Comparative economic analysis of full scale MABR configurations

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Abstract
The membrane-aerated biofilm reactor (MABR) is a technology that can deliver oxygen at high rates and transfer efficiencies. This paper provides a comparative cost analysis of the MABR compared to the activated sludge process. Membrane cost and electricity cost were found to be the critical parameters determining the relative feasibility of the conventional process to the membrane based process. The general downward trend in the market price of membranes and the steady increase in energy costs in recent years may prove to be a strong driver for the further development of this technology.

Keywords membrane, biofilm, oxygen, energy, cost.

INTRODUCTION
One of the key advantages of biofilm-based wastewater treatment processes is the potentially high volumetric reaction rate that can be attained due the high specific biomass concentration. Unfortunately, this advantage is rarely exploited in full-scale processes as a result of oxygen transfer limitations into thick biofilms. In the membrane aerated biofilm reactor (MABR), the biofilm is immobilized on an oxygen permeable membrane. Oxygen diffuses through the membrane into the biofilm where oxidation of pollutants, supplied from the biofilm side of the membrane takes place. Although full-scale implementation of the MABR has not yet occurred, it is evident, based on laboratory scale data, that the MABR has the potential to outperform several high-rate processes in current use(Syron and Casey 2008). A key benefit of the MABR is the high oxygen utilization efficiency attainable which may confer an economic advantage in terms of aeration energy requirements. Because the energy requirements for aeration and mixing comprise a very significant fraction of the operating costs of aerobic biotreatment processes, the MABR has the potential to offer operational cost savings. This important aspect of MABR performance appraisal has, surprisingly, been scarcely dealt with in the literature. One of the most promising application areas of the MABR is in total nitrogen removal. As a result of the unique microbial stratification profile in MABRs, the potential exists for simultaneous nitrification, denitrification and COD removal in a single biofilm. Nitrifiers, are preferentially located in the oxygen rich region adjacent to the membrane biofilm interface while denitrifiers grow in the anoxic region at the biofilm liquid interface where the COD concentration is typically at its highest value. There are now a significant number of reports confirming the performance of the MABR as a technology

that can achieve simultaneous carbon substrate oxidation, nitrification and denitrification in a single biofilm(Terada et al. 2003; Timberlake et al. 1988).

This study aims to appraise the capital and operating costs of prospective full scale MABR installations applied to COD and nitrogen removal. Central to the analysis is a previously published multispecies mathematical model of the MABR(Shanahan and Semmens 2004). This model is employed to predict COD and dissolved nutrient removal rates for simulated MABR configurations. Comparison is made with the activated sludge process. Using a common basis flowrate and composition, representative of typical municipal wastewater, the model is applied to estimate the process requirements in terms of reactor sizing, membrane area requirements and ancillary equipment sizing. Using established methods for the economic analysis of processes, we estimate capital and operating costs for a number of cases.

METHODS

Design basis

The design basis under investigation is a 10,000 m^3/d flow of wastewater. In case 1 a comparison was made between (a) the conventional activated sludge (CAS) process and (b) an activated sludge system augmented with membrane aeration (referred to as M-AS). Influent COD and ammonia concentrations were 800 mg/L and 50 mg NH_3-N/L respectively and 90% removal was specified. The purpose of the membrane is to provide additional biooxidation capacity via the membrane aerated biofilm whilst maintaining the same suspended biomass concentration as the CAS process. The oxygen transfer efficiency is expected to be higher in the MAS process and the objective is to investigate if this process can achieve overall cost savings when the cost of membranes is considered.

In case 2 we compare systems for total nitrogen removal. Here we consider (a) a complete mix activated sludge system with a preanoxic basin and (b) a membrane aerated biofilm reactor where the oxygen permeable membrane is the sole source of oxygen and the bulk liquid is anoxic. Influent COD and ammonia concentrations were 250 mg/L and 40 mg NH_3-N/L respectively and 90% removal was specified.

Table 1 Kinetic and stoichiometric constants,

Parameter	Value	Units
μ, specific biomass growth rate	3.5	1/d
$\mu_{,N}$ specific biomass growth rate for nitrification	0.12	1/d
k_d, endogenous decay coefficient	0.1	1/d
Y, biomass yield coefficient	0.4	mg_{VSS}/mg_{bsCOD}
f_d, fraction of the biomass that remains as cell debris	0.15	g/g
K_s half saturation coefficient for COD	20	g/m^3
K_n, saturation coefficient for nitrification	0.06	g/m^3
Y_n, biomass yield for nitrification	0.13	g/g

Design calculations for Case 1

For the CAS process, calculation of aeration tank volume, HRT, and oxygen requirements followed well established procedures (Tchobanoglous et al. 2003). In the preliminary comparative analysis presented here, only soluble COD and nitrogenous constituents were considered. The MLVSS value was specified as 2000 g/m^3 in all cases. The area required for secondary clarification were estimated assuming a hydraulic application of 24 m^3/m^2 d(Tchobanoglous et al. 2003). In the case of the membrane augmented process, the CAS design was modified to take into account the role of the membrane attached biofilm, accordingly, the reaction rates associated with the biofilm were evaluated by the Aquasim based model (Shanahan and Semmens 2004). In order to calculate aeration tank volume a mass balance on COD was performed

$$S_0 - S = HRT\left[\frac{kX_L S}{S + K_S} + r_{SB}\right]$$

Where S is the substrate concentration, r_{SB} is the rate of substrate removal by biofilm, X_L is the concentration of suspended biomass in the aeration tank and HRT is the hydraulic retention time. This equation is solved for HRT where all other parameters as specified by the design basis or from Table 1. For a given flow rate the aeration tank volume can be calculated

V=Q (HRT)

The total oxygen requirements for COD oxidation are

$$R_0 = Q\left(S_0 - S\right) - 1.42P_{X,bio}$$

where $P_{X,bio}$ is the biomass as VSS wasted per day.

$$P_{X,bio} = \frac{QY\left(S_0 - S\right)}{1+\left(k_d\right)SRT} + \frac{\left(f_d\right)\left(k_d\right)QY\left(S_0 - S\right)}{1+\left(k_d\right)SRT} + QX_{o,i} + Yr_{SB}V$$

where SRT is the solids retention time.

The aeration demand is split into that supplied by the membranes and that supplied by the diffusers. The aeration capacity required by the air diffusers, r_{OD}, is the total aeration demand minus that supplied by the membranes

$$r_{OD} = R_0 - r_{OB}aV$$

where r_{OB} is the specific aeration rate determined by the model and a is the membrane specific area

For the bubble diffusers the volumetric flowrate of air is:

$$Q_A = \frac{R_0\left(OTE\right)}{24\rho_A\alpha\beta\phi}$$

where ρ_A is the density of air, OTE oxygen transfer efficiency, $\alpha, \beta \,\&\, \phi$ are aeration correction factors, accounting for biomass, wastewater constituents, and temperature.

The blower power consumption is given by

$$P = \frac{P_{A1}T\lambda}{2.73\times10^5\varepsilon\left(\lambda - 1\right)}\left[\left(\frac{P_{A2}}{P_{A1}}\right)^{1-1/\lambda} - 1\right]Q_A$$

where P_{A1} & P_{A2} are the inlet and outlet absolute pressures, ε is the blower efficiency, assumed to be 60%, λ is the ratio of C_P to C_V and has a value of 1.4, T is the inlet temperature. For the membranes, the energy requirements for aeration were calculated according to equations previously described in detail(Semmens 2007). Briefly, power requirements are determined by the gas flow and the compression required to raise the gas pressure. During aeration, oxygen is removed all along the fiber length, and the gas composition within the fiber changes. If an inadequate gas flow is provided, then the oxygen will be depleted within the membrane. The required flowrate, Q, may be calculated from the desired oxygen flux across the membrane, the membrane area, and the oxygen transfer efficiency. The pressure drop required for aeration is linearly related to the oxygen flux across the membrane and the square of membrane length, and inversely related to the third power of the membrane fiber diameter. Smaller fiber diameters will provide a larger surface area/volume for gas transfer but will significantly increase the headloss for gas flow.

Design calculations for Case 2

For the activated sludge process with a preanoxic basin the aeration tank volume, HRT, and oxygen requirements were calculated using the same methods as for case 1, but included additional steps for the calculation of the volume of the anoxic basin and the net oxygen demand(Tchobanoglous et al. 2003). In this case, the oxygen used for nitrification is partially recovered because the produced nitrate is used as an electron acceptor, reducing the required oxygen. For the MABR the total energy requirements comprises both the compressor and liquid mixing, the latter assumed to be 20 kW/1000m^3(Tchobanoglous et al. 2003).For the MABR, a COD removal flux of 30 g m^{-2} day^{-1} was assumed based on expected performance of an optimally configured MABR. For the required COD removal rate it was possible to calculate the membrane area necessary to achieve the treatment objectives in terms of both COD and nitrogen removal. The tank volume and corresponding HRT was calculated by specifying the specific membrane surface area.

Estimation of capital and operating costs

Capital costs were estimated from factors developed for the activated sludge process(Gillot et al. 1999) and were corrected 2008 values. All capital expenditure was expressed on an annualized basis using an average interest rate of 3%, a lifetime of 20 years for construction items, 10 years for pumps, instrumentation and membranes. Operating costs comprised energy for aeration and mixing, assuming a range of electricity costs. A sludge disposal cost of €20/wet tonne (Rosso and Stenstrom 2005) was used throughout. In order to investigate the impact of footprint on overall costs, land costs were assumed to be 1,000,000 € per hectare, typical of an urban location.

RESULTS AND DISCUSSION

Table 2 gives a representative overview of the key process parameters and associated costs for Cases 1 and 2. The data presented were calculated assuming a membrane cost of €40/m^2 and electricity costs of 0.10€/kWh. It should be noted that the membrane delivers approximately one third of the total oxidation capacity in case 1 and 100% in case 2. In terms of the economic feasibility of the various options presented, the most important finding is that significant energy cost savings can be achieved by the use of membrane aeration, however membrane costs comprise a significant portion of the overall process cost and may not be recovered by the savings in energy. Sensitivity analysis revealed that, as far as lifetime total costs are concerned, membrane cost and electricity costs are the critical parameters in defining the relative feasibility of the conventional process to the membrane based process.

Table 2 Representative design figures and summary cost breakdown, all costs in €

	Case 1		Case 2	
	CAS	M-AS	AS with denitrification	MABR
MABR COD removal capacity	-	28	-	30
membrane area (m^2)	-	80,412	-	75,000
specific area of membrane (m^2/m^3)	-	100	-	200
Aeration tank/reactor volume (m^3)	1170	814	3,154	375
Design SRT (day)	5	5	12.5	-
HRT (hr)	2.8	1.9	8	0.9
Total oxygen requirement (kg/d)	6613	6,329	2,343	
Power requirement for aeration (kW)	165.3	101.3	85.9	9
Annualized capital: membranes	0	282,803	0	351,692
Annualized capital: aeration/anoxic tanks	34,233	28,626	62,268	14,043
Annualized capital: secondary settling	36,828	19,916	46,969	5,100
Annualized capital: sludge recycle	6,283	5,536	6,283	0
Annualized capital: land	18,597	15,807	22,934	10,939
Annualized capital: instrumentation	4,037	4,037	4,037	4037
electricity (blowers)	144,785	89,274	75,235	7,848
other electrical	96,583	63,681	50,157	3,000
sludge disposal	60,327	89,527	5,895	5,895
Labor	20,000	20,000	20,000	20,000
Total annualized capital	140,669	385,784	213,767	365,753
Total operational	261,308	271,935	151,256	7,848
Total cost	401,977	558,719	365,054	373,582

At current electricity prices (*ca*. €0.10 per kWh) the membrane based processes show overall cost saving when the membrane cost is less than or equal to €20/m^2 and €40/m^2 for cases 1 and 2 respectively. Figures 1 and 2 illustrate the relationship between these parameters on overall process costs. The total MABR cost is insensitive to electricity cost because operating cost is small relative to the capital investment. In case 2, where the bulk liquid is anoxic and the COD to NH_4 ratio is optimized for total nitrogen removal, there appears to significant cost saving achievable in the MABR in terms of energy, construction and land cost.

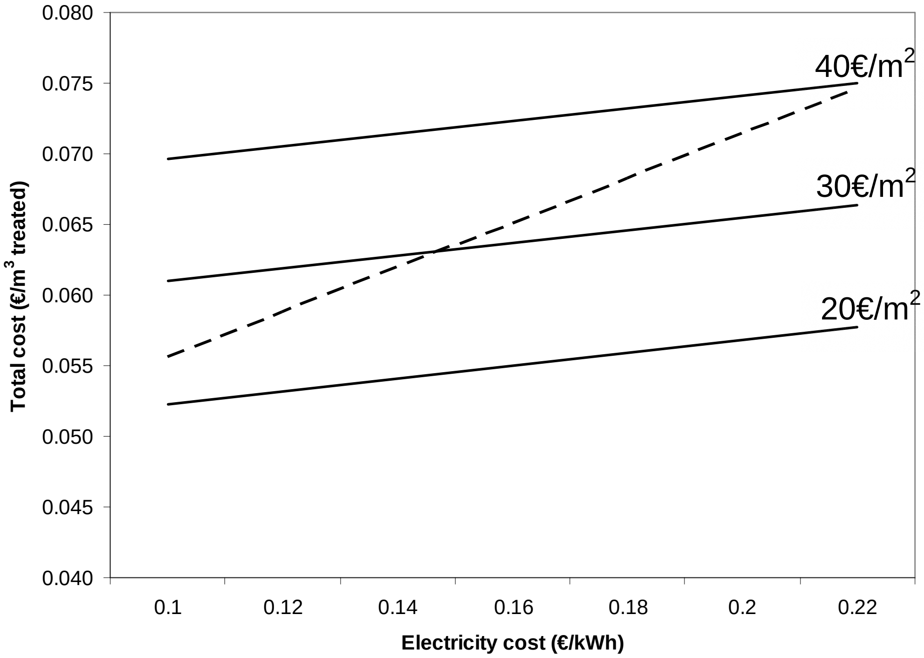

Figure 1. Total costs for the CAS process (----) and M-AS process (——) in terms of electricity cost and membrane cost.

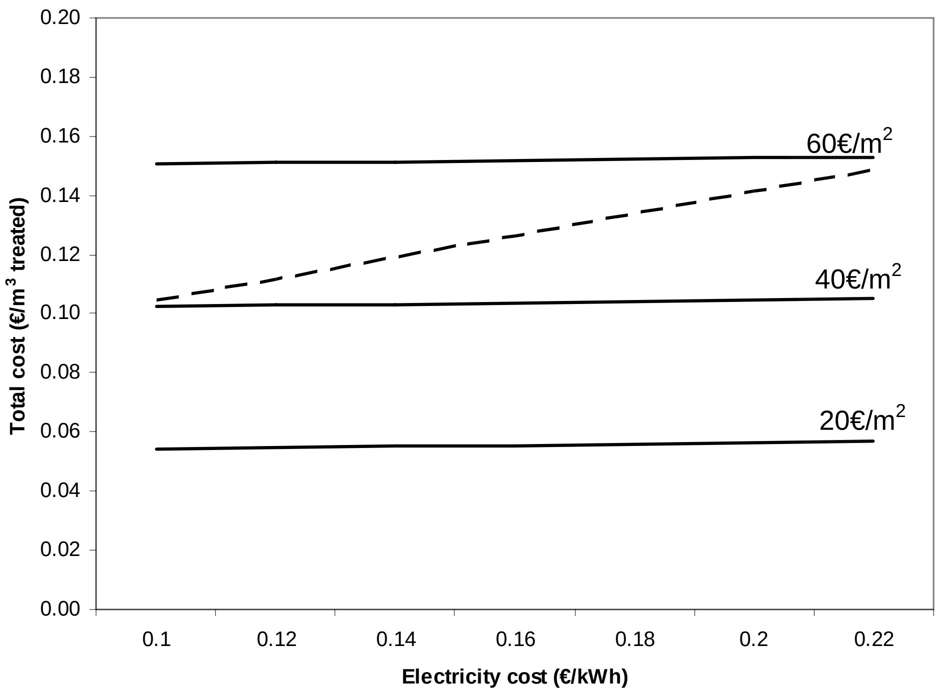

Figure 2. Total costs for the AS process with denitrification (----) and MABR process (——) in terms of electricity cost and membrane cost.

CONCLUSIONS

The MABR is receiving increased attention as a technology that can increase the oxygen transfer efficiency in wastewater treatment processes. This study indicates that membrane replacement cost is the principle economic obstacle to the commercial development of the technology. The cost of membranes has decreased significantly in recent years (Judd 2006) and this trend is expected to continue. A corresponding increase in the cost of electricity suggests that the MABR may, in the coming years, be developed as a technology that can offer cost savings in the wastewater treatment sector. However, at present, it appears to offer the best economic performance when operated as a system for combined COD total nitrogen removal. The results of this study suggest that it is unlikely to offer any cost benefit when operated as a system to supplement aeration in conventional activated sludge processes. In further developing the MABR, particular attention needs to be addressed at developing membrane module designs which are capable of providing optimized conditions for biological oxidation, long lifespan and reliable operation. In addition there is a need to ensure biofilm thickness control measures are in place to ensure reliable long term operation.

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