ND systems except that the sludge ages are 20 to 30% shorter. This is because per kg COD/d organic load, NDBEPR systems generate about 15-25% more MLSS (with >90% aerobic P uptake BEPR, Ekama and Wentzel, 2004) and utilise about 3-5% less oxygen (these percentages vary with sludge age) than ND systems.

For Case 1, treating settled wastewater (Table 3.1) in a MLE ND system the trends are similar to raw wastewater, except that the sludge ages are extra-ordinarily long (100-150d) and the reactor volumes approach those for equivalent SST BNR systems. Because settled wastewaters have low COD concentrations (except in low per capita water consumption countries like South Africa), the sludge age of the system will be governed by the volume requirements of the membranes, with the membrane units usually supplying >100% of the biological oxygen demand. For these situations,
the volume of the reactor is the same for raw and settled wastewater *because it is fixed by the PWWF*. The sludge mass in the reactor also is the same (for the same $X_{taer}$) but because the organic load of settled wastewater is much lower than raw wastewater, the sludge age is much longer than for raw wastewater. Two options are available to reduce the volume of the reactor, (i) additional membranes can be installed in the aerobic zone in double storey layout, which reduces the aerobic zone volume per membrane area (Case 2) or (ii) influent flow balancing can be installed which reduces the PWWF and therefore reduces the membrane area required (Case 3).

Installing double storey membranes (Case 2) in a MLE ND system treating raw wastewater, decreases the sludge age required by about 20% compared with single storey membranes and the volume of the aerobic zone, and therefore the biological reactor, is governed by the OTR even for high $X_{taer}/S_{U}$. This is because (i) with more membrane area per aerobic zone volume, a greater PWWF can be handled, which for a fixed PWWF/ADWF ratio ($f_{q}=2$) results in a higher ADWF, organic load and biological oxygen demand per aerobic zone volume and (ii) a lower airflow rate and therefore a lower membrane unit OTR per membrane area. For double storey membranes (Case 2) treating settled wastewater in a MLE ND system, the sludge ages are significantly decreased compared with single storey, but these are still very long (50-100d), especially at low aerobic mass fractions where they are extremely long (100-150d). Generally, the lower the organic load per PWWF, the longer the sludge age and the greater the volume of the membrane BNR system compared with the equivalent conventional system. For weak settled wastewater and low aerobic mass fractions, this can be greater than 100% of the equivalent SST BNR system! Although NDBEPR systems produce more sludge and require less oxygen than ND systems, resulting in shorter sludge ages, they usually have lower aerobic mass fractions. Hence, the sludge ages for these systems also are very long and the percentage volume compared with the equivalent conventional system lie between 50 to 100%.

A way to significantly reduce the required membrane area, and hence the volume of the reactor, is to introduce influent flow balancing (Case 3). This is possible in areas where PWWFs are of relatively short duration (a few hours) – for areas with sustained wet weather flows lasting several days, this will not work. With flow balancing, the PWWF per organic load is significantly reduced resulting in shorter sludge ages and hence lower volumes for the system. With flow balancing ($f_{q}=1$) and single storey membrane layout, and reducing the maximum membrane flux from 1.0 to 0.7 $\text{m}^3/(\text{m}^2\cdot\text{d})$, the OTR of the membrane and fine bubble aeration systems govern the volume of the aerobic zone and hence also of the biological reactor for ND and NDBEPR systems treating raw wastewater ($X_{taer}/S_{U}$ from 10 to 40). For NDBEPR systems treating settled wastewater with $X_{taer}/S_{U} > 38$ (Figure 3.5a) (for ND systems >43), the membrane units supply >100% of the biological oxygen demand (Figure 3.5b). However, the sludge ages are significantly lower for both ND and NDBEPR systems compared with no flow balancing, resulting in reactor volumes between 40 and 60% of the equivalent SST BNR systems even for fairly low aerobic mass fractions (Figure 3.5c).
3.5 CONCLUSIONS

Installing membranes for solid-liquid separation into biological nutrient removal (BNR) activated sludge (AS) systems makes a profound difference not only to the design of the BNR system itself, but also to the approach to design of the whole wastewater treatment plant (WWTP).

In multi-zone BNR systems with membranes in the aerobic reactor and fixed volumes for the anaerobic, anoxic and aerobic zones (i.e. fixed volume fractions), the mass fractions can be varied (within a range) by varying the inter-reactor recycle ratios. This zone mass fraction flexibility is a significant advantage of membrane BNR systems over conventional BNR systems with SSTs, because it allows changing the mass fractions without changing the zone volumes to optimise biological N and P removal in conformity with influent wastewater characteristics and the effluent N and P concentrations required.

In principle, the design procedure for membrane BNR systems is as follows: With selected zone mass fractions and associated recycle ratios (for desired N&P removal), the peak wet weather flow (PWWF), i.e. hydraulic considerations, determines the surface area of membranes required. This required surface area has associated with it a required aerobic zone volume and oxygen transfer rate (OTR). If the OTR of the membranes can meet the peak biological oxygen demand (i.e. low wastewater strength), then the aerobic zone (and hence biological reactor) volume is governed by...
the volume required to accommodate the membranes. However, if the membranes cannot meet the peak biological oxygen demand, then the aerobic (and hence biological reactor) volume is increased to accommodate additional aeration capacity. Under these circumstances (i.e. high wastewater strength), the OTR of the membrane and additional aeration systems govern the volume of the aerobic zone and hence the biological reactor. Accurate aeration information is therefore essential to correctly size the reactor. With the volume of the biological reactor fixed by these two criteria, the sludge age is calculated such that the applied organic load generates sufficient sludge mass to meet the aerobic zone MLSS concentration required by the membranes (12 to 18 g/l for the Kubota® membranes considered here).

Generally, the longer the sludge age of the membrane system, the greater the biological reactor volume as a percentage of the equivalent conventional system volume at 15d sludge age and 4 g TSS/l aerobic zone concentration. Although nitrification denitrification biological excess P removal (NDBEPR) systems generate more sludge (15-25%) and utilise less oxygen (3-5%) than ND systems and therefore require shorter sludge ages, the difference is not large enough to make a significant difference to the volume of the membrane NDBEPR system as a percentage of the equivalent conventional system volume. The aerobic mass fraction (f_{maer}) has the greatest impact on the sludge age of the system and therefore also the membrane reactor volume as percentage of the equivalent conventional system volume – the lower the aerobic mass fraction, the longer the sludge age and the greater the percentage volume. For f_{maer}<0.50 and low wastewater strengths, the volume of the membrane BNR system increases above 50% of the equivalent conventional system volume, which erodes some of the volume reduction advantage of using membranes. Influent flow balancing significantly reduces the sludge age of the membrane BNR system and hence also the percentage of the equivalent conventional system volume. Double storey membrane units (without flow balancing) also reduces the sludge age and the percentage volume of the equivalent conventional system, but not as significantly as influent flow balancing.

Generally for low aerobic mass fraction (f_{maer}<50%) single storey membrane BNR systems without flow balancing treating low strength (usually settled) wastewater (X_{laer}/S_{li} > 30), the sludge age to achieve the required aerobic zone TSS concentration will be extremely long (>100d). Hence, the percentage volume of the conventional system is high and the membrane aeration system will supply more oxygen than required for the peak total biological oxygen demand, resulting in uneconomical operation. Further, the extremely long sludge ages will result in reduced BEPR (Wentzel et al., 1990). Treating high strength (usually raw) wastewater for this situation results in much shorter but still long sludge ages (>40d). Hence the percentage volume of the conventional system is lower and there is a much greater utilisation of the oxygen transfer of the membrane aeration system. Treating raw wastewater eliminates primary sludge production, and if the aerobic mass fraction and raw wastewater strength are not too high, the sludge age of the system will be sufficiently long to discharge a stable waste activated sludge (WAS) directly to dewatering systems. Therefore, for PWWF/ADWF ratios in the upper range and aerobic mass fractions in the lower range, the indicated mode of operation of membrane BNR systems is extended aeration. The cost of the membranes therefore can be offset not only by the reactor volume reduction but also by savings on sludge treatment costs.

If the PWWF/ADWF (f_{q}) ratio is reduced by flow balancing, between 50 and 100% more ADWF capacity can be achieved for the same membrane surface area, depending on the extent to which the f_{q} ratio is reduced. For systems with low PWWF/ADWF ratios and high aerobic mass fractions (>60%), it becomes feasible to treat settled wastewater because the sludge ages to achieve the high aerobic zone TSS concentration of 12 to 18 g/l are not very long with the result that the percentage volume of the conventional system is not very high. Moving from an influent flow unbalanced raw wastewater system to the flow balanced settled wastewater system can double the ADWF capacity.
of the biological reactor and results also in a much more efficient use of the oxygen transfer of the membranes systems. However, this increased capacity has to be traded-off against the cost of treating primary sludge. The WAS is likely to be stable because the sludge age for settled wastewater systems is still long (>30 days). This obviates having to stabilise the WAS, which for a BEPR system can be problematic due to the significant P (and N) release in the anaerobic and/or aerobic digesters (Pitman et al., 1991). Therefore with flow balancing, the design approach to the WWTP changes away from extended aeration to include primary and possibly secondary sludge stabilisation.

Combining membranes and SSTs for solid-liquid separation in the same BNR system is possible in a side stream membrane reactor with a low sludge return flow (20%) to the reactor to harvest a high grade effluent from the BNR reactor for re-use. Also, by wasting activated sludge directly from such a side stream reactor obviates waste sludge thickening but the costs of this would have to be compared with other thickening systems such as dissolved air flotation.

In this chapter, the impact on the design of BNR activated sludge systems and the encompassing wastewater treatment plant of including membranes for solid-liquid separation in the BNR system have been explored. The intention has been to illustrate the principles and procedures and not to generate numerical values that can be directly applied to design.

3.6 REFERENCES


CHAPTER 4

MBR VERSUS CAS BNR ACTIVATED SLUDGE SYSTEM
STEADY STATE BEHAVIOUR: PHASE 1 – FEASIBILITY

4.1 INTRODUCTION

As described in Chapters 1 and 2, the inclusion of membranes for solid/liquid separation in the biological nutrient removal (BNR) activated sludge system offers several significant potential advantages, such as reduced bioreactor volumes, elimination of secondary settling tanks (SSTs), effluent quality independent of mixed liquor settling and flocculation characteristics, improved effluent quality in terms of microbiological characteristics, higher waste activated sludge concentrations possibly obviating sludge thickening. These advantages, coupled with the robustness, simplicity to operate and increasing affordability of modern membranes, are making membranes increasingly attractive as the solid/liquid separation process of choice in the BNR activated sludge system. However, the membranes impose a set of unique selection properties on the BNR activated sludge biocoenosis; most if not all bacteria and higher organisms are retained in the system, whereas in the conventional activated sludge systems (CASS) with SSTs only those organisms that flocculate and settle are retained; the membrane systems are operated at considerably higher mixed liquor concentrations than the CASS, 8 000-20 000 mg TSS/ℓ versus 3 000-5 000 mg TSS/ℓ. Although the effects of these differences on the design and performance of completely aerobic biologically mediated COD removal and nitrification activated sludge systems are reasonably well established (Brindle et al., 1996 and Stephenson et al., 2000), their possible impacts on the design, operation and performance of the biologically mediated nutrient (nitrogen and/or phosphorus) removal activated sludge system remains to be quantitatively established. The main aim of this research is to address these deficiencies.

In Chapter 3, the impact of including membranes on the design of a membrane BNR activated sludge system was considered. However, what remains is for the impact of the membranes on the kinetics of the biologically mediated processes of the nutrient removal to be quantified. This is the main aim of the experimental part of this research project, and required:

- Evaluation of biological nutrient removal (BNR) performance at typical MBR total suspended solids (TSS) concentrations (12 000-18 000 mg TSS/ℓ), and
- Comparison of the biological N and P removal kinetic rates under MBR conditions (high reactor TSS concentrations, greater than 12 000 mg TSS/ℓ) with those in conventional BNR systems (low reactor TSS concentration, 4 000 mg TSS/ℓ).

To address these objectives required operation of parallel conventional and membrane BNR activated sludge systems at laboratory-scale, monitoring of their behaviour and comparing their performance. This experimental investigation took place in two phases: Phase 1 from April 2003 to July 2004 and Phase 2 from February 2005 to July 2006. Phase 1 focussed on an initial assessment of the BNR performance in MBR and conventional (CAS) BNR activated sludge systems, while Phase 2 continued this evaluation but additionally examined BNR kinetics. This Chapter summarises the results from the Phase 1 investigation (detail in Ramphao et al., 2004). The Chapter describes the set-up, operation and monitoring of the two systems, evaluation of their behaviour and compares their performance. The Phase 2 investigation is summarised in Chapter 5, and the BNR kinetics evaluation in Chapter 6.
4.2 SYSTEMS OPERATED

The research approach adopted for the experimental investigation was to operate two parallel laboratory-scale BNR activated sludge systems in the constant temperature Water Research Laboratory (20°C), one a MBR system with membrane solids-liquid separation and the other a conventional system with SSTs. Both BNR systems were 3 reactor anaerobic, anoxic aerobic UCT configurations (Figs 4.1 and 4.2, Table 4.1). The UCT configuration was selected because it allows the biological excess phosphate removal (BEPR) to operate independently of the nitrogen removal, i.e. zero nitrate recycle to the anaerobic reactor, provided the recycle(s) to anoxic reactor do not overload this reactor with nitrate.

Figure 4.1: Schematic layout of conventional UCT BNR activated sludge system.

Figure 4.2: Schematic layout of MBR UCT BNR activated sludge system.
Table 4.1: Initial MBR and conventional UCT systems’ design and operating parameters.

<table>
<thead>
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<th>SYSTEM PARAMETERS</th>
<th>MBR UCT</th>
<th>CONVENTIONAL UCT</th>
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</thead>
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<td>Sludge age (d)</td>
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<td>20</td>
</tr>
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<td>Anaerobic (R1) Volume (ℓ)</td>
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<td>5.6 (2.8(^1))</td>
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<tr>
<td>Anoxic (R2) Volume (ℓ)</td>
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<td>6.2</td>
</tr>
<tr>
<td>Aerobic (R3) volume (ℓ)</td>
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<td>Anaerobic (R1) mass fraction (%)</td>
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<tr>
<td>Anoxic (R2) mass fraction (%)</td>
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<td>27.9(^3)</td>
</tr>
<tr>
<td>Aerobic (R3) mass fraction (%)</td>
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<td>59.5(^3)</td>
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<td>s-sludge return recycle</td>
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</tr>
<tr>
<td>a-recycle (R3 to R2)</td>
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<tr>
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</tr>
<tr>
<td>MLVSS concentration (mg/l)</td>
<td>12 500</td>
<td>3 600</td>
</tr>
<tr>
<td>MLSS concentration (mg/l)</td>
<td>18 000</td>
<td>4 950</td>
</tr>
<tr>
<td>Influent flow (ℓ/d)</td>
<td>140</td>
<td>15</td>
</tr>
<tr>
<td>Feed COD concentration (mg/l)</td>
<td>1 000</td>
<td>1 000</td>
</tr>
<tr>
<td>OUR (mg O/ℓ/h)</td>
<td>135</td>
<td>37</td>
</tr>
<tr>
<td>Membrane flux (m(^3)/m(^2)/d)</td>
<td>0.239</td>
<td>-</td>
</tr>
</tbody>
</table>

\(^1\) effective volume for an r-recycle = 1:1

\(^2\) The MBR UCT aerobic volume is in fact 32P, the extra 2 P is for a re-aeration reactor where OUR is measured (see Fig. 4.2 below).

\(^3\)For the given a- and r-recycle ratios.

The two systems were set up to have identical design parameters such as anaerobic (\(f_{an} = 0.126\)), anoxic (\(f_{anox} = 0.279\)) and aerobic (\(f_{aer} = 0.595\)) mass fractions, recycles (a+s = 3:1, r = 1:1) and
sludge age (20 d). The only differences were the influent flow (140 P/d and 15 ℓ/d) and total reactor volume (74 ℓ and 25 ℓ), due to the physical constraints imposed by the dimensions of the A4 size membrane panels, and the much higher reactor concentrations in the MBR system. The membranes considered in this investigation were the Kubota® panel type (Figure 4.3).

The conventional UCT system consisted of a total process volume of 25 ℓ, divided into 5.6 ℓ anaerobic, 6.2 ℓ anoxic and 13.2 ℓ aerobic with secondary settling tank (SST), to give an unaerated mass fraction of about 40.5% (i.e. 12.6% anaerobic and 27.9% anoxic). All reactors were in series with an underflow recycle (s-recycle) from the SST to the anoxic reactor of 1:1, a recycle from aerobic to anoxic reactors (a-recycle) of 2:1 and a recycle from anoxic to anaerobic reactors (r-recycle) of 1:1, all recycles with respect to influent flow.

In the parallel MBR UCT system, the volumes were 19 ℓ and 21 ℓ for the anaerobic and anoxic reactors respectively, also giving an unaerated mass fraction of about 40.5%. The aerobic reactor was 34 ℓ, and contained the 5 double sided A4 size Kubota panel membranes with total surface area 0.586 m² (Figure 4.3). The combined “as” recycle (aerobic to anoxic) was set-up at 3:1 and the r-recycle at 1:1.

Both the MBR and CAS UCT systems were operated at 20 d sludge age, by withdrawing the required volume of mixed liquor (including samples) daily from the aerobic reactors.

4.3 SYSTEM FEED

The two experimental UCT systems were fed raw sewage (800 mg COD/ℓ) from Mitchells Plain Wastewater Treatment Plant, augmented with sodium acetate (200 mg COD/ℓ, to accentuate BEPR), ammonia (20 mg N/ℓ, to increase TKN/COD) and phosphorus (sufficient to ensure systems not P limited). The sewage was collected in 2 m³ batches, macerated, stored in 400 ℓ stainless steel tanks refrigerated at 4°C and served as feed for approximately 2-3 weeks. In the Phase 1 experimental period, the experimental investigation used 20 and 21 (raw) sewage batches for the MBR and conventional UCT systems respectively, and extended over a period of 397 days.

4.4 SYSTEM MONITORING

The performances of the two systems were extensively monitored and compared (Table 4.2), to identify and quantify the influence of the membranes on system response.
Table 4.2: Sampling position and parameter measurement.

<table>
<thead>
<tr>
<th>TEST</th>
<th>COD1</th>
<th>TKN2</th>
<th>FSA3</th>
<th>NO$_3$4</th>
<th>NO$_2$5</th>
</tr>
</thead>
<tbody>
<tr>
<td>Influent</td>
<td>F,*</td>
<td>*</td>
<td>F</td>
<td>*</td>
<td>*</td>
</tr>
<tr>
<td>Anaerobic</td>
<td>F</td>
<td>F</td>
<td>F</td>
<td>F</td>
<td>F</td>
</tr>
<tr>
<td>Anoxic</td>
<td>F</td>
<td>F</td>
<td>F</td>
<td>F</td>
<td>F</td>
</tr>
<tr>
<td>Aerobic</td>
<td>*</td>
<td>*</td>
<td>F</td>
<td>F</td>
<td>F</td>
</tr>
<tr>
<td>Final Effluent</td>
<td>F*</td>
<td>F*</td>
<td>F</td>
<td>F</td>
<td>F</td>
</tr>
</tbody>
</table>

F- Filtered through Schleicher & Schull ME 25/21 0.45 μm membrane filters.
* Unfiltered samples.
- Direct measurement taken (filtering not applicable).

1. COD Chemical Oxygen Demand, open reflux method; 5220 (B)
2. TKN Total Kjeldahl Nitrogen, micro-kjeldahl method; 4500 – N$_{org}$ (C)
3. FSA Free and Saline Ammonia, titrimetric method; 4500 – NH$_3$ (B), (E)
4. NO$_3$ Hydrazine reduction (Technicon Auto-Analyzer); 4500 – NH$_3$ (H)
5. NO$_2$ Hydrazine reduction (Technicon Auto-Analyzer); 4500 – NO$_2$ (H)
6. T-P Total Phosphorus; Sulphuric acid/ per sulphate digestion at 100°C followed by molybdate-vanadate colour development for ortho-phosphate (Standard Methods, 1985 – Method 424C III)
7. TSS Total Suspended Solids dried at 103-105°C; 2540 (D)
8. VSS Volatile Suspended Solids ignited at 600°C; 2540 (E)
9. DSVI Dilute Sludge Volume Index; (Ekama and Marais, 1984), 271(D)
10. OUR Oxygen Utilisation Rate; automated (Randall et al., 1991), 271(B)
11. pH pH meter, Hanna Instruments model HI9023; 4500 – H$^+$ (B)

4.5 SYSTEM RESPONSES

From the experimental results measured during Phase 1 of the investigation, Ramphao et al. (2004) compiled a detailed report. Only the conclusions of this report are given below.

4.5.1 Initial membrane tests

Initial test were conducted to determine membrane potential fluxes, transmembrane pressure (TMP), the oxygen transfer rate and the residence time distribution (RTD) in the (aerobic) membrane bioreactor. These indicated:

- The membranes could readily accommodate the required operational fluxes (~0.24 m$^3$/m$^2$/d) and TMP (~100 mm water).
- With activated sludge over a short duration, only a small increase in TMP was required to achieve the same flux as distilled water; however, with time it was expected that the activated sludge would generate a dynamic gel layer on the membrane surface influencing the flux – TMP relationship so TMP was continually monitored (see below).
- The expected OUR (135 mg O/ℓ/h) could be readily sustained at a mixed liquor DO concentration even as high as 5.0 mg O/ℓ.
• RTD tests indicated that the aerobic reactor containing the membranes could be accepted to be completely mixed.

4.5.2 System operation

With regard to operation:

• For the MBR UCT system, long term aerobic and anoxic mixed liquor concentrations were 19-21 g TSS/ℓ and 14-16 g TSS/ℓ respectively. When the a-recycle blocked, the aerobic reactor concentration increased while the anoxic reactor concentration decreased. In the CAS system with SST, the aerobic and anoxic reactor concentrations are closely equal and independent of the a-recycle ratio. Clearly, the distribution of mixed liquor in an MBR BNR system is significantly different from that in a conventional BNR system, and is directly influenced by the a-recycle ratio. This is in agreement with the design framework developed in Chapter 3 for MBR UCT systems.

• The zone mass fraction flexibility above is a significant advantage of membrane BNR systems over conventional BNR system with SSTs, because it allows changing the zone mass fractions without changing the zone volumes to optimise biological N and P removal in conformity with influent wastewater characteristics and the effluent N and P concentrations required.

• Since the MBR UCT system was operated at higher reactor concentrations than the conventional UCT system (a requirement for the membranes) the influent COD mass loading per unit volume was 3.2 times higher than for the CAS UCT system, at 1 867 mg COD/(ℓ reactor.d) versus 592 mg COD/(ℓ reactor.d). This reflects the potential substantial reactor volume savings that can be achieved in an MBR type system. However, as described in the design procedures developed in Chapter 3, these volume reductions are strongly influenced by the nature of the influent, in particular the PWWF/ADWF ratio (fq), the COD strength and the aerobic mass fraction (fmaer).

4.5.3 Trans-Membrane Pressure (TMP)

The observations on the TMP variation during the experimental period strongly suggest that (Figure 4.4):

![Graph showing TMP and DSVI over time](image)

Figure 4.4: Trans membrane pressure (TMP) and diluted sludge volume index (DSVI) with time in the Phase 1 investigation. TMP is independent of sludge settleability (Figure 4.4). In BNR activated
sludge systems, which are prone to poor settling sludges, this would confer a considerable advantage on the MBR systems over conventional systems with SSTs, in that the MBR system operation is essentially independent of the sludge settling characteristics.

- With time there is a progressive increase in TMP at a rate of 0.115 mm/d to maintain a constant flux which cannot be reversed with water or chemical cleans (Figure 4.4). This would indicate a finite life-span for the membranes. Such deterioration in TMP (or equivalently membrane flux) has not been observed at full-scale (Kennedy, pers comment). However, the 0.115 mm/d TMP increase needs to be seen in context – over 20 years this would amount to an increase of only 700 mm, which is acceptable for full scale membrane applications.

4.5.4 Mass Balances

In analyzing the experimental data:

- Good N mass balances were obtained over the two UCT systems operated in this Phase 1 of the investigation – the investigation average (IA) was 103.5 and 95.5% for the MBR and CAS systems respectively. This indicates reliable experimental and analytical techniques and surety in data.

- Relatively low COD mass balances were obtained (IA 90.5 and 87.1% for the MBR and CAS systems respectively). However, these COD mass balances are similar to those obtained previously over NDBEPR systems (84-90%), but lower than in ND only systems (95-105%). This would indicate that the presence of an anaerobic reactor/zone may lead to a loss of COD that is not taken into account in the calculated COD mass balance. Although this COD “loss” has been incorporated into some kinetic simulation models for BEPR systems, the underlying mechanisms for the COD “loss” have not been established so that this incorporation is essentially empirical; this aspect requires further investigation (Hu et al., 2006a,b).

4.5.5 System Removals and Effluent Quality

From the experimental investigation:

- The MBR system exhibited removals that were equivalent or superior to those produced by the CAS system (Table 4.3), viz.
  - COD removal, 96 versus 94%
  - TKN removal, 97 versus 97%
  - FSA removal, 98 versus 97%
  - Total nitrogen removal, 74 versus 75%
  - Phosphorus (P) removal, 66% (27 mg P/ℓ) versus 54% (22 mg P/ℓ).

- Of the removals above and in Table 4.3, the largest difference was in P removal which was substantially better in the MBR UCT system. The reason for the higher P removal was the difference in the nitrate concentration recycled to the anaerobic reactor (MBR 0, CAS 2 mg NO3-N/ℓ) and the consequent anoxic P uptake BEPR (see below).

- N removal was the same both systems. The nitrate that the anoxic reactor of the CAS system could not denitrify was denitrified in the anaerobic reactor, causing the lower P
removal. At the lower mixed liquor concentration, the denitrification in the anoxic reactor of the CAS system is more sensitive to DO in the recycles than that in the MBR system.

- The MBR system produced a solids free effluent, whereas in the CAS system there was a continual loss of solids to the effluent (IA 15 mg TSS/l), quantified as the difference between filtered (IA 57 mg COD/l) and unfiltered (IA 73 mg COD/l) effluent COD concentrations. This confirms that the MBR UCT system effluent quality is independent of the flocculation characteristics of the mixed liquor.

- The MBR UCT system produced an effluent filtered (by the membranes) COD concentration (IA 35 mg COD/l) that was consistently lower than the effluent filtered (0.45 μm) COD concentration in the conventional UCT system (IA 57 mg COD/l). This implies that the membranes retain organics that would be considered “soluble” in a conventional system with SSTs (~ 22 mg COD/l). This is accommodated in the steady-state models as a reduced unbiodegradable soluble COD fraction (fS,us) in the influent, determined as 0.036 (36 mg COD/l) and 0.058 (57 mg COD/l) respectively for the two systems of this investigation.

- The microbiological quality (faecal coliforms) of the MBR UCT system effluent (0 cfu/100 ml) was superior to that of the conventional UCT system (> 30 cfu/100 ml). However, only one test was performed and this needs to be repeated to substantiate the results. In any event, the relative difference is not reflective of full scale plants.

- If the effluent is to be re-used in water reclamation, the improvement in effluent quality from an MBR UCT system is of considerable benefit. It may be possible to use the MBR UCT system effluent directly for non-potable water applications, such as toilet flushing and gardening. Further, if water of potable quality is required, the subsequent treatment unit processes probably will be significantly reduced.

### Table 4.3: Comparison of MBR and conventional UCT system performances.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Influent (common)</th>
<th>UCT System Type</th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>MBR Effluent Efficiency (%)</td>
<td>Conventional Effluent Efficiency (%)</td>
<td></td>
</tr>
<tr>
<td>COD (mg/l)</td>
<td>987</td>
<td>35 96</td>
<td>57¹ (73²) 941</td>
<td></td>
</tr>
<tr>
<td>TKN (mg N/l)</td>
<td>103</td>
<td>3.2 97</td>
<td>3.2 97</td>
<td></td>
</tr>
<tr>
<td>FSA (mg N/l)</td>
<td>84</td>
<td>1.9 98</td>
<td>2.6 97</td>
<td></td>
</tr>
<tr>
<td>NO₃ (mg N/l)</td>
<td>0</td>
<td>23 -</td>
<td>22 -</td>
<td></td>
</tr>
<tr>
<td>TN (mg N/l)</td>
<td>103</td>
<td>26 74</td>
<td>25 75</td>
<td></td>
</tr>
<tr>
<td>TP (mg P/l)</td>
<td>41</td>
<td>14 66</td>
<td>19 54</td>
<td></td>
</tr>
<tr>
<td>TSS (mg/l)</td>
<td>N/A</td>
<td>0 100</td>
<td>15 -</td>
<td></td>
</tr>
<tr>
<td>CFU/100 ml</td>
<td>N/A</td>
<td>0 -</td>
<td>30 -</td>
<td></td>
</tr>
</tbody>
</table>

¹Filtered sample (0.45 μm); ²unfiltered sample
N/A = not available; FSA = free and saline ammonia; TN = total nitrogen; TP = total phosphorus; CFU = colony forming unit

### 4.5.6 Anoxic P uptake

From the experimental investigation:
In both systems some anoxic P uptake BEPR was observed. This anoxic P uptake was more prevalent and of higher magnitude in the CAS system than in the MBR system, 39 and 17% of total P uptake respectively.

With anoxic P uptake by denitrifying PAOs (DPAOs), significantly reduced BEPR has been reported (Ekama and Wentzel, 1999; Hu et al., 2002) and was evident here also, due to less efficient utilisation of the influent RBCOD by the Phosphorus Accumulating Organisms (PAOs) (Hu et al., 2002). Conditions identified that induces anoxic P uptake by denitrifying (D)PAOs include (i) nitrate load on the anoxic reactor exceeding the denitrification potential of the reactor, i.e. positive nitrate concentrations in the anoxic reactor and (ii) low aerobic and high anoxic mass fractions (Hu et al., 2002).

The occurrence of anoxic P uptake BEPR in the MBR system appeared to be linked to the a-recycle ratio; decreasing the as-recycle ratio to 2:1 favoured anoxic P release and restoring it to 3:1 favoured anoxic P uptake. Increasing the as-recycle ratio in the MBR system had three main effects; it (i) increased the concentrations of mixed liquor in the anoxic and anaerobic zones and hence increased the mass fractions, (ii) increased the nitrate load on the anoxic reactor, and (iii) increased frequency of alternation between the anoxic and aerobic zones. Hu et al. (2002) noted that increased frequency of alternation appears detrimental to DPAOs and hence anoxic P uptake. Accordingly, the causes for the shifts in anoxic P uptake appear to be due to (i) and/or (ii) above. Of these, the (i) increased in anaerobic and anoxic mass fractions only marginally and so were unlikely to have stimulated the effects observed. Accordingly, the observed changes in anoxic P uptake were ascribed to the nitrate load – higher nitrate loads tend to favour anoxic P uptake, in agreement with the observations of Hu et al. (2002).

In the CAS system, the nitrate load to the anoxic reactor exceeded its denitrification potential and anoxic P uptake was high.

The observations above, and the fact that anoxic P uptake results in reduced BEPR, would suggest that the nitrate load in both systems should have be reduced by reducing the a-recycle ratio to maximise BEPR. In the CAS system, reducing the a-recycle ratio does not impact on the anoxic mass fraction and hence, provided the nitrate load remains just less than the denitrification potential, the N removal performance will decrease. However, in the MBR system reducing the a-recycle ratio causes a corresponding reduction in the anoxic mass fraction and hence impacts adversely on N removal performance. This highlights the conflicting requirements between N and P removal in BNR activated sludge systems which is by no means unique to the MBR configuration. The balance between optimizing N and P removal performance will be influenced by a number of factors, such as influent TKN/COD ratio, P removal requirements, N removal requirements, influent RBCOD.

At present anoxic P uptake BEPR is not explicitly incorporated in the steady-state design procedures for BEPR systems, because quantitative relationships linking the extent of anoxic P uptake to the system design or operational parameters have not been established (Hu et al., 2002, 2006a,b). In the interim, the lower BEPR with anoxic P uptake can be accommodated by reducing the P content of the PAOs (fXBGP), 0.259 and 0.232 mg P/mg VSS here for the MBR and CAS systems respectively. However, predicting a priori exactly what value to use remains uncertain. Clearly this requires further investigation.
4.5.7 Sludge Production

The MBR UCT system had significantly different sludge production:

- The MBR UCT system had a greater sludge production per COD load (6.5 kg VSS in the bioreactor/kg COD load; 0.32 kg VSS wasted/kg COD load = 0.41 kg TSS/kg COD load) than the conventional UCT system (4.5 kg VSS in the bioreactor/kg COD load; 0.27 kg TSS/kg COD load = 0.22 kg VSS wasted /kg COD load). This increased sludge production can be explained in part by the retention of solids that would normally be lost to the effluent in a conventional system with SSTs, and that the membranes retain unbiodegradable organics that would be considered “soluble” in a system with SSTs and hence also would be lost to the effluent. However, these explanations do not account for the magnitude of increase in sludge production. More comprehensive data at different sludge ages are required to determine if this increase is consistent, and to identify the underlying cause.

- In the steady-state design procedures, the increased sludge production in MBR UCT systems is accommodated by increasing the influent unbiodegradable particulate COD fraction ($f_{S,us}$), determined as 0.224 versus 0.067 for the conventional UCT system.

4.5.8 Sludge settleability

From the experimental investigation:

- During the investigation monthly microscopic examination of the mixed liquors from the two systems determined anoxic – aerobic (AA, Casey et al., 1994) filaments as the dominant ones in both systems (M. parvicella and 0092). Mixed liquor settleability measured with the diluted sludge volume index (DSVI) in the two systems varied over the investigation period, with MBR DSVI ranging from 80 to 135 mL/g, and that in the conventional system, from 80 to 240 mL/g. However, the variations in DSVI in the two systems were not in agreement – the DSVIs did not increase or decrease simultaneously.

4.6 CLOSURE

This initial Phase 1 study clearly demonstrated that BNR in membrane activated sludge systems is entirely feasible, and offers considerable advantages over conventional BNR activated sludge systems with SSTs, such as improved effluent quality, operation independent of sludge flocculation or settling characteristics, etc. From the investigation, a number of aspects requiring attention were identified (Ramphao et al., 2004):

- In the systems operated, the reactors were completely mixed and hence the influence of the membranes on N and P removal kinetics could not be established. This requires separate kinetic studies in which the time dependent responses of the relevant processes can be quantified. This was investigated with batch tests (see Chapter 6).

- The observed higher sludge production in the MBR UCT system does not substantiate whether this is consistently true for MBR systems. More comprehensive data at different sludge ages are required. This was partly addressed (see Chapter 5).

- It should be recognised that the higher sludge production in the MBR, manifesting as a higher unbiodegradable particulate COD fraction, is not that a significant disadvantage. This is because most MBR BNR systems will not be sized on a sludge age – sludge
production basis, as CAS BNR systems are, but on peak wet weather flow (PWWF) or the oxygen transfer rate (OTR) of the aeration systems (Chapter 3). These criteria force long sludge ages onto the MBR BNR systems, significantly longer than equivalent SST BNR systems. It is therefore not that significant that the MBR BNR system produces 10 to 20% more sludge (if this is confirmed by the ongoing research), it just means that the sludge age can be decreased by about 10 to 20%. Since sludge ages tend to be long in high mixed liquor TSS concentration MBR BNR systems, nitrification is not a critical factor in selection of system sludge age (as it is for CAS BNR systems). Therefore, reducing the sludge age by 10 to 20% will not impact nitrification performance (but may be slightly beneficial to P removal, Wentzel et al., 1990). However, the higher sludge production does affect plant economics. Higher sludge production results in higher sludge treatment and disposal costs, but this is compensated for by the lower oxygen demand to conform to the COD mass balance over the system.

- Anoxic P uptake has been consistently observed in this and other investigations. Anoxic P uptake is detrimental to the BEPR performance in a BNR system. However, quantitative links between design and operational parameters and the extent of anoxic P uptake have not been established. This has hindered incorporation of anoxic P uptake in the design and simulation models for BNR systems, with or without membranes, and requires resolution. The additional understanding generated from the batch tests to determine the effect of high mixed liquor TSS concentration on the anaerobic P release and acetate uptake rates and the anoxic and aerobic P uptake rates is presented in Chapter 6.

4.7 REFERENCES


CHAPTER 5

COMPARISON OF CAS AND MBR BNRAS SYSTEM BEHAVIOUR – PHASE 2

5.1 INTRODUCTION

In Phase 1 of the project, two biological nutrient removal (BNR) activated sludge systems in UCT configurations were run in parallel (Chapter 4). The first system was a membrane bioreactor (MBR) UCT system and the other a conventional (CAS) UCT system using secondary settling tanks (SST’s). The design parameters for both systems were the same as far as possible (except the hydraulic retention time) to allow an accurate comparison of the performances of the MBR and CAS systems.

This first phase (Chapter 4) showed that the MBR system could operate effectively and produce an effluent of equal or better quality to that produced by the CAS system. From the conclusions of the Phase 1 investigation, it was recommended that further investigations into the kinetics of BNR in the MBR be undertaken, to better understand the influence of the membranes, and how high mixed liquor TSS concentration impacts on the biological activity and behaviour of the micro-organisms in the system. Specific areas requiring further research included:

- Elucidating the nutrient removal kinetic behaviour of the MBR BNR system, and comparing this with the equivalent kinetics in a CAS system.
- Verifying the higher excess sludge production by the MBR system.
- Qualifying the phosphorus uptake in the system (particularly anoxic-P uptake) and how to optimise BEPR in the system.
- Investigation the interaction between aeration rates, membrane flux and transmembrane pressure (TMP).

The last three areas required the MBR and CAS systems to be run and monitored at steady state for a longer period of time, and the first would necessitate batch testing from a well defined and controlled steady state systems. These formed the objectives of Phase 2 of the project. This Chapter deals specifically with the operation and observations of the parallel MBR and CAS systems over Phase 2 during the time the batch tests to elucidate the impact of high mixed liquor TSS concentration on BNR kinetics were conducted. This chapter summarises the results from the Phase 2 investigation (detail in du Toit et al., 2006). The results of the batch tests are presented in Chapter 6.

5.2 RESEARCH APPROACH

In order to address these objectives two parallel lab-scale membrane (MBR) and conventional (CAS) activated sludge systems (Figs. 4.1 and 4.2) were operated under laboratory conditions allowing their behaviour to be monitored and their performance compared. In order to verify the previous results of Ramphao et al. (2004) (Chapter 4) the same original experimental apparatus and operational conditions were adopted and testing continued. Both systems were UCT configuration NDBEPR systems, which was chosen because it allowed denitrification and P removal to act independently of each other; the anaerobic reactor is protected from recycled nitrate from the anoxic reactor – provided the recycles do not overload the anoxic reactor with nitrate. As far as possible system design and operational parameters such as zone mass fractions and inter-reactor recycles were kept the same. A summary of the system design and operating parameters is presented in
Table 4.1. The operating, monitoring and testing procedures were the same as in Phase 1 (see Chapter 4, Sections 4.2 to 4.4).

In order to investigate the oxygen transfer efficiencies of the system unsteady state aeration testing and steady state aeration testing were conducted in the aerobic reactor with tap water and mixed liquor at various concentrations respectively. Due to the disruptive nature of these tests this testing was conducted once the main investigation (BNR kinetics tests) had been completed.

5.3 EXPERIMENTAL SYSTEM RESULTS

5.3.1 Steady-State Periods

The steady-state investigation was conducted for 449 days with a total of 29 sewage batch periods. Each sewage batch was accepted as a steady-state period. For every sewage batch data outside the range mean ± 1.96 x sample standard deviation (95% confidence interval), were rejected. All remaining data were considered valid and averaged to represent the “average” response of the system for that sewage batch (steady-state) period. These steady-state averages were used to calculate average ratios of system characteristics.

5.3.2 Mass Balances

Nitrogen and COD mass balances were performed for each sewage batch period in order to verify the accuracy and reliability of the analytical data, and to provide an early warning sign if the data was poor. Good N and COD mass balances were achieved for the MBR system of 96% and 103% respectively. However consistently low mass balances were achieved for the CAS system of 80% and 83% respectively. The causes for the poor mass balances in the CAS system were investigated and were attributed mainly to unaccounted for sludge losses from reactor overflows due to interconnecting pipe blockages.

5.3.3 Mixed Liquor Solids

For all the MBR and CAS system reactors the mixed liquor solids parameters, MLSS, MLVSS, COD and TKN were monitored regularly from the beginning of the investigation. The information on the variation of mixed liquor concentrations with time was necessary in order to interpret the BNR performance of the systems.

5.3.4 Sludge Age (Rs)

As far as possible the sludge age of the system was consistently maintained at 20 days by wasting the appropriate mixed liquor volume from the aerobic reactor. In the event of an unintentional mixed liquor loss from the system due to i) a spill from the reactors, ii) a burst peristaltic pump tube, or iii) foam removal from the anoxic reactor from the high denitrification, knowledge of the total solids content of the system allowed the mass of mixed liquor lost to be determined and approximate mixed liquor mass to be wasted reduced accordingly over the following days.

5.3.5 MLSS and MLVSS concentrations

Throughout the investigation there were sporadic unintended mixed liquor losses, typically through spillages or during the monthly system cleaning. Where ever possible sludge was retained, filtered through a 2 mm mesh to break up sludge “clots” and to prevent the accidental addition of foreign objects into the system, and returned to the system. When sludge was lost the total mixed liquor lost
was calculated from the difference in solids concentration from the day prior to the spill, and the loss compensated for by not wasting for the equivalent number of days following the spill.

The MLSS concentration in the aerobic reactor remained within the range 16 000 to 19 000 mg TSS/ℓ and the anoxic and anaerobic concentrations were within the ranges of 12000-14000 mg TSS/ℓ and 6000-8000 g TSS/ℓ respectively for the MBR system. This is less than the design solids concentration, but is attributed to a lower average $S_{i}$ and higher $f_{S,up}$ than were used for design.

Similarly the CAS system showed lower MLSS concentrations than expected with aerobic mixed liquor solids concentrations consistently within the range 2500-3500 mg TSS/ℓ compared with the expected range 4500-5500 mg TSS/ℓ.

5.3.6 Mixed Liquor Characteristics

In order to quantify the mixed liquor in both systems the VSS, TSS, COD and TKN concentrations of the mixed liquor were measured. Investigation average ratios between these parameters are listed in Table 5.1.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Unit</th>
<th>MBR System</th>
<th>CAS System</th>
</tr>
</thead>
<tbody>
<tr>
<td>VSS/TSS</td>
<td>mg VSS/mg TSS</td>
<td>0.809</td>
<td>0.814</td>
</tr>
<tr>
<td>COD/VSS</td>
<td>mg COD/mg VSS</td>
<td>1.402</td>
<td>1.496</td>
</tr>
<tr>
<td>TKN/VSS</td>
<td>mg N/mg VSS</td>
<td>0.085</td>
<td>0.094</td>
</tr>
</tbody>
</table>

From Table 5.1 the following is noted:

1. Both systems exhibited high VSS/TSS ratios which are not characteristic of BEPR systems. In BEPR systems the development of a PAO population is stimulated. PAOs have a low VSS/TSS ratio (0.46 mg VSS/mg TSS) due to the additional intracellular inorganic polyphosphate.

2. Although the COD/VSS ratios differ substantially in the two systems, both fall close to the expected and theoretical $f_{CV}$ values of 1.48 and 1.42 respectively (WRC, 1984).

3. The COD/VSS ratio indicates that the COD incorporated into sludge solids was lower in the MBR system than in the CAS system. However, comparison of mass balances showed a greater proportion of the influent COD mass was removed from the MBR system via the mixed liquor wasted than for the CAS system. In order for this to occur proportionally more mixed liquor would need to be wasted from the MBR system than the CAS system in order to achieve higher COD removals through wasting, particularly with lower COD incorporated in the sludge mass (VSS). This is only possible, at the same sludge age, if the sludge production in the MBR system is substantially greater than that of the CAS system, which in fact was the case.

5.3.7 Sludge Production

In Chapter 4 it was reported that sludge production in the MBR and CAS systems differed significantly (Ramphao et al., 2004). A number of reasons were suggested, but more data was required to validate these reasons. Hence one of the objectives for this investigation was to validate the observed difference in sludge production in the two systems.

1. With the exception of Sewage Batch 4 (of 29) the sludge production in the MBR system was consistently higher than that of the CAS system by on average 50%.
2. Average sludge productions for the MBR and CAS systems were 0.311 and 0.205 (mg VSS/d)/(mg COD/d) respectively. Ramphao et al. (2004) reported similar results, 0.32 and 0.22 (mg VSS/d)/(mg COD/d) respectively.

3. The higher sludge production can be accommodated in the steady state design and dynamic simulation models by increasing the $f_{s,up}$ fraction. This was demonstrated by the high $f_{s,up}$ values observed in the MBR system of 0.224 in the investigation of Ramphao et al. (2004) (Phase 1) and 0.200 in this investigation (Phase 2).

A number of factors contribute to the higher sludge production in the MBR system, but together these only account for it in part.

1. The retention of solids by membranes in the MBR system resulted in approximately 17.2 mg TSS/accumulating in the MBR system that would have been lost through the SST in the CAS system. This would have “increased” sludge production by 0.018 (mg VSS/d)/(mg COD/d).

2. In the MBR system, organics that would be considered soluble in the CAS system are retained. This is demonstrated by the difference in the filtered effluent COD system averages from the MBR and CAS systems of 10 mg COD (Table 2). This accounts for approximately 0.01 difference in $f_{S,up}$ values above.

3. The higher P removal in the MBR system (Table 2) suggests a larger PAO population which produces more sludge per unit influent COD than OHOs due to their lower endogenous respiration rate (Wentzel et al., 1990).

4. Particulate organics that are biodegradable in the CAS system are no longer biodegradable in the MBR system due to factors such as high MLSS concentrations, or different floc morphology.

In the literature previous studies comparing CAS and MBR BNR systems run under the same operating conditions have indicated that the sludge production of the two systems were very similar (Masse et al., 2006, Monti et al., 2006). However, in both these investigations the systems were run at the same COD loading rate per unit reactor volume and Masse et al., (2006) included the sludge lost through the SST in sludge wasting calculations in order to compare sludge productions by the two systems.

Additionally sludge production in nitrification-denitrification (ND) and ND biological excess phosphorus removal (BEPR) systems operated using the same wastewater source, for sludge ages in the region of 20 days, have produced sludge in comparable magnitudes to those observed in the CAS system ranging from 0.18 to 0.31 (mg VSS/d)/(mg COD/d). Thus it would appear that there is an increase in sludge production in the MBR system linked to the increased MLSS concentration in the MBR system and the retention by membranes of all solids.

5.4 SYSTEM REMOVALS AND EFFLUENT QUALITY

The average removals of both the MBR and CAS systems for this investigation are summarised in Table 5.2.
Table 5.2: Summary of the influent and effluent qualities, and the resultant removals, of both UCT systems.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Units</th>
<th>Influent</th>
<th>MBR System</th>
<th>CAS System</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>Effluent</td>
<td>Efficiency %</td>
</tr>
<tr>
<td>COD</td>
<td>mg COD/l</td>
<td>951</td>
<td>41, 115&lt;sup&gt;a&lt;/sup&gt;</td>
<td>95.7, 88.0&lt;sup&gt;b&lt;/sup&gt;, 92.7&lt;sup&gt;b&lt;/sup&gt;</td>
</tr>
<tr>
<td>TKN</td>
<td>mg N/l</td>
<td>106.5</td>
<td>1.52</td>
<td>98.6</td>
</tr>
<tr>
<td>FSA</td>
<td>mg N/l</td>
<td>81.7</td>
<td>0.74</td>
<td>99.1</td>
</tr>
<tr>
<td>NO&lt;sub&gt;3&lt;/sub&gt;</td>
<td>mg -NO&lt;sub&gt;3&lt;/sub&gt;-N/l</td>
<td>0</td>
<td>16.53</td>
<td>-</td>
</tr>
<tr>
<td>TN</td>
<td>mg N/l</td>
<td>106.5</td>
<td>18.05</td>
<td>83.1</td>
</tr>
<tr>
<td>TP</td>
<td>mg P/l</td>
<td>30.3</td>
<td>8.4</td>
<td>22.5&lt;sup&gt;c&lt;/sup&gt;</td>
</tr>
<tr>
<td>TSS</td>
<td>mg TSS/l</td>
<td>N/A</td>
<td>0</td>
<td>-</td>
</tr>
<tr>
<td>E. coli</td>
<td>CFU/100 ml</td>
<td>N/A</td>
<td>0</td>
<td>-</td>
</tr>
</tbody>
</table>

<sup>a</sup> unfiltered sample; <sup>b</sup> 0.45 filtered sample; <sup>c</sup> P removal mg P/l; N/A = value not available

5.4.1 COD Removals

The COD removal efficiency of the MBR system (96%) was superior to that of the CAS system (92% unfiltered, 95% 0.45 µm filtered). These results were comparable to those observed by Ramphao et al. (2004) of 96% COD removal in the MBR UCT system and 93% unfiltered and 94% 0.45 µm filtered in the CAS system.

The difference in filtered COD removals from both systems is attributed to the smaller pore size of the membranes which retain organics that would otherwise be considered soluble in a CAS system. However membrane specifications state that the nominal pore size of the Kubota membranes used in this study were 0.4 µm, while the paper membranes used to filter the conventional effluent are only marginally larger 0.45 µm, thus the improved filterability of the membrane system is attributed rather to the development of a dynamic gel layer which reduces the effective pore size of the membranes.

The MBR unfiltered COD values (measured at the 800 ml mark after 30 min settling in the 1000 ml DSVI test on MBR aerobic sludge) were consistently higher than those in the CAS system which confirmed that the MBR system retains and accumulates unsettleable material which would flow out with the effluent in a CAS system. This was observed in the DSVI test in which the supernatant of the CAS system mixed liquor would become clear in time, whereas the MBR supernatant remained turbid.

The difference between the filtered (50 mg COD/l) and unfiltered (75 mg COD/l) effluent COD measured in the CAS system is attributed to the loss of non-settleable solids through the SST. Approximately 21.5 mg TSS/l were lost as COD in the effluent.

Following Ramphao et al. (2004) (Chapter 4), differences in the MBR and CAS effluent COD concentrations are accommodated in the steady state design and kinetic simulation models as differences in the influent soluble COD fractions (f<sub>s,us</sub>) which were 0.044 and 0.068 respectively.

5.4.2 N Removals

The TKN removal efficiency of the MBR system (98%) was marginally better than that of the CAS system (97% unfiltered, 98% 0.45 µm filtered). This is again attributed to the retention of solids by the membranes that would have been lost in the effluent of the CAS system. FSA removal was also
very similar for both systems 99% in the MBR system and 98% in the CAS system. Thus near complete nitrification was achieved in both systems.

Effluent nitrate concentrations were virtually the same for both systems (18.0 and 18.1 mg NO₃-N/ℓ in the MBR and CAS systems respectively) resulting in similar total nitrogen (TN) removals, (81.5% for the MBR system versus 79.5% for the CAS system).

Nitrogen is removed from BNR systems either by incorporation of nitrogen in mixed liquor and its subsequent removal through wasting, or through nitrification/ denitrification.

- The influent N incorporated in the mixed liquor was lower in the MBR UCT system than in the CAS system. This corresponds to the observation above of lower COD incorporated in the MLVSS.
- Nevertheless, N removal through sludge wasting was higher in the MBR system than the CAS system, which was due to the higher sludge production in the MBR system, and consequent increased sludge mass wasted per unit influent N.
- The MBR system achieved a higher N removal through denitrification even though the anoxic mass fraction in the two systems were the same. The reason for this is the negligible effect of the DO in the recycle flow to the anoxic reactor at high TSS concentration.

5.4.3 P Removals

In both systems TP was dosed in excess of the amount the system could remove in order to demonstrate the upper limit of BEPR. Thus P removal performance is represented by P removals. System average P removals of 21.3 mg P/ℓ and 16.7 mg P/ℓ were achieved. Clearly, the P removal performance of the CAS system was lower than that of the MBR system.

The anoxic P uptake was more prevalent in the CAS system with 22.1% of P uptake taking place in the anoxic reactor, in contrast to only 8.5% anoxic P uptake in the MBR system. Additionally the conventional anoxic reactor was regularly overloaded with NO₃ as evidenced by consistent anoxic NO₃ concentrations >1 mg NO₃-N/ℓ.

The above observations indicate that the CAS system was not operated optimally. Low MLSS concentrations, as reported earlier reduced the denitrification potential of the anoxic reactor, thus allowing NO₃ to be recycled to the anaerobic reactor. Hence, as the CAS system was not performing optimally and exhibited a greater anoxic P uptake BEPR, a comparison cannot be made between the two systems on whether or not the presence of membranes changed the P removal efficiency of the MBR system. It could be argued that the membranes did stimulate a higher BEPR because this is what was observed. However, if the a-recycle ratio on the MBR system were increased to overload the anoxic reactor with nitrate like the CAS system anoxic reactor was, the MBR system would also have exhibited greater anoxic P uptake BEPR and hence lower P removal like the CAS system. The difference in BEPR behaviour was therefore a consequence of a difference in denitrification in the anoxic reactor.

5.4.4 Coliform Removal

Periodic effluent samples were tested from both systems for the indicator micro-organism e-coli using the membrane filtration method. Results indicated pathogen counts were unobservable in the MBR UCT system whereas in the CAS system pathogen counts ranged from 580 to 5600 CFU/100 ml.
Clearly, the MBR UCT system produced an effluent that was equal, if not superior in quality to the CAS system. Due to complete retention of solids, and pathogens, the membrane effluent is more viable for reuse purposes.

5.5 OXYGEN TRANSFER TESTING

Currently, one of the most important considerations for the design of MBR plants is the feasibility of running the systems at high solids concentrations. In previous literature it has been noted that the oxygen transfer efficiency (OTE) of systems decreases substantially (by reducing the oxygen mass transfer coefficient $K_{La}$ relative to tap water $\alpha$) at high solids concentrations (Cornel et al., 2003, Krampe and Krauth, 2003).

Once the intensive study on BNR performance in the MBR system had been completed, the oxygen transfer efficiencies of the activated sludge at high MLSS concentrations were determined by performing tests on the sludge in the aerobic reactor. $K_{La}$ values were determined for a number of air flow rates with tap water in the operating range of the aeration system. Steady state tests were then carried out on the activated sludge once it had reached endogenous conditions (to reduce the interference by differences caused by variations in feed). For a single MLSS concentration a number of readings were taken at different airflow rates and the alpha values observed were averaged. Alpha values of 0.17-0.28 (21000 mg TSS/ℓ), 0.38-0.68 (17000 mg TSS/ℓ) and 0.53-0.80 (11000 mg TSS/ℓ) were observed.

The observed Alpha values varied over a wide range at each TSS concentration. The reason for this is the sensitivity of the alpha value to the measured OUR, DO concentration ($C_L$) and airflow rate. As the system was run at along sludge age under endogenous conditions, the OUR was low which resulted in high DO concentrations making the difference ($C_{Sat}-C_L$) a difference of two large numbers which varied considerably. In order to compensate for high $C_L$ values, only low airflows were selected, at which the air flow rotameter was less accurate. Possibly the measure system was accurate but imprecise so the average a number of readings at the same MLSS concentrations yields a reasonably accurate Alpha value. In future such tests should be run on sludge with a short sludge age which has a high active fraction and hence an higher endogenous OUR. Also, it should be noted that a number of factors affect the oxygen transfer. The geometry of and mixing in the reactor and the characteristics of the aeration (coarse bubble aeration) have a considerable influence on the Alpha value. Thus the results measured in this investigation must be interpreted with care. Further studies in conditions resembling full scale systems are required to accurately calculate Alpha and OTE for design.

5.6 MEMBRANE PERFORMANCE

Due to the constant flow and load conditions, the membrane flux was constant and the TMP was controlled to maintain the flux [0.24 m$^3$/m$^2$.h)]. Because the volume of the MBR system aerobic reactor needed to be constant to not change its VSS concentration, the external U-bend over which the effluent exited the aerobic reactor was lowered below the reactor ML level to increase the TMP as required. From day 50 to 350, the TMP increased from around 20 to 80 mm (0.2 mm/d). While the membranes were rinsed under running water every month, no chemical cleans were applied. From day 350 to 430 the TMP increased to 200 mm (1.5 mm/d) caused by inorganic colloidal solids from construction work at the WWTP, which also increased the influent and reactor ISS concentrations. Possibly the presence of unbiodegradable colloidal material in the mixed liquor adversely influenced the TMP, which was observed by Fleischer et al. (2005). On completion of the investigation a full chemical clean was performed on the membranes using a 1.0% hypochlorite solution. The hypochlorite was drawn through the membranes by applying a vacuum (negative) pressure on the effluent line of the membranes. Within 24 hour of the chemical clean the membrane
TMP had returned to its original TMP at the start of the investigation (55 mm water). This indicated that the increase of TMP was reversible with effective cleaning of the membranes.

5.7 OTHER BNR MBR INVESTIGATIONS

In this investigation 6 case studies are reported from the literature in which various BNR configurations using membranes were proposed and investigated:

- Monti et al. (2006) compared two AS systems in a UCT configuration, one a MBR and the other using a conventional SST, with the same system design and operational parameters. Both systems had the same sludge mass and hence the influence of sludge concentration on BNR performance was not assessed.

- Lesjean et al. (2003, 2005) conducted a 4-year study on pre-denitrification and post-denitrification configurations of MBR wastewater treatment systems at bench and pilot scale. The systems were run at varying sludge ages and high solids concentrations. Excellent nutrient removal was observed in both configurations without additional carbon dosing in the post-denitrification system. However the systems were found to be generally underloaded, and precipitation of P due to calcium and ferric ions was observed which compromised observations of biological P removal.

- Ahn et al. (2003) operated two lab scale MBR systems as a sequencing anaerobic/anoxic MBR (SAM) and the other a MLE system. P removal was observed in the SAM system, but at the cost of poor N removal. BNR removal was observed but not optimised.

- Mouthon-Bello and Zhou (2005) conducted a study on a submerged MBR in an anoxic-anaerobic-aerobic configuration at 20 and 50 day sludge ages. Alum was dosed to the aerobic reactor to aid P removal. This configuration made the anaerobic reactor redundant and biological P removal could not be observed.

- Fleisher et al. (2005) investigated the BNR performance of an MBR system in a 5-stage configuration in order to ascertain whether biological and chemical P removal could be achieved concurrently. They successfully demonstrated that BEPR could be achieved, using chemical precipitation in the MBR reactor to completely remove all remaining P. The 5-stage configuration was also successful in reducing TN to <3 mg N/ℓ. In addition Fleischer et al. (2005) modelled the observed system performance and suggested that current simulations (IWA ASM2d) adequately predicted the BNR performance of the system. Lastly they investigated the solids produced from the membrane system in order to determine if they differed from conventional solids and observed that a higher density cake could be produced from the MBR sludge than from conventional sludge.

- Ramphao et al. (2004) (Chapter 4) investigated the BNR performance of two systems in UCT configurations. In contrast to the study by Monti et al. (2006) the systems were run at their design solids concentrations, i.e. aerobic solids concentrations were 4500 mg TSS/ℓ in the CAS system compared to 18000 mg TSS/ℓ in the MBR system. The MBR system produced an effluent that was consistently equal to, or better than, the conventional effluent. It was found that the current BNR simulations could adequately predict system performance, but solids production in the MBR system was substantially higher than expected.

The research has shown that the inclusion of membranes in the system does not adversely affect the BNR performance, and also that at high concentration sludges, as are characteristic of MBR systems, the BNR performance remains consistent. However these studies have only indicated that MBRs are feasible and have not investigated in depth how the performance is affected – notably how to optimise BNR in MBR systems. The studies have demonstrated the inability to compare systems without the kinetic constants for modelling being established, as each investigation is on
different wastewaters and serves different BNR objectives. Additionally information important to
design such as the oxygen transfer efficiency in high solids concentration sludges remains much

5.8 CONCLUSIONS

From this investigation, the following conclusions were made:

1. Membranes in a BNR system are a feasible nutrient removal solution with excellent organic
and nutrient removal performance. The presence of membranes and consequently operating
the system at high sludge concentrations did not adversely affect BNR performance, but
produced an effluent of equal or superior quality to that produced by a CAS system using
SSTs. In addition pathogen counts indicated that all pathogens were retained by the
membranes. Thus the membrane effluent is safer and more viable for reuse purposes.

2. Higher sludge productions of 0.311 and 0.320 (mg VSS/d)/(mg COD/d) were observed in
the MBR system in both this investigation and by Ramphao et al. (2004). This higher sludge
production is accommodated in steady state design theory by increasing the unbiodegradable
particulate COD fraction ($f_{S,up}$) to 0.200 in this investigation and 0.224 from Ramphao et al.
(2004). The increased sludge production in the MBR is justified in part by the retention of
all solids by the membranes. Similarly the unbiodegradable soluble COD fraction ($f_{S,us}$)
must be decreased to account for the additional retention of “soluble” COD which is
attributed to the finer membrane pore size.

3. A theoretical evaluation of the BNR performance of the MBR system indicated that the
current steady state BNR theory was able to closely predict the system performance for
COD removal and nitrification. However for denitrification the DPP was under predicted
requiring $K_2’T$ to be adjusted from 0.145 to 0.216 mg N/mg VSS/d at 20°C in order to
match observed and predicted values. The BEPR predictions were close to those observed
when the system PAO population reached a steady state (sewage batches 18-25). $f_{XBGP}$
observed in this period (0.376 mg P/mg VSS) was close to that determined theoretically of
0.38 mg P/mg VSS (Wentzel et al., 1990).

4. Aeration testing was performed on the system, in order to determine alpha values for the
high concentration sludge. Alpha values of 0.5-0.6 for ~15 000 mg TSS/ and 0.2-0.3 for
~20 000 mg TSS/ were determined, which are higher than other values reported in the
literature. These values are however specific to the laboratory system run in which factors
such as reactor geometry and high aeration turbulence would have affected oxygen transfer
in the system. Additionally the low sensitivity of the measuring apparatus resulted in
substantial variance of results.

5. The permeability of the membranes can be influenced by fine colloidal material, however
observations indicated that the permeability returns to previous levels once colloids are
removed from solution by assimilation into the mixed liquor.

Hence the combination of membranes in BNR AS is being increasingly applied. Membrane
applications are becoming common in Europe, North America and Asia where much research has
been conducted on the performance of membranes, typically in aerobic systems in order to
determine their life spans, mechanisms of fouling etc. Additionally numerous studies have reported
excellent COD removal and nitrification performances, however few case studies investigating
BNR using membranes have been reported despite speculation that the inclusion of membranes may
indeed affect the nature of the activated sludge biomass (Witzig et al., 2002).
5.9 RECOMMENDATIONS FOR FURTHER STUDY

It is strongly recommended that accurate knowledge of the oxygen transfer rate in high concentration sludge is an important design consideration. However in this investigation the difficulty in measuring this parameter at a lab scale was highlighted. It is recommended that aeration system parameters need to be determined at full scale at a fully operational BNR MBR WWTP.

5.10 REFERENCES


CHAPTER 6

THE IMPACT OF HIGH MIXED LIQUOR CONCENTRATION (3-13 g VSS/ℓ) ON THE KINETICS OF N AND P REMOVAL IN BIOLOGICAL NUTRIENT REMOVAL ACTIVATED SLUDGE SYSTEMS.

6.1 INTRODUCTION

For conventional (with settling tanks) activated sludge (CAS) systems for biological nutrient removal (BNR), considerable knowledge has been accumulated on their performance, design and operation. Design procedures and performance simulation models have been developed based on well structured and researched stoichiometric and kinetic principles of the underlying fundamental biologically mediated processes. It is not certain whether this knowledge developed for CAS BNR systems can be applied directly to membrane bioreactor (MBR) BNR systems, given the significant differences that may arise when membranes are included such as (i) floc structure (Zhang et al., 1997; Cicek et al., 1999; Huang et al., 2001; Yamamoto, 2002; Gao et al., 2004; Manser et al., 2005a), (ii) bacterial communities (Ghyoot et al., 1999; Luxmy et al., 2000; Liebig et al., 2001; Smith et al., 2002; Manser et al., 2005b), (iii) metabolic activities (Witzig et al., 2002; Rosenberger et al., 2002; Lee et al., 2003; Han et al., 2005; Sperandio et al., 2005; Li et al., 2005) and (iv) sludge production (Cicek et al., 1999; Smith et al., 2002; Holbrook et al., 2005; Monti et al., 2005).

Ramphao et al. (2005) concluded that incorporating membranes in BNR AS systems makes a profound difference not only to the design of the BNR system itself, but also to the approach to design of the whole wastewater treatment plant (Chapter 3). This Chapter presents the investigation into whether or not the steady state and kinetic models developed for CAS BNR systems can be applied also with reasonable accuracy to model MBR BNR systems.

Accordingly, the kinetic rates of nitrification, denitrification, anaerobic acetate uptake and P release, anoxic P release/uptake and aerobic P uptake were measured in batch tests over a range of volatile suspended solids (VSS) concentrations (3-13 g VSS/ℓ) on sludge harvested from an MBR BNR system and compared with the corresponding rates measured in a parallel CAS BNR system at 3 g VSS/ℓ. Also, the influence of the limitation of substrate (ammonia, oxygen, nitrate, phosphorus, acetic acid) concentrations on the kinetic rates was investigated in the batch tests. To provide additional information on the anoxic behaviour of phosphate accumulating organisms (PAO), the ability of the PAOs in MBR BNR systems to denitrify under anoxic conditions with simultaneous phosphate uptake was investigated and quantified.

6.2 MATERIAL AND METHODS

Two parallel lab-scale membrane (MBR) and conventional (CAS) activated sludge systems were operated for 450 days allowing their behaviour to be monitored and their performance compared. Their design, operation, monitoring and performance were presented in Chapters 4 and 5. To determine the kinetics rates, batch tests on the mixed liquor harvested from the two BNR systems were conducted (Parco, 2006; Parco et al., 2006, 2007). Particularly on the MBR system the influence of the VSS concentration and of the limitation of ammonia, oxygen, nitrate and acetate concentrations on the kinetic rates was examined.
6.3 Overall MBR and CAS system N and P REMOVAL performance

The characteristics of the source sludge used in the batch tests affect the observed batch test behaviour. Consequently, to give some insight into overall N and P removal behaviour of the two systems, the average N and P removals of the two systems during the time the batch tests were conducted are briefly discussed below (Phase 2, details in Chapter 5).

6.3.1 N and COD mass balances

Nitrogen mass balances were calculated from the wastewater batch average results by comparing the exiting N via the effluent, waste sludge stream and nitrate denitrified (from a nitrate balance over the anoxic and anaerobic reactors) with the N entering the systems via the influent TKN. N balances for the MBR system ranged between 75 and 120% with an overall average of 96%. Similarly, the COD balance was calculated by comparing the exiting COD via the effluent, waste sludge stream and oxygen utilised in the aerobic reactor (corrected for nitrification) with the COD entering the systems via the influent COD. COD balances for the MBR system ranged between 89 and 107% with an overall average of 103%. These mass balances are good and validate that the measured results are reliable. In fact, the COD balances over the MBR system are among the best achieved in NDBEPR systems at UCT (Ekama and Wentzel, 1999), possibly as a result of the membranes.

6.3.2 Biological nitrogen removal

The TKN removal efficiency of the MBR system was marginally better than that of the CAS system (Table 5.2). This is attributed to the retention of solids by the membranes compared with that lost with the effluent from the CAS system. In both systems, nitrification was virtually complete, as indicated by the low residual free and saline ammonia (FSA) concentrations. The two systems achieved similar total N removals indicated by the similar effluent nitrate concentrations.

6.3.3 Biological phosphorus removal

In both systems, TP was dosed in excess of the system removal capacity in order to observe the maximum BEPR possible. Thus P removal performance is represented by total P (TP) removals. System average P removals of 22.5 mg P/ℓ and 17.4 mg P/ℓ were achieved in the MBR and CAS systems showing that the P removal performance of the CAS was inferior to that of the MBR system. Reasons for this are: (1) Anoxic P uptake was higher in the CAS system at 22% of total P uptake in the anoxic reactor compared with the MBR system at 8%. This was also evident in the P removal anoxic batch tests results. With anoxic P uptake by denitrifying PAOs, significantly reduced BEPR has been reported (Ekama and Wentzel, 1999; Hu et al., 2002) probably due to less efficient utilisation of stored polyhydroxyalkanoates (PHA) by PAOs when nitrate serves as electron acceptor (Hu et al., 2002, 2007a,b). (2) The nitrate load on the anoxic reactor of the MBR system was lower than the reactor’s denitrification potential identified by zero nitrate concentration in this reactor. In contrast, denitrification was not complete in the anoxic reactor of the CAS system, identified by > 1 mg NO3-N/ℓ nitrate concentration in this reactor. This is one of the main factors that stimulates anoxic P uptake (Ekama and Wentzel, 1999; Hu et al., 2002). For the CAS system, the recycle of nitrate to the anaerobic reactor from the anoxic reactor reduced the RBO available for PAOs and additionally reduced the P removal. The main reason for the difference in denitrification capacity of the MBR and CAS anoxic reactors was the dissolved oxygen (DO) in the recycles, which had a negligible effect on denitrification at the very high VSS concentration in the MBR system.
6.4 CALCULATING THE BIOPROCESS SPECIFIC KINETIC RATES

In the steady-state design procedures and dynamic kinetic models, the increased sludge production in MBR systems can be accommodated by increasing the influent unbiodegradable particulate COD fraction \( (f_{S_{up}}) \). This was done in this investigation. Fixing the unbiodegradable soluble COD fraction \( (f_{S_{us}}) \) for the MBR and CAS systems at the measures values, i.e. 0.045 and 0.066 respectively, the \( f_{S_{up}} \) fraction for the MBR and CAS systems were calculated to be 0.241 and 0.084 mg COD/mg COD respectively to match the measured average mass of VSS in the systems (Ekama and Wentzel, 1999). Noting that the model takes account of the different masses of PAOs in the two systems, it is a concern that for two systems with the same design and operating parameters fed the same wastewater, different \( f_{S_{up}} \) fractions are obtained. If \( f_{S_{up}} \) is really a wastewater characteristic, \( f_{S_{up}} \) should be the same for both systems. However, higher \( f_{S_{up}} \) in NDBEPR systems than in ND systems fed the same wastewater has been consistently observed in the past (Ekama and Wentzel, 1999). The problem of obtaining different \( f_{S_{up}} \) fractions for the MBR and CAS systems, is that they result in different OHO \( (f_{avOHO}) \) and PAO \( (f_{avPAO}) \) biomass fractions of the VSS in the systems, where \( f_{avOHO} = \frac{X_{BH}}{X_v} \) and \( f_{avPAO} = \frac{X_{BG}}{X_v} \) and \( X_{BH}, X_{BG} \) and \( X_v \) are the OHO, PAO and total VSS concentrations respectively. However, the method of calculating \( f_{S_{up}} \) by matching the calculated mass of VSS in the system with that measured has always has been applied in the past to determine the \( f_{avOHO} \) and \( f_{avPAO} \) active fractions and the OHO and PAO specific kinetic rates (van Haandel et al., 1981; Wentzel et al., 1990; Clayton et al., 1991; Ekama and Wentzel, 1999) and these specific rates have become adopted as default values in the ASM1 and ASM2 kinetic models. So because there is no other way of determining biomass specific kinetic rates from experimental systems fed real wastewater, the uncertainty that different \( f_{S_{up}} \) fractions will have on the kinetic rates, while not ideal, has to be accepted as it has been in the past (Ekama and Wentzel, 1999) because expressing kinetic rates in terms of VSS makes the rates incomparable between different BNR systems. In the end, steady state models aligned with and based on the same but simplified principles as kinetic models are the only interface between experimental systems and the kinetic models.

Because the kinetic rates determined from the batch tests results were assigned to the biomass population mediating the particular bioprocess, and the steady state NDBEPR model (Wentzel et al., 1990) was used to determine the OHO \( (f_{avOHO}) \) and PAO \( (f_{avPAO}) \) active fractions from the measured data on the MBR and CAS systems, it was important for the OHO specific denitrification rate and the PAO specific P release and P uptake rates that the observed and predicted P removal of the systems matched well. This ensured that the OHO and PAO specific kinetic rates were consistent with estimates of the OHO \( (f_{avOHO}) \) and PAO \( (f_{avPAO}) \) active fractions determined in the past. Figure 6.1 shows the wastewater batch average observed and predicted P removal of the MBR system, where the latter is based on the known system operating parameters, dosed acetate (200 mg/l) and measured wastewater RBO concentration (Chapter 5). The nitrification batch tests, for which a close correlation between predicted and measured P removal was not important, were conducted at the beginning of the investigation when the predicted and measured P removal did not match well. The denitrification (anoxic) and P release and P uptake (anaerobic-anoxic/aerobic) batch tests were conducted during wastewater batches 10 to 25, when the predicted and measured P removal did match well. The measured kinetic rates in the MBR and CAS systems can therefore be legitimately compared with rates measured in previous investigations.
Figure 6.1: Wastewater Batch average measured and predicted P removal for the MBR system. The denitrification, P release and P uptake batch tests were conducted between during Wastewater Batch 10 to 25 when the predicted and measured P removal matched well.

6.5 Nitrification Kinetics – Aerobic Batch Tests

6.5.1 Test and calculation procedures

The batch tests were conducted in two parallel reactors constructed of clear cylindrical Perspex (~ 5 mm thick) with volume of 3 ℓ and diameter of 120 mm. The contents of the reactors were completely mixed by means of a motor driven paddle mixer, mounted centrally on the lid of the reactor. To ensure good mixing, each reactor was fitted with a pair of vertical side wall baffles situated opposite each other and extending about 2 cm radially into the bulk solution. The aerobic reactor had two paddles (one with inclined vanes) fitted to the mixing shaft, one situated on the bottom and one 2 cm above it. These were positioned to ensure complete suspension of the reactor contents while avoiding turbulence at the liquid surface, to minimise air entrainment into the bulk solution.

The lid of the reactor had a semicircular opening for retrieval of samples and insertion of pH and dissolved oxygen (DO) probes into the bulk liquid. Each reactor had a single inlet and outlet situated on its bottom plate. Aeration was provided by humidified (by bubbling through a large vessel of water) low pressure compressed air that entered through a small Perspex tube which terminated in a porous stone at the bottom of the reactor vessel. The air flowrate was regulated by throttling a ball valve on the main air supply line and adjusting a hose-clamp on the tube entering reactor. The typical set-up configured for an aerobic batch test is shown in Fig. 6.2.

During the batch test, the inside walls of the reactor were routinely brushed to prevent mixed liquor particles from adhering to the reactor walls above the liquid surface. The batch tests were conducted on sludge harvested from aerobic reactor of the parent MBR and CAS systems.

At the start of an aerobic batch test, measured volumes of sludge from the aerobic reactors were transferred to the batch reactors. Since endogenous conditions were to be created, initially the mixed liquor was aerated for a long period (at least 12 hours) in absence of substrate. This ensured complete degradation of all residual influent biodegradable organics coming from the aerobic reactor of the parent systems and hence the activity of heterotrophic bacteria with oxygen consumption for the substrate oxidation was minimised.
Once the initial test conditions had been established (absence of biodegradable COD), ammonia was added directly to the reactor as a small volume (65-175 ml) of concentrated ammonium chloride solution (5 g NH₄Cl/ℓ) and the OUR determined until endogenous conditions were re-established. The difference between the total OUR and the OUR related to endogenous respiration was considered the oxygen utilised (OU) due to NH₄-N oxidation. This could be verified from the amount of ammonia added for each batch test.

During the test, grab samples were taken manually at specific time intervals; the sampling intervals (in minutes) per hourly period during the test are summarised in Table 6.1. Samples were taken also at the beginning of the endogenous phase, before adding ammonia, and when the respiration rate returned to the initial rates of endogenous respiration (low value of OUR), after 6-12 hours from the start of the nitrification process. Samples (50 ml) were taken from the batch reactor and immediately centrifuged and filtered (0.45 µm) after which 2 drops of 8.6 g/ℓ HgCl₂ were added to the filtrate. The samples were normally stored overnight at 4°C and analyzed the following day for NO₃-N, NO₂-N, PO₄-P, N-NH₄, and soluble COD. In addition 2 mixed liquor samples (50 ml) at the beginning and at the end of the batch tests were also taken for VSS and TSS measurements. All analyses were performed according to Standard Method (1985).

<table>
<thead>
<tr>
<th>Sample Interval (min)</th>
<th>Test Period aerobic condition (h)</th>
</tr>
</thead>
<tbody>
<tr>
<td>15</td>
<td>0-1</td>
</tr>
<tr>
<td>30</td>
<td>1-9</td>
</tr>
</tbody>
</table>

The DO concentration, and concomitantly oxygen utilisation rate (OUR), was measured continuously and automatically using a YSI model 5739 DO probe and HiTech Microsystem DO meter that controlled reactor aeration between high and low DO set points via a solenoid valve on the air line (Randall et al., 1991). When the DO concentration in the mixed liquor reached the low point set point (2 mg O/ℓ), the solenoid valve was opened and the reactor contents aerated. When the DO in the bulk solution reached the high set point (5 mg O/ℓ) the air was switched off automatically and the decrease in DO with time monitored. When the DO reached the low set point again, the aeration was recommenced and the cycle repeated. During each air-off period in the cycle, the slope of the DO-time data was automatically calculated by linear regression to give the
OUR at that time, which, together with regression correlation coefficient, temperature and time was stored by the DO/OUR meter. The OUR results for each conducted test were downloaded from the DO/OUR meter to a PC. The data were imported into a spreadsheet program to analyze and plot the results. The DO/OUR meter and the probe were calibrated for each batch test.

The difference between the total OUR and the OUR associated with endogenous respiration was accepted as the OUR due to ammonia (NH₄-N) nitrification (OURₙᵢᵗᵣ) (Figure 6.3). The oxygen utilised (OU) for nitrification (area under the OURₙᵢᵗᵣ – time curve) matched stoichiometrically to the concentration of ammonia added to the batch test through OU/4.57= ammonia concentration added.

Three groups of aerobic nitrification batch tests were conducted to evaluate the effect of VSS, ammonia and dissolved oxygen (DO) concentration on the nitrification kinetics in the MBR system: Group (1), i.e. ten with 10-20 mg N-NH₄/ℓ, twelve with 30-40 mg N-NH₄/ℓ and seven with 50 mg N-NH₄/ℓ on MBR system ML diluted (with effluent) to different VSS concentrations between 2 and 14 g VSS/ℓ, i.e. eight with 2-3 g VSS/ℓ, two with ~4 g VSS/ℓ, six with ~ 5-6 g VSS/ℓ, five with 7-10 g VSS/ℓ, two with 10-11 g VSS/ℓ and five with 12-14 g VSS/ℓ on MBR system ML, 2 Group (2), i.e. two on MBR system ML at the same VSS concentration (~9 g VSS/ℓ) but at different DO concentrations, 2-5 and 10-15 mg O/ℓ and six Group (3), i.e. in parallel, three on each of MBR and CAS system ML with MBR ML diluted to the same low VSS concentration as that from the CAS system (2-3 g VSS/ℓ) to determine the effect of the membranes.

For all the batch tests, the nitrification, or ammonia utilisation rate [AUR, mg NH₄-N/(ℓ.h)] was calculated from the OUR profile (Fig. 6.3) as the average maximum value of OURₙᵢᵗᵣ and from the slope of the FSA concentration – time profile (Fig. 6.4). The AUR determined from the two methods were found to be in close agreement for all the nitrification batch tests. The VSS specific nitrification or ammonia utilisation rate (SAUR, mg NH₄-N/(g VSS.h)) was defined as AUR/Xᵥ where Xᵥ was the measured VSS concentration in the batch test.

To quantify the maximum specific growth rate of autotrophic nitrifying organisms (ANOs, µₐ), nitrification was considered as a single composite bioprocess, where NH₄-N to NO₂-N oxidation limits the overall transformation of NH₄-N to NO₃-N, as observed under most circumstances in municipal wastewater treatment plants and generally accepted in steady state and kinetic models for
nitrification (WRC, 1984; Henze et al., 1987, 1995). Consequently, the nitrification kinetics were expressed as:

\[
\frac{dNH_4-N}{dt} = \mu_A \frac{NH_4-N}{Y_A K_n + NH_4-N} X_{BA} \tag{6.1}
\]

where

\[
\frac{dNH_4-N}{dt} = \text{ammonia utilisation rate AUR – mg NH}_4\text{-N/(ℓ.h)}
\]

\(\mu_A = \text{maximum specific growth rate of ANOs – mg ANOVSS/(mg ANOVSS.d)}\)

\(Y_A = \text{yield coefficient of ANOs – mg ANOVSS/mg NH}_4\text{-N}\)

\(NH_4\text{-N} = \text{ammonia concentration (mg N/ℓ)}\)

\(K_n = \text{ANO Monod half saturation coefficient for ammonia (mg NH}_4\text{-N/ℓ)}\)

\(X_{BA} = \text{ANO biomass concentration (mg ANOVSS/ℓ)}\)

Stoichiometrically the oxygen requirements for the conversion of ammonia to nitrate is 4.57 mg O/mg NH\(_4\)-N nitrified. Hence, the oxygen consumption rate for nitrification is:

\[
\frac{dO}{dt} = 4.57 \left[ \frac{\mu_A \frac{NH_4-N}{Y_A K_n + NH_4-N} X_{BA}}{X_{BA}} \right] \tag{6.2}
\]

where

\(\frac{dO}{dt} = \text{oxygen utilisation rate (OUR) for nitrification [mg O/(ℓ.h)]}\)

From Eq. 6.1, to determine \(\mu_A\) from the batch test data, the concentration of ANO in the ML (\(X_{BA}\)) needs to be known. From the nitrogen mass balances over the parent systems, the fluxes of nitrate produced in the MBR and CAS systems were calculated. From these nitrate fluxes, the masses of nitrifiers in the systems were calculated (WRC, 1984, Henze et al., 2008), which requires the yield coefficient \((Y_A)\) and endogenous respiration rate \((b_A)\) of the ANOs to be defined. The values accepted were \(Y_A=0.1\) mg ANOVSS/mg NH\(_4\)-N and \(b_A=0.04 /d\) (WRC, 1984). The mass of ANOs divided by the measured mass of VSS in the systems gives the ANO VSS biomass fraction \((f_{ANOVSS})\), which were 0.0127 and 0.0187 mg ANOVSS/mg VSS for the MBR and CAS systems respectively. These fractions remained approximately constant due to the approximately constant influent TKN/COD concentration ratio fed to the parent systems. With \(X_{BA}\) known from \(f_{ANOVSS}X_v\), the \(\mu_A\) values were calculated. Two methods were applied: (1) the slope of the linear portion of the NH\(_4\)-N concentration – time profile (Fig. 6.4), i.e. AUR, was determined by linear regression. Then because the NH\(_4\)-N concentration is much larger than the \(K_n\) for most of the ammonia-time profile, \(\mu_A \approx Y_A AUR/(f_{ANOVSS}X_v)\), where \(X_v\) is the measured VSS in the batch test. (2) From the nitrification OUR (OUR\(_{Nit}\)) – time profile, the AUR versus time was calculated from \(OUR_{Nit}/4.57 = dNH_4-N/dt = AURt\). This AUR\(_t\) was plotted versus the calculated NH\(_4\)-N concentration at time \(t\), which was calculated from \([NH_4-N]_t = [NH_4-N]_{t-1} – AURt \times \Delta t\), using the initial ammonia concentration as the starting ammonia concentration. Using Eq. 6.1, the values for the Monod \(\mu_A\) and \(K_n\) that gave the least squares best fit to the AUR\(_t\) – time plot (Fig. 6.5) were accepted. A probability plot of the \(K_n\) values obtained with this method from the different batch tests is given in Figure 6.6. The average \(K_n\) value was 1.9 mg NH\(_4\)-N/ℓ. This \(K_n\) value is nearly 100% higher than that usually accepted for CAS systems, i.e. \(K_n=1.0\) mg NH\(_4\)-N/ℓ (WRC, 1984; ASM1, Henze et al., 1987). The \(\mu_A\) values obtained from the two methods were close in agreement for all batch tests. The results of the batch tests are listed in Table 6.3. In Table 6.2, the \(\mu_A\) are grouped in VSS concentration and initial ammonia concentration ranges.
### Table 6.2: MBR system $\mu_A$ (/d) ranges at different VSS and initial ammonia concentration ranges.

<table>
<thead>
<tr>
<th>VSS concentration (mg VSS/l)</th>
<th>NH$_4$-N added (mg N/l)</th>
<th>10-20</th>
<th>30-40</th>
<th>50</th>
<th>70</th>
</tr>
</thead>
<tbody>
<tr>
<td>2000-3000</td>
<td></td>
<td>0.342</td>
<td>0.379-0.466</td>
<td>0.532-0.568</td>
<td></td>
</tr>
<tr>
<td>3000-5000</td>
<td></td>
<td>0.249-0.311</td>
<td></td>
<td>0.476</td>
<td></td>
</tr>
<tr>
<td>5000-7000</td>
<td></td>
<td>0.198-0.251</td>
<td>0.310-0.393</td>
<td>0.406-0.446</td>
<td></td>
</tr>
<tr>
<td>&gt; 7000</td>
<td></td>
<td>0.183-0.199</td>
<td>0.246-0.366</td>
<td>0.332-0.428</td>
<td>0.582-0.521</td>
</tr>
</tbody>
</table>

Figure 6.5: A typical ammonia utilisation rate (AUR) versus the NH$_4$-N concentration for aerobic batch test on mixed liquor from the CAS system.
Table 6.3: VSS specific ammonia utilisation rate (SAUR), maximum specific growth rate ($\mu_n$) and half saturation coefficient ($K_n$) of ammonia oxidizing organisms.

<table>
<thead>
<tr>
<th>Batch test</th>
<th>NH4-N added</th>
<th>SSV</th>
<th>NH4-N load</th>
<th>FSA profile</th>
<th>OUR profile</th>
<th>NH4-N measured</th>
<th>NH4-N measured from OUR</th>
<th>MXn</th>
<th>model kinetic</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>mg/l</td>
<td>mg/l</td>
<td>mgNH4-N/SSV</td>
<td>mg/l</td>
<td>mg/l</td>
<td>mgNH4-N/SSV</td>
<td>mg/l</td>
<td>mg/l</td>
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<td>29</td>
<td>13432</td>
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<td></td>
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<td>28</td>
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<td>17/03/05</td>
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<td>63.40</td>
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</table>
6.5.2 Nitrification – results and discussion

The VSS specific ammonia utilisation rate \([\text{SAUR}, \text{mg NH}_4\text{-N/(g VSS.h)}]\) and maximum specific growth rate \((\mu_A)\) of the ANOs were calculated as described above for the 29 nitrification batch tests. The SAUR and \(\mu_A\) results are shown in Figs. 6.7 and 6.8.

![Figure 6.7: VSS specific nitrification rates (SAUR) for the batch tests on MBR system mixed liquor at different VSS concentrations.](image-url)
From Figs. 6.7 and 6.8, as the VSS concentration increases, the SAUR and \( \mu_A \) decrease. It was hypothesised that this decrease arises either from oxygen or ammonia diffusion limitations. In the 2 Group 2 batch tests, by using pure oxygen for aeration in one to elevate the DO concentration throughout the test to 10-15 mg O/l and air in the other parallel test (DO 2-5 mg O/l), it appeared that the VSS concentration effect was not related to DO transfer limitations because the SAUR and \( \mu_A \) were both lower at the high DO than at the low DO (Figs. 6.7 and 6.8). From the increase in SAUR and \( \mu_A \) with increase in initial ammonia concentration evident in Figs. 6.7 and 6.8, it was concluded that the effect was possibly due to ammonia diffusion limitation into the floc at increasing VSS concentration.

Comparing the OUR profiles in the nitrification batch tests at different VSS concentrations clearly demonstrates a change in nitrification behaviour with change in VSS concentration (Figs. 6.9 and 6.10). At low VSS concentration (2.2 g VSS/l, Fig. 6.9), on addition of ammonia, the OUR increases sharply to a plateau, remains at this plateau for a period, and then drops precipitously to the endogenous level. In contrast, at high VSS concentration (8.8 g VSS/l, Fig. 6.10), the increase and decrease in OUR on ammonia addition and depletion is slow. Whether this effect of increasing VSS concentration on the SAUR and \( \mu_A \) is related to ammonia transport limitation is difficult to tell from the results. The batch tests at increasing VSS concentration have increasing OUR, e.g. 35 mg O/(l.h) at 2.2 g VSS/l (Fig. 6.9) and 90 mg O/(l.h) at 8.8 g VSS/l (Fig. 6.10). Stenstrom and Song (1991) show that DO diffusion resistance into the flocs for nitrification increases with increasing nitrification OUR.

The 6 Group 3 parallel nitrification batch tests results on low VSS concentration ML (2-3 g VSS/l) from the CAS and MBR systems demonstrate that the SAUR and \( \mu_A \) for the CAS system are approximately 1.8 and 1.2 times higher than those for the MBR system at the same low VSS concentration. The results are listed in Table 6.4 and plotted in Figs. 6.11 (SAUR) and 6.12 (\( \mu_A \)). The \( \mu_A \) is less different than the SAUR because the sludge production in the MBR system is higher than in the CAS system, so the ANOs are diluted in a greater VSS mass at the same sludge age.
Figure 6.9: Oxygen Uptake Rate (OUR) profile for the batch test on MBR system mixed liquor at low VSS concentration (2.2 g VSS/ℓ).

Figure 6.10: Oxygen Uptake Rate (OUR) – time profile for the batch test on MBR system mixed liquor at high VSS concentration (8.8 g VSS/ℓ).

Table 6.4: Nitrification batch results for the MBR and CAS system ML at the same low VSS concentration (2-3 g VSS/ℓ).

<table>
<thead>
<tr>
<th>Parameter</th>
<th>MBR system</th>
<th>CAS system</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>23/5</td>
<td>31/5</td>
</tr>
<tr>
<td>Ammonia added (mg NH₄-N/ℓ)</td>
<td>42</td>
<td>33</td>
</tr>
<tr>
<td>VSS (mg VSS/ℓ)</td>
<td>2692</td>
<td>2670</td>
</tr>
<tr>
<td>Ammonia/VSS load (mg NH₄-N/g VSS)</td>
<td>15.75</td>
<td>12.26</td>
</tr>
<tr>
<td>AUR [mg NH₄-N/(ℓ.h)]</td>
<td>5.32</td>
<td>5.60</td>
</tr>
<tr>
<td>SAUR [mg NH₄-N/(g VSS.h)]</td>
<td>1.98</td>
<td>2.10</td>
</tr>
<tr>
<td>ANO max specific growth rate (μₐ, /d)</td>
<td>0.374</td>
<td>0.396</td>
</tr>
<tr>
<td>ANO/VSS fraction (fₐ/ANO)</td>
<td>0.0127</td>
<td>0.0127</td>
</tr>
</tbody>
</table>
Figure 6.11: Specific nitrification rates (SAUR) for the MBR and CAS system mixed liquor at similar low VSS concentration (2-3 g VSS/l).

Figure 6.12: Nitrifier (ANO) maximum specific growth rate ($\mu_A$) for the MBR and CAS system mixed liquor at similar low VSS concentration (2-3 g VSS/l).

From all the batch tests, it is evident that the ANOs in the MBR and CAS systems exhibit a different activity, apparently induced by different environments under which the ANOs develop. The reasons for this possibly are: (1) In CAS systems with SSTs, organism loss via the effluent occurs including ANOs. Therefore CAS system may select ANOs with higher maximum specific growth rates ($\mu_A$) than MBR systems. In the MBR system all the ANOs are retained, including slow growing ones. (2) At the high VSS concentrations in the MBR system, oxygen and ammonia transport limitations decrease the observed SAUR and $\mu_A$.

Previous findings that the nitrification rate increased as floc size decrease (Zhang et al., 1997). The smaller size of the flocs may help the nitrifiers to be more in contact with the substrate. As in this investigation, Zhang et al. (1997) measured, for the MBR system, the nitrification activities at different ML concentration and found similar results: the specific nitrification rate decreased as the ML concentration increased. After the dilution, it was found that floc size was decreased markedly in the MBR system, so the floc size distribution changes when ML concentration was changed.

In Table 6.5, the specific nitrification rates obtained in previous investigations on the MBR and CAS systems are reported together with the operating conditions (wastewater characteristic and sludge age of the parent systems).
Table 6.5 Specific nitrification rates (mg N/g VSS h) for MBR and CAS systems

<table>
<thead>
<tr>
<th>Reference</th>
<th>System</th>
<th>Specific nitrification rate (mgN/gVSS h)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fan et al. (2000)</td>
<td>MBR (domestic wastewater)</td>
<td>1,7±2</td>
</tr>
<tr>
<td>Zhang et al. (1997)</td>
<td>MBR (SRT=12 d, mixed wastewater)</td>
<td>2,9</td>
</tr>
<tr>
<td></td>
<td>CAS (SRT=4 d, domestic wastewater)</td>
<td>1,02</td>
</tr>
<tr>
<td>Manser et al. (2005a)</td>
<td>MBR (SRT=20 d, domestic wastewater)</td>
<td>1,99</td>
</tr>
<tr>
<td></td>
<td>CAS (SRT=20 d, domestic wastewater)</td>
<td>2,38</td>
</tr>
<tr>
<td>Ahn et al. (1999)</td>
<td>MBR (leachate)</td>
<td>1,4</td>
</tr>
<tr>
<td>Han et al. (2004)</td>
<td>MBR-ND (SRT=30 d, synthetic wastewater)</td>
<td>0,18</td>
</tr>
<tr>
<td>Lubbecke et al. (1995)</td>
<td>MBR (SRT=8 d, synthetic wastewater-high NH4 load)</td>
<td>4,7</td>
</tr>
<tr>
<td>Li et al. (2005)</td>
<td>MBR (no waste, synthetic wastewater-high NH4 load)</td>
<td>7,8</td>
</tr>
<tr>
<td></td>
<td>CAS (no waste, synthetic wastewater-high NH4 load)</td>
<td>15,6±18,2</td>
</tr>
<tr>
<td>Liebig et al. (2001)</td>
<td>MBR (no waste, synthetic wastewater-high NH4 load)</td>
<td>24</td>
</tr>
<tr>
<td></td>
<td>CAS (synthetic wastewater-high NH4 load)</td>
<td>26</td>
</tr>
<tr>
<td>Wyffel et al. (2003)</td>
<td>MBR (no waste, synthetic wastewater-high NH4 load)</td>
<td>18,3</td>
</tr>
<tr>
<td>Li et al. (2006)</td>
<td>MBR (no waste, synthetic wastewater-high NH4 load)</td>
<td>5,5±24</td>
</tr>
</tbody>
</table>

Comparing the results of this investigation with those listed in Table 6.5 it can be seen that very similar results are obtained in the investigations with domestic wastewater and sludge ages between 12 and 20 days (Fan et al., 2000, Zhang et al. 1997, Manser et al. 2005a). But this does not really mean too much. Because the maximum specific growth rates ($\mu_A$) were not calculated in these investigations, it is difficult to compare the results meaningfully with the results of this investigation. If the influent TKN/COD ratio or the unbiodegradable COD fraction ($f_{\text{S}u\text{p}}$) of the wastewater are different, then different SAUR rates [mg FSA-N/(g VSS.h)] will be obtained that have nothing to do with the ANOs but arise purely from differences in the ANO/VSS ratio ($f_{\text{ANO/VSS}}$). Nitrification rates can only be meaningfully compared between different AS systems at different sludge ages and fed different wastewater on the basis of $\mu_A$.

Parallel nitrification batch tests results on sludges from the MBR and CAS systems if this investigation demonstrate that the SAUR and $\mu_A$ for the CAS BNR system are approximately 1.8 times and 1.2 times higher than those measured in parallel MBR system at the same VSS concentration (Figs. 6.11 and 6.12).

The nitrification batch tests on the MBR and CAS system sludges clearly exhibit different growth behaviour apparently induced by different environments in which the ANOs grow. It is possible to speculate as to the reasons for this. In the CAS system with SSTs, organism loss via the effluent will occur, particularly for the poorly flocculating nitrifiers. This causes that the system selects for nitrifiers with higher specific growth rates. In the MBR system with membranes, all nitrifiers will be retained including the slower growers. As a consequence of the different environmental conditions (organism retention due to settleability versus total organism retention due to membranes), different microorganism groups can be retained and develop. In particular, this may
include predators, and so different predation populations could possibly occur in the two systems, resulting in different predator behaviour, which may negatively affect the ANOs.

From Table 6.5, it is interesting to notice that similar differences between MBR and CAS systems have been observed in the investigations where the CAS and MBR system were operated in parallel at the same sludge age and treated the same wastewater (Li et al. 2005, Manser et al. 2005a). Moreover Luxmy et al. (2000), operating at the same conditions 2 parallel systems: MBR and CAS, found significantly different bacterial community in the two systems as determined by FISH (Fluorescence In Situ Hybridization) and DGGE (Denaturing Gradient Gel Electrophoresis) techniques. Similar results were observed by Liebig et al. (2001), additionally the MBR resulted in a lower relative in situ abundance of ammonia and nitrite oxidising bacteria, only *Nitrospira* (higher substrate affinity $K_n$ and lower growth rate $\mu_A$) was identified to contribute to ammonia oxidation.

In summary for nitrification:

1. At the same VSS concentration, the MBR system exhibits lower VSS specific ammonia utilisation rate (SAUR) and ANO maximum specific growth rates ($\mu_A$) than the parallel CAS system, apparently due to different selection pressures imposed by membranes and SSTs.
2. In the MBR system, as the VSS concentration increases, the SAUR and $\mu_A$ decrease, apparently due to ammonia and/or oxygen transfer limitations.
3. In the MBR system, as the initial ammonia concentration is increased for similar VSS concentrations, the SAUR and $\mu_A$ increase, indicating possible ammonia transport limitation at increasing VSS concentration.

### 6.6 Denitrification kinetics – Anoxic Batch Tests

#### 6.6.1 Batch test and calculation procedures

In nitrification denitrification (ND) systems, the kinetics of biological denitrification have been extensively investigated and a simplified general kinetic model has been developed (van Haandel et al., 1981; Ekama and Wentzel, 1999) as:

$$\frac{dNO_3^- - N}{dt} = -K \cdot X_v = -K_{2OH0} \cdot f_{avHO} \cdot X_v$$

where

- $dNO_3^-/dt$ = nitrate utilisation rate NUR – mg NO$_3$-N/(ℓ,d)
- $K_{2OH0}$ = OHO specific denitrification rate constant utilizing slowly biodegradable particulate organics (BPO) in the primary anoxic reactor – mg NO$_3$-N/(mg OHOVSS.d)
- $f_{avHO}$ = OHO active fraction of volatile settleable solids (VSS)
- $X_v$ = measured VSS concentration (mg VSS/ℓ)

In the NDBEPR system, an initial rapid rate of denitrification by influent readily biodegradable organics (RBO) is usually absent because these organics have been converted to VFAs and taken up by the PAOs in the anaerobic reactor. Therefore, provided the PAOs do not participate in the denitrification, the denitrification proceeds as a single slow rate by the OHOs ($K_{2OH0}$) utilizing the biodegradable particulate organics (BPO) (Clayton et al., 1991, Ekama and Wentzel, 1999). To determine the OHO specific denitrification rate constant utilizing BPO $K_{2OH0}$ [mg NO$_3$-N/(mg OHOVSS.d)] with Eq. 6.3, nitrate concentration versus time profiles (nitrate utilisation rate, NUR) in anoxic batch tests were measured. The anoxic batch tests were conducted in the same batch reactors as the nitrification batch tests, except the air supply was replaced with nitrogen gas during the non-aerated (anoxic) part of the batch tests.
At the start of an anoxic batch test, measured volumes of ML from the anaerobic and aerobic reactors of the MBR and CAS systems were transferred to the batch reactors. Because the VSS concentration in the anaerobic reactor is lower than that in the aerobic reactor, to achieve high VSS concentrations in the batch tests (8-12 g VSS/ℓ), the sludge needed to be concentrated rather than diluted with effluent. An appropriate volume of mixed aerobic and anaerobic reactor mixed liquor was gently centrifuged (500 rpm for 3 min) and half the supernatant discarded. The substrate available for the denitrification is the BPO from the wastewater and endogenous process enmeshed in the anaerobic and aerobic reactor mixed liquor. Because anoxic conditions are required, the mixed liquor was de-oxygenated to remove the DO from the aerobic reactor sludge by the nitrogen gas flow. When the DO concentration was zero, nitrate was added as a small volume (120-240 ml) of concentrated potassium nitrate solution (7.2 gKNO₃/ℓ), sufficient to ensure that nitrate was present for the entire test period (5-6 hours). At regular intervals (see Table 6.6), grab samples were drawn from the batch reactor, immediately centrifuged and filtered (0.45 µm) and 2 drops HgCl₂ (8.6 g/ℓ) added to the filtrate. Filtrates were analyzed for NO₃, NO₂, TP, NH₄-N and soluble COD. Additionally at the beginning and end of the tests duplicate samples were taken for VSS and TSS measurements. During the test the pH of the mixed liquor was controlled at around 7.5 by adding small volumes of 20% HCl solution. After the anoxic period, aeration was commenced for 18-20 h which was controlled with the DO controller/OUR meter used in the nitrification batch tests. Sampling was continued during the aerobic period to observe the ANO and PAO behaviour.

<table>
<thead>
<tr>
<th>Sample Interval (min)</th>
<th>Test Period anoxic condition (min)</th>
</tr>
</thead>
<tbody>
<tr>
<td>5</td>
<td>0-20</td>
</tr>
<tr>
<td>10</td>
<td>20-40</td>
</tr>
<tr>
<td>20</td>
<td>40-60</td>
</tr>
<tr>
<td>30</td>
<td>90-360</td>
</tr>
</tbody>
</table>

Throughout the anoxic period of the batch tests, precautions were taken to ensure no oxygen ingress into the mixed liquor. Nitrogen gas was bubbled through the batch reactor over the whole anoxic period. Not only did this purge any oxygen that might dissolve into the mixed liquor, it served also to form an inert layer of N₂ gas between the mixed liquor surface and the air above, preventing further oxygen dissolution. Small (10 mm diameter) hollow plastic spheres were also floated on the surface of the mixed liquor reactor to isolate the mixed liquor from the N₂ gas/atmosphere above.

Five groups of anoxic batch tests for denitrification were conducted, viz. Group (1): On MBR system ML at different VSS concentrations between 2.5 and 12 g VSS/ℓ with ML from the anaerobic and aerobic reactors mixed in proportion to the recycles entering the anoxic reactor; Group (2): like Group (1) but at different nitrate concentrations; Group (3): like Groups (1) and (2) but with different proportions of anaerobic and aerobic ML (Set 3.1 – 50/50 by VSS mass, Set 3.2 – 100% anaerobic and Set 3.3 – 100% aerobic); Group (4) on MBR and CAS system ML in parallel with the MBR ML diluted to the same low VSS concentration as that from the CAS system (2-3 g VSS/ℓ) and with ML from the anaerobic and aerobic reactors mixed in proportion to the recycles entering the anoxic reactor and Group (5): like Group (4) but with wastewater added.

6.6.2 Denitrification kinetics – results and discussion

From the 14 Groups (1) and (2) anoxic batch tests, a general denitrification behaviour could be defined (Fig. 6.13), i.e. nitrate reduction took place via a single linear rate, in agreement with observations on BNR sludge by others (Clayton et al., 1991; Ekama and Wentzel, 1999). Thus denitrification with RBO via the K₁ rate was absent (Fig. 6.13) which confirms that all RBO was removed in the anaerobic reactor of the MBR system. The P concentration remained essentially
constant (Fig. 6.13), indicating that apparently, the PAOs did not take part in the denitrification process, even though PAOs were known to be present in the mixed liquor from the high P removal exhibited in the MBR system. This implied that the observed denitrification rate (NUR) could be ascribed solely to the OHOs, i.e. zero contribution by PAOs. So the OHO specific denitrification rate $K_{2OHO}$ was calculated from Eq. 6.3, i.e. $K_{2OHO}=NUR/(f_{vOHO} X_v)$ mg NO$_3$-N/(mg OHOVSS.d).

### Denitrification Batch Test

$X_v=10824$ mgVSS/l

NO$_3$-N added = 55 mgN/l

$y = -12,772x + 76,555$

$R^2 = 0.994$

![Figure 6.13: Nitrate (NO$_3$) and total soluble phosphorous (P$_{tot}$) concentration time profiles](image)

The $K_{2OHO}$ rates obtained at different MBR system VSS concentrations (2.5-12 g VSS/l) and different initial nitrate concentrations ranging from 30 to 90 mg N/l (Groups 1 and 2) are shown in Fig. 6.14. From the nitrate concentration-time profiles (Fig. 6.13), the nitrate utilisation rate (NUR) was zero order with respect to nitrate concentration, in agreement with past work (van Haandel et al., 1981, Clayton et al., 1991; Ekama and Wentzel, 1999). The $K_{2OHO}$ rates observed at different VSS concentrations indicate no effect of VSS on the rate (Fig. 6.14), in contrast to the observed effect of VSS on the nitrification rate. From all the anoxic batch tests, an average $K_{2OHO} = 0.264$ mg NO$_3$-N/(mg OHOVSS.d) was obtained. This is very close to the average $K_{2OHO}$ rate reported in the literature for conventional (low VSS) BNR systems with SSTs, i.e. 0.255 from Ekama and Wentzel (1999).

For the one Group 3 batch test with no anaerobic reactor ML (Set 3.3), the nitrate (NO$_3$-N) and filtered total P ($P_{tot}$) concentration time profile was the same as those of Groups (1) and (2), i.e. denitrification took place at a single linear rate [slope=NUR, mg NO$_3$-N/(l.h)] and the filtered
total P concentration remained essentially constant (Fig. 6.15). Because the PHA of the PAOs in the aerobic reactor is low, most likely only the OHOs mediate this “post” denitrification (K3OHO), where the energy source is derived from the BPO generated by organism death and cell lysis. So the K3OHO rate was calculated from K3OHO=24.NUR/(favOHOXv).

Figure 6.15: Nitrate (NO3-N) and total soluble P (Ptot) concentration time profiles for the anoxic batch tests on MBR system ML from the aerobic reactor only.

For the three Group 3 batch tests with a high proportion of anaerobic reactor ML (two Set 3.1 with 50/50 anaerobic and aerobic reactor ML and one Set 3.2 with 100% anaerobic reactor ML), the general nitrate (NO3-N) and filtered total P (Ptot) concentration time profiles (Fig. 6.16) were different to those with low or no anaerobic reactor ML. In these three batch tests, denitrification took place via two sequential linear rates – an initial fast rate for less than an hour followed by a second slower rate for the remainder of the anoxic period. Because P uptake terminated at about the same time as the change in denitrification rate, it seemed that PAOs also denitrified during the initial fast rate but stopped within 1 h. Then the OHOs continued to denitrify during the slower rate for the remainder of the anoxic period while nitrate was present. A similar behaviour was observed in the P removal batch tests (see below). It would appear that the high proportion of anaerobic ML in these anoxic batch tests resulted in a high concentration of PHA, which made denitrification by PAOs become visible. In contrast, in the batch tests with a low or zero (Groups 1 and 2 and Set 3.3) proportion of anaerobic ML, the PHA concentration is diluted by the aerobic reactor sludge with low PHA so that denitrification by the PAOs is not detectable. The PHA profiles measured in the P removal batch tests confirmed this (see below). To determine the OHO (K2OHO) and PAO (KPAO) specific denitrification rates, the NUR of the second slower rate [slope = NUROHO, mg NO3-N/ℓh] was deducted from the NUR of the first faster rate (slope = NURPAO+OHO) to obtain NURPAO. The OHO and PAO specific denitrification rates were then calculated from K2OHO=24.NUROHO/(favOHOXv) and KPAO=24.NURPAO/(favPAOXv), where Xv is the measured VSS concentration in the batch test in mg VSS/ℓ. The calculated specific denitrification rates (K2OHO, KPAO, K3OHO) from the Group 3 anoxic batch tests were added to Figure 6.14 and are plotted in Fig. 6.17.
Figure 6.16: Typical nitrate and total soluble phosphorous concentration time profiles for anoxic batch tests on MBR system mixed liquor with high proportion of anaerobic reactor ML.

Figure 6.17: OHO and PAO specific denitrification rate constants ($K_{\text{OHO}}$, $K_{\text{PAO}}$) on MBR system ML at different VSS and initial nitrate concentrations and different proportions anaerobic and aerobic reactor ML (Set 3.1 – 50/50 by VSS mass, Set 3.2 – 100% anaerobic and Set 3.3 – 100% aerobic).

From Fig. 6.17 the following can be observed:

- The $K_{\text{OHO}}$ rates fall within the same band of 0.2 to 0.3 mg NO$_3$-N/(mg OHOVSS.d) with an average of 0.25. Also no effect of VSS or initial nitrate concentration was apparent.
- At a mean of 0.104 mg NO$_3$-N/(mg PAOVSS.d), the $K_{\text{PAO}}$ is 2.5 times lower than $K_{\text{OHO}}$. This conforms to the observation of Hu et al. (2002) that the PAOs denitrify with PHA significantly slower than the OHOs with slowly biodegradable particulate organics (BPO). This is quantified further with the anaerobic-anoxic/aerobic batch tests discussed below.
- At a rate of 0.278 mg NO$_3$-N/mg OHOVSS d), the $K_{\text{OHO}}$ rate was in the same range of the $K_{\text{OHO}}$ rates. From the $K_{\text{OHO}}$ rate of 0.10 reported by Ekama and Wentzel (1999), this was unexpectedly high. The reason for this high “post” denitrification rate is difficult to establish because the question arises what the source of the organics is that drives the rate to be so high. Possibly it is due to extra-cellular polymeric substances (ESP) or soluble microbiological products (SMP) in the aerobic reactor of the MBR system. This and
previous investigations (Sperandio et al., 2005) observed the accumulation of these products in MBR systems. However, if these products are not utilised in the aerobic reactor with oxygen, why should they be utilised in the anoxic reactor with nitrate and cause a high denitrification rate?

The results of the Group (4) batch tests on ML from the MBR and CAS systems at low VSS concentration (with ML from the aerobic and anaerobic reactors mixed in proportion to the flows entering the anoxic reactor) are listed in Table 6.7. In conformity with Groups (1), (2) and (3) Set 3.3, the denitrification took place at a single linear rate [slope=NUR, mg NO₃-N/(ℓ.h)] and the filtered total P concentration remained essentially constant. The average OHO specific denitrification rate for the MBR system ML was 0.242 mg NO₃-N/(mg OHOVSS.d) and that for the CAS system was 0.206 mg NO₃-N/(mg OHOVSS.d). Although the MBR system average rate is 20% higher than that of the CAS system, the two rates are not statistically different due to the low number of tests (3 on each system) and the high variability in the rates.

Table 6.7: Specific OHO denitrification rates (K₂OHO) for parallel batch tests on MBR and CAS system mixed liquor at the same low VSS concentration (2.5-3.5 g VSS/ℓ)

<table>
<thead>
<tr>
<th>Batch Test</th>
<th>System</th>
<th>NUR</th>
<th>VSS</th>
<th>VSS Specific Denit. Rate</th>
<th>OHO Active Fraction</th>
<th>OHO Specific Denit. Rate</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>mg NO₃-N</td>
<td>mg VSS/ℓ</td>
<td>mg NO₃-N/(g VSS.h)</td>
<td>favOHO</td>
<td>mg NO₃-N/(mg OHOVSS.d)</td>
</tr>
<tr>
<td>17Jun05</td>
<td>CAS</td>
<td>2.86</td>
<td>2294</td>
<td>1.25</td>
<td>0.120</td>
<td>0.249</td>
</tr>
<tr>
<td>6Jul05</td>
<td>CAS</td>
<td>3.16</td>
<td>2830</td>
<td>1.12</td>
<td>0.156</td>
<td>0.172</td>
</tr>
<tr>
<td>12Sep05</td>
<td>CAS</td>
<td>2.39</td>
<td>2302</td>
<td>1.04</td>
<td>0.126</td>
<td>0.198</td>
</tr>
<tr>
<td>17Jun05</td>
<td>MBR</td>
<td>2.37</td>
<td>2550</td>
<td>0.93</td>
<td>0.095</td>
<td>0.235</td>
</tr>
<tr>
<td>6Jul05</td>
<td>MBR</td>
<td>3.37</td>
<td>3534</td>
<td>0.95</td>
<td>0.106</td>
<td>0.216</td>
</tr>
<tr>
<td>12Sep05</td>
<td>MBR</td>
<td>2.27</td>
<td>2324</td>
<td>0.97</td>
<td>0.085</td>
<td>0.275</td>
</tr>
</tbody>
</table>

In the Group (5) batch tests, which were the same as Group (4) except with added wastewater comprising RBO and SBO but not acetate, the denitrification took place in 2 phases – see Figs. 6.18a (MBR) and 6.18b (CAS), i.e. an initial rapid NUR phase due to the utilisation of influent RBO and BPO followed by a second slower NUR phase due to the utilisation of the only the BPO. During both phases the filtered total P concentration remained essentially constant, indicating that the PAOs did not take part in the denitrification.
To determine the OHO specific denitrification rates utilizing RBO ($K_{1OHO}$) and SBCOD ($K_{2OHO}$), the NUR of the second slower rate [slope = NUR$_{2OHO}$, mg NO$_3$-N/(ℓ.h)] was deducted from the NUR of the first faster rate (slope = NUR$_{1OHO+2OHO}$) to obtain NUR$_{1OHO}$. The OHO specific denitrification rates were then calculated from $K_{2OHO}$ = 24.NUR$_{2OHO}$/(f$_{w}$OHOX$_v$) and $K_{1OHO}$ = 24.NUR$_{1OHO}$/(f$_{w}$OHOX$_v$), where X$_v$ is the measured VSS concentration in the batch test in mg VSS/ℓ. The results are listed in Table 6.8. The average $K_{1OHO}$ and $K_{2OHO}$ rates are 0.256 and 0.413 mg NO$_3$-N/(mg OHOVSS.d) respectively. Usually for ND systems, the $K_{1OHO}$ rate is much higher than the $K_{2OHO}$ rate (van Haandel et al., 1981), but this is not the case here. The $K_{1OHO}$ is also much slower than 0.70 mg NO$_3$-N/(mg OHOVSS.d) observed by Clayton et al. (1991) in their NDBEPR systems. However, what the actual $K_{1OHO}$ rate is in NDBEPR systems makes little difference because the influent RBO is usually converted to VFA and taken up by the PAOs in the anaerobic reactor and so is not utilised by OHOs for denitrification. The $K_{2OHO}$ rate is higher than the 0.25 mg NO$_3$-N/(mg OHOVSS.d) observed in Group (1) and (2) batch tests because in the Group (5) batch tests the BPO/VSS ratio is much higher from the added wastewater BPO.
Table 6.8: Specific OHO denitrification rates (K_{1OHO} and K_{2OHO}) for parallel batch tests on MBR and CAS system mixed liquor at the same low VSS concentration (2.5-3.5 g VSS/ℓ).

<table>
<thead>
<tr>
<th>Batch Test</th>
<th>System</th>
<th>NUR</th>
<th>VSS</th>
<th>VSS Specific Denit. Rate</th>
<th>OHO Active Fraction</th>
<th>OHO Specific Denit. Rate</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>mg NO3-N / (ℓ.h)</td>
<td>mg VSS/ℓ</td>
<td>mg NO3-N / (g VSS.h)</td>
<td>favOHO</td>
<td>mg NO3-N / (mg OHOVSS.d)</td>
</tr>
<tr>
<td>23Jun05</td>
<td>CAS</td>
<td>3.78</td>
<td>2994</td>
<td>1.26</td>
<td>0.156</td>
<td>0.194</td>
</tr>
<tr>
<td>29Jun05</td>
<td>CAS</td>
<td>3.85</td>
<td>2786</td>
<td>1.38</td>
<td>0.156</td>
<td>0.212</td>
</tr>
<tr>
<td>23Jun05</td>
<td>MBR</td>
<td>3.14</td>
<td>4210</td>
<td>0.75</td>
<td>0.106</td>
<td>0.169</td>
</tr>
<tr>
<td>29Jun05</td>
<td>MBR</td>
<td>4.82</td>
<td>2424</td>
<td>1.99</td>
<td>0.106</td>
<td>0.450</td>
</tr>
<tr>
<td>23Jun05</td>
<td>CAS</td>
<td>7.97</td>
<td>2994</td>
<td>2.66</td>
<td>0.156</td>
<td>0.410</td>
</tr>
<tr>
<td>29Jun05</td>
<td>CAS</td>
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<td>2786</td>
<td>2.29</td>
<td>0.156</td>
<td>0.353</td>
</tr>
<tr>
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<td>MBR</td>
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<td>4210</td>
<td>1.75</td>
<td>0.106</td>
<td>0.396</td>
</tr>
<tr>
<td>29Jun05</td>
<td>MBR</td>
<td>5.26</td>
<td>2424</td>
<td>2.17</td>
<td>0.106</td>
<td>0.491</td>
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</tbody>
</table>

6.7 BIOLOGICAL P REMOVAL KINETICS – ANAEROBIC-ANOXIC/AEROBIC BATCH TESTS

6.7.1 Batch test and calculation procedures

For these batch tests, MBR system anoxic reactor ML (in which nitrate was < 1 mg NO3-N/ℓ) was subjected first to anaerobic and then to anoxic and/or aerobic phases (Figure 6.19). Because the VSS concentration in the anoxic reactor is lower than that in the aerobic reactor, to achieve the high VSS concentrations in the batch tests (8-12 g VSS/ℓ), the sludge needed to be concentrated rather than diluted. An appropriate volume of anoxic reactor mixed liquor was gently centrifuged (500 rpm for 3 min) and half the supernatant discarded. At the start of the anaerobic phase, defined masses of concentrated sodium acetate and potassium phosphate or real wastewater (without acetate) were added. The anaerobic phase was maintained for 4 to 5 h, which was sufficient for complete acetate uptake and P release. Then the batch volume was halved – one half was aerated with online OUR measurement and nitrate was dosed to the other half without aeration to establish anoxic conditions. The pH in the batch tests was maintained between 7.0 and 7.2 by dosing concentrated HCl or NaOH throughout the batch test as required.

The batch tests were conducted in 3 reactors constructed of clear cylindrical Perspex (~ 5 mm thick): the anaerobic reactor with volume of 5 liters and diameter of 200 mm and the aerobic and anoxic reactors with volume of 3 liters and diameter of 120 mm (the same ones used for the aerobic and anoxic batch tests). The contents of the reactors were completely mixed by means of motor driven paddle mixers mounted centrally on the lid of the reactor. To ensure good mixing, each reactor was equipped with a pair of vertical side wall baffles situated opposite each other and extending about 2 cm radially into the bulk liquid. The reactors had two paddles (one with inclined vanes) fitted to the mixing shaft, one situated on the bottom and one 2 cm above it. These were positioned to ensure complete suspension of the reactor contents while avoiding turbulence at the liquid surface, to minimise air entrainment into the bulk solution. The lid of the anaerobic reactor had two portholes for taking of samples and insertion of a pH probe. Each reactor, situated on its bottom plate, had a single inlet for the nitrogen or air supply and outlet for the emptying of the reactor. The nitrogen gas in the anaerobic and anoxic reactors was provided by a liquid nitrogen bottle that entered through a small Perspex tube which terminated in a porous stone at the bottom of the reactor vessel.
Aeration was provided by low pressure compressed air which entered through a small Perspex tube which terminated in a porous stone at the bottom of the reactor vessel. The nitrogen and air flowrates were regulated by throttling a ball valve on the main nitrogen/air supply line and adjusting a hose-clamp on the tube entering reactor.

During the batch test, the inside walls of the reactors were routinely brushed to prevent mixed liquor particles from adhering to the reactor walls above the liquid surface.

Throughout the anaerobic and anoxic phases, precautions were taken to ensure no oxygen ingress into the mixed liquor. Nitrogen gas was bubbled through the batch reactor over the anaerobic and anoxic periods. Not only did this purge any oxygen that might dissolve into the mixed liquor, it served also to form an inert layer of N₂ gas between the mixed liquor surface and the air above, preventing oxygen dissolution. Small (10 mm diameter) hollow plastic spheres were also floated on the surface of the reactor contents to isolate the mixed liquor from the N₂ gas/atmosphere.

The typical set-up configured for an anaerobic and aerobic/anoxic batch tests are shown in Figs. 6.20 and 6.21.
Once the anaerobic condition was established, a specific volumes of concentrated sodium acetate (CH₃COONa 10 g/ℓ) and concentrated potassium phosphate (K₂PO₄ 10 g/ℓ), heated to 20°C, were added.

During the anaerobic phase, grab samples were taken manually at specific time intervals; the sampling intervals (in minutes) for the different periods during the test are summarised in Table 6.9.

![Batch reactor set up for aerobic and anoxic phases of the batch test.](image)

**Figure 6.21: Batch reactor set up for aerobic and anoxic phases of the batch test.**

<table>
<thead>
<tr>
<th>Sample Interval (min)</th>
<th>Test Period anaerobic condition (min)</th>
</tr>
</thead>
<tbody>
<tr>
<td>5</td>
<td>0-15</td>
</tr>
<tr>
<td>10</td>
<td>15-45</td>
</tr>
<tr>
<td>15</td>
<td>45-60</td>
</tr>
<tr>
<td>20</td>
<td>60-100</td>
</tr>
<tr>
<td>30</td>
<td>100-190</td>
</tr>
<tr>
<td>50</td>
<td>190-240</td>
</tr>
</tbody>
</table>

Samples (30-40 ml) were taken from the batch reactor and immediately centrifuged (3500 rpm for 10 min) and filtered (0.45 µm), after which 2 drops of 8.6 g/ℓ Hg Cl were added to the filtrate. The samples were normally stored overnight at 4°C and analyzed the following day for HAc, NO₃-N, NO₂-N, PO₄-P, NH₄-N and soluble COD. All analyses were conducted according to Standard Methods (1985). Additionally HAc was analysed with a HPLC-Beckman Coulter System Gold (UV detector 215 nm, mobile phase: 0,1% H₃PO₄, flow rate: 0,5 ml/min; Temperature: 60°C; Column: Water Fast Fruit Juice Column). VSS and TSS were measured at the start, phase change and end of the batch tests. The solids fraction of all the samples was analyzed for polyhydroxybutyrate (PHB) with the method of Satoh et al. (1996).

From the batch test results the following slopes were calculated from the concentration-time profiles: Anaerobic maximum P release and acetate (HAc) uptake rates, the P release to acetate uptake (P/HAc) ratio and the aerobic and anoxic maximum P uptake rates. Also from the anoxic NUR, the denitrification rates and from the aerobic AUR, the nitrification rates were determined.
From a direct comparison of the P uptake rate under aerobic and anoxic conditions, the proportion of denitrifying PAOs ($X_{DNPAO}/X_{PAO}$) was estimated with the method of Meinhold et al. (1999), i.e.

$$X_{PAO} = X_{DNPAO} + X_{APAO}$$ and  $$\frac{q_{anx}}{\eta_{NO3}} = \frac{1}{q_{aer}} = \frac{X_{DNPAO}}{X_{PAO}}$$

(6.4)

where

$X_{PAO} =$ concentration of PAOs = $f_{avPAO} X_v$ (mg PAOVSS/ℓ)

$X_{DNPAO} =$ concentration of denitrifying PAOs (mg DNPAOVSS/ℓ)

$X_{APAO} =$ concentration of aerobic PAOs (mg APAOVSS/ℓ)

$q_{anx} =$ specific PAOVSS anoxic P uptake rate – mg P/(g PAOVSS.h)

$q_{aer} =$ specific PAOVSS aerobic P uptake rate – mg P/(g PAOVSS.h)

$$\eta_{NO3} =$$ reduction factor for anoxic conditions relative to aerobic conditions.

The initial slope of the anoxic or aerobic P uptake profiles, i.e. where the P uptake proceeded without PHA limitation, was used to determine the P uptake rates ($q_{anx}$, $q_{aer}$). The $\eta_{NO3}$ value reflects the observation that under anoxic conditions nitrate is less efficient than oxygen as electron acceptor (Payne 1981, Casey et al., 1994, 1999; Kuba et al. 1993, Orhon et al. 1996, Sperandio et al. 1999, Meinhold et al., 1999; Muller et al. 2003). The OHO and PAO active fractions ($f_{avOHO}$ and $f_{avPAO}$) were calculated from the MBR system results as described above. Unfortunately, the MBR anaerobic batch tests were not comparable with those from the CAS system due to the differing P removal behaviour in the CAS system mentioned earlier.

### 6.7.2 Anaerobic P release and anoxic/aerobic P uptake behaviour

Altogether fifteen anaerobic batch tests were conducted, 13 (BTs 1 to 13) with low to moderate acetate dosages varying from 0.009 to 0.043 mg HAcCOD/mg VSS and VSS concentrations ranging from 2.7 to 11.2 g VSS/ℓ, one (BT14) with excess acetate addition at 0.166 g HAcCOD/g VSS at 6.37 g VSS/ℓ and one (BT15) with wastewater addition at 5.52 g VSS/ℓ. The results are listed in Tables 6.10a and 6.10b.

Typical anaerobic-aerobic/anoxic batch test behaviour with low to moderate acetate dose (BT 1 to 13) is shown in Fig. 6.22 (BT8). The P release took place at a high rate while acetate was present. After the acetate was all taken up, a slow P release took place while conditions were still anaerobic, presumably due to some PAO maintenance processes. After the anaerobic phase, P uptake took place in the aerobic and anoxic phases. The aerobic P uptake rate was considerably faster than the anoxic P uptake rate (Fig. 6.22). In the anoxic phase, the P uptake gradually diminished and stopped after ~2 h. In the tests where an aerobic phase followed the anoxic phase, P uptake started again under aerobic conditions and the final P concentration at 24 h reached a low value but not as low as the half of the batch volume that was aerated immediately after the anaerobic phase (Fig. 6.22).
### Table 6.10a: Anaerobic – anoxic/aerobic batch tests results (Batch Tests 1 to 8).

<table>
<thead>
<tr>
<th>Date</th>
<th>18/10/05</th>
<th>21/11/05</th>
<th>28/11/05</th>
<th>04/12/05</th>
<th>06/02/06</th>
<th>09/02/06</th>
<th>13/02/06</th>
<th>29/03/06</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Batch Test No</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>VSS concentration (mg/l)</td>
<td>1</td>
<td>8530</td>
<td>3357</td>
<td>2697</td>
<td>3001</td>
<td>3896</td>
<td>8643</td>
<td>6360</td>
</tr>
<tr>
<td>Organic Load (mg HAC/mg VSS)</td>
<td>2</td>
<td>0.034</td>
<td>0.04</td>
<td>0.043</td>
<td>0.042</td>
<td>0.035</td>
<td>0.034</td>
<td>0.039</td>
</tr>
<tr>
<td>PAOActive fraction (f_{PAO}) (mg PAOVSS/mg VSS)</td>
<td>3</td>
<td>0.237</td>
<td>0.247</td>
<td>0.243</td>
<td>0.243</td>
<td>0.282</td>
<td>0.282</td>
<td>0.282</td>
</tr>
<tr>
<td>OHOActive fraction (f_{OHO}) (mg OHOVSS/mg VSS)</td>
<td>4</td>
<td>0.125</td>
<td>0.073</td>
<td>0.136</td>
<td>0.136</td>
<td>0.148</td>
<td>0.148</td>
<td>0.148</td>
</tr>
<tr>
<td>Prelease/Acetate uptake ratio (mg P/mg HAc)</td>
<td>5</td>
<td>0.51</td>
<td>0.546</td>
<td>0.638</td>
<td>0.583</td>
<td>0.571</td>
<td>0.509</td>
<td>0.642</td>
</tr>
<tr>
<td>Prelease/Acetate uptake ratio (mol/mol)</td>
<td>6</td>
<td>0.5</td>
<td>0.53</td>
<td>0.619</td>
<td>0.565</td>
<td>0.554</td>
<td>0.494</td>
<td>0.622</td>
</tr>
<tr>
<td>Anaerobic P Release rate (mg P/(g VSS.h))</td>
<td>7</td>
<td>12.05</td>
<td>23.69</td>
<td>21.08</td>
<td>23.07</td>
<td>19.36</td>
<td>16.9</td>
<td>20.33</td>
</tr>
<tr>
<td>Anaerobic acetate uptake rate (mg HAc/(g VSS.h))</td>
<td>8</td>
<td>9.41</td>
<td>29.19</td>
<td>31.79</td>
<td>36.99</td>
<td>31.6</td>
<td>27.93</td>
<td>31.65</td>
</tr>
<tr>
<td>Anaerobic acetate uptake rate (mg HAc/(g PAOVSS.h))</td>
<td>9</td>
<td>81.9</td>
<td>118.2</td>
<td>130.8</td>
<td>152.2</td>
<td>112.1</td>
<td>99.0</td>
<td>112.2</td>
</tr>
<tr>
<td>Aerobic P Uptake rate (mg P/(g VSS.h))</td>
<td>10</td>
<td>11.12</td>
<td>11.1</td>
<td>5.48</td>
<td>9.5</td>
<td>13.19</td>
<td>11.07</td>
<td>11.55</td>
</tr>
<tr>
<td>Aerobic P Uptake rate (mg P/(g PAOVSS.h))</td>
<td>11</td>
<td>46.92</td>
<td>44.94</td>
<td>22.55</td>
<td>39.09</td>
<td>46.77</td>
<td>39.26</td>
<td>40.96</td>
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<td>Anoxic P Uptake rate (mg P/(g VSS.h))</td>
<td>12</td>
<td>4.53</td>
<td>3.85</td>
<td>2.68</td>
<td>3.2</td>
<td>3.22</td>
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<td>19.11</td>
<td>15.59</td>
<td>11.03</td>
<td>13.17</td>
<td>11.42</td>
<td>7.66</td>
<td>11.49</td>
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<tr>
<td>DNPAO/PAO ratio (DNPAO/PAO, %)</td>
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<td>51</td>
<td>43</td>
<td>61</td>
<td>42</td>
<td>31</td>
<td>24</td>
<td>35</td>
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<tr>
<td>Anx/Aer P uptake rate ratio (g_{anx}/g_{aer}, %)</td>
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<td>41</td>
<td>34</td>
<td>49</td>
<td>34</td>
<td>25</td>
<td>19</td>
<td>28</td>
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<tr>
<td>OHO Denitrification rate (mg NO\textsubscript{3}-N/(g VSS.h))</td>
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<td>1.74</td>
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<td>1.25</td>
<td>1.14</td>
<td>1.76</td>
<td>1.12</td>
<td>1.52</td>
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<td>PAO Denitrification rate (mg NO\textsubscript{3}-N/(g VSS.h))</td>
<td>17</td>
<td>1.57</td>
<td>20.6</td>
<td>2.87</td>
<td>2.63</td>
<td>1.14</td>
<td>2.04</td>
<td>1.9</td>
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<td>% PAO Denitrification</td>
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<td>19</td>
<td>12.04</td>
<td>24.61</td>
<td>21.56</td>
<td>32.51</td>
<td>17.69</td>
<td>52.28</td>
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<td>NO\textsubscript{2} formation Anoxic (mg NO\textsubscript{2}-N/ℓ)</td>
<td>19</td>
<td>0.85</td>
<td>4.46</td>
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<td>0</td>
<td>1.81</td>
<td>0</td>
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<td>Nitrification Rate (mg FSA/(g VSS.h))</td>
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<td>0.71</td>
<td>1.82</td>
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<td>1.09</td>
<td>1.34</td>
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<td>78</td>
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Table 6.10b: Anaerobic – anoxic/aerobic batch tests results (Batch Tests 9 to 15).

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<th>Date</th>
<th>03/05/06</th>
<th>08/11/05</th>
<th>15/11/05</th>
<th>01/12/05</th>
<th>18/04/06</th>
<th>04/04/06</th>
<th>15/02/06</th>
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<td>Batch Test No</td>
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<td>10</td>
<td>11</td>
<td>12</td>
<td>13</td>
<td>14</td>
<td>15</td>
</tr>
<tr>
<td>VSS concentration mg/ℓ</td>
<td>8721</td>
<td>6258</td>
<td>5026</td>
<td>11182</td>
<td>7503</td>
<td>6369</td>
<td>5522</td>
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<tr>
<td>Organic Load mg HAC/mg VSS</td>
<td>2.034</td>
<td>0.02</td>
<td>0.02</td>
<td>0.01</td>
<td>0.02</td>
<td>0.166</td>
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<tr>
<td>PAOActive fraction (f_{PAO}) mg PAOVSS/mg VSS</td>
<td>0.264</td>
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<td>0.247</td>
<td>0.243</td>
<td>0.288</td>
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<td>0.148</td>
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<tr>
<td>Prelease/Acetate uptake ratio mg P/mg HAc</td>
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<td>0.515</td>
<td>0.606</td>
<td>0.588</td>
<td>0.753</td>
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<td>Prelease/Acetate uptake ratio mol/mol</td>
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<td>0.499</td>
<td>0.587</td>
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<td>0.73</td>
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<td>Anaerobic P Release rate mg P/(g VSS.h)</td>
<td>18.24</td>
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<td>34.92</td>
<td>28.99</td>
<td>27.87</td>
<td>19.36</td>
<td>29.29</td>
<td>22.4</td>
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<td>118.9</td>
<td>162.4</td>
<td>117.4</td>
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<td>67.2</td>
<td>106.1</td>
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<td>9.77</td>
<td>8.62</td>
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<td>8.28</td>
<td>10.49</td>
<td>5.74</td>
<td>6.91</td>
</tr>
<tr>
<td>Aerobic P Uptake rate mg P/(g PAOVSS.h)</td>
<td>37.01</td>
<td>40.09</td>
<td>54.13</td>
<td>34.07</td>
<td>36.42</td>
<td>24.50</td>
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</tr>
<tr>
<td>Anoxic P Uptake rate mg P/(g VSS.h)</td>
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<td>18.91</td>
<td>8.33</td>
<td>6.77</td>
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<tr>
<td>Anoxic P Uptake rate mg P/(g PAOVSS.h)</td>
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<td>7.40</td>
<td>18.91</td>
<td>8.33</td>
<td>6.77</td>
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<td></td>
</tr>
<tr>
<td>DNPAO/PAO ratio DNPAO/PAO</td>
<td>0.97</td>
<td>1.15</td>
<td>1.5</td>
<td>1.01</td>
<td>1.05</td>
<td>1.15</td>
<td>1.02</td>
</tr>
</tbody>
</table>
During the anoxic phase, nitrate reduction took place via two linear rates with a reduction in slope (rate) at about the same time as the anoxic P uptake stopped (Fig. 6.23), like the Group (3) denitrification batch test with a high proportion of anaerobic ML (Set 3.1 and 3.2, Fig. 6.16). The reason for this behaviour is the same – PAO denitrification due high PHA in the batch test. In the anaerobic batch tests, a high PHA is stimulated by the acetate dose.

The observations in Figs 6.22 and 6.23 may be explained by dividing PAOs into two groups – one capable of utilizing nitrate and oxygen as electron acceptor (DNPAO) and one group capable of utilizing only oxygen (APAO) as electron acceptor (Meinhold et al., 1999). When P uptake ceases under anoxic conditions, the DNPAO have utilised their stored PHB. Then only the APAO still have PHB, so when aeration starts, only the APAO take up P. In the anaerobic-aerobic tests, both groups of PAO have stored PHB and consequently both groups take up P. This was confirmed by the PHB profiles, which mirror the P release and P uptake profiles (Fig. 6.24). In the first part of anoxic and aerobic phases, where there is no PHB limitation for both PAO groups, the maximum anoxic P uptake rate is slower than aerobic P uptake, which according to Eq. 6.4 is due to a lower DNPAO VSS concentration and a lower energy capture when the nitrate is utilised as an electron acceptor compared with oxygen, which reduces the yield coefficient under anoxic conditions compared to aerobic conditions (Kuba et al., 1996, Beun et al., 2000; Muller et al., 2003) and is taken into account via the $\eta_{\text{NO}_3}$ factor.
From the observations obtained from all the batch tests and the literature information, it seemed reasonable to accept that both PAOs and OHOs participate to the denitrification process ($K_{2OHO} + K_{PAO}$) in the first part (first slope) of the anoxic phase while only the OHO bacteria continue to denitrify (second slope) for the remainder of the anoxic phase while nitrate was present (Fig. 6.23). A comparison of the PAO and OHO denitrification rates is presented below after the affect of VSS concentration on the acetate uptake, P release and P uptake rates has been ascertained.

6.7.3 Anaerobic acetate uptake and P release kinetics

The specific VSS and specific PAOVSS anaerobic acetate (as COD) uptake and P release rates obtained at the different VSS concentrations for the 15 batch tests are shown in Figs. 6.25 and 6.26 respectively. For Batch Tests 1 to 13, the acetate uptake and P release rates vary over quite a wide range from 19 to 37 mg COD/(g VSS.h) or 67 to 162 mg COD/(g PAOVSS.h) and 12 to 24 mg P/(g VSS.h) or 51 to 96 mg P/(g PAOVSS.h) respectively. No effect of VSS or initial acetate concentration on the rates is apparent from the results (Figs. 6.25 and 6.26). The results obtained with different concentrations of acetate added (corresponding in the graph to different load mg COD/mg VSS) show the HAc uptake rates to be zero order with respect to HAc concentration (Figs. 6.25a and b) in agreement with the literature studies (Wentzel et al., 1985, 1989), and in conformity with the observations in BT14 with excess HAc addition where similar anaerobic HAc uptake and P release rates were observed as in the low to moderate acetate dose BTs 1 to 13. The average (and standard deviation) acetate uptake and P release rates obtained over the VSS concentration range (2.7-11.2 g VSS/ℓ) for the 14 batch tests with acetate dose including the excess acetate dose BT14 are 29.8±5.3 mg COD/(g VSS.h) or 116.1±24.7 mg COD/(g PAOVSS.h) and 18.6±3.1 mg P/(g VSS.h) or 72.3±13.7 mg P/(g PAOVSS.h) respectively. The P release to acetate uptake ratio in the 14 batch tests varied between 0.47 and 0.75 with an average of 0.57±0.07 mg P/mg HAc and also shows no effect with acetate dose and VSS concentration (Fig. 6.27).
Figure 6.25a and b: Anaerobic VSS specific maximum acetate uptake rate mg HAc/(g VSS.h) (Fig. 6.25a, left) and PAOVSS specific maximum acetate uptake rate mg HAc/(g PAOVSS.h) (Fig. 6.25b, right).

Figure 6.26a and b: Anaerobic VSS specific maximum P release rate mg P/(g VSS.h) (Fig. 6.26a, left) and PAOVSS specific maximum P release rate mg P/(g PAOVSS.h) (Fig. 6.26b, right).

Figure 6.27: P release to acetate uptake ratios in the anaerobic phase for the 13 batch tests with low to moderate acetate addition (0.01 to 0.043 mg COD/mg VSS)
6.7.4 Aerobic and anoxic P uptake rates

The specific VSS and specific PAOVSS aerobic and anoxic P uptake rates obtained in the 15 batch tests over the VSS concentration range from 2.7 to 11.2 g VSS/ℓ are shown in Figs. 6.28a and b and Figs. 6.29a and b respectively. For Batch Tests 1 to 14, aerobic and anoxic P uptake rates vary over quite a wide range from 5.5 to 13.4 mg P/(g VSS.h) or 22.6 to 54.1 mg P/(g PAOVSS.h) and 0.5 to 4.7 mg P/(g VSS.h) or 1.9 to 19.1 mg P/(g PAOVSS.h) respectively. Again no effect of VSS or initial acetate concentration on these rates is apparent. The average (and standard deviation) aerobic and anoxic P uptake rates obtained over the VSS concentration range (2.7-11.2 g VSS/ℓ) for the 13 batch tests with acetate dose (excluding the excess acetate dose BT14 in which anoxic P release was observed) are 10.0±2.3 mg P/(g VSS.h) or 40.0±7.6 mg P/(g PAOVSS.h) and 3.0±1.1 mg P/(g PAOVSS.h) or 11.6±4.7 mg P/(g PAOVSS.h) respectively.

Figures 6.28a and b: Aerobic VSS specific maximum P uptake rate mg P/(g VSS.h) (Fig. 6.28a, left) and PAOVSS specific maximum P uptake rate mg P/(g PAOVSS.h) (Fig. 6.28b, right).

Figures 6.29a and b: Anoxic VSS specific maximum P uptake rate mg P/(g VSS.h) (Fig. 6.29a, left) and PAOVSS specific maximum P release rate mg P/(g PAOVSS.h) (Fig. 6.29b, right).

6.7.5 Fermentation of readily biodegradable organics (RBO)

For comparison, also shown in Figs. 6.25 to 6.26 are the specific VSS and specific PAOVSS anaerobic soluble COD uptake and P release rates for the Batch Test 15 with wastewater addition (○). Although shown in Fig. 6.25, it is not correct to compare the soluble COD “uptake” rate of BT15 with the specific VSS and specific PAOVSS anaerobic acetate (as COD) uptake rate because the wastewater soluble readily biodegradable organics (RBO) are first hydrolysed and fermented to volatile fatty acids (VFAs) by the OHOs in the anaerobic reactor before the VFAs can be taken up
by the PAOs. So the decrease in soluble (0.45 µm filtered) COD concentration is associated with the hydrolysis/fermentation of RBO to VFAs by OHOs. However, the specific VSS or specific PAOVSS anaerobic P release rate (Fig. 6.26) is comparable because this rate is associated with the uptake of the VFAs by the PAOs. It can be seen that the initial P release rate with wastewater RBO is much slower (only ~1/3rd) than with acetate (Figure 6.26). This is because the hydrolysis rate of RBO by OHOs is much slower than the acetate uptake rate by PAOs so in BT15 the P release rate is constrained by the production rate of VFAs by OHOs. Also, the P release per COD uptake ratio is comparable because the soluble COD reduction rate is linked to the P release rate via the RBO fermentation and VFA uptake rates. The P release per COD uptake ratio for BT15 was 0.59, which is close to the average of 0.57 obtained for BTs 1 to 14.

The hydrolysis/fermentation rate of readily biodegradable organics (RBO) to acetate mediated by the OHOs, and the associated P release by the PAOs from the acetate uptake, is modelled as a first order rate with respect to the RBO COD concentration (Wentzel et al., 1985, 1992; Henze et al., 1995), viz.

\[ \frac{dS_{bs}}{dt} = K_{ferm} S_{bs} f_{avOHO} X_v \text{ mg COD/(ℓ.d)} \]  

\[ \frac{dP}{dt} = C_{sp} K_{ferm} S_{bs} f_{avOHO} X_v \text{ mg P/(ℓ.d)} \]

where

\[ K_{ferm} = \text{OHO hydrolysis/fermentation rate of RBO – ℓ/(mg OHOVSS.d)} \]

\[ C_{sp} = \text{P release per acetate taken up ratio (mg P/mg COD)} \]

The average \( K_{ferm} \) rate and \( C_{sp} \) ratio measured by Wentzel et al. (1985) in CAS BNR systems fed real wastewater were 0.06 ℓ/(mg OHOVSS.d) and 0.5 mg P/mg COD respectively.

Applying the Wenzel et al. (1985) fermentation model to the results of BT15 yielded \( K_{ferm} = 0.036 \) ℓ/(mg OHOVSS.d) based on the measured soluble COD – time profile and 0.027 ℓ/(mg OHOVSS.d) based on the P release – time profile. This is considerably lower than the rate of 0.06 measured by Wentzel et al. (1985). Considering the wide scatter in the other kinetic rates measured in this investigation, and in the individual \( K_{ferm} \) rates measured by Wentzel et al. (1985), it is possible that the \( K_{ferm} \) rate measured in this one batch test with wastewater is a low value in a population with an average around 0.06 ℓ/(mg OHOVSS.d). In hindsight, fewer batch tests with acetate addition and more with wastewater addition should have been done so the effect of high VSS concentration on the fermentation rate could be established.

However, considering that…

1. The considerable scatter in the individual \( K_{ferm} \) rate measured by Wentzel et al. (1985) of which 0.06 ℓ/(mg OHOVSS.d) was the average,

2. the VSS concentration in the anaerobic reactor of MBR BNR systems are not much higher than in CAS BNR systems – from Ramphao et al. (2005) the VSS concentration in the anaerobic reactor will not be greater than half that in the aerobic reactor due to the diluting effect of the upstream reactor inflow. If the aerobic reactor is say 10 g VSS/ℓ, then the anaerobic reactor VSS concentration will be less than 5 g VSS/ℓ, which is not much greater than the VSS concentration in the anaerobic reactor of CAS systems;

3. the acetate uptake and P release rates by PAOs showed no VSS concentration effect up to 11 g VSS/ℓ, and

4. the OHOs, which mediate the RBO fermentation to acetate, showed no decrease in denitrification rate with VSS concentration,
...it is likely that the fermentation rate ($K_{ferm}$) by OHOs also is not affected by high VSS concentrations so that the rate observed in CAS systems also can be accepted for MBR systems.

### 6.7.6 Comparing kinetic rates with those of other investigations

The acetate uptake, P release and P uptake kinetic rates determined in this investigation are compared in Tables 6.11 and 6.12 with rates determined in previous investigations with batch tests on sludge harvested from different BNRAS systems at conventional BNR system VSS concentrations, such as multi-reactor UCT systems and sequencing batch reactors (SBR) fed real wastewater (mixed OHO and PAO cultures) or acetate only (enhanced PAO cultures).

#### Table 6.11: Average P removal kinetic rates and stoichiometry measured in this investigation compared with those from other investigations.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>This investigation</th>
<th>Other (Refs)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Anaerobic P release rate with acetate</td>
<td>mg P/(g VSS.h)</td>
<td>18.6 ± 3.1</td>
</tr>
<tr>
<td></td>
<td>mg P/(g PAOVSS.h)</td>
<td>72.3 ± 13.7</td>
</tr>
<tr>
<td>Anaerobic HAc uptake rate</td>
<td>mg HAcCOD/(g VSS.h)</td>
<td>29.8 ± 5.3</td>
</tr>
<tr>
<td></td>
<td>mg HAcCOD/(g PAOVSS.h)</td>
<td>116.1 ± 24.7</td>
</tr>
<tr>
<td>Aerobic P uptake</td>
<td>mg P/(g VSS.h)</td>
<td>10.0 ± 2.3</td>
</tr>
<tr>
<td></td>
<td>mg P/(g PAOVSS.h)</td>
<td>40.0 ± 7.6</td>
</tr>
<tr>
<td>Anoxic P uptake</td>
<td>mg P/(g VSS.h)</td>
<td>3.0 ± 1.1</td>
</tr>
<tr>
<td></td>
<td>mg P/(g PAOVSS.h)</td>
<td>11.6 ± 4.7</td>
</tr>
<tr>
<td>$P_{release}/HAc_{uptake}$</td>
<td>mg P/mg COD</td>
<td>0.57 ± 0.07</td>
</tr>
</tbody>
</table>


The VSS (mixed culture) specific rates from this investigation are within the rather wide range of VSS specific values in the literature for CAS BNR systems. The literature rates vary widely because the PAO active fraction ($f_{avPAO}$) differs widely depending on the P removal achieved in the systems relative to the influent COD concentration – the higher the P removal, the higher the $f_{avPAO}$ of the measured VSS and therefore the higher the VSS specific rate. The PAOVSS specific rates measured in this investigation are also within the range of PAOVSS specific rates measured on enhanced PAO cultures (Wentzel et al., 1989; Kuba et al., 1994; Kuba et al., 1997; Wachtmeister et al., 1997, Hu et al. 2003). The range of literature PAOVSS specific rates is wide and is due to some enhanced PAO culture systems being influent P limited rather than influent COD limited (i.e. had more P been available in the influent, then a higher P removal would have been achieved). Accepting that in this investigation (in which the two systems were not P limited), the $f_{avPAO}$ was estimated reasonably accurately, the conformity of the acetate uptake, P release and P uptake kinetic rates to literature rates from CAS BNR systems, confirms that of these PAO kinetic rates are not affected by high VSS concentrations in MBR systems. The investigation also highlights the importance of specifying kinetic rates of bioprocesses with respect to the actual biomass concentration mediating the particular bioprocess. Rates with respect to a lumped parameter like VSS are not comparable between different BNR systems and not useful for kinetic models. It also highlights the importance of recognizing that the specific conditions in the parent systems in which the biomass develops – conditions like P limitation or not, acetate or real wastewater feed – have a major affect on how the biomass behaves in batch tests, much more so than when solid-liquid separation is achieved by membranes or sedimentation.
Table 6.12: Summary of P removal stoichiometry and kinetic rates in a Batch Tests using sludge drawn from BNRAS, ENBNR and MBR systems

<table>
<thead>
<tr>
<th>Source</th>
<th>System</th>
<th>Aereobic max P release mgP/gVSS h</th>
<th>Aerobic max HAc consumption mgHAc/mgC h</th>
<th>$q_{Pal}$ Aerobic max P uptake mgP/mgC h</th>
<th>$q_{Pan}$ Anoxic max P uptake mgP/mgC h</th>
<th>$q_{Pal}/q_{Pan}$ %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Kuba et al. (1993)</td>
<td>SBRA/A</td>
<td>40</td>
<td>110</td>
<td>1,2</td>
<td>40</td>
<td>90</td>
</tr>
<tr>
<td></td>
<td>SBRA/O</td>
<td>40</td>
<td>80</td>
<td>1,3</td>
<td>45</td>
<td>40</td>
</tr>
<tr>
<td>Kerr-Jespersen et al. (1994)</td>
<td>Bio-DENPHO (A/A fixed film)</td>
<td>1,27±0,5</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Kuba et al. (1994)</td>
<td>SBRA/O</td>
<td>106</td>
<td>256</td>
<td>0,90</td>
<td>70</td>
<td>27</td>
</tr>
<tr>
<td></td>
<td>SBRA/A</td>
<td>62</td>
<td>125</td>
<td>1,16</td>
<td></td>
<td></td>
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<tr>
<td>Born et al. (1998)</td>
<td>SBRA/O</td>
<td>81</td>
<td></td>
<td>4,3</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>UCT WWTP</td>
<td>9,5</td>
<td></td>
<td>4,2</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Wachtmeister et al. (1997)</td>
<td>SBRA/O</td>
<td>70</td>
<td>10</td>
<td>15</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>UCT MBR</td>
<td>13</td>
<td>6</td>
<td>50</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Kuba et al. (1997)</td>
<td>Genemuiden BNR WWTP</td>
<td>5</td>
<td>7</td>
<td>1,9</td>
<td>5,7</td>
<td>1,2</td>
</tr>
<tr>
<td></td>
<td></td>
<td>9</td>
<td>31</td>
<td>0,8</td>
<td>3,8</td>
<td>1,6</td>
</tr>
<tr>
<td></td>
<td></td>
<td>19</td>
<td>50</td>
<td>1</td>
<td>13</td>
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<td></td>
<td></td>
<td>123</td>
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<td>16,9</td>
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<td></td>
<td></td>
<td>12</td>
<td></td>
<td>16,1</td>
<td>12,4</td>
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<td></td>
<td></td>
<td>96</td>
<td></td>
<td>17,9</td>
<td>13</td>
<td></td>
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<tr>
<td></td>
<td>SBRA/AO</td>
<td>44</td>
<td></td>
<td>15,1</td>
<td>2,8</td>
<td></td>
</tr>
<tr>
<td></td>
<td>SBRA/O</td>
<td>7</td>
<td></td>
<td>30</td>
<td>1,9</td>
<td></td>
</tr>
<tr>
<td></td>
<td>SBRA/AO</td>
<td>149</td>
<td></td>
<td>26,8</td>
<td>3</td>
<td>11</td>
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<tr>
<td>Lao et al. (2001)</td>
<td>SBRA/A</td>
<td>58,6</td>
<td>0,84</td>
<td>72,02</td>
<td>55,3</td>
<td></td>
</tr>
<tr>
<td></td>
<td>SBRA/O</td>
<td>57,6</td>
<td>0,83</td>
<td>45,5</td>
<td>15,4</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Local WWTP</td>
<td>10,1</td>
<td>0,37</td>
<td>6,65</td>
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<tr>
<td>Hu et al. (2003)</td>
<td>SBRA/A</td>
<td>18,1</td>
<td></td>
<td>9,8</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>SBRA/O</td>
<td>20,4</td>
<td></td>
<td>9,6</td>
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<tr>
<td>Shoji et al. (2003)</td>
<td>A2N system</td>
<td>17,78</td>
<td>29,27</td>
<td>1,39</td>
<td>9,77</td>
<td>2,89</td>
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<td></td>
<td>UCT MBR</td>
<td>29,58</td>
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</tbody>
</table>

6.7.7 Comparing PAO and OHO denitrification behaviour in this and with other investigations

Although anoxic P uptake and denitrification by PAOs can vary significantly with time in a particular system, from this experimental investigation and literature (Kuba et al., 1997; Ekama and Wentzel, 1999; Hu et al., 2002), it appears the magnitude of anoxic P uptake relative to the total (anoxic and aerobic) P uptake depends on mainly two factors for significant anoxic P uptake to take place, viz. (1) the duration of exposure of the activated sludge to aerobic and anoxic conditions, i.e. high anoxic and low aerobic mass fractions and (2) the nitrate load on the anoxic reactor in relation to its denitrification potential, i.e. high nitrate load which exceeds the denitrification potential so that nitrate is present in the anoxic reactor outflow. Different methods have been developed in order to determine the fraction of denitrifying PAOs, $X_{DNPAO}/X_{PAO}$ (Meinhold et al. 1999, Hu et al. 2003), the PAO contribution to denitrification (Hu et al. 2002) or the level of PAO activity in the anoxic zone (Wachmeister et al. 1997). With these methods, the relative magnitude of PAO denitrification can be determined for a specific sludge drawn from a BEPR system.

In this investigation, the PAO denitrification rate ($K_{PAO}$) was determined from the slopes of the nitrate profile where the first slope corresponded to $K_{PAO}+K_{2OHO}$ and the second flatter slope to the $K_{2OHO}$ only (Fig. 6.23). The PAO denitrification in the MBR system was calculated from the system P uptake and denitrification performance according to the method of Hu et al. (2002). This
method assumes that the nitrate concentration denitrified by PAOs is proportional to the % PHB utilised by the DNPAOs in the anoxic reactor which in turn is proportional to the % anoxic P uptake relative to the total anoxic and aerobic P uptake (calculated from P mass balances around the anoxic and aerobic reactors). So the nitrate concentration denitrified by the PAOs is the catabolic fraction \([(1-fcvYG,ANS)/2.86 = 6.2 \text{ mg COD/mg NO3-N}]\) of the influent RBO taken up by PAOs in the anaerobic reactor \((\Delta Sbsi = \text{PHB COD})\) multiplied by the % anoxic P uptake. The nitrate concentration denitrified by OHOs is the difference between the observed nitrate concentration denitrified in the anoxic reactor and the nitrate concentration denitrified by PAOs. With the concentration of nitrate denitrified by the OHOs and PAOs known, the VSS specific and PAOVSS and OHOVSS specific denitrification rates were calculated from the PAO and OHO active fractions \((\text{favPAO, favOHO})\).

Comparing the % anoxic P uptake determined for the MBR system with that measured in the batch tests over the same period in which the batch tests were conducted (Table 6.13a and b), it can be seen that this was much lower in the MBR system than in the batch tests. This was probably due to the low nitrate load on the anoxic reactor of the MBR system relative to its denitrification potential \(< 1 \text{ mg N/ℓ}, \text{Table 6.13}\), which indicates the anoxic reactor was under loaded with nitrate. Consequently, the PAOs had little opportunity to use the limited nitrate available in the anoxic reactor of the MBR system. In contrast, in the batch tests with excess acetate dosing during the anaerobic phase leading to high PHB and excess nitrate during the anoxic phase (both conditions not prevalent in the MBR system), the DNPAO can utilise their PHB with nitrate without competition from OHOs. So the conditions in the batch tests are not the same as the conditions in the MBR system. This is the reason for the large difference between the PAO denitrification rate \([\text{KPAO}, \text{mg NO3-N/(mg PAOVSS.d)}\) measured in the batch tests compared with those calculated from the MBR system performance results (Table 6.13). For the batch tests, the KPAO varied between 0.097 and 0.284 mg NO3-N/(mg PAOVSS.d) with a mean of 0.16 whereas in the MBR system the KPAO varied between 0.008-0.033 mg NO3-N/(mg PAOVSS.d) with a mean of 0.02. In contrast, the OHO denitrification rates \([\text{K2OHO}, \text{mg NO3-N/(mg OHOVSS.d)}\) measured in the batch tests were similar to those calculated from the MBR system performance results (Table 6.13). In the batch tests, the K2OHO varied between 0.182 and 0.454 mg NO3-N/(mg OHOVSS.d) with a mean of 0.33 and in the MBR system, it varied between 0.158 and 0.314 mg NO3-N/(mg PAOVSS.d) with a mean of 0.27. The batch test and MBR system K2OHO rates are neither statistically different from each other nor to those measured in the anoxic-aerobic batch tests described earlier.
Table 6.13a: Comparison of batch test 1 to 8 and MBR system OHO and PAO VSS specific denitrification rates.

<table>
<thead>
<tr>
<th>Batch Test Date</th>
<th>18/10/05</th>
<th>21/11/05</th>
<th>28/11/05</th>
<th>04/12/05</th>
<th>06/02/06</th>
<th>09/02/06</th>
<th>13/02/06</th>
<th>29/03/06</th>
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</thead>
<tbody>
<tr>
<td>Batch Test No</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>OHOActive fraction ($f_{aOHO}$) mg OHOVSS/mg VSS</td>
<td>1</td>
<td>0.125</td>
<td>0.073</td>
<td>0.136</td>
<td>0.136</td>
<td>0.148</td>
<td>0.148</td>
<td>0.148</td>
</tr>
<tr>
<td>PAOActive fraction ($f_{aPAO}$) mg PAOVSS/mg VSS</td>
<td>2</td>
<td>0.237</td>
<td>0.247</td>
<td>0.243</td>
<td>0.243</td>
<td>0.282</td>
<td>0.282</td>
<td>0.282</td>
</tr>
<tr>
<td>OHO Denitrification rate mg NO$_3$-N/(g VSS.h)</td>
<td>3</td>
<td>1.74</td>
<td>1.33</td>
<td>1.25</td>
<td>1.14</td>
<td>1.76</td>
<td>1.12</td>
<td>1.52</td>
</tr>
<tr>
<td>OHO Denitrification rate mg NO$_3$-N/(mg OHOVSS.d)</td>
<td>4</td>
<td>0.334</td>
<td>0.437</td>
<td>0.221</td>
<td>0.201</td>
<td>0.285</td>
<td>0.182</td>
<td>0.246</td>
</tr>
<tr>
<td>PAO Denitrification rate mg NO$_3$-N/(g VSS.h)</td>
<td>5</td>
<td>1.57</td>
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<td>1.14</td>
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<td>PAO Denitrification rate mg NO$_3$-N/(mg PAOVSS.d)</td>
<td>6</td>
<td>0.159</td>
<td>0.200</td>
<td>0.283</td>
<td>0.260</td>
<td>0.097</td>
<td>0.174</td>
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<td>MBR System</td>
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<tr>
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<td>15</td>
<td>17</td>
<td>18</td>
<td>18</td>
<td>22</td>
<td>22</td>
<td>22</td>
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<tr>
<td>Batch Period</td>
<td>2</td>
<td>13/10-27/10</td>
<td>11/11-28/11</td>
<td>12/11-28/11</td>
<td>29/11-19/12</td>
<td>31/01-14/02</td>
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<td>Anoxic P uptake %</td>
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<td>86.9</td>
<td>93.9</td>
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<td>Anx/Aer P uptake ratio %</td>
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<td>28.8</td>
<td>15.8</td>
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<td>OHO denitrification rate mg NO$_3$-N/(mg OHOVSS.d)</td>
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<td>0.239</td>
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<td>0.173</td>
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<td>PAO denitrification rate mg NO$_3$-N/(mg PAOVSS.d)</td>
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<tr>
<td>Anoxic reactor nitrate conc mg NO$_3$-N/l</td>
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<td>0.27</td>
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Table 6.13b: Comparison of batch test 9 to 15 and MBR system OHO and PAO VSS specific denitrification rates.

<table>
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<tr>
<th>Date</th>
<th>Batch No</th>
<th>OHO Active Fraction (f_OHO)</th>
<th>mg OHOVSS/mg VSS</th>
<th>PAO Active Fraction (f_PAO)</th>
<th>mg PAOVSS/mg VSS</th>
<th>OHO denitrification rate (mg NO3-N/(g VSS.h))</th>
<th>mg NO3-N/(g VSS.d)</th>
<th>PAO denitrification rate (mg NO3-N/(g VSS.h))</th>
<th>mg NO3-N/(g VSS.d)</th>
<th>Anoxic P uptake %</th>
<th>Aerobic P uptake %</th>
<th>Anoxic/Aer P uptake ratio</th>
<th>% = Anox/(Aer)(1/ηNO3)</th>
<th>MBR System</th>
<th>Wastewater Batch No</th>
<th>Anoxic Reactor Nitrate Conc mg NO3-N/ℓ</th>
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<tr>
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<td>9</td>
<td>0.095</td>
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<td>0.028</td>
<td>0.028</td>
<td>0.228</td>
<td>15.2</td>
<td>94.8</td>
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<td>0.010</td>
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<td>0.007</td>
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<td>14.8</td>
<td>85.2</td>
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<tr>
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<td>1.16</td>
<td>0.25</td>
<td>14.8</td>
<td>85.2</td>
<td>0.217</td>
<td>0.022</td>
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<tr>
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<td>0.159</td>
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<td>85.2</td>
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<tr>
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<td>1.16</td>
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<td>14.8</td>
<td>85.2</td>
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<td>0.159</td>
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<td>85.2</td>
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<td>0.217</td>
<td>0.022</td>
<td>0.04</td>
<td>0.33</td>
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</table>

Batch Test No: 9, 10, 11, 12, 13, 14, 15
In a previous investigation comparing conventional BNRAS systems with external nitrification (EN) BNRAS systems, Hu et al. (2002) observed a similar large difference between the $K_{\text{PAO}}$ and $K_{\text{2OHO}}$ rates as that in the MBR system of this investigation. In the conventional BNRAS systems $K_{\text{2OHO}}$ varied from 0.05 to 0.32 mg NO$_3$-N/(mg OHOVSS.d) with a mean of 0.14 while $K_{\text{PAO}}$ varied from 0.005 to 0.071 mg NO$_3$-N/(mg PAOVSS.d) with a mean of 0.028. In the ENBNRAS systems, which promoted anoxic P uptake BEPR due to its high anoxic mass fraction and overloaded anoxic reactor, the $K_{\text{PAO}}$ varied from 0.029 to 0.071 mg NO$_3$-N/(mg PAOVSS.d) with a mean of 0.051. The $K_{\text{2OHO}}$ varied from 0.122 to 0.176 mg NO$_3$-N/(mg OHOVSS.d) with a mean of 0.15. So the specific PAOVSS denitrification rate of the PAOs on internally stored PHB was 3 to 5 times slower than that of the OHOs on slowly biodegradable organics (SBO) in conventional and EN BNR systems. In this investigation, the specific PAOVSS denitrification rate of the PAOs on internally stored PHB was 14 times lower than that of the OHOs on slowly biodegradable organics (SBO) in the MBR system.

During the investigation, in the MBR system, the PAOs achieved an average % anoxic P uptake of 9% and contributed 11% to the total denitrification. In the batch tests, the PAOs achieved an average % anoxic P uptake of 37% and contributed 22% to the total denitrification. The relationship between the proportion of DNPAOs to total PAOs ($X_{\text{DNPAO}}/X_{\text{PAO}}$) and the PAO contribution to the total denitrification is shown plotted in Fig. 6.30. As expected, the greater the proportion of DNPAOs, the greater the % denitrification by PAOs. For the results of this investigation, the $X_{\text{DNPAO}}/X_{\text{PAO}}$ ratio varied from 8 to 61% while the contribution by the PAOs to denitrification varied from 12 to 25%. This indicates, as found by Hu et al. (2002), that the contribution by the PAOs to denitrification was small compared with that of the OHOs. The large reduction in P removal (20 to 30%) with significant anoxic P uptake BEPR compared with aerobic P uptake BEPR appears to be a high price to pay for the low additional denitrification.

![Figure 6.30: Denitrification by PAOs (as a percentage of the total denitrification) versus proportion of denitrifying PAO (as a percentage of the total PAOs, $X_{\text{DNPAO}}/X_{\text{PAO}}$).](image)

Finally, with regard to the nitrification rates obtained during the aerated part of the anaerobic batch tests, VSS specific nitrification rates [mg FSA/(g VSS.h)] with VSS concentration are shown in Fig. 6.31. As found in the aerobic nitrification tests above, as the VSS concentration in the batch tests increases, the specific nitrification rate decreases. This confirms the VSS concentration effect on the VSS specific nitrification rate.
6.8 CONCLUSIONS

To assess the impact of high VSS concentration in membrane bioreactor biological nutrient removal (BNR) activated sludge (AS) systems on the bioprocess kinetic rates that mediate biological N and P removal, two identical (except for the hydraulic retention time) parallel laboratory scale University of Cape Town (UCT) nitrification denitrification (ND) biological excess phosphorus removal (BEPR) systems fed the same real wastewater were operated for 450 days, one at a low VSS concentration (3 g VSS/l) and solid liquid separation with a secondary settling tank (CAS system), the other at a high VSS concentration (13 g VSS/l) and solid liquid separation with submerged panel membranes (MBR system). From the BNR performance of these two systems and from aerobic, anoxic-aerobic and anaerobic-anoxic-aerobic batch tests on sludge harvested from the two systems the following conclusions were drawn.

Both systems achieved similar in N removals (MBR 83%, CAS 81%). Nitrification was complete in both systems – effluent free and saline ammonia (FSA) concentration from the MBR system was 0.7 mg FSA-N/l and from the CAS system 0.9 mg FSA-N/l. Denitrification was better in the MBR system (effluent nitrate MBR 18.0 mg NO₃-N/l and CAS 20.0 mg NO₃-N/l) due to the negligible impact of the dissolved oxygen in the recycle to the anoxic reactor at the high VSS concentration of the MBR system. The P removal in the MBR system (22.5 mg P/l) was higher than that in the CAS system (17.4 mg P/l). This was due to the recycle of nitrate from the anoxic reactor to the anaerobic reactor and greater anoxic P uptake in the CAS system due to the non-zero nitrate concentration in the anoxic reactor. This made the kinetic rates associated with BEPR measured in the batch tests incomparable between the two systems.

In order to assign the observed bioprocess kinetic rates to the organism group mediating it, the measured VSS concentration in the batch tests was fractionated into ordinary heterotrophic (OHO) and phosphorus accumulating (PAO) organism active concentrations by applying the steady state BNR AS model to the measured performances of the MBR and CAS systems. Due to the higher sludge production in the MBR system [0.31 (g VSS/d)/(g COD/d)] than in the CAS system [0.20 (g VSS/d)/(g COD/d)], the influent unbiodegradable particulate COD fraction (f₃up) of the MBR system was higher (0.241) than that of the CAS system (0.084). This affected the fractionation of the VSS into the ordinary heterotrophic organism (OHO) and phosphate accumulating organism (PAO) active fractions in the two systems with the steady state BNR models, which also affected the observed OHO and PAO VSS specific kinetic rates calculated from the results of the batch tests on sludge harvested from two systems. This affect was unavoidable because kinetic rates expressed in terms of VSS are not comparable between different BNR systems and steady state models.
aligned with and based on the same but simplified principles as kinetic models are the only interface between experimental systems and the kinetic models.

From the aerobic nitrification batch tests: (1) At the same low VSS concentration, the MBR system exhibited lower VSS specific ammonia utilisation rate (SAUR) and autotrophic nitrifier organism (ANO) maximum specific growth rates ($\mu_A$) than the parallel CAS system, apparently due to different selection pressures imposed by membranes and SSTs. (2) For the MBR system, as the VSS concentration increased, the SAUR and $\mu_A$ decreased, apparently due to ammonia and/or oxygen transfer limitations. (3) For the MBR system at the VSS concentration, as the initial ammonia concentration increased, the SAUR and $\mu_A$ increased, indicating possible ammonia transport limitation at increasing VSS concentration.

From the above, it was evident that the ANOs in the MBR and CAS systems exhibited different behaviour, apparently induced by different environments under which the ANOs develop. The reasons for this possibly are: (1) In CAS systems with SSTs, organism loss via the effluent occurs including ANOs. Therefore CAS system may select ANOs with higher maximum specific growth rates ($\mu_A$) than MBR systems. In the MBR system all the ANOs are retained, including slow growing ones. (2) At the high VSS concentrations in the MBR system, oxygen and ammonia transport limitations decreased the observed SAUR and $\mu_A$.

From the anoxic-aerobic batch tests, the OHOVSS specific denitrification rate by OHOs ($K_{2OHO}$) utilizing slowly biodegradable organics (SBO) obtained at different MBR system VSS concentrations (2.5-12 g VSS/l) and different initial nitrate concentrations ranging from 30 to 90 mg N/l showed no effect to initial nitrate concentration, in agreement with past work (van Haandel et al., 1981, Clayton et al., 1991; Ekama and Wentzel, 1999) and no effect to VSS concentration. From all the anoxic batch tests, the average $K_{2OHO}$ was 0.264 mg NO$_3$-N/(mg OHOVSS.d), which is very close to the average $K_{2OHO}$ rate reported in the literature for conventional (low VSS) BNR systems with SSTs, i.e. 0.255 from Ekama and Wentzel (1999).

From the anaerobic-anoxic-aerobic batch tests, the specific VSS and specific PAOVSS anaerobic acetate (as COD) uptake and P release rates showed no effect of VSS or initial acetate concentration. Also, the results obtained with different concentrations of acetate added showed the acetate uptake rate to be zero order with respect to acetate concentration, which is in agreement with literature studies (Wentzel et al., 1985, 1989). The P release to acetate uptake ratio also showed no effect with acetate dose and VSS concentration. The specific VSS and specific PAOVSS aerobic and anoxic P uptake rates also showed no effect of VSS concentration. The average PAOVSS specific anaerobic acetate uptake and P release rates and the aerobic P uptake rate obtained over the VSS concentration range were within the range of literature rates observed on enhanced PAO culture systems, confirming that within experimental variation, high VSS concentration does not affect the rates.

In the anaerobic-anoxic/aerobic batch tests with acetate uptake, the PAOs showed significantly higher anoxic P uptake and denitrification rates than in the MBR system itself, where high acetate and excess nitrate did not occur. In the former the PAOs denitrified 22% of the nitrate whereas in the MBR system only 11%. The OHOVSS specific denitrification rates were within the same 0.2 to 0.3 mg NO$_3$-N/(mg OHOVSS.d) range in all the batch with an anoxic phase. While the PAOVSS specific denitrification rate in the anaerobic-anoxic/aerobic batch tests was about half of the OHOVSS specific denitrification rate, in the MBR system, the PAOVSS specific denitrification rate was only 1/14th of the OHOVSS specific denitrification rate because the conditions in the anaerobic-anoxic/aerobic batch tests (high acetate and nitrate) were not prevalent in continuous flow BNR systems fed real wastewater. The large reduction in P removal resulting from significant
anoxic P uptake BEPR seems counter-productive for the very small PAO contribution to denitrification.

The results from this investigation show that the BNRAS steady state and kinetic models developed for low VSS concentration BNRAS systems with secondary settling tanks can be applied with reasonable confidence to predict the performance of high VSS concentration BNRAS systems with membranes, except for the maximum specific growth rate of the nitrifiers, which was observed to be significantly lower in the MBR system.

6.9 REFERENCES


MAHARAJ S., DU TOIT G.J.G., WENTZEL M.C. and BUX F. (2007) Molecular approaches to study the dynamics of nitrifying bacteria in a conventional activated sludge system and
membrane bioreactor. 4th International Water Association Leading-Edge Conference and Exhibition on Water & Wastewater Technology, Singapore, 3-6 June (Poster).


CHAPTER 7
CONCLUSIONS AND RECOMMENDATIONS

7.1 DESIGN OF BNR MBR SYSTEMS

Installing membranes for solid-liquid separation into biological nutrient removal (BNR) activated sludge (AS) systems makes a profound difference not only to the design of the BNR system itself, but also to the design approach for the whole wastewater treatment plant (WWTP). In multi-zone BNR systems with membranes in the aerobic reactor and fixed volumes for the anaerobic, anoxic and aerobic zones (i.e. fixed volume fractions), the mass fractions can be controlled (within a range) with the inter-reactor recycle ratios. This zone mass fraction flexibility is a significant advantage of membrane BNR systems over conventional BNR system with SSTs, because it allows changing the mass fractions to optimise biological N and P removal in conformity with influent wastewater characteristics and the effluent N and P concentrations required.

In contrast to conventional BNRAS systems with secondary settling tanks, the size of which is governed by organic (COD) load and system sludge age, the size of MBR BNR systems is governed by hydraulic load (peak wet weather flow, PWWF) and oxygen transfer rate (OTR). So for design of MBR systems accurate information on the membrane flux-trans-membrane pressure (TMP) relationship and oxygen transfer rate (OTR) and efficiency (OTE) are required. For PWWF to Average dry weather flow (ADWF) ratios ($f_q$) in the upper range ($f_q \sim 2.0$), aerobic mass fractions in the lower range ($f_{maer} < 0.60$) and high (usually raw) wastewater strengths, the indicated mode of operation of MBR BNR systems is as extended aeration WWTPs. However, the volume reduction compared with equivalent conventional BNR systems with secondary settling tanks will not be large (40-60%), but the cost of the membranes can be offset against sludge thickening and stabilisation costs. Moving from a flow unbalanced raw wastewater system to a flow balanced ($f_q = 1$) low (usually settled) wastewater strength system can double the ADWF capacity of the biological reactor, but the design approach of the WWTP changes away from extended aeration to include primary sludge stabilisation. The cost of primary sludge treatment then has to be paid from the savings from the increased WWTP capacity.

While some OTR, OTE and alpha values at different TSS concentrations (3 to 18 g TSS/l) were measured in this investigation on the laboratory MBR system, it is recommended that full scale tests are conducted to measure these parameters. This information is difficult to obtain because vendors of membranes tend to keep this information confidential.

7.2 KINETICS OF BNR IN MBR SYSTEMS

To assess the impact of high VSS concentration in membrane bioreactor biological nutrient removal (BNR) activated sludge (AS) systems on the bioprocess kinetic rates that mediate biological N and P removal, two identical (except for the hydraulic retention time) parallel laboratory scale University of Cape Town (UCT) nitrification denitrification (ND) biological excess phosphorus removal (BEPR) systems fed the same real wastewater were operated for 450 days, one at a low VSS concentration (3 g VSS/l) and solid liquid separation with a secondary settling tank (CAS system), the other at a high VSS concentration (13 g VSS/l) and solid liquid separation with submerged panel membranes (MBR system). From the BNR performance of these two systems and from aerobic, anoxic-aerobic and anaerobic-anoxic-aerobic batch tests on sludge harvested from the two systems the following conclusions were drawn.
From the aerobic nitrification batch tests: (1) At the same low VSS concentration, the MBR system exhibited lower VSS specific ammonia utilisation rate (SAUR) and autotrophic nitrifier organism (ANO) maximum specific growth rates ($\mu_A$) than the parallel CAS system, apparently due to different selection pressures imposed by membranes and SSTs. (2) For the MBR system, as the VSS concentration increased, the SAUR and $\mu_A$ decreased, apparently due to ammonia and/or oxygen transfer limitations. (3) For the MBR system at the VSS concentration, as the initial ammonia concentration increased, the SAUR and $\mu_A$ increased, indicating possible ammonia transport limitation at increasing VSS concentration.

From the above it was evident that the ANOs in the MBR and CAS systems exhibited different behaviour, apparently induced by different environments under which the ANOs develop. The reasons for this possibly are: (1) In CAS systems with SSTs, organism loss via the effluent occurs including ANOs. Therefore CAS system may select ANOs with higher maximum specific growth rates ($\mu_A$) than MBR systems. In the MBR system all the ANOs are retained, including slow growing ones. (2) At the high VSS concentrations in the MBR system, oxygen and ammonia transport limitations decrease the observed SAUR and $\mu_A$.

From the anoxic-aerobic batch tests, the OHOVSS specific denitrification rate by OHOs ($K_{2OHO}$) utilizing slowly biodegradable organics (SBO) obtained at different MBR system VSS concentrations (2.5-12 g VSS/l) and different initial nitrate concentrations ranging from 30 to 90 mg N/l showed no effect to initial nitrate concentration, in agreement with past work (van Haandel et al., 1981, Clayton et al., 1991; Ekama and Wentzel, 1999) and no effect to VSS concentration. From all the anoxic batch tests, the average $K_{2OHO}$ was 0.264 mg NO$_3$-N/(mg OHOVSS.d), which is very close to the average $K_{2OHO}$ rate reported in the literature for conventional (low VSS) BNR systems with SSTs, i.e. 0.255 from Ekama and Wentzel (1999).

From the anaerobic-anoxic-aerobic batch tests, the specific VSS and specific PAOVSS anaerobic acetate (as COD) uptake and P release rates showed no effect of VSS or initial acetate concentration. Also, the results obtained with different concentrations of acetate added showed the acetate uptake rate to be zero order with respect to acetate concentration, which is in agreement with literature studies (Wentzel et al., 1985, 1989). The P release to acetate uptake ratio also showed no effect with acetate dose and VSS concentration. The specific VSS and specific PAOVSS aerobic and anoxic P uptake rates also showed no effect of VSS concentration. The average PAOVSS specific anaerobic acetate uptake and P release rates and the aerobic P uptake rate obtained over the VSS concentration range were within the range of literature rates observed on enhanced PAO culture systems, confirming that within experimental variation, high VSS concentration does not affect the rates.

In the anaerobic-anoxic/aerobic batch tests with acetate uptake, the PAOs showed significantly higher anoxic P uptake and denitrification rates than in the MBR system itself, where high acetate and excess nitrate did not occur. In the former the PAOs denitrified 22% of the nitrate whereas in the MBR system only 11%. The OHOVSS specific denitrification rates were within the same 0.2 to 0.3 mg NO$_3$-N/(mg OHOVSS.d) range in all the batch with an anoxicic phase. While the PAOVSS specific denitrification rate in the anaerobic-anoxic/aerobic batch tests was about half of the OHOVSS specific denitrification rate, in the MBR system, the PAOVSS specific denitrification rate was only 1/14th of the OHOVSS specific denitrification rate because the conditions in the anaerobic-anoxic/aerobic batch tests (high acetate and nitrate) were not prevalent in continuous flow BNR systems fed real wastewater. The large reduction in P removal resulting from significant anoxic P uptake BEPR seems counter-productive for the very small PAO contribution to denitrification.
The results from this investigation show that biological N and P removal are not negatively affected by high suspended solids concentrations (up to 13 g VSS/P) typical of MBR activated sludge plants. Hence the BNRAS steady state and kinetic models developed for low VSS concentration BNRAS systems with secondary settling tanks can be applied with reasonable confidence to predict the performance of high VSS concentration BNRAS systems with membranes, except for the maximum specific growth rate of the nitrifiers, which was observed to be significantly lower in the MBR system and decreases as VSS concentration increases. However, for BNR system design and operation, because the maximum specific growth rate of nitrifiers varies considerably from one wastewater to another, it is regarded a wastewater characteristics rather than a kinetic constant, so it needs to be known for any BNR (MBR or conventional) system. For MBR BNR systems, in order to attain high reactor TSS concentration, sludge ages will usually be long enough for nitrification even with the reduced maximum specific growth rate of nitrifiers. Provided the models are given the appropriate input information, they will predict the performance of MBR BNR systems at high VSS concentration equally reliably (or unreliably) as that of conventional BNR systems at low VSS concentration with secondary settling tanks for solid liquid separation. This investigation indicated no reason to not build MBR BNR systems.

7.3 REFERENCES


