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Effect of flow rate, draw solution concentration and temperature on the performance of TFC FO membrane, and the potential use of RO reject brine as a draw solution in FO–RO hybrid systems

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ABSTRACT

The main objective of this research study was to investigate the effect of feed and draw solution flow rate, draw solution concentration (2.5–7.7 wt% NaCl) and draw solution temperature (23°C–60°C) on the performance of a commercial polyamide thin-film composite forward osmosis (FO) flat sheet membrane in the active-layer-facing-draw solution (AL-DS) membrane orientation. Increasing the feed and draw solution flow rate improved the membrane flux by mitigating concentration polarization effects on both sides of the membrane. The membrane flux also increased at higher draw solution concentration due to higher osmotic pressure. Additionally, it was found that increasing the draw solution temperature slightly improved the membrane flux but the temperature effect was negligible due to the severe effect of concentration polarization. It was observed from experimental results that the salt rejection was maintained above 98% at all operating conditions, and an optimal water flux of 11.4 L m^{-2} h^{-1} was reported at a flow rate of 48 mL min^{-1} with deionized water as feed and 7.7 wt% NaCl as draw solution at 40°C. The water flux across the membranes decreased when brine at 7.7 wt% NaCl was used as the draw solution and raw seawater as the feed solution in active-layer-facing-feed solution (AL-FS) membrane orientation because of concentration polarization effects and lower osmotic pressure. The results showed that the performance of FO membranes is influenced by the operating conditions. Therefore, optimizing these conditions is essential and can significantly improve the performance of FO membranes.

Keywords: Forward osmosis; Thin-film composite membrane; Water flux; Reverse salt flux; Brine dilution; Desalination

1. Introduction

The exponential growth of the world population has significantly burdened the existing water and energy resources. The complexity of the water-energy nexus occurs because a lot of energy may be required to obtain fresh water in some cases, whereas, a significant amount of water may be needed to generate power in others. The inability to meet the rising demands for fresh water from the available freshwater resources requires consumable water to be produced from saline water or wastewater. For instance, arid regions such as the Middle East heavily rely on energy-intensive thermal processes such as multi-stage flash distillation for seawater/brackish water desalination. Reverse osmosis (RO) is also used for freshwater production and has the largest share of desalination market, but it requires a significant amount of energy to produce external hydraulic pressure across the membrane as the driving force. On the other hand, forward osmosis (FO) is a membrane separation technology that is driven by the osmotic pressure gradient existing between the

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high concentration solution (draw) and the low concentration solution (feed) across a semi-permeable membrane [1]. In the FO process, water diffuses from feed to draw solution across the membrane. Since FO does not require hydraulic pressure to operate, it has gained a lot of attention due to its lower fouling propensity, energy requirement and operational cost [2]. Additionally, FO can be used for many applications such as product concentration (food processing), wastewater concentration and power generation [3]. Compared with standalone RO process, desalination of high salinity waters is much more economically feasible via FO hybrid systems where FO is used initially to dilute the seawater [4].

Despite the desirable characteristics of FO, there are several challenges associated with the types of draw solutions and membranes that hinder the implementation of FO in large-scale processes. An ideal draw solution must be non-toxic, sustain a high osmotic gradient, economically viable, and allow easy recovery of draw solute at marginal energy consumption [1,5]. In addition, a good FO membrane should be able to minimize concentration polarization (CP), fouling and reverse salt diffusion to improve water flux across the membrane [3,6]. The asymmetric nature of FO membranes results in CP effects, which reduce the driving force and, hence, the water flux across the membrane [7]. The occurrence of CP hinders the process efficiency and results in higher capital costs [8]. CP can be categorized into two types: (1) external concentration polarization (ECP) that occurs on the active layer of the membrane, and (2) internal concentration polarization (ICP) that occurs within the porous support layer of the membrane. Depending on the membrane orientation, ECP and ICP can be further classified as dilutive external concentration polarization (DECP), concentrative external concentration polarization (CECP), dilutive internal concentration polarization (DICP) and concentrative internal concentration polarization (CICP). DECP and CICP take place when the active layer is oriented towards the draw solution (AL-DS); whereas, CECP and DICP occur when the active layer faces the feed solution (AL-FS) [9]. The dilutive phenomenon occurs due to the dilution of draw solution concentration at the membrane surface by the incoming water flux. The incoming water flux not only dilutes the draw solution at the membrane boundary layer, but the convective force diffuses the draw solutes away from the membrane boundary. This dilutive phenomenon on the DS alone contributes to up to 80% of the flux decline during the FO process [10,11]. Several studies have tried to mitigate CP effects in the FO membranes by optimizing FO operating conditions, using spacers, fabricating thin hydrophilic membranes with porous and less tortuous substrates, and developing highly soluble draw solutions with high diffusion coefficients and low molecular weight [8,10–18]. Performance of commercial FO membranes has been studied using bench-scale experiments for several applications in arid regions. For instance, commercial TFC membrane was tested for volume reduction of process water from oil and gas operations in Qatar. The results indicated that the FO process was able to reduce the volume of pretreated process water by 50% [19]. Another study in Saudi Arabia used FO coupled with low-pressure reverse osmosis (LPRO) for indirect desalination of seawater using secondary wastewater effluent as feed. It was observed that the energy consumption of the process was 50% less than that used for high-pressure RO. It was concluded from the cost analysis that FO coupled with LPRO technology might prove to be economically viable [20].

Brine generation from desalination plants in the United Arab Emirates (UAE) is a challenge both on the environmental and economic front [21]. Concentrated waste brine from large-scale desalination plants are directly disposed of in neighbouring seawater bodies; whereas, inland and smaller capacity desalination plants carry out land disposal of concentrated brine in evaporated ponds or impoundments [6]. It is also widely understood that the energy profile in the UAE, as well as the harsh seawater in the UAE characterized by high salinity and temperature, has encouraged the proliferation of thermal-based desalination processes as opposed to the more economical and environmental friendly membrane based option, that is, RO [8,22]. One of the earlier studies reported on the possible concentration of brine using FO commercial membrane and the need for more process optimization [6]. However, the energy needed to operate a standalone FO system would be higher than an RO system. Nonetheless, an FO–RO hybrid system can be potentially more energy-efficient and economical in treating high saline water by reducing source salinity in the FO dilution stage [4,20,23].

In this study, the effect of feed and draw solution flow rate (14.4 and 48 mL min^{-1}), draw solution concentration (2.5–7.7 wt% NaCl) and draw solution temperature (23°C–60°C) on the FO performance of commercial thin-film composite (TFC) FO flat sheet membrane was experimentally studied to enhance the FO membrane water flux. Additionally, the potential of using brine as draw solution and seawater as feed solution in FO process was explored. We believe that this research study will contribute to the advancement of FO research by providing information on the FO membrane performance for seawater and brine dilution in arid regions such as the UAE.

2. Experimental

2.1. Materials

Polyamide TFC FO flat sheet membranes (HTI OsMem™ TFC-ES) from Hydration Technology Innovation (HTI) were tested for performance in this study. The commercial HTI TFC membrane comprised of a polyamide active layer formed via interfacial polymerization on top of a polysulfone substrate, which was embedded on a polyester mesh as backing support. The maximum operating transmembrane pressure and temperature for the membrane was 0.7 atm and 71°C, respectively and can be operated within the solution pH range between 2 and 11. To increase the wettability of the TFC membranes, they were soaked in 50% ethanol solution for 5 min and then rinsed with deionized (DI) water before testing to ensure complete saturation of the porous support layer of the membrane with water. Synthetic draw solutions of different concentrations (2.5, 3.5 and 7.7 wt% NaCl) were made by dissolving reagent grade NaCl (VWR BDH Prolabo®, Leuven, Belgium) in DI water. DI water was used as a feed solution for all experiments in the AL-DS mode, while raw seawater was used as a feed solution for experiments in the AL-FS mode. Raw seawater employed in this study was collected from Khor Al-Raha beach in Abu Dhabi (UAE).

2.2. FO experimental setup

The bench-scale FO experimental setup used in this work is illustrated in Fig. 1. The FO membrane test cell consists of two channels on each side of the membrane with an effective membrane area of 36 cm^2 (9 cm length, 4 cm width and 3 mm depth). Polyester spacers are installed in both flow channels of the FO cell to support the membrane and elevate turbulence for minimizing ECP. The membrane cell was sealed with O-rings to prevent leakage and hold the membrane in place. Feed and draw solutions are introduced co-currently into the membrane cell. A two-channel variable-speed peristaltic pump (Masterflex®, Cole-Parmer, USA) was used to circulate feed and draw solutions in a closed loop system at the desired flow rates. A water bath (WB-22, Witeg Labortechnich GmbH) was employed to create an elevated temperature environment. The membrane cell was left outside the water bath when tested at room temperature. For experiments conducted at higher temperatures of 40°C and 60°C, the membrane cell was immersed in a water bath and the draw solution was placed on a hot plate magnetic stirrer. The feed solution temperature was maintained at room temperature for all experiments. The volume of draw solution (1,800 mL) used was higher than that of feed solution (700 mL) in order to maintain approximately steady draw solution concentration throughout the 4-h experiment. The feed solution was placed on a balance with the built-in RS-232 bi-directional communication interface, which was connected to a data logging system to record the change in feed solution weight with time to calculate the water flux. The total dissolved solids of both feed and draw solution were measured using a conductivity meter (YSI, Model 52, USA) to determine the salt flux.

2.3. Transport properties

Water flux (J_w) across the FO membrane is obtained using Eq. (1):

$$J_w = \frac{W_0 - W_e}{\rho \cdot \Delta t \cdot A_m} \tag{1}$$

where W_0 and W_e are the initial weight and final weight at time t of the feed solution (g), ρ is the water density (g L^{-1}), Δt is the duration of performance test (h) and A_m is the effective surface area of the membrane (m^2).

Reverse salt flux (J_s) across the FO membranes is calculated using Eq. (2):

$$J_s = \frac{W_e \cdot C_{f,e} - W_0 \cdot C_{f,0}}{\rho \cdot \Delta t \cdot A_m} \tag{2}$$

where $C_{f,e}$ and $C_{f,0}$ are the initial concentration and final concentration at time t of the feed solution (g L^{-1}), respectively.

Salt rejection (R) of the FO membranes is calculated using Eq. (3):

$$R = \left(1 - \frac{\frac{J_s}{J_w}}{\frac{\left(C_{d,0} + C_{d,e}\right)}{2}}\right) \times 100\% \tag{3}$$

where $C_{d,0}$ and $C_{d,e}$ are the initial concentration and final concentration at time t of the draw solution (g L^{-1}), respectively.

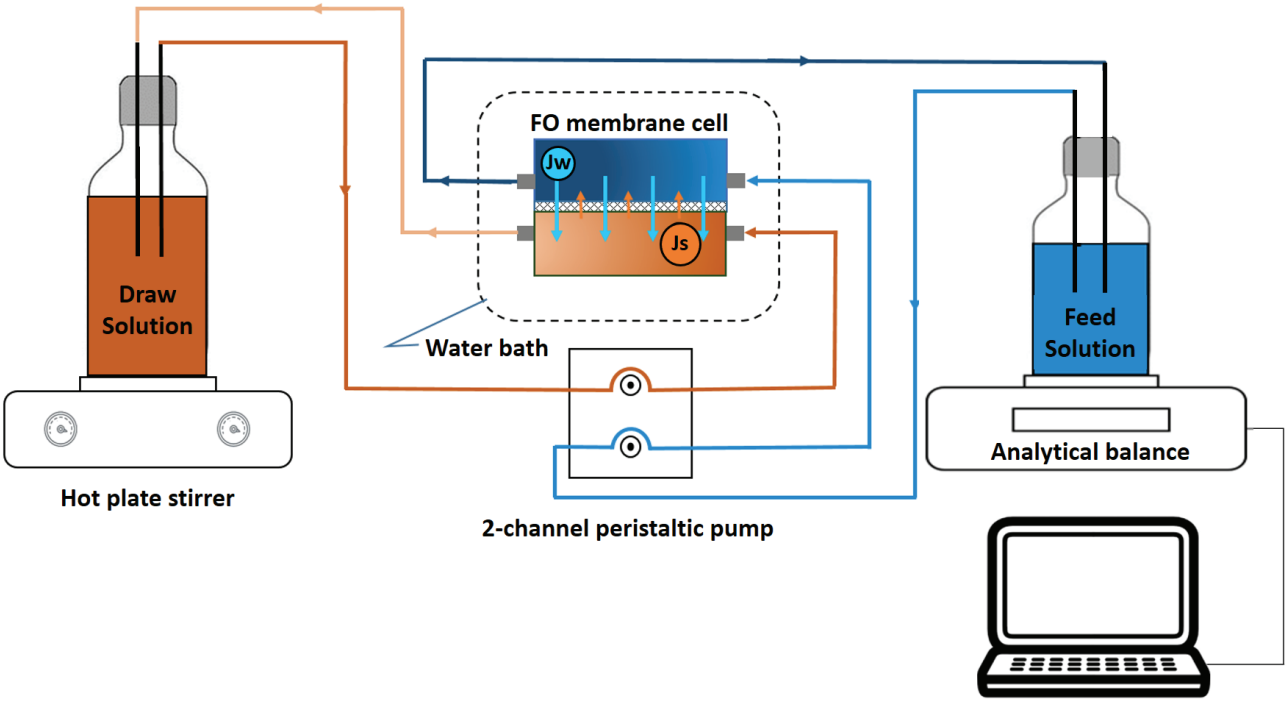

Fig. 1. A schematic diagram of the bench-scale FO experimental setup that was used for testing the performance of FO commercial membranes.

3. Results and discussion

3.1. Effect of flow rate on water and salt flux

The effect of increasing feed and draw solution flow rates on the membrane performance was studied using DI water and 3.5 wt% NaCl as feed and draw solution, respectively. The FO tests were conducted in AL-DS orientation at a temperature of 40°C. It can be observed from Fig. 2(a) that the water flux increased by about 84% from 5.1 to 9.4 L $m^{-2} h^{-1}$ (LMH) when the flow rate was increased from 14.4 to 48 mL min^{-1}. The improvement in water flux resulted from the occurrence of turbulence at a higher flow rate or cross-flow velocity that helped to mitigate the occurrence of DECP at the membrane surface by mixing the permeate water faster in the bulk draw solution. DECP occurs on the dense active layer due to the incoming permeate flux that dilutes the draw solution close to the active layer on the permeate side of the membrane and decreases the effective driving force due to diminished osmotic pressure on the side of draw solution [7,24]. Since the flow rate of both draw and feed solutions was increased, the CP effects were simultaneously reduced on both sides of the membrane. The reverse salt flux for the TFC membrane was measured at different feed and draw solution flow rates, and it was observed to increase from 1.4 to 2.9 g $m^{-2} h^{-1}$ (GMH) when the flow rate was increased (Fig. 2(b)). This confirms that reverse salt flux increases with increasing water flux due to a tradeoff between selectivity and permeability. The salt rejections at both flow rates were comparable and were approximately maintained at 99%.

3.2. Effect of draw solution concentration on water and salt flux

The effect of increasing draw solution concentration on the membrane performance was investigated using DI water as feed solution in AL-DS orientation with a flow rate of 48 mL min^{-1} and a temperature of 40°C. As can be observed in Fig. 3(a), increasing the draw solution concentration from 2.5 to 7.7 wt% increased the water flux by about 52% from 7.5 to 11.4 LMH due to the increase of the osmotic pressure difference between the feed and draw solution. The reverse salt flux also increased at higher draw solution concentration because the draw solute leakage from draw to feed solution increases at higher salt concentration gradient (Fig. 3(b)). The salt rejection was found to increase slightly at higher draw solution concentration. However, the salt rejections obtained at all draw solution concentrations can be said to be comparable and was observed to be higher than 98.6%. It is important to note that increasing the draw solution concentration will not proportionally increase membrane water flux because of the ICP, which is an intrinsic property of the membrane. ICP cannot be mitigated such as ECP by manipulating operating conditions, but it can be partially addressed by improving the design of the membrane substrate structure. An optimal membrane should be thin, have low tortuosity and high porosity to reduce ICP [23].

3.3. Effect of draw solution temperature on water and salt flux

The effect of increasing draw solution temperature on the membrane performance was investigated using DI water and 3.5 wt% NaCl as feed and draw solution, respectively, in AL-DS orientation with a flow rate of 48 mL min^{-1}. Draw solution temperature was studied instead of feed solution because seawater (feed) temperature is usually constant. The lowest draw solution temperature was fixed at the ambient temperature of 23°C; whereas the highest temperature was chosen as 60°C to stay below 71°C, which is the maximum temperature rating of the membrane. Based on the Van't Hoff equation, the membrane flux was expected to increase on increasing the draw solution temperature due to an increase in the osmotic pressure. Nonetheless, increasing the draw solution temperature did not significantly improve water flux because of only a slight increase in osmotic pressure (Fig. 4(a)). Water flux increased by about 6.5% when the draw solution temperature was increased from 23°C to 40°C. The reduced viscosity of the draw solution at higher temperature increases water permeability coefficient by facilitating diffusion between draw solution

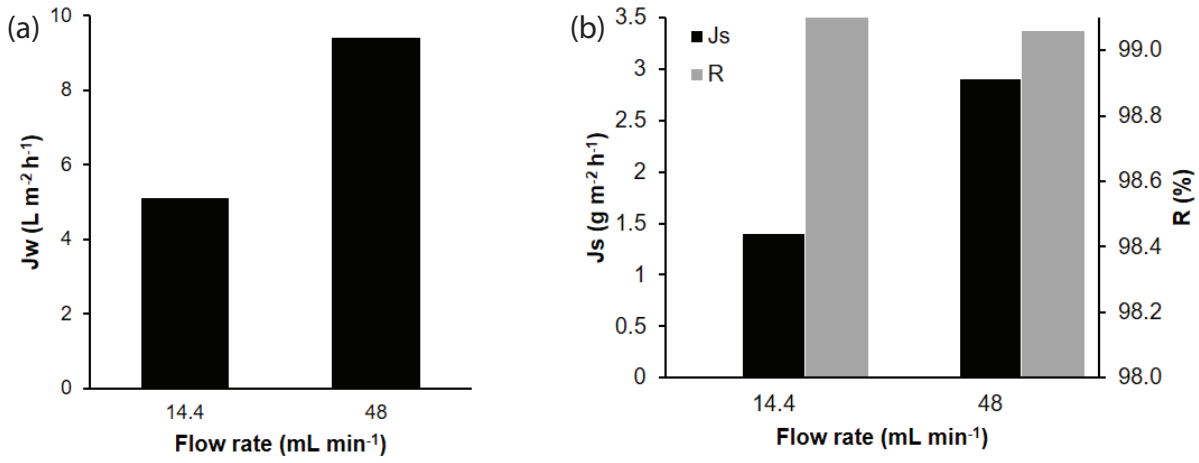

Fig. 2. Effect of increasing feed and draw solution flow rates on the (a) water flux, (b) salt flux and salt rejection at draw solution concentration of 3.5 wt% NaCl and temperature of 40°C, using DI water as feed in AL-DS mode.

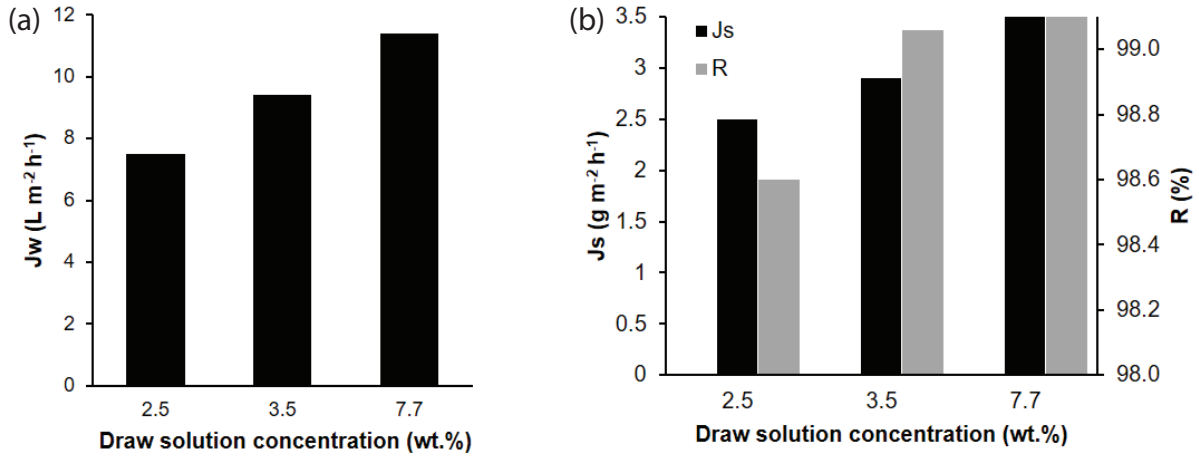

Fig. 3. Effect of increasing draw solution concentration on the (a) water flux, (b) salt flux and salt rejection at a flow rate of 48 mL min^{-1} and temperature of 40°C using DI water as feed in AL-DS mode.

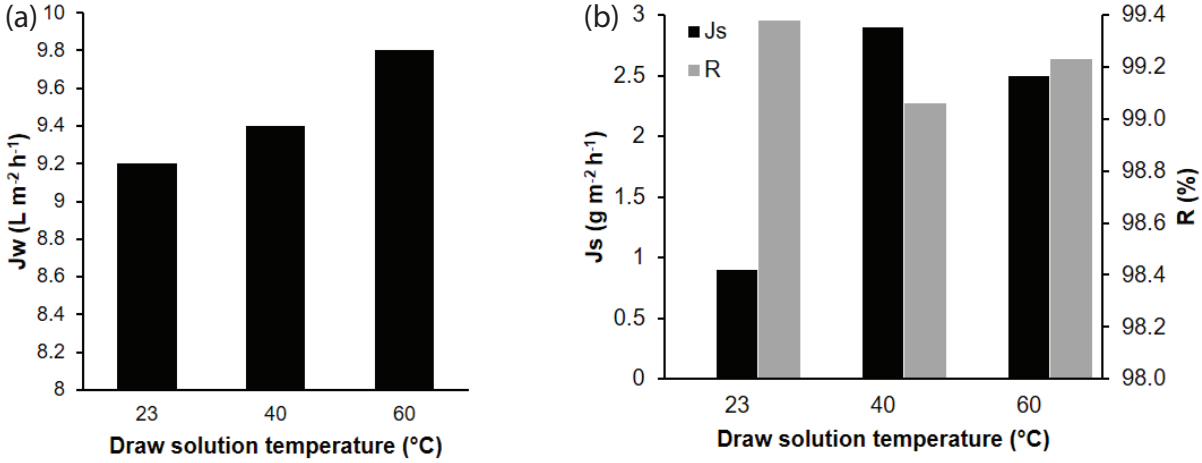

Fig. 4. Effect of increasing draw solution temperature on the (a) water flux, (b) salt flux and salt rejection at a flow rate of 48 mL min^{-1} using 3.5 wt% NaCl as draw solution and DI water as feed in AL-DS mode.

present in the bulk solution and in the porous support layer to partially lower the effect of ICP [25,26]. Nonetheless, the impact of temperature on water flux was minimal because the effect of ICP and ECP are severe at higher water fluxes. Since ICP and ECP moduli are exponential functions of water flux and increase with the membrane's water permeability coefficient, the temperature of draw solution will have a negligible impact on the membrane flux in the presence of CP.

Additionally, salt rejection may have decreased due to the increase in the diffusion coefficient of NaCl draw solution at a higher temperature [11,19]. Reverse salt flux increased from 0.9 to 2.9 GMH when the draw solution temperature was raised from 23°C to 40°C (Fig. 4(b)). However, the change in temperature between 40°C and 60°C showed minimal impact on the reverse salt flux reporting 2.9 and 2.5 GMH, respectively. The salt rejection was maintained above 99.1% at all temperatures. The potential for FO for seawater and brine dilution was investigated based on the above observations.

3.4. FO seawater and brine dilution in Abu Dhabi, UAE

The potential application of FO membrane for seawater desalination in the UAE using FO–RO hybrid desalination process was also explored using seawater as a draw solution agent so that it can be diluted before the RO process. Both seawater and NaCl were used as draw solutions at a similar salinity level (3.5 wt% NaCl) to establish a baseline. The draw solutions were tested at a temperature of 40°C against DI water as feed at a flow rate of 48 mL min^{-1} in AL-FS mode. Fig. 5 demonstrates the experimental results obtained using the two draw solutions. The water flux was comparable for the two types of draw solution, where 9.4 and 9.6 LMH were obtained for 3.5 wt% NaCl and seawater, respectively (Fig. 5(a)). The salt flux recorded for both draw solutions did not exceed 4 GMH and was slightly lower for the synthetic 3.5 wt% NaCl solution. As can be seen from Fig. 5(b), the recorded salt flux was 2.9 GMH and 3.8 GMH for synthetic NaCl draw solution and seawater, respectively. Salt rejection was maintained above 98% using both the draw solutions.

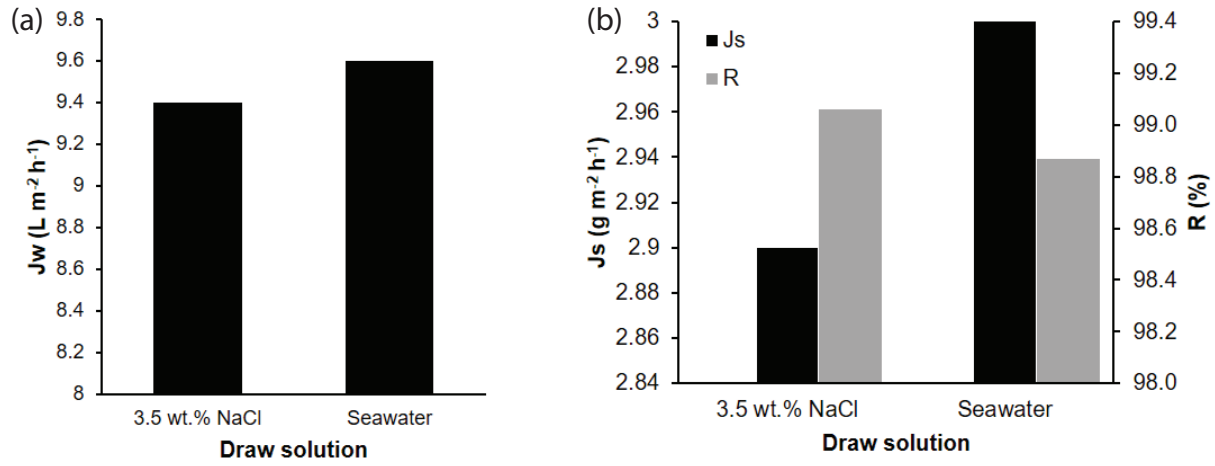

Fig. 5. Effect of using seawater and 3.5 wt% NaCl as draw solution on the (a) water flux, (b) salt flux and salt rejection of the membrane at a flow rate of 48 mL min^{-1} and temperature of 40°C using DI water as feed in AL-FS mode.

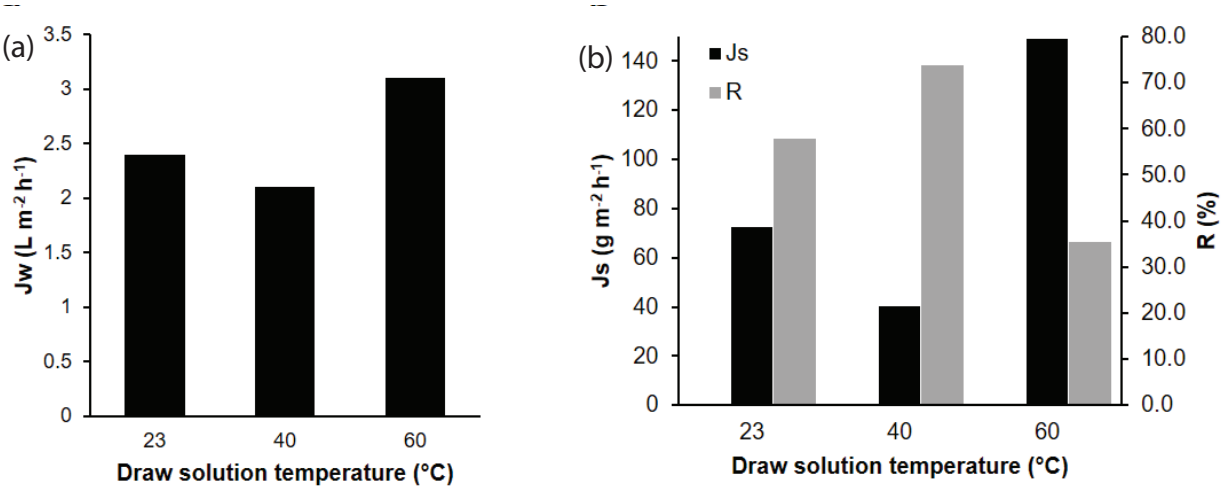

Fig. 6. Effect of using seawater and 7.7 wt% NaCl as feed and draw solution, respectively on the (a) water flux, (b) salt flux and salt rejection of the membrane at a flow rate of 48 mL min^{-1} in AL-FS mode at different draw solution temperatures.

The feasibility of FO TFC membrane for brine dilution was tested using 7.7 wt% NaCl as draw solution (synthetic brine) at various temperatures and seawater as feed solution at a flow rate of 48 mL min^{-1} in AL-FS mode. It was observed that the transmembrane water flux was overall low, and the highest water flux was obtained at 60°C as 3.1 LMH (Fig. 6(a)). The salt flux was also very high especially at 60°C, resulting in low salt rejection of 35% (Fig. 6(b)). The low transmembrane water flux can be justified due to the lower osmotic pressure difference and the occurrence of severe DICP that limited the separation performance of FO membranes [22].

Furthermore, the movement of salt molecules towards both sides of the membrane and their accumulation at the surface could have lowered the apparent osmotic potential gradient across the membrane and in turn resulted in poorer water flux performance [8]. The high reverse salt flux further supports these observed results (Fig. 6(b)). It is clear that the membrane capacity at generating high water flux while maintaining low salt reversal is still limited; therefore, more membrane designs need to be explored that have better separation properties and can withstand high temperature to allow effective use of this integration in the future.

4. Conclusion

In this study, the influence of draw/feed solution flow rate, draw solution concentration and temperature on the FO performance was investigated to get a better understanding of the influence of operating conditions on membrane performance. Water and reverse salt flux were observed to increase at a higher flow rate and draw solution concentration due to ECP mitigation and higher osmotic driving force, respectively. Higher draw solution temperatures also enhanced water flux across the membrane due to reduced water viscosity and higher diffusion rate; however, the enhancement in water flux was not very significant because the effect of temperature on membrane flux is negligible in the presence of severe CP effects. The potential of FO for seawater

desalination and brine dilution in the UAE was also explored by using commercial TFC membranes. Results revealed that the water flux obtained were similar when using seawater and synthetic NaCl as draw solution and DI water as feed solution. However, the application of synthetic brine as draw solution against seawater as feed resulted in a reduction in water flux due to the lower osmotic pressure difference and severe ICP effects. Based on the results, this study concludes that operating conditions can significantly influence FO performance by influencing CP; and that novel draw solutions with high osmotic pressure and FO membranes with good separation properties that can withstand the high temperature and salinity of UAE's seawater are required to mitigate CP effects and improve the FO process viability.

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Symbols

A_m — Membrane surface area, m^2
$C_{f,0}$ — Initial draw solute concentration in the feed solution
$C_{f,e}$ — Final draw solute concentration in the feed solution, g L^{-1}
J_s — Reverse salt flux across FO membrane, g m^{-2} h^{-1}
J_w — Water flux across FO membrane, L m^{-2} h^{-1}
R — Salt rejection, %
W_0 — Initial feed solution weight, g
W_e — Final feed solution weight, g
Δt — Duration of performance test, h
ρ — Water density, g L^{-1}

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