Factors contributing to flux improvement in vacuum-enhanced direct contact membrane distillation

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Abstract

Low water flux in membrane distillation (MD) is a concern for full-scale application. In the past decades, attempts have been made to improve water flux in MD and vacuum-enhanced direct-contact MD (VEDCMD) has proven to be an effective configuration to achieve this. However, only qualitative assessments of the factors that might improve water flux have been reported in the literature. In this study, a mechanistic investigation of the factors contributing to higher water flux in VEDCMD was performed. Direct-contact MD (DCMD) and pressure-enhanced DCMD (PEDCMD) configurations were also investigated for comparison. Less membrane compaction was identified as one dominant factor contributing to improved water flux in VEDCMD as very little compaction occurred in VEDCMD compared to that which occurred in DCMD and PEDCMD. Lower air pressure inside the membrane pores was found to be the other dominant factor contributing to improved water flux in VEDCMD; the magnitude of the air pressure was estimated to be approximately the average of the feed and distillate pressures in VEDCMD and approximately the distillate-side pressure in DCMD and PEDCMD. Pressure difference, as is present in both PEDCMD and VEDCMD, was found to have a minimal effect on water flux.

Keywords

Direct-contact membrane distillation, Flux improvement, Membrane compaction, Membrane pore pressure

1. Introduction

Membrane distillation (MD) is a thermally-driven membrane process in which separation occurs through a phase change. The driving force in MD is the vapor pressure difference resulting from the temperature difference across the membrane. Because of the high latent heat of evaporation of water, MD is an energy-intensive process [1]. However, it can be combined with low-grade (“waste”) heat to reduce energy costs [2-8].

MD has advantages over conventional membrane processes (e.g., reverse osmosis) because the driving force of MD does not decrease significantly with increasing feed-water salinity [9]. This has inspired interest in MD for treatment of high salinity brines [10-12] and complex feed waters such as produced waters in the oil and gas industry. More recent interest in MD results from the higher rejection rates of MD membranes over reverse osmosis membranes; theoretically, MD can achieve 100% rejection of salts and non-volatile organics. For this reason, MD is well suited to remove salts and low-molecular-weight contaminants (e.g., boron, trace organic compounds, and urea) that may pass through other treatment methods such as reverse osmosis [13-16]. MD could be considered as a replacement treatment option or as a side-stream treatment whereby a side stream of reverse osmosis permeate could be further polished for these contaminants and then blended with the bulk permeate to achieve treatment objectives. In all cases, maintaining membrane hydrophobicity is key for achieving high rejection of the membrane [17].

When treating more complex feedwaters, it has been found that high water fluxes can cause faster transport of fouling and scaling material to the membrane surface and thus, high water fluxes are detrimental to membrane flux over time [10]. On the other hand, when treating feed waters with relatively low fouling and scaling potentials, obtaining high water flux is desirable [9]. Approaches to improve water flux have included developing novel membranes [18], deaerating feed waters [2, 19], and using vacuum in the membrane module [20, 21]. Schofield et al. [19] and Cath et al. [20] employed equal vacuum on both the feed and distillate sides of the membrane and observed higher fluxes compared to the traditional DCMD configuration (without hydraulic pressure or vacuum employed). Cath et al. [20] also used vacuum only on the distillate side while keeping the feed side at atmospheric pressure and this led to even greater flux improvement. This process, called vacuum-enhanced DCMD (VEDCMD), was patented in 2010 [22].

Increased flux with the application of vacuum has been partially or fully attributed to decreased air pressure inside the membrane pores, decreased temperature polarization, decreased membrane conductive heat loss, and increased pressure difference across the membrane [19-21]. However, the mechanisms for improved flux have not been systematically evaluated and the dominant factors affecting flux enhancement have not been clearly identified. It is possible that only one or two factors may explain flux improvement. In addition, of these factors, mechanistic evaluations of air pressure inside the membrane pores have been inconclusive. In two different studies, one where vacuum was employed on the distillate side of the DCMD membrane [21] and the other where vacuum was employed on both the feed and distillate sides [19], the pore pressure (sum of air pressure and water vapor pressure inside the pores [23, 24]) was assumed to be equal to the vacuum pressure of the distillate stream. It is unlikely that the same assumption is valid for both scenarios. Therefore, in determining the dominant factors affecting water flux in VEDCMD, analyses must go beyond what has been done previously in the literature.

Furthermore, there is not consensus in the literature on the role of membrane compaction on water flux. One study from the literature that employed hydraulic pressures of equal magnitude on the feed and distillate sides of the membrane observed flux reduction compared to traditional DCMD systems; membrane compaction was suggested to be the reason [25], especially for membranes with low porosities [26]. Another study that employed hydraulic pressure only on the feed side observed no flux change [20], and it was postulated that membrane compaction was one of the factors that increased water flux (opposite to [25]), while other factors reduced water flux, leading to the near constant overall water flux.

The overall objective of this study is to evaluate the dominant factors responsible for improved water flux when vacuum is employed in DCMD. Because employing vacuum only on the distillate side of the membrane (as in VEDCMD) leads to greater flux improvement than any other configuration, VEDCMD is the main focus of this study. To achieve the overall objective, first, the dominant factors contributing to higher water flux in VEDCMD are evaluated theoretically. Second, the magnitude of air pressure inside the membrane pores in VEDCMD is quantified and its effect on water flux is evaluated. Third, membrane compaction is measured and the effect of membrane compaction on water flux is evaluated. Results from this study will aid in better understanding of the factors leading to enhanced water flux in VEDCMD and in turn, facilitate application of VEDCMD especially in treatment of water with low fouling and scaling potential where high water flux is desirable.

2. Materials and methods

2.1. Experimental setup

Three DCMD configurations were evaluated: traditional DCMD without hydraulic pressure or vacuum applied; VEDCMD with vacuum applied on the distillate side; and pressure-enhanced DCMD (PEDCMD) with hydraulic pressure applied on the feed side. The PEDCMD configuration was used for comparative purposes by creating the same hydraulic pressure difference as used in VEDCMD. A bench-scale system with a modified acrylic membrane cell was used to study these configurations. The membrane cell utilized a flat-sheet membrane with 118 cm² (13.5 cm by 8.7 cm) of effective membrane surface area. DCMD and PEDCMD used the same setup (Fig. 1a) with a needle valve at the outlets of both the feed and distillate sides to adjust the pressure inside the membrane cell. In VEDCMD (Fig. 1b), the distillate-side pump was on the outlet of the membrane cell to create a vacuum on the distillate side by drawing water from the cell. The pressures on the feed and distillate sides were maintained at 20/20 (feed/distillate) kPa in DCMD; 60/20, 80/20, and 100/20 kPa in PEDCMD; and 20/-20, 20/-40, and 20/-60 kPa in VEDCMD. All pressures in the current study were reported as gauge pressures. All tests were performed using the same feed temperature (40 °C), distillate temperature (20 °C), and fluid flow rates (1.0 L/min) on both the feed and distillate sides. A 35 g/L NaCl solution was used as the feed solution and deionized (DI) water was used on the distillate side. The feed solution and the DI water were re-circulated counter-currently on their respective sides of the membrane. As water evaporated through the membrane, excess water from the distillate reservoir overflowed into a beaker on an analytical balance. The overflow rate was used to calculate the water flux. The test was stopped when the flux was stable for 30 min. To create turbulent flow and reduce temperature polarization, spacers were placed in the flow channels on both the feed and distillate sides. A bench conductivity meter (Traceable™, VWR International, USA) was used to measure the feed and distillate conductivities. Temperatures, pressures, and dissolved

oxygen (DO) concentrations were monitored using dual-channel digital thermometers, pressure gauges, and DO probes at the inlet and outlet of the membrane cell. Average DO values were calculated and used to investigate oxygen (i.e., air) transport across the membrane. The DO probes (YSI 556 MPS, YSI Environmental, Yellow Springs, OH) had a response time of 2-8 seconds according to the manufacturer.

2.2. Membrane characterization

A single-layer flat-sheet polytetrafluoroethylene membrane (GE Water & Process Technologies, Minnetonka, MN) was used in this investigation. A new membrane coupon was used for each test. Membrane properties (thickness, porosity, tortuosity, and average pore size) were determined for the unused membrane and the membranes after flux testing.

2.2.1. Membrane thickness and compaction measurements

Membrane thickness (δ) was measured from scanning electron micrographs of membrane cross sections. Three membrane coupons were taken from the same roll of membrane. For each coupon, membrane thickness was measured at three different locations to calculate the average membrane thickness and standard deviation. Each membrane coupon was frozen in liquid nitrogen and cut with a blade.

Multiple regression analysis was performed using Minitab® 17.1.0 to statistically evaluate membrane compaction. The new (unused) membrane was set as the control and the various pressure scenarios (e.g., 60/20, 20/-20) were set as the variables. A p value less than 0.05 was considered significant [27].

2.2.2. Membrane porosity and tortuosity measurements

Membrane porosity (ε) was determined by [28]:

$$\varepsilon = 1 - \frac{m}{\rho_p \times A \times \delta} \quad (1)$$

where m and A are the mass and surface area of the membrane coupon (assumed constant before and after flux testing); ρ_p is the reported density of the polymer material (2.2 g/cm³ [29, 30]). Although additives may be included during the membrane manufacturing process, and could cause ρ_p to be different from 2.2 g/cm³, the effect of additives on ρ_p is assumed negligible here due to the much less mass added than the mass of the polymer material [31]. Membrane tortuosity (τ) was calculated as [32, 33]:

$$\tau = \frac{(2 - \varepsilon)^2}{\varepsilon} \quad (2)$$

2.2.3. Average membrane pore size determination

Average membrane pore diameter (d) was determined by the gas permeation test with compressed air [34, 35]. The experimental apparatus used for this measurement was reported in [35]. Specifically, a membrane disk with an effective membrane area of 8.5 cm² was placed at the bottom of a stainless steel cylinder. The permeation flux of air through the dried membrane was measured at various transmembrane pressures in the range of 10-100 kPa at room temperature using an air flow meter and two digital monometers (one before the membrane and one after the membrane). By plotting a linear dependence between the air permeance (y axis) and

the mean transmembrane pressure (x axis), the intercept (B_o) and the slope (K) were determined, and consequently, the pore diameter was calculated using the following equation:

$$d = \frac{16}{3} \left(\frac{B_o}{K} \right) \left(\frac{2RT}{\pi M_j} \right)^{0.5} \mu_g \quad (3)$$

where R is the universal gas constant; T is the membrane temperature during the measurement (assumed the same as the ambient temperature [35]); M is the molecular weight; and μ_g is the gas viscosity [35]. Here, subscript j is used to represent air (subscript i will later be used to represent water vapor).

2.3. Flux prediction using mass transfer models

In DCMD, viscous flow, where the air and water vapor mixture behaves as a continuous fluid [36] can be neglected because no pressure difference exists [37]. Because of this, both the Schofield model (Fig. 2a) and the simplified dusty-gas model (surface diffusion neglected; Fig. 2b) can be represented with the same thermal resistance circuit. In PEDCMD and VEDCMD, a pressure difference exists; therefore, the simplified dusty-gas model and the Schofield model include viscous flow and have unique thermal resistance circuits.

The flux attributed to Knudsen diffusion (N_k) was determined by [3]:

$$N_k = \frac{2}{3} \frac{r\varepsilon}{\tau\delta} \left(\frac{8M_i}{\pi RT_m} \right)^{0.5} \Delta P_i \quad (4)$$

where r is membrane pore radius; T_m is the average temperature of the membrane surface on the feed side (T_{mf}) and distillate side (T_{md}); and ΔP_i is the vapor pressure difference across the membrane, which equals the feed-side vapor pressure (P_{if}) minus the distillate-side vapor pressure (P_{id}). By neglecting the salt concentration in the distillate side due to the very low distillate conductivity ($< 10 \mu\text{S/cm}$ at the end of the test), the vapor pressure of the distillate side (P_{id}) was calculated by the Antoine equation [38]:

$$P_{id} = \exp\left(23.328 - \frac{3841}{T_{md} - 45}\right) \quad (5)$$

Considering the effect of salts in the feed solution on the vapor pressure of the feed side (P_{if}), P_{if} was determined by [39, 40]:

$$P_{if} = \exp\left(23.328 - \frac{3841}{T_{mf} - 45}\right) (1 - x_{NaCl}) (1 - 0.5x_{NaCl} - 10x_{NaCl}^2) \quad (6)$$

where x_{NaCl} is the molar fraction of NaCl in the bulk feed solution.

T_{mf} was determined by [24, 41, 42]:

$$T_{mf} = T_f - \frac{Q}{h_f} \quad (7)$$

where T_f is the bulk temperature of the feed solution and h_f is the heat transfer coefficient in the boundary layer on the feed side. Q is the total heat input in the system, which was determined by:

$$Q = C_p \dot{m} (T_{dout} - T_{din}) \quad (8)$$

where C_p and $\dot{m}$ are the heat capacity and fluid flow rate of the distillate, respectively; and T_{dout} and T_{din} are the distillate-side temperatures at the outlet and inlet of the membrane cell, respectively.

T_{md} was determined by [24, 41, 42]:

$$T_{md} = T_d + \frac{Q}{h_d} \quad (9)$$

where T_d is the bulk temperature of the distillate and h_d is the heat transfer coefficient in the boundary layer on the distillate side.

The heat transfer coefficient (h ; h_f in Eq. 7 and h_d in Eq. 9) was determined by:

$$h = \frac{Nu k_f}{d_h} \quad (10)$$

where k_f is the thermal conductivity of the fluid; d_h is the hydraulic diameter of the flow channel on the feed or distillate side; and Nu is the Nusselt number. For a spacer-filled channel, d_h was determined by [39, 43]:

$$d_h = \frac{4\varepsilon_s}{2/H + 4(1 - \varepsilon_s)d_f} \quad (11)$$

where H is the channel height; d_f is the spacer filament size, which was measured by a digital caliper; and ε_s is the spacer porosity, which was determined by [39, 43]:

$$\varepsilon_s = 1 - \frac{\pi d_f^2}{2l_m H \sin \theta_s} \quad (12)$$

where l_m is the mesh size (grid length) of the spacer; and θ_s is the hydrodynamic angle of the spacer, which is the angle formed by the grids of the spacer. Nu in Eq. 10 was determined by [44]:

$$Nu = 0.664 k_{dc} Re^{0.5} Pr^{0.33} \left(\frac{2d_h}{l_m} \right)^{0.5} \quad (13)$$

where Re is the Reynolds number and Pr is the Prandtl number. For a spacer-filled channel, Re was calculated by [43]:

$$Re = \frac{\dot{m} d_h \rho_f}{A_c \mu_f} \quad (14)$$

where ρ_f is the fluid density; A_c is the cross-section area of the module channel; and μ_f is the fluid viscosity. k_{dc} in Eq. 13 was determined by [43, 44]:

$$k_{dc} = 1.654 \left(\frac{d_f}{H} \right)^{-0.039} \varepsilon_s^{0.75} \left(\sin \frac{\theta_s}{2} \right)^{0.086} \quad (15)$$

The flux attributed to molecular diffusion (N_m) was determined by [3]:

$$N_m = \frac{\varepsilon}{\tau \delta} \frac{P_{pore} D}{RT_m} \frac{1}{|P_{air}|_{ln}} \Delta P_i \quad (16)$$

where P_{air} is the air pressure inside the membrane pores; P_{pore} is the membrane pore pressure, which equals the sum of P_{air} and P_i ; and D is the diffusion coefficient. $P_{pore} D$ was calculated as [24, 45]:

$$P_{pore} D \text{ (kPa m}^2\text{/s)} = 1.895 \times 10^{-8} \times T_m^{2.072} \quad (17)$$

The flux attributed to viscous flow (N_v) was determined by [3]:

$$N_v = \frac{P_i}{8RT_m \mu} \frac{r^2 \varepsilon}{\tau \delta} \Delta P \quad (18)$$

where μ is the vapor dynamic viscosity and ΔP is the transmembrane pressure difference.

According to the structure of the thermal resistance circuits (Fig. 2), the overall inverse flux ($1/N$) for the Schofield model is [36]:

$$\frac{1}{N} = \frac{1}{N_k + N_v} + \frac{1}{N_m} \quad (19)$$

and the overall flux (N) for the dusty-gas model (without surface diffusion) is [36]:

$$N = N_v + \frac{N_k N_m}{N_k + N_m} \quad (20)$$

In the discussion below, membrane thickness, δ , was chosen as a representative membrane property because the other three membrane properties (r , ε , and τ) depend on δ . This is easily seen in Eq. 1 for ε ; Eq. 2 shows how τ depends on ε ; and the following Eq. shows how r depends on both τ and ε :

$$r = r_o \left(\frac{\rho_p A \delta - m}{\rho_p A \delta + m} \right) \sqrt{\frac{\tau_o}{\varepsilon_o}} \quad (21)$$

where r_o , ε_o , and τ_o are the pore radius, porosity, and tortuosity of an unused membrane, respectively. Eq. 21 was derived by combining Eqs. 1 and 2 with the expressions of membrane porosity before and after membrane compaction [25]:

$$\varepsilon_o = N_{pore} \frac{\pi r_o^2 \tau_o}{A} \quad (22)$$

$$\varepsilon = N_{pore} \frac{\pi r^2 \tau}{A} \quad (23)$$

where N_{pore} is the membrane pore number, which was assumed constant during membrane compaction [25].

According to the mass transfer models, factors that may affect membrane flux include temperature, feed concentration, fluid flow rate, ΔP , P_{air} , and membrane properties that may be altered by ΔP . Because DCMD, PEDCMD, and VEDCMD were all tested under the same experimental conditions except the liquid pressures on the feed and distillate sides, the only factors that may affect water flux are: ΔP , changes in membrane properties (r , ε , τ , and δ) due to membrane compaction, and P_{air} .

3. Results and discussion

3.1. Pressure difference across the membrane

Flux test results are shown in Table 1 for DCMD, PEDCMD, and VEDCMD. The feedwater conductivity was measured near 54.5 mS/cm while the distillate conductivity remained near constant throughout the experiment. Based on the water flux results, two observations can be made. First, DCMD and PEDCMD had similar fluxes (12-13 L/m²-hr) despite the ΔP that exists in PEDCMD. Second, VEDCMD had higher fluxes (16-21 L/m²-hr) than PEDCMD although the same ΔP values of 40, 60, and 80 kPa were used in both. Similar observations were also reported by Cath et al. [20]. The different flux results in the presence of the same ΔP suggest that the enhanced flux in VEDCMD cannot simply be attributed to pressure difference.

3.2. Changes in membrane properties

Membrane properties of the unused membrane and the membranes after flux testing are shown in Fig. 3. The statistical significance data are shown in Table 2. The unused membrane has a

thickness of $67 \pm 10 \mu\text{m}$ and porosity of 80%. By comparing the thickness data for the unused membrane with the membrane after testing in VEDCMD, it can be seen that although the membrane thickness had a relatively large variability, no significant thickness reduction was observed ($p > 0.05$), thus no membrane compaction occurs in VEDCMD. It also can be seen that the thickness was slightly reduced after testing in DCMD ($p = 0.026$), thus the membrane is compacted; similar thickness reduction (10-33%) in DCMD has also been observed by Zhang et al. [25]. For the membranes tested in PEDCMD, thickness reduction greater than 50% was observed ($p = 0.000$). Lawson et al. [37] reported 13-20% membrane thickness reductions for different membranes when the membrane surface was pressurized to 35 kPa (similar to PEDCMD with a ΔP of 35 kPa). Data for membrane porosity are also shown in Fig. 3 and they follow the same trend as the thickness with minimal porosity change in VEDCMD, less than 10% decrease in DCMD and near 30% decrease in PEDCMD. The decrease of membrane porosity will lead to greater tortuosity in DCMD and PEDCMD per Eq. 2, which is also consistent with the literature [25, 26]). The average pore size of each membrane after use was measured with the gas permeation test and no difference was observed in membrane pore size before and after testing. Based on membrane compaction alone (Fig. 3), it is postulated that DCMD flux should be similar to VEDCMD flux and greater than the PEDCMD flux, which was not what was observed experimentally (Table 1). Thus, it is likely that other factors, in addition to membrane compaction, account for the higher water flux in VEDCMD.

3.3. Air pressure inside the membrane pores

3.3.1. Estimation of P_{air} based on literature

Because P_{air} cannot be measured directly, it must be estimated. In a previous DCMD study that employed equal magnitudes of hydraulic pressures on both the feed and distillate sides of the membrane, it was assumed that the air inside the membrane pores was stagnant and that the ideal gas law could be used to calculate P_{pore} (sum of P_{air} and P_i) [25]. For PEDCMD and VEDCMD, it is likely that the stagnant air assumption is not valid due to the transmembrane pressure difference.

To better understand the air pressure inside the membrane pores, the DO concentrations of the feed and distillate streams were measured. As can be seen in Fig. 4, the concentration of DO in DCMD (triangle symbols) and PEDCMD (star symbols) were relatively constant; however, in VEDCMD, the concentration of DO decreased with time – more for the distillate (open circle symbols) than for the feed (solid circle symbols). Based on the VEDCMD pump configuration (Fig 1b), it is likely that both the feed and distillate streams were degassed, with air from the feed side being drawn across the membrane into the distillate stream due to the applied vacuum. This was confirmed by the decreasing DO concentrations in both streams and the observation of air bubbles in the distillate loop in VEDCMD, but not in DCMD or PEDCMD. Based on these observations, the assumption of stagnant air inside the pores for DCMD and PEDCMD may be used as a first estimate for P_{air} in DCMD and PEDCMD, but not for VEDCMD. In VEDCMD, the bubbling of air out from the membrane pores suggests that the assumption of stagnant air inside the pores cannot be used; alternatives to using the ideal gas law are given in Section 3.3.3 for VEDCMD.

Modeled water fluxes for DCMD and PEDCMD using the ideal gas law assumption are given in Fig. 5. Only the Schofield model was used for flux prediction because both the Schofield and dusty-gas models predicted similar water fluxes (results not shown) due to the fact that water flux was only minimally affected by ΔP . For PEDCMD, comparing predicted water fluxes with

measured water fluxes at a thickness near 30 μm (the membrane thickness after the PEDCMD test; Fig. 3), the model under-predicts water flux by approximately 40%. For DCMD, comparing the predicted water flux to the measured water flux at a thickness of 50 μm (the membrane thickness after the DCMD test; Fig. 3), the model over-predicts water flux by approximately 30%. Thus, using the ideal gas law to estimate P_{air} does not lead to good flux predictions in DCMD and PEDCMD; estimation of P_{air} using force balance analysis was then considered.

3.3.2. Estimation of P_{air} for DCMD and PEDCMD using force balance analysis

3.3.2.1. Force balance analysis at the pore surface

According to the force balance analysis, the following relationship exists [46, 47]:

$$P_{fluid} - P_{pore} = \frac{2\sigma \cos \theta}{r} \quad (24)$$

where σ is the surface tension of the liquid and θ is the angle formed between the liquid and the gas relative to the membrane material (Fig. 6) under the applied hydraulic pressure. θ_f was used at the feed side and θ_d at the distillate side. If the θ values are known, P_{pore} (sum of P_{air} and P_i) can be calculated, thus P_{air} can be estimated as P_i can be calculated using Eqs. 5 and 6.

To estimate θ at steady state for DCMD and PEDCMD, the movement of air inside the membrane pores and the structure of the liquid-gas interface were considered. In DCMD, membrane compaction was observed (Fig. 3), leading to a P_{pore} of approximately 100 kPa (estimated using the ideal gas law), which is greater than P_f and P_d (20 kPa). Thus the liquid-gas interface on the feed and distillate sides may extend into the respective liquid stream [25].

Because of the shear stress of the flowing liquid on the membrane surface [46], the air extended into the liquid may be removed, leading to a near flat liquid-gas interface, where θ_f and θ_d reach near 90° (Fig. 6a) and P_{pore} is close to P_f and P_d (Eq. 24). PEDCMD may follow the same process as DCMD until steady state is reached at the distillate side, where P_{pore} is also close to P_d and θ_d reaches approximately 90° (Fig. 6b). Because $P_f > P_{pore}$, the feed water may extend into the membrane pores, leading to θ_f less than 90° (Fig. 6b) to maintain equilibrium at the feed side [25]. Therefore, the same conclusion ($P_{pore} = P_d$) may be reached for both DCMD and PEDCMD.

3.3.2.2. Estimation of P_{air} for DCMD and PEDCMD using the $P_{pore} = P_d$ assumption

The modeled water fluxes for DCMD and PEDCMD using the $P_{pore} = P_d$ assumption are given in Fig. 7. Modeled water fluxes are given for a ΔP of 60 kPa. Comparing the measured water fluxes in PEDCMD to the predicted water fluxes at a thickness near 30 μm , it can be seen that the Schofield model gives an excellent fit for PEDCMD but over-predicts water flux by approximately 30% for DCMD, similar to that for the ideal gas law assumption. Because P_d can be controlled experimentally and the magnitude can be measured easily, the $P_{pore} = P_d$ assumption can be considered as a fast and reliable estimation of P_{air} for both DCMD and PEDCMD.

3.3.3. Estimation of P_{air} in VEDCMD

To account for the observed flow of air inside the pores in VEDCMD (discussed with Fig. 4), a first approximation for P_{pore} is to set it equal to P_d (as was done by Zhang et al. [21]). Modeled fluxes for VEDCMD using the $P_{pore} = P_d$ assumption are given in Fig. 8 (blue lines). As can be seen, higher water fluxes are predicted by this model with greater vacuum pressures on the distillate side, which is consistent with the change of the measured water flux. The increased

water flux is attributed to the decreased P_{pore} (thus P_{air}) according to the $P_{pore} = P_d$ assumption, thus this model also supports the view in the literature that air pressure inside the membrane pores plays an important role on water flux in MD [19-21]. However, the predicted water fluxes in the membrane thickness range after the VEDCMD testing (55 to 65 μm ; Fig. 3) are greater than the experimentally measured water fluxes by 40-70%, suggesting that the $P_{air} = P_d$ assumption is not valid under VEDCMD conditions despite what has been suggested in the literature [21]. Thus, a better approximation for the pressure inside the membrane pores is required.

Because a pressure difference was present across the membrane and air flux was observed in VEDCMD (Fig. 4), a new estimate of the P_{pore} value was taken as the average of the feed and distillate side pressures ($P_{pore} = (P_f + P_d)/2$). Modeled fluxes using this assumption are given in Figure 8 (black lines). As was observed with $P_{pore} = P_d$, higher water fluxes are predicted with increased distillate-side vacuum pressures, but in this case the prediction errors obtained assuming the average pressure are much lower (<20%) in the thicknesses corresponding to VEDCMD conditions (55-65 μm ; Fig. 8). Therefore, the $P_{pore} = (P_f + P_d)/2$ assumption may be used as an easy and fast estimation of water flux in VEDCMD once the hydraulic pressures and vacuums on the feed and distillate sides are known. Because the calculated P_{air} using the $P_{pore} = (P_f + P_d)/2$ assumption for VEDCMD (0, -10, and -20 kPa, respectively, at P_f/P_d of 20/-20, 20/-40, and 20/-60 kPa) is lower than the calculated P_{air} using the $P_{pore} = P_d$ assumption for DCMD and PEDCMD (20 kPa for both DCMD and PEDCMD), the lower P_{air} was identified as another dominant factor (besides less membrane compaction) contributing to the improved water flux in VEDCMD.

4. Conclusion

Compared to traditional DCMD, VEDCMD offers opportunities for higher fluxes that may be particularly useful in treatment of water with low fouling and scaling potential. Transmembrane pressure difference was ruled out as providing significant contribution to the higher flux through testing with PEDCMD. The factors that were found to result in the higher water flux in VEDCMD were less membrane compaction and lower air pressure inside the membrane pores. Therefore, approaches to further improve the VEDCMD process should focus on reducing air pressure inside the membrane pores without compacting the membrane or significantly increasing the energy requirements of the process.

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Nomenclature and Units

A	Membrane surface area (m^2)
A_c	Cross-section area of module channel (m^2)
B_o	Intercept of a linear plot
C_p	Heat capacity of distillate (kJ/kg K)
d	Membrane average pore size (m)
d_f	Diameter of spacer filament (m)
d_h	Hydraulic diameter of the flow channel (m)
D	Diffusion coefficient (m^2/s)
h	Heat transfer coefficient ($\text{W/m}^2 \text{ K}$)
$h_{(f,d)}$	Heat transfer coefficient in the boundary layer on the feed or distillate side ($\text{W/m}^2 \text{ K}$)
H	Channel height (m)
k_{dc}	Parameter in Eq. 13
k_f	Thermal conductivity of the fluid (W/m K)
K	Slope of a linear plot
l_m	Mesh size of the spacer (m)
m	Mass of membrane sample (kg)
M	Molecular weight (kg/kmol)
$M_{(i,j)}$	Molecular weight of vapor or air (kg/kmol)
N	Overall flux ($\text{L/m}^2 \text{ h}$)
$N_{(k,m,v)}$	Water flux due to Knudsen diffusion, Molecular diffusion, or viscous flow ($\text{L/m}^2 \text{ h}$)
N_{pore}	Membrane pore number
Nu	Nusselt number
P_{air}	Air pressure inside the membrane pores (Pa)
P_{fluid}	Hydraulic pressure of the liquid stream (Pa)
P_i	Water vapor pressure (Pa)
$P_{i(f,d)}$	Water vapor pressure at the feed or distillate side (Pa)
P_{pore}	Membrane pore pressure (sum of P_{air} and P_i , Pa)
Pr	Prandtl number
r	Membrane pore radius (m)
r_o	Original membrane pore radius (m)
R	Universal gas constant (J/mol K)
Re	Reynolds number
T	Membrane temperature during the gas permeation test
$T_{(dout, din)}$	Distillate-side temperature at the outlet or inlet of the membrane cell (K)
$T_{(f,d)}$	Bulk temperature of the feed or distillate solution (K)
T_m	Average membrane temperature (K)
$T_{(mf, md)}$	Membrane surface temperature on the feed or distillate side (K)
x_{NaCl}	Molar fraction of NaCl inside the solution
δ	Membrane thickness (m)
ΔP	Pressure gradient (Pa)
ΔP_i	Pressure gradient of water vapor (Pa)
ε	Membrane porosity
ε_o	Original membrane porosity
ε_s	Spacer porosity
θ	Angle formed among liquid, membrane material, and gas inside the membrane pores ($^\circ$)

$\theta_{(f,d)}$	θ at the feed or distillate side ($^{\circ}$)
θ_s	Hydrodynamic angle of the spacer ($^{\circ}$)
μ	Vapor dynamic viscosity (kg/m s)
$\mu_{(f,g)}$	Liquid and gas dynamic viscosity (Pa s)
$\dot{m}$	Fluid flow rate (L/min)
$\rho_{(p,f)}$	Density of membrane polymer material or fluid (kg/m ³)
τ	Membrane tortuosity
τ_o	Original membrane tortuosity

List of Tables

Table 1: Flux test results for DCMD, PEDCMD, and VEDCMD (T_f : 40 °C, T_d : 20 °C, flow rate: 1.0 L/min).

Table 2: Statistical analysis for membrane thickness comparison after testing in DCMD, PEDCMD, and VEDCMD.

TABLES

Table 1

Configuration	Feed/distillate pressure (kPa)	Pressure gradient (kPa)	Flux (L/m ² h)
DCMD	20/20	0	12.1±0.6
	60/20	40	12.3±1.1
PEDCMD	80/20	60	12.1±1.1
	100/20	80	13.1±1.5
VEDCMD	20/-20	40	16.4±0.5
	20/-40	60	17.6±0.5
	20/-60	80	20.8±0.7

Table 2

Configuration	Feed/distillate pressure (kPa)	<i>p</i> value
DCMD	20/20	0.026
	60/20	0.000
PEDCMD	80/20	0.000
	100/20	0.000
VEDCMD	20/-20	0.813
	20/-40	0.251
	20/-60	0.235

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Fig. 1. Schematic drawings of the bench-scale MD system configurations for (a) DCMD and PEDCMD ([9]) and (b) VEDCMD (adapted from [20]).

Fig. 2. Thermal resistance circuits of (a) the Schofield model and (b) the simplified dusty-gas model [45].

Fig. 3. Membrane thickness and porosity before and after testing in DCMD, PEDCMD, and VEDCMD.

Fig. 4. Dissolved oxygen concentration inside the feed and distillate streams in DCMD, PEDCMD, and VEDCMD.

Fig 5. Tested and modeled water fluxes as a function of membrane thickness using the ideal gas law assumption for DCMD and PEDCMD (line: modeled fluxes for DCMD and PEDCMD; symbol: tested fluxes).

Fig. 6. Force balance at the liquid-gas interface in (a) DCMD and (b) PEDCMD (note: the figure is a schematic representing the pressure effects at the pore surface and does not reflect any inter-pore behavior).

Fig. 7. Tested and modeled water fluxes as a function of membrane thickness using $P_{pore} = P_d$ assumption for DCMD and PEDCMD (line: modeled fluxes; symbol: tested fluxes).

Fig. 8. Tested and modeled water fluxes as a function of membrane thickness using the $P_{pore} = P_d$ and $(P_{pore} = (P_f + P_d)/2)$ assumptions for VEDCMD (line: modeled fluxes; symbol: tested fluxes).

FIGURES

(a)

(b)

Figure 1

(a)

(b)

Figure 2

Figure 3

Figure 4

Figure 5

(a)

(b)

Figure 6

Figure 7

Figure 8