Evaluation of Forward Osmosis as a Pretreatment Process for Multi Stage Flash Seawater Desalination

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Abstract

The present study evaluates the feasibility of applying forward osmosis (FO) process for the pretreatment of feed solution to a Multi Stage Flash (MSF) desalination plant. For the first time, real brine reject and real seawater were used as the draw solution and the feed solution, respectively in the FO process. The FO pretreatment is expected to dilute the brine reject and reduce the concentration of divalent ions, which are responsible for scale formation on the surface of heat exchanger in the MSF evaporator unit. The FO experiments were performed at different draw solution temperatures ranging between 25 – 40°C, different draw and feed solutions flowrates and different membrane orientations. A maximum average membrane flux of 22.3 $L/m^2.h$ was reported at a draw solution temperature of 40°C and 0.8 and 2.0 LPM flow rate of draw and feed solutions, respectively. The experimental results also revealed the process sensitivity to the feed solution temperature. It was found that the average membrane flux in the FO process operating at 0.8 and 2 LPM draw and feed solution flow rates, respectively was 16.9 $L/m^2.h$ at 25°C brine temperature but increased to 22.3 $L/m^2.h$ at 40°C brine temperature. These membrane fluxes resulted in 3% and 8.5% dilution of the draw solution at 25°C and 40°C temperatures, respectively. The average membrane flux in the FO mode was equal to that in the PRO mode at low flow rates but it was lower than that in the PRO mode at high flow rates of the feed and draw solutions. The outcomes of the study are very promising with regard to membrane flux and dilution of draw solution.

Keywords: Forward Osmosis (FO); Multi Stage flash (MSF); Pretreatment; Scaling; Membrane flux.

1. Introduction

Thermal desalination processes such as multi stage flash (MSF) and multi effect distillation (MED) demonstrated a high performance in a harsh environment without the requirements for seawater pretreatment. Conventionally, the top brine temperatures (TBT- the operating temperature in thermal desalination plant) in the MSF (operating in brine recycle mode) and MED are 65°C and 112 °C, respectively. These temperatures would achieve a recovery rate equal to 30% in both technologies [1-3]. However, these processes experience a major drawback represented by scale formation and deposition on the surface of heat transfer tubes. Scale formation reduces the heat transfer efficiency of the heat exchangers and adversely impact the performance of the thermal plant [1–4]. Alkaline scales, mainly $CaCO_3$, were reported in MED plants while non-alkaline scales, such as $MgSO_4$ and $CaSO_4$, were the main scale formations in the MSF plants [3]. The main strategies for scale minimization and removal in thermal plants are the use of antiscalants and periodic cleaning. Technically, these strategies are not effective in preventing scale deposition that builds up over time [2,5–8]. Alternatively, unconventional processes such as pretreatment using Nanofiltration (NF) membrane was proposed for the removal of scale ions from the feed solution to thermal plants [9,10]. Integrating NF with thermal desalination demonstrated a high efficiency in the removal of divalent ions, such as Ca^{2+}, Mg^{2+} and SO_4^{-2}, from seawater and allowed thermal plants to operate at elevated top brine temperatures [11]. An experimental work by Hasan et al. [10] demonstrated the advantage of NF pretreatment of seawater for the removal of scale ions and feasibility to increase the TBT of the MSF plant to 130°C. Operating at a TBT equal to130°C increased the recovery rate in the MSF plant and produced a gained output ratio (GOR) of 13. Despite the successful application of NF process for pretreatment of seawater, economically the process was unfeasible due to the high-energy cost [12–14]. The specific power consumption in the NF process was 1 kWh/m^3 when operating at a 65% recovery rate which is considered relatively high for an economic desalination process [15,16].

A previous study investigated the feasibility of FO pretreatment of feed solution to the MSF plant [17]. Brine reject was used as the draw solution while seawater was the feed solution. The study revealed the potential of the FO process to reduce the concentration

of divalent ions in the brine reject [17]. At a 32% recovery rate, 62% reduction in the concentration of Ca^{2+}, Mg^{2+} and SO_4^{2-} ions was achieved by the FO process. Moreover, the feasibility of the FO process combined to MSF plant is approved to reduce the concentration of divalent ions in the feed solution at 130 °C [15]. The estimated water flux for 45 g/L seawater salinity was in the range of 4 $L/m^2.h$ to 9.6 L/m^2h depending on the recovery rate of the FO process. For 45 g/L seawater salinity and 16% FO recovery rate, the concentration of divalent ions in the draw solution decreased by 13% after the FO treatment. The corresponding value for 32% FO recovery rate was 23.5% reduction in the concentration of divalent ions [15]. Unlike the NF process, the driving force in the FO process is the osmotic pressure gradient across the FO membrane instead of the hydraulic pressure [17,18]. Therefore, the cost of such pretreatment process is expected to be lower than that of the NF process.

Previous FO studies on the pretreatment of feed solution to the MSF plant were theoretical and there is no experimental data, yet to proof the concept. Experimental data, therefore, are essential to demonstrate the feasibility of the FO-MSF system for seawater desalination. In this study, we experimentally investigated the feasibility of FO process for the treatment of the feed solution to the MSF plant. Real MSF brine reject and seawater, the draw and feed solution, respectively, were collected from a thermal desalination plant in Doha City, Qatar. The study investigated the impact of different operating parameters on the performance of the FO process. Concentration of divalent ions in the brine reject was measured before and after the FO process to determine the dilution ratio.

2. Materials and Methods
2.1. Draw and Feed solutions characteristics

In the FO process a brine reject (BR) collected from a MSF thermal desalination plant in Qatar was used as the draw solution (DS) while the feed solution (FS) was seawater collected from the west bay in Qatar. The characteristics and composition of the DS and

133   the FS in terms of cation and anion concentrations, total dissolved solids (TDS),

134   conductivity, pH and turbidity are illustrated in Table 1.

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137   **Table 1:** Characteristics of seawater and brine reject of MSF

	FS (SW)	DS (BR)	Method
Cl⁻(ppm)	22183.7	35377.9	APHA 4110 DETERMINATION OF ANIONS BY ION CHROMATOGRAPHY Equipment: Metrohm 850 Professional IC
Br⁻(ppm)	74.93	118.79	
SO₄²⁻(ppm)	3153.6	5041.5	
Na⁺(ppm)	12952.6	20993.4	APHA 3120 METALS BY PLASMA EMISSION SPECTROSCOPY Equipment: iCAP 6500-ICP-OES CID Spectrometer (Thermo Scientific)
K⁺(ppm)	458.7	739.7	
Ca²⁺(ppm)	485.8	725.4	
Mg²⁺(ppm)	1535.0	2504.8	
TDS (ppm)	43474	81492	APHA 2540 C. Total Dissolved Solids Dried at 180°C
Conductivity ($\mu s/cm$)	60750	93650	APHA 2510 B. Conductivity
pH	8.40	9.07	APHA 4500-H+ B. Electrometric Method
Turbidity (NTU)	1.45	0.35	APHA 2130 B. Nephelometric Method

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2.2. FO membrane

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141   In this study, a flat sheet Thin Film Composite (TFC) FO membrane made by HTI

142   (USA). According to the manufacturer the membrane can tolerate temperatures up to 40

143   °C and has a rejection rate for sodium chloride ions of 90%. The membrane was placed

144   inside the FO cell and washed for 30 minutes with distilled water for pre-conditioning

145 and removal of any impurities from the surface. A 1 mm Sepa CF high fouling spacer (8

146 x 3.5 cm) was always placed on the support side of the FO membrane.

2.3. FO Bench-scale unit

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149 A CF042D-FO cell provided by Sterlitech was used in this study. The exterior

150 dimensions of the cell are 12.7 x 10 x 8.3 cm (5 x 4 x 3.25 in) with a membrane active

151 area of 42 cm^2 (6.5 in^2). CF042D FO cell has a shape of a cube, made of acetal

152 copolymer and can withstand a maximum temperature and pressure of 82°C (180 °F) and

153 1000 psi (69 bar), respectively. The FO system has two stainless steel tanks of 9 L

154 capacity for the FS and DS supply provided by Sterlitech (USA). The flow rates of the FS

155 and the DS were measured using two panel mount flow meters F-550 from Blue-White

156 industries Ltd. These flow meters have a maximum reading ability up to 4 LPM (1 GPM)

157 and a minimum of 0.4 LPM (0.1 GPM). The flows of the FS and the DS were set in a

158 countercurrent flow mode. The FS and the DS were circulated in the system using two

159 Mount Gear Pumps with Console Drive 115 V PEEK Gears/PTFE seals provided by

160 Sterlitech Company. The maximum motor rotation per minute for the pump is 5000 rpm

161 where the pump can achieve a maximum flow rate of 3200 mL/min (3.2 LPM). The

162 pressure of the FS and the DS were measured using pressure gauges type PEM series

163 provided by Winters Company in the range of 0 – 3 bar (0 - 45 psi). A MX-CA11B

164 immersion circulator (PolyScience Co-USA) controlled the temperature of the DS during

165 the FO experiments. This controller measured the temperature of solutions which can

166 read up to 135 °C. The conductivity of the FS and the DS were measured by OAKTON

167 PCD650 multi meter. This device has the ability to measure the conductivity, salinity and

168 total dissolved solids (TDS) in the range of 0–500 mS, 0-800 ppt and 0–500 ppt,

169 respectively. Ohaus Ranger^{TM} bench-scale balance connected to a computer to record the

170 change in the FS weight. Water flux was calculated by measuring the change in weight of

171 the FS over time divided by the membrane area multiplied by water density according to

172 Equation (1):

173 $$J_w(LMH) = 1000x\frac{\Delta W}{A\,t\,\rho} = \frac{1000(W_f - W_i)}{A\,t\,\rho} \qquad\qquad \text{Equation (1)}$$

174   ΔW is the change in the FS weight (kg) before and after permeation time, A is the

175   membrane effective area (m^2), t is the time (h) and ρ is water density (kg/m^3). The initial

176   volume of the DS and the FS were 6 L and the FS and the DS were recycled back to the

177   same tanks. Figure 1 depicts a schematic diagram of the FO system used in this study.
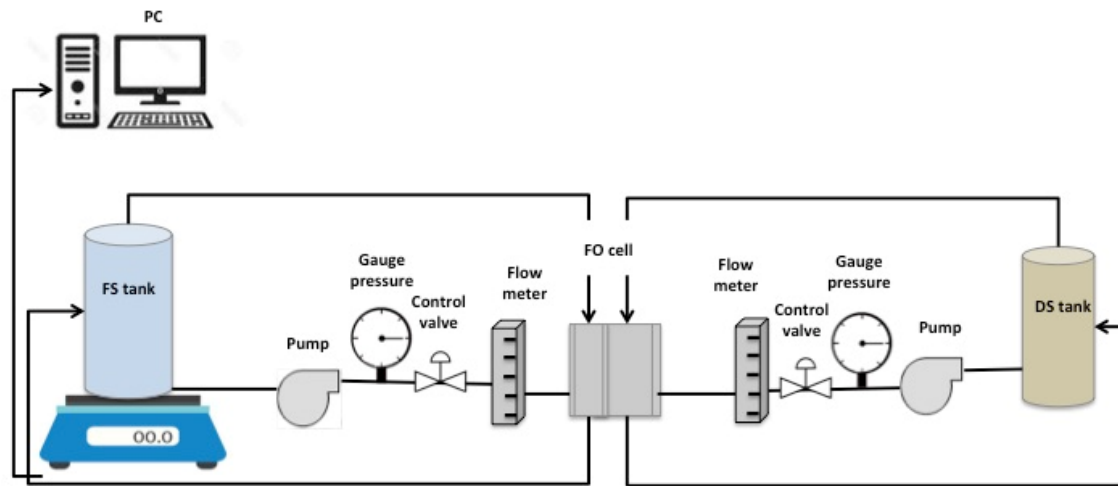
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Figure 1. A schematic diagram of the FO system used in this study

179   **2.4. Experimental procedure**

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181   The experiments were designed to operate for 150 minutes. Firstly, the impact of

182   changing the circulation flow rate of the FS and the DS was studied. Three different flow

183   rates were studied, 0.8:0.8 LPM, 2.0:2.0 LPM, 0.8:2.0 LPM for the draw solution and the

184   feed solution, respectively. Secondly, the influence of increasing the temperature of the

185   DS from 25 °C to 40 °C by a 5 °C increment was studied. Finally, the impact of the

186   membrane orientation was studied where firstlty the membrane active layer (AL) faced

187   the DS (AL-DS), PRO mode and then the membrane AL faced the FS (AL-FS), FO

188   mode. After each run and before the next run the system was washed with distilled water

189   for 30 minutes.

190   **3. Results and discussion**

191   **3.1. Impact of Feed Solution (FS) and Draw Solution (DS) flow rates**

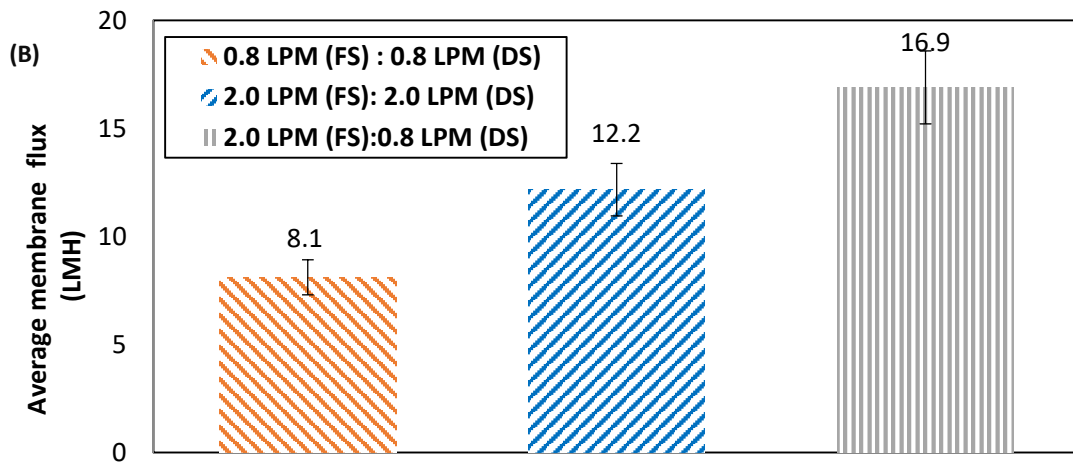192   **3.1.1 Membrane flux**

193   The impact of increasing the FS flow rate and the DS flow rate on the membrane flux

194   was evaluated at room temperature (i.e. 25ºC). The experiments were conducted at a flow

195   rate of 0.8 LPM for the FS and the DS then the flow rates were elevated to 2 LPM for

196   both the FS and the DS. Finally, the membrane flux was evaluated at a FS and DS flow

197   rate of 2.0 and 0.8 LPM, respectively. Each experiment lasted for 150 minutes and the

198   membrane active layer was facing the DS (i.e. PRO mode). Figure 2 (A) shows the

199   change of membrane flux with time at the different flow rates of the FS and the DS.

200   Figure 2 (A) shows the three consecutive runs where the membrane was washed for 30

201   minutes with distilled water between each run. Figure 2 (B) illustrates the average

202   membrane flux in the FO process at the different flow rates of the FS and the DS. In

203   general, membrane flux decreased over time due to the dilution and concentration of

204   draw and feed solution, respectively, that resulted in reducing the osmotic pressure

205   driving force across the FO membrane (Figure 2(A)).

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(A)

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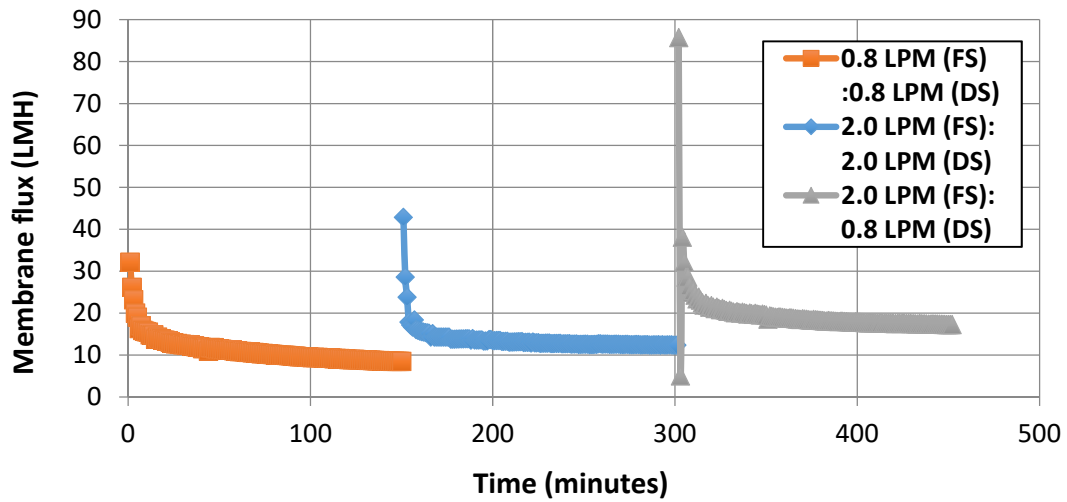
Figure 2. (**A**) Change of membrane flux with time at different FS and DS flow rates (**B**) The average membrane flux at different FS and DS flow rates at room temperature (25°C)

As shown in Figure 2 (A) the membrane flux increased at higher flow rates of the FS and the DS. This is due to increasing the rate of turbulence inside the module which reduces the concentration polarization. The average membrane flux at 0.8 LPM flow rate of FS and DS was 8.1 $L/m^2.h$, which increased to 12.2 $L/m^2.h$ at 2.0 LPM flow rate of FS and DS (Figure 2B). This suggests that the high flow rates of FS and DS reduced the effect of concentration polarization and increased the membrane flux [20]. Interestingly, the highest membrane flux was 16.9 $L/m^2.h$ at a flow rate of 2.0 LPM for the FS and 0.8 LPM for the DS. The reason for the high average membrane flux at these flow rates is due to the development of a small positive hydraulic pressure across the FO membrane in

220 the same direction of the draw solution. A value of 0.15 bar hydraulic pressure was

221 measured on the feed solution side. This hydraulic pressure was created by the higher

222 flow rate of the FS, 2.0 LPM, compared to FS, 0.8 LPM, inducing an additional

223 permeation flow across the membrane. Increasing the flow rate of the FS and the DS

224 from 0.8 LPM: 0.8 LPM to 2.0 LPM: 2.0 LPM resulted in 33.6% increase in the average

225 membrane flux while over 52% increase in the average membrane flux was achieved at

226 2.0:0.8 LPM flow rate of the FS and the DS, respectively. It is apparent that operating the

227 FO system at higher flow rates does not necessarily result in the most desirable

228 membrane flux despite its advantage in reducing the effect of concentration polarization.

229 Operating the FO system at different feeds flow rates with the flow rate of the FS being

230 higher than that of the DS would result in a better membrane flux in the FO process.

231 The drop in the membrane flux over time can be attributed to either the reduction in the

232 osmotic pressure driving force due to the dilution and concentration of the draw and the

233 feed solution as mentioned earlier or it could be due to the fouling of the membrane.

234 Figure 3 shows the Scanning Electron Microscope (SEM) images for the active and

235 support layer of the FO membrane before and after the FO experiments. Figure 3 (A) and

236 Figure 3 (B) show the clean (i.e. before use) active layer and support layer of the

237 membrane, respectively. Figure 3 (C) and Figure 3 (D) show the membrane active layer

238 and support layer, respectively, after the experimental run and before washing the

239 membrane with distilled water. Figure 3 (E) and Figure 3 (F) show the membrane active

240 layer and support layer, respectively, after washing the membrane with distilled water for

241 30 minutes. It is clear from Figure 3 (C) and Figure 3(D) that salts accumulated on both

242 sides of the membrane forming a fouling layer on the membrane surface. However, it

243 should be noted that the same membrane was used for three consecutive runs. The

244 membrane was washed after every experiment for 30 minutes with distilled water. In fact,

245 all tests were repeated three times and the membrane flux results were very close in all

246 runs after membrane washing. This indicates that the fouling materials were washed out

247 during the washing process. Figure 3 (E) and Figure 3 (F) show the membrane active

248 layer and support layer were clean after washing with distilled water for 30 minutes.

249 Generally, fouling of the FO membrane seems to be reversible by membrane cleaning

250 with distilled water. This finding is in an agreement with previous work carried out by

251 Modern Water in Oman [9]

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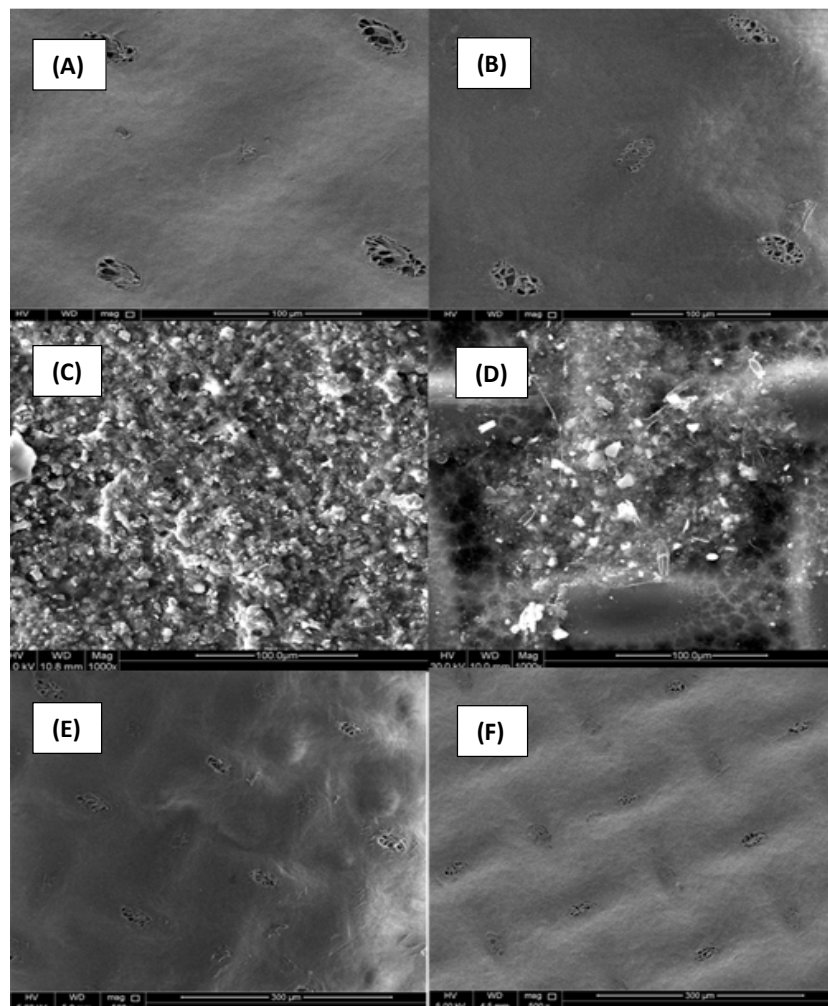
Figure 3. Fouling on the FO membrane surface after permeation during PRO mode **(A)** clean active layer, **(B)** clean support layer, **(C)** active layer after use, **(D)** support layer after use, **(E)** active layer after washing with distilled water for 30 minutes, **(F)** support layer after washing with distilled water for 30 minutes.

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255 **3.1.2 Recovery rate**

256 Figure 4 shows the achieved recovery rates at the studied flow rates after 150 minutes.

257 The recovery rate percentage is calculated using Equation (2) [39]:

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259 $$\%R = \frac{\frac{\Delta V_p}{\Delta t}}{\frac{\Delta V_F}{\Delta t}} 100 \qquad\qquad\qquad\qquad\qquad \text{Equation (2)}$$

260 ΔV_p and ΔV_F are the change in volume of permeate and feed solution (L), respectively,

261 and Δt is time of the experiment (h). Equation 2 can be described in terms of V_p and V_F

262 since Δt is equal for both permeate and feed solution:

263

264 $$\%R = \frac{V_p}{V_F} \times 100\% \qquad\qquad\qquad\qquad\qquad \text{Equation (3)}$$

265 V_p is the permeate volume at the end of each run and V_F is the initial volume of the feed

266 solution. The maximum recovery rate at 0.8:0.8 LPM flow rate of the feed and the draw

267 solution was 1.4%, increased to 2.1% at 2.0:2.0 LPM flow rate of the feed and the draw

268 solution. The maximum recovery rate of 3% was achieved at 2.0 LPM: 0.8 LPM flow

269 rate of the feed and the draw solution, respectively. As explained earlier the higher

270 recovery rate at 2.0 LPM: 0.8 LPM was due to the development of a positive hydraulic

271 pressure in the same direction of the draw solution, which promoted further permeation

272 flow across the membrane.

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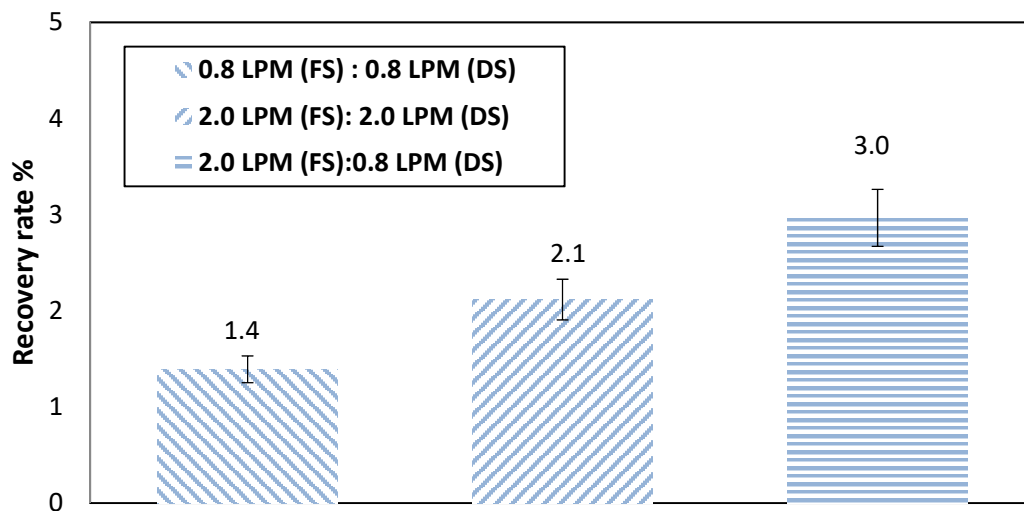
274 **Figure 4**. Impact of increasing FS and DS flow rates on recovery rate percentage at
275 isothermal conditions (25°C).

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3.1.3 Dilution of draw solution

The concentration of Ca^{2+}, Mg^{2+} and SO_4^{2-} ions were measured in the DS before and after the experimental runs. As mentioned earlier, these ions are responsible for scale formation and deposition in the MSF desalination plant. As shown in Table 2 at 0.8 LPM: 0.8 LPM flow rate of the feed solution and the draw solution and at room temperature, the concentration of SO_4^{2-}, Mg^{2+}, and Ca^{2+} ions in the draw solution was 5203.9, 2539.8 and 733.6 ppm, respectively. After the FO experimental run these concentrations became 5163.3, 2519.9 and 714.1 ppm for SO_4^{2-}, Mg^{2+} and Ca^{2+} ions, respectively. It was noticed that the draw solution was diluted by 0.78%, 0.78% and 1.1% for SO_4^{2-}, Mg^{2+}, and Ca^{2+} ions, respectively. The corresponding dilution in ions concentration at 2.0:2.0 LPM flow rate of the feed solution and the draw solution was 2.6%, 1.3% and 1.7% for SO_4^{2-}, Mg^{2+} and Ca^{2+} ions, respectively. Where the concentration of SO_4^{2-}, Mg^{2+} and Ca^{2+} ions in the draw solution was 5052.9, 2493.7 and 729 ppm, respectively. After the FO experimental run these concentrations became 4923.4, 2460.9 and 716.8 ppm for SO_4^{2-}, Mg^{2+} and Ca^{2+} ions, respectively. A further dilution of 2.9%, 1.8% and 2.2% for SO_4^{2-}, Mg^{2+} and Ca^{2+} ions, respectively was achieved at 2.0:0.8 LPM flow rate of the feed solution and the draw solution. Where the concentration of SO_4^{2-}, Mg^{2+} and Ca^{2+} ions in the draw solution was 5712.2, 2601.6 and 726.2 ppm, respectively. After the FO experimental run these concentrations became 5547.6, 2555.3 and 715.5 ppm for SO_4^{2-}, Mg^{2+} and Ca^{2+} ions, respectively. It was noticed that the SO_4^{2-} and Ca^{2+} ions exhibited higher dilution percentage than the Mg^{2+} ions; this was attributed to several factors such as the FO membrane rejection to ions and to the ions diffusion coefficient. Generally, SO_4^{2-} is highly rejected by FO membranes because of its negative charge and its large molecular weight [10, 18]. FO membrane rejection to Ca^{2+} was higher than Mg^{2+} because of the larger molecular weight of calcium ions compared to magnesium ions. It should be also noted that SO_4^{2-} has a diffusion coefficient of $1.07*10^{-9}$ m^2/s, which is higher than that of Ca^{2+} (i.e. $0.793*10^{-9}$ m^2/s) and Mg^{+2} (i.e. $0.703*10^{-9}$ m^2/s). This will affect ions reverse salt diffusion across the membrane and hence concentration in solution. A high rejection of SO_4^{2-} ions by FO membranes is very important in the FO pretreatment of seawater to prevent scale formation in the MSF plant since $MgSO_4$ and $CaSO_4$ are the main non-alkaline scale formations. It should be mentioned here that the initial concentrations of

308 the studied ions slightly varied between each set of experiments due to the fact that fresh
309 samples were used for every set of experiments. New seawater samples and new brine
310 reject samples were used for every set of experiments.

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317 **Table 2:** Concentration of SO_4^{2-}, Mg^{2+}, and Ca^{2+} ions in the DS before and after the FO
318 experiments at different DS and FS flowrates and at room temperature.

| Ion (ppm)	Flow rates of FS and DS (LPM)	Initial ions concentration in DS (ppm)	Ions concentration in DS at the end of the run (ppm)	Reduction (ppm)	Reduction (%)
SO_4^{2-}	0.8 LPM (FS): 0.8 LPM (DS)	5203.9	5163.3	40.6	0.78
	2.0 LPM (FS): 2.0 LPM (DS)	5052.9	4923.4	129.5	2.6
	2.0 LPM (FS): 0.8 LPM (DS)	5712.2	5547.6	164.6	2.9
Mg^{2+}	0.8 LPM (FS): 0.8 LPM (DS)	2539.8	2519.9	19.9	0.78
	2.0 LPM (FS): 2.0 LPM (DS)	2493.7	2460.9	32.8	1.3
	2.0 LPM (FS): 0.8 LPM (DS)	2601.6	2555.3	46.3	1.8
Ca^{2+}	0.8 LPM (FS): 0.8 LPM (DS)	733.6	714.1	8.06	1.1
	2.0 LPM (FS): 2.0 LPM (DS)	729	716.8	12.2	1.7
	2.0 LPM (FS): 0.8 LPM (DS)	726.2	715.5	16.1	2.2

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3.2. Impact of Draw Solution Temperature

Practically, the temperature of brine reject from an MSF desalination plant will be around 40°C. Therefore, we investigated the impact of the draw solution temperature on the performance of the FO process. Increasing the temperature of the draw solution was evaluated at different flow rates of the feed and the draw solutions. The experiments were conducted at draw solution temperatures of 25°C, 30°C, 35°C, and 40°C while the temperature of the feed solution remained constant at 25°C. Typically, the temperature of MSF brine is about 40°C but using different brine temperature would help to understand the impact of DS temperature on the performance of the FO process. The duration of each experiment was 150 minutes in which the membrane active layer was facing the DS (i.e. PRO mode). Figure 5 (A) presents the average membrane flux at the different draw solution temperatures and flow rates of feed and draw solutions. It was noticed that for any given flow rates, the average membrane flux increased with the increase in the temperature of the draw solution from 25°C to 40°C. At 0.8: 0.8 flow rate of the feed and the draw solution, the average membrane flux was 8.1 L/m^2.h at a DS temperature of 25°C which increased to 13.9 L/m^2.h at a DS temperature of 40°C, achieving 72% higher average membrane flux due to the temperature increase of the DS. The corresponding increase in the average membrane flux for 2.0 LPM: 2.0 LPM and 2.0 LPM: 0.8 LPM flow rates for the FS and the DS was 52% and 32%, respectively. The increase of the membrane flux at high draw solution temperature could be due to the lower water viscosity and greater diffusivity across the membrane at higher temperatures. Furthermore, according to Van't Hoff equation, the osmotic pressure increases with the increase of the draw solution temperature, therefore, the driving force in the process increases. Figure 5(A) also shows that the impact of feeds' flow rates on the average membrane flux remained unaffected with the variation in the temperature of the feed solution. The FO process at flow rate equal to 2.0 LPM: 0.8 LPM for the feed solution and the draw solution, respectively, exhibited the highest average membrane flux, which reached 22.3 L/m^2.h at 40°C. According to a previous study by Hawari et al. [21], the membrane flux in a FO process increased with increasing the draw solution temperature. The membrane flux reached a maximum value at a draw solution temperature around 26 °C before it started to decrease again with further temperature increase of the draw

351 solution up to 32°C. On the contrary, the membrane flux in this study kept increasing as

352 the temperature of the DS increased. Where at the highest draw solution temperature of

353 40°C the highest membrane flux was obtained. This could be due to the increase of the

354 flow rates of the feed solution and the draw solution that resulted in i) reducing the

355 impact of concentration polarization which could have overcame the thermal-osmosis

356 effect ii) developing a positive hydraulic pressure across the FO membrane that promoted

357 permeation flow. Where at flow rates 2.0 LPM for FS and 0.8 LPM of DS a 0.15 bar

358 hydraulic pressure was measured across the membrane. This can be clearly seen in Figure

359 5 (A) where at the flow rates of 0.8 LPM: 0.8 LPM for the FS and DS there was no

360 enhancement in the membrane flux when the temperature increased from 35°C to 40°C.

361 In effect, the effect of CP decreased at high flow rates of the feed and draw solution,

362 leading to a noticeable increase in the water flux from the feed to draw solution which

363 reduced the counter water flux due to the thermal-osmosis. It should be noted that the

364 obtained average membrane flux was found to be very promising and it is 2.3 to 5.5 times

365 more than the theoretically anticipated water flux in previous studies [22–26]. However,

366 water flux between 25 and 33 L/m2h was reported in FO experiments treating feed water

367 at 32 oC [20]. Figure 5B also show that, depending on the temperature of draw solution,

368 increasing the flow rate of feed solution from 0.8 LPM to 2.0 LPM resulted in 33 to

369 100% increase in water flux. On the other hand, increasing the temperature of draw

370 solution (using same flow rates) resulted in up to 71% increase in water flux. As such,

371 increasing flow rate of feed solution resulted in a larger improvement in the water flux.

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(A)

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(B)

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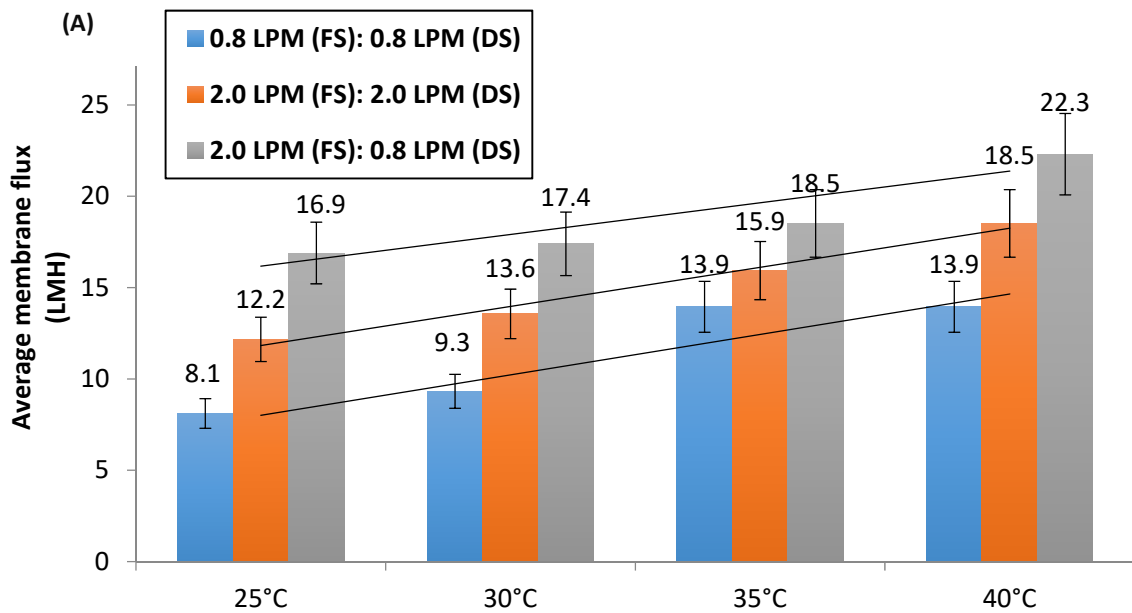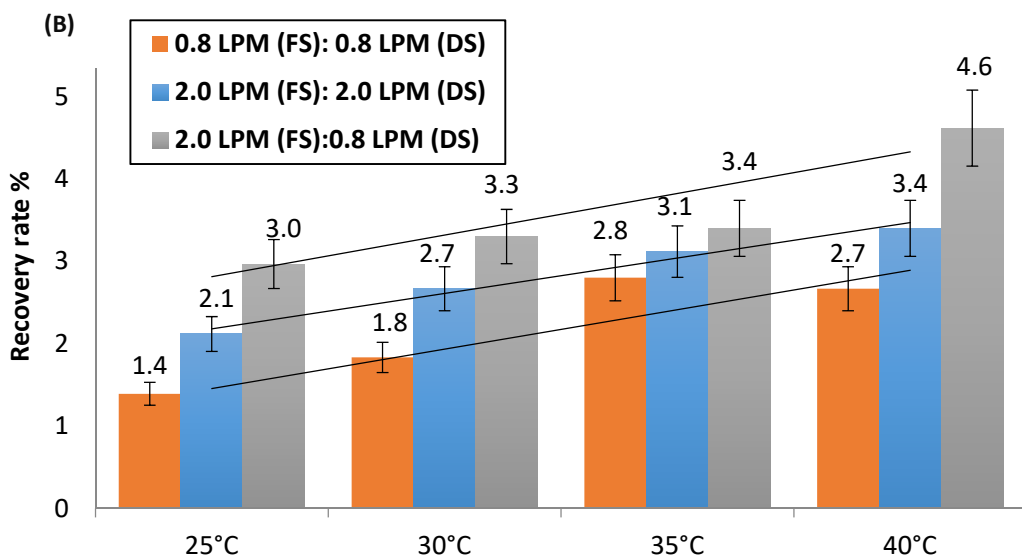
Figure 5. (A) Average membrane flux at different DS temperatures for the different
studied flow rates **(B)** Water recovery percentage at different DS temperature for the
different studied flow rates

Figure 5 (B) shows the recovery rate of the FO process at different flow rates of the draw
solution and the feed solution. The recovery rate of the FO process increased with
increasing the temperature of the draw solution from 25°C to 40°C and was always higher
at 2.0:0.8 LPM flow rate of the feed and the draw solution. At the 2.0:0.8 LPM flow rate,

the recovery rated increased by 35%, from 3% to 4.6%, due to the increase of DS

temperature from 25°C to 40°C. The reason for a greater membrane flux at a draw

solution temperature of 40°C is due to the higher diffusivity of water molecules at

elevated temperatures and higher osmotic pressure of draw solution. The FO process

resulted in 2.9%, 1.8% and 2.2% dilution of SO_4^{2-}, Mg^{2+}, and Ca^{2+} ions in the draw

solution, respectively at 25°C and 2.0:0.8 LPM flow rate. The dilution of SO_4^{2-}, Mg^{2+},

and Ca^{2+} ions in the draw solution increased to 7.7%, 4.9%, and 8.5%, respectively at

40°C and 2.0:0.8 LPM flow rate. As indicated before, SO_4^{2-} and Ca^{2+} ions exhibited

higher dilution by the FO than Mg^{2+}ion but most important that SO_4^{2-} ions are reduced in

the solution since it is the main component in non-alkaline scale formations in the MSF

plant.

3.3. Impact of membrane orientation

The membrane orientation is one of the factors that would affect the performance of the

FO process [20]. The membrane flux was evaluated in the PRO mode (DS-AL) and in the

FO mode (FS-AL). Figure 6 shows that the membrane flux was generally higher in the

PRO mode than in the FO mode. This is due to the lower effect of concentration

polarization in the PRO mode. This is in agreement with previous studies, which

demonstrated better controllable concentration polarization effects when the FO process

operates in the PRO mode [20,27–29]. Figure 6 shows that the average membrane flux in

the FO mode was equal to that in the PRO mode at low flow rates, i.e. 0.8:0.8 LPM, and

feed temperatures 30°C to 40°C. Whilst the average membrane flux in the PRO mode was

higher than that in the FO mode at 2.0:2.0 LPM and 2.0:0.8 LPM. This suggest that at

low feeds' flow rates, the FO process has no preference to operate on either mode, i.e. FO

or PRO, whereas at high feeds' flow rate it is desirable to operate the process at the PRO

mode to increase the membrane flux. In case of seawater contains high fouling materials,

it is suggested to operate the process in the FO mode and hence 0.8:0.8 LPM would be

more energy-efficient and easier to clean the membrane. However, 2.0:0.8 LPM FO

process operating at the PRO mode generated 1.6 times more water flux than 0.8:0.8

LPM FO process and should be considered when seawater quality does not promote

membrane fouling.
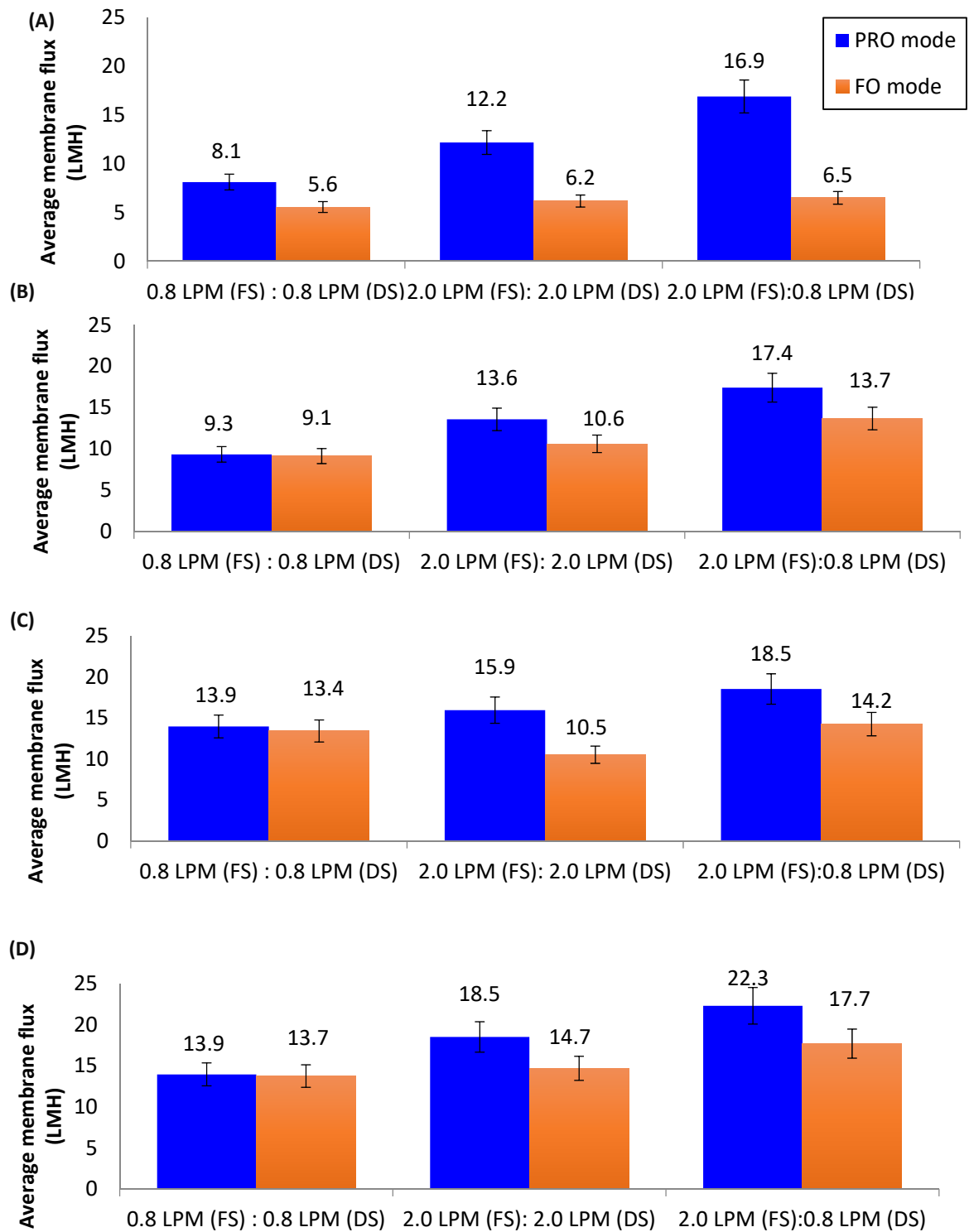
Figure 6. Impact of membrane orientation on average membrane flux (**A.** At DS temperature of 25ºC, **B.** At DS temperature of 30ºC, **C.** At DS temperature of 35ºC, **D.** At DS temperature of 40ºC)

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4. Conclusions

The feasibility of employing the forward osmosis (FO) process in the pretreatment of Multi Stage Flashing (MSF) desalination plant feed solution was evaluated in the present study. Brine reject from real MSF desalination plant and seawater were the draw and feed solutions. The impact of increasing the FS flow rate and the DS flow rate on the membrane flux was evaluated. It was found that the highest membrane flux was 16.9 $L/m^2.h$ at a flow rate of 2.0 LPM for the FS and 0.8 LPM for the DS. The maximum membrane flux was 22.3 $L/m^2.h$ at 40^oC draw solution temperature, this actually three times more than the theoretical flux expected in the previous studies that suggested FO process for treatment of feed solution to the MSF. The study showed that water flux was generally higher in the PRO mode than in the FO mode. In fact, this is not a problem in the FO process for seawater pretreatment for the MSF desalination plant since results demonstrated that the membrane flux can be fully recovered by washing the membrane with distilled water for 30 minutes. This indicates that the fouling was not severe and mainly reversible. In general, the study showed the promising application of FO process for pretreatment of seawater to MSF plant. A pilot plant test is planned to take place in the very near future.

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