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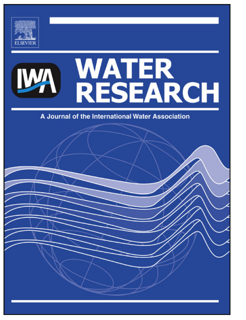

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Model-based energy optimisation of a small-scale decentralised membrane bioreactor for urban reuse

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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.

Keywords: Energy, Reuse, Model-based optimisation, Scenario analysis, MBR calibration

Symbols and abbreviations

AOTR	Actual oxygen transfer rate in $gO_2 \cdot d^{-1}$
ASM2d	Activated sludge model no.2d
BOD_5	5 day biological oxygen demand in $mg \cdot l^{-1}$
BOD_f	5 day biological oxygen demand of a sample filtered through 0.45 µm, in $mg \cdot l^{-1}$
b_{PAO}	Rate constant for lysis of X_{PAO} in d^{-1}
CAS	Conventional activated sludge
COD	Chemical oxygen demand in $mg \cdot l^{-1}$
COD_f	Chemical oxygen demand of a sample filtered through 0.45 µm in $mg \cdot l^{-1}$
$C_{rsat\_average}$	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 ℃ and 1 atm
CSTR	Continuously stirred tank reactor
C_{tank}	Actual oxygen concentration in the aeration tank in $gO_2 \cdot m^{-3}$
DO	Dissolved oxygen in $mgO_2 \cdot l^{-1}$
F	Correction factor for fouling of the air diffusers (1 for clean diffusers)

F_{coarse}	Correction factor for fouling of the coarse bubble air diffusers (1 for clean diffusers)
F_{fine}	Correction factor for fouling of the fine bubble air diffusers (1 for clean diffusers)
HRT	Hydraulic retention time in hours
K_O	Half saturation coefficient for oxygen, in $mgO_2 \cdot l^{-1}$
MBR	Membrane bioreactor
$MLSS$	Mixed liquor suspended solids in $mg \cdot l^{-1}$
NH_4-N	Ammonia-nitrogen in $mgN \cdot l^{-1}$
NO_2-N	Nitrite-nitrogen in $mgN \cdot l^{-1}$
NO_3-N	Nitrate-nitrogen in $mgN \cdot l^{-1}$
O_{air}	Fraction of oxygen in the air in %
ON	Organic nitrogen in $mgN \cdot l^{-1}$
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}
PO_4-P	Ortho-phospate in $mgP \cdot l^{-1}$
Q_{air}	Airflow rate in $Nm^3 \cdot h^{-1}$
$Q_{air,coarse}$	Coarse bubble airflow rate in $Nm^3 \cdot h^{-1}$
$Q_{air,fine}$	Fine bubble airflow rate in $Nm^3 \cdot h^{-1}$
S_A	Fermentation products, considered to be acetate, in $mgCOD \cdot l^{-1}$
SCA	Scenario analysis
S_F	Fermentable, readily biodegradable organic substrates in $mgCOD \cdot l^{-1}$
S_I	Inert soluble organic material in $mgCOD \cdot l^{-1}$
S_{NH4}	Ammonium plus ammonia nitrogen in $mgN \cdot l^{-1}$
$SOTR$	Standard oxygen transfer rate in $gO_2 \cdot d^{-1}$
S_{PO4}	Inorganic soluble phosphorus, primarily ortho-phosphates in $mgP \cdot l^{-1}$
SRT	Solids retention time in days
SS	Suspended solids in $mg \cdot l^{-1}$
T	Temperature of the mixed liquor in ℃
$tCOD$	Total COD in $mg \cdot l^{-1}$
TKN	Total Kjeldahl nitrogen in $mgN \cdot l^{-1}$
TN	Total nitrogen in $mgN \cdot l^{-1}$
TON	Total oxidised nitrogen in $mgN \cdot l^{-1}$
TP	Total phosphorous in $mgP \cdot l^{-1}$
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_{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, dimensionless
μ_{PAO}	Maximum growth rate of X_{PAO} in d^{-1}
ρ_{air}	Density of air at standard conditions in $kg \cdot m^{-3}$
φ	Temperature correction factor for oxygen transfer
ω	α-factor exponent coefficient, dimensionless

1 Introduction

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 (Sperandio 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.

2 Materials and Methods

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 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 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.

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 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 1).

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 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 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 2).

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 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 ℃. A number of process upsets occurred ove r 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:

$$SOTR = k_L a \cdot \left(DO - DO_{sat}\right) \cdot V \tag{1}$$

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*:

$$AOTR = SOTR \cdot \frac{(\beta \cdot C_{rsat\_average} - C_{\tan k})}{C_{ssat}} \cdot \varphi^{(T-20)} \cdot \alpha \cdot F \tag{2}$$

$$SOTR = 24 \cdot Q_{air} \cdot \rho_{air} \cdot OTE \cdot y \cdot O_{air} / 10000 \tag{3}$$

$$\alpha = e^{-\omega \cdot MLSS} \tag{4}$$

In this model the influence of MLSS concentration on the AOTR is accounted for through the α-factor (Eq. 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).

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 $kWh \cdot (Nm^3)^{-1}$. 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.

3 Results and discussion

3.1 Hydraulic profile

Figure 3 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.

3.2 Model calibration

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 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 4).

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 4 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 4). Predicted PO_4-P concentrations show acceptable values and dynamic behaviour (Figure 5) 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.

3.2.3 On/off controller for aeration model for predictive $Q_{air,fine}$ values in energy profiling

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 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 4 and Figure **5**). Any differences can be attributed to slight deviations from reality using the implemented on/off controller.

3.3 Scenario analysis

The evolution of biological aeration demand and maximum effluent NH_4-N concentration as a function of the SRT (Figure 6) demonstrates that lowering the SRT by increasing the wastage rate has a beneficial effect on demand for $Q_{air,fine}$. However, Figure 6 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 7), 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 8 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 $mg \cdot l^{-1}$ 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.

It is acknowledged that over the range of operating conditions studied in the SCA, the biological processes and kinetics may change. For instance, Sperandio 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 3.4).

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 $Nm^3 \cdot h^{-1}$.

3.4 Model application for optimisation

Results from the scenario analysis were used in the selection of better operational parameter values (Table 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 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).

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.

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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Table 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

Table 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 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

Table 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·(g COD)$^{-1}$
NH_4-N	g·m^{-3}	0.07	0.337	0.338
NO_3-N	g·m^{-3}	21.4	21.9	21.68
PO_4-P	g·m^{-3}	4.35	9.65	5.18
MLSS	g·m^{-3}	7,832	6,584	7,869

Table 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 zone, permeate pump and recirculation pump)

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

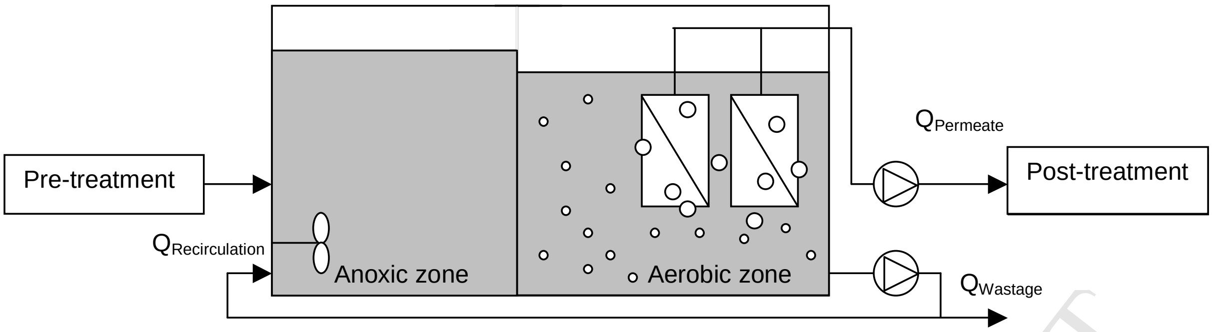


Figure 1: Schematic overview of the wastewater recycling plant

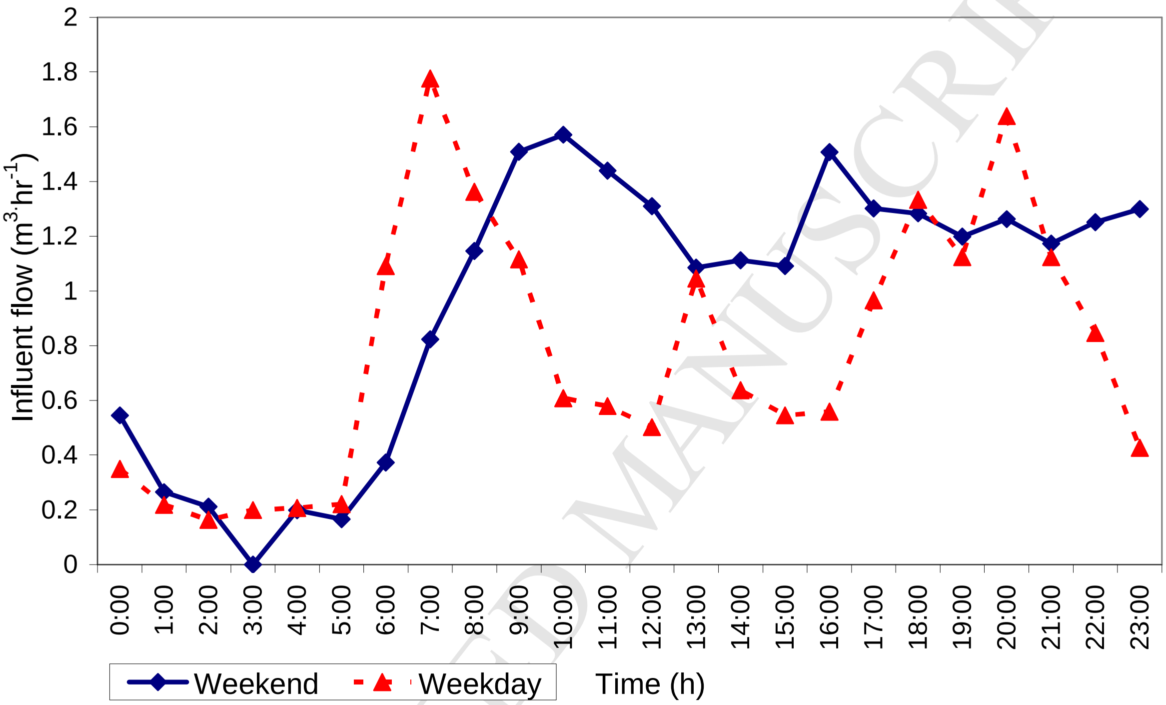


Figure 2: Comparison of typical diurnal flow profiles during a weekday and a day in the weekend

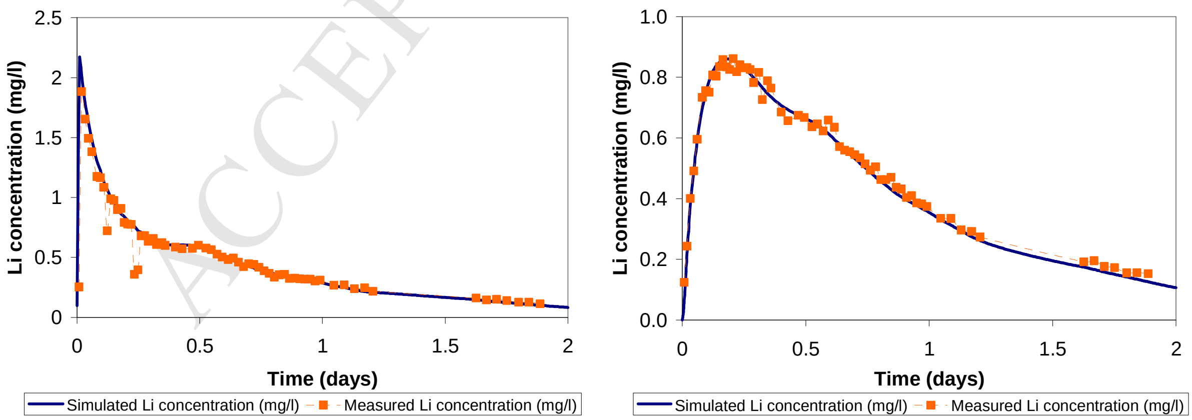


Figure 3: Predicted and actual Li concentrations in (a) anoxic, and (b) aerobic tanks during the tracer test

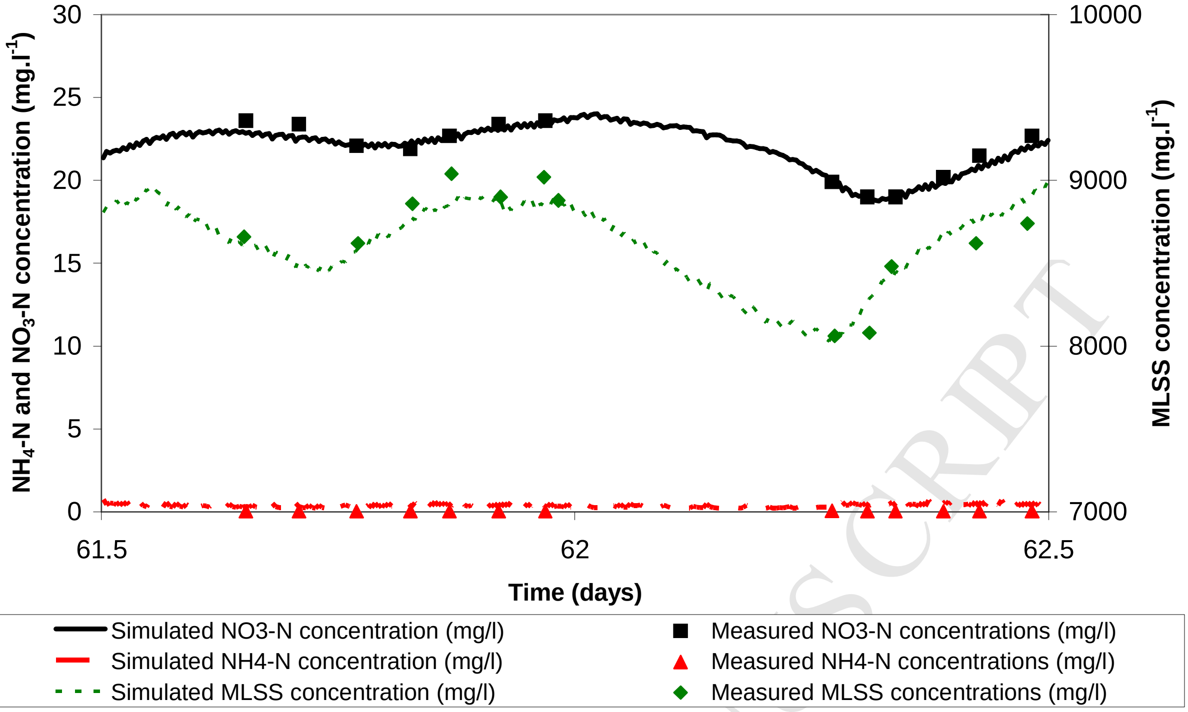


Figure 4: Simulated and recorded NH_4-N, NO_3-N and MLSS concentrations using measured $Q_{air,fine}$, averaged per 15 minute interval, as input

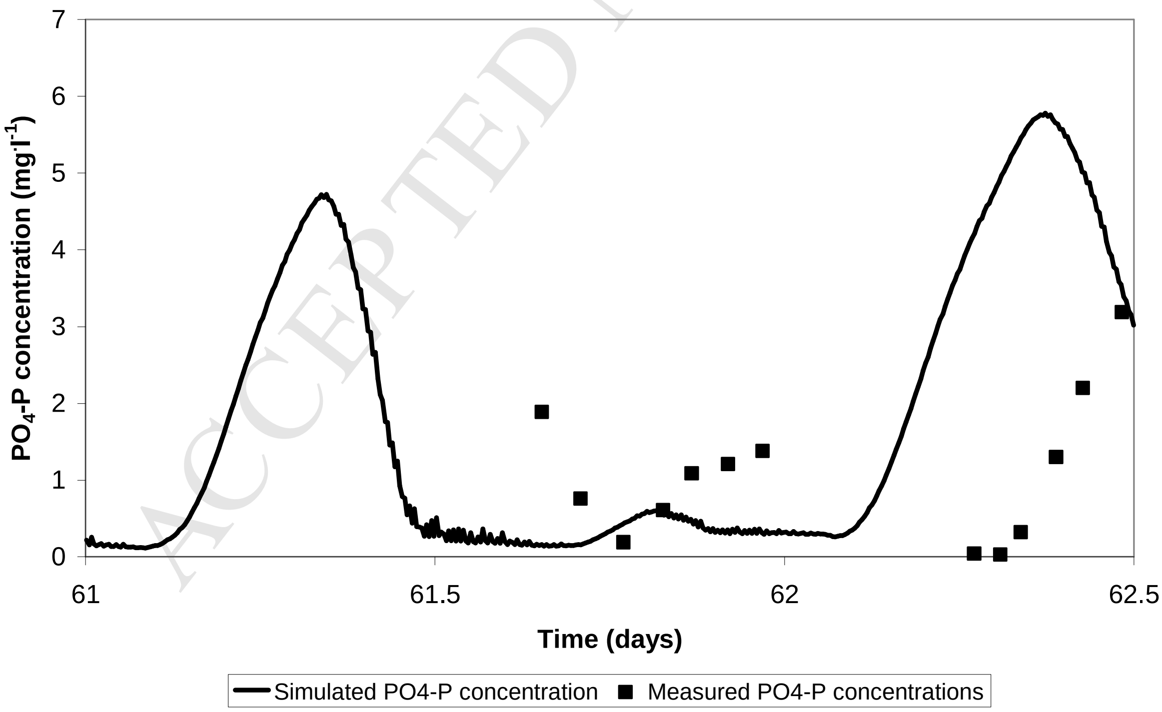
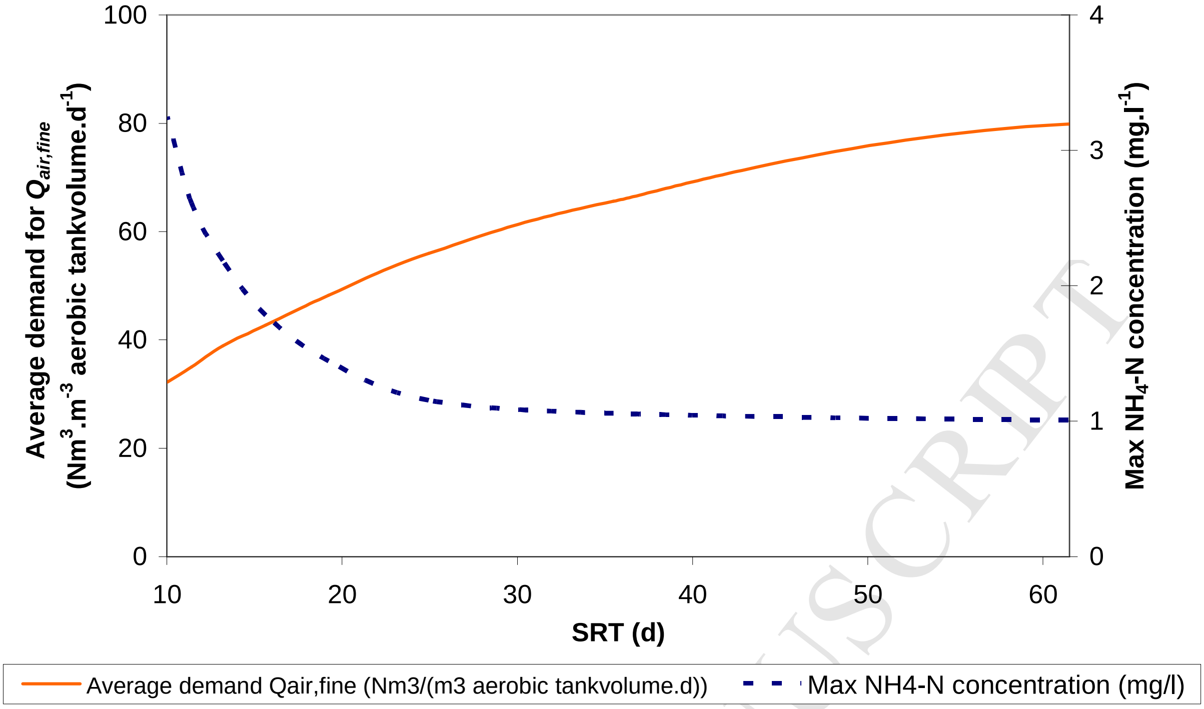


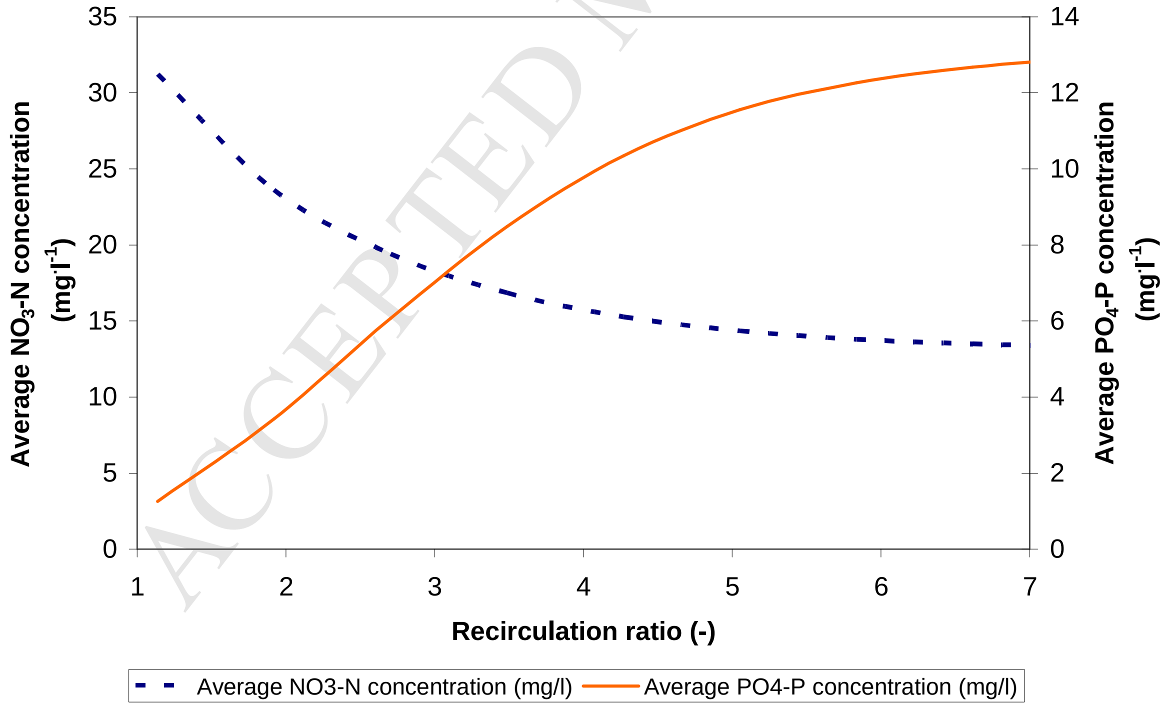
Figure 5: Simulated and measured PO_4-P values using measured measured $Q_{air,fine}$, averaged per 15 minute interval, as input



592

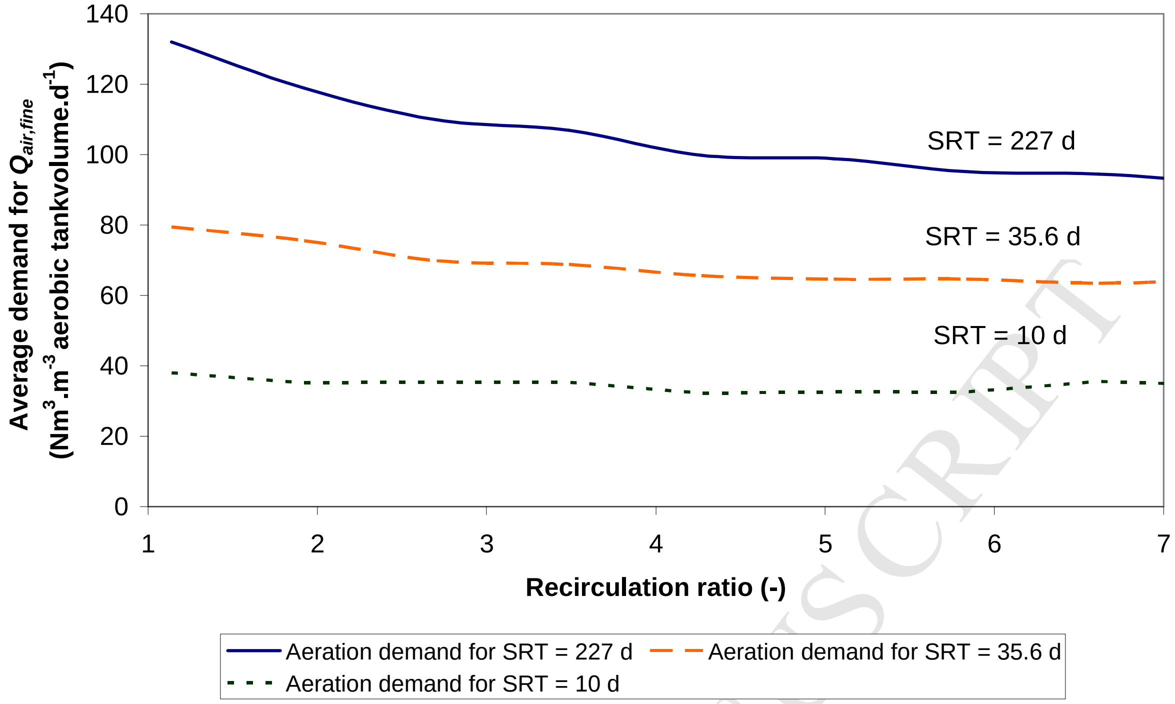
593 Figure 6: Influence of wastage rate on the total demand for biology aeration ($Q_{air,fine}$) and the maximum
594 occurring effluent NH_4-N concentration during the 35-day simulation ($Q_{air,coarse}$ = 42 $Nm^3 \cdot h^{-1}$; DO
595 setpoint = 1.25 $mg \cdot l^{-1}$; recirculation flow = 108 $m^3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

596



597

598 Figure 7: Influence of recirculation flow rate on the average effluent NO_3-N and PO_4-P concentrations
599 during the 35-day simulation ($Q_{air,coarse}$ = 42 $Nm^3 \cdot h^{-1}$; DO setpoint = 1.25 $mg \cdot l^{-1}$)



600

601 Figure 8: Influence of recirculation rate and SRT on total demand for $Q_{air,fine}$ ($Q_{air,coarse}$ = 42 $Nm^3 \cdot h^{-1}$; DO
602 setpoint = 1.25 $mg \cdot l^{-1}$)

603