Energy-efficient wastewater treatment via the air-based, hybrid membrane biofilm reactor (hybrid MfBR)

M. Aybar, G. Pizarro, J. P. Boltz, L. Downing and R. Nerenberg

ABSTRACT

We used modeling to predict the energy and cost savings associated with the air-based, hybrid membrane-biofilm reactor (hybrid MfBR). This process is obtained by replacing fine-bubble diffusers in conventional activated sludge with air-supplying, hollow-fiber membrane modules. Evaluated processes included removal of chemical oxygen demand (COD), combined COD and total nitrogen (TN) removal, and hybrid growth (biofilm and suspended). Target concentrations of COD and TN were based on high-stringency water reuse scenarios. Results showed reductions in power requirements as high as 86%. The decrease mainly resulted from the dramatically lower air flows for the MBfR, resulting from its higher oxygen-transfer efficiencies. When the MBfR was used for COD and TN removal, savings up to US$200/1,000 m^3 of treated water were predicted. Cost savings were highly sensitive to the costs of the membrane modules and electrical power. The costs were also very sensitive to membrane oxidation flux for ammonia, and the membrane life. These results suggest the hybrid MBfR may provide significant savings in energy and costs. Further research on the identified key parameters can help confirm these modeling predictions and facilitate scale-up.

Key words | biofilm, energy, hollow-fiber, MABR, MBfR, membrane

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INTRODUCTION

The conventional activated sludge (CAS) treatment process is robust, reliable, and effective, and has become the industry standard for wastewater treatment in the USA. However, activated sludge is extremely energy intensive, consuming around 2% of all electrical power in the USA (Tchobanoglous *et al.* 2003). In a typical wastewater treatment plant, around 40–60% of the energy is devoted to aeration, typically by mechanical or diffused air devices (Tchobanoglous *et al.* 2003; WERF 2011). However, in diffused aeration only 5–25% of oxygen supplied by blowers is transferred to the liquid phase, with the rest evolving as bubbles (Tchobanoglous *et al.* 2003).

Hollow-fiber membranes can provide a highly efficient means for oxygen transfer (Buer *et al.* 2008; Li *et al.* 2010), and membrane biofilm reactors (MBfR), also known as membrane-aerated biofilm reactors (MABRs), equipped with gas-supplying membranes can achieve essentially 100% gas transfer efficiency (Ahmed & Semmens 1992; Pankhania *et al.* 1994; Semmens 2007; Syron & Casey 2008; Martin & Nerenberg 2012). The high efficiency, along with the ability to match oxygen supply with oxygen demand, can provide significant energy savings. Cassettes or modules of membranes can retrofit existing CAS processes in a modular fashion, or can be used in new centralized or decentralized treatment systems.

When air or pure oxygen is supplied to the lumen of a membrane, and when the far end of the membrane is sealed, 100% of the oxygen passively diffuses through the membrane wall to the surrounding biofilm or bulk liquid (Martin & Nerenberg 2012). When chemical oxygen demand (COD) is present in the bulk liquid, a biofilm forms on the membrane's outer wall (Essila *et al.* 2000). In this system, the biofilm serves as a 'reactive barrier', consuming oxygen before it reaches the bulk

doi: 10.2166/wst.2014.086

liquid. Hollow-fiber membranes can be made with outside diameters as small as 75 μm, providing specific surface areas as high as 5,000 m^2/m^3 (membrane surface area/reactor volume).

MBfRs are especially effective for total nitrogen (TN) removal, due to the counter-gradient diffusion of substrates (Essila *et al.* 2000; Downing & Nerenberg 2008b). In a conventional biofilm (co-diffusional transport of substrates), COD, NH_4^+ and dissolved oxygen (DO) diffuse from the bulk liquid into the biofilm, where their concentrations decrease as they penetrate the biofilm (Figure 1(a)). Slow-growing nitrifying bacteria (NB) accumulate close to the attachment surface. As the biofilm thickness increases, NB are exposed to lower DO concentrations. This situation is unfavorable for nitrification, due to NB's relatively high half-saturation constant for DO. The situation is compounded at high COD loadings, as heterotrophic bacteria (HB) deplete DO and increase mass-transfer resistance for DO and NH_4^+ penetration into the biofilm. Unlike conventional biofilms, in the MBfR the DO is highest at the attachment surface (Figure 1(b)) and drops to low levels in the bulk liquid. In this situation, NB are exposed to high DO levels, leading to higher nitrification rates. At the same time, the outer biofilm and bulk liquid are anoxic, allowing HB to reduce NO_3^- with influent COD as an electron donor. As a result, this configuration can achieve COD removal, nitrification, and denitrification within a single biofilm.

The objective of this study was to perform a quantitative evaluation of hollow fiber membranes' mass transfer advantages for oxygen supply in comparison with conventional systems based on fine pore diffusers. We used mathematical modeling for cost and energy estimation to compare the MBfR with conventional aeration for processes for COD removal and for combined COD and TN removal.

METHODS

The energy consumption model includes energy for air supply, flow recirculation, and reactor mixing. The costs include membrane modules and energy. It was assumed that both CAS and hybrid MfBR systems required similar recycle and settling features, so the energy consumption would be similar. The only pumping cost considered in simulations was the internal recycle for the CAS-modified Ludzack Ettinger (MLE) process. Viscosity and pressure loss effects were not taken into account in the model. Different influent COD and total Kjeldahl nitrogen (TKN) loadings were assessed, as well as different oxygen transfer efficiencies (OTEs) and anoxic-reactor mixing energies. The MBfR was considered in hybrid (suspended and attached biofilm growth) (Downing & Nerenberg 2008b) configurations and was set up as a well-mixed reactor in which membranes were submerged. As a benchmark for energy and costs, the CAS process was used in the MLE process configuration (Tchobanoglous *et al.* 2003). Figure 2 shows the schematics of both CAS and MBfR processes.

CAS and MBfR models' description

Both systems were designed for either biodegradable COD removal only, or for COD and TN removal. The treatment goals were 8 mgCOD/L for COD removal alone and 8 mgCOD/L and 3 mgN/L for COD and TN removal processes. These goals represent a high stringency scenario for water reuse context (FDEP 1996). Systems were based on a 10,000 m^3/d (2.6 million gallons per day (MGD)) average daily flow. The influent COD was variable between 150 and 800 mgCOD/L, where the COD was fully degradable, and the TKN was either 35 or 50 mgN/L.

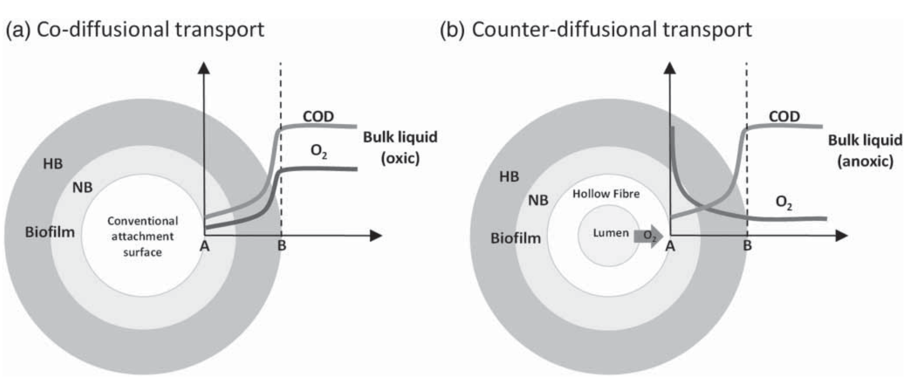

Figure 1 | COD and DO cross-section profiles from point A to B inside a biofilm formed at the outer wall of a circular attachment surface (NH_4^+ profile not shown). (a) Conventional biofilm, co-diffusion of substrates. (b) MBfR, counter-diffusion of substrates.

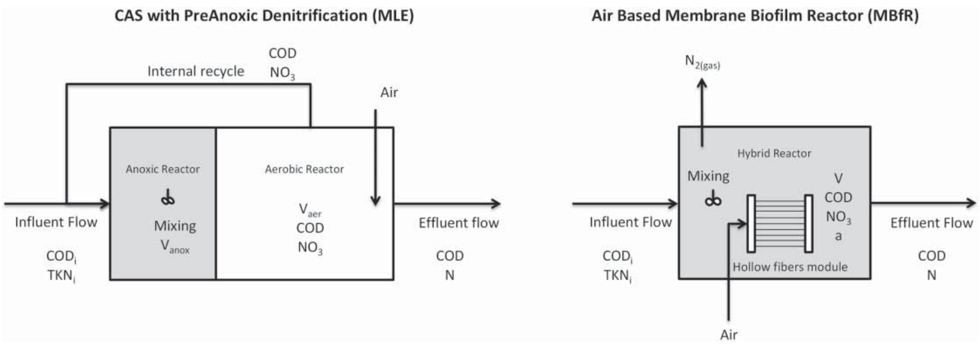

Figure 2 | Schematics of CAS in an MLE configuration (left) and MBfR (right). Inlet in both systems contains COD and TKN (as ammonia in the models). In the MBfR, *a* represents the specific surface area of membranes.

Oxygen requirements for COD biological oxidation and nitrification were determined using the methodology proposed by Metcalf and Eddy Wastewater Engineering for process design (Tchobanoglous *et al.* 2003). The CAS and hybrid MBfR unit processes' definition is shown in Table 1.

For the hybrid MBfR for COD removal only, the membrane area was determined based on the membrane oxygen transfer capacity, with suspended growth determined based on an solids retention time (SRT) of 3 days (Ahmed *et al.* 2004). For the MBfR for COD and TN removal, the membrane area was determined based on the nitrification flux. The required membrane area (main variable for estimating membrane module cost) was determined by dividing the nitrogen loading by the removal flux rate. The reactor volume was then determined by dividing the required membrane area by the specific surface area of membrane a. An a value of 115 m^2/m^3 was obtained from the literature (Downing *et al.* 2010). The a value was fixed for all the simulations carried out in this study; however, the use of higher values of this parameter could result in less required volume of reactors and decreased cost of construction of membrane cassettes. A flux of 1.5 $mgN/(m^2 \cdot d)$ was used (Downing & Nerenberg 2008b). It was assumed that ammonium was transformed to nitrate, although research suggests that nitrite may be formed, decreasing oxygen demand (Downing & Nerenberg 2008a). All the simulations had sufficient COD to reduce the nitrate; therefore no external electron donor was necessary to add to reactors.

Energy required for aeration depends on biological needs of oxygen for degradation of organic matter, the OTE, the energy required by the blower to achieve the required air flow or pressure, and the blower motor efficiency. In this work, the total power consumption is estimated based on the methodology described by Tchobanoglous *et al.* (2003) and Mueller *et al.* (2002). For the simulations, different input COD concentrations (150, 300, 500 and 800 mgCOD/L), Specific OTE (SOTE) parameter values (20 and 40%), influent TKN (35 and 50 mgN/L), and mixing energy of MBfR (15 and 25 kW/1,000 m^3) were considered.

Table 1 | CAS and hybrid MBfR processes considered in simulations

Process/Feature	CAS	Hybrid MBfR
Biofilm	No	Yes
Suspended growth	Yes	Yes
Nitrification	Suspended biomass in aerobic reactor	Biofilm
Denitrification	Suspended biomass in anoxic unit	Suspended biomass
COD oxidation	Suspended biomass in aerobic and anoxic units	Suspended biomass
Oxygen supply method	Fine pore diffusers	Hollow-fiber membrane

RESULTS

Energy modeling results

The hybrid configuration had a high potential for energy savings with a smaller reactor size capable of sustaining only membrane nitrification and COD oxidation with low SRTs.

Figure 3(a) shows the individual power consumption of both systems. The MBfR is shown for two mixing energies, 15 and 25 kW/1,000 m^3 of treated wastewater). The MBfR consumes significantly less power than CAS, although the difference is smaller if a higher mixing energy is used. For the CAS line (upper black line), the increase in power consumption from 300 mgCOD/L results from increasing

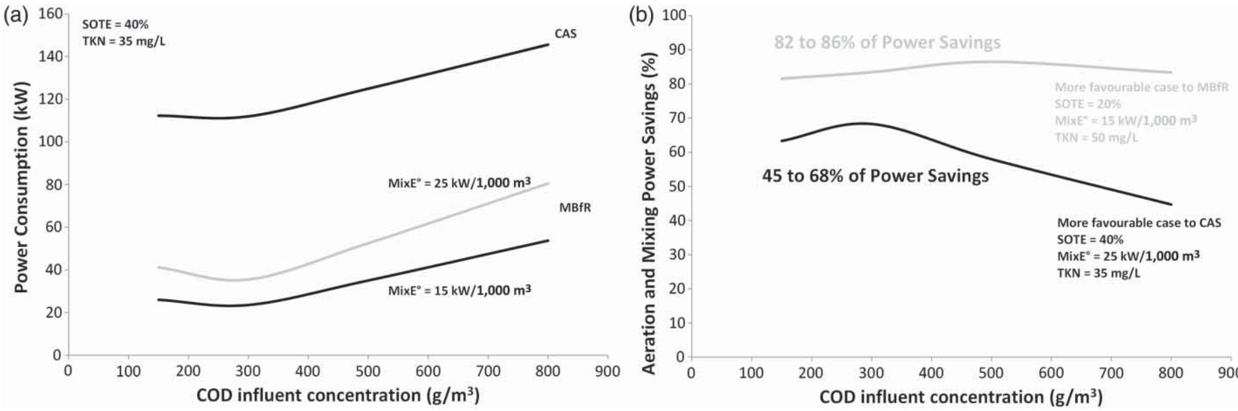

Figure 3 | Power consumption and savings with the Hybrid MBfR, relative to CAS, for COD and TN removal. (a) Variation of power consumption for CAS and MBfR processes. (b) Range of energy savings for a favorable case for MBfR and CAS (upper gray and lower black line, respectively).

blower energy for COD oxidation. This increment offset the decrease in energy consumption associated with less internal recycle pumping and anoxic reactor mixing (due to less nitrogen oxidized to nitrate as more is used for biomass growth). For the MBfR line (lower gray and black lines), the increase in power consumption with COD increments also results from increasing blower energy for COD oxidation. The reactor mixing energy depends on the reactor volume (more volume requires more energy for mixing). From 150 to 300 mgCOD/L, the reactor volume decreases because the membrane area is based on nitrification requirements. With the higher COD, there is more TKN incorporated into biomass and less available for nitrification. Above 300 mgCOD/L, the reactor volume is defined based on the suspended biomass requirements (COD removal); thus the volume increases with increasing COD.

The influent COD/TN ratio also affects the overall design and energy consumption. The power savings with the MBfR increases with increasing influent TKN concentration at high levels of COD. At low levels and high concentrations of TKN, the MBfR system consumes more power. This is because the main reactor volume size, which directly affects mixing energy, is defined by the influent nitrogen that needs to be oxidized (volume estimated with fiber membrane area density); hence the power saving is similar to the CAS system. The CAS system uses more energy with more TKN for all COD values because the anaerobic reactor size is proportional to nitrification; hence more mixing is required for higher volumes.

The main factor for energy savings is the oxygen transfer efficiency. Plotting results for the most favorable case for the MBfR system and the most favorable case for CAS, upper and lower limits of savings can be estimated. The range of energy savings in both cases is estimated comparing the CAS and MBfR system costs. Figure 3(b) shows the range of savings from the lower black curve, which corresponds to the favorable case for the CAS system (SOTE = 40%, high mixing energy for the MBfR reactor 25 kW/1,000 m^3 and influent TKN = 35 mgN/L), until the upper gray curve that corresponds to the favorable case for the MBfR system (SOTE = 20%, low mixing energy for the MBfR reactor 15 kW/1,000 m^3 and influent TKN = 50 mgN/L) with maximum savings of 68 and 86% respectively.

For 150 to 300 mgCOD/L, the decrease in consumption for the MBfR system, because of the less mixing energy requirements, results in a slight increase in energy savings for this COD range. Above 300 mgCOD/L, the increase in energy consumption for the CAS system is less, relative to the MBfR, because for the CAS process the lower energy requirements of internal recycle pumping and anoxic reactor mixing offset a portion of overall consumption.

Results demonstrate that there are significant energy savings associated with aeration and mixing of reactors (up to 86%). However, it is important to differentiate the contribution of each consumption separately. For the CAS reactor above COD of 300 mgCOD/L, the mixing energy demands are insignificant compared to internal recycle pump and compressor demands.

When COD concentration increases, the blower consumption becomes more relevant as more organic matter needs to be oxidized. For the MBfR system, over 90% of consumption (from 300 mgCOD/L of initial COD) corresponds to mixing of the reactor (case specific energy 25 kW/1,000 m^3). Therefore, the energy savings simulated with this model could be substantially improved if the energy for mixing is optimized.

Cost modeling results

Cost comparison was performed considering the differences between a CAS and MBfR in two configurations, hybrid system with removal of COD and nitrogen, and hybrid system with COD removal only. For this comparison oxygen and energy requirement inputs were obtained from previous simulations. Variables considered for energy consumption (electricity costs) and material consumption (cost of membrane modules) are shown in Table 2.

This scenario compares costs of traditional CAS (with fine bubble diffusers) and CAS 'retrofitted' as a MBfR. The costs of infrastructure and equipment (e.g., diffusers and aeration basins) are not considered as they are assumed to be sunken costs. Figure 4 shows the comparison between the costs of the two systems for different electricity and membrane cost scenarios. Each graph shows the area or zone which represents the most cost-effective system through 'equal cost' lines (US$/1,000 m^3 of treated water) representing the cost difference between the two systems. The X axis represents the cost of electricity (US$/kWh) and the Y axis the cost of the membrane modules (US$/$m^2$). Electricity pricing varies widely between countries, from approximately US$0.07/kWh to US$0.2/kWh. The cost of the membrane module depends on the membranes themselves and in the construction of the cassette that holds the membranes. Casey *et al.* (2008) assumed membrane costs ranging from US$30 to US$80/m^2. According to Visvanathan *et al.* (2000) and Gander *et al.* (2000), prices of plate and frame MBR membrane units vary from US$50 to US$130/m^2. However, these membrane units are less competitive than hollow-fiber configurations (Lesjean *et al.* 2004). More economic values could be expected for MABR systems based on hollow-fiber membranes made with new dense polymeric materials.

Figure 4(a) shows that for the hybrid system with COD plus TN removal and 500 mgCOD/L, savings for unitary costs of between US$100 and US$150/m^3 are achieved for electricity cost close to US$0.2/kWh and membrane module cost less than 15 US$/$m^2$. For example, if we consider a 2.6 MGD (10,000 m^3/d) treatment plant, approximately US$500,000/year could be saved. However, the MBfR system costs between US$50 and US$100/m^3 more than the CAS system for energy cost values near US$0.05/kWh and membrane costs over US$30/$m^2$. Figure 4(b) (800 mgCOD/L), shows that there are savings for all the scenarios considered in the simulations (US$0.05 to US$0.2/kWh cost of electricity and US$10 to US$40/m^2 cost of membranes). In the latter case, greater savings are observed in comparison to the case of 500 mgCOD/L because the microorganisms require a greater amount of nitrogen for cell growth, to remove the higher concentration of organic matter; therefore there is a lower concentration of nitrogen to be removed by membranes, resulting in a lower requirement for membrane surface area. In this case, the savings from less membrane area are the main parameter within overall costs.

Table 2 | Variables considered in cost simulations

CAS reactor	Hybrid MBfR reactor
Anoxic reactor mixing energy	Reactor mixing energy
Blower power energy	Blower power energy
Internal recycle pump energy	Membrane material cost

The costs of the hybrid system for COD removal only (Figures 4(c) and 4(d)) are different to those of the system for COD and TN removal. In this configuration for MBfR, the required amount of membrane depends on the flow of oxygen required for the removal of COD by the suspended heterotrophic microorganisms. An increase in concentration of organic matter generates slightly more savings. There is no significant cost variation for different COD values because energy savings are only proportional to COD concentration in CAS and MBfR systems, and membrane area depends only on oxygen transfer capacity to the bulk.

MBfR total cost (cost of electricity + cost of membrane modules) sensitivity was evaluated as a function of different parameters. Results showed that for a hybrid system with COD and nitrogen removal, the parameters that strongly affect the total cost are the concentration of COD and TKN (COD/TKN ratio), the membrane specific nitrification rate, the cost of membranes and the membrane life. The ammonia removal rate is a critical parameter for the hybrid system, greatly affecting costs. Decreasing the ammonia flux from 1.5 gN/(m^2·d) (reference value in simulations) to 0.75 gN/(m^2·d) (50% decrease) increased the cost by about 80%. For the case shown in Figure 4(a), a decrease in the ammonia rate would move down the CAS/MBfR limit (thick black line), increasing the area where CAS is more cost effective. Simulations using different ammonia rate values showed an average decrease in cost savings from US$45 to US$100 per unit of rate decreased. Additionally, membrane life has a significant impact on the operational cost (Lesjean *et al.* 2004). Decreasing the lifetime from 10 to 5 years caused an increase of 70% in the total cost of operation. On the other hand, increasing the lifetime from 10 to 15 years decreased the total cost by

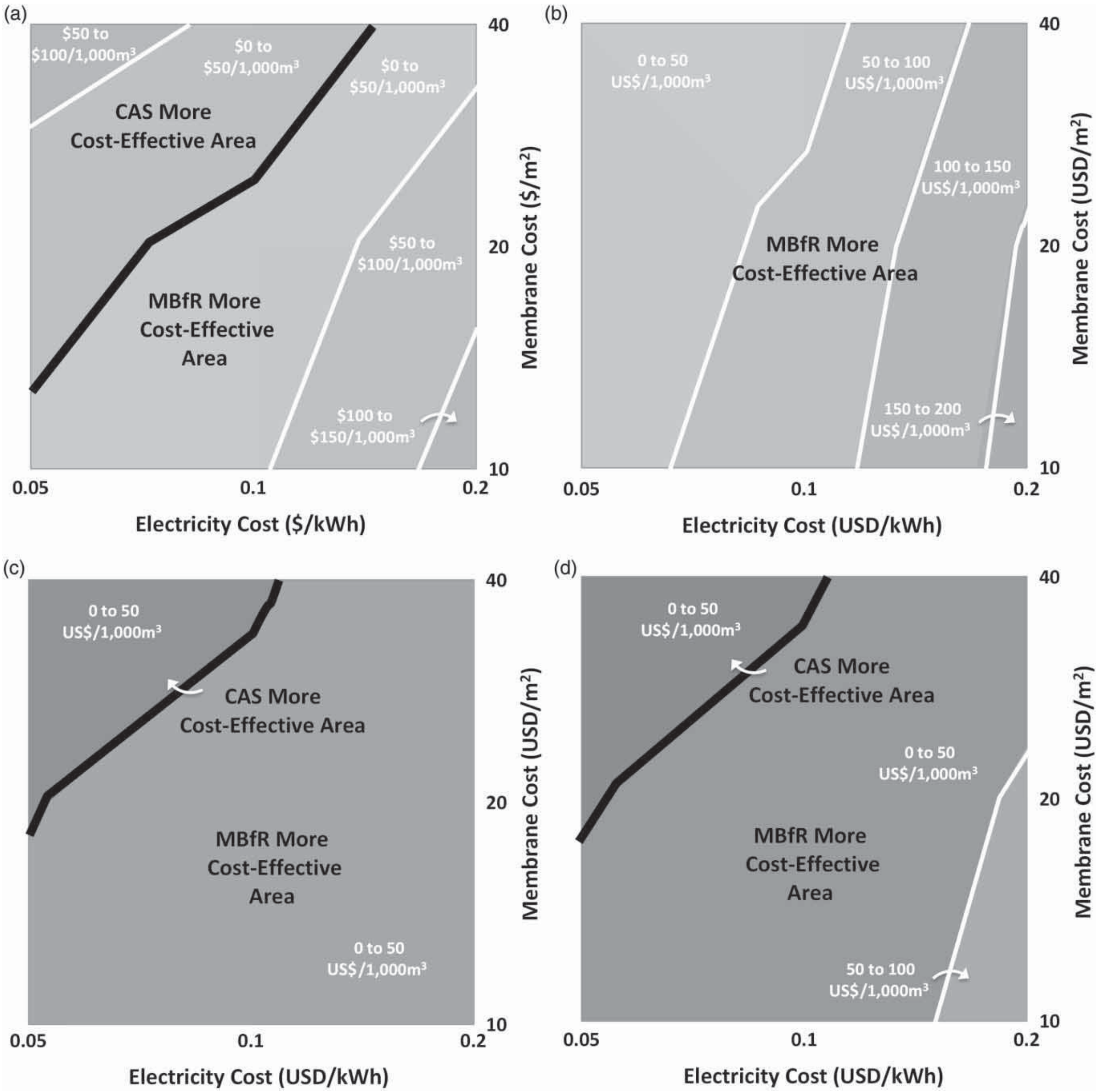

Figure 4 | Isocost lines for total cost difference between CAS and hybrid MBfR systems. CAS SOTE = 20%, MBfR mixing energy = 15 kw/1,000 m^3. (a) Hybrid COD + N removal (500 mgCOD/L). (b) Hybrid COD + N removal (800 mgCOD/L). (c) Hybrid COD only removal (300 mgCOD/L). (d) Hybrid COD only removal (500 mgCOD/L).

only approximately 25%. Currently, there's little information reported about real membrane lifetime. Long-term pilot scale testing is necessary to improve modeling assumptions. In the case of the hybrid system with COD removal only, the parameters that affect the cost are the concentration of COD, the COD membrane oxidation rate, the cost of membranes and membrane life. In this latter configuration, the energy consumption costs associated with the mixing energy and the outlet pressure and the cost of electricity have a greater effect because in this configuration the membrane area is smaller and therefore the cost of this input has a lower relative weight in the calculation.

CONCLUSIONS

In this work, we used modeling and cost estimating to simulate the potential energy and total economic savings generated by implementation of improvements in oxygen supply in conventional treatment systems. Simulations were made considering COD and TN target concentrations based on a high-stringency water reuse scenario. Oxygen supply by membranes has great potential for energy savings. According to the results, MBfRs may save up to 85% of electrical energy when compared with CAS systems. The main savings are associated with higher oxygen transfer efficiency of membranes compared with fine-bubble diffusers, which are considered to have the highest energy efficiency for CAS. The extent of savings depends mainly on the efficiency of the fine bubble systems (SOTE up to 40%), the COD/TKN ratio and MBfR mixing energy consumption.

Considering costs, the economic savings are strongly dependent on membrane and electricity prices. Comparing systems with removal of COD and N, savings up to US$200/1,000 m^3 of treated water could be achieved, depending on the costs mentioned above. Critical factors affecting the cost estimate of required membrane fibers are oxidation rates of COD and N, the cost of membranes and membrane lifetime.

Simulations of energy and economic savings show great potential for improvements that can be implemented in conventional treatment systems or new treatment plants. These improvements do not depend only on cost parameters, but also on influent and effluent water quality, treatment settings and contaminant removal rates. Bench and pilot scale studies are needed to better quantify the parameters with the greatest impact on energy and cost, such as COD and nitrification fluxes, membrane specific surface areas, and membrane costs and life.

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First received 15 October 2013; accepted in revised form 5 February 2014. Available online 19 February 2014