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REVIEW ARTICLE

Modelling the energy demands of aerobic and anaerobic membrane bioreactors for wastewater treatment

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A modelling study has been developed in which the energy requirements of aerobic and anaerobic membrane bioreactors (MBRs) are assessed in order to compare these two wastewater treatment technologies. The model took into consideration the aeration required for biological oxidation in aerobic MBRs (AeMBRs), the energy recovery from methane production in anaerobic MBRs (AnMBRs) and the energy demands of operating submerged and sidestream membrane configurations. Aeration and membrane energy demands were estimated based on previously developed modelling studies populated with operational data from the literature. Given the difference in sludge production between aerobic and anaerobic systems, the model was benchmarked by assuming high sludge retention times or complete retention of solids in both AeMBRs and AnMBRs. Analysis of biogas production in AnMBRs revealed that the heat required to achieve mesophilic temperatures (35°C) in the reactor was only possible with influent wastewater strengths above 4–5 g COD L^{-1}. The general trend of the submerged configuration, which is less energy intensive than the sidestream configuration in aerobic systems, was not observed in AnMBRs, mainly due to the wide variation in gas demand utilized in anaerobic systems. Compared to AeMBRs, for which the energy requirements were estimated to approach 2 kWh m^{-3} (influent up to 1 g COD L^{-1}), the energy demands associated with fouling control in AnMBRs were lower (0.80 kWh m^{-3} for influent of 1.14 g COD L^{-1}), although due to the low fluxes reported in the literature capital costs associated with membrane material would be three times higher than this.

Keywords: energy; membrane bioreactors; submerged; sidestream; crossflow

1. Introduction

Aerobic membrane bioreactors represent a specific subset of bioreactor technology in which the membrane replaces alternative means of solid–liquid separation, such as a gravity sedimentation tank. The inclusion of a membrane results in the complete uncoupling of the hydraulic and sludge retention times, providing greater operational flexibility and the potential to intensify the biological process. The advantages and disadvantages of MBRs are often quoted [1], but perhaps their key advantages are improved effluent quality and lower sludge production. Their disadvantage is membrane fouling, with an associated high energy demand. Despite this disadvantage, several studies have shown that the membrane bioreactor market is expected to grow in industrial and municipal applications, both in Europe [2] and in North America [3]. The principal applications driving market growth relate to situations where tight effluent consents have to be met, a small footprint is required, robust disinfection is required or the water is to be reused [1].

Anaerobic biological processes are mainly applied to high strength industrial wastewaters at mesophilic temperatures as an alternative to aerobic treatment. Their use results from the lower energy demand due to the absence of aeration [4], the possibility of recovering energy from the methane in the biogas produced, and reduced biomass production and its associated disposal costs. However, the main drawbacks of anaerobic treatment are the lower quality effluent generated, especially when operating with low strength wastewaters (0.3–0.7 g COD L^{-1}) at low temperatures (8–25°C), the possibility of generating odours, and the need for downstream nutrient removal. Traditionally, anaerobic reactors have utilized granular sludge as a method of biomass retention [5], although in more recent studies the potential for using membranes has been discussed [6]. In particular, benefit has been reported from their high solids

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retention, even at low temperatures, and the rejection of high molecular weight organics, which are further degraded and which would otherwise be lost in the effluent.

Translating the concept to municipal wastewater treatment, the adoption of anaerobic membrane bioreactors (AnMBRs) will result in a reduction both in energy usage and in sludge production. Given the current demand to reduce the energy and carbon footprint of wastewater treatment, consideration of AnMBRs seems timely. Recent reviews concerned with AnMBRs have looked at the impact of operational factors on biological performance [7] and the parameters affecting membrane flux [8]. The main conclusion drawn from these reviews is the need to assess the feasibility of both sidestream and submerged configurations to arrive at optimum fouling control strategies and minimize the overall energy demand.

The current study addresses this point by comparing energy balances for AnMBRs and aerobic membrane bioreactors (AeMBRs) for the treatment of low strength municipal wastewater. The aim of the study has been to establish the overall changes in energy that are to be expected, as well as identifying the critical components controlling the overall energy balance, in order to indicate where future improvements may be achieved.

2. Model development

The overall assessment compares the energy balance across. AeMBRs and AnMBRs. The main components of the model have included: biological aeration (AeMBRs), energy recovery from methane production (AnMBRs) and energy demands associated with the operation of either a sidestream or a submerged system. The aeration and membrane energy demands of submerged systems have been estimated based on model studies previously developed [9,10]. In the case of sidestream configuration energy demands for sludge pumping, these were estimated based on pressure loss calculations along the membrane modules using a previously validated rheological model [11]. Wherever possible the new models have been populated with published operational data from the literature, and standard assumptions have been made where necessary.

2.1. Biological aeration

Aeration demand for AeMBRs were based on the model of Verrecht *et al.* [10], adapted to using literature data as input parameters instead of direct kinetic modelling of the biological reactor. Aeration energy was calculated with respect to the oxygen consumed (MO_2) by heterotrophic (organic pollutant oxidation) and autotrophic (nitrifying) bacteria, taking into account the oxygen reduction due to denitrification in the anoxic zones of the bioreactor (Equation (1)).

$$\begin{aligned} MO_2 = Q(\mathrm{COD_{IN}} - \mathrm{COD_{EFF}}) + 4.33 \\ Q(\mathrm{TN_{IN}} - \mathrm{TN_{EFF}}) - 2.83Q\mathrm{NO_{3,EFF}}, \end{aligned} \tag{1}$$

where Q represents the influent flow (L d^{-1}) and COD_{IN} and COD_{EFF} are the influent and effluent COD concentrations, respectively (mg COD L^{-1}). The remaining terms are related to the oxygen required for nitrification estimated from the difference between total influent and effluent nitrogen (TN_{IN} and TN_{EFF}, respectively), and also taking into account the amount of oxygen saved by denitrification calculated on the effluent nitrate concentration, $NO_3^-{}_{EFF}$ (mg NO_3^-–N L^{-1}). Sludge wastage is

Table 1. AeMBR case studies and respective operational conditions utilized to calculate energy demand [12–15].

Parameter	Units	Innocenti *et al.* [12]	Laera *et al.* [13]	Rosenberger *et al.* [14]	Teck *et al.* [15]
Volume	L	1400	6	3500	20
HRT	h	14	8	14	8
SRT	d	∞	∞	∞	300
MLSS	g L^{-1}	16.6	22.9	16	18
MLVSS	g L^{-1}	8.7	17.2	11.2	16
COD_{IN}	mg L^{-1}	300	400	790	1000
COD_{EFF}	mg L^{-1}	19	57	10	5
TN_{IN}	mg L^{-1}	42.2	49.3	65.8	10.0
TN_{EFF}	mg L^{-1}	2	0.8	13	0.3
$NO_3^-{}_{EFF}$	mg L^{-1}	11.3	40.6	13	0.3

HRT: hydraulic retention time; SRT: sludge retention time; MLSS: mixed liquid suspended solids; MLVSS: mixed liquid volatile solids; COD_{IN}: chemical oxygen demand of the influent; COD_{EFF}: chemical oxygen demand of the effluent; TN_{IN}: total nitrogen in the influent; TN_{EFF}: total nitrogen in the effluent; $NO_3^-{}_{EFF}$: total nitrate in the effluent.

not included in the COD balance, since AeMBRs operating at high sludge retention times (SRTs) or with complete retention of solids, were principally considered in order to provide a direct benchmark for the anaerobic systems [12–15] (Table 1).

In the case of submerged MBRs, biological aeration demands, represented by $Q_{air,bio}$ (Equation (2)), were estimated considering the contribution of membrane gas scouring ($MO_{2,MEM}$) (Equation (3)) to the overall oxygen requirement (MO_2), taking into account that the air flow required to control fouling ($Q_{air,MEM}$) was provided by coarse bubble diffusers, which are more effective in their scouring effect than fine bubble diffusers, but offer lower oxygen transfer efficiency [10].

$$Q_{\mathrm{air,bio}} = \frac{MO_2 - MO_{2,\mathrm{MEM}}}{0.21 \cdot 1000 \cdot \alpha \cdot \beta \cdot \gamma \cdot \mathrm{OTE}}, \tag{2}$$

$$MO_{2,\mathrm{MEM}} = 0.21 \cdot 1000 \cdot \alpha \cdot \beta \cdot \gamma \cdot \mathrm{OTE}_{\mathrm{MEM}} \cdot \mathrm{Q}_{\mathrm{air,MEM}}, \tag{3}$$

where β and γ are oxygen transfer efficiency factors, taken as 0.95 and 0.89, respectively [10]. Oxygen transfer efficiencies (OTEs) of 0.02 and 0.05 were assumed for membrane coarse bubble aeration and biological fine bubble diffusers, while the factor, α, which accounts for the oxygen transfer efficiency, relates to the mixed liquid suspended solids (MLSS) concentration, according to Equation (4) [16].

$$\alpha = e^{-0.082\,\mathrm{MLSS}} \tag{4}$$

The power requirements associated with biological aeration ($E_{air,bio}$) were obtained from the power consumption of a blower (Equation (5)) delivering the corresponding air flow at the static pressure of the liquid column in the membrane tank (h), which was considered to be 2 m wherever it was not reported directly.

$$E_{\mathrm{air,bio}} = \frac{108{:}748 \cdot \lambda}{\xi \cdot (\lambda - 1)} \left[\left(\left(\frac{p \cdot g \cdot h + 101325}{101321} \right)^{(1-\frac{1}{\lambda})} - 1 \right) \right] \cdot Q_{\mathrm{air,bio}}. \tag{5}$$

In Equation (5), ξ represents the blower efficiency (60%), ρ is the air density, g is the gravity constant and λ is the heat capacity ratio, which has a value of 1.4 for air and 1.3 for biogas.

2.2. *Methane production*

Under anaerobic conditions, biodegradation of organics takes place without the need for oxygen or nitrate as electron acceptor. The end-products are mainly methane and carbon dioxide, which can either be recovered as biogas or are dissolved in the effluent. The energy associated with methane production in AnMBRs has been calculated from the biogas production data reported in the literature, assuming an energy content of 36500 kJ m^{-3} in the biogas produced [17]. The highest methane yields reported for AnMBRs treating low strength wastewater range between 0.29 and 0.33 L CH_4 g^{-1} COD [18,19], and are given in studies in which soluble and completely biodegradable substrate were employed as influent at mesophilic temperatures (35–37°C). Lower methane yields of 0.12 and 0.08–0.09 L CH_4 g^{-1} COD [20,21] have been reported for synthetic wastewaters, but at temperatures ranging between 11–25°C and 25–30°C, respectively, highlighting the importance of temperature in the production and recovery of biogas. Variable degrees of methanisation in AnMBRs treating actual influents have ranged between 0.20–0.23 L CH_4 g^{-1} COD for screened wastewater [22], 0.27 L CH_4 g^{-1} COD for raw wastewater [23] and 0.09–0.12 L CH_4 g^{-1} COD for black water [24].

2.3. *Membrane energy demands in submerged and sidestream configuration*

Energy demands associated with membrane operation in AeMBRs and AnMBRs were divided into components related to permeate pumping (E_{PER}), and to fouling control (E_{FC}). In submerged MBRs, membranes are immersed in the mixed liquor and gas is sparged below the membrane module in the form of air or biogas, in the case of AeMBRs and AnMBRs, respectively. In sidestream pumped crossflow operation the membrane module is located outside the bioreactor and the mixed liquor is pumped through the membrane module and recycled back to the bioreactor. This provides enough turbulence to enhance the back transport of foulants from its surface, thus reducing membrane fouling. Although sidestream MBRs are usually operated at constant pressure, with the pump generating both the liquid crossflow and the driving force for permeation, for direct comparison all systems were assumed to utilize permeate pumping, irrespective of system configuration. The energy required for E_{PER} was calculated as the product of permeate flow (Q_p) and the transmembrane pressure (TMP) (Equation (6)).

$$E_{\mathrm{PER}} = Q_{\mathrm{p}}\,\mathrm{TMP} \tag{6}$$

The energy required for fouling control (E_{FC}) was further subdivided into pumping ($E_{FC,P}$) and gas sparging ($E_{FC,G}$) for pumped crossflow and submerged configurations, respectively. The power consumption required for fouling control in submerged configuration ($E_{FC,G}$) was obtained by replacing the biological aeration requirements ($Q_{air,bio}$) in Equation (5) by the membrane gas demand ($Q_{GAS,MEM}$), calculated according to Equation (7):

$$Q_{GAS,MEM} = SGD_m \cdot A_m, \quad (7)$$

where SGD_m denotes the specific gas demand normalized against membrane area ($m^3\ m^{-2}\ h^{-1}$), which is equivalent to the specific aeration demand ($SAD_{m)}$ employed in aerobic systems, and A_m represents membrane filtration area. Although different relationships between flux and specific gas demand required (SGD_m) have been proposed in AeMBRs [1,10], standard 50% intermittent gas sparging intensities of 0.3 and 0.5 $m^3\ m^{-2}\ h^{-1}$ were employed for fluxes below and above 15 L $m^{-2}\ h^{-1}$ (LMH), respectively, for the aerobic case studies shown in Table 1, all of which employed hollow fibre membranes. In the case of AnMBRs, the SGD_m reported in the different case studies was used to estimate the energy demands for fouling control in the submerged configuration.

In the pumped sidestream configuration the energy associated with fouling control was calculated as the product of the tangential flow (Q_{CFV}) ($m^3\ s^{-1}$) through the membrane, and the pressure loss ΔP (Pa), assuming a pump efficiency (ξ of 60% (Equation (8)).

$$E_{CF,P} = \frac{Q_{CFV} \cdot \Delta p}{\xi} = \frac{CFV \cdot S_m \cdot \Delta p}{\xi}, \quad (8)$$

where Q_{CFV} was obtained from the crossflow velocity (CFV, m s^{-1}), and the cross-sectional area (S_m) was calculated from the geometric characteristics of the membrane modules reported in different case studies. Pressure losses (ΔP) were estimated using the Darcy–Weisbach equation (Equation (9)), using the diameter (D) and length (L) of the membrane module, and the Fanning friction factor (f) was calculating according to Colebrook's relationship (Equation (10)). Chilton and Stainsby [11] introduced a modified Reynolds number (RE) (Equation (11)), in which the effective viscosity was calculated at the wall and which could be applied to the general Herschel–Bulkley model based on the previously defined generalized Reynolds number [25]. The parameter X obtained by solving Equation (11) represents the ratio between the yield stresses of the sludge (τ_B) and the shear stress at the membrane wall (τ_w). Comparison of theoretical results based on rheological characterization of wastewater sludges and kaolin slurries were able to predict to within 15% the experimentally measured pressure drop for crossflow velocities between 0.1 m s^{-1} and 6 m s^{-1}, and therefore appear appropriate for the present case.

$$\Delta P = \frac{4\,p\,f\,CFV^2}{2D} L \quad (\text{turbulent conditions}), \quad (9)$$

$$f^{-0.5} = 4\log_{10}(RE'\,f^{0.5}) - 0.4, \quad (10)$$

$$RE = \frac{RE}{(1-X)^4}, \quad (11)$$

$$\tau_W = \mu_B(1-X)^{-1}(1-\frac{1}{3}X-\frac{1}{3}X^2-\frac{1}{3}X^3)^{-1}. \quad (12)$$

The rheological parameters were obtained from the characterization of AeMBR and AnMBR sludge given by Laera *et al.* [26] and Pevere *et al.* [27], respectively, in which the Bingham shear stress (τ_B) (Pa) and viscosity (μ_B) (Pa s) were reported to increase with biomass concentration according to Equations (13) [26] and (14) [27]. A Bingham plastic rheological model was preferred to represent the behaviour of MBR sludge based on constant viscosity presented in different studies at shear rates exceeding 100–500 s^{-1} [26,27,29]. Additionally, adoption of the power law models which have been also been proposed [26–29], resulted in a decrease in frictional pressure losses with increasing solid concentrations due to the more prominent shear thinning behaviour that both aerobic and anaerobic sludges present when fitted to this rheological model.

$$\mu_B = 0.02894\ MLSS$$
$$\tau_B = 0.001(0.233\ MLSS + 1), \quad (13)$$

$$\mu_B = 0.001e^{0.04\,MLSS} \qquad \tau_B = 0.067e^{0.07\,MLSS}. \quad (14)$$

3. Results and discussion

3.1. Energy balances in aerobic and anaerobic MBRs

Analysis of the energy demands in AeMBRs with complete sludge retention shows that, irrespective of the organic load applied, the total energy demand approaches 2 kWh m^{-3} for wastewater of 0.4 g COD L^{-1} strength or above (Figure 1).

The aeration required for biological pollutant oxidation represents as much as 88–93% of the total energy

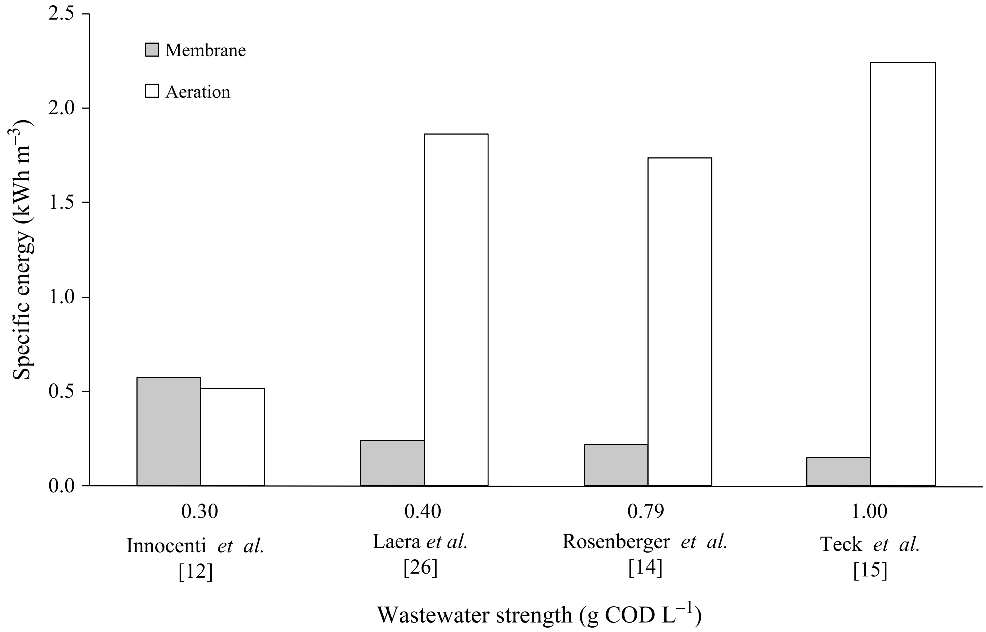

Figure 1. Membrane (grey bars) and aeration (white bars) energy demands in AeMBRs with complete sludge retention, calculated using published operational data [12,14,15,26] for wastewater strength between 0.30 and 1.00 g COD L^{-1}.

demand, except for the lowest influent COD concentration modelled, 0.3 g COD L^{-1}, in which membrane aeration provided 40% of the biological oxygen requirements and accounted for 60% of the total energy demand (1.2 kWh m^{-3}). Operation of AeMBRs at high sludge age, or even complete retention of solids, would avoid the high cost of sludge treatment and disposal, which can represent between 40% and 60% of total cost [30], without having a negative effect on biological performance [13–15,31]. The high levels of MLSS accumulated in the system result in a high energy requirement, due to the reduction in aeration efficiency to below 20% [16,31]. As a result, in order to optimize operational costs, aerobic membrane bioreactors treating municipal wastewater are usually operated at hydraulic retention times below 12 h and variable sludge retention times of between 15 and 50 days, which are intended to achieve mixed liquor concentrations between 8 and 12 g MLSS L^{-1} [1,10]. Different surveys have estimated their energy requirements to range between 0.6 and 1.2 kWh m^{-3} [32] distributed between membrane (60–70%) and biological aeration (30–40%). Consequently, an additional 0.8–1.4 kWh m^{-3} of energy is required to operate an AeMBR at low biomass production.

Results from model calculations reveal that the energy demands in submerged AnMBRs range from 0.03 to 5.7 kWh m^{-3} (Figure 2). Such variability in energy requirements for fouling control arises as a result of the wide range of gas demands, reported in submerged configuration between effectively no gas sparging [20,22,33] and 3 MLH [18,34]. In sidestream AnMBRs energy demands range between 0.23 and 16.52 kWh m^{-3} (Figure 3), the variation being attributed to the impact of crossflow velocity and bioreactor MLSS on flux and pressure losses. In contrast to AeMBRs, energy demands for submerged and sidestream AnMBR systems are within the same range, due to the higher fluxes reported for crossflow systems, and also to the uncertainty of appropriate gas sparging rates that result in sustainable membrane operation in the submerged configuration, which will be discussed in the following section.

However, data given in the literature indicate that the available (electrical) energy produced ranges from 0.15 to 0.3 kWh m^{-3} as the wastewater strength increases from 0.24 to 1.14 g COD L^{-1} [22,24], demonstrating that the methane generated (expressed as heat in Figures 2 and 3) is sufficient to recover a significant proportion of the total energy demand for wastewater treatment, and even offset the energy demand associated to fouling control.

It is important to note that the reported methane production in some studies [18,19,23,24,35] has been obtained at mesophilic temperatures, and therefore lower biogas production can be expected at ambient wastewater temperatures (8–25°C) in temperate climates, given that biogas production from low strength wastewater is insufficient to provide enough energy to allow the reactor to be heated to 35°C. This is illustrated by horizontal lines in Figures 2 and 3, which indicate the amount of energy required to heat the reactor to 35°C from an influent wastewater temperature of 15°C, taking into account that 50% of the energy could

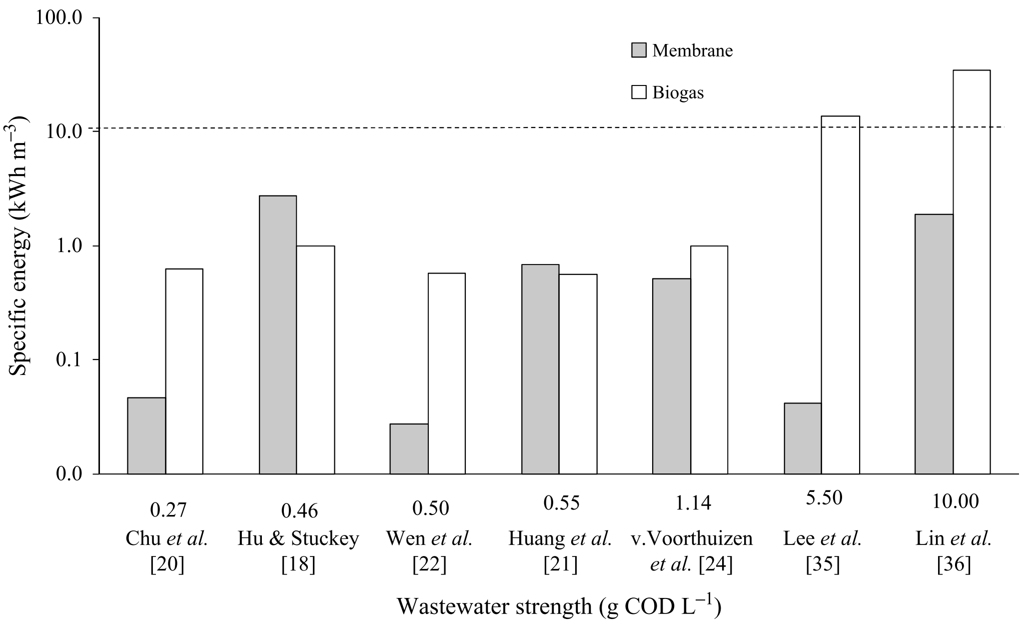

Figure 2. Membrane energy demand (grey bars) and energy produced from biogas converted to heat (white bars) in submerged AnMBRs with complete sludge retention, for wastewater strength between 0.27 and 10.00 g COD L^{-1} [18,20–22,24,35,36]. The horizontal dashed line indicates the amount of energy required to heat the reactor to 35°C from an influent wastewater temperature of 15°C, taking into account that 50% of the energy could be recovered by heating the influent with the permeate.

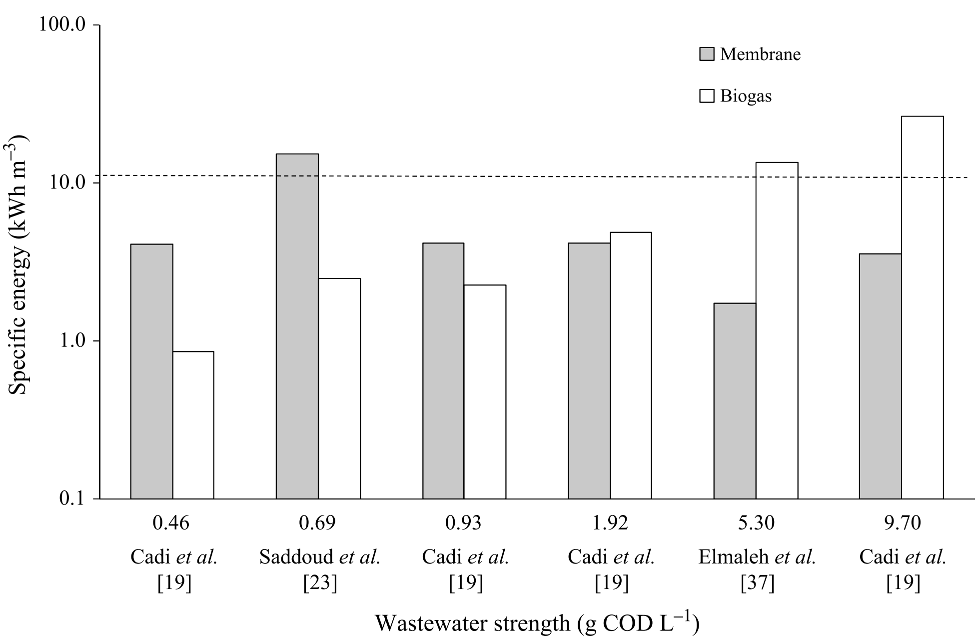

Figure 3. Membrane energy demand (grey bars) and energy produced from biogas converted to heat (white bars) in sidestream AnMBRs with complete sludge retention, for wastewater strength between 0.27 and 10.00 g COD L^{-1} [19,23,37]. The horizontal dashed line indicates the amount of energy required to heat the reactor to 35°C from an influent wastewater temperature of 15°C, taking into account that 50% of the energy could be recovered by heating the influent with the permeate.

be recovered by heating the influent with the permeate. The heat recovered from biogas increases from 0.62 to 34.8 kWh m^{-3} as wastewater strength increases from 0.24 g to 10 g COD L^{-1} [20,36]. For domestic wastewater applications it therefore is energetically preferable to operate without heating, as the energy required to heat the reactor is only achieved for influent COD concentrations above 4–5 g COD L^{-1}.

For high strength wastewaters, methane production not only covers the heat balance but can also generate 5–20 kWh m^{-3} of electrical power for export [38]. Indeed, commercial applications of AnMBR technology, employing both sidestream and submerged membranes, have been developed for different industrial and agricultural wastes. For example, the Biorek process developed in Denmark by BIOSCAN A/S [39] comprises an anaerobic digester, coupled to sidestream crossflow membrane filtration to remove organic pollutants, and the effluent is treated by reverse osmosis to recover clean water and a nutrient-rich concentrate. More recently, Kubota has commercialized a submerged anaerobic membrane fermentation unit (KSAMBR) for application to distillery wastewaters, solid waste bio-gasification facilities and food processing factories, with the advantage of reducing the carbon footprint of the digester to between one-third and one-fifth that of a conventional digester, while reducing the organic load of aerobic post-treatment polishing [40]. A specific case study of the treatment of high strength Shochu wastewater with an organic and volatile solids concentrations of 182 g COD $Kg^{-1}_{residue}$ and 34 g VS L^{-1}, respectively, showed that 12 GJ d^{-1} were generated, and that electricity consumption accounted for a quarter of this energy.

3.2. *Impact of operational parameters on the energy requirements of submerged AeMBRs and AnMBRs*

Specific energy demand in submerged configurations is determined by the relationship between the applied gas sparging rate and permeate flux. Although results do not usually correlate between different studies, a general trend exists in aerobic systems, consisting of a linear increase in flux up to a certain gas sparging intensity, following which the attainable flux remains constant. At the upper end of the reported range, Le Clech *et al.* [41] have found an increase in critical flux from 88 to 121 m^3 m^{-2} h^{-1} as SGD_m is increased from 0.73 to 2.3 m^3 m^{-2} h^{-1}, equating to energy demands of 0.08 and 0.17 kWh m^{-3} with tubular membranes (Table 2). Similarly, in a pilot scale study in which full scale flat sheet membrane modules were employed [42], specific gas demands of 0.5 to 0.94 m^3 m^{-2} h^{-1} resulted in critical fluxes ranging from 28 to 40 LMH, corresponding to energy requirements of 0.20 and 0.26 kWh m^{-3}, respectively. A parallel study with hollow fibre membranes [43] showed that a SGD of 0.5 m^3 m^{-2} h^{-1} was sufficient to achieve a flux of 31 LMH, while a reduction to 0.3 m^3 m^{-2} h^{-1} yielded a critical flux of 25 LMH, resulting in energy requirements of 0.14 and 0.18 kWh m^{-3}, respectively.

There is an overall consensus in the literature concerning the range of gas sparging intensities which have been shown to impact on permeate flux at 0.52–0.88 and 0.32–0.5 m^3 m^{-2} h^{-1} applied to flat sheet and hollow fibre configurations, respectively [44]. Fluxes varied between 8 and 30 LMH, with 80% of plants running below 20 LMH, which is consistent with the observed critical flux range of 30–40 LMH. However, while the critical flux represents the transition between fast and slow fouling rate, long-term operation requires only a fraction of the maximum flux to be applied, which extends the filtration cycle before chemical cleaning is required. As an illustration, fluxes of 10 and 22 LMH, representing a reduction of 50–60% with respect to critical flux, and for SGD_m of 0.5 and 0.94 m^3 m^{-2} h^{-1}, would appear to conserve operational conditions, leading to energy demands of 0.56 and 0.48 kWh m^{-3} in hollow fibre and flat sheet systems, respectively, as an increase in resistance to filtration was not observed over prolonged periods [42,43]. On the other hand, a flux reduction of 25–30% (below the corresponding critical flux) at the same gas demand resulted in fouling rates of 5 and 15 mbar d^{-1} for the flat sheet and hollow fibre membranes. This illustrates the flux window to which membrane operation is subjected.

Table 2. Categorization of energy demand associated with membrane operation in submerged AeMBRs [42–47].

Area (m^2)	SGDm (m^3 m^{-2} h^{-1})	Flux (LMH)	TMP (kPa)	E_{CF} (kWh m^{-3})	E_{PER} (kWh m^{-3})	E_{TOT} (kWh m^{-3})	Reference
0.21	0.73	88^{CF}	–	0.08	–	0.08	Le Clech *et al.* [45]
0.21	2.3	121^{CF}	–	0.17	–	0.17	
40	0.94	28^{SF}	7–15	0.38	$3.2–6.8 \times 10^{-3}$	0.38	Guglielmi *et al.* [43]
40	0.94	23^{SF}	6	0.48	2.8×10^{-3}	0.48	
69.6	0.5	10^{SF}	8	0.56	3.7×10^{-3}	0.57	Guglielmi *et al.* [42]
69.6	0.5	22^{SF}	17–25	0.26	$7.9–12 \times 10^{-3}$	0.26	
0.24	0.75	20^{SF}	7.5	0.42	3.5×10^{-3}	0.43	Zhang *et al.* [46]
0.5	0.3	18^{CF}	12	0.15	5.6×10^{-3}	0.16	Bouhabila *et al.* [47]
0.5	0.8	25^{CF}	15	0.29	6.9×10^{-3}	0.30	

SGD_m: specific gas demand, normalized; TMP: transmembrane pressure; E_{CF}: Energy required to control fouling; E_{PER}: energy required to pump permeate; E_{TOT}: total energy requirement; CF: Critical flux; SF: Sustainable flux.

Table 3. Categorization of energy demands associated with membrane operation in submerged anaerobic MBRs [18,20–22,24,35,36,48].

Area (m^2)	SGDm ($m^3\ m^{-2}\ h^{-1}$)	Flux (lmh)	TMP (kPa)	E_{CF} ($kWh\ m^{-3}$)	E_{PER} ($kWh\ m^{-3}$)	E_{TOT} ($kWh\ m^{-3}$)	Reference
0.10	0	10.4	100	0.00	4.6E–02	0.05	Chu *et al.* [20]
0.30	0	5.0	60	0.00	2.8E–02	0.03	Wen *et al.* [22]
0.24	0.40	5.3	0.07	0.69	3.2E–03	0.69	Huang *et al.* [21]
0.10	3.00	8.0	38	3.41	1.8E–02	3.43	Hu and Stuckey [18]
0.04	0.54	10.0	50	0.49	2.3E–02	0.51	van Voorthuizen *et al.* [24]
0.04	1.35	8.0	50	1.53	2.3E–02	1.56	
0.05	1.80	25.0	30	0.65	1.4E–02	0.67	Wu *et al.* [48]
0.03	1.50	7.2	10	1.89	4.6E–03	1.90	Lin *et al.* [36]
0.03	1.50	2.4	27	5.68	1.3E–02	5.70	
0.10	3.00	5.0	20	5.46	9.3E–03	5.47	Lee *et al.*[35]

SGD_m: specific gas demand normalized; TMP: transmembrane pressure; E_{CF}: Energy required to control fouling; E_{PER}: energy required to pump permeate; E_{TOT}: total energy requirements; CF: Critical flux; SF: Sustainable flux.

Information concerning the operational significance of sparging rates in submerged AnMBRs is currently less clear. Critical flux analysis has revealed that gas sparging is less effective for enhancing permeate flux compared to AeMBRs, due to the presence of high concentrations of colloidal matter and fine solids in anaerobic biomass [24,36,49]. Increasing specific gas demands from 0.78 to 3.76 $m^3\ m^{-2}\ h^{-1}$ increased the critical flux from 6 to 10 LMH. This is consistent with the literature data (Table 3), which show that while fluxes range only between 5 and 10 LMH, specific gas demands as high as 3 $m^3\ m^{-2}\ h^{-1}$ have been quoted [18,34]. Although such high gas sparging intensity levels have extended membrane operation, they do not result in an increase in permeability. At the highest SGD_m of 3 $m^3\ m^{-2}\ h^{-1}$, equivalent to an energy demand of 3.43 kWh m^3, a stable permeability of 20 LMH bar^{-1} was observed over 90 days' operation [18], while a higher permeability of 60–70 LMH bar^{-1} was reported during 450 h in a mesophilic AnMBR treating ethanol at a specific gas demand of 1.5 $m^3\ m^{-2}\ h^{-1}$ [36], equating to an energy demand of 1.9 kWh m^{-3}. A fouling resistance of 0.5 to 3% in the membrane was reported after 150 days' operation by Huang *et al.* [21] in two AnMBRs at SRT of 30 to 60 days, by applying a SGD of 0.4 $m^3\ m^{-2}\ h^{-1}$, which equates to an energy consumption of 0.69 kWh m^{-3}.

The lowest energy demands, of 0.02–0.05 kWh m^{-3}, reported for submerged AnMBRs (Table 3) result from permeate suction and correspond to studies in which membrane filtration has been coupled with high rate anaerobic reactors in which fouling is controlled without using gas sparging, by only applying intermittent filtration [20,22] or by relying on the shear provided by upflow velocity [33]. Results have shown that although sustainable operation could not be achieved, fouling rates between 10 mbar d^{-1} [33] and 100 mbar d^{-1} [20] were observed due to the low solid and colloidal load on the membrane [24]. These fouling rates are of the same order of magnitude as residual fouling rates (15–35 mbar d^{-1}) and one to two orders of magnitude lower than the 2000–2600 mbar d^{-1} value reported for cake layer fouling in backwash cycles in aerobic systems [42]. Implementation of backwashing, together with low gas sparging intensity in high rate anaerobic reactors coupled to membrane filtration, could therefore result in an efficient fouling control strategy for AnMBRs, as suggested in a previous study involving a low solid anoxic membrane system [50]. Duration of filtration cycles of 10 min experimentally determined from analysis of the critical mass deposited on the membrane, which leads to irreversible fouling, resulted in fluxes of 20 LMH maintained over 20 days, equating to specific energy demands of 0.05 kWh m^{-3} for gas sparging.

3.3. *Impact of operational parameters on the energy demands of sidestream AeMBRs and AnMBRs*

Analysis of the pressure drop per m generated when pumping both aerobic and anaerobic sludges (Figure 4) reveals that when compared at the same MLSS and crossflow velocity (CFV) aerobic biomass leads to higher frictional pressure losses than does anaerobic biomass, reflecting the higher viscosity of the former. As an illustration of this, the curve corresponding to the AeMBR sludge at the lowest MLSS concentration, 5 g MLSS L^{-1}, overlaps with the maximum concentration of 20 g MLSS L^{-1} for the anaerobic sludge. Similarly, the rate of increase of pressure loss with increasing CFV is higher in aerobic systems than in AnMBRs. The

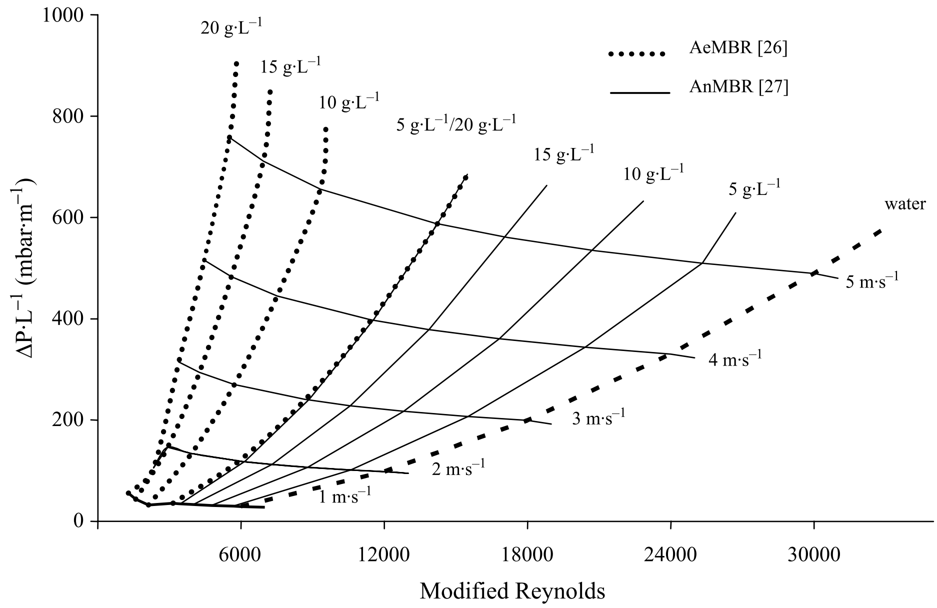

Figure 4. Influence of mixed liquid suspended solids (MLSS) and crossflow velocity (CFV) on unitary pressure drop (ΔP) for aerobic and anaerobic MBR sludges, assuming a 6 mm diameter lumen [26,27].

effect of CFV on pressure drop (ΔP) is more prominent than the increase in MLSS over the range considered. Taking a reference point of 10 g MLSS L^{-1} and 3 m s^{-1} in the AeMBR, increasing CFV to 4 m s^{-1} results in a unitary pressure loss of 444 mbar m^{-1}, while the resulting value for a biomass concentration of 15 g MLSS L^{-1} is 293 mbar m^{-1}. However, the rate of increase in pressure loss with solids concentration is more prominent if MLSS is above 15 g L^{-1}.

Energy demands for fouling control in AeMBRs ranged from 0.67 to 7.42 kWh m^{-3}, depending on crossflow velocity and MLSS concentration (Table 4). Increasing crossflow velocity and biomass concentration results in higher energy expenditures in sidestream systems due to a combination of higher volumetric flow and pressure losses in the first case, and to the lower critical fluxes reported at higher MLSS in the second [51,52]. For instance, at a solid concentration of 10 g MLSS L^{-1} [52], specific power requirements range from 0.67 to 4.25 kWh m^{-3}, with CFV varying from 1 to 4 m s^{-1}, while for a higher MLSS of 15 g MLSS L^{-1} there is a more pronounced increase in energy demand, from 0.38 to 5.14 kWh m^{-3} [51]. To further illustrate the influence of biomass concentration, at a crossflow velocity of 3 m s^{-1} energy demand increases from 0.67 to 3.10 kWh m^{-3} as MLSS increases from 2.1 to 15 g MLSS L^{-1}, due to a decrease in critical flux from 270 to 80 LMH [51].

In comparison the energy demands for fouling control in sidestream AnMBRs range between 0.23 and 16.50 kWh m^{-3} showing similar trends to AeMBRs with respect to crossflow velocity and biomass concentration (Table 5).

However, despite presenting lower flux than their aerobic counterpart, lower energy demands are predicted in AnMBRs at similar biomass concentration and crossflow velocity due to the lower pressure losses arising from the lower viscosity. For instance, at a solid concentration of 15 g MLSS L^{-1}, CFV varying from 1.6 to 3.4 m s^{-1} results in an increase in energy demand from 0.88 to 3.01 kWh m^{-3} in an anaerobic system [56] and from 0.38 to 5.14 kWh m^{-3} in AeMBRs operating at the same biomass concentration [51]. The lowest energy demands in sidestream AnMBRs occur either in suspended growth reactors operating at low solid concentrations or when crossflow filtration is coupled with attached growth systems [57], due to a combination of low frictional pressure losses and high flux. Fluxes of 252 and 120 LMH [37,56] at corresponding crossflow velocities of 2.6 and 3.5 m s^{-1}, are among the highest reported for AnMBRs and result in energy demands of 0.48 kWh m^{-3} and 1.45 kWh m^{-3}, respectively, although a further reduction to 0.23 kWh m^{-3} is predicted for crossflow velocities in of 0.93 m s^{-1} [57].

3.4. Implications of the anaerobic flow-sheet

Key findings from the modelling assessment of the literature reporting AnMBR case studies lead to a number of conclusions:

(1) The energy required to heat the bioreactor to mesophilic conditions can only be compensated at wastewater strengths above 4–5 g COD L^{-1}. When considering the treatment of municipal

wastewater in AnMBRs it is therefore economically unfavourable to operate under mesophilic conditions. In this situation net energy recovery from wastewater can only be feasible at low temperatures.

(2) A number of features of the energy associated with membrane operation are unique to AnMBRs, which prohibits direct transfer of knowledge from study of AeMBRs. The general trend of submerged membrane systems as less energy intensive is not observed in AnMBRs, due to the lower fluxes observed and the uncertainty of the appropriate gas sparging rates required for sustainable operation.

(3) From an energy point of view, the most effective AnMBR configuration appears to be a combination of high rate anaerobic systems and membranes. Assessment of energy demands of systems in which biomass retention is not solely controlled by the membrane have shown either very low energy demands and moderate fouling at low fluxes, or moderate energy demands together with high fluxes in the sidestream configuration. Low solid membrane feed could be the key to achieving low energy demands.

(4) Comparison of energy demands of AeMBRs and AnMBRs, assuming complete retention of solids, suggests that although the latter would have a lower energy demand, the capital cost of the membrane material would be as much as three times higher, as a result of the lower fluxes encountered in anaerobic systems.

Table 4. Categorization of energy demand associated with membrane operation in sidestream AeMBRs [31,51–55].

MLSS (g L^{-1})	CFV (m s^{-1})	Flux (LMH)	TMP (kPa)	ΔP (kPa)	$E_{CF,P}$ (kWh m^{-3})	E_{PER} (kWh m^{-3})	E_{TOT} (kWh m^{-3})	Reference
2.1	3	270	nr	28.4	0.67	–	0.67	Cicek *et al.* [51]
4.6		180	180	31.0	1.10	0.08	1.18	
8.6		95	nr	34.3	2.3	–	2.3	
15.4		80	nr	38.9	3.10	–	3.1	
10	1	20	10	6.3	0.67	< 0.01	0.67	Defrance *et al.* [52]
	2	40	25	11.0	1.30	0.01	1.32	
	3	80	60	21.4	2.99	0.03	3.02	
	4	115	80	65.5	4.54	0.04	4.58	
	5	130	120	107.2	7.37	0.06	7.42	
1.8	1.6	10	1–3	12.5	3.3	< 0.01	3.3	Ognier *et al.* [53]
8	4	100	60–90	24.7	4.59	0.04	4.63	Tardieu *et al.* [54]
10–50	1–5	40	20–30	–	–	0.01	0.6–0.75^{+}	Muller *et al.* [31]
4.5–6	0.4	20–100	100	–	–	0.05	2^{+}	Zhang *et al.* [55]

MLSS: mixed liquid suspended solids; CFV: crossflow velocity; TMP: transmembrane pressure; ΔP: pressure losses; $E_{CF,P}$: energy required for pumping and to control fouling; E_{PER}: energy required to pump permeate; E_{TOT}: total energy requirements; nr: not recorded.

Table 5. Categorization of energy demands associated with membrane operation in sidestream anaerobic MBRs [19,23,37,49,50].

MLSS (g L^{-1})	CFV (m s^{-1})	Flux (LMH)	TMP (kPa)	ΔP (kPa)	$E_{CF,P}$ (kWh m^{-3})	E_{PER} (kWh m^{-3})	E_{TOT} (kWh m^{-3})	Reference
0.4	2.6	252	90	21.5	0.44	0.04	0.48	Beaubien *et al.* [56]
7	2.6	108	80	23.1	1.33	0.03	1.36	
15	2.6	72	80	25.1	1.81	0.03	1.84	
10.8–14.5	2	15–18	40–45	15.2–15.7	3.51–4.04	0.02	3.53–4.06	Cadi *et al.* [19]
10	3	9	200	45.2	16.42	0.1	16.52	Saddoud *et al.* [23] *
0.18	3.5	120	50	10.4	1.43	0.02	1.45	Elmaleh and Abdelmoumni [37]
0.15–0.5	0.93	30	40	0.3	0.21	0.02	0.23	Cho and Fane [57]

MLSS: mixed liquid suspended solids; CFV: crossflow velocity; TMP: transmembrane pressure; ΔP: pressure losses; $E_{CF,P}$: energy required for pumping and to control fouling; E_{PER}: energy required to pump permeate; E_{TOT}: total energy requirements; * based on 8 mm lumen diameter.

4. Conclusions

Potential savings in sludge treatment and disposal costs can be achieved in both AeMBRs and AnMBRs, since operation under complete retention of solids is possible. Assessment of model calculations has shown that in AeMBRs low biomass production is attainable with an energy input of 2 kWh m^{-3}, due to the low oxygen transfer efficiency at high biomass concentrations. In AnMBRs fouling control is the determining factor in the energy demand of the process, and sludge production and energy requirements are therefore less directly linked, although extended SRT is likely to influence biomass characteristics, and hence membrane performance. A wide variation in energy demand in submerged AnMBRs, ranging from 0.03 to 3.57 kWh m^{-3} has been found, highlighting the need to investigate further the gas intensity required to control fouling in order to ascertain the full potential of such technology for mainstream wastewater treatment.

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