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**Energy analysis and optimization of hollow fiber membrane contactors for recovery of
dissolve methane from anaerobic membrane bioreactor effluent**

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Abstract

This work presents an energy analysis and optimization of the hollow fiber membrane contactors for the recovery of dissolved methane (CH_4) in effluents of anaerobic membrane bioreactor wastewater treatment processes. The obtained CH_4 could be merged with biogas for further purification or used with a micro-turbine for electricity generation to achieve an energy self-sufficient wastewater treatment process. A mathematical model considering simultaneous CH_4 and carbon dioxide (CO_2) desorption was used to estimate the membrane area required to remove the dissolved CH_4 , as well as quality of the outlet gas from the membrane contactor. Energy balance between electrical energy obtained from the recovered CH_4 and energies consumed by vacuum and liquid pumps for the operation of membrane contactor were investigated and reported as a *Net Electricity* obtained per m^3 of effluent or simply *Net E*. Results revealed that a combination of a high strip gas flow rate and slightly low vacuum condition closed to the atmospheric pressure can provide the highest *Net E* at 0.178 MJ/m^3 . This value is 85.37% of the total electrical energy that can generated from a 90% recovery of dissolved CH_4 using an effluent saturated with a 60% vol CH_4 biogas and flow rate at $2 \text{ m}^3/\text{day}$. The calculation was made based on the assumptions that 1) the membrane contactor is operated in a non-wetting mode where membrane properties remain constant, 2) flux decline due to the membrane fouling is not considered and 3) the energy required for membrane cleaning and other relevant activities are not factored into the energy analysis. Based on our results, to obtain a high CH_4 mole fraction at the gas outlet, a low strip gas flow rate is recommended, however, the operating gas pressure needs to be lowered by applying a vacuum condition to improve the *Net E*. In addition, it was found that the *Net E* could be improved by increasing the number of membrane fibers, and lowering the liquid flow rate. The CH_4 recovery efficiency could also be optimized to obtain an optimal *Net E*.

Keywords

Biogas; Membrane contactor; Anaerobic treatment; Net energy; Hollow Fiber

1. Introduction

Methane (CH_4) is a combustible gas and one of the major energy sources used in several important applications, such as electricity generation, transportation, and as a precursor for high-value chemical products. The demand for CH_4 is escalating as driven by the constant search for a cleaner and more sustainable energy resource to replace the fossil fuels. As CH_4 can be obtained from anaerobic digestions of biomass, wastewater treatment by anaerobic processes starts to receive tremendous research attention in the past decades. In addition to the production of CH_4 , the anaerobic treatment of wastewaters offers other competitive advantages over aerobic treatments, such as higher process stability, lower energy expenditure, and sludge production [1]. To date, the anaerobic treatments and energy recovery from several types of domestic and industrial wastewaters have been successfully developed and investigated.

The anaerobic treatment utilizes bacteria and other microorganisms to digest organic matters in the wastewater for production of CH_4 in the form of biogas (a gas mixture contains mostly CH_4 and carbon dioxide, CO_2). Since the treatment process is operated in a completely closed environment, the dissolved gases in the produced effluent is in an equilibrium with the biogas in the headspace, resulting in a significant amount of gases, including CH_4 , lost in the effluent solution. Lui et al.[2] pointed out that for an anaerobic treatment of municipal wastewater with an averaged chemical oxygen demand (COD) at 200 mg/L and 303K, 45% of the produced CH_4 would be theoretically in the dissolved form and the number will be higher at lower

temperatures. A review study on the significance of dissolved CH_4 in anaerobic effluents by Crone et al. [3] revealed that the amount of dissolved CH_4 in the effluents depends on the type of anaerobic processes. The degree of CH_4 saturation or ratio between the actual dissolved CH_4 per theoretical value calculated from the Henry's constant was found to vary between 1.34 to 6.9 for effluent of an upflow anaerobic sludge blanket (UASB) and 1 to 1.5 for effluent of an anaerobic membrane bioreactors (AnMBR). This is due to the difference in the liquid retention time of the systems. Potentially, the dissolved CH_4 is an energy source, which upon recovery, can help to increase the total energy production of the anaerobic treatment processes. Otherwise, it would be discharged into the environment and, without proper disposal, the leaked CH_4 is a major greenhouse gas that has a global warming potential of 21 times more powerful than that of CO_2 . Therefore, treating dissolved CH_4 are strongly recommended in order for the wastewater treatment to stay as a carbon neutral process [4].

Several technologies have been applied thus far to treat the dissolved CH_4 in the effluents of anaerobic wastewater treatment processes. One of them uses methane oxidizing bacteria to biologically treat the dissolved CH_4 to form final products such as hydrocarbons and CO_2 [5]. However, for this technology, the CH_4 would be lost and cannot be recovered. Alternatively, there are three other mainstream techniques for extraction of CH_4 from the anaerobic effluents. They are aeration, gas stripping and membrane processes. The advantages and disadvantages of each technique and equipment used are summarized in Table 1. Among these techniques, CH_4 recovery by the membrane contactors appears very promising. This is because membrane contactors possess very high surface area per unit volume and allow liquid and gas phases to come into contact, without dispersing of one phase into the other, thus avoiding problems present

in techniques which use phase mixing processes like foaming and flooding. Figure 1 shows the membrane contactor system for the CH₄ recovery from an effluent of anaerobic digestion process. Microporous hollow fiber membranes are usually applied for degassing applications since they offer a higher gas desorption rate than those of dense membranes. However, dense membranes can also be selected for the treatment of effluents with high residual organic materials to prevent membrane pore clogging problems. A sweep gas or vacuum condition is applied to one side of the membrane and flows counter-currently to the effluent flow which is feeding from the other side of the membrane. In this way, the dissolved gases in the effluent can then be desorbed and diffused through the membrane pores driven either by a concentration or partial pressure gradients.

Table 1. Techniques and technologies for the extraction of dissolved CH₄ from effluents of anaerobic waste treatment processes.

Technique	Pros	Cons	Process equipment	Details
Aeration: Mass transfer area is created by introducing bubbles of carrier gases into solution	- Simple	- Phase mixing	Bubble columns	The bubbles are created at the bottom of the reactor and flow upward.
	- High mass transfer area	- Difficult to scale up and operate	Multi-stage bubble column cross-flow cascades	Several bubble columns connect in series. Carrier gas and solution flow counter currently.
		- Difficult to control quality of product gas - Low CH ₄ concentration in the gas outlet	Forced draft aeration	Vacuum is applied at the top of the bubble column
Gas stripping: Mass transfer area is provided by the packing materials	- Scale up is straight forward	- Relatively low mass transfer area	Air stripping packed column	A cylindrical column is packed with small subjects either in random packing or structured packing. Carrier gas and solution flow counter currently.
	- Easy to operate	- Flooding and foaming problems - High gas/liquid flow rate ratios are required	Vacuum packed tower	A vacuum condition is applied in the packed column
			Down-flow hanging sponge reactor	Sponge cubes are used as the packing material.

Membrane:	- High mass transfer area	- Additional mass transfer resistance	Membrane contactors	Effluent and gas are fed counter
Micro porous or dense membranes are used as phase barriers.	- Easy to operate and scale up	- Membranes need to be periodically replaced		currently in opposite sides of the membrane. Vacuum or sweep gas mode are commonly used.

Fig. 1. A membrane contactor for recovery of CH₄ from an effluent of anaerobic digestion process

Despite its promise, to this end, there are only a few research articles on using the membrane contactor for recovery of CH₄ from effluents of anaerobic processes. Bandara et al. [6] demonstrated a successful integration of a membrane contactor unit into a bench-scale UASB reactor and obtained an averaged dissolved CH₄ recovery efficiencies of 77%. Since membranes hold the key to the membrane contactor performances, Cookney et al. [7] further evaluated two types of membranes, including micro-porous and dense hollow fiber membranes for CH₄ recovery from effluents of both AnMBR and UASB reactors, respectively. Results showed that the micro-porous membrane provided higher CH₄ removal efficiencies than the dense membrane when compared within a same range of liquid velocities. McLeod et al. later found that 97% CH₄ removal efficiency by a micro-porous membrane contactor was achievable. This far exceeded the

88% CH₄ recovery that was necessary for the process to reach carbon neutrality [4]. Following this, the membrane contactor process was employed for CO₂ desorption from a liquid effluent of UASB reactor by Luo et al. [8] to increase the CH₄ concentration in the produced biogas by removing CO₂ from the system. Recently, Wongchitphimon et al. [9] developed highly hydrophobic hollow membranes especially targeted at CH₄ recovery application. The fibers showed a more superior performance than a commercial polypropylene membrane. In addition, the energy requirement to operate the membrane contactor for CH₄ recovery from effluents of an anaerobic process was roughly estimated by Cookney et al. [10]. However, the data reported was for one specific experimental condition and only the energy requirement for a vacuum pump was considered in the calculation, which was far from realistic. To operate the membrane contactors, a liquid pump is also required to deliver the solution to the system and this should be included in the energy analysis to provide a closer energy requirement of the membrane contactor.

In this work, a detailed energy analysis of the membrane contactors for the recovery of CH₄ from the effluent of an anaerobic waste treatment process was performed. To obtain the most accuracy estimation of the energy consumptions for both the vacuum and liquid pumps for the membrane contactor, a mathematical mass transfer model was used to determine the membrane area required to remove a given dissolved CH₄ concentration. A plug flow model considering simultaneous desorption of dissolved CH₄ and CO₂, which was successfully validated with experimental results using a real effluent of AnMBR in our previous work [11] was employed. This model takes into account all the effects from relevant physical parameters and uses local mass transfer driving forces to estimate the gas desorption along the length of the membrane modules. The energy analysis was reported as *Net Electricity* obtained per m³ of effluent or

simply *Net E*. Several operating parameters, including gas pressures, strip gas and effluent flow rates were optimized in order to investigate their effects and to obtain a highest possible *Net E* from the recovered CH₄ under various operating conditions.

2. Modeling

2.1 Modeling for simultaneous desorption of dissolved CH₄ and CO₂ in the membrane contactor

Fig. 2 shows the mass balance in the gas and liquid side along a length of membrane module.

Two different approaches using either a vacuum or strip gas condition can be applied to supply a mass transfer driving force necessary for the gas desorption. In the vacuum mode, a vacuum pump is installed in the gas outlet line to create a difference in the partial pressure between the bulk gas and at the gas-liquid interface. For the sweep gas mode, a carrier gas is fed to the gas inlet and flows counter-currently to the liquid phase to create gas desorption by concentration gradient differences. Using the sweep gas mode requires less energy consumption as compared to the vacuum mode, thus making it more attractive for the recovery of dissolved CH₄. However, high gas flow rates are required, which would significantly reduce the product gas concentration. Therefore, to get the best of both worlds, we have chosen in this work to operate the membrane contactor in a combo mode where both vacuum and a strip gas are employed as shown in Fig. 3.

Fig. 2. Mass balance in a small element of the membrane contactor

Fig. 3. The combo mode using both a strip gas and vacuum pump to recovery dissolved CH_4 .

The system of differential equations for the mass balance of dissolved CH_4 and CO_2 in liquid phase and striped gases in gas phases in the Z direction (as shown in Fig. 2) can be written as;

Gas phase

$$\frac{dF_{\text{CH}_4,G}}{dZ} = n_f \pi d_o N_{\text{CH}_4} \quad (1)$$

$$\frac{dF_{\text{CO}_2,G}}{dZ} = n_f \pi d_o N_{\text{CO}_2} \quad (2)$$

Liquid phase

$$\frac{dF_{\text{CH}_4,L}}{dZ} = n_f \pi d_o N_{\text{CH}_4} \quad (3)$$

$$\frac{dF_{\text{CO}_2,L}}{dZ} = n_f \pi d_o N_{\text{CO}_2} \quad (4)$$

where F_i is the molar flow rate of gas i in gas phase (G) or dissolve gas in liquid phase (kmol/s), N_i is the desorption flux of gas i (kmol/m².s), d_o is the diameters of the hollow fiber membranes (m) and n_f is the number of fiber.

N_i can be written in terms of the driving force and the overall mass transfer coefficient based on the liquid phase (K_{OV} , m/s) is as follows:

$$N_i = K_{OV}(C_{i,L} - H_i p_i) \quad (5)$$

where H is the Henry's constant (kmol/atm.m³), C_i (kmol/m³) and p_i (atm) are the concentration and partial pressure of gas i in the liquid and gas phases, respectively.

p_i of CH₄ and CO₂ and $C_{i,L}$ as the function of liquid flow rate (L_m , m³/s) are calculated as follows:

$$p_i = \left(\frac{F_{iG}}{F_{CH_4,G} + F_{CO_2,G} + F_{inert,G}} \right) P_g \quad (6)$$

$$C_{i,L} = F_{i,L} / L_m \quad (7)$$

The overall mass transfer resistances (K_{OV}) can be expressed using a resistance-in-series model for a physical desorption as shown in the following equation:

$$\frac{1}{K_{OV}d_o} = \left[\frac{1}{k_L d_i} + \frac{RT_G H}{k_M d_{ln}} + \frac{RT_G H}{k_G d_o} \right] \quad (8)$$

where k_L , k_M and k_G are the individual liquid, membrane and gas phases (m/s) mass transfer coefficients, respectively. d_o , d_{ln} , and d_i are the outer, log mean and inside diameters of the hollow fiber membranes (m), respectively, R is the gas constant (kmol.K/m³.atm), T_G is the temperature in the gas side (K), N_i is the desorption flux of gas i (kmol/m².s).

2.2 Individual mass transfer coefficients

2.2.1 Liquid side mass transfer coefficient (k_L)

The Lvque solution ($Sh = 1.62Gz^{1/3}$) has been widely applied to predict the liquid phase mass transfer coefficient in the tube side (k_L) of the membrane contactor by using Graetz number (Gz) to describe the fluid behaviors [12]. However, as reported by McLeod et al. [4], there was an overestimation of the Sherwood number (Sh) for CH_4 desorption at low Gz numbers (Fig. 4). For CH_4 desorption, high recovery efficiency will be obtained by using long membrane modules or in a low Gz number range ($Gz < 5$). Hence, in order to obtain the best accuracy prediction of k_L at a low Gz , the Lvque solution was modified by applying extension terms and using the curve fitting technique in Microsoft Excel to determine their new coefficients. The modified equation is as shown below:

$$Sh = \frac{k_L d_i}{D_{i,L}} = 2.162Gz^{1/3} - 0.0154Gz - 1.235 \quad (9)$$

Results for the above equation was compared to the experimental results from McLeod et al. [4] and other correlations in the literature as shown in Fig. 4. It can be seen that the proposed equation gives satisfying results compared with the experimental data when $Gz > 0.4$ given that the mean square error (MSE) is 0.053. In addition, the calculated Sh from Eq. 9. were also in good agreement with Sh from the Lvque and Kreulen equations when $20 > Gz > 100$. Therefore, Eq. 9 is valid when Gz is in the range of 0.4 to 100.

Fig. 4. Comparison between experimental and calculated Sherwood numbers from Eq. 9 and those from the literature.

2.2.2 Gas side mass transfer coefficient (k_G)

For the gas flowing in the shell side of membrane module, Yang and Cussler [16] proposed the following correlation to predict the gas side mass transfer coefficient for both gas absorption and stripping:

$$Sh_{i,G} = \frac{k_{i,G} d_e}{D_{i,G}} = 1.25 \left(\frac{d_e}{L} Re_G \right)^{0.93} Sc_G^{0.33} \quad (10)$$

where $D_{i,G}$ is the gas diffusivity in gas phase (m^2/s) and d_e is the hydraulic diameter (m). Eq. 10 is applicable for $0 < Re_G < 500$ as mentioned in Zheng et al. [17].

2.2.3 Membrane mass transfer coefficient (k_M)

In the case of a non-wetted mode where only gases fill the membrane pores, the membrane mass transfer coefficient (k_M) can be calculated using the following equation:

$$k_M = \frac{D_{G,eff}\varepsilon_M}{\tau_M\delta_M} \quad (11)$$

$D_{G,eff}$ is the effective diffusion coefficient of the gas in the pores (m²/s), which is calculated by

the combination of molecular and Knudsen diffusivities as shown below:

$$\frac{1}{D_{G,eff}} = \frac{1}{D_M} + \frac{1}{D_{Kn}} \quad (12)$$

where D_M and D_{Kn} are the molecular and Knudsen diffusion coefficients (m²/s), respectively.

2.3 Energy analysis of the membrane contactor for the CH₄ recovery

2.3.1 Energy consumption for vacuum pump

The energy consumption of a vacuum pump can be calculated by the following equation [18]:

$$W_{VP} = \frac{F_{TOT,g}RT_G\kappa}{1000*(1-\kappa)\eta_{VP}} * \left(\frac{P_{G,in}}{P_{G,out}} \right)^{(\kappa-1)/\kappa} - 1 \quad (13)$$

where, W_{VP} is the energy consumption for the vacuum pump (kW), κ is the adiabatic constant, η_{VP} is the vacuum pump efficiency, $P_{g,in}$ and $P_{g,out}$ are the pressures in the suction and exhaust ports of the vacuum pump (Pa), respectively. κ and η_{VP} can be calculated as follows:

$$\kappa = \kappa_{CH_4} * y_{CH_4} + \kappa_{CO_2} * y_{CO_2} \quad (14)$$

$$\eta_{VP} = 0.1058 \ln \left(\frac{P_{G,in}}{P_{G,out}} \right) + 0.8746 \quad (15)$$

The vacuum pump requires a cooling system to operate effectively. Therefore, the energy consumption for a cooling system ($W_{VP,cooling}$) is also included [18]. The energy required for the cooling system can be calculated as a function of the vacuum pump efficiency as shown below:

$$W_{VP,cooling} = 0.054\eta_{VP}W_{VP} \quad (16)$$

As a result, the total energy required to operate a vacuum pump is calculated by Eq. 19 and represented as:

$$W_{VP,tot} = W_{VP} + W_{VP,cooling} \quad (17)$$

2.3.2 Energy consumption for liquid pump

The energy consumption is equal to the energy required for the liquid pump to overcome the pressure drop of the liquid flowing in hollow fiber modules. This can be calculated from the following equation [19];

$$W_{LP} = Q\Delta P_L / 36\eta_{pump}\eta_{tran}\eta_{motor} \quad (18)$$

where Q is the pump capacity which is set to equate the volumetric liquid flow rate (m^3/hr), η_{pump} , η_{tran} , and η_{motor} are the efficiencies of the pump, transmission systems and electric motor, respectively, and are assumed to be equal at 70%, ΔP_L is the liquid pressure drop (Pa) in the lumen side of membrane module and is given based on the Hagen-Poiseuille equation:

$$\Delta P_{liquid} = \frac{8\mu_{H2O}V_L Z}{d_i^2} \quad (19)$$

where μ_{H2O} is the viscosity of water ($kg.m/s$) and V_L is the liquid velocity (m/s).

The total energy consumption of the vacuum, and liquid pumps in the form of electricity required for the treatment of $1 m^3$ of effluent (MJ/m^3) can then be calculated from the following equation:

$$E_{pumps} = \frac{86.4*(W_{VP,tot}+W_{LP})}{L_{m,day}} \quad (20)$$

The constant in the above equation is the conversion factor for converting the energy consumption (kW) to energy consumed per day (MJ/day). $L_{m,day}$ is the liquid flow rate per day (m^3/day).

2.3.3 Net Electricity obtained per m^3 of treated effluent (*Net E*)

In this work, the recovered energy from the dissolved CH_4 in the effluent of anaerobic wastewater treatment process is converted to electricity using a micro-turbine by the following equation:

$$E_{MT} = \eta_{MT} * \left(y_{CH_4} * F_{tot,day} * \omega_{CH_4} / L_{m,day} \right) \quad (21)$$

where η_{MT} is the micro-turbine efficiency assumed at 30%, ω_{CH_4} is the energy content of CH_4 at a value of 38 MJ/m^3 , $F_{tot,day}$ is the product gas flow rate (m^3/day).

The difference in energy obtained and consumed as electricity in the membrane contacting process or *Net E* is calculated by the following equation:

$$Net E = E_{MT} - E_{pumps} \quad (22)$$

where *Net E* is the net electricity obtained from the membrane contactor for the treatment of 1 m^3 of effluent.

2.4. Model validation, basic operating conditions, and modelling parameters

The described mass balance model in section 2.1 was successfully validated in our previous work in which an anaerobic membrane bioreactor effluent was used in experiments and where model implementation and equations for physical parameters of both gas and liquid phases can

be found [11]. In the model validation experiment, Gz numbers of the liquid phase and Re_G were between 15 to 70 and 1.2 to 2.3, respectively, which were also within the validity ranges for k_L in Eq. 9 and $k_{i,G}$ in Eq. 10. Although the correlation for k_L was changed to the new modified equation (Eq. 9), the simulation results in this work were very closed to our previous work on the model validation experiments given that both of them operated using a Gz number ranging from 15 to 70 in the liquid phase. Accordingly, our modified equation (Eq. 9) and the Lvque solution gave insignificant difference in the calculated Sh numbers (see Fig 4).

The membrane used was an in-house hollow fiber hydrophobic microporous membrane fabricated using polyimide (Matrimid, MT) as a substrate, and followed by cyclic imide ring-opening reaction and modification with fluorinated silica nanoparticles [9]. The membrane properties and basic operating conditions are listed in table 2. The module packing density was set at 0.14 for all simulations. Therefore, the module diameter was altered to allow a constant packing density when the number of fibers were changed to study their effects. Table 3 displays the values of important gas and effluent properties used in this simulation at 298.15 K. The correlations for the calculations of these parameters were also shown in our previous work [11].

It should be noted that the following assumptions have been made for the calculation of energy analysis:

1. Membrane fouling and drop in the performance in time were not considered.
2. Energy required for membrane cleaning and relevant activities was not considered in this energy analysis.
3. Non-wetting mode of membrane performance as ideal case scenario was considered.
4. Optimization of hollow fiber membrane contactors did not include the optimization of the membrane itself, which was fixed and its properties were taken from [9].

When a target CH₄ recovery efficiency is fixed (constant E_{MT}), the Net E is greatly affected by three parameters including height of the membrane module required to remove the dissolved CH₄ that is influencing the W_{LP} , gas outlet pressure, and flow rate that is relating to the $W_{VP,tot}$. Applying a higher strip gas or lower gas pressure to reduce the membrane height and W_{LP} also results in an increase of the gas outlet flow rate and $W_{VP,tot}$. In addition, the system is made more comprehensive by simultaneous desorption of CO₂ that directly enhance the gas outlet flow rate. Therefore, the system is now more complicated, and the operating parameters need to be carefully adjusted to reach an optimum point with the highest Net E possible.

Table 2. Membrane and module properties as well as the basic operating conditions.

Parameters	Values
Fiber o.d./i.d. (μm)	945/670
Pore size (nm)	46.5
Porosity	0.66
Tortuosity	2
Module packing density	0.14
Gas and liquid temperature (K)	298.15
Liquid pressure (atm)	1
Liquid flow rate (m ³ /day),(L/min)	2, 1.39

Table 3. Values of important gas and liquid parameters at 298.15 K.

Parameters	Specifications	Values
Liquid phase	density (kg/m ³)	997.27
	viscosity (kg/m.s)	8.96 x 10 ⁻⁴
	diffusivity of CH ₄ (m ² /s)	1.46 x 10 ⁻⁸
	diffusivity of CO ₂ (m ² /s)	5.9 x 10 ⁻⁹
Gas phase	diffusivity of CH ₄ (m ² /s)	2.2 x 10 ⁻⁵

	diffusivity of CO ₂ (m ² /s)	1.52 x10 ⁻⁵
Henry's constant	CH ₄ (kmol/atm.m ³)	1.386 x 10 ⁻³
	CO ₂ (kmol/atm.m ³)	3.41 x 10 ⁻²

3. Results and discussion

3.1 Optimization of *Net E*

In this section, an attempt is made to optimize operating parameters in the membrane contactor in order to obtain a maximum *Net E*. For all the simulations, the solution is saturated with biogas containing 60%vol CH₄ in balance of CO₂ (the dissolved CH₄ and CO₂ concentrations are 0.00083 and 0.0136 mol/L, respectively). The gas and liquid temperatures are at 298.15 K and the operating liquid pressure is set constantly at 1 atm. The number of fibers and packing density of the membrane module are 650 and 0.14, respectively. These numbers are chosen to give the same packing density and obtain a liquid velocity at 0.1 m/s which is in the range of our model validation experiments [11]. Under each operating condition, the length of the membrane module is varied for the membrane contactor to achieve 90% CH₄ recovery efficiency. These basic operating parameters and those shown in Table 2 are used in all the simulations in this section, unless otherwise stated. The energy gained (E_{MT}) without considering the energy consumed by the pumps is 0.2085 MJ/m³ and the recovered CH₄ flow rate is approximately at 0.026 L/min.

Fig. 5. Effects of gas pressure on *Net E* and CH₄ mole fraction in the gas outlet. (Liquid flow rate 2 m³/day, liquid pressure 1 atm, gas and liquid temperatures 298.15 K, number of fibers 650, and packing density 0.14)

Fig. 6. Effect of gas pressure on energy consumed by vacuum and liquid pumps. (Inert gas flow rate: 0.07 L/min, liquid flow rate: 2 m³/day, liquid pressure: 1 atm, gas and liquid temperatures: 298.15 K, number of fibers: 650, and packing density: 0.14)

3.1.1 Effect of gas pressures

The effect of operating gas pressure on the *Net E* and the outlet CH₄ mole fraction in the product gas are displayed in Fig. 5. The solid and dotted black lines are for the inlet strip gas flow rate at 0.07 and 0.1 L/min, respectively. *Gz* numbers in the liquid phase and *Re_G* are varied from 0.49 to 1.46, and 0.23 to 0.29, respectively. It can be seen that the *Net E* is significantly affected by varying the gas pressure. The *Net E* increases with the gas pressure to a certain point (0.65 and 0.8 atm for using the inert gas flow rates at 0.07 and 0.1 L/min, respectively), after which, the opposite trend is found at the higher gas pressures as the *Net E* reduces.

The increase in the operating gas pressure from the vacuum condition to the atmospheric pressure also rises the *y*_{CH₄} in the gas outlet which implies that CO₂ can be desorbed to a greater extent at lower gas pressures. This is due to the higher CO₂ polarity that allows it to be more dissolved in water and creates a larger mass transfer driving force for the desorption. In the simultaneous desorption of CH₄ and CO₂, the dissolved CO₂ enables faster transfer to the gas phase and rapidly reaches its equilibrium concentration with dissolved CO₂ remaining in the effluent (values of $H_{CO_2} * p_{CO_2}$ and $C_{CO_2,L}$ are equal in Eq. 5). This limits further desorption of CO₂ to the gas phase, while CH₄ as the target gas can be desorbed continuously throughout the membrane module. The reduction of gas pressure lowers the equilibrium concentration in the gas phase of CO₂ ($H_{CO_2} * p_{CO_2}$), thus significantly enhance the total desorbed CO₂, and lowers the CH₄ concentration in the gas phase since the amount of desorbed CH₄ was fixed at 90% recovery efficiency. It should be mentioned that *y*_{CH₄} in the gas outlets of these simulations are in between 0.10 to 0.16, which are already higher than the lower flammable limit of CH₄ (0.05) but still lower than the minimum requirement for direct use in a micro-turbine (0.35) [4]. Therefore,

we recommend to combine this product gas with the biogas line coming from the headspace of the anaerobic bioreactor (see Fig. 1) to achieve a higher CH₄ concentration for use in the micro-turbine.

The energy consumptions for the liquid (E_{LP}) and gas pumps ($E_{VP,tot}$), as well as the total energy consumed for the treatment of 1 m³ effluent (E_{pumps}) using a strip gas flow rate of 0.07 L/min at different operating gas pressures are displayed in Fig. 6. The lower gas pressure decreases the bulk gas partial pressures and hence increases the mass transfer driving force and reduces the length of membrane module required to remove the dissolved CH₄. Consequently, this results in a reduction of the energy consumption for the liquid pump (E_{LP}) while increasing the energy consumption for the vacuum pump ($E_{VP,tot}$). It is observed that the E_{pumps} is lowest at 0.6 atm and increases when the gas pressure decreases further. This is because of the notable increase in the $E_{VP,tot}$ at low gas pressures, which is attributed to an increase in the total product gas flow rate coming from the extra CO₂ desorption at low gas pressures. This directly increases the energy requirement for the vacuum pump according to Eq. 13. Moreover, it can also be seen that E_{pumps} increases when the gas pressure is higher than 0.7 atm. This is due to a significant increase in the membrane area from 2.99 to 4.11 m² (37.42% increase) required for CH₄ recovery at the high gas pressure range of 0.7 to 0.9 atm. Contrastingly, the membrane area is only required to increase from 2.4 to 2.99 m² (24.5% increase) at the gas pressure range of 0.5 to 0.7 atm. At the high gas pressure range, there is a large segment in the middle of the membrane module where the CH₄ desorption rate is low given that the CH₄ concentration in the gas phase is close to its equilibrium concentration in the liquid phase. As a result, a larger mass transfer area is required

to provide the same removal efficiency and, thus increasing the E_{LP} at a rate that is higher than that in the low gas pressure range.

Furthermore, the operating gas pressure to achieve the maximum $Net E$ is shifted from 0.65 to 0.8 atm and the maximum $Net E$ is higher when the sweep gas flow rate increases from 0.07 to 0.10 L/min (Fig. 5). As previously explained, E_{LP} is increased significantly when the system almost reaches the equilibrium at an operating pressure close to 1 atm. The increase in the strip gas flow rate directly helps to increase the mass transfer driving force and reduce the length of membrane module where the driving force is small. Thus, this results in a significant decrease in the E_{LP} as well as an increase in the $Net E$.

Fig. 7. Effects of gas pressure on $Net E$ and CH_4 mole fraction in the gas outlet. (Liquid flow rate: 2 m³/day, liquid pressure: 1 atm, gas and liquid temperatures: 298.15 K, number of fibers: 650, and packing density: 0.14)

Fig. 8. Comparison of the energy requirement for vacuum ($E_{VP,tot}$) and liquid pumps (ELP) between operating gas pressure of 0.6 and 0.7 atm at different strip gas flow rate.

3.1.2 Effect of strip gas flow rates

Fig. 7 shows the effect of strip gas flow rate on the *Net E* and CH_4 mole fraction in the gas outlet. Two operating gas pressures at 0.6 and 0.7 atm are used for comparison. The selected gas/liquid flow rate ratio is between 0.043 to 0.011. This ratio needs to be higher than 0.034, a minimum theoretical value reported by Cookney et al. [7], for the removal of 98% dissolved CH_4 . Gz numbers in the liquid phase and Re_G are varied from 1.83 to 2.61, and 0.22 to 0.45, respectively. An optimum point for the *Net E* can also be obtained when varying the strip gas flow rate. Although the increase in strip gas flow rate has little effect on the CH_4 mass transfer coefficient given that the mass transfer is controlled by the resistance in the liquid phase [7, 11], it can directly improve the mass transfer driving force, which significantly shortens the length of the membrane module and reduces the energy requirement for the liquid pump. However, using too high a strip gas flow rate would not only dilute the CH_4 concentration in the gas outlet but also

enhance the energy requirement for the vacuum pump. The comparison of the effect of strip gas flow rates at different operating gas pressures on the *Net E* reveals an interesting result. The drop in *Net E* by decreasing the gas flow rate before the optimum point is steeper when using 0.7 atm gas pressure while the drop in *Net E* with an increase in the gas flow rate after the optimal point when using 0.6 atm gas pressure is larger than that of 0.7 atm. This could be explained by the comparison between the energy requirement for the vacuum ($E_{VP,tot}$) and liquid pumps (E_{LP}) under these two operating gas pressures (Fig. 8). The difference in the E_{LP} between 0.6 and 0.7 atm operating gas pressures is higher at the low strip gas flow rates while for $E_{VP,tot}$, it is higher at the high strip gas flow rates. The additional of gas flow rate from the CO₂ desorption when using a lower operating gas pressure is the main reason for this phenomenon. It decreases the CH₄ gas phase partial pressure which increases the mass transfer driving force significantly, when using low strip gas flow rates leading to a further shortening of the length of the membrane module and E_{LP} . Nevertheless, the additional gas at the lower gas pressure also leads to a higher $E_{VP,tot}$, causing a more profound impact on the total energy requirement and *Net E*. On this account, the $E_{VP,tot}$ increases higher with a strip gas flow rate at the lower operating pressure range and this results in a larger reduction in the *Net E* as can be seen from Fig.7 of that using the operating pressure at 0.6 atm.

Fig. 9. Optimization plot for *Net E*. (Number of fibers: 650, effluent flow rate: 1.39 L/min, effluent is saturated by biogas contained 60% vol CH₄ in balance of CO₂)

Fig. 10. CH₄ mole fraction of the product gases from the membrane contactor at different strip gas flow rates and gas pressures. (Number of fibers: 650, effluent flow rate: 1.39 L/min, effluent is saturated by biogas contained 60% vol CH₄ in balance of CO₂)

3.1.3 Optimal operating gas pressure and strip gas flow rate

The optimization plot for *Net E* using a membrane contactor to recover dissolved CH₄ from the anaerobic effluent is illustrated in Fig. 9. It can be seen that a maximum *Net E* is achieved when using high strip gas flow rates together with an operating gas pressure that is close to 1 atm. At the strip gas flow rate and operating gas pressure of 0.11 L/min and 0.9 atm, respectively, the *Net E* obtained is 0.178 MJ/m³. This is 85.37 % of the total energy that could be potentially recovered from the dissolved CH₄ (E_{MT}). The pressure drop in the lumen side of hollow fiber is 0.097 atm in this condition. It should be noted that the membrane contactor in this work is operated in the non-wetted mode thus the membrane mass transfer resistance itself is insignificant to the overall process [11]. Therefore, even though an in-house made membrane is used in the present study, the reported data stay largely relevant and could represent potential performance of membrane contactors for the recovery of dissolve CH₄ application as long as the pre wetting is effectively prevented. This was accomplished by a membrane contact angle and liquid entry pressure of 122.6°, and 36.2 psi, respectively, of our selected membrane which demonstrates at least 300 hours of stability toward membrane wetting when tested with tap water [9]. The number of Net E reported is also based on the assumption that there is no membrane fouling occurred in the system, therefore, the membrane characteristics are constant throughout the operation and energy penalties associated with the membrane cleaning processes are not included into the consideration of Net E.

Although our above result suggests to use a high strip gas flow rate to increase the *Net E*, the reduction in the CH₄ mole fraction of the outlet gas is of concern (Fig. 10). Therefore, it is important to know the required y_{CH_4} in the product gas so that the combined gas (biogas and product gas from the membrane contactor) has sufficient CH₄ content to be used in a micro-

turbine for electricity generation (35%vol). Table 4 shows the CH_4 content of the combined gas as a function of COD in the wastewater. The calculation was based on the assumption that 1 g of COD generated 0.35 L of CH_4 gas [20]. Parameters such as 60%vol CH_4 biogas, 1 atm and 298.15 K system pressure and temperature, respectively, 90% recovery efficiency of the dissolved CH_4 , a 22.22 g CH_4/m^3 CH_4 solubility and a 0.656 g/L CH_4 density are fixed and utilized. It also indicates the percentage amount of dissolved CH_4 in the effluent as compared to the total CH_4 production. As the concentration of dissolved CH_4 does not depend on the total amount of biogas produced but rather depends only on the biogas composition as well as the gas and liquid temperatures and pressures, the significance of the dissolved CH_4 will be lowered as the inlet COD of wastewater increases. It should be noted that the membrane contactor in this work is used to treat a MBR effluent. Therefore, no super saturation of dissolved CH_4 can occur. According to Crone et al. [3], super saturation of dissolved CH_4 occurs mainly in effluents of anaerobic UASB processes and this often results in a higher significance of dissolved CH_4 to the total CH_4 production of the anaerobic processes. As shown in Fig. 5, even after optimization, a y_{CH_4} of only 0.11 is achieved at the gas outlet. Hence, the product gas ought to combine with the biogas generated from the wastewater with COD higher than 700 mg/L to have a final gas that meets the requirement of the micro-turbine. It should be emphasized that using only the strip gas mode for CH_4 recovery is not recommended for the treatment of low COD wastewater where a high y_{CH_4} in the combined gas is needed. Since using low operating gas pressures with a low strip gas flow rate could obtain higher *Net E* as can be seen from Fig. 9.

Table 4. Composition of the combined gas between generated biogas as the function with COD and product gas from the membrane contactor with different CH_4 mole fraction.

COD of wastewater (mg/L)	CH ₄ produced (L _{CH4} /m ³)	Biogas produce (y _{CH4} =0.6, L _{biogas} /m ³)	% CH ₄ loss in effluent	y _{CH4} of combined gas (produced biogas + product gas from MC)		
				0.1	0.14	0.17
				y _{CH4} from MC	y _{CH4} from MC	y _{CH4} from MC
350	110.25	76.37	30.73	0.25	0.31	0.35
500	157.5	123.62	21.51	0.30	0.36	0.40
650	204.75	170.87	16.55	0.34	0.40	0.43
750	236.25	202.37	14.34	0.36	0.42	0.45

T= 298.15, P = 1 atm, CH₄ density = 0.656 g/L, CH₄ solubility = 22.22 g CH₄/m³

3.2 Effect of relevant operating parameters on the optimization

3.2.1 Effect of number of fibers

In this section, the number of fibers are varied and the packing density is kept constant at 0.14 so that the diameters of the membrane module are adjusted accordingly to obtain the same gas velocity in all conditions. The strip gas flow rate and operating gas pressure are 0.1 L/min and 0.8 atm, respectively. The liquid flow rate is fixed at 1.39 L/min (2 m³/day) and is saturated with 60 vol% CH₄ biogas. The gas and liquid temperature are maintained at 298.15 K and the operating liquid pressure is at 1 atm. Fig. 11 shows that the *Net E* increases with the number of fibers. The change in the number of fibers directly affects the length of membrane module required to remove the dissolved CH₄. The lower the numbers of fibers, the longer the module height and thus the higher the energy consumption for the liquid pump. Moreover, a decrease in number of fibers increases the liquid velocity inside the membrane tubes leading to a higher pressure drop which rises the energy consumption for the liquid pump according to Eq. 19 and 18, respectively. It can also be seen that the membrane area required to recover 90% dissolved CH₄ decreases with the number of fibers. This is because the increase in the desorption rate at the liquid entry stage of the membrane contactor as a result of an increase in the contact surface area

with the number of fibers shorten the later stage of membrane module where the mass transfer driving force is lower. Even though the result suggests the increase in the *Net E* with the number of fibers, unreasonably high numbers of fibers may lead to a non-uniform liquid flow contribution in the lumen side which lowers the performance of the membrane contactors and increases the membrane module costs.

Fig. 11. Effect of number of fibers on *Net E* and membrane area required to remove 90% dissolved CH₄. (Strip gas flow rate 0.1 L/min and operating gas pressure 0.8 atm)

Fig. 12. Optimization plot for *Net E* of using numbers of fibers at (a) 350 and (b) 1000. (effluent flow rate: 1.39 L/min and saturated by biogas contained 60% vol CH₄ in balance of CO₂)

The optimization plots for *Net E* of using numbers of fibers at 350 and 1000 are displayed in Fig. 12. For 350 fibers, the maximum *Net E* is not at high operating gas pressures as in Fig. 9 but rather is at an operating gas pressure of 0.6 atm using a strip gas flow rate of 0.1 L/min. This is as demonstrated in Fig. 11 that the membrane area has to be increased when using the low number of fibers due to the smaller driving force along the length of the membrane module. This increases the significant of E_{LP} to the total energy requirement (E_{pumps}) thus a lower operating gas pressure to increase the driving force and lower the E_{LP} can significantly improve *Net E* at the low operating pressure range of using the number of fiber at 350.

When using a higher number of fibers, the mass transfer is more efficient and the energy requirement for the liquid pump is heavily reduced but the energy consumption for the vacuum pump becomes more significant. Therefore, using a higher strip gas flow rate as a means to replace the vacuum condition to reduce $E_{VP,tot}$ can increase the *Net E* and shift the maximum *Net E* toward using a higher strip gas flow rate and gas pressure as illustrated in Fig. 12b.

Table 5. Comparison of *Net E* at different liquid flow rates.

Parameters	Liquid flow rate (m ³ /day)		
	2		1.3
N_f	650	1000	650
V_L (m/s)	0.1	0.066	0.066
y_{CH_4}	0.15	0.15	0.11
Total gas flow rate (L/min)	0.21	0.21	0.19

$Net\ E$ (MJ/m ³)	0.172	0.191	0.192
Membrane area to remove 90% CH ₄ (m ²)	3.46	3.44	1.77

3.2.2 Effect of liquid flow rates

The effects of liquid flow rates on $Net\ E$ and operating conditions of the membrane contactor are shown in Table 5. The strip gas flow rate and gas pressure were 0.07 L/min and 0.8 atm, respectively. The CH₄ recovery efficiency is again fixed at 90%. Gz numbers in the liquid phase and Re_G of using 1000 hollow fibers are 1.74, and 0.18, respectively. It is obvious that the $Net\ E$ increases with a reduction in the liquid flow rate at a fixed number of fibers. This is due to the lower membrane area required to remove a smaller amount of dissolved CH₄ entering the system. In addition, the pressure drop along the length of membrane module which directly enhances the W_{LP} according to Eq. 18 is also decreased. As previously discussed, the $Net\ E$ could be improved with a greater number of fibers. Table 5 reveals that the $Net\ E$ at the same liquid velocity (m/s) but different liquid flow rates are very closed. This also shows that the y_{CH_4} in the product gas from the membrane contactor can be increased with an increase in the liquid flow rate given a reduced gas/liquid flow rate ratio of the system.

Fig. 13. Optimization plot for *Net E* of the solution saturated with biogas contained 50% vol CH₄ in balance with CO₂. (Number of fibers: 650, effluent flow rate: 1.39 L/min)

Fig. 14. Effects of dissolved biogas composition and gas pressure on the energy consumption of vacuum and liquid pumps. (Number of fibers 650, effluent flow rate 1.39 L/min, strip gas flow rate 0.1 L/min)

3.2.3 Effect of dissolved biogas compositions

The optimization plot for *Net E* using a solution saturated with a biogas containing 50% vol CH₄ in balance with CO₂ is shown in Fig. 13. Compared with Fig. 9, the *Net E* from the dissolved biogas containing only 50% vol CH₄ are relatively lesser given the lower energy content in the biogas. It is also noticed that a decrease in gas pressure has a greater impact in reducing the *Net E* as compared to the solution that is saturated with biogas containing 60% vol CH₄. Since CO₂ is more soluble in water than CH₄, the higher CO₂ composition in the biogas with 50% vol CH₄ implies a higher total concentration of dissolved gases in the effluent. Thus, more gases can be desorbed at a lower gas pressure and this increases the energy consumption for the vacuum

pump. The energy consumption for liquid and vacuum pumps for the recovery of dissolved CH_4 from different dissolved biogas compositions is displayed in Fig 14. Interestingly, the total energy consumptions to recover CH_4 from effluent of different dissolved biogas compositions are very closed. Although the membrane length has to be lengthened for the higher dissolved CH_4 concentration, which causes an increase in the energy consumed for the liquid pump, the energy consumption for the vacuum pump is, however, reduced because of the reduction in total desorbed gas flow rate as mentioned earlier.

Fig. 15. Effect of percentage CH_4 recovery efficiency of *Net E*. (Number of fibers: 650, effluent flow rate: 1.39 L/min, effluent is saturated by biogas contained 60%vol CH_4 in balance of CO_2)

3.2.4 Effect of dissolved CH_4 recovery efficiencies

Fig. 15 shows the effect of the percentage (%) CH_4 recovery efficiency on the *Net E*. The liquid flow rate is kept at $2 \text{ m}^3/\text{day}$, the effluent is saturated with biogas containing 60%vol CH_4 and the strip gas flow rate is at 0.1 L/min. The number of fibers is 650, liquid pressure and system temperature are 1 atm and 298.15 K, respectively. Gz numbers in the liquid phase and Re_G are

varied from 1.03 to 3.87, and 0.32 to 0.36, respectively. The *Net E* increases with % CH₄ recovery efficiency until an optimal point before it gradually decreases (Fig. 15). This is because the membrane length required to further increase the recovery efficiency beyond the optimal point is very large due to the very small mass transfer driving force arising from the tiny amount of CH₄ that remains in the liquid phase. In this case, the increase in energy requirement for the liquid pump is higher than the energy improved for the recovered CH₄. Thus, *Net E* is reduced.

Table 6 Percentage change of Net E when the input parameters are increased and decreased by 20%.

Input parameters	% change of Net E	
	Value +20%	Value -20%
Gas pressure: 0.8 atm	- 0.77	- 1.14
Strip gas flow rate: 0.1 L/min	+ 0.76	- 1.51
number of fibers: 650	+ 4.75	- 8.75
liquid flow rates: 1.39 L/min	- 8.65	+ 6.00
Inlet CH ₄ concentration: 8.3×10^{-4} mol/L	+ 23.64	- 23.59

* Dissolved CH₄ recovery efficiency is fixed at 90%. It is assumed that dissolved biogas is composed of 60 vol% CH₄ and 40 vol% CO₂.

3.2.5 Sensitivity analysis

Results from the previous sections reveal that several operating parameters can affect the Net E. Therefore, a sensitivity analysis was carried out to obtain and compare the significance of each parameter to the energy analysis. Table 6 displays the percentage change of the Net E when the value of each input parameter was changed by $\pm 20\%$ while keeping other parameters constant. The percentage of dissolved CH₄ recovery efficiency was fixed at 90% in all conditions. The

operating gas pressure and strip gas flow rate are selected as 0.8 atm and 0.1 L/min, respectively, as these operating conditions give a high Net E as shown in Fig 9.

It can be seen that the Net E is more affected by varying the operating parameters of the liquid phase than those of the gas phase. This is in agreement to the energy analysis in Fig. 6 which highlights that the energy consumption for the liquid pump (ELP) is the dominating factor when the operating gas pressure employed is close to 1 atm. The change of inlet dissolved CH₄ concentration is the most sensitive parameter to the Net E. The variations are higher than 20% due to an increment/decrement of mass transfer driving force results in a substantial change of the membrane area required to recover the dissolved CH₄.

4. Conclusion

The energy analysis of the membrane contactor for CH₄ recovery from an effluent of **anaerobic membrane bioreactor** wastewater treatment process revealed that the *Net E* or the electrical energy recovered from the dissolved CH₄ in 1 m³ effluent can be optimized by varying the strip gas flow rate and operating gas pressures. The maximum *Net E* is obtained when using a high strip gas flow rate and slightly low vacuum condition that is close to 1 atm. At the strip gas flow rate and operating gas pressure of 0.11 L/min and 0.9 atm, respectively, the *Net E* obtained is 0.178 MJ/m³ which is 85.37 % of the total energy recovered from the dissolved CH₄. However, the CH₄ mole fractions in the outlet gas are low at these conditions. Hence, it is recommended to combine the product gas from the membrane contactor with the headspace biogas to afford a final gas with sufficient CH₄ content to be used in a micro-turbine. The *Net E* can be further improved by using a high number of fibers or low liquid flow rate to reduce the liquid pressure

drop inside the membrane fibers, which affects the energy consumption for the liquid pump as well. In addition, the total energy consumption for the membrane contactor changes slightly with the dissolved biogas composition due to the energy consumption for the vacuum pump is found to decrease but that of the liquid pump is increased with an increase CH_4 composition in biogas. Although an improvement in the CH_4 recovery efficiency increases the amount of desorbed CH_4 , the *Net E* could be reduced at high recovery efficiencies given the need to increase the membrane module length. It should be noted again that the possible drop decline due to the membrane fouling and possible energy penalties from membrane cleaning and relevant processes were not considered in our work. Also, despite no bio-fouling observed with the use of a relatively cleaner anaerobic membrane bioreactor effluent in this work, there remains a possibility for bio-fouling in long-term operations induced by residual bacteria and organic matters in the effluent. In such an event, it is noteworthy to mention that the Net E will decrease, given the energy required for membrane cleaning activities.

Nomenclature

A_m	Mass transfer area (m^2)
C_i	Concentration of gas i dissolved in liquid phase (kmol/m^3)
$D_{G,eff}$	Effective gas diffusion coefficient in membrane pores (m^2/s)
D_i	Diffusion coefficient of gas i (m^2/s)
D_{Kn}	Knudsen diffusion coefficient (m^2/s)
D_M	Molecular diffusion coefficient (m^2/s)
F_i	Molar gas flow rate of gas i (kmol/s)
$F_{tot}, F_{tot,day}$	Total molar gas flow rate (kmol/s , and m^3/day , respectively)

d_e	Hydraulic diameter (m)
d_i	Inside membrane diameter (m)
d_{ln}	Log mean membrane diameter (m)
d_m	Membrane module diameter (m)
d_o	Outside membrane diameter (m)
E_{LP}	Electric energy consumed by liquid pump per 1 m ³ of effluent (MJ/m ³)
E_{MT}	Electric energy obtained for recovered CH ₄ per 1 m ³ of effluent (MJ/m ³)
E_{pumps}	Electric energy required for treatment of 1 m ³ of effluent (MJ/m ³)
$E_{VP,tot}$	Electric energy consumed by vacuum pump system per 1 m ³ of effluent (MJ/m ³)
Gz	Graetz number (dimensionless, $d_i^2 V_L / D_{i,L} Z$)
H	Henry's constant (kmol/m ³ .atm)
k_G	Gas phase mass transfer coefficient (m/s)
k_L	Liquid phase mass transfer coefficient (m/s)
k_M	Membrane phase mass transfer coefficient (m/s)
K_{OV}	Overall mass transfer coefficient base on liquid phase (m/s)
$L_m, L_{m,day}$	Liquid flow rate (m ³ /s, and m ³ /day, respectively),
N_i	Desorption flux (kmol/m ² .s)
$Net E$	<i>Net</i> Electric energy obtained for treatment of 1 m ³ of effluent (MJ/m ³)
n_f	Number of fiber (dimensionless)
P	Pressure (atm, Pa for Eq. 21)
p_i	Partial pressure of gas i (atm)
Q	Liquid pump capacity (m ³ /hr)

R	Gas constant (kmol.K/m ³ .atm)
Re	Reynold number (dimensionless)
Sc	Schmidt number (dimensionless)
Sh	Sherwood number (dimensionless)
T	Temperature (K)
V_G and V_L	Gas and liquid velocities, respectively (m/s)
W_{LP}	Energy consumption for liquid pump (kW)
W_{VP}	Energy consumption for vacuum pump (kW)
$W_{VP,cooling}$	Energy consumption for cooling system of vacuum pump (kW)
$W_{VP,tot}$	Energy consumption for vacuum pump system (kW)
Z	Membrane length (Z)

Greek letters

ε_M	Membrane porosity (dimensionless)
τ_M	Tortuosity (dimensionless)
δ_M	Thickness of membrane (m)
κ	Adiabatic constant (dimensionless)
η	Pump efficiency (dimensionless)
μ	Viscosity of water (kg.m/s)
ω	Energy content (MJ/m ³)

Subscript

G	Gas phase
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<i>L</i>	Liquid phase
LP	Liquid pump
<i>M</i>	Membrane phase
<i>i</i>	CH ₄ or CO ₂
<i>tot</i>	total
VP	Vacuum pump

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