CRANFIELD UNIVERSITY

Bart Verrecht

Optimisation of a Hollow Fibre Membrane Bioreactor for Water Reuse

School of Applied Sciences
Department of Sustainable Systems
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PhD

Supervisor: Prof. Simon Judd
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This thesis is submitted in partial fulfilment of the requirements for the degree of Doctor of Philosophy

ABSTRACT

Over the last two decades, implementation of membrane bioreactors (MBRs) has increased due to their superior effluent quality and low plant footprint. However, they are still viewed as a high-cost option, both with regards to capital and operating expenditure (capex and opex). The present thesis extends the understanding of the impact of design and operational parameters of membrane bioreactors on energy demand, and ultimately whole life cost. A simple heuristic aeration model based on a general algorithm for flux vs. aeration shows the benefits of adjusting the membrane aeration intensity to the hydraulic load. It is experimentally demonstrated that a lower aeration demand is required for sustainable operation when comparing 10:30 to continuous aeration, with associated energy savings of up to 75%, without being penalised in terms of the fouling rate. The applicability of activated sludge modelling (ASM) to MBRs is verified on a community-scale MBR, resulting in accurate predictions of the dynamic nutrient profile. Lastly, a methodology is proposed to optimise the energy consumption by linking the biological model with empirical correlations for energy demand, taking into account of the impact of high MLSS concentrations on oxygen transfer.

The determining factors for costing of MBRs differ significantly depending on the size of the plant. Operational cost reduction in small MBRs relies on process robustness with minimal manual intervention to suppress labour costs, while energy consumption, mainly for aeration, is the major contributor to opex for a large MBR. A cost sensitivity analysis shows that other main factors influencing the cost of a large MBR, both in terms of capex and opex, are membrane costs and replacement interval, future trends in energy prices, sustainable flux, and the average plant utilisation which depends on the amount of contingency built in to cope with changes in the feed flow.

Keywords:

Aeration energy, Aeration intensity, Activated sludge modelling (ASM), Biokinetics, Capex, Cost sensitivity, Intermittent aeration, Large scale, Life cycle, Model-based energy optimisation, Opex , Sustainable operation, Small scale

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Awards

Award 'Best paper presented by a Young Water Professional':

Optimisation of energy consumption of a decentralised small scale MBR for urban reuse. B. Verrecht, T. Maere, L. Benedetti, R. Birks, S. Hills, E. Germain, P. Pearce, I. Nopens and S. Judd. Proceedings of 7th IWA World Congress on Water Reclamation and Reuse, 20-25 September 2009, Brisbane, Australia.

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BSM-MBR: a benchmark simulation model to compare control strategies for membrane bioreactors (2010) T. Maere, B. Verrecht, S. Moerenhout, S. Judd and I. Nopens. In preparation for submission to *Water Research*

Economical evaluation and operating experiences of a small scale MBR for non-potable reuse (2010). B. Verrecht, C. James, E. Germain, R. Birks, A. Barugh, P. Pearce and S. Judd. Submitted for publication in *Journal of Environmental Engineering*.

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Model-based energy optimisation of a small scale MBR for urban reuse (2010) B. Verrecht, T. Maere, L. Benedetti, I. Nopens and S. Judd, *Water Research* 44 (14),. 4047-4056

An aeration energy model for an immersed membrane bioreactor (2008) B. Verrecht, G. Guglielmi, J.W. Mulder, C. Brepols and S. Judd. *Water Research* 42 (19), 4761-4770

Book contribution

Co-author of the sections on biotreatment, design, biokinetic modelling and cost benefit analysis in *'The MBR book – 2nd edition'* (Judd and Judd, 2010, Elsevier).

Conference contributions

Validation of effluent quality and energy optimisation simulations for a small scale MBR (2010) Verrecht, B., Maere, T., Benedetti, L., Nopens, I. and Judd, S. Platform presentation and paper in conference proceedings at MDIW 2010, Trondheim, Norway, 27-30 June 2010

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ABBREVIATIONS AND SYMBOLS

ABBREVIATIONS

ADF	Average daily flow, in MLD (megalitres per day)
AOTR	Actual oxygen transfer rate, in $gO_2 d^{-1}$
ASM1	Activated sludge model no. 1
ASM2d	Activated sludge model no.2d
BOD_5	5-day biological oxygen demand, in $g.m^{-3}$
BOD_f	5 day biological oxygen demand of a sample filtered through 0.45 µm, in $g.m^{-3}$
BSM1_LT	Long term benchmark simulation model no. 1
CAPEX	Capital expenditures
CAS	Conventional activated sludge plant
CFU	Colony Forming Unit, in $CFU.(100\ ml)^{-1}$
CIP	Cleaning in place
COD	Chemical oxygen demand, in $g.m^{-3}$
COD_f	Chemical oxygen demand of a sample filtered through 0.45 µm, in $g.m^{-3}$
COD_{in}	Influent COD concentration, in $gCOD\ m^{-3}$
COP	Cleaning out of place
CST	Capillary suction time, in s
CSTR	Continuously stirred tank reactor
DO	Dissolved oxygen, in $mgO_2 l^{-1}$
DO_{bio}	DO concentration in the oxidation/nitrification tank, in $mg\ l^{-1}$
DO_{mbr}	DO concentration in the membrane tank, in $mg\ l^{-1}$
EQI	Effluent quality index, in $kg\ PU.d^{-1}$
FS	Flat sheet
GAC	Granular activated carbon
HF	Hollow fibre
HRT	Hydraulic retention time
LCA	Life cycle analysis
LCC	Life cycle costing
LMH	Litres per m^2 per bar
MBR	Membrane bioreactor
MLD	Megalitres per day
MLSS	Mixed liquor suspended solids, in $g.m^{-3}$

NPV	Net present value
OPEX	Operational expenditures
OTE	Oxygen transfer efficiency, in m^{-1}
OTE_{coarse}	Coarse bubble oxygen transfer efficiency, in m^{-1}
OTE_{fine}	Fine bubble oxygen transfer efficiency, in m^{-1}
PDF	Peak daily flow, MLD
PE	People equivalent
PLC	Programmable Logic Controller
PU_x	Pollution unit for effluent component x, in $kg.d^{-1}$
PVDF	Polyvinylidene Fluoride
SCA	Scenario analysis
SCADA	Supervisory Control And Data Acquisition
SOTR	Standard oxygen transfer rate, in $gO_2 \cdot d^{-1}$
SRT	Solids retention time, in d
SS	Suspended solids, in $mg \cdot l^{-1}$
TMP	Transmembrane pressure, bar
US EPA	United States Environmental Protection Agency
UV	Ultraviolet

SYMBOLS

A	Membrane surface area, in m^2
A_x	Open membrane module x-sectional area, in m^2
$b_{h,20}$	Heterotrophic decay rate at 20 ˚C, in d^{-1}
$b_{n,20}$	Autotrophic decay rate at 20˚C, in d^{-1}
b_{PAO}	Rate constant for lysis of X_{PAO}, in d^{-1}
$C_{rsat\_average}$	Average dissolved oxygen saturation concentration, in $gO_2 \cdot m^{-3}$, for clean water in an aeration tank for a given temperature T
C_{ssat}	Dissolved oxygen saturation concentration, in $gO_2 \cdot m^{-3}$, in clean water at 20 °C and 1 atm
C_{tank}	Actual oxygen concentration in the aeration tank, in $gO_2 \cdot m^{-3}$
d_f	Hollow fibre outside diameter, in m
E_A	Specific energy demand for aeration, in kWh m^{-3}
F	Correction factor for fouling of the air diffusers (1 for clean diffusers)
f	Endogenous residue
f_{anox}	Anoxic fraction of the biotank

f_{bp}	Fraction of slowly biodegradable COD
f_{bs}	Fraction of readily biodegradable COD
F_{coarse}	Correction factor for fouling of the coarse bubble air diffusers (1 for clean diffusers)
f_{cv}	COD/VSS ratio, in gCOD $gVSS^{-1}$
F_{fine}	Correction factor for fouling of the fine bubble air diffusers (1 for clean diffusers)
f_n	N-content in VSS, in gN $gVSS^{-1}$
f_{nus}	Soluble unbiodegradable fraction of influent TKN
f_{up}	Fraction of un-biodegradable COD
f_{us}	Fraction of soluble un-biodegradable COD
i	Discount rate, in %
J	Permeate flux, in l m^{-2} hr^{-1}
J_0	Intercept of the J vs. U curve, in $l.m^{-2}.h^{-1}$
J_c	Critical flux, LMH (litres per m^2 per hour)
J_{net}	Netto flux, LMH
J_{sust}	Sustainable flux, LMH
J_x	Flux, $l.m^{-2}.h^{-1}$
K	Permeability, $LMH.bar^{-1}$
$K_{den1,20}$	Denitrification rate, readily biodegradable COD at 20˚C, in gN-NO_3 $gVSS^{-1}$ d^{-1}
$K_{den2,20}$	Denitrification rate, slowly biodegradable COD at 20˚C, in gN-NO_3 $gVSS^{-1}$ d^{-1}
$K_{n,20}$	Half-saturation constant for ammonia nitrogen at 20˚C, in gN m^{-3}
K_O	Half saturation coefficient for oxygen, in $mgO_2 \cdot l^{-1}$
$K_{s,20}$	Half-saturation constant for readily biodegradable COD at 20˚C, in gCOD m^{-3}
$L_{membrane}$	Length of the membrane module, in m
L_{tank}	Tank length, in m
m	Slope of the J vs U curve
m_o	Mass flow of dissolved oxygen, in $gO_2.d^{-1}$
$N_{a,out}$	Effluent ammonia concentration at process temperature, in gN m^{-3}
NH_4-N	Ammonia-nitrogen, in $mgN \cdot l^{-1}$
N-$NO_{3,in}$	Influent nitrate-N concentration, in gN m^{-3}
NO_2-N	Nitrite-nitrogen, in $mgN \cdot l^{-1}$

NO_3-N	Nitrate-nitrogen, in mgN·l^{-1}
O_{air}	Fraction of oxygen in the air, in %
ON	Organic nitrogen, in mgN·l^{-1}
p	Blower inlet pressure, in Pa
PE_{sludge}	Pumping energy required per unit of sludge, in kWh.m^{-3}
pH	Process pH
PO_4-P	Ortho-phospate, in mgP·l^{-1}
P_{sludge}	Power required for sludge pumping, in kW
P_x	Sludge production, in kg.d^{-1}
Q_A	Aeration rate, in m^3 hr^{-1}
$Q_{air,coarse}$	Coarse bubble airflow rate in Nm^3·h^{-1}
$Q_{air,fine}$	Fine bubble airflow rate in Nm^3·h^{-1}
Q_E	Effluent flow, in m^3.d^{-1}
Q_I	Influent flow, in m^3.d^{-1}
Q_{MR}	Membrane recirculation flow, in m^3.d^{-1}
Q_{NR}	Nitrate recirculation flow, in m^3.d^{-1}
Q_p	Permeate flow rate in m^3 hr^{-1}
Q_W	Wastage flow, in m^3.d^{-1}
r_m	Recirculation ratio to anoxic zone
S_A	Fermentation products, considered to be acetate, in mgCOD·l^{-1}
SAD_m	Specific aeration demand per unit of membrane area, in Nm^3.m^{-2}.h^{-1}
SAD_p	Specific aeration demand, in m^3 h^{-1} air per m^3 h^{-1} permeate product
S_F	Fermentable, readily biodegradable organic substrates, in mgCOD·l^{-1}
S_I	Inert soluble organic material, in mgCOD·l^{-1}
S_{NH4}	Ammonium plus ammonia nitrogen, in mgN·l^{-1}
S_{PO4}	Inorganic soluble phosphorus, primarily ortho-phosphates, in mgP·l^{-1}
T	Temperature of the mixed liquor/air, in °C or ºK
tCOD	Total COD, in mg·l^{-1}
TKN	Total Kjeldahl nitrogen, in mgN·l^{-1}
TKN_{in}	Influent total Kkjeldahl nitrogen concentration, in gN m^{-3}
TN	Total nitrogen, in mgN·l^{-1}
TON	Total oxidised nitrogen, in mgN·l^{-1}
TP	Total phosphorous, in mgP·l^{-1}
U	In-module air upflow gas velocity, in m.s^{-1}
V	Total biotank volume, in m^3

W_{tank}	Tank width, in m
X_H	Heterotrophic organisms, in $mgCOD \cdot l^{-1}$
X_I	Inert particulate organic material, in $mgCOD \cdot l^{-1}$
X_S	Slowly biodegradable substrates, in $mgCOD \cdot l^{-1}$
y	Aerator depth, in m
y	Membrane aerator nozzle depth, in m
y_{fine}	Fine bubble aerator nozzle depth, in m
Y_h	Heterotrophic yield, in $gVSS\ gCOD^{-1}\ d^{-1}$
Y_n	Autotrophic yield, in $gVSS\ gCOD^{-1}\ d^{-1}$
Y_{PO}	Polyphosphate (PP) requirement for storage of poly-hydroxy-alkanoates (PHA), in $gP \cdot (gCOD)^{-1}$
α	Clean-to-process water correction factor for oxygen transfer
β	Salinity-surface tension correction factor
β_x	Weighting factor for effluent component x, dimensionless
δ	Flat sheet membrane channel separation, in m
Δh	Head loss, in m
ΔK	Permeability decline, $LMH.bar^{-1}.h^{-1}$
λ	Aerator constant (~1.4)
$\mu_{h,max,20}$	Heterotrophic maximum growth rate at 20˚C, in d^{-1}
$\mu_{n,max,20}$	Autotrophic maximum growth rate at 20˚C, in d^{-1}
μ_{PAO}	Maximum growth rate of X_{PAO}, in d^{-1}
ξ_B	Blower efficiency, dimensionless
ξ_p	Pump efficiency, dimensionless
ρ_{air}	Density of air at standard conditions, in $kg \cdot m^{-3}$
ρ_{air}	Air density, in $kg\ m^{-3}$
ρ_{sludge}	Sludge density, in $kg.m^{-3}$
ς	Blower efficiency
φ	Module packing density, in m^{-1}
ψ	Correction factor for effect of temperature on oxygen transfer: $\psi = 1.024^{T-273}$
ω	α-factor exponent coefficient, dimensionless

CHAPTER 1

INTRODUCTION

1 INTRODUCTION

1.1 BACKGROUND

Membrane bioreactors (MBR) for wastewater treatment combine biological treatment with a membrane separation step for solid-liquid separation, which replaces the sedimentation step in the conventional activated sludge process (CAS) (Stephenson *et al.*, 2000; Judd, 2006, 2008). Due to the synergy between biotreatment and membrane filtration, MBRs offer several widely acknowledged advantages over CAS. The ultrafiltration (UF) or microfiltration (MF) membranes have an effective pore size < 1µm and filtration of the mixed liquor ensures an effluent free of solids, pathogenic bacteria and viruses, and no further disinfection is required. As the settling properties of the mixed liquor are not as important as in CAS, which relies on sedimentation for clarification of the effluent, the hydraulic (HRT) and solids retention time (SRT) can be controlled independently. This allows operation at longer SRTs and higher MLSS concentrations, and more stable and complete nitrification can be reached because of favourable conditions for the growth of autotrophic bacteria (Munz *et al.*, 2008). Another important benefit of operation at elevated MLSS concentrations is process intensification resulting in reduced plant footprint and sludge production.

Since the advent of the submerged systems developed by Kubota and Zenon in the late 1980's and early 1990's, MBR technology has achieved significant commercial success worldwide and on increasingly large scales, as recently reviewed by a number of authors (Yang *et al.*; 2006; Melin *et al.*, 2006; Lesjean *et al.*, 2009; Santos and Judd, 2010; Kraume and Drews, 2010; Judd and Judd, 2010). Besides the above mentioned benefits over CAS, the main drivers for this success include: more stringent effluent legislation, local water scarcity, incentives to encourage recycling (Section 3.1), decreasing investment cost (Section 7.1) and increasing confidence in and acceptance of MBR technology, resulting in ever decreasing time-to-market for new market entrants (Santos and Judd, 2010). The separate control of the HRT and SRT allows further optimisation of the biological process, e.g. for enhanced nutrient removal (Kraume *et al.*, 2005; Vocks *et al.*, 2005, Daigger *et al.*, 2010). Several studies have also indicated the potential of MBR as pre-treatment for reverse osmosis (RO) due to its superior and consistent effluent quality (Tao *et al.*, 2005; Qin *et al.*, 2006), and it is considered the most robust technology for effluent reuse (Winward *et al.*, 2008), which is becoming increasingly attractive in water scarce regions and/or densely populated urban areas. MBR technology is thus gaining momentum for urban non-potable reuse purposes, such as toilet flushing and irrigation (Boehler *et al.*, 2007; Clerico, 2007; Meuler *et al.*, 2007), as well as cooling tower make-up water (Clerico, 2007; Ogoshi *et al.*, 2001). This has resulted in a growing number of small scale MBRs ($<200m^3.d^{-1}$) for niche reuse applications, ranging from single household (Matulova *et al.*, 2010; Abegglen *et al.*, 2008), to holiday resort buildings/hotels (Boehler *et al.*, 2007; Meuler *et al.*, 2007), apartment/office blocks (Clerico, 2007), and cruise ships (Chapter 3).

However, the advantages of membrane bioreactors come at a cost. Their hydraulic potential is limited by membrane fouling, the deposition of solids onto the membrane surface or in the membrane pores, which leads to a loss of permeability. Membrane fouling has formed the focus of numerous studies, and has been extensively reviewed by Le-Clech *et al.* (2006) and Meng *et al.* (2009). In submerged MBRs, fouling is controlled by inducing shear at the membrane surface through membrane aeration. This inherent need for membrane aeration results in significantly higher energy demand compared to CAS (Fenu *et al*., 2010), leading to higher operational expenses (opex), while membrane installation and replacement costs contribute significantly to capital expenditures (capex). They are thus still viewed as a high-cost option, both with regards to capex and opex. However, Brepols *et al.* (2010) have recently indicated that MBR technology is becoming cost-competitive when compared to CAS followed by tertiary filtation and disinfection. To further the success of this technology, it is essential to better understand the determining factors for capex and opex of membrane bioreactors, both on the small and large scale, as this could lead to further cost reductions.

1.2 AIMS AND OBJECTIVES

The present thesis reports work funded by Thames Water. This thesis aims to assess the suitability of small scale membrane bioreactor technology for urban reuse purposes, and looks into the impact of design and operational parameters on energy consumption and costing of MBRs on both small and large scale. Accordingly, the following objectives were identified:

1. To assess the current knowledge on aeration in membrane bioreactors and identify areas that require further understanding.

2. To study the relationship between membrane aeration and fouling, from (a) a general modelling approach benchmarked against two full-scale plants, and (b) a practical study, applying the outcomes to optimisation of MBR energy consumption.

3. To assess and verify the applicability of activated sludge (ASM) modelling to MBRs, and gain a better understanding about the impact of hydraulic and biological conditions in MBR on biological performance.

4. To determine and quantify the parameters impacting on capital investment and operating and maintenance (O&M) costs of MBRs on (a) a small community-scale installation, and (b) a large centralised plant.

1.3 THESIS PLAN

This thesis is presented in paper format, with all intellectual input provided by the first author with Professor Simon J. Judd as the corresponding author. All experimental work was undertaken by the first author, except for the determination of basic water quality parameters, which was carried out by the Thames Water labs in Reading.

Chapter 2 identifies the impact of aeration on membrane fouling, energy demand and costing of membrane bioreactors, according to a review of the literature.

Chapters 3 to 7 cover the technical content of the thesis. Chapter 3 describes a small full scale membrane bioreactor for urban reuse and determines the main factors influencing operational expenditures (opex) for small scale systems (Submitted to *Journal of Environmental Engineering* (2010)). Research and experiments supporting chapters 5 and 7 have been carried out on this case study.

Chapter 4 introduces an aeration energy model for an immersed membrane bioreactor based on a flux:aeration relationship identified from literature (Published in *Water Research* (2008), 42 (19), 4761-4770: Verrecht, B., Judd, S., Guglielmi, G., Brepols, C., Mulder, J. W.: An aeration energy model for an immersed membrane bioreactor)

Chapter 5 quantifies the benefits of intermittent aeration for the small scale MBR, and contrasts the conclusions against heuristic data obtained from large HF MBR (Submitted for publication in *Water Science and Technology* (2010): Verrecht, B., James, C., Germain, E., Judd, S. Intermittent aeration of a hollow fibre MBR)

Chapter 6 presents a model-based energy optimisation of the small-scale MBR and thus provides a full scale validation of the ASM models to MBR. (Published in *Water Research* (2010), 44 (14, 4047-4056): Verrecht, B., Maere, T., Benedetti, L., Nopens, I., Judd, S. Model-based energy optimisation of a small-scale decentralised membrane bioreactor for urban reuse).

Chapter 7 discusses and quantifies the main factors impacting both capital and operational expenditures for large scale HF MBRs, based on a modelling approach (Accepted for publication in *Water Research* (2010): Verrecht, B., Maere, T., Nopens, I., Brepols, C., Judd, S. The cost of a large-scale hollow fibre MBR).

The thesis road map (Figure 1-1) summarises the development of this thesis. Conclusions are reached in chapter 8, in which further work is also proposed. A design methodology for an immersed MBR is developed in Appendix A.

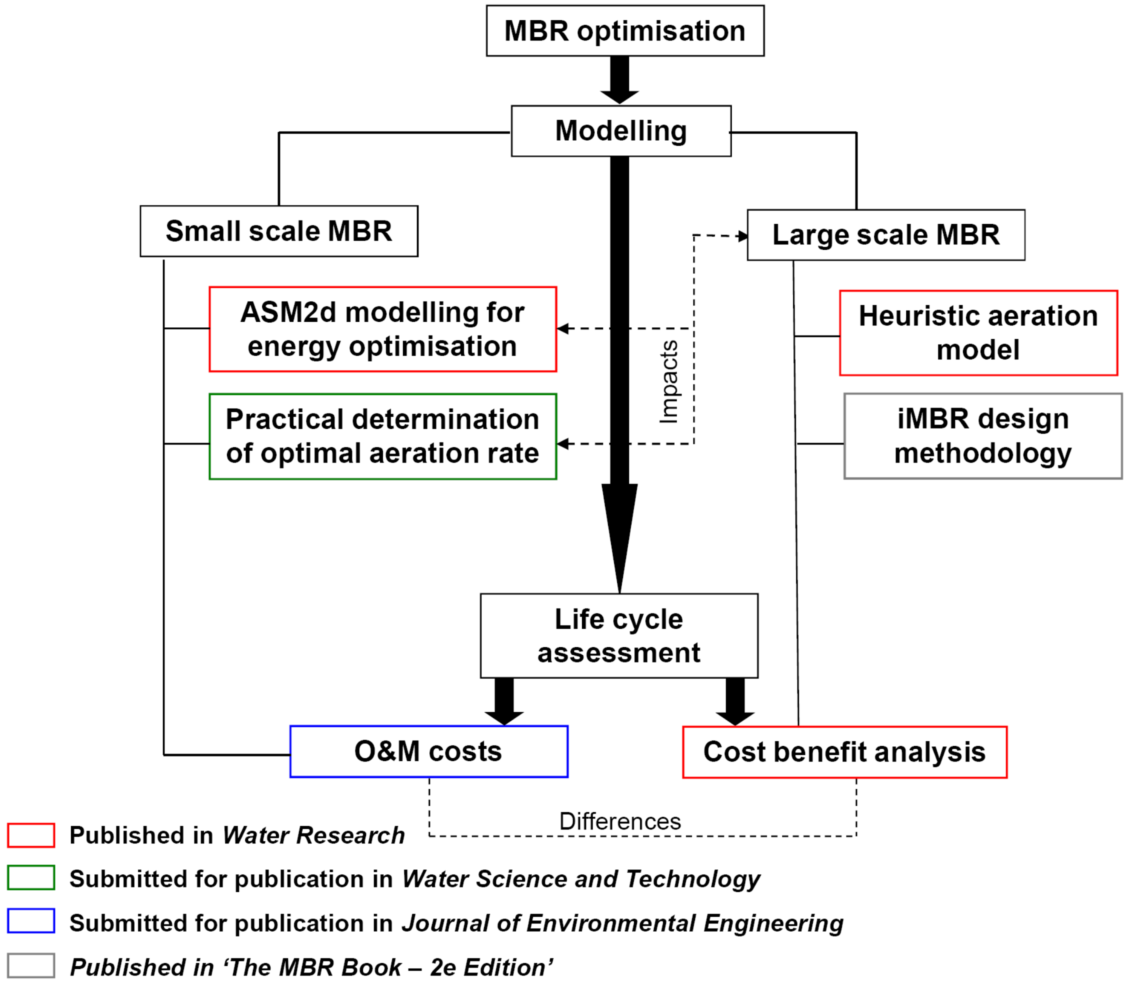

Figure 1-1: Thesis road map

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CHAPTER 2

SCOPING LITERATURE REVIEW:

IMPACT OF AERATION ON FOULING, ENERGY DEMAND AND COSTING OF MEMBRANE BIOREACTORS

2 SCOPING LITERATURE REVIEW: IMPACT OF AERATION ON FOULING, ENERGY DEMAND AND COSTING OF MEMBRANE BIOREACTORS

2.1 INTRODUCTION

Over the last decade, membrane bioreactors have gained widespread acceptance as costs have decreased and reference sites increased in size and number. Full scale applications are now plentiful (Lesjean *et al.*, 2009; Judd and Judd, 2010), and the worldwide market is expected to grow at a compound annual growth rate of 10.5 %, increasing in value from $296.0 million in 2008 to $488.0 million by 2013 (Kraume and Drews, 2010). However, as with all membrane processes, their performance is ultimately viewed as being constrained by membrane fouling, which has formed the focus of considerable research efforts over the past 15 years. In an extensive review on this topic by Le-Clech *et al.* (2006), the factors affecting membrane fouling were classified into four groups: membrane materials, biomass characteristics, feedwater characteristics and operating conditions. Complex interactions between these aspects govern the fouling process, which complicates its understanding. A complete overview of the impact of these factors is outside the scope of this literature review; a recent review by Meng *et al.* (2009) provides recent developments with regards to membrane fouling behaviour and fouling factors (Table 2-1).

Long-term sustainable operation of MBRs evidently demands effective fouling suppression through hydraulic, biology and chemical control (Meng *et al.*, 2009). Chemical fouling control strategies that have attracted attention over the last few years include the addition of powdered activated carbon (PAC), which reduces the EPS concentration and cake layer resistance (Ying and Ping, 2006) as well as irremovable fouling (Ng *et al.*, 2006). Remy *et al.* (2009, 2010) found dosing PAC at low concentrations (0.5 $g.l^{-1}$ MLSS) to increase the critical flux, the flux beyond which fouling becomes significant (Section 2.2.1), by 10%. Promising research has also been carried out on the addition of bespoke flux enhancing/fouling reducing chemicals: Koseoglu *et al.* (2008) reported a critical flux increase of up to 46% through addition of cationic polymers. Whilst an increased critical flux reduces the membrane area requirement - a significant contributor to capital cost - the authors acknowledged that further research into biotoxicity and larger scale tests would be necessary. Other chemical approaches include the use of metal salts to enhance bioflocculation of organic matter (Zhang *et al.*, 2008) and the use of chemically enhanced backwashing using ozone to reduce fouling (Kim *et al.*, 2007). Biological fouling control, on the other hand, mainly focuses on selecting appropriate operational biological parameters, such as SRT, and consequently MLSS concentration and viscosity. Trussell *et al.* (2006) reported that fouling resistance decreases as the F/M ratio decreases, indicating that it may be beneficial to work at higher MLSS concentrations. However, this has to be balanced against the increased viscosity, which counteracts the scouring effects of membrane aeration, leading to cake layer build up and increased fouling tendency (Chang and Kim, 2005). Sludge bulking caused by filamentous bacteria also has to be avoided, as this increases the fouling tendency due to the release of bound EPS (Meng *et al.*, 2006).

Notwithstanding the above developments, many of the above mentioned chemical fouling control methods are not yet commonly implemented at full scale. Fouling mitigation in most installed MBR plants is achieved through a combination of control of hydrodynamic conditions, and chemical cleaning of the membrane to maintain its permeability. In crossflow MBRs, the crossflow velocity induces shear at the membrane surface, which limits the build up of a cake layer on the membrane surface. In the more widespread submerged MBRs, shear at the membrane surface is provided by membrane aeration, which also keeps the solids in suspension. As membrane aeration not only controls fouling but also accounts for a large share of MBR energy consumption, and so operating costs, it is unsurprising that many publications have concerned the impact of aeration on MBR operation.

Table 2-1: Relationship between various fouling factors and membrane fouling. (Reprinted from Meng *et al.*, 2009)

Sludge Condition	Effect on membrane fouling	Ref.
MLSS	- MLSS↑ → normalized permeability↓	(Trussell *et al*. 2007)
	- MLSS↑ → fouling potential↑	(Psoch & Schiewer, 2006a)
	- MLSS↑ → Cake resistance↑, specific cake resistance↓	(Chang *et al*. 2005)
Viscosity	- Viscosity↑ → membrane permeability↓	(Li *et al*. 2007)
	- MLSS/Viscosity↑ → membrane permeability↓	(Trussell *et al*. 2007)
	- Viscosity↑ →membrane resistance↑	(Chae *et al*. 2006)
F/M	- F/M↑ → fouling rates↑	(Trussell *et al*. 2006)
	- MLSS (2-3 g/L): F/M↑ → irremovable fouling↑ MLSS (8-12 g/L): F/M↑ →removable fouling↑	(Watanabe *et al*. 2006)
	- F/M↑ → Protein in foulants↑	(Kimura *et al*. 2005)
EPS	- polysaccharide↑ → fouling rate↑	(Drews *et al*. 2006)
	- bound EPS influences on specific cake resistance	(Cho *et al*. 2005b)
	- polysaccharide↑ → fouling rate↑	(Lesjean *et al*. 2005)
	- bound EPS↑ → membrane resistance↑	(Chae *et al*. 2006)
	- The tightly bound EPS is important to fouling	(Ramesh *et al*. 2007)
SMP	- SMP is more important than MLSS	(Zhang *et al*. 2006)
	- colloidal TOC relates with permeate flux	(Fan *et al*. 2006)
	- filtration resistance is determined by SMP	(Jeong *et al*. 2007)
	- SMP is probably responsible for fouling	(Spérandio *et al*. 2005)
	- polysaccharide is a possible indicator of fouling	(Le-Clech *et al*. 2005)
	- SMP↓→fouling index↓	(Jang *et al*. 2006)
	- fouling rates correlate with SMP	(Trussell *et al*. 2006)
Filamentous bacteria	- filamentous bacteria↑ → sludge viscosity↑	(Meng *et al*. 2007)
	- bulking sludge could cause a severe fouling	(Sun *et al*. 2007)
	- filamentous bacteria↓ → cake resistance↓	(Kim *et al*. 2006)
SRT	- SRT decrease from 100 to 20 d → TMP↑	(Ahmed *et al*. 2007)
	- SRT decrease from 30 to 10 d → fouling↑	(Zhang *et al*. 2006)
	- SRTs↑ → fouling potentials of SMP↑	(Liang *et al*. 2007)
	- SRT decrease from 5 to 3 d → fouling↑	(Ng *et al*. 2006b)
HRT	- HRT↓ → membrane fouling↑	(Meng *et al*. 2007)
	- HRT↓ → membrane fouling↑	(Chae *et al*. 2006)
	- HRT↓ → membrane fouling↑	(Cho *et al*. 2005a)
Aeration	- aeration intensity ↑ → permeability↑	(Trussell *et al*. 2007)
	- air sparging improves membrane flux	(Psoch & Schiewer, 2006a)
	- larger bubbles for fouling control are preferable	(Phattaranawik *et al*. 2007)
	- air backwashing for fouling control is preferable	(Chae *et al*. 2006)
	- Bubble-induced shear reduces fouling significantly	(Wicaksana *et al*. 2006)
	- Air scouring can prolong membrane operation	(Sofia *et al*. 2004)
Permeate flux	- sub-critical flux mitigates irremovable fouling	(Lebegue *et al*. 2008)
	- sub-critical flux mitigates fouling	(Guo *et al*. 2007)

2.2 MEMBRANE AERATION IN SUBMERGED MEMBRANE BIOREACTORS

2.2.1 Aeration as part of hydrodynamic conditions to control fouling

Table 2-2 reviews trends in aeration research, based on and extended from Meng *et al.* (2009). From early reports (Ueda *et al.*, 1997), several studies have indicated the positive impact of increasing aeration intensity on permeability through the application of shear, with improved control of the cake layer build up (Bouhabila *et al.*, 2001), lateral movement fibre movement (Wicaksana *et al.*, 2006) and increased membrane scouring (Han *et al.*, 2005), while pore blocking is apparently unaffected. The use of air sparging improves the achievable membrane flux (Psoch and Schiewer, 2006a), but no beneficial effect on irremovable fouling has been observed (Delgado *et al.*, 2008). In terms of bubble morphology, studies by Phattaranawik *et al.* (2007) and Prieske *et al.* (2008) have shown that larger bubbles cause more turbulence and are thus more efficient. Recently, computational fluid dynamics have also been applied to achieve a better understanding of shear stresses on the membrane surface (Ratkovich *et al.*, 2009).

One of the most important parameters for long term sustainable MBR operation is the critical flux J_c. This concept was first introduced by Field *et al.* (1995) as the optimum operational flux that results in zero fouling, and is usually determined by a flux stepping method (Le-Clech *et al.*, 2003). However, subsequent research has indicated that the original formulation of critical flux is not valid for MBRs, since even at very low fluxes (below J_c), an increase of the TMP can be observed (Le-Clech *et al.*, 2003). This led to the introduction of the concept of sustainable flux, below which the fouling rate is acceptable for long term plant operation (Le-Clech *et al.*, 2006; Bacchin *et al.*, 2006). A strict definition of sustainable flux is impossible, as an acceptable interval between chemical cleaning steps is usually decided on by the plant operators. However, the sustainable flux is by definition always lower than the critical flux.

Numerous studies suggest critical or sustainable flux to increase roughly linearly (Le Clech *et al.*, 2003; Yu *et al.*, 2003; Xu and Wu, 2008; *Wu et al.*, 2008) with aeration rate up to some threshold value, beyond which little or no further improvement in permeability is observed (Le Clech *et al.*, 2003; Xu and Wu, 2008; Howell *et al.*, 2004). The increased flux has been generally attributed to the associated increase in crossflow velocity of the air-lifted liquid (Ueda *et al.*, 1997; Liu *et al.*, 2003; Xu and Yu, 2008). It is also known that the local flow pattern around an air bubble rising through a channel is very complex (Ghosh and Cui, 1999, Cui *et al.*, 2003), exerting significant transient shear at the membrane surface and increasing the flux attained over that from liquid flow alone. However, most studies have been carried out at laboratory scale; it is widely reported that fouling rates measured at such small scales are inappropriate to describe long-term operation at full-scale (Kraume *et al.*, 2009). This is corroborated by Pollice *et al.* (2005) in a review on sub-critical flux fouling in MBRs, where greater permeability decline rates have been reported for lab scale experiments than for large scale plants. Quantitative information on sub-critical behaviour obtained at the bench scale thus cannot be transferred to full scale plants.

Quantitative data on J_c for pilot and large scale plants are listed in Table 2-3, together with factors impacting on J_c. Based on experiments on a large pilot scale HF MBR, Guglielmi *et al*. (2007b) reported an increase in J_c from 24.9 to 30.1 LMH as membrane aeration increased from 0.3 to 0.5 $Nm^3.m^{-2}.h^{-1}$, above which no further increase in Jc is observed, corroborating the results obtained by earlier authors working at smaller scales (Ueda *et al*., 1997, Bouhabila *et al*., 2001, Le Clech *et al*., 2003, Ndinisa *et al*., 2006, McAdam *et al*., 2010). In a further study on a large FS pilot plant, Guglielmi *et al*. (2008) similarly observed that J_c increased from 22 to 31 LMH with aeration up to a threshold value (0.88 $Nm^3.m^{-2}.h^{-1}$), and that more frequent but milder chemical cleanings (200 $mg.l^{-1}$ NaOCl monthly vs. 2000 $mg.^{l-1}$ every three months) resulted in significantly higher J_c values for FS modules at the same SAD_m. The authors also reported that fouling rates below the critical flux increased exponentially with the flux, confirming results by Brookes *et al.*, 2006, Le Clech *et al.*, 2003, Germain *et al*., 2005 and, most recently, McAdam *et al.*, 2010. Fan *et al.* (2006) attempted to correlate critical flux to several parameters indicative for sludge quality (MLSS, EPS, TTF, DSVI, temperature, colloidal particle content). A correlation could only be found for colloidal TOC: a decrease in colloidal TOC from 50 to 10 $g.m^{-3}$ resulted in an increased J_c from 15 to 45 LMH. Guglielmi *et al.* (2007a) also noticed that an increase in temperature had a positive effect on J_c.

Table 2-2: Impact of aeration on filtration (Adapted and extended from Meng *et al.*, 2009)

Aeration factor	Reference
Aeration intensity↑	
→ Cake layer↓, TMP↓	Ueda *et al.*, 1997, Ndinisa *et al.*, 2006
→ Cake layer↓, no effect on internal fouling	Bouhabila *et al.*, 2001
→ Flux decline↓	Hong *et al.*, 2002, Liu *et al.*, 2003
→ Scouring effect↑, Resistance decline↓	Han *et al.*, 2005
→ Permeability↑	Psoch and Schiewer, 2006a, Trussell *et al.*, 2007
→ Fibre movement↑	Wicaksana *et al.*, 2006
→ Cake removing efficiency↑	Chang and Judd, 2003
→ Cake resistance↓	Psoch and Schiewer, 2006b, Fan and Zhou, 2007
→ Influences type and composition of EPS in flocs, fouling↓	Ji and Zhou, 2006
→ Increases shear in tubular membranes	Ratkovich *et al.*, 2009
→ Critical flux J_c↑	Le-Clech *et al.*, 2003, Howell *et al.*, 2004, Guglielmi *et al.*, 2007b, 2008
→ No positive effect of increasing aeration beyond threshold value	*Inter alia* Ueda *et al.*, 1997, Ndinisa *et al.*, 2006, Hong *et al.*, 2002, Liu *et al.*, 2003, Guglielmi *et al.*, 2007b, 2008
Excessive aeration	
→ Severe floc breakage → Pore blocking↑	Fan and Zhou, 2007, Meng *et al.*, 2008
Air-sparging	
→ Improves membrane flux	Psoch and Schiewer, 2006a
→ Cake build up↓	Delgado *et al.*, 2008
→ Irremovable fouling unaffected	Delgado *et al.*, 2008
Larger bubbles	
→ Preferable for fouling control	Phattaranawik *et al.*, 2007
→ Higher crossflow velocity, more efficient scour	Prieske *et al.*, 2008
Air backwashing	
→ Preferable for fouling control	Chae *et al.*, 2006
Bubble-induced shear	
→ Reduces fouling significantly	Wicaksana *et al.*, 2006
Air scouring	
→ Prolongs membrane operation	Sofia *et al.*, 2004

Notwithstanding reported correlations with sludge/water quality determinants, reports have generally indicated that membrane aeration energy demand can be reduced through adjusting membrane aeration proportional to flux for both FS and HF modules. Moreover, based on lab scale experiments, Meng *et al.* (2008) report that both too small and too large aeration intensities have a negative impact on permeability; too low an aeration intensity is insufficient to effectively remove the membrane foulants, leading to build up of a cake layer, while excessive aeration intensities break up the sludge flocs and increase pore blocking. This suggests a possible decrease in sustainable flux above a threshold aeration intensity. Analogous phenomena have been reported for sidestream MBRs, where fouling is reduced by increasing crossflow velocities and excessive crossflow velocities have been found to cause floc breakage and lead to rapid permeability declines (Park *et al.*, 2005). The findings from the above studies suggest an optimal aeration rate in function of operating flux, forming the premise for the aeration energy model developed in Chapter 4.

Table 2-3: Overview of reported critical flux values for pilot and large scale plants and impacting factors

Scale	Membrane	MLSS $g.m^{-3}$	J_c LMH	Aeration $Nm^3.m^{-2}.h^{-1}$	Remarks	Reference
Aerobic; 240 $m^3.d^{-1}$	HF	10-21	15-45	Not specified	Colloidal TOC↓→Jc↑	Fan *et al.* (2006)
Aerobic; 40 $m^3.d^{-1}$	HF	7.7	28-31	0.35	T↑→Jc↑	Guglielmi *et al.* (2007a)
Anoxic 2.8m^3 + aerobic 5.1m^3	HF; 69m^2	10±0.5	24.9–30.1	0.3 – 1	Aeration intensity↑→Jc↑; up to threshold→no further improvement	Guglielmi *et al.* (2007b)
Anoxic 2.3m^3 + Aerobic 4.4m^3	FS; 40m^2	20±1	22-31	0.5 – 1	Aeration intensity↑→Jc↑; up to threshold→no further improvement; Chemical cleaning frequency↑ + chemical concentration↓→Jc↑	Guglielmi *et al.* (2008)

2.2.2 Impact of aeration on energy consumption

It is widely acknowledged that MBRs command a higher energy consumption compared to CAS, due to the inherent need for additional membrane aeration. For full scale plants, coarse bubble membrane aeration typically consumes 35 – 50% of the total plant energy demand, which generally ranges from 0.6 to 2.0 $kWh.m^{-3}$ depending on plant utilisation (Judd and Judd, 2010, Fenu *et al.*, 2010a; Brepols *et al.*, 2010). Membrane aeration is thus the primary target for reducing energy demand. Consequently, over the past 15 years membrane aeration demand per unit of permeate produced has decreased dramatically, from 0.70 $kWh.m^{-3}$ to less than 0.10 $kWh.m^{-3}$ for HF modules, due to advances in module design and aeration intermittency (Figure 2-1). The introduction of intermittent aeration, specifically by limiting aeration for 10s every 20s ("10:10" aeration) or every 40s ("10:30" aeration), has been one of the main improvements in energy consumption for HF membranes (Chapter 5). Besides the energy benefit, Van Kaam *et al.* (2008) also shown intermittent aeration to prevent damage to the mixed liquor flocs, decreasing fouling potential.

Fenu *et al.* (2010a) analysed the energy profile of a MBR with an average flow of 230 $m^3.h^{-1}$. Total energy consumption was 0.64 $kWh.m^{-3}$, or roughly twice the value for reference CAS installations in Flanders, Belgium (0.30 $kWh.m^{-3}$). The authors also conclude that an MBR is still more energy intensive than CAS with tertiary UF and UV, a treatment train that can deliver comparable effluent quality. The main contributors to the excessively high energy consumption of the MBR are the filtration process, which accounted for 56% of total energy consumption, and the elevated mixed liquor concentration resulting in higher mixing energies. Coarse bubble aeration used three times more energy than the fine bubble aeration, while contributing less than one third to the total biological oxygen demand. The authors stressed the importance designing the MBR plant so as to enhance oxygen usage supplied by the coarse bubble aeration. In this regard, the IWA ASM models under dynamic conditions (Section 2.3) can be used to quantify and account for the impact of coarse bubble aeration on biological oxygen demand.

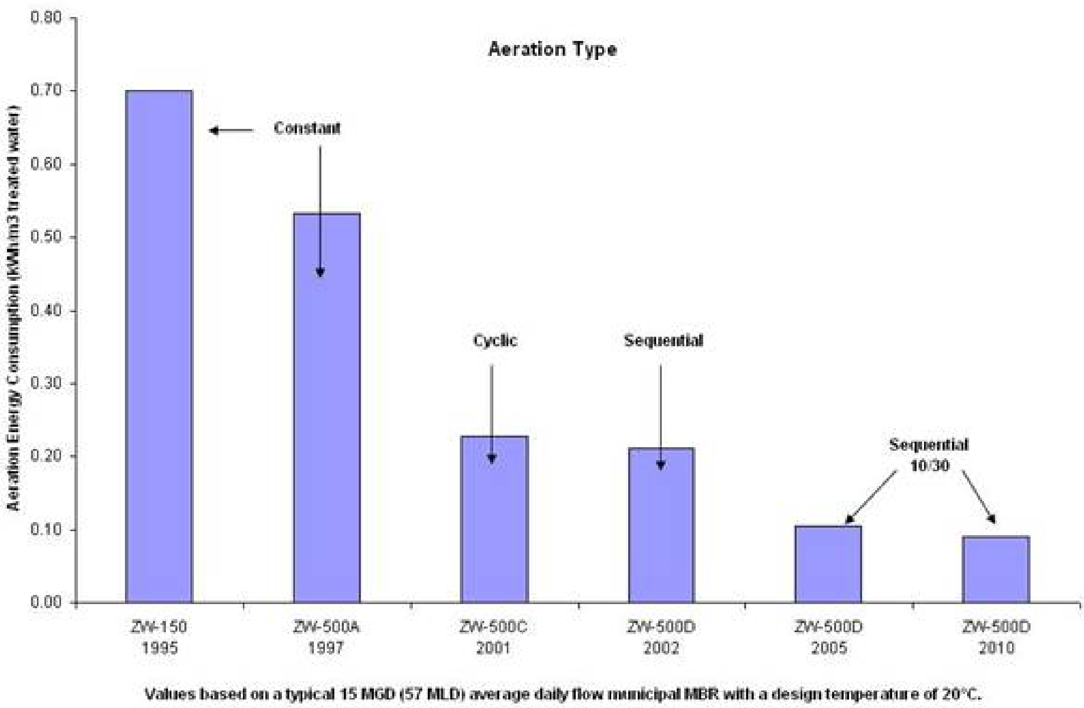

Figure 2-1: Impact of module design and aeration intermittency on membrane aeration energy demand for HF modules, kindly provided by GE Zenon

2.3 APPLICATION OF ASM-BASED MODELLING TO MBR

The activated sludge models (ASMs) of Henze *et al.* (2000) were created with the purpose of describing the biological dynamics of conventional activated sludge (CAS) systems. Applying these ASM to MBRs demands that the differences between MBR and CAS systems be recognised (Ng and Kim, 2007), viz.: a) microbiological composition, leading to different stoichiometric and kinetic parameters (Wen *et al*., 1999; Jiang *et al*., 2005; Lobos *et al*., 2005), b) biomass concentration, leading to changes in oxygen transfer and uptake (Krampe and Krauth, 2003; Germain *et al*, 2007), and c) accumulation of soluble microbial products (SMP) rejected by the membrane step (Drews *et al.*, 2007), and d) requirement for additional aeration for membrane scouring (Judd, 2006).

Fenu *et al.* (2010b) have recently carried out an extensive review into ASM based modelling with special regard to the above mentioned MBR specificities. A clear division was made between modified and unmodified ASM models for MBR. The unmodified models are the plain ASM models, as described in Henze *et al.* (2000), and only parameter estimations have been performed for application to an MBR. They can be used when the modelling purpose is similar to the original ASM modelling goals - process design, effluent characterisation, prediction of sludge production and oxygen demand - making them immediately relevant to practitioners. Modified ASM models, on the other hand, extend the ASM models in terms of biokinetic process models and/or SMP/EPS models. These models are mainly used in academia to improve process understanding, e.g. for prediction of membrane fouling (*inter alia* Lu *et al.*, 2001; Di Bella *et al.*, 2008). A discussion of the modified ASM models is outside the scope of this thesis, and the reader is referred to Fenu *et al.* (2010b) for more information. Table 2-4 gives a qualitative overview of the impact of the above mentioned MBR specificities on influent fractionation and process kinetics for the unmodified ASM models, based on the review by Fenu *et al.* (2010b). Influent fractionation is necessary in modelling exercises to obtain a detailed characterisation in terms of biodegradability and physical state of the influent COD, and warrants further attention in applying ASM modelling to MBRs. The high SRT typically encountered in MBRs can result in biodegradation of the fraction (X_I) of the influent COD that is normally considered inert, leading to an overestimation of the sludge production (Spérandio and Espinosa, 2008; Rosenberger *et al.*, 2006). Jiang *et al.* (2005) compared two methods for influent fractionation, and concluded that the use of a physical-chemical method (Roeleveld and van Loosdrecht, 2002) compared to a chemical-biological method based on respirometry (Vanrolleghem *et al.*, 1999) resulted in an overestimation of X_I and consequently sludge production. However, Fenu *et al.* (2010b) also report that a trial-and-error approach is often used for modelling applications on the full scale, tuning the influent COD fractions to fit predicted sludge concentrations to observed MLSS values in the plant (Erftverband, 2001).

Table 2-4: influence on MBR process specificities on influent fractionation and process kinetics (after Fenu *et al.*, 2010b)

Influent fractionation	Reference
→ Overestimation of X_I possible with physical-chemical method compared to chemical-biological → Overestimation of sludge production	Jiang *et al.*, 2005
→ High SRT → Biodegradation of X_I → Overestimation of sludge production	Sperandio and Espinosa, 2008; Rosenberger *et al.*, 2006
Process kinetics	
Nitrification kinetics	
→ More complete nitrification in MBR	Munz *et al.*, 2008; Parco *et al.*, 2006;
→ Hydrodynamic conditions and SRT → Influence floc size	Manser *et al.*, 2005; Sarioglu *et al.*, 2008, 2009
→ Halfsaturation coefficient K_{OA} and K_{OH} lower than in CAS	Parco *et al.*, 2006; Manser *et al.*, 2005
Denitrification kinetics	
→ Denitrification rate similar to CAS	Parco *et al.*, 2007
→ O_2 recirc from membrane to anoxic zone → denitrification potential↓	Sarioglu *et al.*, 2008; Daigger *et al.*, 2010
Phosphorous removal kinetics	
→ Kinetic parameters comparable in CAS and MBR	Parco *et al.*, 2007
→ Default ASM2d parameters underestimate PO4-P	Jiang *et al.*, 2008
→ Full scale plants: biological PO-P removal better than expected	Silva *et al.*, 2009; Fenu *et al.*, 2010
Oxygen transfer rate (α-factor)	
→ MLSS↑; viscosity↑ → α-factor↓	Krause *et al.*, 2003; Krampe and Krauth, 2003; Germain *et al.*, 2007
→ Bound carbohydrates↑ → α-factor↑	Germain *et al.*, 2007
→ soluble COD↑→ α-factor↓	Germain *et al.*, 2007
Sludge production	
→ Biodegradation of X_I at high SRT → Overestimation of MLSS	Sperandio and Espinosa, 2008; Rosenberger *et al.*, 2006
→ Very sensitive to Y_H, b_H and f_p	Jiang *et al.*, 2005; Sperandio and Espinosa, 2008

In terms of process kinetics, hydrodynamic and biological conditions typically encountered in MBR impact most on nitrification performance (Fenu *et al.*, 2010b). Nitrification is more stable and complete in MBRs (Munz *et al.*, 2008; Parco *et al.*, 2006; Manser *et al.*, 2005), which is often attributed to a decrease in the half-saturation coefficients K_{NH} and K_{OA} which impact on the ammonia concentration. Manser *et al.* (2005) attribute this to the smaller floc size in MBRs due to increased shear. However, recent investigations by Sarioglu *et al.* (2008, 2009) on MBRs with limited shear forces have indicated larger floc sizes than those typical of a conventional activated sludge plant, with values for K_{NH} higher than the default ASM1 values reported. Thus, there is as yet no agreement on general kinetic parameters for nitrification as such values depend on SRT, MLSS, viscosity, dissolved oxygen and floc size (Fenu *et al.*, 2010b), and are therefore case-specific. Denitrification kinetics are less affected by the MBR specificities, and Parco *et al.* (2007) report denitrification rates similar to CAS. However, denitrification may be negatively affected by recirculation of oxygen-rich mixed liquor from the membrane zone to the anoxic zone (Sarioglu *et al.*, 2008; Daigger *et al.*, 2010). With regards to phosphorous removal, conflicting results are reported: Parco *et al.* (2007) mention values for the kinetic parameters similar to CAS for ASM1 and ASM3, while Jiang *et al.* (2008) state that default ASM2d parameters underestimate the PO_4-P concentrations in the effluent. However, recent full-scale investigations report better biological phosphorous removal than expected even in absence of anaerobic zones (Silva *et al.*, 2009; Fenu *et al.*, 2010a; Section 6.3.2.2). A possible explanation could be the occurrence of localised anaerobic zones due to insufficient mixing, but this needs further study. One of the most important consequences of the elevated mixed liquor concentrations in MBR is the severe impact on the oxygen transfer rate, quantified by the α-factor: it is widely reported that increasing MLSS concentrations and increasing viscosity lead to a decreasing α-factor (Krause *et al.*, 2003; Krampe and Krauth, 2003; Germain *et al.*, 2007). The latter authors also mention that an increase in bound carbohydrates and soluble COD respectively have a positive and negative effect on the α-factor.

The above shows there to be no consensus on values for biokinetic parameters generally applicable to MBRs. However, literature on the application of the ASMs to full scale MBRs is scarce or not readily accessible (Erftverband, 2001; Erftverband, 2004). Research focuses mainly on sludge production through application of ASM1 and ASM3 to lab and bench-scale MBRs (Spérandio and Espinosa, 2008; Lubello *et al.*, 2009), often using synthetic feed wastewater, making it impossible to extrapolate the results for full scale applications. The requirement for full scale validation of the ASMs for MBR applications is thus self-evident, and recently identified as an urgent research need (Fenu *et al.*, 2010b). Chapter 6 provides a case study of the calibration and application of ASM2d to a community-scale MBR, incorporating an aeration model accounting for oxygen mass transfer at the operational biomass concentration and differentiates between coarse and fine bubble aeration.

2.4 MBR COSTING

MBR capex and opex have decreased considerably over the past two decades, as evidenced by Figure 2-2, which charts the trend in MBR process costs vs. time for flat sheet modules, and Figure 2-3, showing the evolution in operating and maintenance costs vs. time (Kennedy and Churchouse, 2005). While membrane replacement costs were the main contributor to O&M costs for early MBRs, energy and sludge disposal have now taken over as the main costs (Brepols *et al.*, 2010) due to decreasing membrane costs and longer than expected membrane life (Judd, 2006). Reduction of energy costs is thus paramount and forms an important research topic (Section 2.2.2). However, quantitative cost information in literature is scarce. Available data used for the cost analysis of a large-scale MBR are discussed in section 7.1, whilst the determining factors for costing of small scale MBRs (< 500 $m^3.d^{-1}$), which differ significantly from large scale applications, are discussed in Chapter 3.

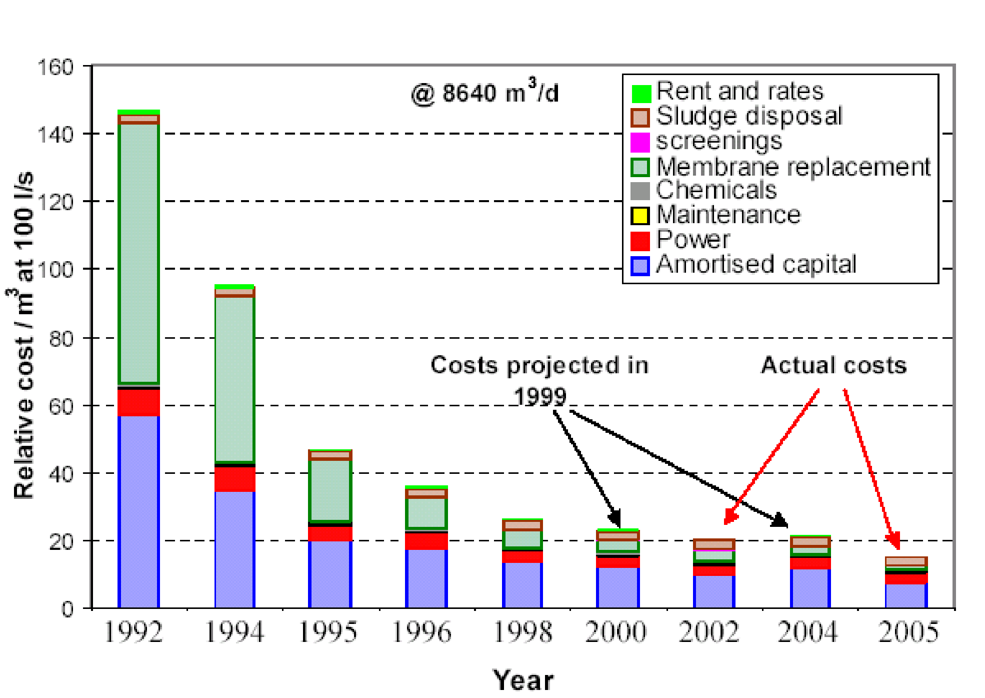

Figure 2-2: MBR process costs (Kubota) vs. time (Kennedy and Churchouse, 2005)

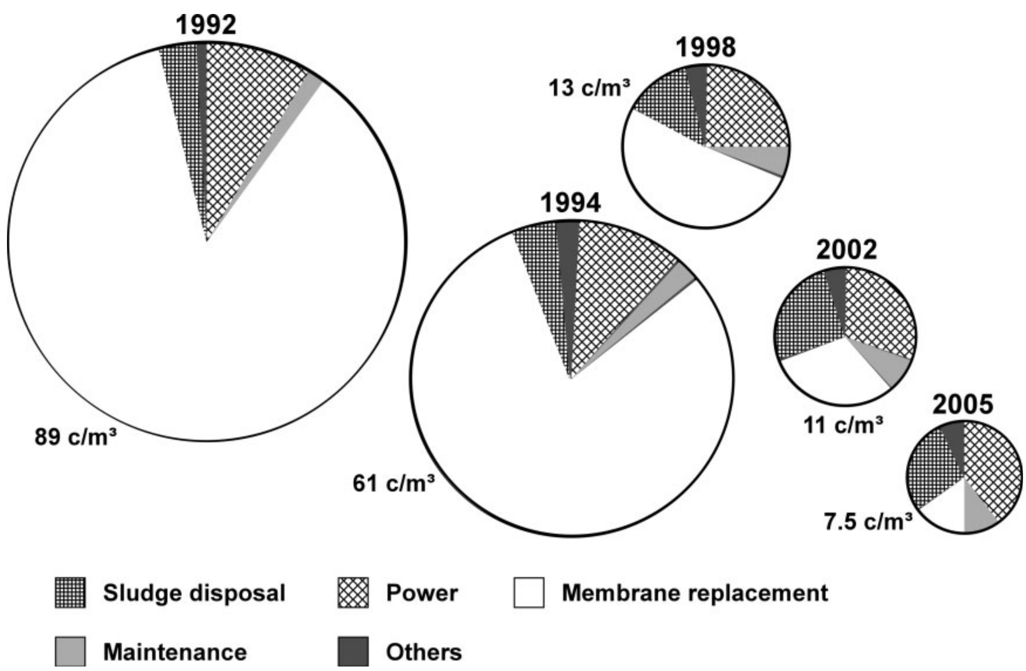

Figure 2-3: Development of operating and maintenance costs (Kennedy and Churchouse, 2005)

2.5 CONCLUSIONS

A literature review on the impact of membrane aeration on hydraulic control of fouling, energy consumption and costing of MBRs has revealed:

- Membrane aeration is commonly used for fouling mitigation in full scale MBRs. An optimum membrane aeration rate exists as a function of flux, making it possible to devise control strategies to limit membrane aeration during low flow periods.
- Membrane aeration has a significant impact on O&M costs, as it typically accounts for 35-50% of the total plant energy demand. Improvements in module design and the advent of intermittent aeration have resulted in a significant decrease in energy consumption for membrane aeration.
- Unmodified activated sludge models (ASMs) can be applied to MBRs to optimise process design, and for prediction of effluent quality, sludge production and oxygen demand. However, the hydrodynamic and biological conditions typically encountered in MBRs, such as high SRT and MLSS concentrations, combined with vigorous aeration, have to be taken into account since they impact on biological performance. There is no consensus yet on biokinetic parameters generally applicable to MBRs, and literature suggests they may be case-specific. However, full scale model validation is scarce.
- MBR capex and opex have significantly decreased over the last two decades. However, quantitative costing information in literature is scarce.

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CHAPTER 3

ECONOMICAL EVALUATION AND OPERATING EXPERIENCES OF A SMALL SCALE MBR FOR NON-POTABLE REUSE

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3 ECONOMICAL EVALUATION AND OPERATING EXPERIENCES OF A SMALL SCALE MBR FOR NON-POTABLE REUSE

B. Verrecht[1], C. James[2], E. Germain[2], R. Birks[2], A. Barugh[2], P. Pearce[2] and S. Judd[1]

[1]Centre for Water Science, Cranfield University, Cranfield, Bedfordshire MK43 0AL, UK

[2]Thames Water R&D, Innovation Centre, Island Road, Reading, Berkshire RG2 0RP, UK

ABSTRACT

Due to their consistently high effluent quality, small footprint and robustness to variations in influent quality, membrane bioreactors (MBRs) have become the technology of choice for small-scale reuse applications such as in office buildings, hotels and on cruise ships. The emergence of these systems arises from a number of drivers: lack of sewerage infrastructure, requirement for planning permission, subsidies, new guidelines for green buildings, and the public profile of recycling generally. This paper details the design and operation of a small scale MBR providing 25 $m^3.d^{-1}$ of reclaimed water for toilet flushing and irrigation. Operational experience and outcomes from a two year evaluation period are included. An economic analysis of operational costs (opex) is also presented, revealing that for a plant of this scale staffing costs account for the largest component (53%) of the opex followed by energy consumption (28%). The optimum design of these systems should therefore be focused on reducing operational complexity to minimise manual intervention.

3.1 INTRODUCTION

The increasing global population places an ever-growing pressure on water resources. In water scarce regions and/or densely populated urban areas, water reuse is becoming increasingly attractive. Due to its small footprint, superior and consistent effluent quality, and robustness to changes in influent wastewater strength (Winward *et al.*, 2008; Judd and Judd, 2010), membrane bioreactor (MBR) technology is gaining momentum for urban non-potable reuse purposes, as evidenced by recent implemented schemes over a whole spectrum of plant sizes.

Whilst high-profile large water reuse MBR installations exist (Ernst *et al.*, 2007), a growing number of small scale MBR (<200$m^3.d^{-1}$) are being employed for niche reuse applications, ranging from single household (Abegglen *et al.*, 2006, 2008), to holiday resort buildings/hotels (Boehler *et al.*, 2007; Meuler *et al.*, 2007; Paris *et al.*, 2008), apartment/office blocks (Clerico, 2007), and cruise ships (MER, 2006). MBRs have been shown to produce reliably high-quality effluent under conditions of highly variable loads, both seasonally (especially for tourist resorts) and diurnally, associated with these small plants. The product water is often used for toilet flushing (Boehler *et al.*, 2007; Clerico, 2007; Meuler *et al.*, 2007; Merz *et al.*, 2007) as well as for irrigation and cooling tower make-up water (Clerico, 2007; Ogoshi *et al.*, 2001).

Incentives and drivers for the emergence of these reuse schemes differ according to application and region. In areas with water scarcity the main driver is water conservation through reuse for purposes such as golf course irrigation (Meuler *et al.*, 2007). Outside of cities, the main driver for installating water reuse technology is often the lack of sewerage, such that no planning permission is given without an installed reuse system (Clerico, 2007). In areas where water scarcity is less critical, e.g. in metropolitan cities, such as New York City (NYC), the main driver is the "green" agenda linked with the reuse of blackwater. Several green-building schemes have emerged over the past decade, such as LEED (Leadership in Energy and Environmental Design; USGBC, 2010) and the Code for Sustainable Homes (CSH, 2010) in the UK, demanding decreased in-building water consumption to achieve improved environmental credentials. To obtain planning permission in Battery Park City, an area in New York City, developers must comply with the LEED standards, such that installation of a water reclamation system is imperative. Buildings with a high LEED rating can command higher rents, and an additional financial incentive was introduced in NYC in 2004 whereby water and wastewater charges are reduced by 25% for buildings which can reduce their water consumption commensurately (Clerico, 2007). Similar drivers have been reported for Japan: the Tokyo Metropolitan Government requires large new buildings to adopt water saving measures, including rainwater harvesting and in-building greywater treatment for reuse for toilet flushing (Asano, 2007). As early as 1997, 1475 on-site individual building and block-wide water reclamation and reuse systems existed in Japan (Ogoshi *et al.*, 2001), and 20 years of experience in Fukuoka city have proven water reuse for toilet flushing to be economically justifiable in many water-scarce urban areas.

Despite these drivers and the numerous reference applications, decentralised reuse systems are still subject to major drawbacks. In areas with very high land values it can be more profitable to use the space required for a reuse system in the basement for other purposes, notwithstanding the relatively small footprint of a MBR system. Furthermore, required effluent quality for reuse purposes is significantly higher than that demanded from conventional treatment plants discharging into the environment. Post-treatment is thus required to provide residual disinfection and remove odours and colour (e.g. GAC, ozone, UV, chlorination or a combination), leading to increased costs and footprint (Clerico, 2007, Abegglen *et al.*, 2009). However, available quantitative literature data in this area, as pertaining to small scale systems costs, is scarce.

This paper presents a case study detailing the design and operation of a small scale MBR for decentralised reuse in the UK. Plant performance and operational experience from 2 years of operation are presented. An economic analysis of operational costs was performed and the main factors influencing operational expenditure (opex) identified; suggestions for improving system robustness and to limit operational complexity of small scale plants are provided.

3.2 MATERIALS AND METHODS

3.2.1 Plant description

The wastewater reclamation plant (Figure 3-1) is installed at a sustainable development in south London (UK), consisting of over 100 properties split into 8 housing blocks and a community centre. Besides residential properties, the site also houses several offices, a nursery, an exhibition centre and a show home for visitors. The buildings are fitted with water efficient appliances, and the wastewater reclamation plant produces an average reclaimed water flow of 25 $m^3.d^{-1}$ for toilet flushing and irrigation.

The treatment process comprises the following (Figure 3-1, Table 3-1):

Pre-treatment: Domestic wastewater from the dwellings is collected via two pumping stations and septic tanks, which provide flow equalisation and primary settling. The tanks have a residence time of up to 6 days; they were in place before the MBR system was installed, and were not designed specifically as pre-treatment for the MBR. Further pre-treatment is provided by 3 mm disposable sac screens (Copasac, Eimco UK) to remove hairs and fibres that could otherwise damage or clog the MBR membranes.

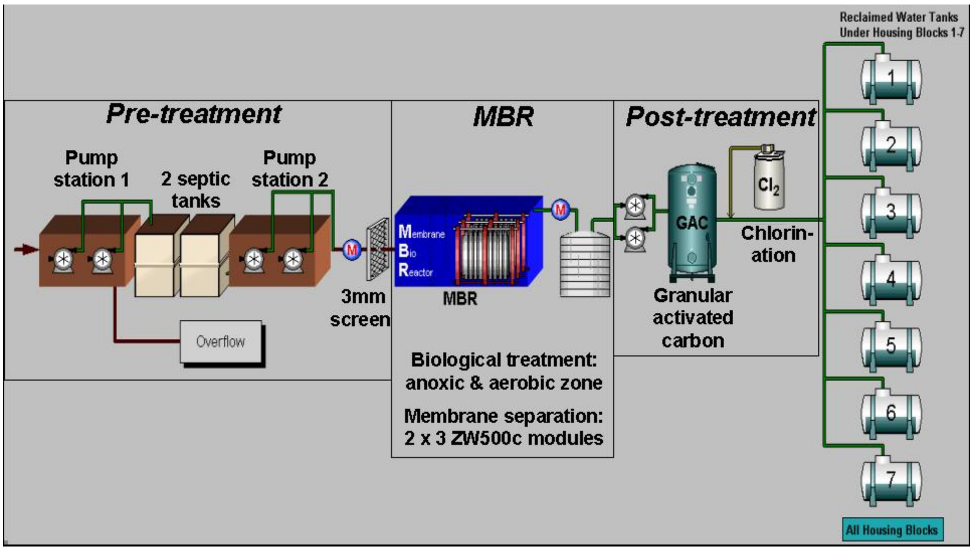

Figure 3-1: Schematic overview of the wastewater recycling plant

Membrane bioreactor: The MBR, a package plant designed by GE Zenon (Canada), contains both an anoxic (10.1 m^3) and aerobic (12.8 m^3) zone for nitrification and denitrification respectively. The anoxic zone is equipped with a submerged agitator (3021, ABS, Germany) to keep the solids in suspension. Inflow of settled and screened sewage into the anoxic zone is controlled by the liquid level in the aerobic zone. The mixed liquor overflows through a weir from the anoxic zone to the aerobic zone, where the dissolved oxygen concentration is maintained at around 2 $mg.l^{-1}$ through on/off control of a blower (GM3S DN 50, Aerzen, Germany) with a maximum capacity of 90 $Nm^3.h^{-1}$. The fine bubble aeration provided for DO control also keeps the contents of the aerobic tank mixed. Biomass is recirculated from the aerobic tank to the anoxic tank by means of a centrifugal pump (Sewabloc F50-250, KSB, Germany) with a maximum capacity of 8 $m^3.h^{-1}$, corresponding to a maximum recirculation ratio of 7.7. The solids retention time (SRT) is controlled by a timer-controlled automatic wastage valve, and sludge is wasted to the local sewer.

The ultrafiltration membrane separation step is provided by 2 x 3 ZW500c (GE Zenon, Canada) hollow fibre membrane modules, made from PVDF with a pore size of 40nm. The membrane cassettes are submerged in the aerobic zone, and provide a total membrane surface area of 139 m^2. A lateral channel blower (Becker, Germany) provides air for membrane scouring to the coarse bubble diffusers incorporated in the module design, at a maximum flow rate of 115 $Nm^3.h^{-1}$, corresponding to a maximum specific aeration demand (SAD_m) of 0.82 $Nm^3.m^{-2}.h^{-1}$. Air cycling between the two cassettes is made possible by intermittent aeration valves, controlled by an adjustable timer. Under normal operation, one cassette is aerated at a time for 10 seconds every 20 seconds.

Table 3-1: MBR characteristics and range of MBR operational parameters over the 2 year evaluation period

Parameter	Unit	Value
Influent flow	$m^3.d^{-1}$	25
Volume anoxic zone	m^3	10.1
Volume anoxic zone	m^3	12.8
Recirculation ratio	-	2.3 – 4.3
Hydraulic retention time	*d*	1
Solids retention time	*d*	35 - 50
MLSS	$g.m^{-3}$	7554 ± 1773
Temperature	*°C*	14 - 27
Filtration parameters		
Membrane surface	m^2	69.6 & 139.2
Instantaneous filtration flux	*LMH*	10.8 - 28.4
Filtration time	*s*	600
Relaxation time	*s*	30
Backwash time	*s*	30
Instantaneous backwash flux	*LMH*	10.8 – 28.4
SAD_m	$Nm^3.m^{-2}.h^{-1}$	0.11 – 1.25
SAD_p	-	4.6 - 110
Aeration intermittency	-	Continuous; intermittent 10 s on – 10 s off; intermittent 10 s on – 30 s off
Permeability	$LMH.bar^{-1}$	~ 100

Post-treatment: To ensure reclaimed water quality, post-treatment consists of filtration through a mixture of granular activated carbon and hydroxyapatite to remove residual colour and odour, followed by chlorination for further disinfection and suppression of bacterial regrowth in the distribution pipework (Karim *et al.*, 2005). The GAC vessel has a bed volume of 200 litres, and is normally run at an empty bed contact time of 10-20 minutes. A dose of 3 $mg.l^{-1}$ NaOCl is required to achieve a 1 $mg.l^{-1}$ chlorine residual after 24h. The housing development has an existing dual pipe network to accommodate both the reclaimed and potable water supply to the houses, and reclaimed water is stored in tanks under each of the seven housing blocks.

The wastewater reclamation plant is automatically controlled by a PLC, and is monitored online with a dedicated SCADA system displaying all relevant flows, levels, temperature, pressure and concentrations. Grab samples are collected twice weekly from the influent, the aerobic zone mixed liquor, the MBR effluent, post GAC and final effluent. Samples were analysed according to the standard methods (APHA, 2005), and influent wastewater characterisation and fractionation has been published elsewhere (Verrecht *et al.*, 2010a).

3.2.2 Calculation of operational costs

A cost sensitivity analysis was carried out, including energy consumption, staff cost for maintenance and plant attendance, chemicals and activated carbon usage, and sludge treatment and disposal. Table 3-2 displays the plant characteristics, derived from an evaluation period of two years, and assumptions used in the calculation of operational costs. Costs for the GAC adsorption media and chemicals were obtained from the suppliers, while costs for sludge treatment and disposal were derived from Ginestet *et al.* (2006) who based their analysis on collection, thickening, digestion and dewatering plus average values among different disposal/reuse routes including hauling. Sludge production P_x was estimated from (Fletcher *et al.*, 2007):

$$P_x = \frac{V \cdot MLSS}{SRT} \quad \textbf{(3.1)}$$

where V is the total biotank volume (m^3). A maintenance clean (cleaning-in-place; CIP) with 500 ppm NaOCl every two weeks was sufficient to maintain permeability at around 100 $LMH.bar^{-1}$.

Table 3-2: Plant characteristics and assumptions for calculation of operational costs (OPEX)

Plant characteristics used in opex calculation					
Parameter	**Unit**	**Value**	**Parameter**	**Unit**	**Value**
Pre-treatment			NaOCl used per CIP	l	2
PS 1 - kW rating	kW	1.6	MLSS	$g.m^{-3}$	8000
PS 1 – Flow	$l.s^{-1}$	4.25	SRT	d	50
PS 2 - kW rating	kW	1.1	**Post-treatment**		
PS 2 – Flow	$l.s^{-1}$	3.2	Energy consumption	kW	1.4
Membrane bioreactor			GAC capacity	BV	6000
Energy consumption	kW	4.03	Chlorine dosing	$mg.l^{-1}$	3
NaOCl CIP frequency	$1.y^{-1}$	26	**Maintenance / plant attendance**		
CIP NaOCl concentration	ppm	500	Weekly staff attendance	$h.wk^{-1}$	8

Cost assumptions for OPEX calculation			
Parameter	**Unit**	**Value**	**Reference**
Electricity	$£.kWh^{-1}$	0.11	UK value, Energy EU, 2010
Labour costs	$£.h^{-1}$	25	-
Granular activated carbon	$£.kg^{-1}$	2.98	Supplier
Sludge management	$£.tnDS^{-1}$	423 ± 252	Ginestet et al., 2006
NaOCl 14%	$£.l^{-1}$	0.3	Supplier

3.3 RESULTS AND DISCUSSION

3.3.1 Effluent quality

Since no guidelines currently exist for unrestricted urban reuse in the UK, the US EPA standards for unrestricted urban reuse (EPA, 2004) were adopted. Table 3-3 shows that the reclaimed water quality produced consistently meets and exceeds these standards, which is in line with the performance of other reuse MBRs (Clerico *et al.*, 2006, Winward *et al.*, 2008). The chlorine residual was higher than that required under US EPA guidelines, since the length of the distribution pipework and the residence time (> 30 days) provided by the product water storage tanks made the ensuring of a chlorine residual challenging. Undetectable levels of coliforms could not be guaranteed at all times in the tanks, despite coliforms being undetectable in the final effluent. This was addressed by shock dosing with sodium hypochlorite. These tanks were in place before installation of the water reclamation plant, are oversized for their purpose, and suffer from contamination from rainwater infiltration. Similar problems with bacterial regrowth in the distribution pipework were reported by Merz *et al.* (2007). The biological performance of the MBR in terms of nutrient removal has been discussed in detail (Verrecht *et al*., 2010a), and was in line with widely reported trends for MBRs, both on the large and small scale (Fan *et al*., 2006; Abegglen *et al.*, 2008; Gnirss *et al.*, 2008a, Judd and Judd, 2010).

Table 3-3: Comparison of reclaimed water quality with the US EPA guidelines for unrestricted urban reuse

		US EPA recommended guidelines for unrestricted urban reuse	**Product water quality**
Parameter	**Unit**	**Value**	**Value**
BOD_5	$g.m^{-3}$	<10	<1*
Suspended solids	$g.m^{-3}$	No suggestion	<2*
Faecal coliforms	$CFU.(100ml)^{-1}$	No detectable	No detectable
PH	-	6-9	7.3 ± 0.2
Turbidity	NTU	≤2	0.14 ± 0.12
Cl_2 residual	$g.m^{-3}$	1 (after 30 min contact time)	1 (after 24 h contact time)

* Below limit of detection

3.3.2 Analysis of operational costs

Figure 3-2 shows a breakdown of the operational costs for the wastewater reclamation plant. The total operational cost is £2.15 per m^{-3} of reclaimed water produced, 16-27 times higher than opex values reported for large scale MBR by Côté *et al.* (2004), who calculated a value of £0.09.m^{-3} for a 38,000 m^{-3} d^{-1} plant and DeCarolis *et al.* (2004), who reported values between £0.08 and £0.13 per m^3 of permeate produced for plant sizes of 37,000 down to 700 m^3 d^{-1} respectively, illustrating the influence of economies of scale on operational costs. Both studies included labour costs, and DeCarolis *et al.* (2004) also included costs for effluent disinfection with chlorine, which accounted for less than 3% of total opex.

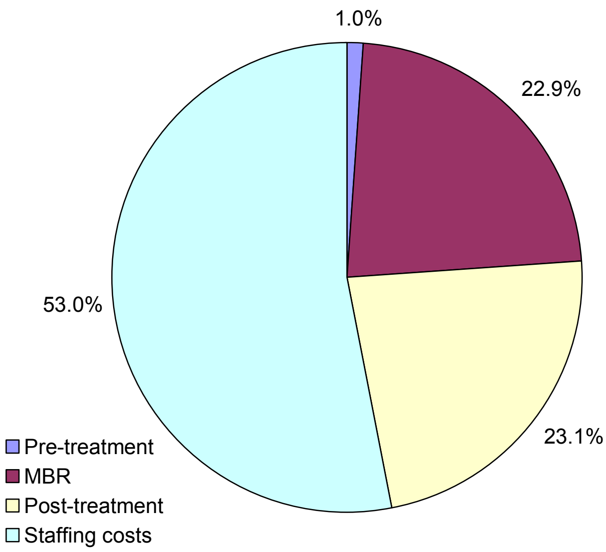

Figure 3-2: Breakdown of operational costs for the wastewater reclamation plant, £2.15.m^{-3} total opex

The contribution of the pre-treatment to total opex is negligible (0.9%), while the MBR and the post-treatment both account for about 23%. However, these contributions are significantly lower than the cost of staff required for maintenance and plant attendance, which accounts for 53% (£1.14.m^{-3}) of total opex, compared to 13-32% of opex for a large scale plant, as reported by DeCarolis *et al.* (2004). However, due to economies of scale, their absolute staffing costs are considerably lower (£0.01-0.04.m^{-3}). Figure 3-2 also shows that the post-treatment train for colour removal, mainly for esthetical reasons, increases total opex by 30%. This is in line with findings by Abegglen *et al.* (2009), who stated that a requirement for colour removal increases opex by 10 to 30%. Thus, in a domestic environment and especially in sustainable developments, where inhabitants may tolerate colour in toilet flushing water, the need for post-treatment of the MBR effluent could be eliminated, leading to substantial savings in capex and opex.

Table 3-4 shows the major contributors to running costs for the MBR (£0.49.m^{-3}) and the post-treatment (£0.50.m^{-3}), excluding staffing costs. Energy consumption makes up 92% of the total operating costs for the MBR. Research at this plant has thus focused on reducing energy demand through intermittent membrane aeration, which has shown that sustainable operation can be achieved when running at a SAD_p of 9.2 under 10:10 aeration (Verrecht *et al.*, 2010c). Further, a modelling approach was followed to identify better operational parameters, resulting in a reduction of the MBR energy consumption from 4.03 to 3.11 kWh.m^{-3}, without compromising biological performance (Verrecht *et* *al.*, 2010a). This reduces the running costs (excluding staffing costs) of the MBR by 20% but has only a minor impact on the operational costs of the entire plant (-4.4%). Table 3-4 also shows that the replacement cost of the granular activated carbon accounts for 69% of the total cost for post-treatment.

Table 3-4: Break-up of operational costs for the MBR and post-treatment

		MBR	Post-treatment
Parameter	**Unit**	**Value**	**Value**
Energy	£.m^{-3}	0.430	0.149
Chemicals	£.m^{-3}	0.002	0.006
GAC	£.m^{-3}	-	0.348
Sludge treatment	£.m^{-3}	0.062	-
Total	**£.m^{-3}**	**0.494**	**0.503**

A simple cost sensitivity analysis (Table 3-5) shows that halving the plant attendance (to 4h per week) reduces opex by 27%. Assuming that the small scale MBR could operate at an energy consumption of 1 kWh.m^{-3}, as typically reported for large scale plants (Brepols *et al.*, 2010), opex would decrease by 15%. Conversely, this value would increase by 30% for an energy consumption of 10 kWh.m^{-3}, corresponding to the high end of values reported for small scale MBR plants which range from 3 kWh to 11.5 kWh.m^{-3} (Boehler *et al.*, 2007, Gnirss *et al.*, 2008b, Verrecht *et al.*, 2010a). An increase in sludge treatment and disposal cost of 60% (423 ± 252, as reported by Ginestet *et al.*, 2006) would increase total plant opex by only 2%. The influence of economies of scale is illustrated though varying the plant capacity: if plant capacity was 4 times higher (100 m^3.d^{-1}), opex per m^3 of reclaimed water produced would decrease by 40%, mainly due to the fact that staffing costs are static with respect to plant capacity up to a certain threshold. This demonstrates the importance of minimising required attendance for small plants. Under the assumptions made, energy consumption would overtake staffing costs as the largest contributor to opex at a plant size of 51.2 m^3.d^{-1} and a specific energy demand for the MBR of 4 kWh.m^{-3}.

Table 3-5: Cost sensitivity analysis (Base operational cost: £2.15.m^{-3})

Parameter	Opex £.m-3	Difference vs. base scenario %
Maintenance - 4 h.wk^{-1} (-50%)	1.58	-27%
Energy consumption MBR		
1 kWh.m^{-3} (~ conservative value large MBR)	1.83	-15%
10 kWh.m^{-3}	2.79	+30%
Sludge treatment and disposal cost - £675.tnDS^{-1}	2.19	+2%
Plant capacity – 100 m^3.d^{-1}	1.26	-41%

The above analysis can be contrasted against large scale MBRs, where energy consumption is the largest contributor to operational costs (Brepols *et al.*, 2010, Verrecht *et al.*, 2010b) and has formed the focus of recent research and development (Garcès *et al.*, 2007, Verrecht *et al.*, 2008, 2010a). For small scale MBRs, however, it is imperative that the plant design is robust and operational complexity avoided so as to minimise manual intervention. From 2 years of operational experiences on the wastewater reclamation plant, several design choices and operational parameters were identified that have a major impact on the amount of plant attendance required:

- *Built-in contingency:* Since small scale plants inherently have to cope with large daily influent variations (Gnirss *et al.*, 2008a, Abegglen *et al.*, 2008), they are generally designed to handle the maximum instantaneous influent flow. Consequently, they are overdesigned compared to their average influent flow, resulting in higher capex as larger plants have to be installed, and larger opex due to inefficiencies and plant underutilisation. Installation of a buffer tank can address some of these concerns. However, a large amount of built-in contingency can also be beneficial to ensure smooth operation: e.g. excess membrane area ensures that the plant can operate at low fluxes, reducing membrane fouling and the need for labour intensive recovery chemical cleaning. High hydraulic and solids retention times, respectively 24 hours and 50 days in this case study, also lead to stable biological performance.
- *Membrane aeration:* Since energy consumption in small scale plants is not the main factor contributing to opex, optimisation of membrane aeration is less important than for large plants. High aeration rates ensure stable membrane performance and reduce maintenance cleaning frequency. It may also be preferable to keep the aeration control to a minimum: valves for intermittent

aeration may reduce energy consumption but present a possible cause of failure.

- *Screens:* Handling of screenings and cleaning screens is one of the most labour intensive tasks on site. However, due to the presence of the excessively large septic tanks, having a hydraulic retention time of about 6 days, most fibres and rags that could potentially block the screens are retained and the 3 mm copasac screens are redundant. Installation of a large septic tank could therefore present a good option for small scale plants, thus eliminating the need for additional screening. However, this potential reduction in opex is countered by the increased capex incurred by septic tank construction.
- *Influent pumps:* Blockage of the influent pumps with rags, fibres and sanitary towels is a regularly occurring problem. It is thus imperative that the influent pumps are easily accessible for cleaning purposes. Installation of oversized influent pumps, possibly with mascerator capacity, may help in reducing the number of blockages and so eliminate a source of frequent plant outages.
- *Remote monitoring:* Attendance/staffing costs can also be reduced by installation of remote monitoring and control, which can also benefit effluent quality and biological performance (Abegglen *et al.*, 2008).

These operational issues show that a trade-off generally exists between capex and opex for small scale plants, as previously discussed for <50 people equivalent package plant MBRs (Fletcher *et al.*, 2007). It is generally the case that capital-intensive plants provide low operational costs because they include design elements that increase efficiency and reduce the need for maintenance and plant attendance.

Based on a model-based approach on the economic feasibility of on-site greywater reuse, Friedler *et al.* (2006) concluded that MBR-based systems were economically unrealistic, only becoming feasible when the building (or cluster of buildings) contained more than 160 apartments, if no subsidies were provided for installation of such systems. This is confirmed on the example of the Solaire, a green building in New York City, where a LCC study by Arpke *et al.* (2006) shows that a decentralised water reuse system is more expensive over a 25-year period, despite an incentive plan that includes a 25% rate reduction for such systems. However, an LCA indicated that the decentralised water reuse system has a lower environmental impact than the conventional centralised approach.

3.4 CONCLUSIONS

A small scale wastewater reclamation plant providing 25 $m^3.d^{-1}$ reclaimed water for toilet flushing and irrigation has been evaluated over a 2 year period, and an economic analysis performed to assess the main factors influencing operational costs. This has revealed:

- Operational costs are 16-27 times higher than those reported for large scale MBRs without post-treatment, due to operational inefficiencies inherent in small scale plants and the disproportionate amount of staff time required.
- Staffing costs incurred by plant attendance and maintenance are the largest contribution (53%) to total opex, followed by energy consumption (28%). This is contrary to findings for large scale plants, where energy costs are the dominating contributor to opex. The main focus for design and operation of small scale plant should be on process robustness, limiting operational complexity so as to minimise manual intervention.
- Post-treatment of the MBR effluent, required mainly for aesthetic reasons, adds significantly to opex (30%). If reclaimed water colour is acceptable for toilet flushing in a domestic environment the need for post-treatment can be eliminated, providing considerable capex and opex savings.
- If post-treatment and labour are excluded, opex costs are about five times higher than those reported for large scale plants, commensurate with the higher specific energy consumption of smaller plants.

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CHAPTER 4

AN AERATION ENERGY MODEL FOR AN IMMERSED MEMBRANE BIOREACTOR

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4 AN AERATION ENERGY MODEL FOR AN IMMERSED MEMBRANE BIOREACTOR

B. Verrecht[1], G. Guglielmi[2], J.W. Mulder[3], C. Brepols[4] and S. Judd[1]

[1]Centre for Water Science, Cranfield University, Cranfield, Bedfordshire MK43 0AL, UK

[2]Trento University, Trento, Italy

[3]Water Authority of Hollandse Delta, the Netherlands

[4]Erftverband, Am Erftverband 6, 50126 Bergheim/Erft, Germany

ABSTRACT

A simple model for evaluating energy demand arising from aeration of an MBR is presented based on a combination of empirical data for the membrane aeration and biokinetic modelling for the biological aeration. The model assumes that aeration of the membrane provides a proportion of the dissolved oxygen demanded for the biotreatment. The model also assumes, based on literature information sources, a linear relationship between membrane permeability and membrane aeration up to a threshold value, beyond which permeability is unchanged with membrane aeration. The model was benchmarked against two full-scale plant to obtain the most appropriate and conservative value of the slope of the flux:aeration curve and the blower efficiency. Benchmarking in this way produced a match to within 20% of all key process plant operational parameters.

The model demonstrated that significant reductions in aeration energy could be obtained through operation at lower flux and reducing the membrane aeration requirement accordingly, so-called ‘‘proportional aeration’’ at lower flows. Similarly, increasing oxygen transfer from membrane aeration would also be expected to decrease energy demand. A sensitivity analysis of some of the key parameters revealed that, of the key operating parameters, loading, SOTE and MLSS concentration remain the most critical in determining energy demand. It is suggested that a key parameter representing membrane aeration in MBRs is the mean in-module air upflow velocity U, since this gives a reasonable representation of the shear applied

through membrane aeration. U was found to vary between 0.04 and 0.1 m/s across a number of modern large pilot and full scale plant.

An analysis reveals that significant reductions in energy demand are attained through operating at lower MLSS levels and membrane fluxes. Evidence provided from recent controlled pilot trials implies that halving the flux can reduce the aeration is suggested whereby the number of membrane tanks on line and/or the membrane aeration intensity is adjusted according to the flow, and thus flux, so as to reduce the overall aeration energy demand.

4.1 INTRODUCTION

The advantages offered by membrane bioreactors (MBRs) over conventional treatment are well-known (Judd, 2006, 2008). The technology incurs a small footprint and provides product water of high quality in a single step, and thus has a significant role to play in wastewater treatment, as discharged water quality standards become increasingly stringent, and in wastewater recycling in particular (Jefferson *et al.*, 2001; Qin *et al.*, 2006). However, the technology is also more costly, both in capital (capex) and operating (opex) expenditure, than the activated sludge process (ASP) on which it is based.

Two of the most significant components of MBR opex are membrane replacement and energy demand (Kennedy and Churchouse, 2005; Judd, 2006), both of which are made more onerous by membrane surface fouling. There is insufficient information available about impacts of operation on MBR membrane life, though anecdotal evidence suggests that the most established commercial products are innately very robust (Kennedy and Churchouse, 2005): membrane replacement is normally associated with process failure of some description.

Membrane surface fouling, and the less well investigated phenomenon of membrane channel clogging, are both ameliorated ostensibly through the use of coarse bubble aeration, applied beneath the MBR membrane module. For an immersed MBR approximately 30-40% of the energy demand arises from aeration of the membrane with a further 10-50% - depending on feedwater strength - demanded for biotreatment (Kennedy & Churchouse, 2005; Judd, 2006; Garces *et al.*, 2007; Stone and Livingston, 2008); it is the membrane aeration which is primarily responsible for promoting permeate flux and/or maintaining membrane permeability. Reducing the energy demand in a conventional MBR thus relies on an understanding of the total aeration requirements of the process, and the balance between the aeration demanded by the membrane compared to that of the biology. In this paper, a simple mathematical model is presented for MBR aeration and available data from two full-scale plant used for benchmarking. A sensitivity analysis is presented for key variable parameters to assess their impact on the specific energy demand for aeration (E_A in kWh per m^3 permeate product).

4.2 MEMBRANE AERATION

Aeration imparted to the membrane is denoted SAD_m, the specific aeration demand in normalised $m^3.h^{-1}$ air per unit membrane area. The flux through the membrane is denoted J in $m^3.h^{-1}$ permeate per m^2 membrane area). The ratio of these two quantities yields a unitless parameter SAD_p, the ratio of volume of air applied per unit permeate volume attained:

$$SAD_p = SAD_m/J \qquad \textbf{(4.1)}$$

For a given aerator system at a fixed depth in the tank, SAD_p relates directly to specific energy demand for membrane aeration (E_A, in kWh per m^3 permeate) (Judd, 2006):

$$E_A = kSAD_p \qquad \textbf{(4.2)}$$

where

$$k = \frac{pT\lambda}{1.73x10^5 \varsigma(\lambda - 1)}\left[\left(\frac{1000y + p}{p}\right)^{1-\frac{1}{\lambda}} - 1\right] \qquad \textbf{(4.3)}$$

and	p	=	blower inlet pressure in Pa
	T	=	air temperature in ºK
	ς	=	blower efficiency
	λ	=	aerator constant (~1.4)
	y	=	membrane aerator depth in m

It has generally been observed that the sustainable flux increases – roughly linearly (Le Clech *et al*, 2003; Yu *et al.*, 2003; Xu and Wu, 2008; Wu *et al.*, 2008) - with aeration rate up to some threshold value, beyond which little or no further improvement in permeability is observed (Ueda *et al.*, 1997; Le Clech *et al.*, 2003; Meng *et al.*, 2008; Xu and Wu, 2008; Howell *et al.*, 2004). The increased flux has been generally attributed to the associated increase in crossflow velocity of the air-lifted liquid (Ueda *et al.*, 1997; Liu *et al.*, 2003; Xu and Yu, 2008). However, it is also known that the local flow pattern around an air bubble rising through a channel is very complex (Ghosh and Cui, 1999), exerting significant transient shear at the membrane surface and increasing the flux attained over that from liquid flow alone.

If the key attribute of the membrane aeration is the shear imparted, then it may be postulated that a key parameter representing aeration is the mean in-module upflow aeration velocity U:

$$SAD_p = \frac{Q_A}{Q_P} = \frac{UA_x}{JA} \quad \textbf{(4.4)}$$

where Q_A = aeration rate in $m^3.h^{-1}$

Q_p = permeate flow rate in $m^3.h^{-1}$

U = mean air flow velocity in channels in $m.s^{-1}$

A_x = open x-sectional area in m^2

A = membrane area in m^2

Combining Equations (4.2) - (4.4) produces the overall equation for specific energy demand associated with membrane aeration:

$$E_A = \frac{pT\lambda}{2.73x10^5 \varsigma(\lambda-1)}\left(\frac{UA_x}{JA}\right)\left[\left(\frac{10^4 y + p}{p}\right)^{1-\frac{1}{\lambda}} - 1\right] \quad \textbf{(4.5)}$$

The above equation reveals a number of interesting facets about the design of an MBR membrane aeration system:

- Energy demand increases linearly with A_x, the free cross sectional area (Judd, 2006), which is always higher for a flat sheet (FS) system than for a hollow fiber (HF) one. The limiting lower value of A_x is imposed by the propensity of the membrane channels to clog.
- The membrane area A can be increased by increasing the length of the module without detriment to the required volumetric aeration rate UA_x, leading to decreasing SAD_p and SAD_m values with constant design flux. On the other hand, increasing module length increases the aerator depth y, impacting negatively on energy demand.

A consideration of the geometries of the two membrane configurations (flat sheet, FS, and hollow fibre, HF) used for MBRs dictates that SAD_m and U follow the relationships:

Flat sheet $U = \frac{2LSAD_m}{\delta}$ Hollow fibre $U = \frac{SAD_m L}{\left(\frac{1}{\varphi} - \frac{d}{4}\right)}$ **(4.6a,b)**

where:

L	=	length of membrane in module, m
δ	=	channel separation (i.e. spacing of flat sheet membrane panels), m
φ	=	packing density (HF membrane area per unit module volume)
d	=	Hollow fibre outside diameter, m

For any given value of U and for a specific membrane length, and aerator depth, the relative energy demand of a FS and HF module is given by:

$$\frac{E_{A,FS}}{E_{A,HF}} = \frac{2\delta J_{HF}}{\left(\frac{1}{\varphi} - \frac{d}{4}\right) J_{FS}} \qquad \textbf{(4.7)}$$

Calculation from first principles of the actual aeration demanded for maintaining steady-state membrane permeation is thus far not possible. However, a study of membrane aeration impacts in terms of critical flux conducted at pilot plant scale on full-scale membrane modules (Guglielmi *et al*., 2007; Guglielmi *et al*., 2008, Table 4-1) suggests that the assumption of a linear correlation between the permeability and aeration rate to be reasonable, and that an inflection point arises at some threshold value beyond which the permeability is constant (Figure 4-1).

Table 4-1: Principal operating conditions for pilot study (Guglielmi *et al.*, 2007; Guglielmi *et al.*, 2008)

Parameter	Unit	FS1/FS2	HF
Membrane module	-	Kubota, E 50, single-deck	Zenon, ZW500c
Nominal pore size	μm	0.8	0.04
Membrane surface area	m^2	40	69.6
Module size (L x W x H)	Mm	(1000 x 600 x 2020)	(992 x 320 x 2085)
Biological process configuration	-	Pre-denitrification with internal recycle from aerobic to anoxic zone	Pre-denitrification with internal recycle from aerobic to anoxic
Membrane position	-	Immersed in aerobic tank	Immersed in aerobic tank
Anoxic volume	m^3	2.3	2.8
Aerobic volume	m^3	4.4	5.1
Filtration cycle	-	540 s filtration + 60 s relaxation	540 s filtration + 60 s relaxation
Net flow-rate, sub-critical	$m^3\ h^{-1}$	0.36-0.54	0.63-0.94
Flux J, sub-critical	$l\ m^{-2}\ h^{-1}$	10-15	10-15
HRT	H	11-17	8.4-12.6
Sludge age, SRT	D	12-15	12
MLSS in the biotank	$kg\ m^{-3}$	20 ± 1	10 ± 0.5
Chemical cleaning protocol	-	FS1: "strong" chemical clean (2000 g m^{-3} as Cl_2) every three months FS2: "weak" chemical clean (200 g m^{-3} as Cl_2) monthly	One "weak" chemical cleaning (200 g m^{-3} as Cl_2) every month
SAD_m	$Nm^3\ m^{-2}\ h^{-1}$	0.75-1.2	0.3-1, intermittent: 10 s ON/10 s OFF

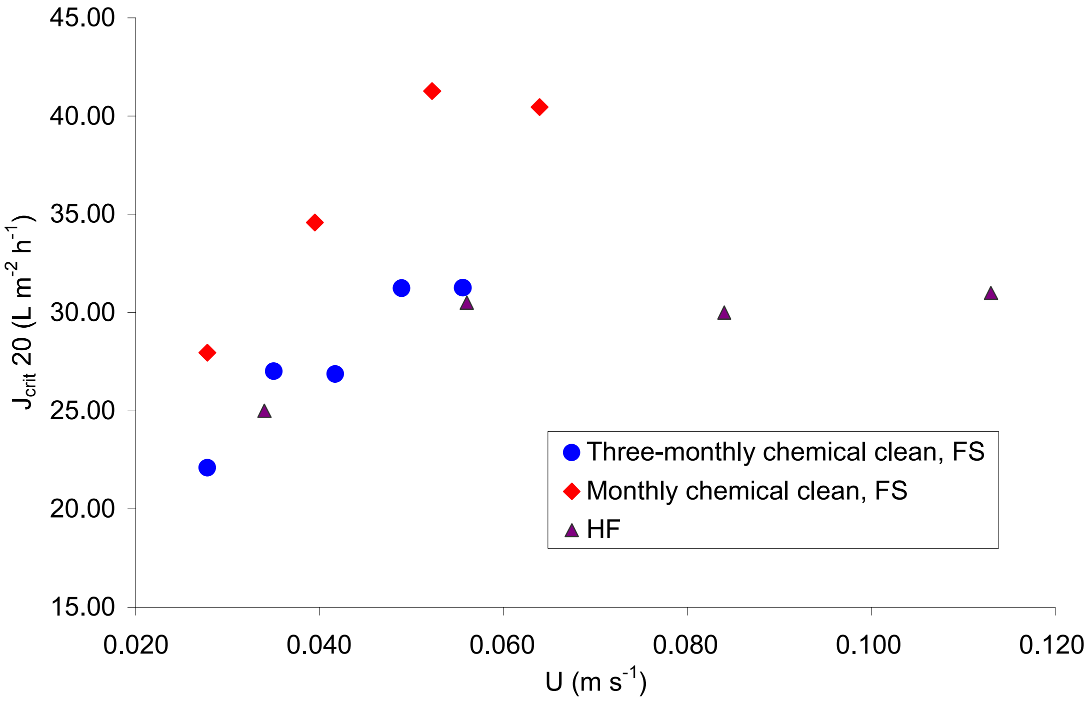

Figure 4-1: Net flux vs. permeability, pilot plant data (adapted from Guglielmi *et al.*, 2007, 2008)

Since energy would seem to relate to $U/(mU + J_0)$, it is critically dependent of the values of the two empirical constants m and J_0 (respectively the slope and intercept of the J vs U plot), and ultimately manifested as SAD_p (Equation 4.4). From the data in Figure 4-1 coupled with available literature information, and based on a reasonably conservative assumption, the following correlations can be used to represent the inter-relationship between J (in $l.m^{-2}.h^{-1}$)) and U (in $m.s^{-1}$):

$$J = mU + J_o \text{ for } J<25 \text{ l.m}^{-2}.\text{h}^{-1}) \qquad \textbf{(4.8)}$$

$$J = 25 \text{ for } U > U_{max} \text{ m.s}^{-1} \qquad \textbf{(4.9)}$$

It should be stressed that Equations 4.8 and 4.9 are derived empirically, and that published trends in permeability vs. aeration do not necessarily provide sufficient supplementary information to permit calculation of U. On the other hand, available information (Table 4-2) from demonstration (i.e. plant employing full-scale membrane modules, Table 4-1) and full-scale plant challenged with municipal wastewater suggest the range of U values (36% standard deviation) employed in practice to be smaller than that of either SAD_p (52% SD) or SAD_m (60% SD), and that these values can be presumed to be at or beyond the threshold value U_{max} in Equation 4.9. These data, all taken from plant operating under "optimal" conditions (i.e. the highest net flux sustained over a period of several weeks) indicate U value between 0.04 and 0.11 m/s. Differences in packing density mean that SAD_m values vary by a factor of 3.5 across all the plants for which data are listed in Table 4-2. It should be noted, however, that in the case of the HF module U is less meaningful since in this case the air is applied intermittently and its impact is less obviously related to shear. U is perhaps then best viewed as a representative normalised aeration parameter, in much the same way as SAD_m or SAD_p. For most full-scale immersed MBR installations currently in operation SAD_p generally exceeds 10, and can be as high as 50 (Judd, 2006), though recent studies have shown that membrane aeration can be reduced to below 5 by either more intermittent application (Garcès *et al.*, 2007) or redesign of the membrane module (Hai *et al.*, 2008).

Table 4-2: Summary design and operating data

	FS1 F, sd	**FS2 D, sd**	**FS3 F, dd**	**FS4 F, dd**	**FS 4 D, dd**	**HF 1 D**	**HF 2 F**	**Ave**	**SD %**
L, m	0.95	0.95	1.9	3.04	2.2	2.09	2.09		
δ, m	0.007	0.007	0.007	0.0065	0.007				
φ, $m^2.m^{-3}$						300	300		
d, m						0.0019	0.0019		
SAD_m	**0.75**	**0.88**	**0.27**	**0.42**	**0.21**	**0.25**	**0.43**	**0.45**	**60%**
U, $m.s^{-1}$	**0.057**	**0.066**	**0.041**	**0.109**	**0.037**	**0.051**	**0.087**	**0.064**	**36%**
Aeration	100%	100%	100%	100%	100%	50%	50%		
J, $l.m^{-2}.h^{-1}$)	**25**	**31**	**25**	**26**	**26**	**30**	**24**	**24**	**33%**
SAD_p	**30**	**28**	**11**	**16**	**8**	**8**	**18**	**20**	**52%**
REF	Judd, 2006	Table 1	Stone and Livingston, 2008	Table 3	Grélot *et al.*, 2007	Table 1	Table 3		

FS flat sheet; HF hollow fibre; F full-scale; D demonstration; sd single deck; dd double deck

4.3 BIOLOGICAL AERATION

Mathematical modelling of biological aeration is well established. For biotreatment the key aeration-related parameters pertain to the efficiency with which oxygen from the air is dissolved in the mixed liquor and then utilised by the microorganisms to degrade the organic material. The aerator is thus selected on the basis of oxygen transfer efficiency, which necessarily demands a fine bubble diffuser or jet aerator. This is to be distinguished from the coarse bubble aerator, producing larger bubbles with greater scouring efficacy, demanded for membrane aeration which none-the-less provides a proportion of the dissolved oxygen requirement. The total aeration demand thus proceeded through determination of the membrane aeration demand and calculation of the dissolved oxygen provided by the coarse bubble aeration. The total aeration demand for biotreatment can be calculated from modified classical Monod biokinetics, and the oxygen provided through membrane aeration subtracted from this. Determination of oxygen demanded for biomass degradation proceeds through the method proposed by Ekama *et al*. (1984), which accounts for oxygen requirements for active heterotrophic respiration, endogenous heterotrophic respiration, nitrification and denitrification. Values for constants used for calculations can be found in Table 4-3.

4.4 BENCHMARKING OF MODEL

The benchmarking of the model proceeded through comparison with two existing sites for which comprehensive information was available for the module design and the key operational parameters of SAD_m, SAD_p and E_A for both membrane and biological aeration, as well as product water quality (and in particular COD and ammoniacal and total N). Benchmarking proceeded through assuming appropriately conservative constant values of 100kPa and 5 $l.m^{-2}.h^{-1}$) for p and J_o respectively and then determining the values of ς and m using Equations 4.3 and 4.8. Note that blower energy losses could also be accounted for by adjusting p, but the net impact on specific aeration demand would be the same.

The two full-scale plants used comprised a double-deck FS plant and a HF plant. Both plants were operational for over two years and the data provided refer to operation under optimal conditions. In this case, optimal conditions refer to the highest flux sustained (i.e. with negligible permeability decline demanding unscheduled remedial cleaning) at full design flow. In both cases membrane aeration was fixed and the biological aeration intermittent and determined by the dissolved oxygen concentration set point. Summary data from benchmarking are shown in Table 4-4. According to these data the key values for the specific demand for both aeration and aeration energy are all within 20% of the plant values for both the FS and HF plants. From these data mean values of 247 and 56% can be taken for m and ς respectively. Other baseline parameter values used for the correlations were taken from textbook literature (Metcalf and Eddy, 2003) and are summarised in Table 4-5.

Table 4-3: Biological parameter values

Parameter	Units	Value
COD fractionation		
Fraction of readily biodegradable COD, f_{bs}	-	0.2
Fraction of slowly biodegradable COD, f_{bp}	-	0.5
Fraction of soluble unbiodegradable COD, f_{us}	-	0.05
Fraction of particulate unbiodegradable COD, f_{up}	-	0.25
Other biological input parameters		
Endogenous residu, f	-	0.2
COD/VSS ratio, f_{cv}	gCOD $gVSS^{-1}$	1.48
N content in VSS, f_n	gN $gVSS^{-1}$	0.1
Soluble unbiodegradable fraction of influent TKN, f_{nus}	-	0.03
Design parameters		
Anoxic fraction, f_{anox}	-	0.4
Process temperature, T	K	293
Process pH	-	7.2
DO concentration in the oxidation/nitrification tank, DO_{bio}	mg l^{-1}	2
DO concentration in the membrane tank, DO_{mbr}	mg l^{-1}	1.5
MLVSS/MLSS ratio, x_v/x	-	0.85
Recirculation ratio to anoxic zone, r_m	-	4
Biokinetics and stoichiometry		
Heterotrophic maximum growth rate at 20˚C, $\mu_{h,max,20}$	d^{-1}	6
Heterotrophic decay rate, $b_{h,20}$	d^{-1}	0.24
Autotrophic maximum growth rate at 20˚C, $\mu_{n,max,20}$	d^{-1}	0.36
Autotrophic decay rate at 20˚C, $b_{n,20}$	d^{-1}	0.04
Half-saturation constant for readily biodegradable COD at 20˚C, $K_{s,20}$	gCOD m^{-3}	20
Half-saturation constant for ammonia nitrogen at 20˚C, $K_{n,20}$	gN m^{-3}	1
Half-saturation constant for dissolved oxygen K_o	gO_2 m^{-3}	0.1
Denitrification rate over readily biodegradable COD at 20˚C, $K_{den1,20}$	gN-NO_3 $gVSS^{-1}$ d^{-1}	0.72
Denitrification rate over slowly biodegradable COD at 20˚C, $K_{den2,20}$	gN-NO_3 $gVSS^{-1}$ d^{-1}	0.1008
Heterotrophic yield, Y_h	gVSS $gCOD^{-1}.d^{-1}$	0.45
Autotrophic yield, Y_n	gVSS $gN.d^{-1}$	0.1

All correlations refer to 20˚C, temperature correction coefficients are excluded.

The protocol for benchmarking was as follows:

1. Set flux J.
2. From module characteristics determine A_x and aerator depth y (Judd, 2006).
3. Determine air flow per unit membrane area (SAD_m) and combine with y and J to provide specific membrane aeration energy demand from Equation 4.5, selecting the most conservative value of ς and m to match with specific aeration energy demand per m^3 air.
4. From air flow rate and physical characteristics determine mass flow of oxygen provided by membrane aeration, assuming the MLSS concentration in membrane tank to be 40% higher than that in the biotank.
5. From feedwater and effluent quality and flow, determine all biological parameters/oxygen requirement for biological degradation according to the method proposed by Ekama *et al.* (1984).
6. Determine total mass flow of dissolved oxygen provided by membrane aeration by multiplying $m_{o,\text{membrane}}$ by A_{req}, the required total membrane area, that can be determined as the ratio of the designflow Q_p and the flux J as specified in Equation 4.9. Key values for calculation of $m_{o,\text{membrane}}$ can be found in Table 4-5.

$$m_{0,membrane} = Q_A \cdot \rho_{air}(OTE_{coarse} \cdot y) \cdot \alpha \cdot \beta \cdot \psi(\frac{\%O_2 inair}{100}) \qquad \textbf{(4.10)}$$

where:

$$\alpha = e^{-\omega X} \qquad \textbf{(4.11)}$$

7. Subtract total mass flow of dissolved oxygen provided by membrane aeration from total mass flow of oxygen required calculated from biokinetics.
8. Determine $Q_{A,\text{biotank}}$ by difference and from air properties (Table 4-5), and determine specific biotank aeration energy demand.
9. Add to energy demand associated with permeation to provide total energy.

Table 4-4: Predicted model and reported values

Parameter	Units	FS			HF		
		Pred.	**Actual**	**%diff**	**Pred.**	**Actual**	**%diff**
Membrane							
SAD_m	$Nm^3.m^{-2}.h^{-1}$)	0.439	0.439	0	0.430	0.430	0
SAD_p	-	17.6	16.4	6	17.2	17.2	0
E'_A (per m^3 air)	$kWh.Nm^{-3}$	0.022	0.022	0	0.013	0.013	0
E_A (per m^3 permeate)	$kWh.m^{-3}$	0.390	0.365	6	0.229	0.200	15
m*		228	-	-	264	-	-
ς*	%	58	-	-	54	-	-
Biology							
E_A (per m^3 permeate)	$kWh.m^{-3}$	0.194	0.165	18	0.028	0.035	-20
Total E	$kWh.m^{-3}$	0.584	0.530	10	0.257	0.235	~9
Product water quality							
COD	$mg.l^{-1}$	23.4	24	3	4.8	4.8	0
Total N	$mg.l^{-1}$	6.4	2.9-5.3	-	3.2	3-9	-
Tank volumes							
Total biotank volume	m^3	693	N/a**		9,986	9,314	7
Anoxic volume	m^3	243	N/a**		3,545	3,525	1
Aerobic volume	m^3	451	N/a**		6,441	5,789	10
Overall HRT	h	6.8	7	-2	4.7	4.7	0

*Values computed from normalisation; **Cannot be determined due to hybrid CAS – MBR plant configuration.

4.5 RESULTS AND DISCUSSION

The impact of HRT, membrane configuration, stacking and membrane aerator nozzle depth on flux and aeration energy demand is shown if Figure 4-2 and Figure 4-3. Baseline parameter values resulting from benchmarking against existing plants or literature (Günder, 2001; Krampe and Krauth, 2003; Metcalf and Eddy, 2003) are listed in Table 4-5. The trends are reported for a HRT range from 4 hrs to 12 hrs, reflecting operating conditions found in operating full scale plants. At an HRT of 4h the flux was assumed to be 25 LMH, corresponding to the conservative maximum value as specified in Equation 4.8. Variations in SRT resulting from variations in HRT at fixed MLSS concentration lead to corresponding changes in effluent ammonia $N_{a,out}$ concentration, since the SRT calculation is based on the growth rate of the slow growing nitrifiers (Ekama *et al.*, 1984).

Table 4-5: Baseline parameter values for calculation of correlations

Parameter	Units	Value
Membrane		
Flux J	$l\ m^{-2}\ h^{-1}$	*25*
m		*247*
ς	*%*	*56*
Module parameters		
$A_x/A_{FS,\ single\ deck}$	-	*0.00417*
A_x/A_{HF}	-	*0.00157*
Aeration		
OTE_{coarse}	$\%\ m^{-1}$	*2.5*
OTE_{fine}	$\%\ m^{-1}$	*4.5*
Fine bubble aerator depth y_{fine}	*m*	*5*
ω	-	*0.084*[1]
β	-	*0.95*
$\%O_2$ *in air*	*%*	*21*[2]
ρ_{air}	$kg\ m^{-3}$	*1.23*
ψ *(T=273K)*	-	*1.024*[2]
Influent characterisation[2]		
CODin	$mg\ l^{-1}$	*430*
TKNin	$mg\ l^{-1}$	*40*
$(N\text{-}NO_3)in$	$mg\ l^{-1}$	*0*
Design parameters		
MLSS in biotank	$mg\ l^{-1}$	*8,000*
SRT	*D*	*19*
$N_{a,out}$	$mg\ l^{-1}$	*0.81*
Anoxic fraction f_n	-	*0.4*
Process temperature	*K*	*293*
Process pH	-	*7.2*
Fine bubble aerator blower efficiency ς_f	*%*	*48.7*

[1] Günder, B. (2001), Kramp and Krauth (2003); [2]Medium strength sewage (Metcalf and Eddy, 2003)

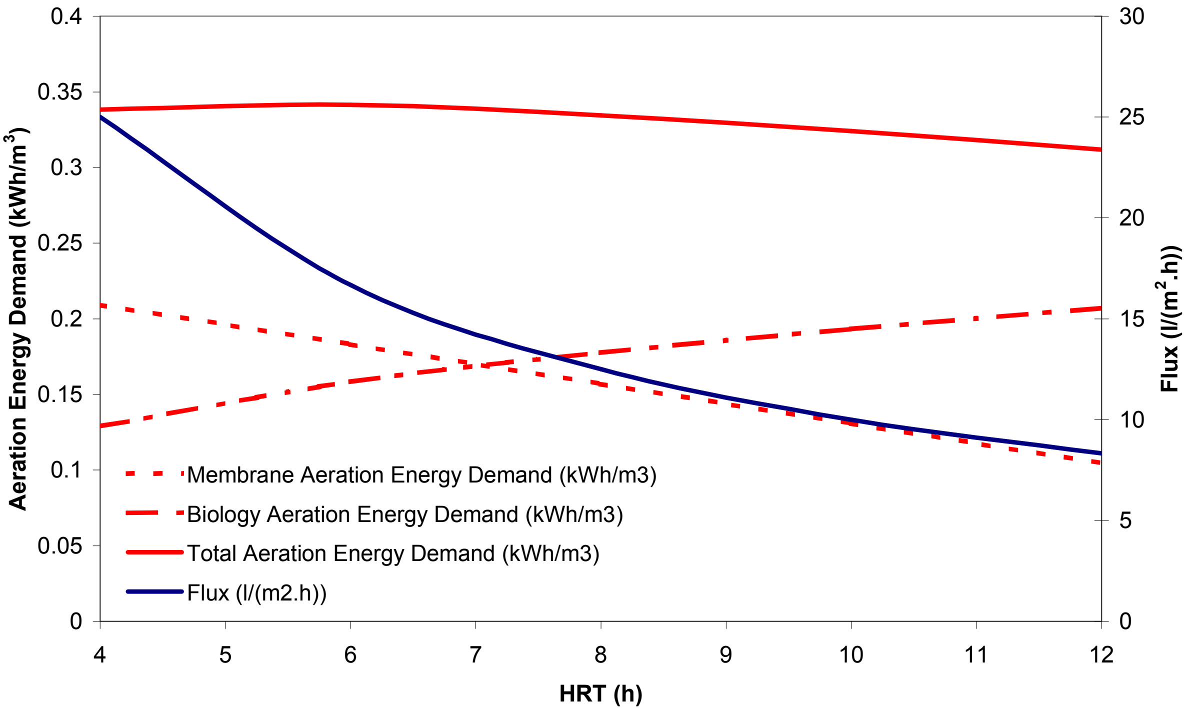

Figure 4-2: Projected aeration energy demand with HRT for Hollow Fibre modules: Contribution from predicted Membrane and Biological aeration energy demand to Total Aeration Energy Demand

Figure 4-2 demonstrates that the impact of the increased biological aeration demand at longer HRTs due to endogenous respiration is countered by the decrease in membrane aeration demand at the lower SAD_m values required at lower fluxes incurred. Thus, a decrease in specific membrane energy demand could be expected if membrane aeration can be adjusted according to the feed flow rate and the blower energy demand decreased proportionately. Overall, the change in total aeration energy demand with HRT is negligible for both HF and FS modules, contrary to reported trends of increased energy demands at longer HRTs for an FS module at ~12 g l^{-1} MLSS (Stone and Livingston, 2008).

The impact of membrane configuration is demonstrated in Figure 4-3. A standard single-deck FS MBR is around 20% higher in specific aeration demand than an HF, with this figure only decreasing slightly for a FS double-deck plant with an aerator depth y of 5m. This demonstrates the impact of both A_x/A and y on energy demand. The impact of the former is demonstrated by Equation 4.7; for the same flux and overall effective membrane length and for typical values of 0.007 m, 300 m^{-1}, 0.002 m for the membrane separation δ, HF module packing density φ and fibre diameter d

respectively, the ratio of HF to FS membrane energy demand is ~0.8. Increasing the volumetric aeration efficiency by using a double-deck plant decreases SAD_m and SAD_p by a factor of 2 but almost commensurately increases the energy demand due to the increased hydrostatic head. This demonstrates the value of reducing the gap between the upper and lower decks.

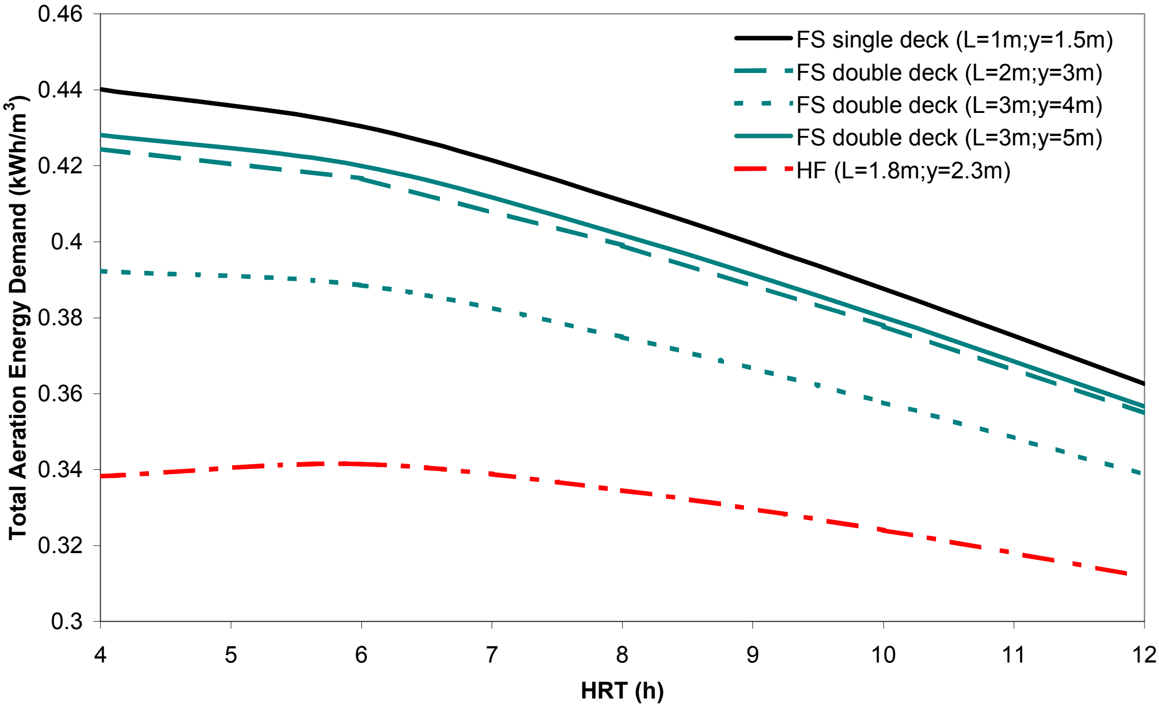

Figure 4-3: Projected Total Aeration Energy Demand with HRT: Impact of membrane aerator nozzle depth y and membrane length L for Flat Sheet and Hollow Fibre modules

The impact of the key empirical parameters of slope *m* and intercept J_o of the *J* vs *U* plot on the membrane aeration energy demand for FS modules correlated against *J* is shown in Figure 4-4. In this case a J_{max} of 30 $l.m^{-2}.h^{-1}$) was assumed, this being the maximum sustainable flux recorded according to the sample data in Table 4-2. It is shown that the energy demand decreases with increasing (less conservative) values for *m* and J_o, which can be attributed to a decreasing value for the upflow gas velocity *U*. Note that, according the data in Figure 4-1, *m* varies between 391 and 544 whilst J_0 is around 12-13 l m^{-2} h^{-1}. Accordingly, the membrane aeration energy for this system would approach 0.1 kWh m^{-3} at 25 l m^{-2} h^{-1} at the highest *m* value.

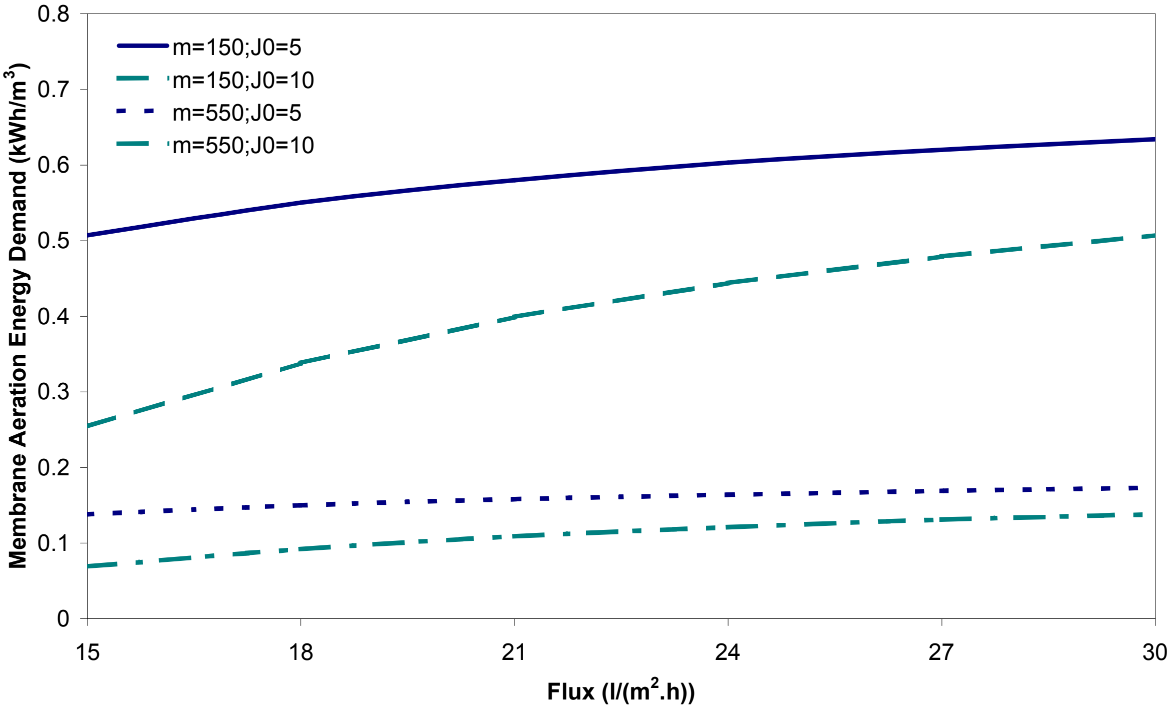

Figure 4-4: Impact of key empirical model parameters slope m and intercept J_o of the *J* vs *U* plot on the Membrane Aeration Energy Demand for Flat Sheet single deck modules correlated against flux *J*

A simple sensitivity analysis was conducted for the key biokinetic and baseline parameter values. For this analysis, an 8hr HRT, considered to be representative for realistic average conditions, was assumed to correspond to the maximum flux of 25 LMH. Table 4-6 shows that of the key operating parameters, variations in MLSS concentration have the most profound influence on aerator energy demand due to its influence on the α-factor. The increase in $E_{A,total}$ caused by increasing SRT or increasing HRT (at fixed SRT) can also be attributed to the corresponding increase in MLSS concentrations, while increasing HRT at fixed MLSS concentrations shows that the influence of variations in SRT is minimal. The oxygen transfer efficiency, for both coarse and fine bubble aeration, has a profound effect on aeration energy demand, while the influence of aerator depth *y* seems negligible since the increase in oxygen transfer at greater depths is offset by the higher hydrostatic head. Using less conservative values for *m* and J_0 decreases projected energy demand, as also shown in Figure 4-4, while increasing *U* demonstrates the impact of aeration beyond the optimum threshold value, resulting in extra aeration energy demand without an increase in possible sustainable flux as proposed in equation 4.9. As can be expected, more efficient blowers lead to a proportionate decrease in energy demand. Influent strength impacts on energy demand as expected from biological treatment, and changes in the biokinetic constants have a negligible impact on the aeration energy demand compared to changes to the process parameters. Of all controllable parameters, it appears that SRT and flux have the most significant impact on aeration energy demand.

Table 4-6: Sensitivity analysis for HF modules

Parameter	Change in parameter		
	10%	30%	50%
Baseline specific aeration energy demand = 0.391 kWh.m^{-3}	%diff	%diff	%diff
Key operating parameters			
HRT – MLSS fixed	-0.01	-0.3	-1.0
HRT – SRT fixed	-5.5	-13.8	-20.0
SRT	5.5	17.0	29.5
MLSS	6.3	20	35.1
r_m	0	0	-0.01
Aeration parameters			
OTE_{coarse}	-2.4	-7.2	-12.0
OTE_{fine}	-4.2	-10.7	-15.4
Fine bubble aerator depth y	-0.6	-1.6	-2.6
M	-2.7	-6.9	-9.9
J_0	-0.7	-2.2	-3.7
U	3.0	8.9	14.9
Blower characteristics			
P	0.9	2.2	3.3
Coarse bubble blower efficiency ς	-4.9	-12.4	-17.9
Fine bubble blower efficiency ς_f	-4.2	-10.7	-15.4
Biomass characteristics			
Ω	5.6	17.7	31.2
Influent characteristics			
Sewage strength* - MLSS fixed	-29.2	0.0	35.6
Sewage strength* - SRT fixed	-44.0	0	164.3
Fraction of soluble and particulate unbiodegradable COD, $f_{us}+f_{up}$	-2.8	-7.8	-13.0
Biokinetic constants			
DO_{bio}	0.00	0.01	0.01
Soluble unbiodegradable fraction of influent TKN	-0.1	-0.2	-0.3

*According to values for low, medium and high strength municipal sewage (Metcalf and Eddy, 2003)

4.6 CONCLUSIONS

Using a simple model to assess the impact of key design and operating parameters on the aeration energy demand of a submerged membrane bioreactor, it can be concluded that significant reductions in energy demand are attainable through operating at:

- lower MLSS levels, the precise impact depending on the nature of the α-factor:MLSS relationship which published evidence suggests is an exponential decline (Günder, 2001; Krampe and Krauth 2003).
- lower fluxes, the influence of which depends on the values of m and J_0 in Equation 4.8; evidence provided (Figure 4-1) implies that halving the flux can reduce the aeration energy demand by as much as 45%.

There are, however, other implications of flux and SRT reduction. Reducing the flux commensurately increases the required membrane area and thus the capital cost. Decreasing the MLSS increases the sludge production, and can also increase the fouling propensity (Trussell *et al.*, 2007). There is therefore a direct financial penalty for operating under more conservative conditions, unless the latent energy of the sludge can be recovered on site.

On the other hand, there is a direct financial benefit to be gained through the adoption of a membrane aeration device and/or regime which (a) enhances oxygen transfer to increase its utilisation for aerobic treatment and/or (b) adjusts in intensity according to the hydraulic load. Enhancing oxygen transfer may only be achievable through creating smaller air bubbles which would then be to the detriment of membrane scouring. Adjusting the intensity of aeration, however, would appear to be more viable. Flows through sewage treatment works vary significantly both diurnally and seasonally, yet the current practice is to provide membrane and membrane aeration capacity for the highest flows and, generally, provide little buffering capacity. A rudimentary analysis using the baseline conditions provided in Table 4-5 and assuming a reasonable conservative flux:aeration relationship of $J = 350U + 7.5$ reveals that the energy demand can reduced by around 20% if the membrane aeration is applied proportional to the flow and the latter is at half the plant capacity for half of the time. This figure increases as the sewage strength decreases. A protocol can be envisaged whereby the number of membrane tanks on line and/or the membrane aeration intensity (in effect the number of blowers in service) can be adjusted according to the flow. Whilst this would add another layer of process complexity onto what is already regarded as a somewhat Byzantine process, ever increasing energy costs may ultimately make such adaptations inevitable.

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CHAPTER 5

EXPERIMENTAL EVALUATION OF INTERMITTENT AERATION OF A HOLLOW FIBRE MEMBRANE BIOREACTOR

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5 EXPERIMENTAL EVALUATION OF INTERMITTENT AERATION OF A HOLLOW FIBRE MEMBRANE BIOREACTOR

B. Verrecht[1], C. James[2], E. Germain[2], W. Ma[3], and S. Judd[1,]*

[1]Centre for Water Science, Cranfield University, Cranfield, Bedfordshire MK43 0AL, UK

[2]Thames Water R&D, Innovation Centre, Island Road, Reading, Berkshire RG2 0RP, UK

[3]Shenyang Research Institute of Chemical Industry, Shenyang, China

ABSTRACT

Intermittent membrane aeration provides a substantially improved energy efficiency in hollow fibre-based immersed membrane bioreactors (HF iMBRs). The benefits of intermittent aeration have been assessed with respect to sustaining a target flux and/or limiting the fouling rate to a sustainable level based on a small plant using full-scale HF modules. Results show that for the same specific aeration demand per unit of permeate produced (SAD_p), fouling rates were significantly lower for 10s filtration, 30s relaxation ("10:30" intermittent aeration) compared to 10:10 and continuous aeration. At a net flux (J_{net}) of 23.3 litres m^{-2} h^{-1} (LMH), a SAD_p of 4.6 was found sufficient to sustain operation, this value being up to 75% and 50% lower compared to continuous and 10:10 aeration respectively. This empirical data was compared with heuristic data from 5 large scale HF iMBR plants, which revealed that 10:30 aeration can sustain a relatively high flux (up to 25.3 LMH) under dry weather conditions in warm climates, with the recorded SAD_p ranging from 5.3-10.9.

5.1 INTRODUCTION

Membrane bioreactors (MBR) are now well established for treating wastewaters to provide a high effluent quality whilst incurring a low footprint, but are nonetheless limited by the high energy demand incurred by aeration (Verrecht *et al*. 2008; Judd and Judd, 2010). Aeration, required both for the biological process and for membrane scouring, can contribute significantly to operating costs, along with other process facets intended to maintain membrane permeability such as chemical and physical cleaning and membrane replacement (Verrecht *et al.*, 2010b). However, it is aeration which provides the primary adjustable parameter for reducing energy demand and has formed the focus of a number of studies at demonstration and/or full scale (Garcès *et al.*, 2007, Verrecht *et al.*, 2010a; Tao *et al.*, 2009; Pawloski *et al.*, 2008).

Several studies have focused on the impact of aeration on critical (J_c) or sustainable flux (J_{sust}). Based on experiments on a large pilot scale HF MBR, Guglielmi *et al.* (2007) reported an increase in J_c with increasing membrane aeration up to a certain membrane area specific aeration demand (SAD_m) threshold value (between 0.3-0.5 $Nm^3.h^{-1}$ per m^{-2} membrane area), above which no further increase in J_c is observed, corroborating results obtained by other authors for flat sheet (Ueda *et al.*, 1997; Ndinisa *et al.*, 2006; Guglielmi *et al.*, 2008) and tubular modules (Le Clech *et al.*, 2003). These data all indicate that membrane aeration energy demand can be reduced through adjusting membrane aeration proportional to flux. Moreover, Meng *et al.* (2008) report that both small and large aeration intensities have a negative impact on permeability; low aeration intensities do not effectively remove membrane foulants, leading to build up of a cake layer, whilst excessive aeration intensities break up the sludge flocs which increase pore blocking. This suggests a possible decrease in J_{sust} above a threshold aeration intensity.

One of the main improvements in energy consumption for HF membranes has been the introduction of intermittent aeration, specifically by limiting aeration for 10s every 20s ("10:10" aeration) or every 40s ("10:30" aeration), as patented by GE Zenon (2007). However, published research in this area is limited, and is focused primarily at laboratory and pilot scale. Jiang *et al.* (2005) studied the impact of different aeration cycles on MBR fouling with real municipal wastewater and a *ZeeWeed 500C* module on the pilot scale. Over a 5 week study at a flux of 30 LMH and a temperature of 17-18°C, the reported filtration resistance increased by 42.6% on changing the aeration frequency from 10:10 (SAD_m = 0.42) to 10:30 (SAD_m = 0.21) and a further 9.5% when changing to 10:60 aeration (SAD_m = 0.14). Mansell *et al.* (2006) studied the fouling rate increase with flux at different aeration frequencies (10:10 and 10:30) over a 7 month period. At a mean temperature of 23°C and a flux of 20.4 LMH and 10:10 aeration at a SAD_m of 0.36, fouling was insignificant over a two month period, while fouling rates of approximately 0.18 and 0.66 $LMH.bar^{-1}.h^{-1}$ were measured at fluxes of 25.5 and 30.6LMH respectively. Under 10:30 aeration conditions and a SAD_m of 0.18, the fouling rates were 0.66 and 1.32 $LMH.bar^{-1}.h^{-1}$ respectively at a flux of 17.7 and 20.4 LMH. However, in both the above studies both the intermittency of the aeration and the aeration demand were varied, SAD_m decreasing with intermittency. Contrary to these findings, Monclùs *et al.* (2010) identified a critical specific aeration demand ($SAD_{m,crit}$) on a pilot scale HF MBR of 0.19 vs 0.06 $Nm^3.m^{-2}.h^{-1}$ under 10:10 and continuous aeration respectively at the same flux.

It is clearly necessary to uncouple aeration intermittency and demand and establish the relative impact of each of these parameters under comparable conditions on a representative system, i.e. based on a full-sized membrane module. This can be achieved through comparing fouling rates, i.e. the rate of increase in transmembrane pressure (TMP) with time for different aeration regimes, under conditions of:

a) different SAD_m values for continuous aeration, and
b) constant SAD_m values at aeration intermittencies of 50% (10:10 aeration) and 25% (10:30 aeration).

Results from an empirical study are presented and compared with data from heuristic data from five established, large, full-scale HF MBRs.

5.2 MATERIAL AND METHODS

5.2.1 Plant description

A small full-scale MBR is located at a sustainable development in South London and produces an average reclaimed water flow of 25 $m^3 \cdot d^{-1}$ for toilet flushing and irrigation (Figure 5-1). Domestic wastewater from the residences is collected via a pumping station and septic tanks, which provide buffering volume and primary settling. Influent from the septic tanks flows through 3 mm screens to the MBR, which contains both anoxic (10.1 m^3) and aerobic zones (12.5 m^3) for nitrification and denitrification respectively. The membrane separation step is provided by 2 x 3 ZW500c (GE Zenon, Canada) membrane modules of 139.2 m^2 total membrane surface area, submerged in the aerobic zone. More detailed information about the plant and its performance can be found in Verrecht *et al.* (2010a).

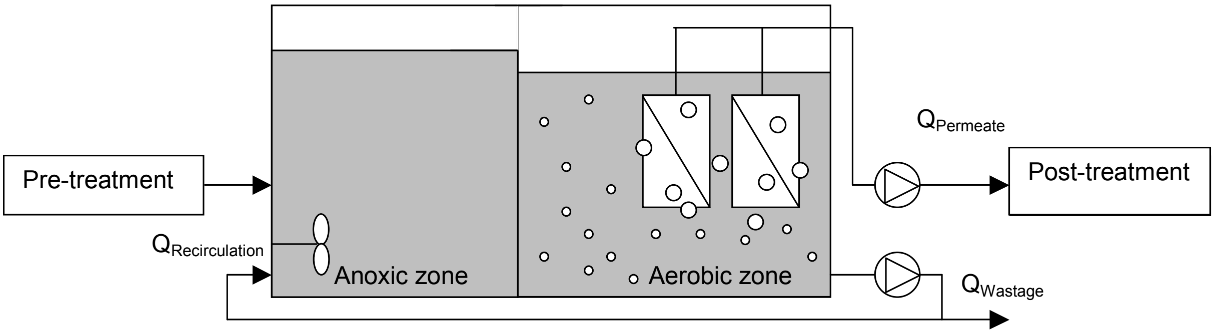

Figure 5-1: Schematic overview of the wastewater recycling plant

5.2.2 Experimental procedure for gathering of empirical data

Empirical data regarding the impact of aeration on sustainable permeability was gathered over a three month period. Increased fluxes were achieved by taking one membrane cassette out of operation, resulting in a reduction in effective membrane surface area to from 139.2 to 69.6 m^2. Table 5-1 provides an overview of the plant operating conditions over the experimental period, according to the following protocol:

- Prior to the start of an experiment, the membranes were relaxed for a minimum of one hour followed by a backflush at a J_{net} of 23.3 LMH for a duration of 5 minutes.
- The desired J_{net} and membrane aeration, in terms of specific aeration demand (SAD_m in Nm^3.m^{-2}.h^{-1}) were set. Membrane aeration was either continuous, intermittent 10'' on – 10'' off (10:10 aeration), or intermittent 10'' on – 30'' off (10:30 aeration). When operating with intermittent aeration, the air was diverted away from the adjacent unused membrane cassette during the 'off' periods to ensure that no interference occurred from aeration of the nearby cassette.
- The trial duration was either of 6 hours duration or until the permeability K (LMH.bar^{-1}) had declined to below 60 LMH.bar^{-1}.
- After each experiment, membrane aeration was increased to 75 Nm^3.h^{-1} (SAD_m = 1.25 Nm^3.m^{-2}.h^{-1}) overnight to remove potential cake layer build up on the membrane surface, and the flux was set at a J_{net} of 17.2 LMH. These standard conditions were maintained overnight to return the initial membrane permeability to a target value of 94.4 +/- 2.2 LMH/bar.

Experiments were carried out at three different net fluxes (J_{net} of 23.3, 16.2 and 11.3 LMH), while membrane aeration ranged from a SAD_m of 0.02 to 0.56 Nm^3.m^{-2}.h^{-1} for continuous aeration conditions, and from 0.05 to 0.22 Nm^3.m^{-2}.h^{-1} for intermittent aeration. Following each set of experiments at a certain J_{net}, a maintenance clean (cleaning-in-place; CIP) was carried out by backflushing with 500 ppm NaOCl to maintain the initial permeability at around 98.6 +/- 2.1 $LMH.bar^{-1}$ before conducting the next series of experiments, corresponding to chemical cleaning roughly every two weeks.

The MLSS concentration was measured daily according to standard methods (APHA, 2005), and capillary suction time (CST) was measured twice weekly using a Triton 304B CST analyser (Triton Electronics Ltd., UK). Measurements for both CST and MLSS were carried out in duplicate, with daily values generally within 5% of each other, and an average was taken (Table 5-1). The wastewater strength, as determined by standard methods, was high (Table 5-2), as the MBR was fed with domestic wastewater without rainwater dilution from dwellings with average water consumption of 80-100 $l.capita^{-1}.d^{-1}$. Effluent quality (Table 5-2) was consistently high and in line with widely reported values for MBR (Judd and Judd, 2010). The developed sludge was readily filterable and of consistent quality, according to the CST data (Table 5-1).

Table 5-1: Operational parameters

Filtration parameters			**Plant and sludge characteristics**		
Parameter	Unit	Value	Parameter	Unit	Value
Membrane surface	m^2	69.6	MLSS	$g.m^{-3}$	6378 ± 653
Filtration time	S	600	CST		73.2 ± 14.1
Relaxation time	S	30	SRT	d	50
Backwash time	S	30	HRT	h	24
J_{net}	LMH	23.3-16.2-11.3	Temperature	°C	18-26.5
SAD_m; Continuous	$Nm^3.m^{-2}.h^{-1}$	0.02 – 0.56	Recirculation ratio	-	2.3
SAD_m; Intermittent	$Nm^3.m^{-2}.h^{-1}$	0.05 – 0.22			

Table 5-2: Mean wastewater and effluent quality determinants over course of trials, twice weekly grab samples

Parameter	Influent Value, $g.m^{-3}$	Effluent Value, $g.m^{-3}$
BOD_5	352 ± 44	1.2 ± 0.9
COD	714 ± 65	28.5 ± 10.5
NH_4-N	94 ± 5	0.6 ± 2.1
NO_3-N	< 0.3*	23 ± 11.0
PO_4-P	12 ± 1	7.9 ± 6.3
SS	143 ± 17	<2*

* Below detection limit

5.2.3 Data processing

An average value for the permeability K (temperature corrected to 20°C) of each ten minute filtration cycle was determined for the duration of the experiments (Le Clech *et al.*, 2003). A polynomial curve was fitted to the permeability vs. time plots, allowing determination of the permeability decline $\Delta K/\Delta t$ (LMH.bar^{-1}.h^{-1}) either after 4 hours, or at the moment when permeability had declined to below 60 LMH.bar^{-1}. To assess whether operation was sustainable under the chosen combination of J_{net} and SAD_m, it was assumed that full scale plants undergo a CIP maintenance cleaning twice weekly, such that the maximum permeability decline rate for sustainable operation was 0.48 LMH.bar^{-1}.h^{-1}; this corresponded to a decline in permeability from 100 to 60 LMH.bar^{-1} over a 3.5 day period.

5.3 RESULTS AND DISCUSSION

5.3.1 Empirical data: benefits of intermittent aeration

Figure 5-2 shows the permeability profile vs time for a net flux J_{net} of 23.3 LMH and three different aeration modes: continuous, 10:10 and 10:30, at the same overall SAD_m of 0.11 Nm^3.m^{-2}.h^{-1}. The permeability decline rate decreased from 19.3 LMH.bar^{-1}.h^{-1} for continuous aeration to 5.45 and 0.33 at 10:10 and 10:30 aeration respectively. More vigorous aeration at shorter time periods is thus more effective in maintaining permeability than continuous aeration at a lower air flow rates. In all cases the permeate volume-normalised membrane aeration demand (SAD_p) was 4.6 Nm^3 air per m^3 of permeate; SAD_p, the ratio of SAD_m to flux, is directly proportional to specific energy demand in kWh/m^3.

The correlation of overall permeability decline rate $\Delta K/\Delta t$ with SAD_p for the three aeration modes is provided in Figure 5-3 for a net flux J_{net} of 23.3 LMH. This figure indicates the permeability decline rate decreases roughly exponentially with SAD_p. The figure also shows that the target maximum permeability decline rate of 0.48 LMH.bar^{-1}.h^{-1} is achieved at the lowest SAD_p value of 4.6 for 10:30 aeration, whereas this is clearly not the case for 10:10 or continuous aeration. For the latter, the most extensively tested aeration mode, the required SAD_p value appears to be around 21 – at least four times higher than the aeration demand for 10:30 aeration. For 10:10

aeration the threshold value is indeterminate but is between 4.6 and 9.2, and evidently closer to the upper limit.

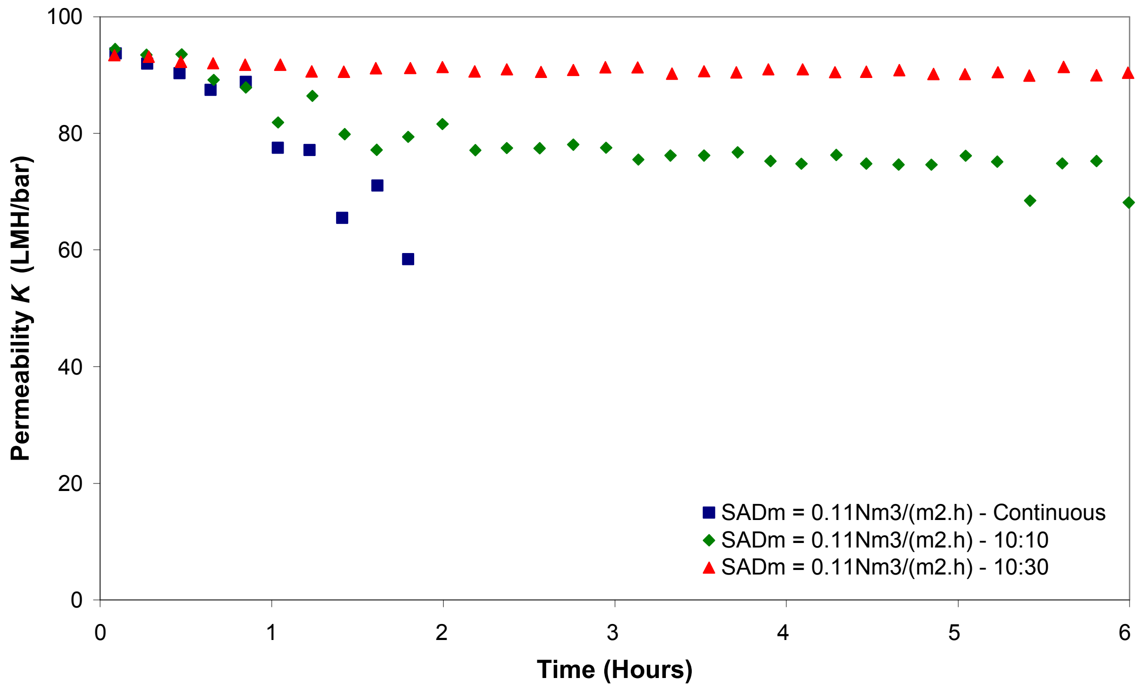

Figure 5-2: Permeability vs. time for J_{net} = 23.3 LMH: comparison of continuous aeration, 10:10 and 10:30 at the same overal membrane aeration (SAD_m=0.11Nm^3.m^{-2}.h^{-1})

A correlation of $\Delta K/\Delta t$ against J_{net} at constant SAD_m (Figure 5-4) indicates a linear dependency for all three aeration modes. According to this figure, the fouling rate for continuous aeration is 16-60 times higher than that for 10:30 aeration over fluxes of 11-24 LMH; the benefit of intermittent aeration becomes more significant as flux increases. In previous studies of immersed HF MBRs by Garcès *et al.* (2007) and Manser *et al.* (2006) unsustainable permeability declines for 10:30 aeration were reported at some threshold flux value. A fairly modest permeability decline rate of 0.51 LMH.bar^{-1}.h^{-1} at an unusually high J_{net} value of 34 LMH and a SAD_m of 0.187 was reported by Garcès *et al.* (2007), compared to a higher value decline rate of 0.66 LMH.bar^{-1}.h^{-1} at a SAD_m of 0.18 and a much lower J_{net} of 17.7 LMH recorded by Manser *et al.* (2007). There is thus significant variation in fouling propensity between studies, presumably because of the nature of the sludge generated.

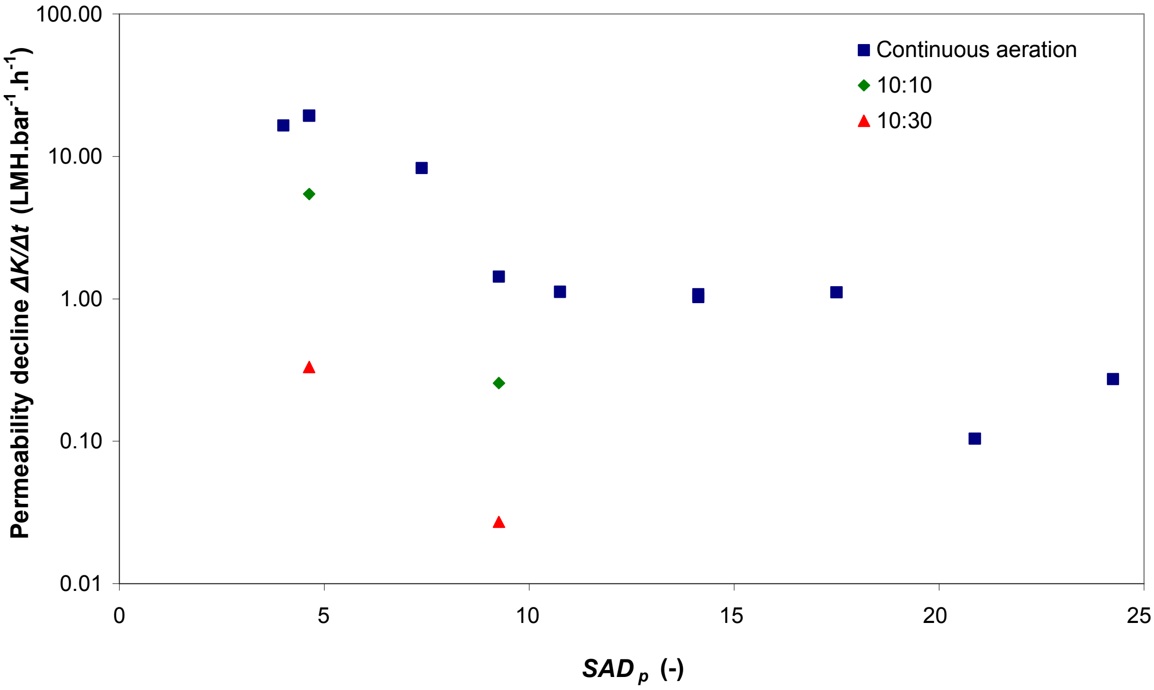

Figure 5-3: Permeability decline *ΔK/Δt* vs. membrane aeration demand per unit of permeate produced SAD_p for J_{net} = 23.3 LMH

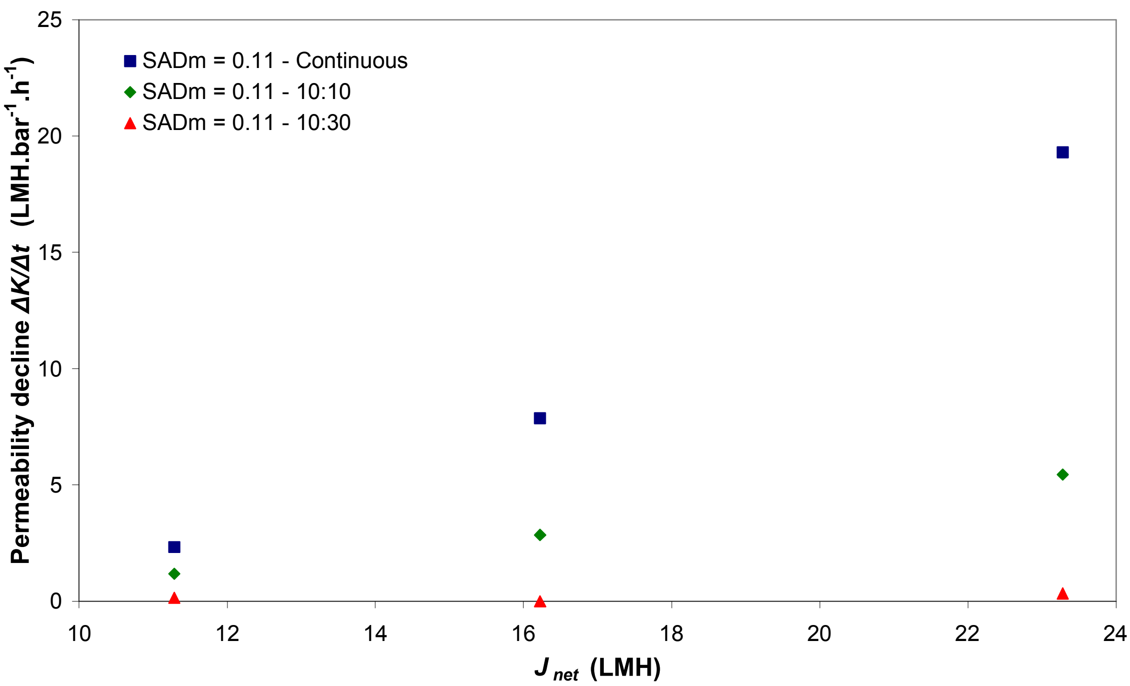

Figure 5-4: Permeability decline vs J_{net} at constant SAD_m = 0.11 $Nm^3.m^{-2}.h^{-1}$; comparison of continuous aeration vs. 10:10 and 10:30.

The results from the trials are generally in keeping with more recently reported industrial practice regarding intermittent aeration (Pawloski *et al.*, 2008; Tao *et al.*, 2009). Since membrane aeration energy generally accounts for 30-50% of the total plant energy demand depending on plant utilisation (Garcès *et al.*, 2007, Verrecht *et al.*, 2008, 2010b), minimising membrane aeration clearly has the largest single impact on total plant energy costs. Pawloski *et al.* (2008) reported a reduction of 5% in the total plant energy bill for a 12 MLD MBR when changing from 10:10 to 10:30 aeration under average daily flow (ADF) conditions without compromising hydraulic performance. Garcès *et al.* (2007) reported a decrease in membrane aeration requirements of 35% when using 10:30 aeration under low flow conditions, which could result in a reduction in total plant energy demand of up to 10.2%, while Mansell *et al.* (2006) estimated that total plant power requirements could decrease by 20% on switching from 10:10 to 10:30 aeration.

5.3.2 Comparison with heuristic data from 5 large scale HF MBRs

Heuristic data obtained from 5 large (>20,000 $m^3.d^{-1}$ average daily flow) full-scale HF MBRs (Table 5-3) indicate that a SAD_p of 10.6-16.7 is required at peak flows using 10:10 aeration and 5.3-10.9 at dry-weather flows using 10:30 aeration. 10:30 aeration protocols are already in use at large scale plants and can apparently sustain a relatively high flux (up to 25.3 LMH) under the appropriately benign conditions of temperature and sludge quality (Judd and Judd, 2010). The lowest required SAD_p values (5.3 and 5.5) are obtained for MBR plants that operate in a hybrid configuration with a conventional activated sludge plant (Plant B and Plant C). These plants are designed to run at relatively high and stable average fluxes (and thus lower SAD_p), since they treat a constant daily flow and are thus not subject to influent flow variations.

Table 5-3: Heuristic aeration and sustainable flux data for 5 large scale HF MBRs (adapted from Judd and Judd, 2010)

	Flow MLD		A_m m^2	Flux LMH		SAD_m $Nm^3.m^{-2}.h^{-1}$		SAD_p -	
	PDF	ADF		PDF	ADF	PDF	ADF	PDF	ADF
Plant A	48		84480	23.7		0.43		18.2	
Plant B	23	23	37920	25.3		0.28	0.14	11.1	5.5
Plant C	42	42	73442	23.8		0.25	0.13	10.6	5.3
Plant D	39	27	68091	23.9	16.5	0.36	0.18	15.1	10.9
Plant E	57	47	91020	26.1	21.5	0.4	0.2	15.3	9.3

*based on 10:30 aeration applied at low-average flows and 10:10 aeration at peak flows.

Figure 5-5 summarises all available data from demonstration/full scale plant, including the data from Mansell *et al.* (2006) and Garcès *et al.* (2007), for 10:10 and 10:30 aeration. The empirical SAD_m values obtained in the empirical study at J_{net} of 23.3 LMH for both 10:10 and 10:30 operation are slightly lower than values reported for full scale plants, possibly due to interference from the biological aeration in the aerobic zone. Moreover, the assumption of a sustainable flux is based on six hours of permeability decline data. Notwithstanding this, it appears to be generally the case that intermittent aeration may be employed, reducing SAD_m, without detriment to flux sustainability. 44% of the data points lie in a range of mean net flux values between 20 and 25 LMH, whilst corresponding the SAD_m values range from 0.11 to 0.43 LMH.bar^{-1}.h^{-1}. It remains unclear, from all these data, as to the nature of the J_{net}:SAD_m relationship at lower aeration rates – i.e. from further increasing aeration intermittency.

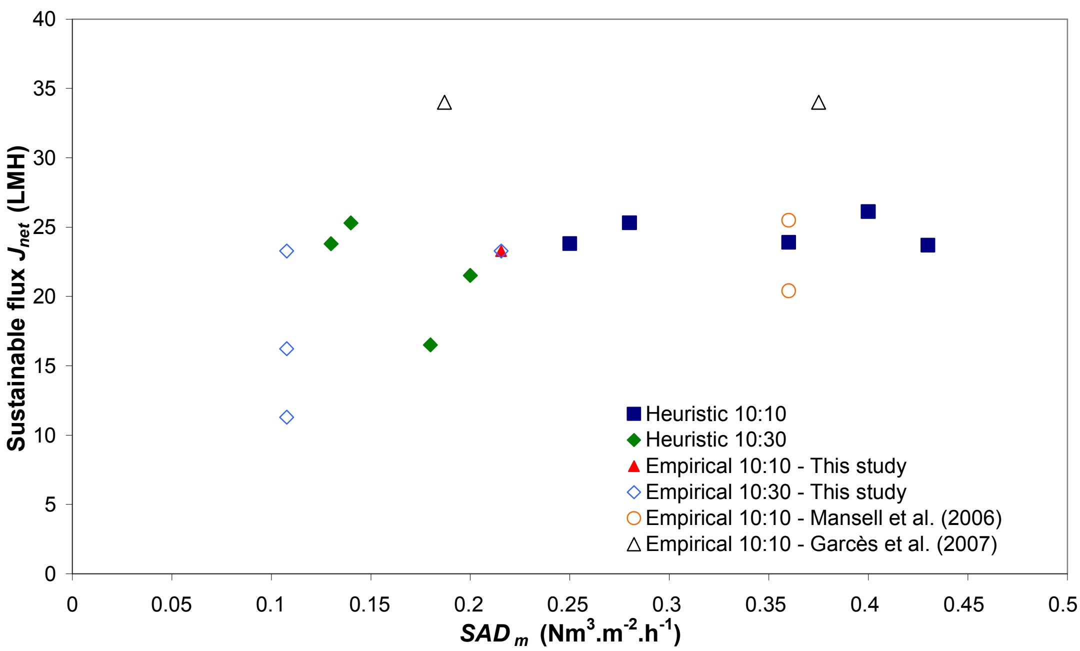

Figure 5-5: Sustainable flux J_{net} vs. SAD_m: comparison of empirical and heuristic data, intermittent aeration.

5.4 CONCLUSIONS

An empirical study into the effects of intermittent aeration on the rate of permeability on a small full-scale HF MBR has revealed:

- The use of 10:10 or 10:30 intermittent aeration results in lower fouling rates when compared to continuous aeration when working under the same overall airflow rate. A lower SAD_p is required for sustainable operation, with associated membrane aeration energy savings of up to 75% when comparing 10:30 to continuous aeration, with no significant impact on fouling rate. A flux of 23.3 LMH under conditions of 10:30 aeration was sustained at a SAD_p as low as 4.6. These findings are corroborated by recent industrial practice.
- Heuristic data across five large scale HF MBRs show that a SAD_p of 10.6-16.7 is required at peak flows using 10:10 aeration and 5.3-10.9 at dry-weather flows using 10:30 aeration. 10:30 aeration can apparently sustain a relatively high flux (up to 25.3 LMH).
- There remains a need to establish the relationship between the attainable net flux and the aeration demand under increasingly intermittent aeration conditions, i.e. for aeration at <25% of the filtration time. Whilst the lower fluxes attainable would increase capital costs, this would almost certainly be offset by the lower operating costs over the plant lifetime.

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CHAPTER 6

MODEL-BASED ENERGY OPTIMISATION OF A SMALL SCALE DECENTRALISED MEMBRANE BIOREACTOR FOR URBAN REUSE

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6 MODEL-BASED ENERGY OPTIMISATION OF A SMALL SCALE DECENTRALISED MEMBRANE BIOREACTOR FOR URBAN REUSE

B. Verrecht[1], T. Maere[2], L. Benedetti[2], I. Nopens[2] and S. Judd[1]

[1]Centre for Water Science, Cranfield University, Vincent Building, Cranfield, Bedfordshire MK43 0AL, UK

[2]BIOMATH, Department of Applied Mathematics, Biometrics and Process Control, Ghent University, Coupure Links 653, B-9000 Gent, Belgium

ABSTRACT

The energy consumption of a small-scale membrane bioreactor, treating high strength domestic wastewater for community level wastewater recycling, has been optimised using a dynamic model of the plant. ASM2d was chosen as biological process model to account for the presence of phosphate accumulating organisms. A tracer test was carried out to determine the hydraulic behaviour of the plant. To realistically simulate the aeration demand, a dedicated aeration model was used incorporating the dependency of the oxygen transfer on the mixed liquor concentration and allowing differentiation between coarse and fine bubble aeration, both typically present in MBRs. A steady-state and dynamic calibration was performed, and the calibrated model was able to predict effluent nutrient concentrations and MLSS concentrations accurately. A scenario analysis (SCA) was carried out using the calibrated model to simulate the effect of varying SRT, recirculation ratio and DO set point on effluent quality, MLSS concentrations and aeration demand. Linking the model output with empirically derived correlations for energy consumption allowed an accurate prediction of the energy consumption. The SCA results showed that decreasing membrane aeration and SRT were most beneficial towards total energy consumption, while increasing the recirculation flow led to improved TN removal but at the same time also deterioration in TP removal. A validation of the model was performed by effectively applying better operational parameters to the plant. This resulted in a reduction in energy consumption by 23% without compromising effluent quality, as was accurately predicted by the model. This modelling approach thus allows the operating envelope to be reliably

identified for meeting criteria based on energy demand and specific water quality determinants.

6.1 INTRODUCTION

6.2 MATERIALS AND METHODS

Membrane bioreactors (MBR) offer a low-footprint option with high quality effluent for recycling municipal wastewater. For applications at small community level, small MBRs are required (Fletcher et al., 2007; Gnirrs *et al.*, 2008, Abegglen *et al*.,2008), which are then inherently less energetically efficient due to wide variations in flows and commensurately large peak loading factors demanding more conservative design. Given that the energy demand contributes significantly to the running costs, it is important to optimise process energy consumption to make the technology more competitive (Judd, 2006).

Mathematical models are widely recognized as providing a useful tool for optimising biological treatment, and several semi-empirical models for the optimisation of MBRs are described in literature (Verrecht *et al*., 2008; Wen *et al*., 1999; Yoon *et al*., 2004). These models, however, have limited predictive power regarding biological performance and total energy demand under dynamic conditions, or else focus mainly on sludge production. The activated sludge models (ASMs) by Henze *et al*. (2000), created with the purpose of describing the biological dynamics of conventional activated sludge (CAS) systems, have been successfully used in the past to optimise full scale CAS plants (Dochain and Vanrolleghem, 2001). However, literature on the application of the activated sludge models to full scale MBR is scarce or not readily accessible (Erftverband, 2001; Erftverband, 2004), and research focuses mainly on sludge production through application of ASM1 and ASM3 to lab and pilot scale MBR (Spérandio and Espinosa, 2008; Lubello *et al*., 2009). The requirement for full scale validation of the ASM models for MBR applications has recently been identified as an urgent research need (Fenu *et al*., 2010). Applying these ASM to MBRs demands that the differences between MBR and CAS systems be recognised, viz.: a) microbiological composition, leading to different stoichiometric and kinetic parameters (inter alia Wen *et al*., 1999; Jiang *et al*., 2005; Lobos *et al*., 2005), b) biomass concentration, leading to changes in oxygen transfer and uptake (Krampe and Krauth, 2003; Germain *et al*, 2007), and c) requirement for additional aeration for membrane scouring (Judd, 2006).

In this paper, the application of ASM2d to a small community-scale MBR for reuse has been appraised with the key objective of optimising energy demand without compromising nutrient removal. The study uses the BIOMATH calibration protocol (Vanrolleghem *et al*., 2003), proceeding through a hydraulic characterisation of the system and employing both steady state and dynamic model calibration to predict water quality. The paper thus provides a case study of the calibration and application of ASM2d to a community-scale MBR. The MBR model incorporates an aeration model accounting for oxygen mass transfer at the operational biomass concentration and differentiates between coarse and fine bubble aeration. Energy consumption values for the different unit processes are derived empirically. A scenario analysis is conducted to link the predicted biological performance for different operational parameters with the empirically derived energy consumption values. The scenario analysis thus allows identification of better operational parameters, and the predicted energy saving and biological removal performance are verified on the full scale plant.

6.2.1 Plant description

The wastewater recycling plant produces an average reclaimed water flow of 25 $m^3 \cdot d^{-1}$ for toilet flushing and irrigation (Figure 5-1). Domestic wastewater from the residences is collected via a pumping station and septic tanks, which provide buffering volume and primary settling. Influent from the septic tanks flows through 3 mm screens to the MBR, which contains both anoxic and aerobic zones for nitrification and denitrification respectively (Table 6-1). Although no anaerobic tank is provided, some of the influent phosphorous is biologically removed, suggesting that part of the anoxic tank may be (intermittently) anaerobic. The membrane separation step is provided by 2 x 3 ZW500c (GE Zenon, Canada) membrane modules with a total membrane surface area of 139 m^2, submerged in the aerobic zone.

6.2.2 Hydraulic profile

A tracer test was carried out using a 22.1 g spike of LiCl dosed into the anoxic zone, with samples taken from the anoxic to aerobic tank overflow weir, the effluent and the sludge recirculation loop every 20 to 30 minutes for the next 40 hours (corresponding to ~1.5 times the hydraulic residence time, HRT). Samples were analysed for Li by atomic emission spectroscopy at 670.784 nm (iCAP 6500 Dual View; Thermo Scientific). To validate the results and determine the number of tanks in series according to the tanks-in-series model (Levenspiel, 1998), the MBR was implemented (Figure 5-1) as an anoxic tank followed by an MBR unit (aerated tank with submerged membrane modules) in the modelling and simulation platform WEST® (MOSTforWATER N.V., Kortrijk, Belgium; Vanhooren *et al.*, 2003). Both tanks were assumed to be completely mixed. During the tracer test, the plant was run under normal flow conditions (Table 6-1).

Table 6-1: Plant dimensions and operational parameters during the tracer test

Parameter	Unit	Value
Anoxic zone		
Volume anoxic zone	m^3	10.09
Aerobic zone / MBR		
Membrane surface	m^2	139.2
Membrane flux during filtration	$l \cdot m^{-2} \cdot h^{-1}$	10.78
Filtration time	s	600
Relaxation time	s	30
Backwash time	s	30
Backwash flux	$l \cdot m^{-2} \cdot h^{-1}$	10.78
Minimum tank volume	m^3	12.21
Maximum tank volume	m^3	12.78
Recirculation flow	$m^3 \cdot d^{-1}$	57.6

6.2.3 Influent characterisation

The MBR was fed with domestic wastewater without rainwater dilution from dwellings with average water consumption of 80-100 $l \cdot capita^{-1} \cdot d^{-1}$. The wastewater strength was thus high (Table 6-2), and comparable to values reported for a single household MBR by Abegglen *et al.* (2008). The septic tanks were estimated to remove 20-30% of the COD, and 0-10% of the N and P (VSA, 2005), as well as buffering the variations in influent concentration to the benefit of biological performance (Gnirss *et al.*, 2008).

Table 6-3 compares the community wastewater characterised according to the STOWA protocol (Roeleveld and van Loosdrecht, 2002) to data for a typical wastewater (Henze *et al.*, 1999), and indicates this wastewater to be 48%, 324% and 81% higher in concentrations of total COD, TKN and TP respectively. The relative quantity of readily biodegradable substrates (S_F and S_A) is also higher due to hydrolysis in the septic tanks (Zaveri and Flora, 2002), which enhances bio-P removal for which the presence of fermentation products such as acetate (S_A) is required (Henze *et al*, 1999; Gernaey and Jørgensen, 2004). Flow variation was between 0 and 1.8 $m^3 \cdot hr^{-1}$, with substantially larger loads (up to 25%) over the weekend (Figure 6-1).

Table 6-2: Average characteristics of influent to the MBR (after septic tanks + screening; samples taken twice per week from January to May 2009)

Variable	Unit	Average	St.Dev.	Variable	Unit	Average	St.Dev
BOD_5	$mg \cdot l^{-1}$	228.17	21.31	TON	$mg \cdot l^{-1}$	0.30	0.00
BOD_f	$mg \cdot l^{-1}$	114.60	14.37	NO_2-N	$mg \cdot l^{-1}$	0.02	0.01
COD	$mg \cdot l^{-1}$	480.50	36.67	PO_4-P	$mg \cdot l^{-1}$	9.29	0.41
COD_f	$mg \cdot l^{-1}$	247.67	48.11	TP	$mg \cdot l^{-1}$	10.87	0.54
TN	$mg \cdot l^{-1}$	81.58	3.51	SS	$mg \cdot l^{-1}$	107.32	9.29
ON	$mg \cdot l^{-1}$	12.21	3.31	pH	-	7.14	0.09
NH_4-N	$mg \cdot l^{-1}$	69.10	5.52				

Table 6-3: Treatment plant wastewater fractionation vs. typical wastewater composition (Henze *et al.*, 1999)

		MBR influent composition in this study (COD=480 $mg \cdot l^{-1}$, TKN=81 $mg \cdot l^{-1}$, TP=11 $mg \cdot l^{-1}$)		**Typical wastewater composition (COD=260 $mg \cdot l^{-1}$,TKN=25 $mg \cdot l^{-1}$, TP=6 $mg \cdot l^{-1}$)**	
Soluble					
Variable	Unit	Value	% of tCOD	Value	% of tCOD
S_F	$mg \cdot l^{-1}$	126.86	26.4%	30	11.5%
S_A	$mg \cdot l^{-1}$	88.89	18.5%	20	7.7%
S_{NH4}	$mg \cdot l^{-1}$	69.10	-	16	-
S_{PO4}	$mg \cdot l^{-1}$	9.29	-	3.6	-
S_I	$mg \cdot l^{-1}$	21.56	4.5%	30	11.5%
Particulate					
Variable	Unit	Value	% of tCOD	Value	% of tCOD
X_I	$mg \cdot l^{-1}$	41.57	8.7%	25	9.6%
X_S	$mg \cdot l^{-1}$	191.26	39.8%	125	48.1%

* Symbols used according to Henze *et al.*, 1999

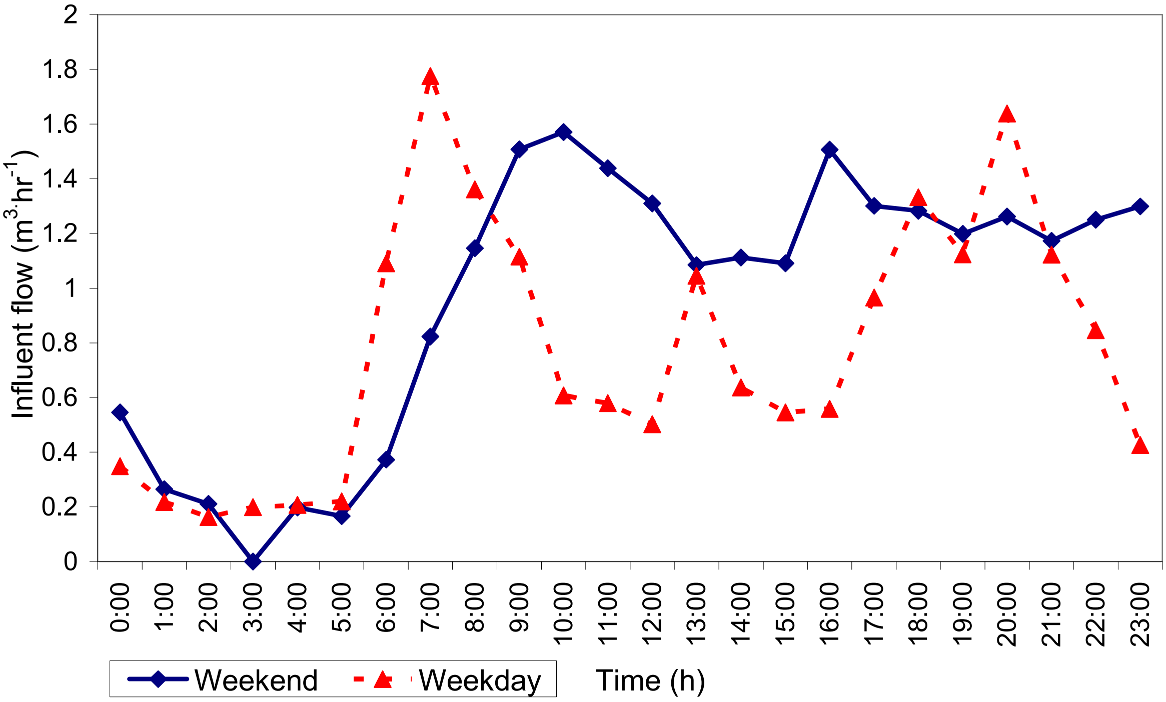

Figure 6-1: Comparison of typical diurnal flow profiles during a weekday and a day in the weekend

6.2.4 Steady state and dynamic plant modelling using ASM2d

For the steady state and dynamic simulations of the plant, ASM2d was chosen as the bio-chemical model since it includes enhanced biological P removal in addition to COD and N removal (Henze *et al.*, 1999). To obtain better representation of P removal, the ASM2d biomass decay rates modifications proposed by Gernaey and Jørgensen (2004) were adopted.

For the model calibration, influent, mixed liquor and effluent data was taken collected from January till May 2009, totalling 93 days, which corresponds to approximately twice the SRT (47 days). A steady-state calibration of the full model was performed based on average data over this period (Table 6-2), and a DO set point of 2 $mg \cdot l^{-1}$ was used, reflecting the average DO value in the aerobic zone. For the dynamic calibration, a high frequency measurement campaign was carried out, and an influent file was produced through analysis of SCADA data, containing a data recorded every 15 minutes for 93 days for the following parameters: influent flow, influent COD, COD_f, BOD_5, TSS, NH_4-N, TKN, PO_4-P, TP, recirculation flow, DO value and wastage flow. During the sampling period, the temperature ranged from 15.8 to 20.7 °C. A number of process upsets occurred over this period, such as a mixer failure, resulting in a necessary manual increase in the recirculation flow to keep the anoxic zone mixed and a blower failure resulting in low DO levels for a number of days. These upsets were included in the model.

Since the model predictions were used for energy consumption calculations, the use of an adequate aeration model was of utmost importance. Basic aeration models, such as the one used in Benchmark Simulation Model 1 (BSM1, Copp, 2002) and many ASM model applications do not account for the detrimental effect of elevated MLSS concentrations on oxygen transfer, and control the oxygen transfer rate by controlling the oxygen transfer coefficient k_La:

$$SOTR = k_L a \cdot \left(DO - DO_{sat}\right) \cdot V \qquad \textbf{(6.1)}$$

To account for the effect of elevated MLSS concentrations on oxygen transfer and for other dependencies of oxygen transfer, e.g., the difference in oxygen transfer from coarse and fine bubble aeration, typical for MBR, a more extensive model as described in Maere *et al.* (2009) was used (Metcalf and Eddy, 2003; Henze *et al.*, 2008; Verrecht *et al.*, 2008, Krampe and Krauth, 2003; Germain et al, 2007, Stenstrom and Rosso, 2008), *viz*:

$$AOTR = SOTR \cdot \frac{(\beta \cdot C_{rsat\_average} - C_{\tan k})}{C_{ssat}} \cdot \varphi^{(T-20)} \cdot \alpha \cdot F \qquad \textbf{(6.2)}$$

$$SOTR = 24 \cdot Q_{air} \cdot \rho_{air} \cdot OTE \cdot y \cdot O_{air} / 10000 \qquad \textbf{(6.3)}$$

$$\alpha = e^{-\omega \cdot MLSS} \qquad \textbf{(6.4)}$$

In this model the influence of MLSS concentration on the AOTR is accounted for through the α-factor (Eq. 6.4), and the effect of using different types of diffusers for biological and membrane aeration can be incorporated by calculating the SOTR for each type of diffuser individually, with appropriate values of oxygen transfer efficiency (*OTE*) and fouling factor *F*. More details about the aeration model can be found in Maere *et al.* (2009).

6.2.5 Scenario analysis

A scenario analysis (SCA) was carried out to determine the optimum operating conditions by varying the experimentally-adjustable degrees of freedom (DOF) that were regarded as having the greatest impact on effluent quality and energy consumption:

- SRT: 9 values for the wastage rate, equally spaced between 0.1 to 2.278 $m^3 \cdot d^{-1}$ yielding SRT values ranging from 10 to 228.7 days
- Recirculation rate: 9 values, equally spaced between 28.8 $m^3 \cdot d^{-1}$ to 187.2 $m^3 \cdot d^{-1}$ (upper range of recirculation pump) yielding recirculation ratios to the influent flow of 1.13 to 7.78
- Dissolved oxygen set point: 5 values, equally spaced between 0.75 and 2 $mg \cdot l^{-1}$

For inputting to the SCA, a data set containing 35 days of influent was taken from the plant when operating under normal influent conditions. The scenario analysis was duplicated using two different membrane aeration rate values (84 and 42 $Nm^3 \cdot h^{-1}$), corresponding to the maximum and minimum realistic values for coarse bubble air flow ($Q_{air,coarse}$), since this parameter accounts for a large part of the total energy consumption.

The SCA grid, using the values described above, resulted in 486 different scenarios. The impact on activated sludge aeration, nutrient removal and MLSS concentration was studied. To calculate the energy consumption for each degree of freedom, empirical correlations for energy consumption of the unit processes (membrane aeration, biology aeration, recirculation pumping, permeate pumping and mixing) were derived from measurements on the plant, at an MLSS of 8,000 $mg \cdot l^{-1}$. Membrane aeration energy was 0.029 to 0.034 $kWh \cdot Nm^{-3}$ for $Q_{air,coarse}$ of 84 and 42 $Nm^3 \cdot h^{-1}$ respectively, indicating that the blower becomes less efficient at lower air flow rates. Energy demand for the recirculation pump varied linearly with the flow rate, up to a maximum of 0.037 $kWh \cdot m^{-3}$ of sludge pumped. Since the activated sludge blower for biology aeration is controlled by an on/off controller at around 2 $mgO_2 \cdot l^{-1}$ (or around the different DO set points, as described above) and runs at fixed speed when in operation, the energy consumption per Nm^3 is constant at 0.0289 $kWh \cdot Nm^{-3}$. For the scenario analysis, the mixing energy was considered constant at 4.6 $kWh \cdot d^{-1}$. Since mixing accounts for less than 5% of the total energy demand, changes in mixing energy arising from changes in viscosity at different MLSS concentrations were considered negligible. The permeate pump was constantly running at 1.8 $m^3 \cdot h^{-1}$ when in operation, resulting in an energy consumption of 0.056 $kWh \cdot m^{-3}$ of permeate. Sludge handling costs were ignored since these depend on site-specific sludge management strategies.

6.3 RESULTS AND DISCUSSION

6.3.1 Hydraulic profile

Figure 6-2 displays measured versus predicted Li concentrations in the anoxic and aerobic zone during the tracer test. The correlation between the measured and predicted data for both the anoxic and aerobic zone corroborates the assumption of perfect mixing. The recovery of Li, defined as the ratio of Li added to Li recovered in the effluent, determined through integration of the effluent Li flux, was 87%, and would have been higher for an extended campaign. The measured Li concentrations in the effluent were always about 7.5 ± 3.5% lower than the Li concentrations measured in the recirculation sludge, suggesting some Li adsorption onto the flocs arose but not to a significant extent. The tanks could thus each be considered CSTRs for the remainder of the modelling exercise.

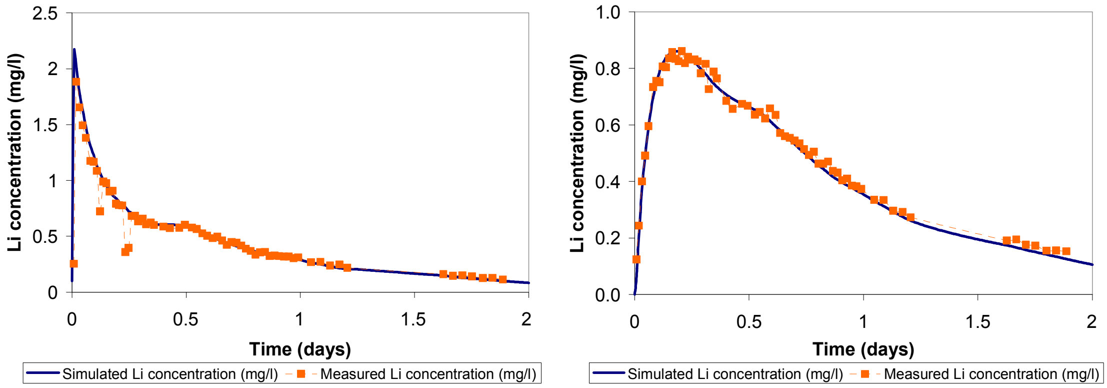

Figure 6-2: Predicted and actual Li concentrations in (a) anoxic, and (b) aerobic tanks during the tracer test

6.3.2 Model calibration

6.3.2.1 Steady state calibration

A steady state calibration was performed to achieve an accurate simulation of the MLSS concentration, this being instrumental in correctly predicting the aeration energy demand due to its effect on oxygen transfer (via the α-factor). However, as shown in Table 6-4, using default values as reported by Henze *et al.* (1999) for all stoichiometric and biokinetic parameters, leads to an underestimation of sludge production (MLSS concentration) by about 15%, as the growth of X_{PAO} (and consequently bio-P removal) could not be simulated correctly in steady state. This can be attributed to the fact that anaerobic conditions, required for the growth of X_{PAO}, do not occur during steady state simulation, indicating the need for a dynamic calibration taking into account the influent variations. In steady state, a correct representation of MLSS concentrations could only be achieved by making substantial and unrealistic changes to μ_{PAO} (2 d^{-1} vs. default value of 1 d^{-1}), b_{PAO} (0.1 d^{-1} vs. default value of 0.2 d^{-1}) and Y_{PO} (0.2 $gP \cdot (g\ COD)^{-1}$ vs. default value of 0.4 $gP \cdot (g\ COD)^{-1}$ (Table 6-4).

Table 6-4: Steady state simulation results compared with average measured values

Parameter	Units	Measured Values	- Default ASM2d values (Henze *et al.*, 1999) - Bio-P module (Gernaey and Jørgensen, 2004)	- Default ASM2d values (Henze *et al.*, 1999) - Bio-P module (Gernaey and Jørgensen, 2004) - μ_{PAO} = 2 d^{-1} - b_{PAO} = 0.1 d^{-1} - Y_{PO} = 0.2 $gP \cdot (g\ COD)^{-1}$
NH_4-N	$g \cdot m^{-3}$	0.07	0.337	0.338
NO_3-N	$g \cdot m^{-3}$	21.4	21.9	21.68
PO_4-P	$g \cdot m^{-3}$	4.35	9.65	5.18
MLSS	$g \cdot m^{-3}$	7,832	6,584	7,869

6.3.2.2 Dynamic calibration

When the dynamic influent file was applied to the model, the concentration of X_{PAO} started to increase without the adjustments to μ_{PAO}, b_{PAO} and Y_{PO} that were necessary in the steady state calibration. Upon reaching dynamic equilibrium, MLSS concentrations were represented accurately using the default parameter values as reported by Henze *et al.* (1999), thereby eliminating the need to adjust μ_{PAO}, b_{PAO} and Y_{PO} as was necessary under steady-state conditions.

To calibrate the aeration model, the measured $Q_{air,fine}$ (averaged over a 15 minute period) was used as the input for the aeration model, while $Q_{air,coarse}$ was fixed at 84 $Nm^3 \cdot h^{-1}$, to mimic the prevailing operational conditions during the calibration period. The values for OTE_{fine} (0.045 m^{-1}), OTE_{coarse} (0.015 m^{-1}) were taken from Metcalf and Eddy (2003), the value for ω (0.084) was the mean value derived from the data of Germain *et al*. (2007), Krampe and Krauth (2003), and Metcalf and Eddy (2003). F_{coarse} (0.8) and F_{fine} (0.8) were calibrated to closely match the measured DO profile. Calibrating the fouling factors could be justified since an inspection of the diffusers had shown visible fouling. Moreover, more advanced techniques for measuring the *α-factor* and OTE_{coarse} and OTE_{fine} were unavailable.

Despite the plant upsets during the evaluated period, the used parameter set allowed for a satisfactory reproduction of the NH_4-N, NO_3-N and MLSS concentration trajectories; Figure 6-3 compares the simulated nitrogen removal profiles (NH_4-N and NO_3-N) and MLSS concentrations with concentrations measured during an intensive sampling period on Days 61-62 of the 93 day campaign. Predicted NH_4-N concentrations were consistently slightly higher than the measured values (~0.25 $mg \cdot l^{-1}$ simulated vs ~0.04 $mg \cdot l^{-1}$ measured and confirmed by using two different analytical techniques). MBRs tend to achieve more stable and complete nitrification than CAS systems (Munz *et al.*, 2008), a fact that is apparently not well incorporated into the various CAS ASM models. Despite this shortcoming, when looking at the total nitrogen removal, the prediction is still very accurate (Figure 6-3). Predicted PO_4-P concentrations show acceptable values and dynamic behaviour (Figure 6-4) though consistently a few hours ahead of those measured. It is postulated that this is caused by the oversimplification of the actual hydraulics by the tanks-in-series concept, which may be unable to accurately predict the occurrence of localised anaerobic zones under certain conditions. However, a CFD model study and on-line data at different locations in the tank would be required to confirm this. In general it can be concluded that the calibrated model predicts nutrient and MLSS concentrations accurately using the default values for ASM2d (Henze *et al.*, 1999) and its modification (Gernaey and Jørgensen, 2004), and the model used along with the energy demand calculations in the subsequent scenario analysis.

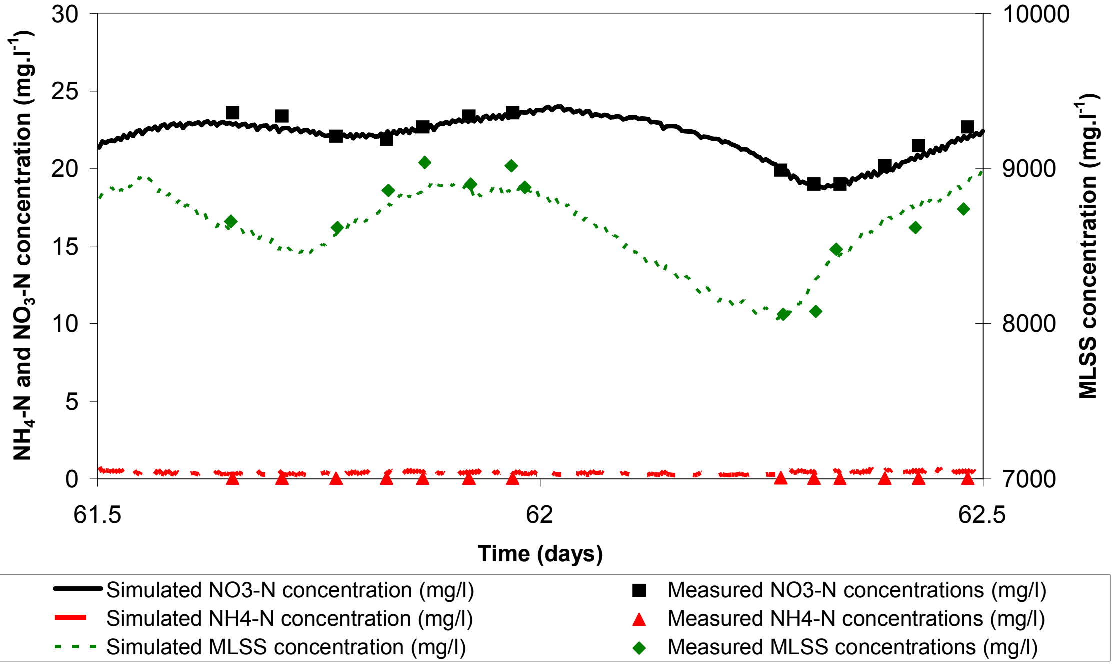

Figure 6-3: Simulated and recorded NH_4-N, NO_3-N and MLSS concentrations using measured $Q_{air,fine}$, averaged per 15 minute interval, as input

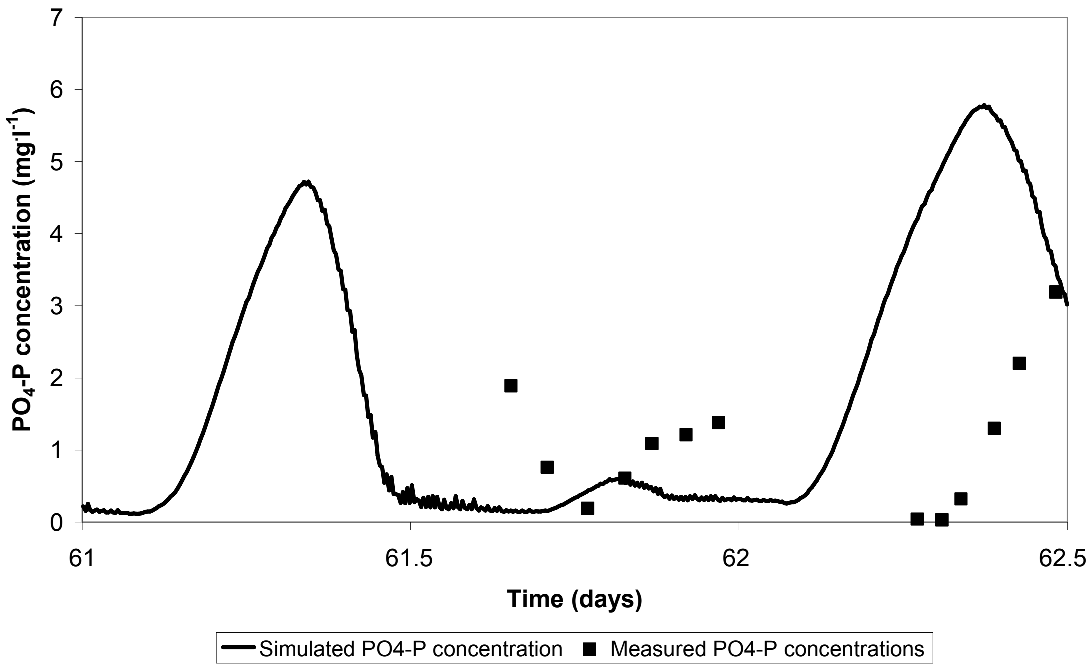

Figure 6-4: Simulated and measured PO_4-P values using measured measured Qair,fine, averaged per 15 minute interval, as input

6.3.3 On/off controller for aeration model to predict $Q_{air,fine}$ values

To lend predictive value to the model, the fine bubble aeration flow rate ($Q_{air,fine}$) demanded an extra on/off controller, switching on aeration at DO < 1.5 $mg \cdot l^{-1}$ with $Q_{air,fine}$ at 90 $Nm^3 \cdot h^{-1}$ and switching off at DO > 2.5 $mg \cdot l^{-1}$, simulating the actual blower operation at the plant. The integral of the predicted $Q_{air,fine}$ value was within 3% difference from the actual measured value when using the parameters as calibrated in Section 6.3.2.2, indicating that aeration demand could be predicted accurately through this approach. Nutrient and MLSS concentrations were reproduced well, with predicted values generally well within 10% of the measured ones (Figure 6-3 and Figure 6-4). Any differences can be attributed to slight deviations from reality using the implemented on/off controller.

6.3.4 Scenario analysis

The evolution of biological aeration demand and maximum effluent NH_4-N concentration as a function of the SRT (Figure 6-5) demonstrates that lowering the SRT by increasing the wastage rate has a beneficial effect on demand for $Q_{air,fine}$. However, Figure 6-5 also shows that this also leads to higher maximum effluent NH_4-N concentrations, indicating a trade-off between minimising the aeration demand (and thus energy consumption) and achievable effluent quality. Operation at SRTs below 23 days leads to a deterioration in nitrification, because of a decrease in MLSS and autotrophs concentration, and to an increase in F/M ratio, similar to observations by Cicek *et al.* (2001). Lowering the DO setpoint had a similar but less pronounced effect on nitrification.

There is a similar phenomenon regarding phosphate and nitrate (Figure 6-6), in that an increase in the recirculation ratio leads to respectively lower and higher effluent NO_3-N and PO_4-P concentrations. This arises because the denitrification and bio-P removal processes compete for the same carbon source (Metcalf and Eddy, 2003) and anaerobic conditions is less sustainable at higher recirculation ratios.

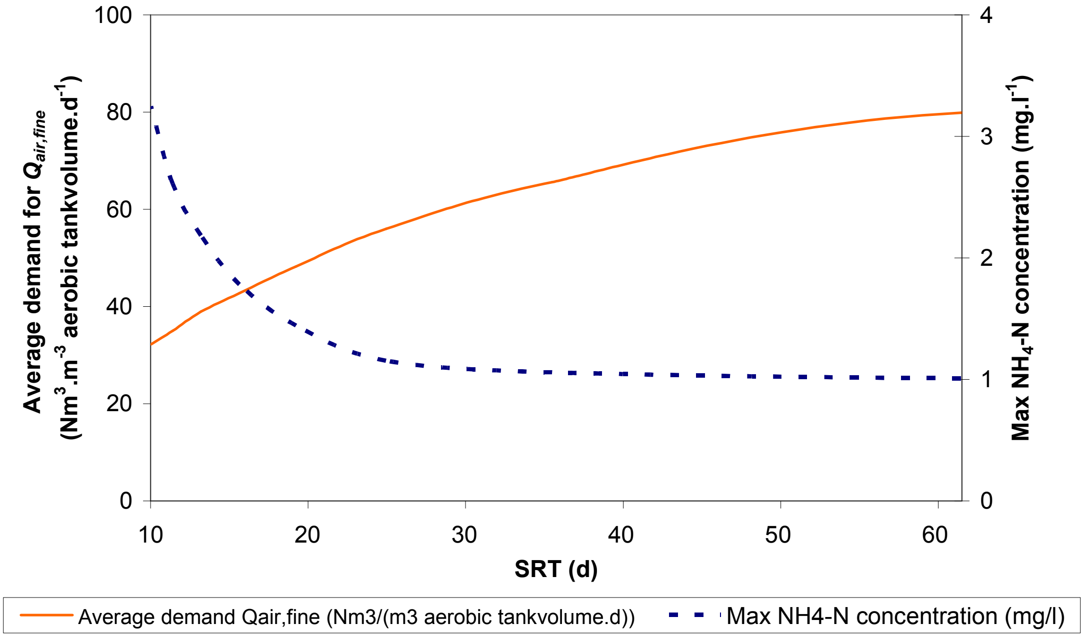

Figure 6-5: Influence of wastage rate on the total demand for biology aeration ($Q_{air,fine}$) and the maximum occurring effluent NH_4-N concentration during the 35-day simulation ($Q_{air,coarse}$ = 42 $Nm^3.h^{-1}$; DO setpoint = 1.25 $mg.l^{-1}$; recirculation flow = 108 $m^3.d^{-1}$)

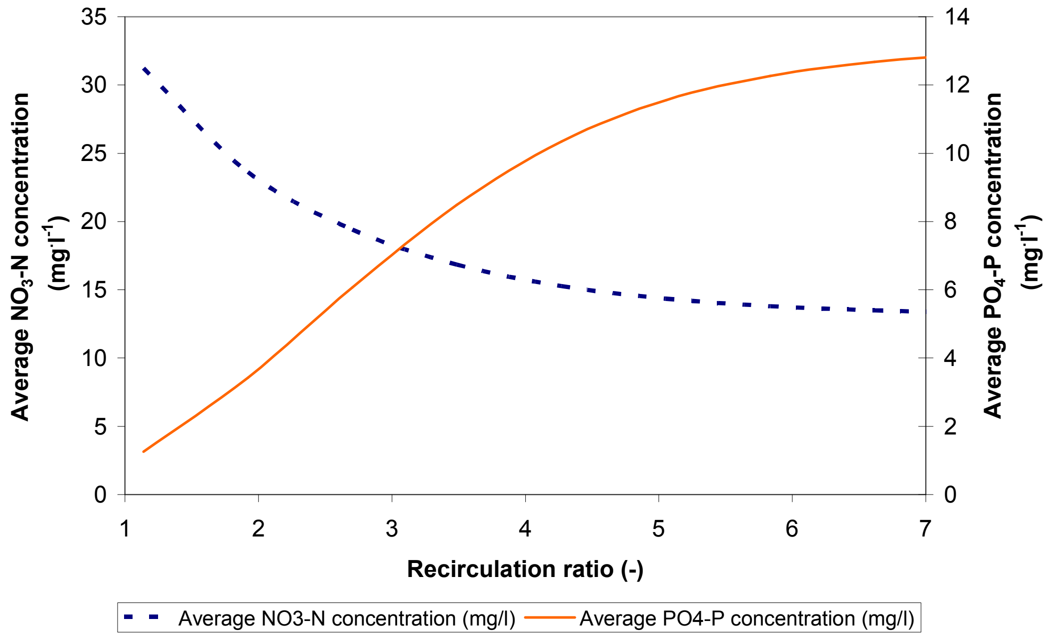

Figure 6-6; Influence of recirculation flow rate on the average effluent NO_3-N and PO_4-P concentrations during the 35-day simulation ($Q_{air,coarse}$ = 42 $Nm^3.h^{-1}$; DO setpoint = 1.25 $mg.l^{-1}$)

Figure 6-7 shows that a change in the SRT (through variation in wastage rate), and thus MLSS concentration, has a much larger impact on total aeration energy demand than changing the recirculation ratio. At a DO setpoint of 1.25 mg·l^{-1} and fixed recirculation ratio, the total fine bubble aeration demand ($Q_{air,fine}$) can change by up to 342% depending on the wastage rate, while this change is limited to 44% when varying the recirculation ratio at fixed SRT and DO set point. This confirms the importance of incorporating the MLSS dependency of the oxygen transfer into the aeration model. The model thus allows the operating envelope to be identified for meeting criteria based on energy demand and/or specific water quality determinants.

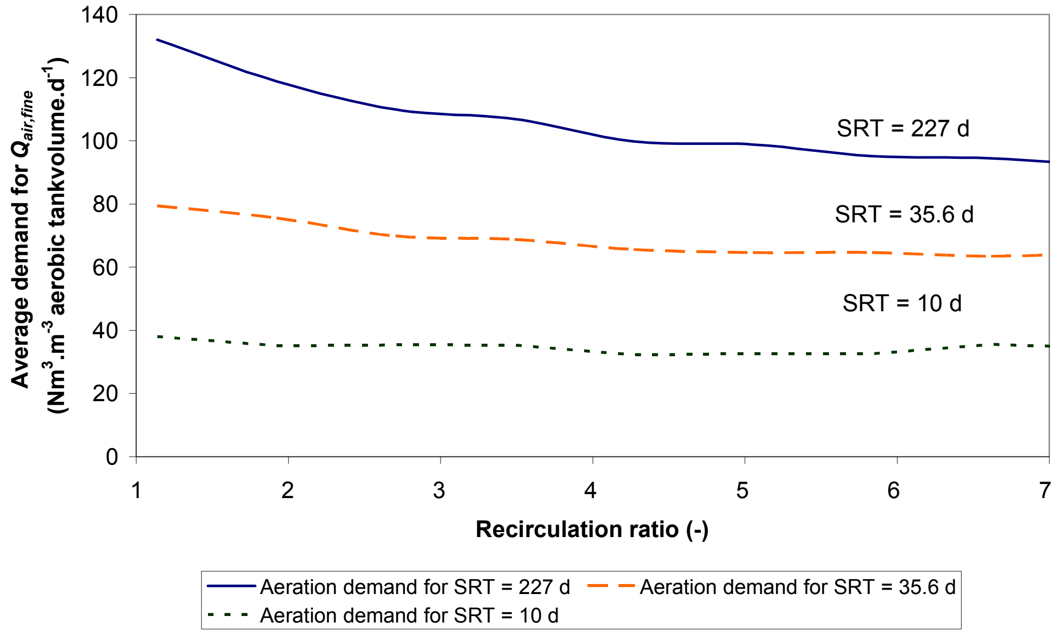

Figure 6-7: Influence of recirculation rate and SRT on total demand for $Q_{air,fine}$ ($Q_{air,coarse}$ = 42 Nm^3.h^{-1}; DO setpoint = 1.25 mg.l^{-1})

It is acknowledged that over the range of operating conditions studied in the SCA, the biological processes and kinetics may change. For instance, Spérandio and Espinosa (2008) suggest that at elevated SRT some of the influent X_I should be considered as X_S, which has implications on the overall sludge balance. Also, simultaneous nitrification and denitrification may occur at low DO set points. The model accounts for this by using oxygen half-saturation coefficients K_O for X_H and X_{PAO}. The effect of floc size on the value of K_O is still debated, the small flocs of an MBR compared to those from CAS would be expected to yield lower values for the halfsaturation constants (*inter alia* Manser *et al.*, 2005). However, no clear consensus has been reached on the impact of specific operational conditions on kinetics. Hence, rather than varying the biokinetic parameters in the model over the studied range of operational parameters, all biokinetic parameters were assumed constant, and an *a posteriori* model validation carried out by confronting the obtained model predictions of the scenario analysis with experimental data independent of the calibration data set (Section 6.3.5).

The outcomes of the scenario analysis were linked with the empirical energy consumption calculations, and ranked in terms of energy consumption while compliant with effluent quality standards of <0.5 $mg \cdot l^{-1}$ NH_4-N, <20 $mg \cdot l^{-1}$ TN < 20 $mg \cdot l^{-1}$, and 5,000 $mg \cdot l^{-1}$ minimum MLSS. Since reuse regulations - such as the US EPA guidelines for unrestricted urban reuse (EPA, 2004) - generally do not include stringent NH_4-N or TN guidelines, these parameters were chosen to achieve consistent effluent quality under conservative operating conditions that could be achieved in a real system.

When comparing the different parameter sets for the two studied air flow rates displayed, the average energy consumption was 13.1 ± 4.7% lower at a membrane coarse bubble aeration of 42 $Nm^3 \cdot h^{-1}$ compared to 84 $Nm^3 \cdot h^{-1}$. The maximum difference in energy consumption between simulations for the different membrane airflow values was 28%, while the minimum was 4.6%. When the membrane aeration airflow rate was set at 42 $Nm^3 \cdot h^{-1}$, the minimum and maximum predicted energy consumption was 2.25 $kWh \cdot m^{-3}$ and 3.83 $kWh \cdot m^{-3}$ respectively. These values increased to 2.74 $kWh \cdot m^{-3}$ and 4.46 $kWh \cdot m^{-3}$ when the membrane aeration was kept at its original value of 84 $Nm^3 \cdot h^{-1}$.

6.3.5 Model application for optimisation

Results from the scenario analysis were used in the selection of better operational parameter values (Table 6-5). The higher wastage rate (and lower SRT) resulted in a modest decline in MLSS and higher F/M ratio, which previous studies have indicated may increase the sludge fouling propensity (Trussell *et al.*, 2006). However, data collected on the real MBR over a period corresponding to approximately twice the SRT indicated permeability to be maintained at the levels achieved in the original without changing the cleaning regime, notwithstanding the reduction in membrane aeration rate. This is attributable to the low operational fluxes (10-13 $l \cdot m^{-2} \cdot h^{-1}$), well below the operating flux values for most large-scale MBRs. However, the lower membrane aeration set point corresponded to a SAD_m of 0.3 $Nm^3 \cdot m^{-2} \cdot h^{-1}$, which is still within the range of SAD_m values (0.2 – 1.28 $Nm^3 \cdot m^{-2} \cdot h^{-1}$) typically considered sufficient for sustainable operation, even at higher fluxes (Judd, 2006). Changing the parameter values did not compromise the effluent quality in terms of COD and N removal based on data obtained through twice weekly grab sampling, but biological P removal deteriorated due to the increased recirculation ratio, as predicted by the model.

Table 6-5 also displays the resulting energy saving compared to the original values. A substantial reduction in energy consumption per m^3 of permeate produced was achieved (23%), and this value was predicted by the model within 5-10%. The energy consumption value of 3.11 $kWh \cdot m^{-3}$ is at the lower end of values typically reported for small MBRs (Boehler *et al.*, 2007; Gnirss *et al.*, 2008), which can range from 3 to 12 $kWh \cdot m^{-3}$ depending on the design and circumstances. However, this value is high when compared to larger, more efficient plants, which can be as low as 0.62 $kWh \cdot m^{-3}$ for standard intermittent aeration (Garcés *et al.*, 2007). Other reported values for large-scale MBRs range from 0.6 to 2.0 $kWh \cdot m^{-3}$, depending on operational parameters and flow conditions (Brepols *et al.*, 2009).

Table 6-5: Changes in operational parameter values according to the conclusions from the scenario analysis, and comparison in energy consumption between original and optimised system (energy demand of membrane aeration, activated sludge aeration, mixing of anoxic

	Unit	Original	New
Operational parameters			
Membrane aeration	$Nm^3 \cdot hr^{-1}$	84	42
Wastage rate	$m^3 \cdot d^{-1}$	0.485	0.645
i.e. SRT	d	47	35
DO set-point	$mg \cdot l^{-1}$	2	1.25
Recirculation flow	$m^3 \cdot d^{-1}$	57.6	108
i.e. Recirculation ratio	-	2.27	4.25
Energy consumption			
Measurement	$kWh \cdot m^{-3}$	4.03	3.11
Reduction	%		23%
Model prediction	$kWh \cdot m^{-3}$	4.25	2.99
Deviation from real value	%	5.1	3.9

The proposed modelling approach can be readily applied to other MBRs, even when operating under more stringent conditions, which is likely for larger scale plants, since it is widely reported that MBRs achieve good and consistent nutrient removal at lower HRT (*inter alia* Judd, 2006). However, operation at high HRTs is not uncommon for smaller plants, as indicated in Gnirss *et al*. (2008), and the findings of this paper may thus provide useful information for future design and operation of small scale installations. The extent of the reduction in energy consumption that can be achieved by applying the proposed methodology will depend on the influent wastewater composition, desired effluent quality, allowable MLSS range and SRT, and initial operating conditions.

6.4 CONCLUSIONS

- A small MBR for domestic water recycling, running under unusual and challenging influent conditions, was dynamically modelled in ASM2d. The model provided an accurate prediction of the dynamic nutrient removal profile and MLSS concentrations using default ASM2d values for all biokinetic and stoichiometric parameters.
- A dedicated aeration model was used, incorporating the effect of elevated MLSS concentrations on oxygen transfer, and allowing differentiation between coarse and fine bubble aeration such that aeration demand could be accurately simulated.
- To allow realistic modelling of the plant, influent fractionation was carried out based on average influent concentrations obtained over a four-month sampling period. Analysis has shown the wastewater strength to be considerably higher than for a typical wastewater of entirely domestic origin with no dilution or infiltration. The amount of readily biodegradable substrate (45%) was also higher than typically reported values (20%) due to hydrolysis in the septic tank.
- A scenario analysis was conducted to simulate the effect of varying the SRT, the recirculation ratio and the DO set point on effluent quality, MLSS concentrations and aeration demand. Linking the outcomes with empirically-derived calculations for energy consumption allowed quantification and optimisation of the energy demand. Decreasing the membrane aeration flow and SRT had the most profound effect on total operational energy consumption, but there was a trade-off in achievable NH_4-N removal due to diminished nitrification with decreasing SRT. Increasing the recirculation flow led to improved TN removal and to deterioration in TP removal. This modelling approach thus allows the operating envelope to be identified for meeting criteria based on energy demand and/or specific water quality determinants - and nutrients in particular.

- New operational parameter values were applied to the plant, resulting in an on-site reduction in energy consumption by 23%, without compromising effluent quality, as predicted by the model.

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CHAPTER 7

THE COST OF A LARGE-SCALE HOLLOW FIBRE MBR

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7 THE COST OF A LARGE SCALE HOLLOW FIBRE MBR

B. Verrecht[1], T. Maere[2], C. Brepols[3], I. Nopens[2] and S. Judd[1]

[1]Centre for Water Science, Cranfield University, Vincent Building, Cranfield, Bedfordshire MK43 0AL, UK

[2]BIOMATH, Department of Applied Mathematics, Biometrics and Process Control, Ghent University, Coupure Links 653, B-9000 Gent, Belgium

[3]Erftverband, Am Erftverband 6, 50126 Bergheim/Erft, Germany

ABSTRACT

A cost sensitivity analysis was carried out for a full scale hollow fibre membrane bioreactor to quantify the effect of design choices and operational parameters on cost. Different options were subjected to a long-term dynamic influent profile and evaluated using ASM1 for effluent quality, aeration requirements and sludge production. The results were used to calculate a net present value (NPV), incorporating both capital expenditure (capex), based on costs obtained from equipment manufacturers and full scale plants, and operating expenditure (opex), accounting for energy demand, sludge production and chemical cleaning costs.

Results show that the amount of contingency built in to cope with changes in feedwater flow has a large impact on NPV. Deviation from a constant daily flow increases NPV as mean plant utilisation decreases. Conversely, adding a buffer tank reduces NPV, since less membrane surface is required when average plant utilisation increases. Membrane cost and lifetime is decisive in determining NPV: an increased membrane replacement interval from 5 to 10 years reduces NPV by 19%. Operation at higher SRT increases the NPV, since the reduced costs for sludge treatment are offset by correspondingly higher aeration costs at higher MLSS levels, though the analysis is very sensitive to sludge treatment costs. A higher sustainable flux demands greater membrane aeration, but the subsequent opex increase is offset by the reduced membrane area and the corresponding lower capex.

7.1 INTRODUCTION

Over the last two decades, implementation of membrane bioreactors (MBRs) has increased due to their superior effluent quality and low plant footprint (Judd, 2008). However, they are still viewed as a high-cost option, both with regards to capital and operating expenditure (capex and opex), mainly due to membrane installation and replacement costs and higher energy demand compared to conventional activated sludge systems. However, quantification of such impacts is constrained by availability of credible data.

An overview of literature investment cost data (McAdam and Judd, 2006, Figure 7-1) over a range of reported plant sizes reveals costs to increase exponentially with decreasing plant size, and that a large variation in required capex arises according to assumptions made and costs included. DeCarolis *et al.* (2004) provided a comprehensive overview of costing data in terms of capex and opex, both for the MBR system alone (based on quotes from four leading suppliers), and for the complete installation (based on preliminary plant design and assumptions about the location-specific contribution of land costs, contractor overheads, engineering, legal costs, etc). Côté *et al.* (2004) compared capex and opex of an MBR to a conventional activated sludge (CAS) system with tertiary filtration for effluent reuse purposes, demonstrating an integrated MBR to be less expensive than a combination of CAS and tertiary filtration - a conclusion subsequently corroborated by Brepols *et al.* (2009) for German wastewater plants. The latter authors showed energy demand to increase for plants with significant in-built contingency, since the average plant utilisation is low. This has recently led Maurer (2009) to introduce the specific net value (SNPV), which takes into account the average plant utilisation over its lifetime and so reflects the cost per service unit.

Notwithstanding the above, no in-depth analysis has been produced quantifying the impact of key design and operating parameters on both capex and opex over the lifetime of an installation. This paper aims to determine both absolute values of capex and opex and their sensitivity to various influencing parameters such as contingency (to provide robustness to changes in feedwater flow and composition), membrane replacement, net flux, and hydraulic and solids residence time (HRT and SRT). The approach taken is to evaluate the impact of representative dynamic flow and load conditions using ASM1 (Henze *et al.*, 2000) on effluent quality, sludge production and aeration demand, based on various MBR process designs. Dynamic simulation results can then be used as input for specific cost models for both capex and opex, generated using representative heuristic and empirical available cost data. Opex for energy demand (Maere *et al.*, 2009), added to sludge treatment and disposal and chemical cleaning costs, can then be combined with capex to produce the NPV. This then allows the impact of design and operation parameter selection to be quantified.

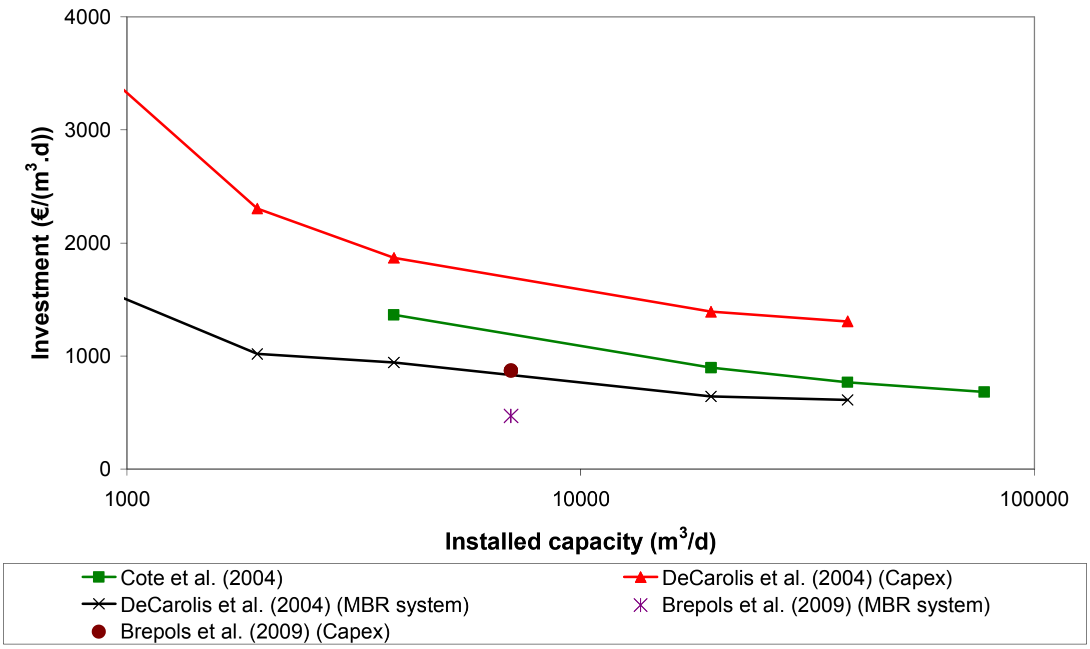

Figure 7-1: Specific investment vs. installed plant capacity, based on literature data (adapted from McAdam and Judd, 2006)

7.2 MATERIALS AND METHODS

7.2.1 Long term influent

The 87 week long BSM1 LT dynamic influent file (Gernaey *et al.*, 2006) was used to evaluate the different plant designs. It includes all phenomena typically observed in a year of full-scale WWTP influent data. Average influent flow (Q_{in}) was 20,851 $m^3.d^{-1}$, while the maximum instantaneous flow was 59,580 $m^3.d^{-1}$. The first 35 weeks of influent data were used to initialise the models; the remaining influent data covering a period of one year (52 weeks) were used for evaluation.

7.2.2 Biological process model

Figure 7-2 depicts the generic nitrifying-denitrifying plant upon which all further design options were based. The ASM1 biokinetic model was selected to study the impact of design and operational parameters on biological performance. Since no consensus exists on updating biokinetic values for an MBR, the default ASM1 biokinetic parameter values, as reported in Henze *et al.* (2000), were used throughout. Simulations were performed using the WEST® simulation and modelling platform (Vanhooren *et al.*, 2003).

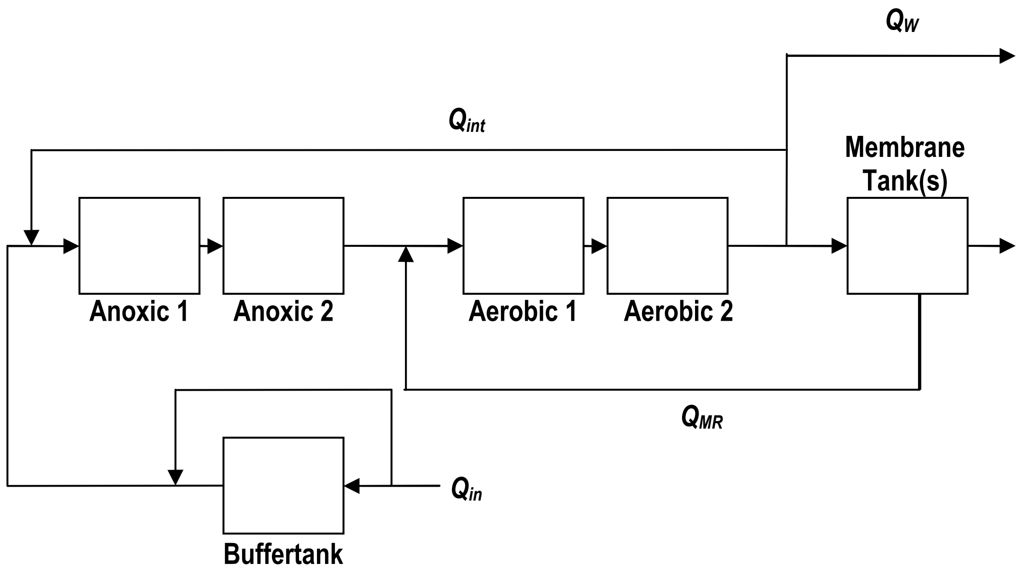

Figure 7-2: Schematic overview of the generic nitrifying/denitrifying MBR design

Biological tank volumes were determined by a required minimum HRT at average influent flow conditions of 8 hours, or a minimum HRT at maximum flow conditions of 4 hours, whichever was the largest, and the default SRT value was 25 days. These design conditions are within reported trends for large MBR in Europe (Itokawa *et al.*, 2008). The anoxic fraction represented 40% of tank volumes. Sludge recirculation was carried out from the membrane tank to the aerobic tanks was taken as four times the average feed flow: Q_{MR} = 83,404 $m^3.d^{-1}$. Internal recirculation from the aerobic tanks to the anoxic tanks was three times the feed flow Q_{int} = 62,553 $m^3.d^{-1}$.

The membrane tank volume, included in the total aerobic volume, was calculated based on a conservative packing density of 45 m^2 membrane area per m^3 of tank volume, which is at the lower end of values reported (Judd and Judd, 2010). The number of membrane tanks required was based on the design parameters for a large scale plant (Brepols *et al.*, 2008), one membrane tank required per 10,000 m^2 membrane area, allowing sufficient flexibility in operation and cleaning.

The required buffer tank volume was dictated by:

- an assumed maximum buffer tank HRT of 2 days - based on the maximum flow from the buffer tank equating to the difference between the conservative net flux and the maximum sustainable flux, corresponding to 40% of plant design flow;
- the combination of plant and buffer tank required to cope with storm flows without bypass.

Taking these constraints into account, the maximum size of the buffer tank was equal to 80% of the daily design plant flow.

7.2.3 Capital costs

To evaluate capital investment costs, pricing information (Table 7-1) was obtained from manufacturers or based on costs provided by end-users for similar items of equipment at full scale MBR plants (Brepols, 2010). Assumptions made were as follows:

Membranes A net design flux of 20 $l.m^{-2}.h^{-1}$ (LMH) was used for calculating membrane area, while the maximum sustainable flux was assumed to be 40% higher, i.e. 28 LMH, which can be considered conservative based on literature values (Judd and Judd, 2010; Garcés *et al.*, 2007). A regime of 10 min filtration followed by 30 s backwashing resulted in an instantaneous flux of 22.1 LMH and maximum instantaneous flux of 30.9 LMH. HF membrane costs were assumed to be €50.m^{-2} (Brepols *et al.*, 2010).

Tanks Tank building costs were based on costs of €220.m^{-3} tank volume (Brepols *et al.*, 2010b).

Plant equipment A 6mm coarse screening step followed by a 0.75 mm fine screen was chosen as a representative pre-treatment for HF membranes (De Wilde *et al.*, 2007a). Screens were sized to treat the maximum instantaneous flow to the plant, with 50% redundancy, ensuring that the whole flow could be treated by 2 sets of fine and coarse screens with one set on standby.

To size the membrane blowers, SAD_m was assumed constant at 0.3 $Nm^3.m^{-2}.h^{-1}$. The number of blowers for membrane aeration installed was based on the number of membrane tanks, with one standby blower. The biology blowers were sized based on the maximum aeration demand to maintain DO at 2 $mg.l^{-1}$ over the final 365 days of simulation, assuming 50% standby capacity and a maximum design temperature of 20 ^{o}C.

Biomass recirculation, permeate pumps and anoxic zone mixers were sized based on those typical of a large scale plant, with one standby in each case. One agitator per 450 m^3 of anoxic tank volume was assumed. Costs of land, civil engineering, other electrical equipment and construction were excluded, these being location specific.

7.2.4 Operational costs

Operational costs were determined using the approach of the control strategy evaluation benchmark community (Copp *et al.,* 2002), which was extended by Maere *et al.* (2009) for MBR applications. The opex analysis was limited to energy demand, sludge treatment and disposal, and chemical usage for membrane cleaning.

7.2.4.1 Energy demand

The individual contributions to energy demand are described below, and a Germany-specific energy cost of €0.0942.kWh^{-1} used throughout (BMWi, 2010).

Aeration energy The influence of MLSS concentration (via the α-factor) and aerator type (fine and coarse bubble) on oxygen transfer was computed using the dedicated aeration model of Maere *et al.* (2009), combining several literature findings (Metcalf and Eddy, 2003; Henze *et al.,* 2008; Verrecht *et al.,* 2008; Krampe and Krauth, 2003; Germain *et al.,* 2007; Stenstrom and Rosso, 2008).

Based on typical practically measured values for blower outlet pressure (106300 Pa; for a typical aerator depth of 5 m and allowing for losses incurred in the pipework) and a blower efficiency ξ_B of 0.60, a value of 0.025 $kWh.Nm^{-3}$ air was determined for the aeration energy demand, corresponding well with literature values (Verrecht *et al.*, 2008) and data from blower manufacturers. The average total aeration energy in $kWh.d^{-1}$ was obtained by summing blower power consumption for both membrane and biology blowers and integrating over the 365 day simulation period (Maere *et al.*, 2009).

Pumping energy Sludge pumping requirements, for internal recirculation (Q_{int}, $m^3.d^{-1}$), membrane recirculation (Q_{MR}, $m^3.d^{-1}$) and wastage (Q_W, $m^3.d^{-1}$) (Figure 7-2), were determined from the expression of Maere *et al.* (2009), using a power requirement of 0.016 $kWh.m^{-3}$ of sludge pumped which was calculated from assuming a simple linear dependency of P_{Sludge} (Power required for sludge pumping) on sludge flow and assuming a total headloss Δh of 3m and a pump efficiency ξ_p of 50%. To calculate additional pumping energy for permeate pumping and backwashing, the expression provided by Maere *et al.* (2009) was applied.

Mixing energy A typical constant mixing power requirement of 8 W per m^{-3} of anoxic tank volume was used (Metcalf and Eddy, 2003), with no supplementary mechanical mixing required for the aerobic, membrane and buffer tanks.

7.2.4.2 Sludge production

Sludge production (in $kg.d^{-1}$) was calculated using the expressions of Copp *et al.* (2002), adapted for MBR use by Maere *et al.* (2009). Reported costs for sludge handling and disposal vary from €43.$tnDS^{-1}$ (Rossi *et al.*, 2002), which accounts for chemicals, labour, treatment and disposal, to €259.$tnDS^{-1}$ (Stensel and Strand, 2004), based on costs for collection, thickening, digestion, dewatering, reuse, but excluding haulage. Sludge handling cost figures across a broad range of values were thus considered.

7.2.4.3 Chemical consumption

A typical membrane cleaning protocol and frequency based on literature data (Brepols *et al.*, 2008; Judd and Judd, 2010) was assumed to provide chemical consumption data. The protocol comprised a weekly clean in place (CIP) with 500 ppm NaOCl and 2000 ppm citric acid, and a cleaning out of place (COP) with 1000 ppm NaOCl and 2000 ppm citric acid, conducted twice yearly. Representative prices for bulk chemicals were obtained from chemical suppliers.

7.2.5 Effluent quality evaluation

Evaluation of effluent quality was based on the approach of Copp *et al.* (2002), which quantifies the pollution load to a receiving water body in a single parameter, the effluent quality index (EQI), in kg pollution $units.d^{-1}$ (kg $PU.d^{-1}$). A larger EQI thus indicates worse effluent quality. The average EQI was determined through integrating the expressions of Copp (2002) over the evaluation period, using the weighting factors β_x as reported by Vanrolleghem *et al.* (1996).

Table 7-1: MBR design parameters and base case costs for the study of operational and design parameters

Parameter	Units	Value	Reference		Units	Value
Assumptions for capex calculation				**Base design, EQI and NPV**		
Membrane cost	€.m^{-2}	50	Judd & Judd, 2010	Design capacity	$m^3.d^{-1}$	30,416
Tank civil cost	€.m^{-3} tank volume	220	Brepols, 2010a,b	Maximum plant capacity*	$m^3.d^{-1}$	42,582
Screens – 0.75 mm	€.$m^{-3}.d^{-1}$ capacity	3.1 – 5.6**	Manufacturers	Total tank volume	m^3	7,097
Screens – 6mm	€.$m^{-3}.d^{-1}$ capacity	0.9 – 2.1**	Manufacturers	Membrane area	m^2	63,366
Blowers	€.$Nm^{-3}.h^{-1}$ capacity	4 – 4.3**	Manufacturers	SRT	d	23.8
Permeate pumps	€.$m^{-3}.h^{-1}$ capacity	58.8	Manufacturers; Brepols, 2010a,b			
Biomass recirculation pumps	€.$m^{-3}.h^{-1}$ capacity	12.1	Manufacturers; Brepols, 2010a,b	Buffer tank size	m^3	14,530
Mixing equipment	€.m^{-3} tank volume	27.8	Brepols, 2010a,b	Maximum flow out of buffertank*	$m^3.d^{-1}$	12,166
Assumptions for opex calculation				Max HRT in buffer tank	d	1.2
Energy cost	€.kWh^{-1}	0.0942	BMWi, 2010			
Sludge treatment cost	€.ton^{-1} of DS	150	-	Effluent quality index	kg PU.d^{-1}	5,430
Citric acid 50%	€/ton	760	Brepols, 2010a,b	NH_4-N	$mg.l^{-1}$	0.52
NaOCl 14%	€/m^3	254	Brepols, 2010a,b	NO_3-N	$mg.l^{-1}$	10.7
Assumptions for NPV calculation				COD	$mg.l^{-1}$	30.1
Membrane life	Year	10	Judd & Judd, 2010	Net present value	M€	26.7
Inflation	%	3%	-			
Discount rate	%	6%	-			

* As determined by the design requirement that maximum sustainable flux = 140% of design flux

** Depending on size of installed equipment

7.2.6 Net present value calculation

The net present value was calculated for a plant lifetime of 30 years, taking into account all capital and operational expenditures during the plant lifetime:

$$NPV = \sum_{t=0}^{29} \frac{(capex)_t + (opex)_t}{(1+i)^t} \qquad \textbf{(7.1)}$$

A membrane life of 10 years was assumed, corresponding to two complete membrane refits during the projected plant lifetime, based on recently reported trends (De Wilde *et al.*, 2007b). Long term inflation was assumed to be 3%, while a discount rate *i* of 6% was used, comparable to values used by Côté *et al.* (2004).

7.3 RESULTS AND DISCUSSION

7.3.1 Effect of contingency: changes in feedwater flow and strength

7.3.1.1 Hybrid plant vs. plant designed for maximum flow

Table 7-2 shows a breakdown of costs for two extreme scenarios:

a) the MBR part of a 'hybrid' plant (i.e. an MBR parallel to a CAS plant; the MBR is designed to treat a constant daily flow, while excess flow is treated by the CAS plant, that is not taken into account in this analysis); and

b) a plant designed to cope with maximum flow conditions (peak flow = 3 x average flow).

The results illustrate that deviating from the ideal 'hybrid' plant scenario leads to severe plant under-utilization, and a resulting cost penalty manifested in a 59% increased NPV value over that of the hybrid plant, despite treating the same cumulative flow over the plant life. The EQI is 3.8% lower for the 'hybrid' plant, due to the constant HRT of 8h, while for the plant designed for maximum flow the HRT can be as low as 4h during peak flows.

Table 7-2: Capex, opex and resulting NPV for an MBR treating steady state influent, as part of a hybrid plant, and a MBR, designed for maximum flow without buffer tanks.

	Unit	MBR part of a hybrid plant	Plant designed for maximum flow
Average plant influent flow	$m^3.d^{-1}$	20,851	20,851
Maximum flow to the MBR	$m^3.d^{-1}$	20,851	59,580
Total tank volume	m^3	6,949	9,930
Average plant utilisation	%	100%	34%
Effluent Quality Index	kg $PU.d^{-1}$	5,035	5,236
$COD_{average}$	$mg.l^{-1}$	29.7	30.15
$NH_4\text{-}N_{average}$	$mg.l^{-1}$	0.46	0.43
$NO_3\text{-}N_{average}$	$mg.l^{-1}$	10.4	9.55
TOTAL CAPEX	**Euro**	**4,634,387**	**7,844,684**
Screens	%	11.8	8.4
Membranes	%	46.9	56.5
Tank construction	%	33.0	27.9
Biology blowers	%	1.4	0.8
Membrane blowers	%	1.5	1.6
Permeate pumps	%	1.5	2.2
Mixing equipment	%	1.9	1.4
Recirculation pumps	%	2	1.2
TOTAL OPEX	**Euro/year**	**618,602**	**891,373**
Energy	%	79.6	84.1
Sludge treatment and disposal	%	17.9	12.3
Chemicals	%	2.5	3.6
NET PRESENT VALUE	**Euro**	**19,047,870**	**30,209,875**

Figure 7-3 shows a breakdown of the energy demand for the same two plants. The values obtained are in line with those reported for full scale plants (Garcés *et al.*, 2007; Brepols *et al.*, 2009). The average energy demand for the 'maximum flow' plant is ~54% higher, mostly due to under-utilisation of the available membrane capacity and the resulting excess aeration. This illustrates that effective control strategies where membrane aeration as applied in proportion to flow conditions could generate significant opex savings.

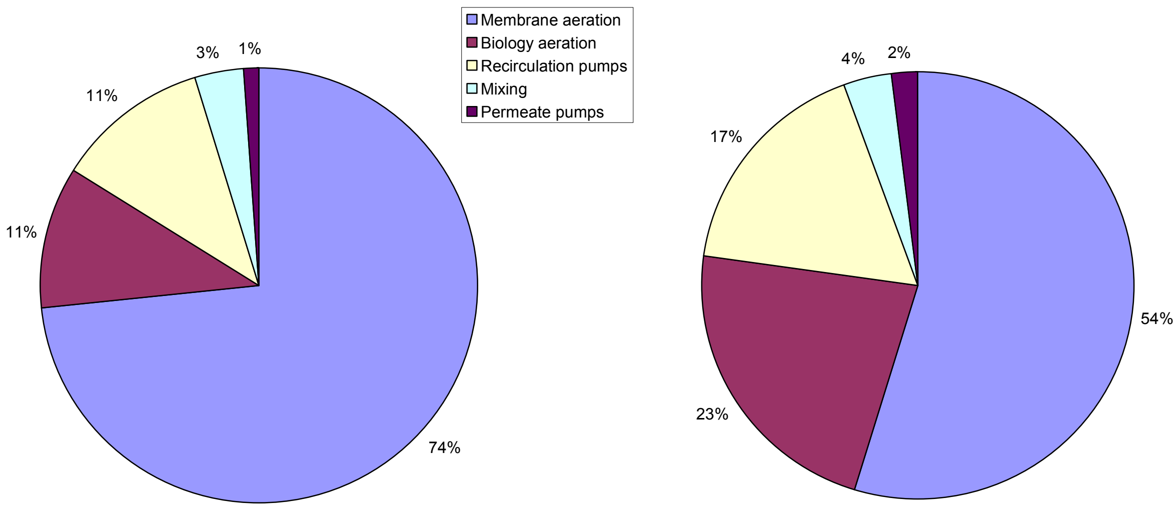

Figure 7-3: Breakdown of energy demand for a) the MBR part of a hybrid plant (Average total energy demand = 0.7 kWh.m^{-3}) and b) a plant designed for maximum flow (Average total energy demand = 1.07 kWh.m^{-3})

The analysis shows NPV and operational efficiency of MBRs to be very susceptible to the extent of built-in contingency, which is mostly determined by the changes in feedwater flow such as during storm events. An example of this is the 48 MLD (megalitres per day) Nordkanal plant in Germany (Brepols *et al.*, 2009), which was designed to treat a peak flow that is 3-4 times higher than the average flow. The plant also has 33% more membrane surface installed than required to treat the peak flow, a requirement under German regulations. Consequently, mean fluxes at the plant are only 8 LMH and specific energy consumption for the MBR is 0.5-1.8 kWh.m^{-3}. Conversely, the hybrid MBR plant at Ulu Pandan in Singapore is designed to continuously treat a flow of 23 MLD, leading to very efficient operation and energy consumption as low as ~0.4 kWh.m^{-3} for the MBR part of the hybrid plant, mainly due to continuous improvement in membrane aeration protocols (Qin *et al.*, 2006, 2007; Seah *et al.*, 2009). Thus, provided there is a constant demand for high quality effluent for reuse, the hybrid plant is the most favoured option. This can be retrofitted to an existing CAS, provided full effluent disinfection is not required (Lesjean *et al.*, 2009; Mulder, 2008).

7.3.1.2 Buffer tank vs. extra membranes

Adding buffering capacity for flow equalisation permits a smaller plant design with a reduced membrane surface requirement and so higher average plant utilization. Figure 7-4 shows the influence of buffer tank size on mean plant utilisation, and contrasts the resulting membrane and buffer tank costs. For practical reasons, the buffer tank size is constrained by an HRT of below 2 days (corresponding to 80% of design flow). Since no influent can bypass the plant under storm conditions the combined capacity of the buffer tank and MBR plant must cope with the maximum flow. Figure 7-4 shows that over the buffer tank size range considered, the cost of adding a buffer tank is only partially offset by the cost savings from a reduction in required membrane surface area due to increased average plant utilization (Figure 7-5). The EQI and NPV trends are both determined by the constraints on tank size imposed by an HRT of 8h at average flow, or a minimum HRT of 4h at maximum design flow. Addition of a buffer tank with the maximum acceptable size results in a NPV decrease from €30.2 million to €27 million, or a saving of 10.5%, due to decreased opex (-21%), which is partly offset by an increase in capex (+32%) (Figure 7-6). A maximum NPV saving of 11.8% can be achieved through addition of buffer tank with the most economical size (i.e 1.2 d HRT; at average plant utilization of 47.9%). Effluent quality, as indicated by EQI, is largely unaffected and deteriorates by a maximum of 4% over the buffer tank size range considered (Figure 7-5).

The cost of land required for the buffer tank is excluded from this NPV analysis. However, provided the additional land required for the buffer tank has a projected value of less than €3.2m, it is always beneficial to build a buffer tank. Assuming a total plant footprint equaling 2.5 times the combined footprint required for the biotanks and buffer tanks (Brepols *et al.*, 2010), a plant with the maximum sized buffer tank (2d HRT) requires 9,715 m^2 extra land compared to a plant without buffer tank. Land costs would have to increase to €324 per m^2 before addition of a buffer tank becomes economically unviable. This value is at least 32% higher than typical reported values for industrial land in the Germany, which range between €17 and €247 per m^2 (Federal Statistics Office, 2010). Assuming a CAS to incur 2.7 times the footprint of an MBR (Brepols *et al.*, 2010), a combined MBR with the maximum sized buffer tank would be ~10% larger than a CAS treating the same flow.

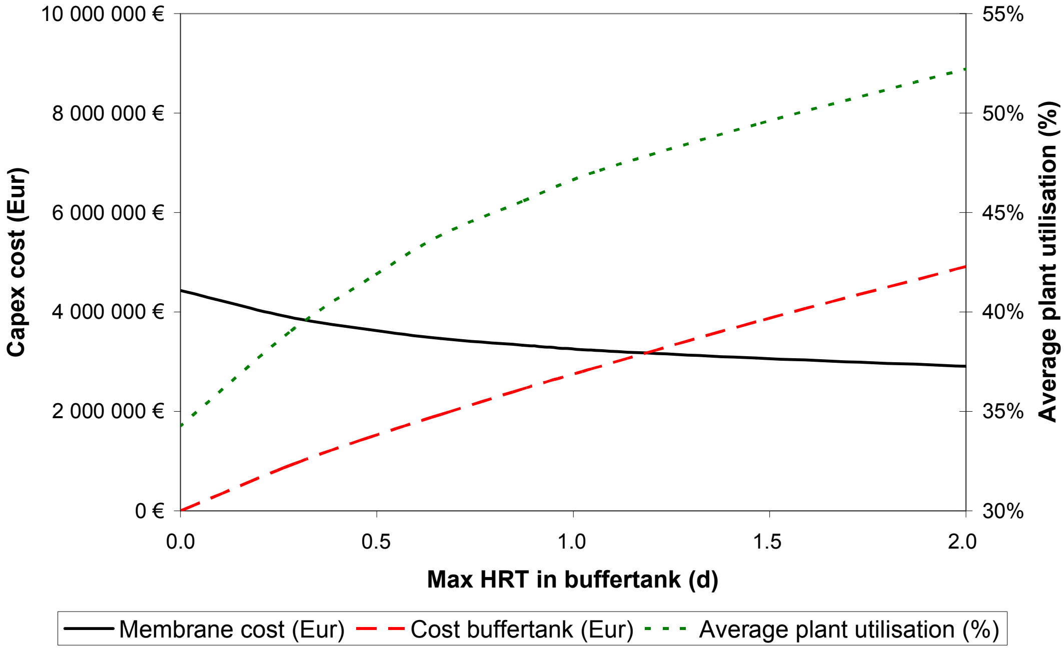

Figure 7-4: Influence of buffer tank size on plant utilisation, and a comparison of the resulting costs for membranes and buffer tank

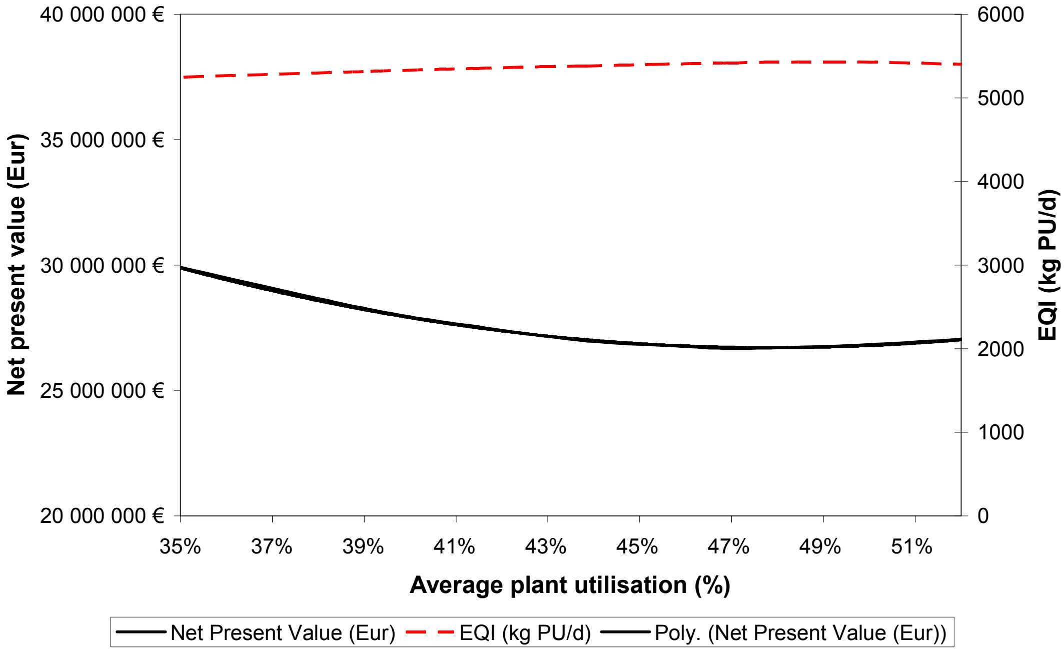

Figure 7-5: Influence of average plant utilisation on net present value and effluent quality index

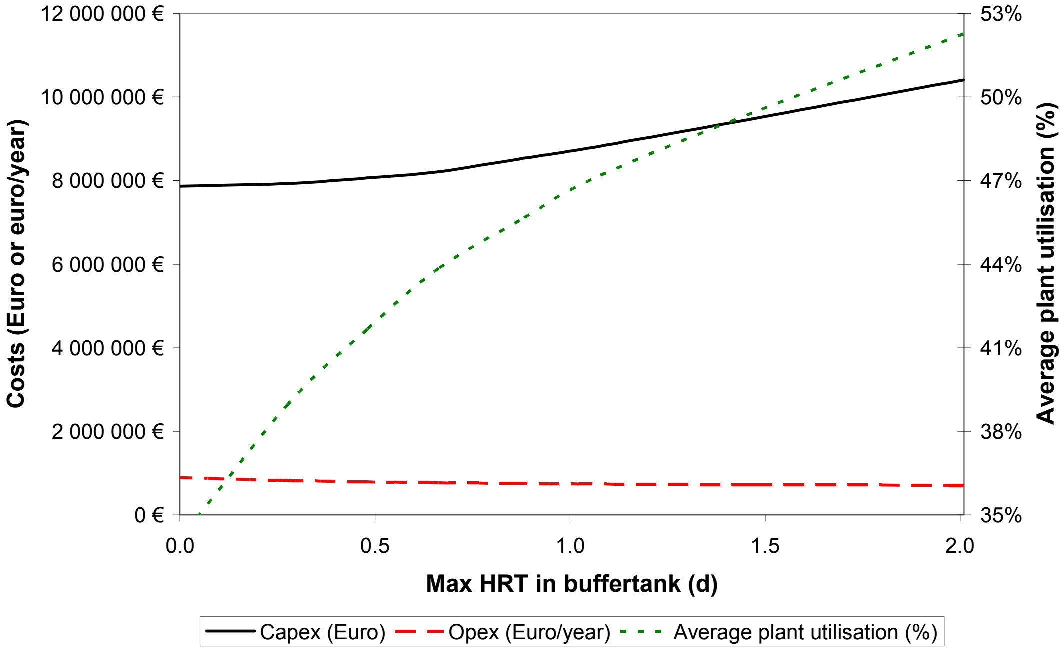

Figure 7-6: Influence of size of buffer tank on capex and opex

7.3.2 Influence of operational and design parameters

The influence of operational and design parameters on NPV and EQI was evaluated for the plant design with the most economically sized buffer tank, thus providing the lowest NPV (Table 7-1). Table 7-3 displays the variation in NPV and EQI resulting from changing parameter values for operation, design and costings within given ranges pertaining to full scale plants.

Influence of SRT A shorter design SRT decreases capex due to decreased installed aerobic tank blower capacity at the lower MLSS concentrations and the resulting decreased aeration demand. However, the cost for the process blowers is less than 2% of total capex (Table 7-2), so the potential influence is negligible. The reduction in NPV is attributed to the effect of SRT on opex. At a conservative sludge treatment and disposal cost of €150.m^{-3} of dry solids, energy consumption accounts for 78-85% of opex, sludge treatment and disposal for 12-19%, and chemical cleaning about 3%. The decreased aeration demand at lower MLSS concentrations and shorter SRT thus outweighs the costs incurred by increased sludge production. This would seem to corroborate recent trends of working at lower MLSS concentrations, particularly in the US (Trussell *et al.*, 2006, 2007), but is contrary to the conclusions of Yoon *et al.* (2004).

The latter study ignored membrane aeration, thus underestimating the total opex since membrane aeration contributes significantly to total energy demand (Figure 7-3).

The influence of SRT is sensitive to sludge treatment and disposal costs. As sludge management costs increase, the cost incurred by sludge treatment and disposal starts to outweigh the opex reduction from decreased energy demand at lower SRT. Table 7-3 also shows that effluent quality requirements place a lower limit on the SRT operating range, since EQI deteriorates as SRT decreases. Selection of SRT is thus based on available sludge processing facilities on site and end disposal costs, as well as the desired effluent quality. Operation at lower SRT and MLSS values may also lead to higher permeability decline rates (Trussel *et al.*, 2006), mitigating against lower SRT operation.

Influence of HRT Longer HRTs increase capex due to the larger tank volume required, but this is partially offset by lower opex at lower MLSS concentrations (10,000 and 6,000 mg/l average MLSS concentrations in aerobic tank at 6 and 10h HRT respectively). The impact on NPV is thus negligible compared to, say, the influence of contingency or choice of SRT. The effect on EQI is more pronounced: an increase in average HRT from 6 to 10 hours improves effluent quality by 9%. A larger MBR thus provides better effluent quality, without detriment to NPV provided land costs are not excessive.

Influence of anoxic fraction Increasing or decreasing the anoxic fraction of total tank volume has a negligible effect on NPV (Table 7-3), but a large impact on EQI. Increasing the anoxic fraction from 30 to 50% improves EQI by 18% due to improved denitrification.

Table 7-3: Sensitivity of NPV and EQI on design and operational parameters and costs. % Change in NPV and EQI is compared with the base conditions as described in Table 7-1

	Net present value		EQI	
	Million euro (M€)	% change	kg PU.d^{-1}	% change
Solids retention time (SRT)				
9.5 days	26.4	-1.1%	5,835	+7.5%
47.6 days	27.8	+4.4%	5,172	-4.7%
Hydraulic residence time (HRT)				
6 hours	26.3	-1.3%	5,628	3.7%
10 hours	26.8	+0.5%	5,214	-4.0%
Sustainable flux at membrane aeration				
15 l.m^{-2}.h^{-1} at SAD_p = 15.3	29.1	9.2%	5,551	+2.2%
30 l.m^{-2}.h^{-1} at SAD_p = 19.1	26.5	-0.5%	5,295	-2.5%
Buffer tank				
0 days HRT (No buffertank)	30.2	+13.4	5,236	-3.6%
2 days HRT (Maximum considered)	27.1	+1.6%	5,401	-0.5%
Anoxic tank volume				
f_{anox} = 30%	26.7	+0.1%	6,313	+16.3
f_{anox} = 50%	26.6	-0.3%	5,146	-5.2
Energy prices				
Rising by 4% annually	28.2	+5.7%	5,430	0%
Rising by 7% annually	34.7	+30.0%	5,430	0%
Sludge treatment costs (excluding hauling)				
43 Eur.ton^{-1} of DS	25.2	-5.6%	5,430	0%
300 Eur.ton^{-1} of DS	28.8	+7.9%	5,430	0%
Membrane costs				
20 Eur.m^{-2} membrane surface	22.4	-15.8%	5,430	0%
100 Eur.m^{-2} membrane surface	33.7	+26.4%	5,430	0%
Membrane costs – halving every ten years	24.2	-9.3%	5,430	0%
Membrane life – 5 years	32.8	+23.1%	5,430	0%

Influence of membrane aeration and sustainable flux Membrane aeration energy contributes significantly to opex (Verrecht *et al.*, 2008; Seah, 2009; Brepols *et al.*, 2009), as confirmed by Figure 7-3. Membrane aeration energy can be related to SAD_p, the specific aeration demand per unit permeate volume. Extensive pilot studies regarding the impact of membrane aeration and sustainable flux (Guglielmi *et al.* 2007, 2008) suggest a neo-linear relationship between sustainable flux J and U, the in-module air flow velocity in m.s^{-1} (Verrecht *et al.*, 2008). For HF geometry, calibrating against two full scale plants (Verrecht *et al.*, 2008), the correlation between J and U can be expressed as:

$$J = m \cdot \frac{SAD_m \cdot L_{membrane}}{\left(\frac{1}{\varphi} - \frac{d_f}{4}\right)} + J_0 \qquad \text{for } J < J_{sust,max}\ (\mathrm{l.m^{-2}.h^{-1}}) \qquad \textbf{(7.2)}$$

$$J = J_{sust,max}\ (\mathrm{l.m^{-2}.h^{-1}}) \qquad \text{for } SAD_m > SAD_{m,max}\ (\mathrm{Nm^3.m^{-2}.s^{-1}}) \qquad \textbf{(7.3)}$$

where J is the flux through the membrane, in $m^3.m^{-2}.h^{-1}$ and SAD_m the specific aeration demand per unit membrane area in $Nm^3.m^{-2}.h^{-1}$, m the slope of the J vs U curve (247, according to Verrecht *et al.*, 2008), L the membrane module length (1.8 m); φ the module packing density (300 m^{-1}), d_f the hollow fibre outside diameter (0.002 m) and J_0 the intercept of the J vs. U curve (5 $l.m^{-2}.h^{-1}$). Thus:

$$SAD_p = \frac{SAD_m}{J} \qquad \textbf{(7.4)}$$

where a minimum SAD_p is required to maintain a sustainable flux (7.2), but increasing SAD_p beyond $SAD_{m,max}$ has no impact on the sustainable flux (7.3) and a higher maximum sustainable flux demands a higher SAD_p. When considering the influence of sustainable flux and SAD_p on NPV (Figure 7-7), higher sustainable fluxes lead to lower NPVs, indicating that the higher operational costs are offset by lower capital expenditures which can mainly be attributed to the reduction in membrane capacity required. An increase in sustainable flux from 15 to 30 $l.m^{-2}.h^{-1}$ results in a decrease in NPV of 9% (at minimum required SAD_p; Table 7-3). It can thus be concluded that higher sustainable fluxes are beneficial to NPV, despite the higher aeration demand and associated increase in opex, provided $J \leq J_{sust,max}$ and J_o takes a positive value.

Energy cost For an annual energy price rise of 4%, in line with the historical average (EIA, 2009), a 5.7% increase NPV arises over the base case for inflation-linked energy costs. A ‘worst case’ of a 7% annual increase, corresponding to a doubling of energy prices roughly every 10 years, increases NPV by 30%.

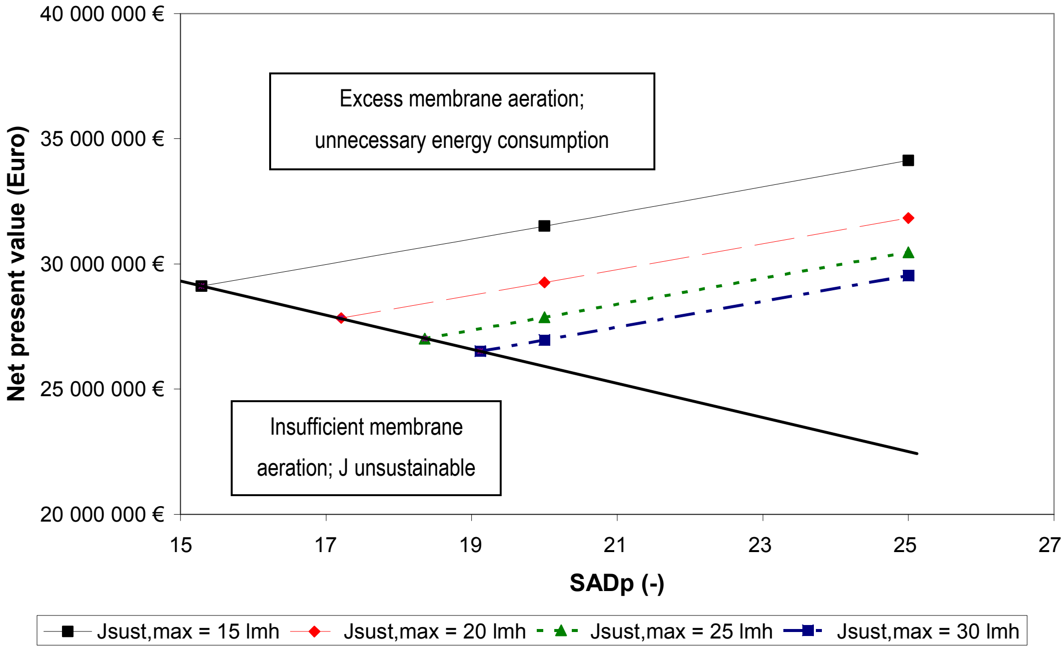

Figure 7-7: Influence of SADp on net present value for a range of sustainable fluxes

Membrane replacement and cost As shown in Table 7-2, membrane costs make up 47-57% of total capex, while the other process equipment combined contributes about 20%. Analysis of component lifetime cost impacts is thus most sensitive to membrane life and costs. A 'worst case' membrane lifetime of 5 years (i.e. 6 membrane replacements in the projected plant lifetime of 30 years) results in a 23% increase in NPV compared to the base cost assuming membrane replacement every 10 years. A halving of membrane costs every 10 years, on the other hand, reduces NPV by 9.2%, whereas an increase in initial membrane cost from €20.m^{-2} to €100.m^{-2} increases NPV by 50% for a 10 year membrane life and by 85% for a 5 year membrane lifetime.

Since membrane replacement is critical in determining NPV, it is unsurprising that considerable attention has been paid to optimisation of membrane lifetime by operating under a sustainable regime and developing adequate cleaning strategies (Brepols *et al.*, 2008). There is increasing evidence that MBR membrane life can reach, or even exceed, a decade for large plants. The Zenon plants at Rodingen (3.2 MLD PDF) and Brescia (42 MLD PDF) are successfully operating with membrane modules which are from 2000 and 2002 years respectively, and the Kubota plant at Porlock still operates with 40% of the panels originally installed in 1997 (Judd and Judd, 2010); predicted replacement intervals of up to 13 years have been reported (De Wilde *et al.*, 2007b).

7.4 CONCLUSIONS

A cost sensitivity analysis, using dynamic simulation results, with respect to design and operational parameters for an MBR over the lifetime of the plant has revealed:

- The contingency provided for changes in feedwater flow and composition impacts significantly on net present value (NPV). The analysis shows that any deviation from the ideal 'hybrid' plant, where the MBR treats a constant influent stream, leads to plant under-utilisation and a resulting cost penalty manifested as an increase of up to 58% in NPV for a plant designed for three times the mean flow.
- Addition of a buffer tank for flow equalisation increases average plant utilisation, leading to more efficient operation and a resulting reduction in opex, whilst capex can also be reduced according to the reduction in membrane area and MBR plant size. In the example presented, a decrease in NPV of up to 11% with increased average plant utilisation from 34 to 48% results.
- Addition of a buffer tank is economically beneficial as long as the cost of land required is less than the NPV saving achieved. In the example presented, addition of a buffer tank is economically viable for increased land costs below €324.m^{-2}, an excessive value for industrial land. An MBR with the maximum sized buffer tank (2d HRT) has a footprint approximately 10% greater than that of a conventional activated sludge plant.
- An increased SRT at constant tank volume increases the NPV since a greater aeration demand is incurred at higher MLSS concentrations. Whilst sludge production is concomitantly reduced, the resulting cost savings do not fully offset the increased energy costs. However, results are very sensitive to sludge treatment and disposal costs. The effect of HRT on NPV is minimal, if land costs are negligible, but a higher average HRT improves effluent quality.
- Higher sustainable fluxes provide a decreased NPV. Although the opex is increased due to the higher aeration demand, this is offset by the reduction in capex and membrane replacement costs since less membrane area is required. An increase in sustainable flux from 15 to 30 LMH decreases NPV by 9%.
- The future trend in energy costs is a determining factor for NPV: a doubling of energy costs every ten years increases the NPV by 30%.

- A membrane lifetime of 5 years results in an NPV 23% higher compared to a 10 year membrane replacement interval, for a constant membrane cost of €50.m^{-2}. If initial membrane costs increase five-fold from €20 per m^2, NPV increases by 85% for a 5 year membrane lifetime and by 50% for a 10 year membrane life.

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CHAPTER 8

CONCLUSIONS AND FURTHER WORK

8 CONCLUSIONS AND FURTHER WORK

8.1 CONCLUSIONS

The present thesis has extended understanding of the impact of design and operational parameters of membrane bioreactors on energy demand and ultimately costing. The following conclusions can be drawn:

1. A literature review (Chapter 2) has revealed that membrane aeration is commonly used for fouling mitigation in full scale MBRs, and that an optimum membrane aeration rate exists as a function of flux. Membrane aeration has a significant impact on O&M costs, as it typically accounts for 35-50% of the total plant energy demand. Improvements in module design and the advent of intermittent aeration have resulted in a significant decrease in energy consumption for membrane aeration (*Objective 1*).
2. If a general algorithm for flux vs. aeration is applicable, significant reductions in energy demand may be attainable through operating at lower MLSS levels and membrane fluxes. A simple heuristic model (Chapter 4) has shown that halving the flux can reduce the aeration energy demand by as much as 45%, illustrating the benefits of adjusting aeration intensity according to the hydraulic load. However, operating under lower MLSS concentrations and fluxes has negative repercussions on capex (*Objectives 2a, 4a and 4b*).
3. The use of 10:10 or 10:30 intermittent aeration results in lower fouling rates when compared to continuous aeration when working under the same overall airflow rate (Chapter 5). A lower SAD_p is required for sustainable operation, with associated membrane aeration energy savings of up to 75% when comparing 10:30 to continuous aeration, without being penalised in terms of the fouling rate. These findings are corroborated by recent industrial practice (*Objective 2b*).
4. The application of activated sludge modelling (ASM) to MBR was verified on a small community-scale MBR (Chapter 6). The model provided a good prediction of the dynamic nutrient removal profile and MLSS concentrations by using default ASM2d values for all biokinetic and stoichiometric parameters, and a physical-chemical influent fractionation method. A dedicated aeration model was used, incorporating the effect of elevated MLSS concentrations on

oxygen transfer, and allowing differentiation between coarse and fine bubble aeration, which has proven necessary to accurately simulate aeration demand *(Objectives 1 and 3)*.

5. A simple model which combines empirical correlations with ASM modelling can be used to accurately predict the performance of an MBR diurnally and seasonally, thereby establishing when the process is at risk of failing consent and allowing optimisation of aeration demand, sludge recirculation and DO set point to reduce energy consumption and costs. This has been demonstrated for a small full scale MBR plant (Chapter 6), resulting in a 23% reduction in energy demand. The developed methodology can be transferred to other full scale MBR plants. (*Objectives 1, 2 and 3*).
6. Cost reduction in small MBRs relies on process robustness with minimal manual intervention to suppress labour costs (Chapter 3). Operational costs are an order of magnitude higher than those reported for large scale MBRs without post-treatment, due to operational inefficiencies inherent in small scale plants and the disproportionate amount of staff time required. In the studied case, staffing costs required for plant attendance and maintenance are the largest contribution to total opex, followed by energy consumption. This is contrary to findings for large scale plants, where energy costs are the dominating contributor to opex. Post-treatment of the MBR effluent, required mainly for aesthetic reasons, adds significantly to opex (30%). If post-treatment and labour are excluded, opex costs are about 5 times higher than costs reported for large scale plants, proportional to the higher energy consumption of small scale plants (*Objectives 1 and 4a*).
7. A cost sensitivity analysis has revealed the design and operational parameters influencing the costing of large MBRs (Chapter 7). The contingency provided for changes in feedwater flow and composition impacts significantly on net present value (NPV), showing the potential of 'hybrid MBR' configurations where the MBR treats a constant influent flow. Addition of a buffer tank to increase average plant utilisation is economically beneficial as long as the cost of land required is less than the NPV saving achieved. An increased SRT at constant tank volume increases the NPV since a greater aeration demand is incurred at higher MLSS concentrations. Whilst sludge production is concomitantly reduced, the resulting cost savings do not fully offset the increased energy costs. However, results are very sensitive to sludge

treatment and disposal costs. The effect of HRT on NPV is minimal, if land costs are negligible, but a higher average HRT leads to improved effluent quality. Higher sustainable fluxes lead to a decreased NPV. Although the opex is increased due to the higher aeration demand, this is offset by the reduction in capex and membrane replacement costs as less membrane area is required. The future trend in energy costs is a determining factor for NPV, as is the membrane replacement interval: a membrane lifetime of 5 years results in an NPV that is 23% higher compared to a 10 year membrane replacement interval (*Objectives 1 and 4b*).

8.2 SUGGESTIONS FOR FUTURE WORK

Further work to address gaps in knowledge identified during the programme is as follows:

1. There remains a need to establish the relationship between the attainable net flux and the aeration demand under increasingly intermittent aeration conditions, i.e. for aeration at <25% of the filtration time. Whilst the lower fluxes necessary would increase capital costs, this could potentially be offset by the lower operating costs over the plant lifetime as indicated by the NPV analysis.
2. PO_4-P removal in full scale plants is often better than expected, even in the absence of anaerobic zones. This is sometimes attributed to insufficient mixing resulting in local anaerobic zones. However, the exact mechanism responsible is not yet clear, and requires further study.
3. Advanced feedback and feed forward control of aeration and recirculation rates in an MBR, based on influent quality and comparison of actual and desired effluent nutrient concentrations, could lead to a significant decrease in operational costs. However, the potential benefits have to be compared against the added process complexity and increased risk of process failure.
4. The contribution of coarse bubble membrane aeration to the aeration demand for biological degradation could potentially be increased through modification of the MBR process layout, decreasing the fine bubble aeration requirement.

APPENDICES

APPENDICES

Appendix A iMBR design methodology: steady-state model

Published in Judd, S.J. and Judd, C. (2010) 'The MBR Book (2^{nd} ed.): Principles and applications of membrane bioreactors in water and wastewater treatment', Elsevier. Crossreferences indicated with an asterisk (*) refer to sections and figures in 'The MBR book (2^{nd} ed).'

The design for an iMBR is generally based on a combination of empirical/heuristic data and biokinetics/biochemical stoichiometry. This then demands information and/or appropriate assumptions regarding the interrelationship between aeration and:

a) permeability and cleaning protocol for the membrane permeation component, and
b) feedwater quality, flows and biokinetics for the biological component.

Whilst the latter can be approached from a biochemical basis, the former cannot reasonably be calculated from first principles. Three distinct but interrelated design phases thus arise, namely the design of: 1) the membrane process, 2) the biological process (and determination of oxygen demand), and 3) the aeration systems.

An example of a design calculation for a nitrifying/denitrifying MBR (Fig. A-1), 25 MLD in capacity and based on HF technology, is detailed below. The plant is intended to treat medium strength wastewater (Table A-1) to a required effluent standard of 0.5 mg/l of NH_4-N (N_e) and 12 mg/l NO_3-N (NO_e). Bio-P removal has not been incorporated in the process design, and more details in this regard are available elsewhere (Metcalf and Eddy, 2003; Grady *et al.*, 2010). The design methodology is schematically illustrated in Figure A-2 with the governing biological and aeration design equations listed in Table A-2.

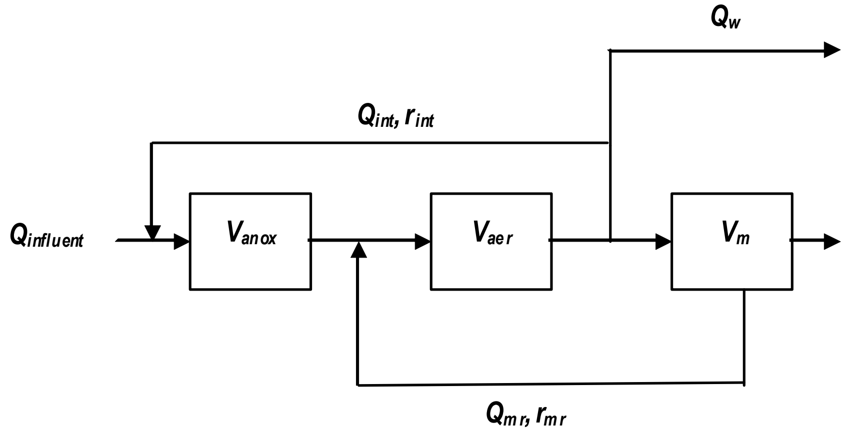

Figure A-1: Schematic overview of generic nitrifying/denitrifying MBR

Table A-1: Feedwater characterisation (adapted from Metcalf and Eddy, 2003)

Parameter	*Unit*	*Value*	*Parameter*	*Unit*	*Value*	*Eq*
Q	m^3/d	25,000	Alkalinity	g/m^3	200	
Q_{peak}	m^3/d	50,000	f_{bs}	-	0.2*	
COD	g/m^3	430	f_{bp}	-	0.5*	
BOD *S*	g/m^3	190	f_{us}	-	0.05*	
TSS	g/m^3	210	f_{up}	-	0.25*	
VSS	g/m^3	160	*bpCOD*	g/m^3	215	A.22
TKN	g/m^3	40	*pCOD*	g/m^3	323	A.23
NH_4-N	g/m^3	25	*iTSS*	g/m^3	50	A.8
NO_3-N	g/m^3	0	*nbVSS*	g/m^3	53.3	A.7

* *Based on typical wastewater fractionation (Ekama et al., 1984)*

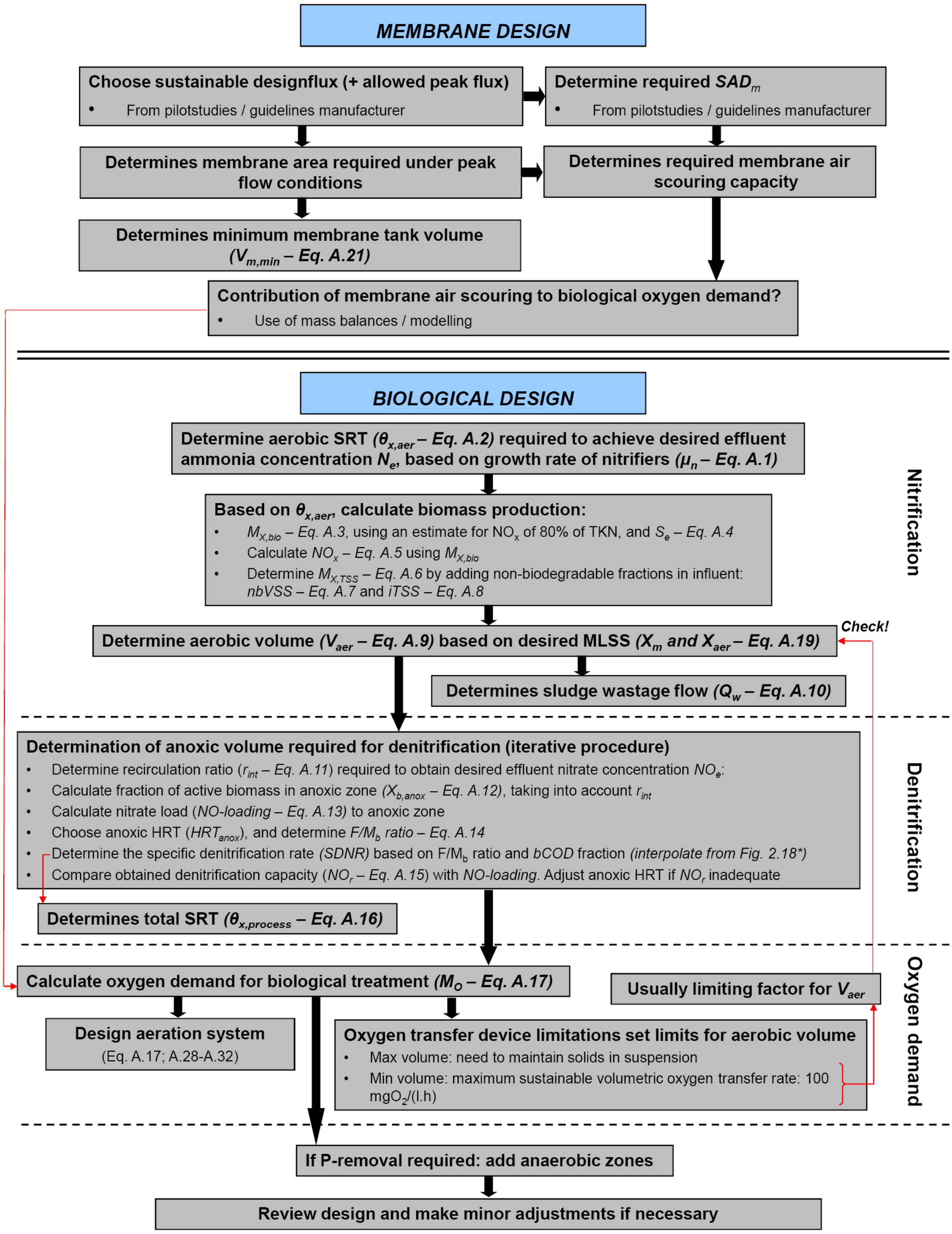

Figure A-2: MBR design calculation

Table A-2: Biological and aeration operating parameters and design equations

Biological parameters	***Equation***	
Specific growth rate of nitrifying bacteria, μ_n in gVSS/(gVSS.d)	$\left(\frac{\mu_{n,m} N_e}{K_n + N_e}\right)\left(\frac{DO}{K_O + DO}\right) - k_{e,n}$	A.1
Aerobic SRT, $\theta_{x,aer}$ in d	$SF \frac{1}{\mu_n}$	A.2
Sludge yield, $M_{X,bio}$ in g/d	$\frac{QY(S-S_e)}{1+k_e\theta_{x,aer}} + \frac{f_d k_e QY(S-S_e)\theta_{x,aer}}{1+k_e\theta_{x,aer}} + \frac{QY_n(NO_x)}{1+k_{e,n}\theta_{x,aer}}$	A.3
Effluent BOD, S_e g/m³	$\frac{K_s(1+k_e\theta_{x,aer})}{\theta_{x,aer}(\mu_m - k_e) - 1}$	A.4
Concentration of TKN oxidised to form nitrate, NO_x in g/m³	$TKN - N_e - 0.12 \, {}^{M_{X,bio}}\!/_{Q}$	A.5
Total sludge yield, $M_{X,TSS}$ in g/d	${}^{M_{X,bio}}\!/_{0.85} + Q(nbVSS + iTSS)$	A.6
Influent non-biodegradable VSS, *nbVSS* in g/m³	$\left(1 - {}^{bpCOD}\!/_{pCOD}\right)VSS$	A.7
Influent inert TSS, *iTSS* in g/m³	$TSS - VSS$	A.8
Aerobic tank volume, V_{aer} in m³	$\frac{M_{X,TSS}\theta_{x,aer}}{X_{aer}}$	A.9
Sludge wastage flow, Q_w in m³/d	$\frac{V_{aer}}{\theta_{x,aer}}$	A.10
Nitrate recirculation ratio, r_{int} -	$\frac{NO_x}{NO_e} - 1$	A.11
Active biomass concentration in anoxic zone, $X_{b,anox}$ in g/m³	$\left(\frac{Q\theta_{x,aer}}{V}\right)\left(\frac{Y(S-S_e)}{1+k_e\theta_{x,aer}}\right)\left(\frac{r_{int}}{r_{int}+1}\right)$	A.12
Nitrate load to anoxic zone, *NO-loading* in g/d	$Qr_{int}\left(TKN - N_e - 0.12 \, {}^{M_{X,bio}}\!/_{Q}\right)$	A.13
Food to active biomass concentration in anoxic zone, F/M_b in g BOD/(g TSS.d)	$\frac{QS}{V_{anox}X_{b,anox}}$	A.14
Denitrification capacity, NO_r, in g/d	$V_{anox}SDNR(X_{b,anox})$	A.15

SRT, $\theta_{x,process}$ in d	$\frac{V_{aer}X_{aer}+V_{anox}X_{anox}}{Q_w X_{aer}}=\theta_{x,aer}+\theta_{x,anox}$	A.16
Oxygen requirement, M_o in kg/d	$Q(S-S_e)-1.42M_{X,Bio}+4.33QNO_X-2.86Q(NO_x-NO_e$ $=M_m+M_b$	A.17
MLSS concentration in anoxic zone, $X_{anox,}$ in g/m³	$X_{aer}\left(\frac{r_{int}}{1+r_{int}}\right)$	A.18
MLSS concentration in aerobic zone, X_{aer}, in g/m³	$X_m\left(\frac{r_{mr}}{1+r_{mr}}\right)$	A.19
Anoxic tank as fraction of total volume, f_{anox}	$\frac{V_{anox}}{V_{process}}$	A.20
Minimum required membrane tank volume, $V_{m,min}$ in m³	$\frac{A_m}{\varphi_{\tan k}}$	A.21
Biodegradable particulate COD, *bpCOD* in g/m³	$f_{bp}COD$	A.22
Total particulate COD, *pCOD*, in g/m³	$(f_{bp}+f_{up})COD$	A.23
Sludge waste per unit permeate, Q'_w m³/m³	$\frac{Q_w}{J_{net}A_m}$	A.24
Hydraulic retention time for tank x, HRT_x, in h	$\frac{V_x}{Q}$	A.25
Total process volume, $V_{process}$, in m³	$V_{anox}+V_{aer}$	A.26
Total process HRT, $HRT_{process}$, in h	$HRT_{anox}+HRT_{aer}$	A.27
Aeration parameters	***Equation***	
Suspended solids correction factor, *α*	$e^{-\omega_x X}$, where *x* refers to fine or coarse bubble	A.28
Temperature correction factor, Φ	$1.024^{(T-20)}$	A.29
Membrane aeration rate, $Q_{A,m}$ in Nm³/h	$SAD_m A_m$	A.30
O_2 transferred by membrane aeration, M_m in kgO_2/d	$Q_{A,m}\rho_A(SOTE_{coarse}y_{coarse})O_{A,m}\alpha\beta\Phi$	A.31
Net air flow for biological requirements, $Q_{A,b}$ in Nm3/h	$\frac{M_0-M_m}{\rho_A(SOTE_{fine}y_{fine})O_{A,m}\alpha\beta\Phi}$	A.32

Symbols

bCOD Biodegradable COD in g/m³

f_{anox}	Anoxic tank as % of aerobic tank size
f_{bp}	Slowly biodegradable COD fraction
f_{bs}	Readily biodegradable COD fraction
f_{up}	Particulate non-biodegradable COD fraction
f_{us}	Soluble non-biodegradable COD fraction
k_e	Heterotrophic endogenous decay coefficient, g VSS/(g VSS d)
$k_{e,n}$	Nitrification endogenous decay coefficient, g VSS/(g VSS d)
K_s	Saturation coefficient (heterotrophic), gBOD/m^3
$MLSS$	Design mixed liquor suspended solids, g/m^3
$O_{A,m}$	Mass percentage of oxygen in air, %
Q	Average flow, in m^3/d
M_m	Oxygen transferred by membrane aeration, kg/d
r_{mr}	Membrane recirculation ratio
S	BOD influent, in g/m^3
$SDNR$	Specific denitrification rate, gNO_3-N/(gMLVSS.d)
$SOTE_{fine}$ and $SOTE_{coarse}$	Standard oxygen transfer efficiency, fine and coarse bubble aeration, %/m
TKN	Nitrogen influent, g/m^3
TSS	Total suspended solids influent in g/m^3
VSS	Volatile suspended solids, g/m^3
X_m	Mixed liquor suspended solids level in membrane tank, g/m^3
Y	Heterotrophic yield coefficient, g VSS/(g BOD)
y_{coarse} and y_{fine}	Coarse and fine bubble aerator depth, m
Y_n	Nitrification yield coefficient, g VSS/(g BOD)
β	Salinity correction factor
φ_{tank}	Membrane packing density in membrane tank, m^2/m^3
μ_m	Maximum heterotrophic specific growth rate, g VSS/(g VSS d)
$\mu_{m,n}$	Maximum nitrification specific growth rate, g VSS/(g VSS d)
ρ_A	Air density, kg/m^3
ω_{fine} and ω_{coarse}	ω-factor for fine and coarse bubble aeration.

A.1 MEMBRANE SYSTEM DESIGN AND OPERATION

Table A-3 details the design and operational parameters for the membrane system, based on average values obtained from 11 HF MBRs for treatment of municipal sewage (Section 5.3*). Key hydraulic design parameters are the net flux (J_{net}), the maximum allowed flux during a limited time period ($J_{net,peak}$) and the peak influent flow (Q_{peak}). It is assumed that Q_{peak} = 2Q (Table A-1), while $J_{net,peak}$ = 140% of J_{net} and can be sustained for the duration of the maximum flow conditions; assumptions will vary according to the anticipated infiltration and trade (i.e. industrial) discharges. On the basis of the stipulated assumptions, the required membrane area is 43% larger than that needed to treat the average flow. Assuming a conservative value for the membrane packing density in the membrane tank (φ_{tank}) allows the minimum membrane tank volume ($V_{m,min}$ – Eq. A.21) to be determined. The larger the discrepancy existing between Q and Q_{peak}, the greater the membrane area demanded and the higher the CAPEX (capital cost) and OPEX (since a larger membrane area will lead to a commensurately greater membrane aeration demand). It may therefore be more economical to install buffering capacity for flow equalisation (Section 7.3.1.2). Coarse bubble membrane aeration contributes to the oxygen required for biological degradation, which requires an adjustment to the biological ("process") aeration.

Table A-3: HF membrane design and operational parameters

Parameter	*Unit*	*Value*	*Ref*	*Parameter*	*Unit*	*Value*	*Ref*
J_{net}[1]	LMH	19.5		SAD_p	Nm^3/m^3	15.4	Table 3.1*
$J_{net,peak}$	LMH	27.3		φ_{tank}	m^2/m^3	45	
K	LMH/bar	104		r_{mr}	-	4	
A_m	m^2	76312	Q/J	X_m	g/m^3	10000	
SAD_m	$Nm^3/(m^2.h)$	0.30		$V_{m,min}$	m^3	1696	Eq A.21

[1]The net flux relates to the operating flux and system downtime incurred by cleaning (Judd and Judd; Table 3.1)

A key constraint of MBRs is the requirement for a relatively short retention time in the membrane tank to limit concentration of the solids and subsequent membrane channel clogging. This demands that the transfer rate between the membrane tank and the biological tank (the return activated sludge or RAS) is in the region of 3-5 times the treated water flow (r_{mr} is assumed to be 4 in Table A-3) – somewhat higher than the equivalent RAS flow for an ASP. It is also relatively high in dissolved oxygen (DO), which makes the anoxic zone to which it may be returned for denitrification less efficient. It is thus usually returned to the aerobic zone of the biotank.

A.2 BIOLOGICAL PROCESS DESIGN AND OXYGEN DEMAND

As with other aerobic treatment processes, the biological component design demands knowledge of appropriate values of the biokinetic constants. A comprehensive listing of biokinetic constants, along with references identifying their origins, is provided in Appendix B*. Since there is no consensus on appropriate biokinetic values for an MBR process, typical values as applicable to ASPs as reported in Metcalf and Eddy (2003) may be used: K_s = 20, k_e = 0.12, μ_m = 6, Y = 0.4, K_n = 0.74, $k_{e,n}$ = 0.08, $\mu_{m,n}$ = 0.75, Y_n = 0.12. These are not necessarily the most appropriate for an MBR, but it has been demonstrated (Verrecht *et al.*, 2010b) that their use appears not to introduce significant error.

The design of the biological system is not greatly different from that of conventional activated sludge (CAS). The simplest procedure employs the stoichiometric correlations introduced in Section 2.2.4* and summarised in Figure 3.4* and Table A-4). However, it is widely recognised (Ekama *et al.*, 1984; Metcalf and Eddy, 2003; Grady *et al.*, 2010) that due to the complex interactions between the different biological processes, a correct process design for nutrient removal can only be obtained through dynamic simulations using the IWA ASM models. A design based on stoichiometry should thus be considered as a useful starting point, which can be refined and optimised by using it as input for the IWA ASM models (Section 3.4*).

The design procedure starts by determining the aerobic SRT ($\theta_{x,aer}$ – Eq. A.2, Table A-2) based on the growth rate of nitrifiers (μ_n – Eq. A.1) required to achieve the desired effluent ammonia concentration N_e. Often, a safety factor *SF* is applied to ensure a sufficiently long SRT to handle TKN peaks in the influent. Temperature has a large impact on the growth rate of nitrifiers and $\theta_{x,aer}$ should be determined for the lowest anticipated temperatures (Ekama *et al.*, 1984). For simplicity, all calculations in the example design have been performed at 20°C, such that no temperature corrections are necessary. Based on $\theta_{x,aer}$, the biomass production in terms of VSS ($M_{X,bio}$ – Eq A.3) can be calculated based on the effluent BOD concentration (S_e – Eq. A.4) and an estimate for NO_x of 80% of the TKN. By adding the influent non-biodegradable VSS (*nbVSS* – Eq. A.7) and inert TSS (*iTSS* – Eq. A.8) the sludge yield based on total solids is obtained ($M_{X,TSS}$ – Eq. A.6), allowing the calculation of the required aerobic volume (V_{aer} – Eq. A.9) and the sludge wastage flow (Q_w – Eq. A.10).

If biological nitrogen removal is required, sufficient anoxic tankage must be added. To determine this volume, an iterative procedure can be followed. The recirculation rate (r_{int} – Eq. A.11) can be calculated based on the effluent nitrate limit (NO_e), which allows calculation of the active fraction of the anoxic biomass ($X_{b,anox}$ – Eq. A.12). $M_{X,bio}$ can be used to calculate a better approximation for the concentration of TKN oxidised to form nitrate (NO_x – Eq. A.5), which allows determination of the nitrate load to the anoxic zone (*NO-loading* – Eq. A.13). Anoxic HRT (HRT_{anox}) has to be estimated, and based on the corresponding food to active biomass concentration in the anoxic zone (F/M_b – Eq. A.14), the specific denitrification rate (SDNR) and denitrification capacity (NO_r – Eq. A.15) can be determined. If NO_r is significantly higher or lower than *NO-loading*, HRT_{anox} has to be adjusted and the procedure reiterated until an adequate value has been found. The total process SRT ($\theta_{x,process}$ – Eq. A.16) can then be calculated. Finally, the oxygen demand required for biological treatment (M_O – Eq. A.17) can be determined. Table A-4 shows the results of this design approach applied to the wastewater characteristics given in Table A-1 at average influent flow *Q*. It is clear that the contribution of *nbVSS* and *iTSS* cannot be ignored for the typical medium strength wastewater of this example, since the sum of these components makes up a large part (63%) of the total sludge production.

Table A-4: Biological operating parameters and design calculations

Aerobic zone				**Anoxic zone**			
Parameter	***Unit***	***Value***	***Eq***	***Parameter***	***Unit***	***Value***	***Eq***
$\theta_{x,aer}$	d	9.3	A.2	$\theta_{x,anox}$	d	3.6	A.16
V_{aer}	m^3	5326	A.9	V_{anox}	m^3	3333	A.25
HRT_{aer}	h	5.1	A.25	HRT_{anox}	h	3.2	
μ_n	gVSS/(gVSS.d)	0.16	A.1	r_{int}	-	1.61	A.11
S_e	$gBOD/m^3$	0.79	A.4	$X_{b,anox}$	g/m^3	1524	A.12
NO_x	$gNO_3\text{-}N/m^3$	31.3	A.5	*NO-loading*	kg/d	1255	A.13
X_{aer}	g/m^3	8000	A.19	X_{anox}	g/m^3	4930	A.18
DO	gO_2/m^3	2		F/M_b	g BOD/(g TSS.d)	1.48	A.14
SF	-	1.5		*SDNR*	g/g.d	0.25	-*
				NO_r	kg/d	1270	A.15
				NO_r/*NO-loading*	-	1.01	
Overall process parameters				**Sludge yield**			
Parameter	***Unit***	***Value***	***Eq***	***Parameter***	***Unit***	***Value***	***Eq***
$\theta_{x,process}$	d	12.9	A.16	$M_{X,bio}$	kg VSS/d	1714	A.3
$V_{process}$	m^3	8659	A.26	$M_{X,TSS}$	kg TSS/d	4599	A.6
$HRT_{process}$	h	8.3	A.27	Q_w	m^3/d	574	A.10
f_{anox}	-	0.38	A.20	Q'_w	m^3/m^3	0.02	A.24
M_O	kgO_2/d	7079	A.17				

Interpolated from Figure 2.18

Even though the above discussion shows that Monod kinetics can be applied to determine biological tank sizes for MBRs, there are generally two other considerations significantly affecting bioreactor sizing which relate to the physical constraints of available equipment. Both are associated with the oxygen demand mass loading demanded by the wastewater to be treated, and the resulting energy input required to meet this oxygen demand. If a specified oxygen supply is needed to meet the process oxygen requirement, and the oxygen transfer equipment can transfer that oxygen with a specified efficiency (kg O_2/kWh), then the energy required is determined by this. This sets the upper and lower limits on the bioreactor volume, viz. (Grady *et al.*, 2010):

a) The maximum bioreactor volume that can be supported by the energy input is associated with the requirement for the oxygen transfer device not only used to transfer oxygen but also to keep solids in suspension. However, since volume is typically minimised in an MBR this is generally not a consideration.
b) The minimum bioreactor volume is determined by the constraint on the oxygen transfer device with respect to the maximum volumetric energy input. When volumetric energy input is excessive, the shear rate becomes high enough to produce deflocculation which can then promote membrane fouling. Also, as a rule of thumb, it is generally accepted that any oxygen transfer device is physically limited regarding the maximum achievable volumetric oxygen transfer rate in mgO_2/(l.hr). For air-based systems (as in an MBR) the practical limit on a sustained basis is generally about 100 mgO_2/(L.hr), peaking as needed to 150 mgO_2/(L.hr). This constraint determines the actual size of many MBRs, as membranes permit higher MLSS concentrations to be maintained than in CAS systems. When this is applied to the design example, the oxygen demand is 295 kg/h (Eq. A.17). This implies a minimum aerobic tank HRT of 2.9 h versus the calculated value (HRT_{aer}) of 5.1h, which can therefore be considered sufficient to overcome the physical constraint of the oxygen transfer devices.

Another consideration for any ASP (including MBRs) is the trade-off between the size of the bioreactor and the solid-liquid separation specification – or the membrane area, in the case of an MBR. This relates to the flux that can be sustained at the solids loading rate to which the flux pertains. This in turn is dictated by the effectiveness of the crossflow or air scour in sweeping the solids away from the surface through the application of shear. If the applied shear is insufficient to remove the colloidal and particulate material from the membrane:solution interface they can deposit and accumulate on the membrane surface and produce fouling and clogging. The solids loading rate is simply the flux multiplied by the solids concentration (the MLSS). Hence reducing the MLSS would be expected to have a significant impact on fouling and clogging. In practice the impact is minor, but conservative sizing of the biotank is nonetheless beneficial in providing buffering of hydraulic surges.

Unlike CASPs, the membrane separation process generates a significant MLSS concentration difference between the membrane and respective anoxic and aerobic biological tanks/zones, as determined by the recirculation ratios r_{int} and r_{mr} respectively. In the design example, a typical maximum MLSS concentration in the membrane tank X_m of 10 kg/m^3 is combined with an assumed membrane recirculation ratio r_{mr} of 4, resulting in a MLSS concentration in the aerobic zone (X_{aer} – Eq. A.19) of 8 g/L. A recirculation rate r_{int} of 1.61 (to obtain an effluent NO_e concentration of 12 mg/L – Eq. A.11) leads to an MLSS concentration in the anoxic zone X_{anox} of 4,930 mg/L (Eq. A.18). The MLSS gradient has to be incorporated in the design procedure to allow correct tank sizing and estimation of required aeration capacity, since the MLSS concentration impacts on the α-factor (Eq. A.28).

A.3 AERATION SYSTEM DESIGN

The design of the aeration system is one of the most important differences between an iMBR and a CASP, since iMBRs typically have both membrane and biology aeration. It is thus necessary to determine how much oxygen arising from membrane aeration contributes to the total oxygen required for carbonaceous degradation and nutrient removal (M_O – Eq. A.17). Dissolved oxygen (DO) concentrations in the membrane tanks can be high due to the prevailing vigorous membrane aeration, and therefore some of it will be available for biological degradation. However, since it is not possible to incorporate this in a simple stoichiometric design, for the purposes of the design calculation it is assumed that all DO from membrane aeration contributes to the total O_2 requirement for biological degradation. This is likely to be an overestimation, and a more accurate representation requires the use of the IWA ASM models under dynamic conditions (Section 3.4*).

Table A-5: Aeration system operating parameters and design calculations

Parameter	*Unit*	*Biology*	*Membrane*	*Eq*
Diffuser type		Fine bubble	Coarse bubble	
Standard oxygen transfer efficiency, $SOTE_x$	%/m	0.05	0.02	
Air density, ρ_A	kg/m^3	1.2	1.2	
Correction factor exponent, ω_x	-	0.084	0.084	
Mass % oxygen in air, $O_{A,m}$	%	23.2	23.2	
Mass transfer correction, solids, α	-	0.51	0.43	A.28
Mass transfer correction, salinity, β	-	0.95	0.95	
Mass transfer correction, temperature, Φ	-	1	1	A.29
Aerator depth, y_x	m	5	2.3	
Air flow rate, membrane tank, $Q_{A,m}$	Nm^3/h	-	22893	A.30
O_2 transferred by membrane aeration, M_m	kg/d	-	2886	A.31
O_2 to be provided by biology aeration, M_b	kg/d	4193	-	A.17
Air flow rate, biotank, $Q_{A,b}$	Nm^3/h	-	5174	A.32

Table A.30 illustrates the design and operating parameters of the aeration system for both fine and coarse bubble aeration. The coarse bubble membrane aeration is evidently much less efficient than the biology aeration for providing dissolved oxygen, as indicated by the much higher airflow rate ($Q_{A,m}$ – Eq. A.30-A.31) required to transfer oxygen into the biomass. This can be attributed to several factors in Eq. A.31: a) the higher MLSS concentration and its impact on the α-factor (Eq. A.28), b) a lower aerator depth y (determined by the membrane module length and placement), and c) a lower oxygen transfer efficiency $SOTE_{coarse}$, all compared to the values for fine bubble aeration.

A.4 REFERENCES

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