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Experimental study of a cellulose triacetate spiral wound forward osmosis membrane for desalination process integration

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ABSTRACT

This paper aimed at assessing the feasibility of the forward osmosis (FO) membrane process step of the FO desalination process, on a laboratory bench-scale test unit using batch mode, for extracting freshwater from feed and simultaneously diluting draw solution (DS). A commercially available spiral wound FO membrane made of cellulose triacetate (CTA) was experimentally tested at different operating conditions. Several affecting parameters, namely, feed concentration, draw solute concentrations, flow rate, and temperature, on water flux and permeate water recovery ratio were investigated. Deionized (DI) water, sodium chloride (NaCl) solutions, Gulf seawater and reverse osmosis (RO) brine were used as feed solutions (FS). Different concentrations of NaCl solutions ranging from 3.5 to 26 wt% were used and tested as DS. The results showed that the water flux increased with increasing DS osmotic pressure. The experimental results indicated that the water flux is dependent on the temperatures of FS and DS. It was observed that the water flux is directly proportional to the temperature of the FS and DS. The experimental results were highly encouraging, and proved that the FO membrane stage could be an efficient desalination system component for either desalting seawater or concentrating highly saline waters including RO brine.

Keywords: Forward osmosis membrane; Desalination technologies; Draw solution; osmotic pressure; Concentration polarization

1. Introduction

Desalination remains the most important and viable source of freshwater in Kuwait and the Gulf Cooperation Council (GCC) countries. The cost of desalination has declined in recent years due to technological advancements and better management. All indications are that desalination technology will play a major role in providing potable water to coastal cities and industries. Among these desalination methods, multi-stage flash (MSF) and reverse osmosis (RO) are the most widely used. MSF is a thermal distillation process that is commonly used for desalinated water production facilities around the world. However, the process is usually coupled with power generation plants and suffers from high capital and operating costs and low recovery ratios compared with RO. RO has become increasingly popular as an alternative seawater desalination technology, as it is currently producing freshwater at lower cost compared with the conventional thermal desalination systems, due to the result of the ongoing, continuous improvements in RO technologies. However, RO still has a number of challenges such as significant concentration polarization, scaling and fouling [1–3]. Additionally, RO is considered as an energy-intensive system because it requires operating pressure greater than 50 atm, and the requirement of a high hydraulic pressure to overcome the osmotic pressure generated by seawater [1,2]. Furthermore, RO has a limited water recovery ratio

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(which is between 30% and 50%) while it generates large volumes of brines [2].

Non-conventional desalination technologies to yield freshwater have received worldwide attention due to water scarcity. Forward osmosis (FO) technology is one of those technologies that have received extensive attention during the last decade as emerging process for seawater desalination. The RO process uses hydraulic pressure as the driving force to transport water through the membrane; whereas, FO takes advantage of naturally induced freshwater transport across a semi-permeable membrane from feed solution (FS) at a lower salt concentration to the aqueous solution at higher salt concentration, known as draw solution (DS). Ideally, the semi-permeable membrane allows only freshwater to pass through the membrane leaving all organic and inorganic salts behind. The DS has higher osmotic pressure than that in FS, to induce freshwater flow across the membrane, and, thus, FO does not rely on a high pressure pump as in a pressure-driven membrane process (i.e., RO) to transport a net water flow across the membrane. Therefore, FO requires less energy in comparison with RO. However, in contrast to RO, the product water of FO technology is unfortunately not a freshwater that can be immediately used as drinking water. The product of FO is diluted DS, that is, a mixture of DS and freshwater. Therefore, a second step of separation unit, known as a regeneration stage, must be utilized to recover DS and freshwater.

Previous studies reported that FO is an emerging process and it has the potential capability to be used for seawater desalination application in a more environmental-friendly context compared with the commercially available desalination technologies [4–6]. Previous and current studies have compared FO with RO technology and reported the following advantages: FO requires between 20% and 30% less consumption energy [7], it has much higher permeate recovery ratio (recovery ratio at least 75%) [4] and discharges lower volume of brine to the environment [8], it has low fouling potential and high cleaning efficiency [9,10], higher boron rejection [11].

FO flux rates (permeate water recoveries) can be significantly greater than RO flux rates due to the osmotic driving forces of FO, which can exceed 250 atm (depending on the DS used) and is expected to produce more than 75% recovery of seawater as reported in the literature [4]. The alternate osmotic agent used as a DS would then be recovered from the high Total Dissolved Solids (TDS) product water to render that water desalinated. This process of individual DS TDS recovery is discussed in other work dependent on system architecture and specific application.

FO can be extended for a number of vital applications, such as seawater desalination [12–14], oil produced water [15], power regeneration [16–18], digested sludge [19,20], liquid food processing [21], protein concentration [22,23], municipal wastewater treatment [24,25], industrial wastewater treatment/reuse [26–28], water softening [11], and many other applications [20,29–32].

In seawater desalination, the principle of the FO process in all various forms basically consists of subsequent systems which are FO membrane and regeneration stages as shown in Fig. 1 [8,11]. In FO stage, the DS withdraws freshwater from the FS across the membrane, while the regeneration stage is utilized to simultaneously concentrate the DS and produce freshwater. The concentrated DS produced from the regeneration stage is then recycled and reused as DS in FO stage. The regeneration stage utilizes either a thermal or membrane separation based process [11].

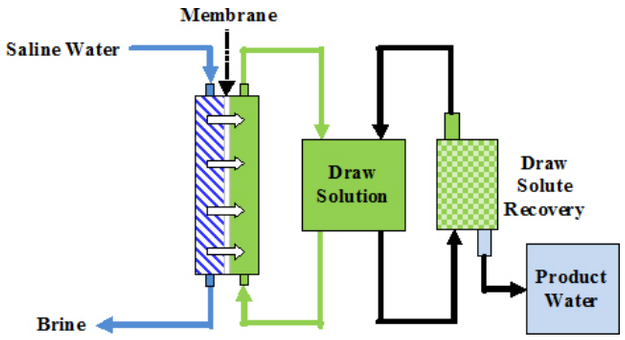


Fig. 1. Schematic drawing of the forward osmosis desalination process [8].

A great progress has been made in developing FO membrane and exploring DS [8]. Zhao et al. [33] reported the membrane developments for a number of applications including seawater desalination. According to Coday et al. [34], the commonly used FO membranes are cellulose triacetate (CTA) and polyamide thin-film composite (TFC) membranes. Ge et al. [1] reported that an ideal FO membrane must have the following characteristics: high water permeability, high rejection of solutes, substantial reduction in internal concentration polarization (ICP), and high chemical and mechanical stability.

Coday et al. [34] reported existing FO developer's worldwide and commercial status for each developer. Modern Water Company (MWC) and Trevi Systems Inc. (TSI) are actively engaged in applied research of innovative FO membrane desalination systems. According to Nicoll [11], MWC built a first FO-RO pilot plant with a capacity of 18 m^3/d, in Gibraltar in 2008. Also, MWC built two FO-RO-based desalination plants with capacities of 100 m^3/d and 200 m^3/d in Oman in 2009 and 2011, respectively. The outcomes of the aforementioned projects reported by Nicoll [11] were as follows: (1) in comparison with RO, the experimental investigation showed that FO has lower fouling potential. FO-RO system did not require chemical cleanings over a number years of operation; in contrast, RO system required frequent chemical cleaning over a duration of few weeks; (2) FO-RO system required less energy consumption than that of RO in a muddy seawater application whereas RO requires an intensive pre-treatment, frequent cleaning, and membrane replacement in a short period of operation; and (3) FO-RO has higher boron rejection in comparison with RO.

On the other hand, TSI has patented an innovative FO technology for desalinating seawater. TSI utilized thermal separation unit as a regeneration stage in their FO technology to recover DS and product water. This technology has not been investigated as thoroughly as other conventional desalination systems. According to the MWC and TSI [11,35], a number of severe limitations in their technologies needed to be investigated which are (1) investigating

non-toxic DSm (2) maintaining a constant high osmotic pressure for the concentrated DS, (3) maintaining high quality of the re-concentrated DS, (4) reducing reverse diffusion of the DS to the FS, (5) investigating continuous operating mode at larger scale applications, and (6) evaluation of techno-economic feasibility study.

Despite the aforementioned considerable advantages of FO technology, FO still has a number of severe limitations in being a cost-effective and sustainable process for seawater desalination and therefore the leading scientist are focusing on the applied research challenges: (1) seeking for the most effective semipermeable-membrane suitable for FO application, (2) discovering an effective and ideal DS, and (3) finding facile regeneration stage.

Although different FO membrane configurations and materials are available, however, these membranes have not been investigated and compared thoroughly in order to find out which of these membranes are the most feasible, in terms of water and salt flux, for saline water applications. Therefore, this study will focus on experimental study of one of the commercially available spiral wound FO membrane to extract freshwater from different salt concentration of saline waters. The investigation on the development of regeneration stage is beyond the scope of this paper, so this study will focus on FO stage only.

On the other hand, desalination technologies for concentrating highly saline brine are currently one of the major challenges of applied research. Unfortunately, well-established membrane and thermal desalination processes are either expensive or technically unfeasible. Zero liquid discharge (ZLD) systems are being considered for such an application. ZLD systems are usually consisting of thermal separation processes including brine concentrators, crystallizers, thermal evaporators, and spray driers. The integration of these processes are aimed at reducing the brine into to a compact solid product/waste that can be either disposed of in landfills or further treated in order to be used as useful products, which provide source of revenue through generating mineral salts as by-products. In addition, ZLD system is potentially capable of recovering high purity distillate at water recovery ratio between 95% and 99% [36,37]. Although ZLD is a powerful process and is proven effective technology for disposing the brine, the capital and operating costs often exceed the cost of the desalination facilities [38–41] and thus, ZLD is not typically utilized. Therefore, reducing the capital and operating costs of ZLD becomes one of the major goals of many leading scientific and commercial sectors around the world in order to make this technology a viable option. Martinetti et al. [36] reported that the costs of ZLD system can be drastically decreased by reducing the incoming volumes of brine and this can be achieved by incorporating FO membrane technology. Therefore, this study will not only cover the assessment of FO membrane for desalination applications but also it will be extended to cover the assessment of FO membrane for brine concentration by using actual RO brine as an example of undesired highly saline brines.

The main objective of this paper was to examine the viability of the FO membrane process step of the FO desalination process, on a laboratory bench-scale test unit using batch mode, for extracting freshwater from feed and simultaneously diluting DS. The study's specific objectives were as follows: (1) to appraise the viability and effectiveness of the commercially available spiral wound FO membrane made of CTA, on a laboratory scale level, for extracting freshwater from different sources of saline waters including Gulf seawater (GS) and RO brine as well as different concentration of NaCl aqueous solution; and (2) to investigate the effect of different concentrations and temperatures of DS and FS upon water flux and permeate water recovery ratio.

2. Materials and methods

2.1. Theory

As for the theory of FO process, the osmotic pressure (π) for solutions is given by Van't Hoff [42] which is the same for the pressure formula of an ideal gas as follows:

$$\pi = cRT \tag{1}$$

where c is the molar concentration of the solute, R is the gas constant (0.082 L.bar/deg.mol), and T is the temperature on the absolute temperature scale (Kelvin).

2.1.1. Water flux

The relationship between osmotic and hydraulic pressures and water flux can be described by Eq. (2) [17,8]:

$$W = A(\Delta\pi - \Delta P) \tag{2}$$

where W is the water flux, A is the hydraulic permeability coefficient of the membrane, $\Delta\pi$ is the difference in osmotic pressure on the two sides of the membrane, and a ΔP is the difference in hydrostatic pressure. For FO, ΔP is zero, while for RO, it has a very high value, depending on the salinity of the FS (Fig. 2).

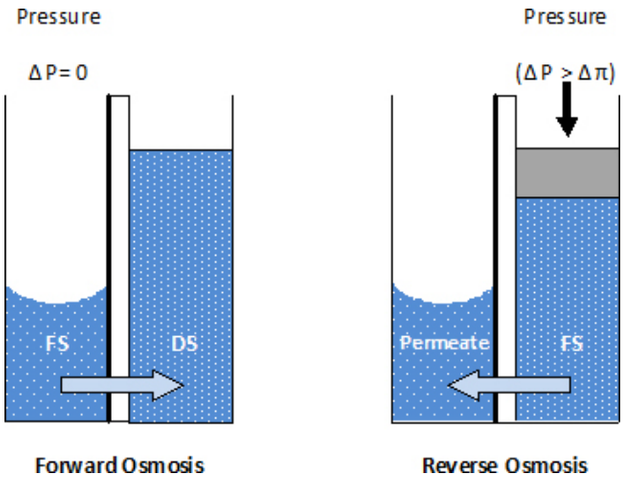


Fig. 2. Solvent flow in forward osmosis and reverse osmosis.

2.1.2. Salt permeability

Due to the concentration gradient across the membrane, a small amount of salt is transported across the membrane from the DS to the FS, which reduces the applied osmotic pressure across the membrane [17]. To calculate the salt permeability coefficient (B) of the FO membrane, Eq. (3) can be used [18]:

$$B = \frac{A(1-R)(\Delta P - \Delta\pi)}{R} \tag{3}$$

where R is the salt rejection of the membrane, which is defined as follows:

$$R = \left[1 - \left(\frac{C_P}{C_F}\right)\right] \times 100 \tag{4}$$

where C_p is the salt concentration of the DS, and C_F is the salt concentration of the FS.

In the FO bench-scale test unit, water will permeate through the membrane from the FS side to the DS side. Hence, during each experimental run, the water flux will be calculated from the decrease in the weight of the FS over time [43]:

$$\text{Water Flux} = \frac{\Delta\text{Weight}}{\text{Water density} \times \text{membrane surface area} \times \Delta\text{time}} \tag{5}$$

Also, the water flux, L/m^2/h (LMH), can be determined by the expression as follows:

$$J = \frac{Vf_2 - Vf_1}{[t_2 - t_1]} \times \left(\frac{1}{A}\right) \tag{6}$$

where Vf_2 is volume of FS at time 2, Vf_1 is volume of FS at time 1, t_2 is time reading 2, t_1 is time reading 1, and A is membrane surface area, 0.5 m^2.

Similarly, the permeate volume can be determined from the decrease in the initial volume of the FS over time. Thus, the FO system recovery would be calculated using Eq. (7):

$$\text{Recovery} = \left(\frac{V_P}{V_F}\right) \times 100 \tag{7}$$

where V_P is permeate volume, and V_F is initial FS volume.

Determination of the salt rejection of the membrane will be calculated by measuring the chloride concentration in the DS during and after each experimental run. Thus, the salt rejection of the membrane would be calculated using Eq. (4).

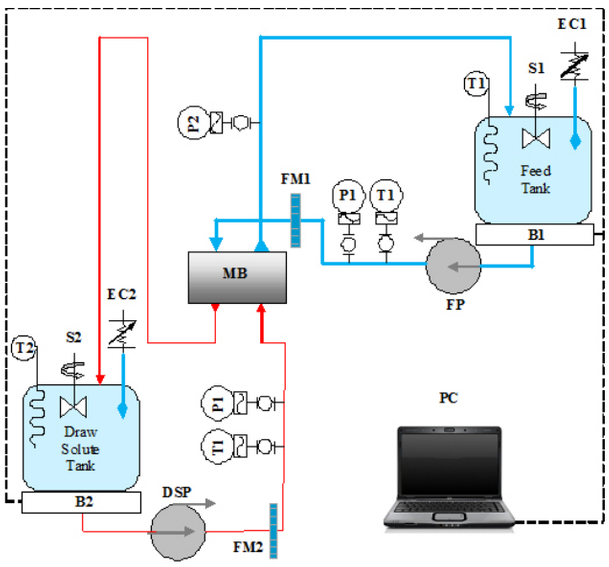
2.2. Experiment setup description

The experimental setup was prepared, constructed, and tested for investigating and verifying the performance of FO membrane process at different operating conditions for treating GS and RO brine, as well as the aqueous solutions of sodium chloride at different salt concentrations. Fig. 3 shows the schematic diagram of the main equipment used for the investigated FO membrane laboratory bench-scale test unit.

The experimental setup comprised of a membrane housing (MB), overhead stirrer assembly for FS (S1), overhead stirrer assembly for DS (S2), inlet pressure gauge indicator (P1), outlet pressure gauge indicator (P2), FS pump (FP), DS pump (DSP), digital recirculating bath for FS (T1), digital recirculating bath for DS (T2), portable conductivity meter for FS (EC1), portable conductivity meter for DS (EC2), flow gauge indicator for FS (FM1), flow gauge indicator for DS (FM2), weighing scale for FS (B1), weighing scale for DS (B2), personnel computer (PC).

The membrane housing vessel (AXEON, Model: 2521) is made of the Polyvinyl Chloride (PVC). FS was sent to the side ports and DS was sent to the end ports. The FS and DS cylindrical tanks (Tamco Model: 3001), with a capacity of 5 gall. The FS flow rate indicators (Blue-White, Model: F-45500L-8) is made of polysulfone and stainless steel wetted materials. The DS flow indicator (King Instrument's, Model: 7510-2-1-2A08) is made of acrylic and stainless steel wetted materials. The pressure gauges (Wika, Model: 233.53) is made of SS wetted materials. The FS pump (AMT, 300 series self-priming pumps) was used to circulate the FS. The DS pump (MP pumps, FRX, Model: 75-SP pup) was used to circulate the DS. The temperature gauges (Wika, Model: TI.50 series) were used to measure the temperature of the DS and FS.

A commercially available spiral wound FO membrane element (Hydration Technology Innovations (HTI), OsMem



MB: Membrane housing
S1: Overhead stirrer assembly for FS
S2: Overhead stirrer assembly for DS
P1: Inlet pressure gauge indicator
P2: Outlet pressure gauge indicator
FP: FS pump
DSP: DS pump
B1: Weighing scale for FS
T1: Digital recirculating bath for FS
T2: Digital recirculating bath for DS
EC1: Portable conductivity meter for FS
EC2: Portable conductivity meter for DS
FM1: Flow gauge indicator for FS
FM2: Flow gauge indicator for DS
PC: Personnel computer
B2: Weighing scale for DS

Fig. 3. Schematic diagram of a forward osmosis bench-scale test unit.

2521FO-MS-CTA-P-3H) made of CTA was investigated in this study. According to Tzahi et al. [44], HTI developer has fabricated unique and very effective FO membranes in comparison with other commercially available FO membranes. Consequently, this study has utilized and investigated the best commercially available spiral wound FO membrane element for research and development.

The recommended design parameters for the investigated FO membrane element are illustrated in Table 1, whereas the standard test conditions provided by the FO developer is presented in Table 2.

2.3. Preparation of feed solutions and draw solution samples

Four different sources of waters, namely, deionized (DI) water, aqueous solutions of sodium chloride (NaCl), GS, and RO brine, were tested individually as FS samples in this study. Barduhn [45] stated that prepared synthetic water using aqueous solutions of sodium chloride (NaCl) gives results very similar to process brines. Therefore, synthetic water using aqueous solutions of NaCl salts (Techno Pharmchem, sodium chloride AR – 33127 with 99.9% purity, analytical reagent grade NaCl) were prepared, used, and examined as FS and DS, in order to validate the potential capability of the FO membrane process for desalting a wide range of liquid streams. The initial salt concentrations of the FS streams, using synthetic waters (NaCl solutions), were ranging from 0.5 to 7 wt% by weight of NaCl salts. Different concentrations of NaCl solutions ranging from 3.5 to 26 wt% by weight of NaCl salt were used as DS. The osmotic pressure of the DS and FS were calculated using Van't Hoff equation and the results are tabulated in Table 3.

The FS samples using aqueous solutions of NaCl were prepared by dissolving a predetermined mass of NaCl salt into a known mass of DI water produced by ultra violet (UV) water purification system (Direct-Q3, Trade Name: Direct-Q).

In addition to this, two different sources of saline water were used and tested individually as FS samples in the laboratory. The examined saline waters were GS (4.9 wt% by weight of dissolved salt) and reject brine (5.6 wt% by weight of dissolved salt), produced from a RO membrane desalination plant. It is important to state that the pH values for all FS and DS were maintained at 7 and its influence was beyond the scope of this study.

Process water such as GS and RO brine were collected from the FS stream and reject brine discharge of the Kadhmah Bottled Water (KBW) commercial plant, respectively. The KBW plant represents one of the main research projects of the Water Research Center (WRC) of the Kuwait Institute for Scientific Research (KISR), located at the Doha Research Plant (DRP) in Kuwait. The KBW plant consists of two series of RO membrane units using the series product staging method. The FS of the first stage of the RO membrane units is GS. The FS samples from the FS stream of the first stage of RO membrane units (i.e., GS) and the dumped RO brines (i.e., the mixed RO brines of two stages were collected individually from the KBW plant and tested as FS in the laboratory investigations. The principles of RO membrane technologies, including the description of reject brines, are described elsewhere [46,47].

Table 1
Manufacturer recommended design parameters for the investigated FO membrane element

Parameter	Value
Membrane diameter, mm	61
Active area, m^2	0.5
Salt rejection, %	99
Water permeation, L/h	4.5
Membrane length, mm	483
Side-port diameter, mm	19
Maximum operating temperature, °C	45
Maximum operating pressure, bar	5
Minimum transmembrane pressure, bar	0.35
SDI	8
pH range	3–8
Maximum chlorine, ppm	2
Maximum NTU	10
Recommended prefiltration, µm	50
Channel height	Not Available
Operating velocity	Not Available
Operating duration	Not Available

Table 2
Standard test conditions

Parameter	Value
FS flow-rate, l/min	4
FS pressure	15 psi (1 bar)
FS temperature, °C	25
FS concentration, ppm	200 (tap water)
DS flow-rate, l/min	0.75
DS pressure	8 psi (0.55 bar)
DS temperature, °C	25
DS concentration, ppm	58,500 (NaCl solution)

Table 3
Osmotic pressure for NaCl solutions at different concentrations

NaCl, wt%	Osmotic pressure, MPa		
	15°C	25°C	40°C
0.5	0.41	0.43	0.45
3.5	2.97	3.08	3.23
7.0	6.17	6.38	6.70
15.0	14.46	14.96	15.72
26.0	28.79	29.79	31.29

2.4. Physicochemical analysis and measuring instruments

Physicochemical analysis was performed for all water samples of FS and DS before and after the completion of each test. The physiochemical analysis included the following key-parameters: the temperature, TDS, electrical conductivity, pH, volume, and mass. Two different types of

salinity measurements are considered in the physiochemical analysis in order to ensure and check the results of the salinity measurements; (1) electrical conductivity and (2) gravimetric method. The accuracy of salinity measurements (which are obtained by a gravimetric method) was also ensured by using a simple mass balance equation. In addition, full chemical analysis was performed for each test when the investigated FS was GS or RO brine. The purpose of conducting full water chemistry analysis was to detect the major components of ionic composition found in all water streams. A DR 5000 Spectrophotometer (Hach, DR 5000) and ion chromatography (Dionex 5000) systems were used to detect the major ionic composition of the aforementioned water samples. The reliability of full chemical analysis of the water samples was analytically ensured by a charge balance. Furthermore, the salinity, in terms of TDS, was also ensured by comparing the results of TDS obtained through physicochemical analysis and gravimetric method.

2.5. Experimental procedure

These experiments were performed in batch mode. The FS and DS reservoirs were filled with a 10 and 5 kg of FS and DS, respectively. For all experiments, the predetermined temperature of the FS and DS was controlled by means of a recirculating bath. When the predetermined operating temperature of FS and DS reached the desired level, the FS circulating pump was manually switched on, and then, simultaneously, the DS circulating pump is turned on to recirculate the FS and DS across a semi-permeable membrane. The FS and DS circulating pumps continuously transported FS and DS from the collecting reservoir to the membrane vessel as shown schematically in Fig. 3. The investigated parameters, namely, salt concentrations of FS and DS, temperature, pressure, and flow rate, were set at the predetermined levels dictated by an experimental envelop. Apart from the salt concentrations of the FS and DS, the predetermined values of the investigated parameters remained constant from starting-point of the test operation until the end of the experiment. By circulating the flow of FS and DS inside the membrane housing, the DS extracted the freshwater from the FS across the membrane surface. Consequently, the mass of the FS in the FS reservoir was gradually decreased; conversely, the mass of the DS level in the DS reservoir was gradually increased. At the same time, the FS concentration in the FS reservoir was gradually increased, and simultaneously, the DS concentration in the DS reservoir was gradually decreased. This is due to the freshwater molecules in the FS being transported to the DS side and mixed with the DS. All values of the investigated parameters were recorded every 5 min while performing the test. The duration of each test was 2 h. Upon completion of the experiment, the operation of the FS and DS pumps was terminated, and simultaneously, the water samples of the FS and DS were collected for laboratory analysis.

3. Results and discussions

For the purpose of simultaneously obtaining the TDS of the tested aqueous solutions of NaCl, the relationship between the electrical conductivity and salt concentration (measured in mg/l) was determined by experimentally measuring the salinity of NaCl solution over a wide range of electrical conductivity values, ranging from 0 to 252 mS/cm. The variations of the theoretical and experimental results of key parameters were plotted on graph as shown in Fig. 4. Based on the experimental results, the empirical polynomial correlations were derived and fitted for the TDS value (ppm) as a function of electrical conductivity (mS/cm). These equations were used to instantly calculate the theoretical results of the TDS parameter from the conductivity measurement.

3.1. FO Flux and water recovery

Figs. 5 and 6 show that a significant increase in water flux and permeate water recovery ratio can be achieved

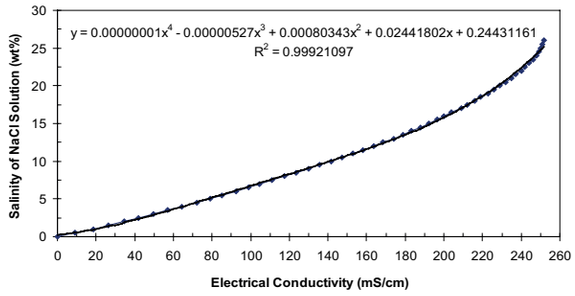


Fig. 4. Relationships between the salinity of NaCl solution and electrical conductivity for the NaCl solution.

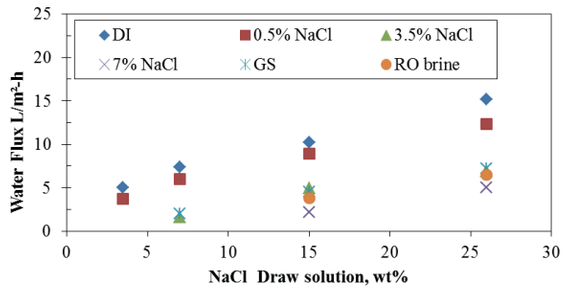


Fig. 5. Water flux obtained with different feed solutions and different concentrations of NaCl draw solution.

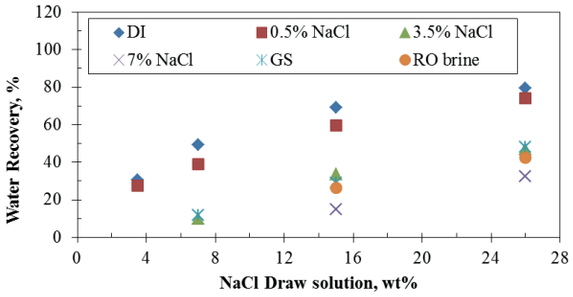


Fig. 6. Water recovery obtained with different feed solutions and different concentrations of NaCl draw solution.

by increasing the osmotic pressure difference (ΔΠ). The increase in the ΔΠ can be accomplished by either increasing the salt concentration of DS or decreasing the salt concentration of FS as shown in Figs. 5 and 6. A dramatic fall was observed in the rate of water permeate with a running time of the experiment and this trend was observed for all investigated FS concentrations as shown in Fig. 7.

The DS that diffused into the porous support layer got diluted during extracting freshwater from feed side, and, thus, the driving force, that is, ΔΠ, is gradually reduced over the period of the experiment. Table 4 summarized the experimental data for the water flux and permeate water recovery at different salt concentration of FS and DS. By maintaining a constant value of DS concentration (e.g., 26 wt.% of NaCl), the water flux and water recovery ratio was decreased by increasing the salt concentration of FS. ICP could also play main role in decreasing the water flux and water recovery ratio. Figs. 5 and 6 suggest that, by increasing the concentration of DS, several advantages can be rendered to the FO system, for instance, a significant increase in production rate, a dramatic fall in the volume of the residual liquid of FO brine.

As shown in Fig. 5, the difference in trend between the investigated DI water and aqueous solutions (including: NaCl solutions, GS, and RO brine) as FS shows that, when DI is used as FS, the water flux in FO system is influenced by only dilutive ICP (support layer DS); however, when aqueous solutions are used, the water flux is affected by both dilutive ICP and concentrative ECP (active layer facing FS). The results in Fig. 5 may show that, concentrative ECP plays an important role in lowering the water flux at higher DS concentrations. This trend observation has been demonstrated in earlier study conducted by Phuntsho et al. [48].

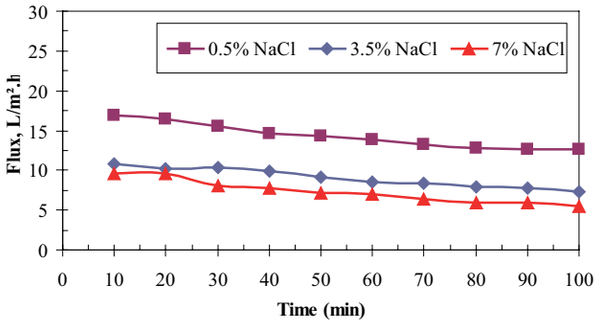


Fig. 7. Water flux over time for various NaCl feed concentrations using 26% NaCl draw solution concentration at 25°C.

Table 4
Water flux and recovery for different feed and draw solutions

Feed	NaCl draw solution, (Wt %)	Water flux at 25°C, l/m².h	Water recovery (%)
DI	3.5	5.0	30.6
(~0 wt %)	7.0	7.4	49.1
	15.0	10.2	69.0
	26.0	15.2	79.4
NaCl	3.5	3.7	27.5
(0.5 wt %)	7.0	6.0	38.8
	15.0	8.9	59.6
	26.0	12.3	74.3
NaCl	7.0	1.6	10.1
(3.5 wt %)	15.0	4.9	33.8
	26.0	6.9	47.1
NaCl	15.0	2.2	14.8
(7.0 wt %)	26.0	5.0	32.3
GS	7.0	2.0	11.7
(4.5 wt %)	15.0	4.5	29.2
	26.0	7.2	48.1
RO brine	15.0	3.8	26.4
(5.5 wt %)	26.0	6.5	42.5

3.2. FO flux profile

Fig. 7 shows the difference in trend between various salt concentrations of FS using 0.5, 3.5, and 7 wt% of NaCl. By increasing the salt concentration of FS, the water flux is reduced because of an increase in the osmotic pressure of the FS, which lead to drastically reduce the driving force (ΔΠ) for driving freshwater from FS into DS side in FO system as demonstrated and reported by Phuntsho et al. [48]. This explains the observed lower water flux for case of high concentrations of FS in Figs. 5 and 6.

For a fixed concentration (26% of NaCl) of the DS, the permeate volumes versus running time for different salt concentrations of FS are shown in Fig. 7. It was clearly observed that rate of accumulated permeate volumes decreased with respect to time. This can be attributed to a phenomenon known as ICP that progressively reduces the flux rate, which was also experienced and reported in other FO studies [49,50]. As freshwater is withdrawn into DS side, a reduction in osmotic pressure difference is expected because of an increase in concentration of FS and decrease in concentration of DS, which has been demonstrated in several earlier studies [48]. This gives a clear indication and explanation why the observed the trend of permeate volumes were decreased with respect to time.

3.3. Effect of FS and DS concentration

Fig. 8 shows the influence of the driving force (ΔΠ) on the water flux. A clear tendency of increasing water flux with rising ΔΠ can be observed. Also, Fig. 8 shows that the water flux was found to be inversely proportional to the salt concentration of FS. As explained previously, the increase in salt concentration of FS is of course undesirable because of increase in the osmotic pressure of the FS, and, thus, it will reduce the driving force (ΔΠ) for transporting water from FS into DS side in FO system. This observation has been demonstrated in several earlier studies [48–50].

It is obvious that the water flux was improved by increasing the salt concentration of DS. Increasing the salt concentration of DS leads to a significant increase in driving force (ΔΠ), and, thus, obtaining higher production rates

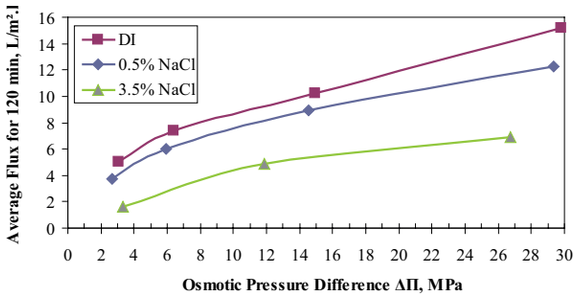


Fig. 8. Water flux over a range of osmotic pressure differences for various feed concentrations and 26% NaCl draw solution concentration at 25°C.

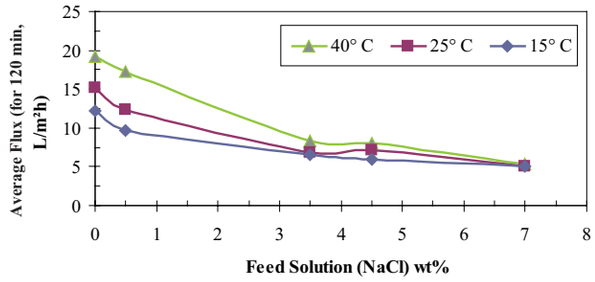


Fig. 9. Water flux various concentrations of NaCl feed at different temperatures using 26% NaCl solution.

from FO stage. The trend of the water flux must be linear with ΔΠ; however, Fig. 8 shows that it is nonlinear and this is due to the influence of the ECP and ICP. Previous studies have demonstrated that both ECP and ICP have significant impacts on the performance of the FO system, in terms of water flux, as they drastically reduce the effective osmotic pressure across the membrane [5,51–53].

As reported by Low [54], the osmotic pressure difference (Δπ) depends mainly on the concentrations of both the FS and DS. According to Eq. (2), with a given water permeability coefficient of the membrane (*A*), an increase in the FS concentration will reduce the water flux as observed in Fig. 8. This gives a clear indication and confirms that an increase in the feed concentration would cause a dramatic fall in the water flux due to reduction in ΔΠ.

3.4. Effect of temperature

Fig. 9 shows the effect of the operating temperature of FS and DS on the average water flux. Furthermore, it shows the influence of the FS concentration upon the water flux. Fig. 9 shows that the water flux can be dramatically improved by increasing the operating temperature of FS and DS. This may be related to the decrease in the solution viscosity achieved by increasing the operating temperature, which leads to a significant increase in the diffusion rate. Fig. 9 shows, in the case of experiments with high ΔΠ, the average water flux was significantly increased from 12.2 to 19.2 L/m².h, as the operating temperature was changed from 15°C to 40°C. The operating temperature reduces the severity of concentrative ECP on the water flux and ultimately enhanced permeate flux at higher DS concentrations as reported by Phuntsho et al. [48]. However, it can be observed that the operating temperature could not improve the water flux as ΔΠ decreased and this is due to an increase in FS concentration. Fig. 9 gives a clear indication that the operating temperature is ineffective for the cases of higher salt concentrations of FS. Low [54] also observed similar trend and reported that the change in water flux due to temperature was almost negligible for cases of low osmotic pressure difference because of high concentration of FS. According to Low [54], the poor flux improvement due to temperature rise for higher concentrations of FS could be due to the following effects: (1) the increase in the concentration of FS caused the osmotic de-swelling effect; and (2) higher concentration of FS escalated the internal CP.

Table 5
Water flux for different feed solutions at different temperatures using 26 wt.% by weight of dissolved NaCl salt as draw solution

Feed solution	Temperature (°C)	Average flux at 120 min, (l/m².h)	Gain in flux, (%)
DI	15	12.2	0
(~0 wt %)	25	15.2	25
	40	19.2	57
NaCl	15	9.6	0
(0.5 wt %)	25	12.3	28
	40	17.3	80
NaCl	15	6.6	0
(3.5 wt %)	25	6.9	5
	40	8.4	27
NaCl	15	4.5	0
(7.0 wt %)	25	5.0	11
	40	5.3	18
GS	15	6.0	0
(4.5 wt %)	25	7.2	20
	40	8.0	33

Table 5 shows the influences of salt concentration and temperature of FS and DS upon the water flux and gain percentage in the water flux. The investigated FS concentration was ranging from 0 ppm, using DI water, up to 7 wt% of NaCl salts. Also, Table 5 illustrated results of the experiments using GS as FS. It is important to note that the investigated operating temperature was ranged from 15 to 40°C, whereas the reported running time was 2 h.

3.5. Effect of flow rate

Fig. 10 shows the influence of the operating flow rate of FS and DS on the average water flux. The investigated flow-rate of FS and DS was varied from 2 to 4 l/min. The investigated FS were NaCl (at salt concentration of 0.5, 3.5, and 7.0 wt.% of NaCl) and GS, whereas investigated DS concentration was 26 wt.% of NaCl salt.

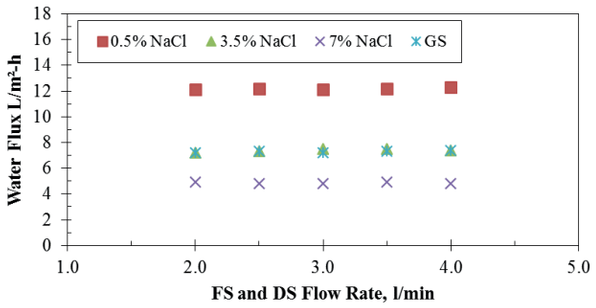


Fig. 10. Water flux obtained for different feed solutions at various flow rates using 26% NaCl draw solution.

For all investigated FS concentrations, the water flux was not changed with the investigated flow rates of FS and DS as shown in Fig. 10. This is due to the fact that the difference in flow rate (ΔQ) of FS and DS was kept constant as per manufacturer instruction to avoid membrane failure. It is therefore recommended to carry out further investigations on the flow rate of FS and DS at different ΔQ in order to provide a clear picture on the influence of the flow rate upon the water flux.

4. Conclusions

The influence of FS and DS concentration, flow-rate, and temperature on the performance of FO membrane stage has been studied utilizing the spiral wound CTA FO membrane module using DI water, NaCl solutions, GS, and RO brine as FS and NaCl as DS. The experimental results showed that the parameters such as: FS and DS concentration, ΔΠ, and FS and DS temperature had significant influence on the separation performance of the FO membrane process in terms of water recovery and permeate flux. The ICP had significant influence on the water flux since the ICP reduced the permeate flux of FO progressively. For the same value of the osmotic pressure difference, the effect of concentrative ICP was greater for the cases of higher concentrations of FS. The permeate flux was improved when the FS and DS temperatures were increased; however, the effect of temperature becomes insignificant for the cases of higher concentrations of FS.

This study proved that the investigated FO membrane element was potentially capable for extracting freshwater from different sources of saline waters including GS and RO brine as well as different concentrations of aqueous solutions. The experimental data obtained in this paper can provide important details which can be used as a reference for designing a pilot scale test unit for further research and development. However, further study is required to examine different types of commercially available FO membrane materials and membrane configurations as well as different types of DS in order to develop an effective and sustainable FO technology for seawater desalination applications. Detailed technical-economic analysis are recommended to be taken into consideration in future study to estimate the actual energy consumption of the investigated FO process and compare the figures obtained to the conventional desalination technologies such as MSF and RO.

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