Accepted Manuscript

Title: The cost of a large-scale hollow fibre MBR

Authors: Bart Verrecht, Thomas Maere, Ingmar Nopens, Christoph Brepols, Simon Judd

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

This is a PDF file of an unedited manuscript that has been accepted for publication. As a service to our customers we are providing this early version of the manuscript. The manuscript will undergo copyediting, typesetting, and review of the resulting proof before it is published in its final form. Please note that during the production process errors may be discovered which could affect the content, and all legal disclaimers that apply to the journal pertain.



1 The cost of a large-scale hollow fibre MBR

2

Bart Verrecht^{*}, Thomas Maere^{**}, Ingmar Nopens^{**}, Christoph Brepols^{***}, and Simon
 Judd^{1,*}

5 *Centre for Water Science, Cranfield University, SIMS, Building 52, Cranfield,

6 Bedfordshire MK43 0AL, UK

7 **BIOMATH, Department of Applied Mathematics, Biometrics and Process Control,

8 Ghent University, Coupure Links 653, B-9000 Gent, BE

9 *** Erftverband, Am Erftverband 6, 50126 Bergheim/Erft, Germany

10 ¹ Corresponding author: <u>s.j.judd@cranfield.ac.uk</u>

11

12

13 Abstract

14 A cost sensitivity analysis was carried out for a full scale hollow fibre membrane 15 bioreactor to quantify the effect of design choices and operational parameters on 16 cost. Different options were subjected to a long-term dynamic influent profile and 17 evaluated using ASM1 for effluent quality, aeration requirements and sludge 18 production. The results were used to calculate a net present value (NPV), 19 incorporating both capital expenditure (capex), based on costs obtained from 20 equipment manufacturers and full scale plants, and operating expenditure (opex), 21 accounting for energy demand, sludge production and chemical cleaning costs.

22

23 Results show that the amount of contingency built in to cope with changes in 24 feedwater flow has a large impact on NPV. Deviation from a constant daily flow 25 increases NPV as mean plant utilisation decreases. Conversely, adding a buffer tank 26 reduces NPV, since less membrane surface is required when average plant 27 utilisation increases. Membrane cost and lifetime is decisive in determining NPV: an 28 increased membrane replacement interval from 5 to 10 years reduces NPV by 19%. 29 Operation at higher SRT increases the NPV, since the reduced costs for sludge 30 treatment are offset by correspondingly higher aeration costs at higher MLSS levels, 31 though the analysis is very sensitive to sludge treatment costs. A higher sustainable 32 flux demands greater membrane aeration, but the subsequent opex increase is offset 33 by the reduced membrane area and the corresponding lower capex. 34

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

WR14476

37 List of symbols and abbreviations

38	ASM1	Activated sludge model no. 1
39	BSM1 LT	Long term benchmark simulation model no. 1
40	Capex	Capital expenditures, in Euro
41	CAS	Conventional activated sludge plant
42	CIP	Cleaning in place
43	COP	Cleaning out of place
44	d _f	Hollow fibre outside diameter. in m
45	EQI	Effluent quality index, in kg PU.d ¹
46	HF	Hollow fibre
47	HRT	Hydraulic retention time, in h
48	i	Discount rate. in %
49	\mathbf{J}_0	Intercept of the J vs. U curve, in $I.m^{-2}.h^{-1}$
50	J.	Flux, in Lm^2 , h^1
51	Lmombrana	Length of the membrane module, in m
52	Ltonk	Tank length, in m
53	m	Slope of the J vs U curve
54	MBR	Membrane bioreactor
55	NPV	Net present value, in Euro
56	Opex	Operational expenditures, in Euro, vear ¹
57	PEsludge	Pumping energy required per unit of sludge, in kWh.m ⁻³
58	Psludge	Power required for sludge pumping, in kW
59	PU	Pollution unit for effluent component x, in ka, d^{1}
60	QF	Effluent flow, in $m^3 d^1$
61	Q,	Influent flow, in m^3 , d^1
62	Q _{MR}	Membrane recirculation flow, in m^3 , d^{-1}
63	Q _{NR}	Nitrate recirculation flow, in m^3 , d^1
64	QW	Wastage flow, in m^3 , d^1
65	SAD _m	Specific aeration demand per unit of membrane area. in Nm ³ .m ⁻² .h ⁻¹
66	SAD	Specific aeration demand per unit of permeate, dimensionless
67	SRT	Solids retention time. in d
68	Ū	In-module air upflow gas velocity, in m.s ⁻¹
69	Wtank	Tank width, in m
70	Δh	Head loss, in m
71	α	Clean-to-process water correction factor for oxygen transfer
72	β _x	Weighting factor for effluent component x, dimensionless
73	ξ_{B}	Blower efficiency, dimensionless
74	$\tilde{\xi}_{p}$	Pump efficiency, dimensionless
75	P _{sludae}	Sludge density, in kg.m ⁻³
76	φ	Module packing density, in m ⁻¹
	C	
	Y	

Verrecht et al, The cost of a large-scale hollow fibre MBR

77 **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.

85

86 An overview of literature investment cost data (McAdam and Judd, 2006, Fig. 1) over 87 a range of reported plant sizes reveals costs to increase exponentially with 88 decreasing plant size, and that a large variation in required capex arises according to 89 assumptions made and costs included. DeCarolis et al. (2004) provided a 90 comprehensive overview of costing data in terms of capex and opex, both for the 91 MBR system alone (based on quotes from four leading suppliers), and for the 92 complete installation (based on preliminary plant design and assumptions about the 93 location-specific contribution of land costs, contractor overheads, engineering, legal 94 costs, etc). Côté et al. (2004) compared capex and opex of an MBR to a conventional 95 activated sludge (CAS) system with tertiary filtration for effluent reuse purposes, 96 demonstrating an integrated MBR to be less expensive than a combination of CAS 97 and tertiary filtration - a conclusion subsequently corroborated by Brepols et al. 98 (2009) for German wastewater plants. The latter authors showed energy demand to 99 increase for plants with significant in-built contingency, since the average plant 100 utilisation is low. This has recently led Maurer (2009) to introduce the specific net 101 value (SNPV), which takes into account the average plant utilisation over its lifetime 102 and so reflects the cost per service unit.

103

104 Notwithstanding the above, no in-depth analysis has been produced quantifying the 105 impact of key design and operating parameters on both capex and opex over the 106 lifetime of an installation. This paper aims to determine both absolute values of capex 107 and opex and their sensitivity to various influencing parameters such as contingency 108 (to provide robustness to changes in feedwater flow and composition), membrane 109 replacement, net flux, and hydraulic and solids residence time (HRT and SRT). The 110 approach taken is to evaluate the impact of representative dynamic flow and load 111 conditions using ASM1 (Henze et al., 2000) on effluent quality, sludge production and 112 aeration demand, based on various MBR process designs. Dynamic simulation

Verrecht et al, The cost of a large-scale hollow fibre MBR

results can then be used as input for specific cost models for both capex and opex,

114 generated using representative heuristic and empirical available cost data. Opex for

- energy demand (Maere et al., 2009), added to sludge treatment and disposal and
- 116 chemical cleaning costs, can then be combined with capex to produce the NPV. This
- 117 then allows the impact of design and operation parameter selection to be quantified.

118Insert Figure 1: Specific investment vs. installed plant capacity, based on literature119data (adapted from McAdam and Judd, 2006)

120

121 **2. Materials and methods**

122 **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 m³.d⁻¹, while the maximum instantaneous flow was 59,580 m³.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.

129 **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).

137

138 Insert Figure 2: Schematic overview of the generic nitrifying/denitrifying MBR design 139

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

Verrecht et al, The cost of a large-scale hollow fibre MBR

146	the average feed flow: $Q_{MR} = 83,404 \text{ m}^3.\text{d}^{-1}$. Internal recirculation from the aerobic
147	tanks to the anoxic tanks was three times the feed flow $Q_{int} = 62,553 \text{ m}^3.\text{d}^{-1}$.
148	
149	The membrane tank volume, included in the total aerobic volume, was calculated
150	based on a conservative packing density of 45 $\ensuremath{m^2}$ membrane area per $\ensuremath{m^3}$ of tank
151	volume, which is at the lower end of values reported (Judd and Judd, 2010). The
152	number of membrane tanks required was based on the design parameters for a large
153	scale plant (Brepols et al., 2008), one membrane tank required per 10,000 m ²
154	membrane area, allowing sufficient flexibility in operation and cleaning.
155	
156	The required buffer tank volume was dictated by:
156 157	The required buffer tank volume was dictated by:an assumed maximum buffer tank HRT of 2 days - based on the maximum flow
156 157 158	 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
156 157 158 159	 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;
156 157 158 159 160	 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
156 157 158 159 160 161	 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.
156 157 158 159 160 161 162	 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
156 157 158 159 160 161 162 163	 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.
156 157 158 159 160 161 162 163 164	 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.

165 **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:

170

171 *Membranes* A net design flux of 20 $I.m^{-2}.h^{-1}$ (LMH) was used for calculating 172 membrane area, while the maximum sustainable flux was assumed to be 40% 173 higher, i.e. 28 LMH, which can be considered conservative based on literature values 174 (Judd and Judd, 2010; Garcés *et al.*, 2007). A regime of 10 min filtration followed by 175 30 s backwashing resulted in an instantaneous flux of 22.1 LMH and maximum 176 instantaneous flux of 30.9 LMH. HF membrane costs were assumed to be €50.m⁻² 177 (Brepols *et al.*, 2010).

178

179 *Tanks* Tank building costs were based on costs of €220.m⁻³ tank volume (Brepols et
180 al., 2010).

Verrecht et al, The cost of a large-scale hollow fibre MBR

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.

187

To size the membrane blowers, SAD_m was assumed constant at 0.3 Nm³.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 °C.

194

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 m³ of anoxic tank volume was assumed. Costs of land, civil engineering, other electrical equipment and construction were excluded, these being location specific.

200

201 **2.4 Operational costs**

202 Operational costs were determined using the approach of the control strategy 203 evaluation benchmark community (Copp *et al.*, 2002), which was extended by Maere 204 *et al.* (2009) for MBR applications. The opex analysis was limited to energy demand, 205 sludge treatment and disposal, and chemical usage for membrane cleaning.

206

207 **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??).

210

211 **Aeration energy** The influence of MLSS concentration (via the α -factor) and aerator 212 type (fine and coarse bubble) on oxygen transfer was computed using the dedicated 213 aeration model of Maere *et al.* (2009), combining several literature findings (Metcalf 214 and Eddy, 2003; Henze *et al.*, 2008; Verrecht *et al.*, 2008; Krampe and Krauth, 2003; 215 Germain *et al.*, 2007; Stenstrom and Rosso, 2008).

Verrecht et al, The cost of a large-scale hollow fibre MBR

217 Based on typical practically measured values for blower outlet pressure (106300 Pa; 218 for a typical aerator depth of 5 m and allowing for losses incurred in the pipework) 219 and a blower efficiency ξ_B of 0.60, a value of 0.025 kWh.Nm⁻³ air was determined for 220 the aeration energy demand, corresponding well with literature values (Verrecht et 221 al., 2008) and data from blower manufacturers. The average total aeration energy in 222 kWh.d⁻¹ was obtained by summing blower power consumption for both membrane 223 and biology blowers and integrating over the 365 day simulation period (Maere et al., 224 2009).

225

226 **Pumping energy** Sludge pumping requirements, for internal recirculation (Q_{int}, 227 $m^{3}.d^{-1}$), membrane recirculation (Q_{MR} , $m^{3}.d^{-1}$) and wastage (Q_{W} , $m^{3}.d^{-1}$) (Insert Figure 228 2), were determined from the expression of Maere et al. (2009), using a power 229 requirement of 0.016 kWh.m⁻³ of sludge pumped which was calculated from 230 assuming a simple linear dependency of P_{Sludge} (Power required for sludge pumping) 231 on sludge flow and assuming a total headloss Δh of 3m and a pump efficiency ξ_{ρ} of 232 50%. To calculate additional pumping energy for permeate pumping and 233 backwashing, the expression provided by Maere et al. (2009) was applied.

234

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.

238

239 **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 \in 43.tnDS⁻¹ (Rossi *et al.*, 2002), which accounts for chemicals, labour, treatment and disposal, to \in 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.

247

248 **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 NaOCI and 2000 ppm citric acid, and a cleaning out of place (COP) with 1000

253 ppm NaOCI and 2000 ppm citric acid, conducted twice yearly. Representative prices

254 for bulk chemicals were obtained from chemical suppliers.

255 **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).

262 **2.6 Net present value calculation**

263The net present value was calculated for a plant lifetime of 30 years, taking into264account all capital and operational expenditures during the plant lifetime:

265

266
$$NPV = \sum_{t=0}^{29} \frac{(capex)_t + (opex)_t}{(1+i)^t}$$
 (1)

267

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).

272

3. Results and discussion

3.1 Effect of contingency: changes in feedwater flow and strength

275 **3.1.1 Hybrid plant vs. plant designed for maximum flow**

276 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
- 279 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 averageflow).

282

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

Verrecht et al, The cost of a large-scale hollow fibre MBR

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.

289

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.

293

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)

304

305 The analysis shows NPV and operational efficiency of MBRs to be very susceptible 306 to the extent of built-in contingency, which is mostly determined by the changes in 307 feedwater flow such as during storm events. An example of this is the 48 MLD 308 (megalitres per day) Nordkanal plant in Germany (Brepols et al., 2009), which was 309 designed to treat a peak flow that is 3-4 times higher than the average flow. The plant 310 also has 33% more membrane surface installed than required to treat the peak flow, 311 a requirement under German regulations. Consequently, mean fluxes at the plant are 312 only 8 LMH and specific energy consumption for the MBR is 0.5-1.8 kWh.m⁻³. 313 Conversely, the hybrid MBR plant at Ulu Pandan in Singapore is designed to 314 continuously treat a flow of 23 MLD, leading to very efficient operation and energy 315 consumption as low as ~0.4 kWh.m⁻³ for the MBR part of the hybrid plant, mainly due 316 to continuous improvement in membrane aeration protocols (Qin et al., 2006, 2007; 317 Seah et al., 2009). Thus, provided there is a constant demand for high quality effluent 318 for reuse, the hybrid plant is the most favoured option. This can be retrofitted to an 319 existing CAS, provided full effluent disinfection is not required (Lesjean et al., 2009; 320 Mulder, 2008).

Verrecht et al, The cost of a large-scale hollow fibre MBR

322 **3.1.2 Buffer tank vs. extra membranes**

323 Adding buffering capacity for flow equalisation permits a smaller plant design with a 324 reduced membrane surface requirement and so higher average plant utilization. 325 Figure 4 shows the influence of buffer tank size on mean plant utilisation, and 326 contrasts the resulting membrane and buffer tank costs. For practical reasons, the 327 buffer tank size is constrained by an HRT of below 2 days (corresponding to 80% of 328 design flow). Since no influent can bypass the plant under storm conditions the 329 combined capacity of the buffer tank and MBR plant must cope with the maximum 330 flow. Figure 4 shows that over the buffer tank size range considered, the cost of 331 adding a buffer tank is only partially offset by the cost savings from a reduction in 332 required membrane surface area due to increased average plant utilization (Figure 333 5). The EQI and NPV trends are both determined by the constraints on tank size 334 imposed by an HRT of 8h at average flow, or a minimum HRT of 4h at maximum 335 design flow. Addition of a buffer tank with the maximum acceptable size results in a 336 NPV decrease from €30.2 million to €27 million, or a saving of 10.5%, due to 337 decreased opex (-21%), which is partly offset by an increase in capex (+32%) (Figure 338 6). A maximum NPV saving of 11.8% can be achieved through addition of buffer tank 339 with the most economical size (i.e 1.2 d HRT; at average plant utilization of 47.9%). 340 Effluent quality, as indicated by EQI, is largely unaffected and deteriorates by a 341 maximum of 4% over the buffer tank size range considered (Figure 5).

342

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

345

346 The cost of land required for the buffer tank is excluded from this NPV analysis. 347 However, provided the additional land required for the buffer tank has a projected 348 value of less than €3.2m, it is always beneficial to build a buffer tank. Assuming a 349 total plant footprint equaling 2.5 times the combined footprint required for the 350 biotanks and buffer tanks (Brepols et al., 2010), a plant with the maximum sized 351 buffer tank (2d HRT) requires 9,715 m² extra land compared to a plant without buffer 352 tank. Land costs would have to increase to \in 324 per m² before addition of a buffer 353 tank becomes economically unviable. This value is at least 32% higher than typical 354 reported values for industrial land in the Germany, which range between €17 and 355 €247 per m² (Ref for industry land – Christoph??). Assuming a CAS to incur 2.7 356 times the footprint of an MBR (Brepols et al., 2010), a combined MBR with the 357 maximum sized buffer tank would be $\sim 10\%$ larger than a CAS treating the same flow.

358

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

361

362 **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.

368

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

370 Insert Table 2: MBR design parameters and base case costs for the study of 371 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

375

376 Influence of SRT A shorter design SRT decreases capex due to decreased installed 377 aerobic tank blower capacity at the lower MLSS concentrations and the resulting 378 decreased aeration demand. However, the cost for the process blowers is less than 379 2% of total capex (Table 1), so the potential influence is negligible. The reduction in 380 NPV is attributed to the effect of SRT on opex. At a conservative sludge treatment 381 and disposal cost of €150.m³ of dry solids, energy consumption accounts for 78-85% 382 of opex, sludge treatment and disposal for 12-19%, and chemical cleaning about 3%. 383 The decreased aeration demand at lower MLSS concentrations and shorter SRT 384 thus outweighs the costs incurred by increased sludge production. This would seem 385 to corroborate recent trends of working at lower MLSS concentrations, particularly in 386 the US (Trussell et al., 2006, 2007), but is contrary to the conclusions of Yoon et al. 387 (2004). The latter study ignored membrane aeration, thus underestimating the total 388 opex since membrane aeration contributes significantly to total energy demand 389 (Figure 3).

390

391 The influence of SRT is sensitive to sludge treatment and disposal costs. As sludge 392 management costs increase, the cost incurred by sludge treatment and disposal

Verrecht et al, The cost of a large-scale hollow fibre MBR

393 starts to outweigh the opex reduction from decreased energy demand at lower SRT.
394 Table 3 also shows that effluent quality requirements place a lower limit on the SRT
395 operating range, since EQI deteriorates as SRT decreases. Selection of SRT is thus
396 based on available sludge processing facilities on site and end disposal costs, as
397 well as the desired effluent quality. Operation at lower SRT and MLSS values may
398 also lead to higher permeability decline rates (Trussel *et al.*, 2006), mitigating against
399 lower SRT operation.

400

401 Influence of HRT Longer HRTs increase capex due to the larger tank volume 402 required, but this is partially offset by lower opex at lower MLSS concentrations 403 (10,000 and 6,000 mg/l average MLSS concentrations in aerobic tank at 6 and 10h 404 HRT respectively). The impact on NPV is thus negligible compared to, say, the 405 influence of contingency or choice of SRT. The effect on EQI is more pronounced: an 406 increase in average HRT from 6 to 10 hours improves effluent quality by 9%. A larger 407 MBR thus provides better effluent quality, without detriment to NPV provided land 408 costs are not excessive.

409

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.

414

415 Influence of membrane aeration and sustainable flux Membrane aeration energy 416 contributes significantly to opex (Verrecht et al., 2008; Seah, 2009; Brepols et al., 417 2009), as confirmed by Figure 3. Membrane aeration energy can be related to SAD_{n} . 418 the specific aeration demand per unit permeate volume. Extensive pilot studies 419 regarding the impact of membrane aeration and sustainable flux (Guglielmi et al. 420 2007, 2008) suggest a neo-linear relationship between sustainable flux J and U, the 421 in-module air flow velocity in m.s⁻¹ (Verrecht et al., 2008). For HF geometry, 422 calibrating against two full scale plants (Verrecht et al., 2008), the correlation 423 between J and U can be expressed as:

424

425
$$J = m \cdot \frac{SAD_m \cdot L_{membrane}}{\left(\frac{1}{\varphi} - \frac{d_f}{4}\right)} + J_0 \qquad \text{for } J < J_{sust,max} (\text{l.m}^{-2}.\text{h}^{-1})$$
(2)

426
$$J = J_{sust,max} (I.m^{-2}.h^{-1})$$
 for $SAD_m > SAD_{m,max} (Nm^3.m^{-2}.s^{-1})$ (3)

Verrecht et al, The cost of a large-scale hollow fibre MBR

(4)

427

428 where *J* is the flux through the membrane, in $m^3.m^{-2}.h^{-1}$ and SAD_m the specific 429 aeration demand per unit membrane area in Nm³.m⁻².h⁻¹, *m* the slope of the J vs *U* 430 curve (247, according to Verrecht et al, 2008), *L* the membrane module length (1.8 431 m); φ the module packing density (300 m⁻¹), *d*_f the hollow fibre outside diameter 432 (0.002 m) and *J*₀ the intercept of the *J* vs. *U* curve (5 l.m⁻².h⁻¹). Thus: 433

434
$$SAD_p = \frac{SAD_m}{J}$$

435

436 where a minimum SAD_{p} is required to maintain a sustainable flux (2), but increasing 437 SAD_p beyond $SAD_{m,max}$ has no impact on the sustainable flux (3) and a higher 438 maximum sustainable flux demands a higher SAD_{p} . When considering the influence 439 of sustainable flux and SAD_p on NPV (Figure 7), higher sustainable fluxes lead to 440 lower NPVs, indicating that the higher operational costs are offset by lower capital 441 expenditures which can mainly be attributed to the reduction in membrane capacity 442 required. An increase in sustainable flux from 15 to 30 l.m⁻².h⁻¹ results in a decrease 443 in NPV of 9% (at minimum required SAD_o; Table 3). It can thus be concluded that 444 higher sustainable fluxes are beneficial to NPV, despite the higher aeration demand 445 and associated increase in opex, provided $J \leq J_{sust,max}$ and J_o takes a positive value.

446

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

448

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%.

453

454 **Membrane replacement and cost** As shown in Table 1, membrane costs make up 455 47-57% of total capex, while the other process equipment combined contributes 456 about 20%. Analysis of component lifetime cost impacts is thus most sensitive to 457 membrane life and costs. A 'worst case' membrane lifetime of 5 years (i.e. 6 458 membrane replacements in the projected plant lifetime of 30 years) results in a 23% 459 increase in NPV compared to the base cost assuming membrane replacement every 460 10 years. A halving of membrane costs every 10 years, on the other hand, reduces 461 NPV by 9.2%, whereas an increase in initial membrane cost from €20.m² to €100.m²

increases NPV by 50% for a 10 year membrane life and by 85% for a 5 yearmembrane lifetime.

464

465 Since membrane replacement is critical in determining NPV, it is unsurprising that 466 considerable attention has been paid to optimisation of membrane lifetime by 467 operating under a sustainable regime and developing adequate cleaning strategies 468 (Brepols et al., 2008). There is increasing evidence that MBR membrane life can 469 reach, or even exceed, a decade for large plants. The Zenon plants at Rodingen (3.2 470 MLD PDF) and Brescia (42 MLD PDF) are successfully operating with membrane 471 modules which are from 2000 and 2002 years respectively, and the Kubota plant at 472 Porlock still operates with 40% of the panels originally installed in 1997 (Judd and 473 Judd, 2010); predicted replacement intervals of up to 13 years have been reported 474 (De Wilde et al., 2007b).

475

476 **4. Conclusions**

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

479

 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.
- 496
 4. An increased SRT at constant tank volume increases the NPV since a greater
 497
 497 aeration demand is incurred at higher MLSS concentrations. Whilst sludge
 498 production is concomitantly reduced, the resulting cost savings do not fully offset

Verrecht et al, The cost of a large-scale hollow fibre MBR

- 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.
- 5025. Higher sustainable fluxes provide a decreased NPV. Although the opex is503increased due to the higher aeration demand, this is offset by the reduction in504capex and membrane replacement costs since less membrane area is required.
- 505 An increase in sustainable flux from 15 to 30 LMH decreases NPV by 9%.
- 506 6. The future trend in energy costs is a determining factor for NPV: a doubling of507 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.m². If
- 510 initial membrane costs increase five-fold from €20 per m², NPV increases by 85%
- 511 for a 5 year membrane lifetime and by 50% for a 10 year membrane life.

512 Acknowledgements

513 The authors are grateful to MOSTforWATER N.V. (Kortrijk, Belgium) for providing the 514 WEST® modelling software, and would also like to thank Thames Water for the 515 resources provided by them for this paper. Thomas Maere is supported by the 516 Institute for Encouragement of Innovation by means of Science and Technology in 517 Flanders (IWT).

518

519 **References**

520 Brepols, C., Drensla, K., Janot, A., Trimborn, M. and Engelhardt, N. (2008). 521 *'Strategies for chemical cleaning in large scale membrane bioreactors'*. Water 522 Science and Technology 57(3) 457-463.

523 Brepols, C. Schäfer, H. and Engelhardt, N. (2009) *'Economic aspects of large scale* 524 *membrane bioreactors'*. Final MBR-Network Workshop: "Salient outcomes of the 525 European projects on MBR technology", 31/03/2009 - 01/04/2009, Berlin, Germany.

526 Brepols, C. (2010) 'Operating Large Scale Membrane Bioreactors for Municipal 527 Wastewater Treatment: Long Term Experiences in Operating Full Scale Membrane 528 Bioreactors for Municipal Wastewater Treatment', IWA Publishing.

529 Brepols, C., Schäfer, H. and Engelhardt, N. (2010) 'Considerations on design and
530 *financial feasibility of full scale membrane bioreactors for municipal applications*'.
531 Water Science and Technology *in press*.

532 Copp, J. B. (2002), '*The COST Simulation Benchmark - Description and Simulator* 533 *Manual*. Office for Official Publications of the European Communities, Luxembourg.

534 DeCarolis, J., Adham, S., Grounds, J., Pearce, B., Wasserman, J. (2004). 'Cost

analysis of MBR systems for water reclamation'. Proceedings of WEFTEC 2004,
 New Orleans, 2-6 October.

537 Côté, P., Masini, M., Mourato, D. (2004) *'Comparison of membrane options for water* 538 *reuse and reclamation'*. Desalination 167 1-11.

539 De Wilde, W., Richard, M., Lesjean, B. and Tazi-Pain, A. (2007a) 'Towards 540 standardisation of MBR technology? A white paper analysing market expectations 541 and technical potential for membrane bioreactor standardisation in Europe'. 542 Published within the framework of AMEDEUS, MBR-Network, EU.

543 De Wilde, W., Thoeye, C. and De Gueldre, G. (2007b) '*Membrane life expectancy* 544 assessment after 3 years of *MBR operation at WWTP Schilde*'. 4th International 545 Water Association Conference on Membranes for Water and Wastewater Treatment, 546 Harrogate, UK

- 547 EIA Energy information administration (2009), website accessed March 2010 548 http://www.eia.doe.gov/emeu/aer/txt/ptb0810.html
- 549 Energy EU (2009) http://www.energy.eu/#industrial ; website accessed March 2010
- 550 Germain, E., Nelles, F., Drews, A., Pearce, P., Kraume, M., Reid, E., Judd, S. J. and 551 Stephenson, T. (2007) '*Biomass effects on oxygen transfer in membrane* 552 *bioreactors*'. Water Research 41(5) 1038-1044.
- Garcés, A., De Wilde, W., Thoeye, C. and De Gueldre, G. (2007), 'Operational cost
 optimisation of MBR Schilde'. Proceedings of the 4th IWA International Membranes
 Conference "Membranes for Water and Wastewater Treatment, 15-17 May 2007,
 Harrogate, UK.
- 557 Gernaey, K., Rosen, C. and Jeppson, U. (2006) '*WWTP dynamic disturbance* 558 modelling – An essential moduel for long-term benchmarking development'. Water 559 Science and Technology 53(4-5) 225-234.
- 560 Guglielmi, G., Chiarani, D., Judd, S., Andreottola, G. (2007). *'Flux criticality and* 561 *sustainability in a hollow fibre submerged membrane bioreactor for municipal* 562 *wastewater treatment* Journal of Membrane Science 289(1-2) 241-248.
- 563 Guglielmi, G., Chiarani, D., Saroj, D.P., Andreottola, G. (2008). '*Impact of chemical* 564 *cleaning and air-sparging on the critical and sustainable flux in a flat sheet* 565 *membrane bioreactor for municipal wastewater treatment*' Water Science and 566 Technology 57(12) 1873-1879.
- 567 Henze, M., Gujer, W., Mino, T., van Loosdrecht, M. (2000), '*Activated Sludge Models* 568 *ASM1, ASM2, ASM2d and ASM3*', IWA Publishing, London.
- Henze, M., van Loosdrecht, M., Ekama, G. A. and Brdjanovic, D. (2008), '*Biological Wastewater Treatment: Principles, Modelling and Design*'. IWA Publishing, London.
- 571 Itokawa, H., Thiemig, C., & Pinnekamp, J. (2008). '*Design and operating experiences* 572 of municipal MBRs in Europe'. Water Science and Technology 58(12) 2319-2327.
- 573 Judd, S. (2008). '*The status of membrane bioreactor technology*'. Trends in 574 Biotechnology 26(2) 109-116.
- 575 Judd, S.J. and Judd, C. (2010) 'The MBR Book (2nd ed.): Principles and applications 576 of membrane bioreactors in water and wastewater treatment', Elsevier.
- 577 Krampe, J. and Krauth, K. (2003), '*Oxygen transfer into activated sludge with high* 578 *MLSS concentrations*', Water Science And Technology, 47(11) 297-303.
- 579 Lesjean, B., Ferre, V., Vonghia, E. and Moeslang, H. (2009). *'Market and design* 580 *considerations of the 37 larger MBR plants in Europe'*. Desalination and Water 581 treatment, 6 227-233
- 582 McAdam, E and Judd, S. (2008), *'Immersed membrane bioreactors for nitrate* 583 *removal from drinking water: Cost and feasibility'*. Desalination 231 52-60.

Verrecht *et al*, The cost of a large-scale hollow fibre MBR

- Maere, T., Verrecht, B., Benedetti, L., Pham, P.T., Judd, S. and Nopens, I. (2009). *Building a Benchmark Simulation Model to Compare Control Strategies for Membrane Bioreactors: BSM-MBR*². 5th IWA specialised membrane technology
 conference for water & wastewater treatment, 01-03 September 2009, Beijing, China.
- 588 Metcalf and Eddy, Tchobanoglous, G., Burton, F. L. and Stensel, H. D. (2003), 589 *'Metcalf and Eddy - Wastewater Engineering – Treatment and Reuse'*, 3rd edition, 590 McGraw-Hill, New York.
- Maurer, M. (2009). 'Specific net present value: An improved method for assessing
 modularisation costs in water services with growing demand'. Water Research 43
 2121-2130
- 594 Mulder, J-W. (2008). 'Hybrid MBR Heenvliet D32 EUROMBRA
- Qin, J., Kekre, K. A., Tao, G., Oo, M. H., Wai, M. N., Lee, T. C. (2006). 'New option of
 MBR-RO process for production of NEWater from domestic sewage' Journal of
 Membrane Science 272(1-2) 70-77.
- Qin, J., Wai, M. N., Tao, G., Kekre, K. A., & Seah, H. (2007). '*Membrane bioreactor*study for reclamation of mixed sewage mostly from industrial sources'. Separation
 and Purification Technology (3) 296-300.
- Rossi, L., Lubello, C., Cammeli, M., Griffini, O. (2002) 'Ultrafiltration compared to
 traditional solid removal for drinking water treatment: Design and economic analysis,
 e21179a'. Proceedings of AWA Conference, Melbourne, Australia, 2002.
- Seah, H. (2009). '*Technology roadmap for Wastewater Treatment in 2030 The Next Frontier in Energy Options*' 7th IWA World Congress on Water Reclamation and
 Reuse. 20-25 September 2009, Brisbane, Australia
- Stensel, H. D. and Strand, S. E. (2004). 'Evaluation of feasibility of methods to
 minimize biomass production from biotreatment, Biosolids & residuals', WERF. IWA
 Publishing
- 610 Stenstrom, M.K. and Rosso, D. (2008) 'Aeration and mixing', Chapter in 'Biological 611 Wastewater Treatment: Principles, Modelling and Design', Henze, M., van 612 Loosdrecht, M., Ekama, G. A. and Brdjanovic, D. (2008), IWA Publishing, London.
- Trussell, R. S., Merlo, R. P., Hermanowicz, S. W. and Jenkins, D. (2006), '*The effect* of organic loading on process performance and membrane fouling in a submerged membrane bioreactor treating municipal wastewater', Water Research 40(14) 2675-2683.
- Trussell, R. S., Merlo, R. P., Hermanowicz, S. W., & Jenkins, D. (2007). 'Influence of
 mixed liquor properties and aeration intensity on membrane fouling in a submerged
 membrane bioreactor at high mixed liquor suspended solids concentrations'. Water
 Research 41(5) 947-958.
- 621 VOA 'Valuation Office Agency' (2007), website accessed March 2010
- 622 <u>http://www.voa.gov.uk/publications/property_market_report/pmr-jan-07/Print-</u>
 623 <u>Version/Industrial-Land.pdf</u>
- Vanhooren, H., Meirlaen, J., Amerlink, Y., Claeys, F., Vangheluwe, H., Vanrolleghem,
 P.A. (2003) 'WEST: Modelling biological wastewater treatment', Journal of
 Hydroinformatics 5(1) 27-50.
- 627 Vanrolleghem, P. A., Jeppsson, U., Carstensen, J., Carlsson, B. and Olsson, G. 628 (1996). 'Integration of wastewater treatment plant design and operation - A
- 629 systematic approach using cost functions'. Water Science and Technology, 34(3-4), 630 159-171.

Verrecht et al, The cost of a large-scale hollow fibre MBR

- 631 Verrecht, B., Judd, S., Guglielmi, G., Brepols, C. and Mulder, J. W. (2008), 'An
- 632 aeration energy model for an immersed membrane bioreactor, Water Research
- 633 42(19) 4761-4770.
- 634 Yoon, S-H., Kim, H-S and Yeom, I-T (2004). 'The optimum operational condition of
- 635 membrane bioreactor (MBR): cost estimation of aeration and sludge treatment.
 636 Water Research 38(1) 37-46.

Verrecht et al, The cost of a large-scale hollow fibre MBR

WR14476

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 -N _{average}	mg.l ⁻¹	0.46	0.43
NO3-Naverage	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

Verrecht et al, The cost of a large-scale hollow fibre MBR

WR14476

parameters						
Parameter	Units	Value	Reference		Units	Value
As	ssumptions for	r capex calcu	lation	Base design,	EQI and NF	٧٧
Membrane cost	€.m-2	50	Judd & Judd, 2010	Design capacity	m ³ .d ⁻¹	30,416
Tank civil cost	€.m⁻³ tank	220	Brepols, 2010b	Maximum plant	m ³ .d ⁻¹	42,582
	volume			capacity*		
Screens – 0.75	€.m- ³ .d ⁻¹	3.1 – 5.6**	Manufacturers	Total tank volume	m ³	7,097
mm	capacity					
Screens – 6mm	€.m- ³ .d ⁻¹	0.9 – 2.1**	Manufacturers	Membrane area	m ²	63,366
	capacity					
Blowers	€.Nm ⁻³ .h ⁻¹	4 – 4.3**	Manufacturers	SRT	d	23.8
	capacity					
Permeate	€.m ⁻³ .h ⁻¹	58.8	Manufacturers;			
pumps	capacity	10.1	Brepols, 2010b			
Biomass	€.m ⁻³ .h ⁻¹	12.1	Manufacturers;	Buffer tank size	m ³	14,530
recirculation	capacity		Brepols, 2010b			
pumps		07.0	D I 00401		2 14	40.400
Mixing	€.m ⁻ ³ tank	27.8	Brepois, 2010b	Maximum flow out	m³.d⁻¹	12,166
equipment	volume			of buffertank*		
А	ssumptions fo	r opex calcul	ation	Max HRT in buffer	d	1.2
				tank		
Energy cost	€.kWh ⁻¹	0.0942	- ret Christophe			
Sludge	€.ton ⁻¹ of	150	-	Effluent quality	kg PU.d-	5,430
treatment cost	DS	700		Index	1	0.50
Citric acid 50%	€/ton	760	Brepois, 2010b	NH4-N	mg.I-	0.52
NaOCI 14%	€/m ³	254	Brepois, 2010b	NO ₃ -N	mg.I	10.7
A	ssumptions fo	or NPV calcula	ation	COD	mg.l ⁻¹	30.1
Membrane life	Year	10	Judd & Judd, 2010	Net present value	Million	26.7
				 	Euro	
	0 /	0 01			(M€)	
Inflation	%	3%		1 1 1		
Discount rate	%	6%	4 /			

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

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

** Depending on size of installed equipment

Verrecht et al, The cost of a large-scale hollow fibre MBR

WR14476

Table 3: Sensitivity of I	NPV and EQI on design and o	operational parameters and costs. %
Change in NPV and EC	OI is compared with the base of	conditions as described in Table 2

Net present value		EQI	
Million euro	% change	kg PU.d¹	% change
(M€)			
26.4	-1.1%	5,835	+7.5%
27.8	+4.4%	5,172	-4.7%
26.3	-1.3%	5,628	3.7%
26.8	+0.5%	5,214	-4.0%
29.1	9.2%	5,551	+2.2%
26.5	-0.5%	5,295	-2.5%
30.2	+13.4	5,236	-3.6%
27.1	+1.6%	5,401	-0.5%
	C		
26.7	+0.1%	6,313	+16.3
26.6	-0.3%	5,146	-5.2
28.2	+5.7%	5,430	0%
34.7	+30.0%	5,430	0%
25.2	-5.6%	5,430	0%
28.8	+7.9%	5,430	0%
22.4	-15.8%	5,430	0%
33.7	+26.4%	5,430	0%
24.2	-9.3%	5,430	0%
32.8	+23.1%	5,430	0%
	Net present Million euro (M€) 26.4 27.8 26.3 26.4 27.8 26.5 30.2 27.1 26.7 26.6 28.2 34.7 25.2 28.8 22.4 33.7 24.2 32.8	Net present value Million euro (M€)% change % change26.4-1.1% 	Net present value Million euro (M€)% change % changeE kg PU.d¹26.4-1.1% 5,8355,83527.8+4.4%5,17226.3-1.3% 5,6285,62826.8+0.5%5,21429.19.2% 5,5515,55126.5-0.5%5,29530.2+13.4 +1.6%5,23627.1+1.6%5,40126.7+0.1% 5,4016,313 5,14626.6-0.3%5,14628.2+5.7% 5,4305,43025.2-5.6% 5,4305,43028.8+7.9%5,43022.4-15.8% 5,4305,43024.2-9.3% 5,4305,43024.2-9.3% 5,4305,43024.2-9.3% 5,4305,430

Verrecht et al, The cost of a large-scale hollow fibre MBR



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





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

WR14476



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



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



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



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

WR14476



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

CERES Research Repository

https://dspace.lib.cranfield.ac.uk/

School of Applied Sciences (SAS) (2006-July 2014)

Staff publications (SAS)

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 Downloaded from CERES Research Repository, Cranfield University