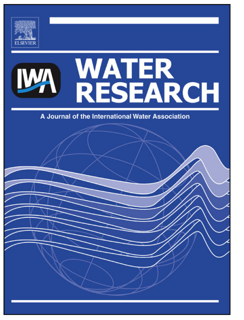
Water Research, Volume 44, Issue 18, October 2010, Pages 5274-5283

Accepted Manuscript



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PII: S0043-1354(10)00441-0

DOI: 10.1016/j.watres.2010.06.054

Reference: WR 8109

To appear in: *Water Research*

Received Date: 31 March 2010

Revised Date: 18 June 2010

Accepted Date: 22 June 2010

Please cite this article as: Verrecht, B., Maere, T., Nopens, I., Brepols, C., Judd, S. The cost of a large-scale hollow fibre MBR, Water Research (2010), doi: 10.1016/j.watres.2010.06.054

The cost of a large-scale hollow fibre MBR

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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.

Keywords *Membrane bioreactor, cost sensitivity, life cycle, biokinetics, aeration*

List of symbols and abbreviations

ASM1	*Activated sludge model no. 1*
BSM1 LT	*Long term benchmark simulation model no. 1*
Capex	*Capital expenditures, in Euro*
CAS	*Conventional activated sludge plant*
CIP	*Cleaning in place*
COP	*Cleaning out of place*
d_f	*Hollow fibre outside diameter, in m*
EQI	*Effluent quality index, in kg PU.*d^{-1}
HF	*Hollow fibre*
HRT	*Hydraulic retention time, in h*
i	*Discount rate, in %*
J_0	*Intercept of the J vs. U curve, in* $l.m^{-2}.h^{-1}$
J_x	*Flux, in* $l.m^{-2}.h^{-1}$
$L_{membrane}$	*Length of the membrane module, in m*
L_{tank}	*Tank length, in m*
m	*Slope of the J vs U curve*
MBR	*Membrane bioreactor*
NPV	*Net present value, in Euro*
Opex	*Operational expenditures, in* $Euro.year^{-1}$
PE_{sludge}	*Pumping energy required per unit of sludge, in* $kWh.m^{-3}$
P_{sludge}	*Power required for sludge pumping, in kW*
PU_x	*Pollution unit for effluent component x, in* $kg.d^{-1}$
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_W	*Wastage flow, in* $m^3.d^{-1}$
SAD_m	*Specific aeration demand per unit of membrane area, in* $Nm^3.m^{-2}.h^{-1}$
SAD_p	*Specific aeration demand per unit of permeate, dimensionless*
SRT	*Solids retention time, in d*
U	*In-module air upflow gas velocity, in* $m.s^{-1}$
W_{tank}	*Tank width, in m*
Δh	*Head loss, in m*
α	*Clean-to-process water correction factor for oxygen transfer*
β_x	*Weighting factor for effluent component x, dimensionless*
ξ_B	*Blower efficiency, dimensionless*
ξ_p	*Pump efficiency, dimensionless*
ρ_{sludge}	*Sludge density, in* $kg.m^{-3}$
φ	*Module packing density, in* m^{-1}

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, Fig. 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.

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

2. Materials and methods

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.

2.2 Biological process model

Figure 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 througout. Simulations were performed using the WEST® simulation and modelling platform (Vanhooren *et al.*, 2003).

Insert Figure 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.

2.3 Capital costs

To evaluate capital investment costs, pricing information (Table 2) 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., 2010).

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 ^oC.

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.

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.

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 (Energyref - Christoph??).

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}) (Insert Figure **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.

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.

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.

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).

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:

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

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).

3. Results and discussion

3.1 Effect of contingency: changes in feedwater flow and strength

3.1.1 Hybrid plant vs. plant designed for maximum flow

Table 1 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.

Insert Table 1: 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.

Figure 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.

Insert Figure 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 = 1.08 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).

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 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 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 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 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 5).

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

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 (Ref for industry land – Christoph??). 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.

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

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 2). Table 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.

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

Insert Table 2: MBR design parameters and base case costs for the study of operational and design parameters

Insert Table 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 2

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 1), 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 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 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 3), but a large impact on EQI. Increasing the anoxic fraction from 30 to 50% improves EQI by 18% due to improved denitrification.

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 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:

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

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

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:

$$SAD_p = \frac{SAD_m}{J} \qquad (4)$$

where a minimum SAD_p is required to maintain a sustainable flux (2), but increasing SAD_p beyond $SAD_{m,max}$ has no impact on the sustainable flux (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), 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 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.

Insert Figure 7: Influence of SAD_p on net present value for a range of sustainable fluxes

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%.

Membrane replacement and cost As shown in Table 1, 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).

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:

1. 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.
2. 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.
3. 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.
4. 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.

5. 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%.
6. 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%.
7. 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.

Acknowledgements

The authors are grateful to MOSTforWATER N.V. (Kortrijk, Belgium) for providing the WEST® modelling software, and would also like to thank Thames Water for the resources provided by them for this paper. Thomas Maere is supported by the Institute for Encouragement of Innovation by means of Science and Technology in Flanders (IWT).

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Table 1: 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**

Table 2: MBR design parameters and base case costs for the study of operational and design parameters

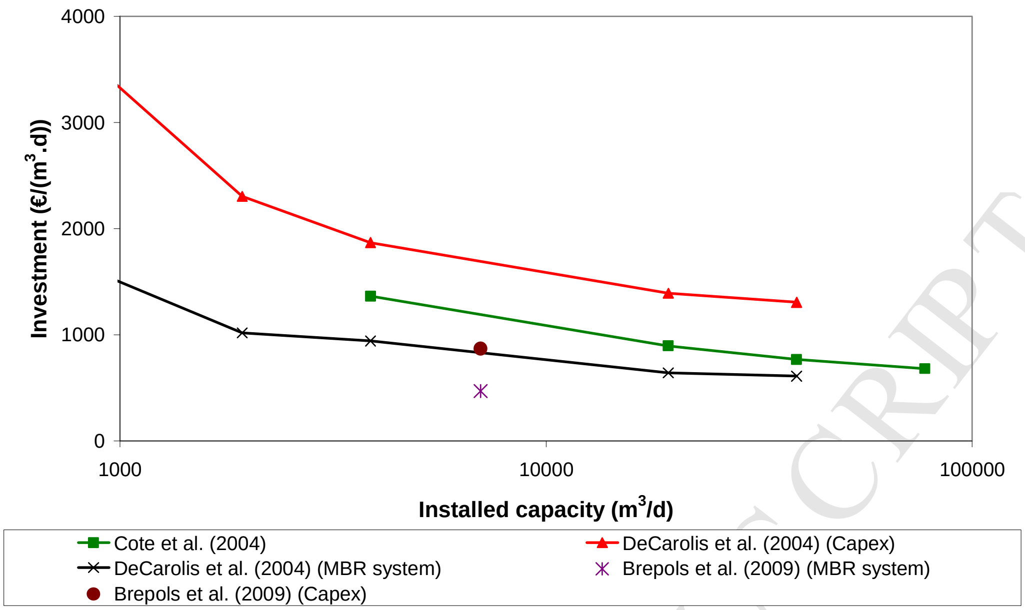
Parameter	Units	Value	Reference		Units	Value
Assumptions for capex calculation				*Base design, EQI and NPV*		
Membrane cost	€.m^{-2}	50	Judd & Judd, 2010	*Design capacity*	$m^3.d^{-1}$	30,416
Tank civil cost	€.m^{-3} tank volume	220	Brepols, 2010b	*Maximum plant capacity**	$m^3.d^{-1}$	42,582
Screens – 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, 2010b			
Biomass recirculation pumps	€.$m^{-3}.h^{-1}$ capacity	12.1	Manufacturers; Brepols, 2010b	*Buffer tank size*	m^3	14,530
Mixing equipment	€.m^{-3} tank volume	27.8	Brepols, 2010b	*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	- ref Christophe			
Sludge treatment cost	€.ton^{-1} of DS	150	-	*Effluent quality index*	kg PU.d^{-1}	5,430
Citric acid 50%	€/ton	760	Brepols, 2010b	*NH_4-N*	$mg.l^{-1}$	0.52
NaOCl 14%	€/m^3	254	Brepols, 2010b	*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*	Million Euro (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

Table 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 2

	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%

1 

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

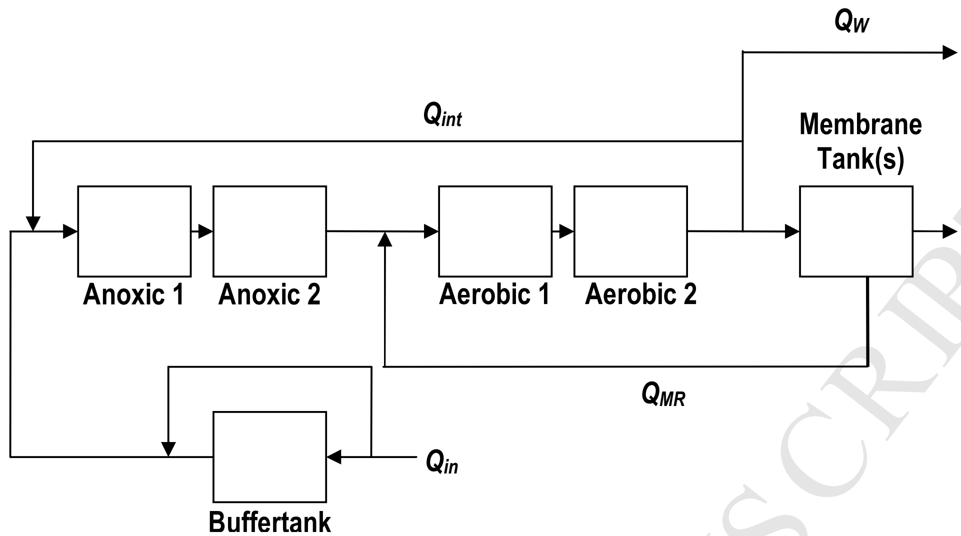


Figure 2: Schematic overview of the generic nitrifying/denitrifying MBR design

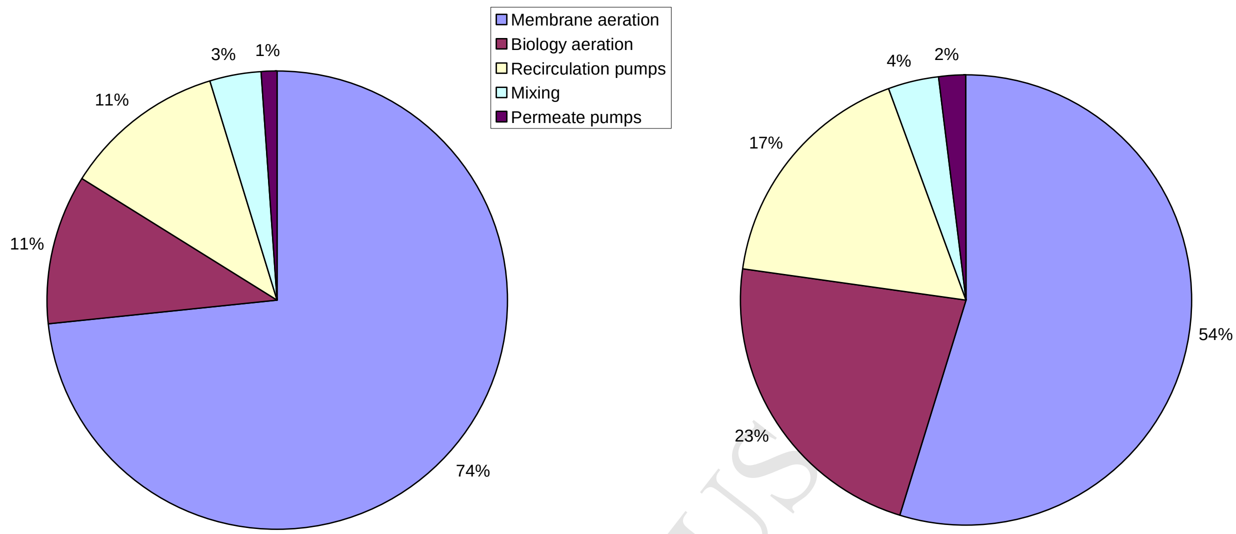


Figure 3: Breakdown of energy demand for a) the MBR part of a hybrid plant (Average total energy demand = 0.7 kWh.m^{-3}) and b) a plant designed for maximum flow (Average total energy demand = 1.07 kWh.m^{-3})

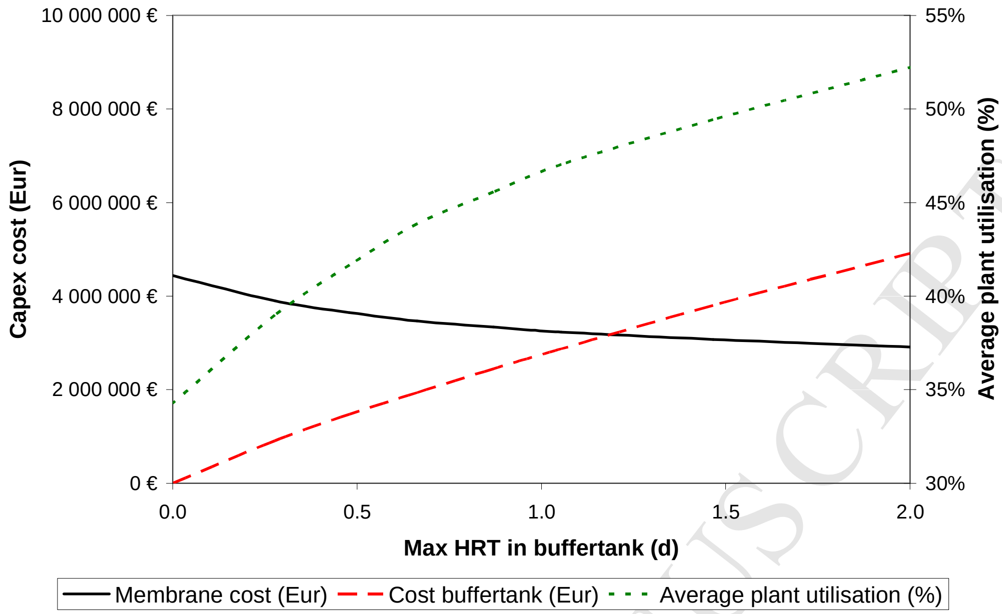


Figure 4: Influence of buffer tank size on plant utilisation, and a comparison of the resulting costs for membranes and buffer tank

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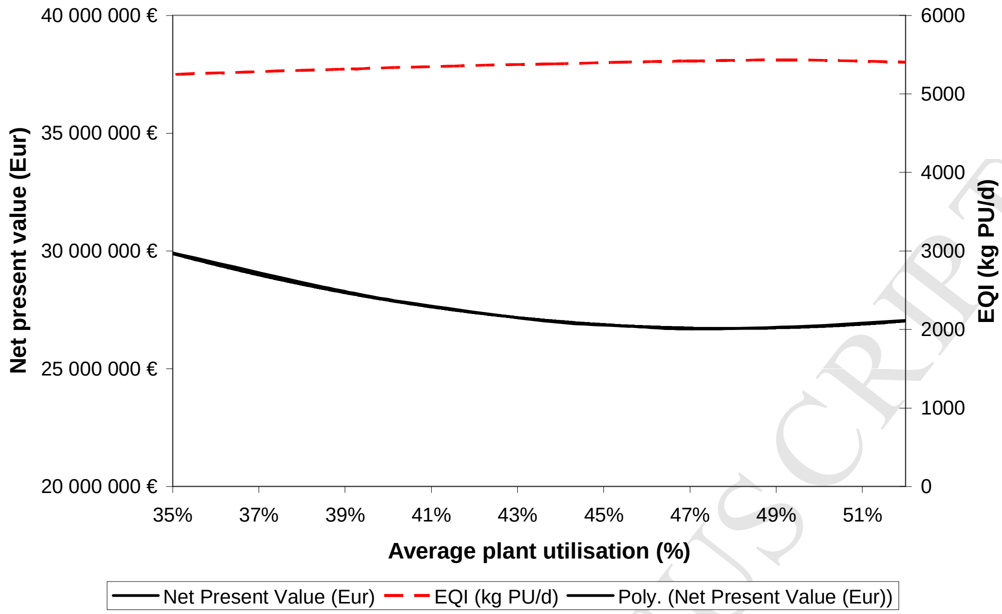

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18 Figure 5: Influence of average plant utilisation on net present value and effluent quality index

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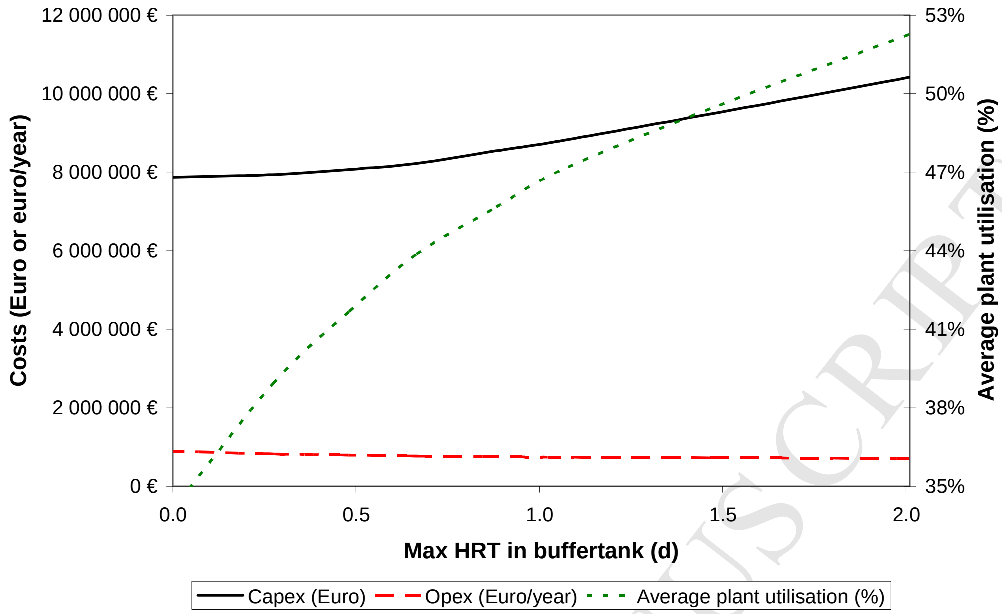

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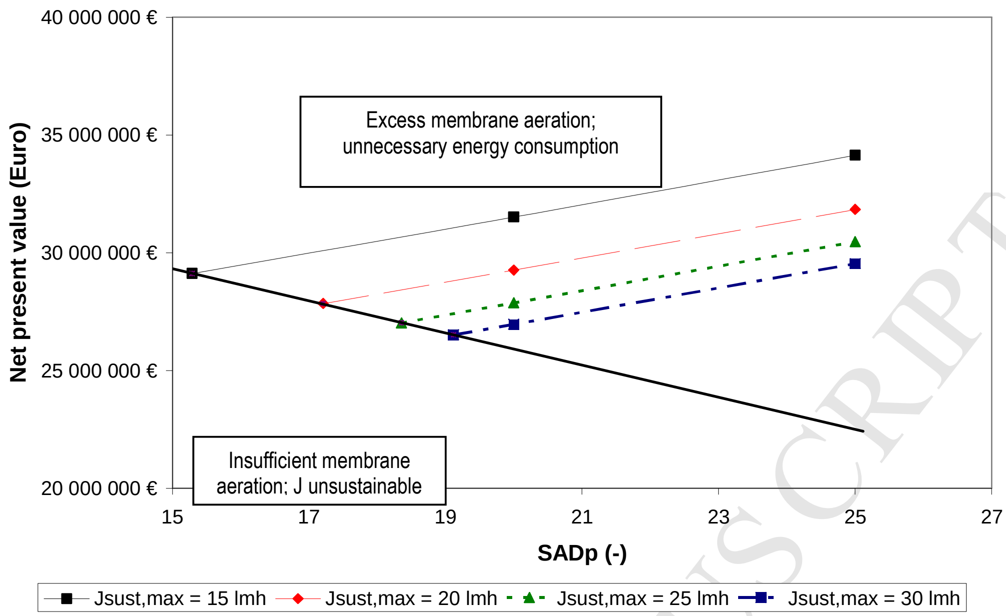
21 Figure 6: Influence of size of buffer tank on capex and opex



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23 Figure 7: Influence of SAD_p on net present value for a range of sustainable fluxes

The cost of a large-scale hollow fibre MBR

Verrecht, Bart

2010-10

Bart Verrecht, Thomas Maere, Ingmar Nopens, Christoph Brepols and Simon Judd, The cost of a large-scale hollow fibre MBR, Water Research, Volume 44, Issue 18, October 2010, Pages 5274-5283

http://dx.doi.org/10.1016/j.watres.2010.06.054