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Dilution of seawater using dewatered construction water in a hybrid forward osmosis system

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ABSTRACT

In this study, dewatered construction water was used for the first time as the feed solution in a combined pretreatment-forward osmosis process to dilute seawater (i.e. draw solution) for further desalination. It was found that at a feed solution and a draw solution flow rate of 2.2 L min^{-1} gave the optimum membrane flux with minimal fouling effects. The addition of a spacer in the membrane feed side was effective at low flow rates (0.8 and 1.5 L min^{-1}). The feed solution was then pretreated using two methods: settling and multimedia filtration and used in the forward osmosis unit at a low flow rate of 0.8 L min^{-1} using a spacer at the feed side. Results revealed a significant increase in the forward osmosis membrane flux by 64.3% when multimedia filtration was carried out with a flux reduction of 7.7%. While the settling method achieved only 13.5% increase in the permeate flux and 12.5% flux reduction. The multimedia filtration process removed most of the particles that would cause fouling which resulted in an elevated and more consistent membrane flux. Results also showed that the water flux was 1.3 times higher when the membrane's active layer was facing the draw solution than when it was facing the feed solution. Cost analysis showed that forward osmosis treatment of dewatered construction water was 7.88 \$.day^{-1} and it was slightly cheaper when the forward osmosis operates in the pressure retarded osmosis mode.

1. Introduction

Dewatered construction water (DCW) is a by-product wastewater generated at construction sites. DCW has low salinity (i.e. conductivity of 3456 μS/cm), trace concentration of heavy metals and low total suspended solids (i.e. turbidity of 350 NTU) (Table 1). Therefore, DCW requires minimum treatment before discharge. Building and construction works in Qatar have resulted in the generation of large amounts of DCW which usually get treated on site before being discharged to the sea or injected into deep aquifers. Alternatively, DCW can be reclaimed for reuse on site or for general use such as irrigation at a lower cost as to the use of desalinated water because of its low salinity and pretreatment requirements (Angel et al., 2015; Powers et al., 2007). This study investigates the possibility of using DCW as a feed solution (FS) in forward osmosis (FO) to reduce the salinity of seawater which is used as the draw solution (DS) in the process. The seawater will get diluted before further desalination by reverse osmosis (RO) at a reduced cost compared to the conventional RO desalination process. The proper reuse of DCW is expected to reduce the adverse environmental impact of discharging such waters to the environment.

Forward osmosis is a new emerging osmotic process that involves a semipermeable membrane separating two solutions of different concentrations; the membrane permits water molecules to pass through but has high rejection to ionic species. This process causes the concentrated solution termed as the draw solution with high osmotic pressure to become more diluted and the less concentrated solution which is called feed solution with low osmotic pressure to become more concentrated (Shiqiang et al., 2017; Qasim et al., 2015). Unlike reverse osmosis (RO), FO does not require the application of hydraulic pressure; instead it only utilizes the osmotic pressure difference with minimum energy requirements and lower membrane fouling (Shiqiang et al., 2017; Zhao et al., 2012).

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Table 1
Chemical characteristics of the DCW used as the FS in the FO process.

Parameter (unit)	obtained value	Parameter (unit)	obtained value	Parameter (unit)	obtained value
pH	7.59	Nitrite-N (ppm)	0.182	Mg (ppm)	80
Turbidity (NTU)	350	Hardness (ppm)	1273	Mn (ppm)	<0.05
COD (ppm)	17	P (ppm)	1.22	Mo (ppm)	<0.05
Alkalinity (ppm)	149	As (ppm)	<0.05	Na (ppm)	297
$HCO3^-$(ppm)	181	B (ppm)	0.6718	Ni (ppm)	<0.05
SO4 (ppm)	1020	Ba (ppm)	0.0572	Pb (ppm)	<0.05
CL (ppm)	441	Ca (ppm)	378	Sb (ppm)	<0.05
Ammonia Nitrogen-N (ppm)	1.615	Co (ppm)	<0.05	Se (ppm)	<0.05
Nitrate-N (ppm)	3.1	Cr (ppm)	<0.05	Sr (ppm)	9.648
EC (μS/cm)	3456	Cu (ppm)	<0.05	V (ppm)	0.247
K (ppm)	26.4	Fe (ppm)	<0.05	Zn (ppm)	<0.05

Several research groups have implemented new concepts on the use of FO technology combined with other processes for the treatment of wastewater. For instance, Cornelissen et al. (2008) developed an osmotic membrane bioreactor (OMBR) for the recovery of wastewater which combines between activated sludge treatment and forward osmosis membrane separation. Instead of using the ultrafiltration or microfiltration membrane as in conventional membrane bioreactors (MBRs) the FO membrane was used directly in contact with the activated sludge. As such, high water fluxes were obtained using 0.5 M NaCl draw solution with low fouling propensity. A hybrid FO-RO process implemented by Cath et al. (2010) was used to transfer water from impaired water source such as wastewater with low salinity to seawater via osmotic gradient in order to dilute the seawater before further desalination with RO process (SWRO). The authors demonstrated that this approach lowers the energy use for the SWRO desalination, allows recycling of wastewater with recoveries of water up to 63%, reassures multi-barrier protection of drinking water and reduces RO membrane fouling due to reduced pollutant load in the diluted seawater. Thiruvenkatachari et al. (2016) applied integrated FO-RO system for coal mine wastewater treatment. FO process was able to recover 80% of the total volume of brackish mine wastewater and producing a dischargeable quality treated solution. Reversible FO fouling was reported by flashing the membrane with clean water to restore water flux. Hickenbottom et al. (2013) evaluated FO process for the treatment of drilling wastewater. Experimental results showed high rejection rate of FO to organic and inorganic matters and capability to achieve >80% recovery rate without membrane fouling. Impact of spacers on the rejection rate of trace antibiotic in wastewater was investigated by Liu et al. (2015). The rejection rate of antibiotic by FO increased when spacer was added and the draw solution was facing the membrane active layer. Researchers concluded that adding spacer promoted turbulence flow that improved the back diffusion of antibiotics from the support layer to the bulk solution. Boo et al. (2013) investigated fouling control in FO for wastewater reclamation using seawater or RO brine as draw solutions. Results revealed that support layer fouling by seawater or RO brine was insignificant while fouling of the membrane active layer occurred due to the accumulation of fouling matters. Fouling minimization by controlling hydrodynamics parameters such as increasing feed flow velocity, employing pulsed flow and using spacers effectively reduced the FO fouling.

In this study, dewatered construction water (DCW) is used for the first time as the feed solution in a combined pretreatment-FO process to dilute seawater for further desalination. The objective of this study is to determine the best pretreatment requirements of DCW for FO process and to study the effect of design and testing parameters such as flow rates of FS and DS, orientation of membrane and the placement of spacer on the FO process. Two pretreatment processes are performed namely, sedimentation and multimedia filtration. For the first time, the impact of membrane orientation on the cost of the FO process was estimated in this study. Furthermore, the study investigated the effect of design and testing parameters on the performance and cost of the FO process.

2. Materials and setup

2.1. Feed solution and draw solution characterization

Dewatered construction water samples were collected from a construction site in Doha City, State of Qatar. Table 1 shows a summary of the chemical characteristics of the DCW to be used as the feed solution in the FO process. The feed water conductivity is 3456 μS/cm compared to 54000 μS/cm for a standard seawater (35000 ppm). Sweater draw solution was made of 0.6 M Sodium Chloride solution. The chemical analysis were performed using Inductively Coupled Plasma-Optical Emission Spectrometry (ICP-OES) (ICP-OES Optima 5000 series) and Ion Chromatography (IC) (Metrohm 850).

2.2. FO unit setup

A schematic diagram for the experimental set up is given in Fig. 1. Sepa CF forward osmosis cell unit, supplied by SteritechTM Corporation, has been used in this study. The outer dimensions of the cell are $9 \times 12 \times 8.5$ cm. The cell is formed of two distinct compartments that are separated by an FO membrane. One compartment permits the flow of draw solution and other compartment is used for the feed solution. Both draw and feed solutions will flow in a counter current mode.

Two cup style feed tanks made of stainless steel with a 9 L capacity, also supplied by SteritechTM Corporation, were used for the DCW feed solution and the seawater draw solution. Moreover, two Cole-Parmer Micro-pumps A Mount Gear pump with Console Drive, PEEK Gears/PTFE seals were used to circulate and control the draw and the feed solutions flow. Two flow meters have been installed in the draw as well as the feed lines in order to measure the desired flow rates. A magnetic stirrer was used to ensure complete homogeneity in the DS and FS tanks.

Initially, the volume of the draw and the feed solutions were 6.0 L each. Solutions leaving the FO cell were recycled back to their respective tanks. The FO unit was operated for almost 1000 min for each experimental run. A new membrane was inserted in the FO cell after each run.

2.3. FO membrane

This study used a cellulose tri-acetate (CTA) forward osmosis membrane supplied by Hydration Technology Innovation (HTI). The

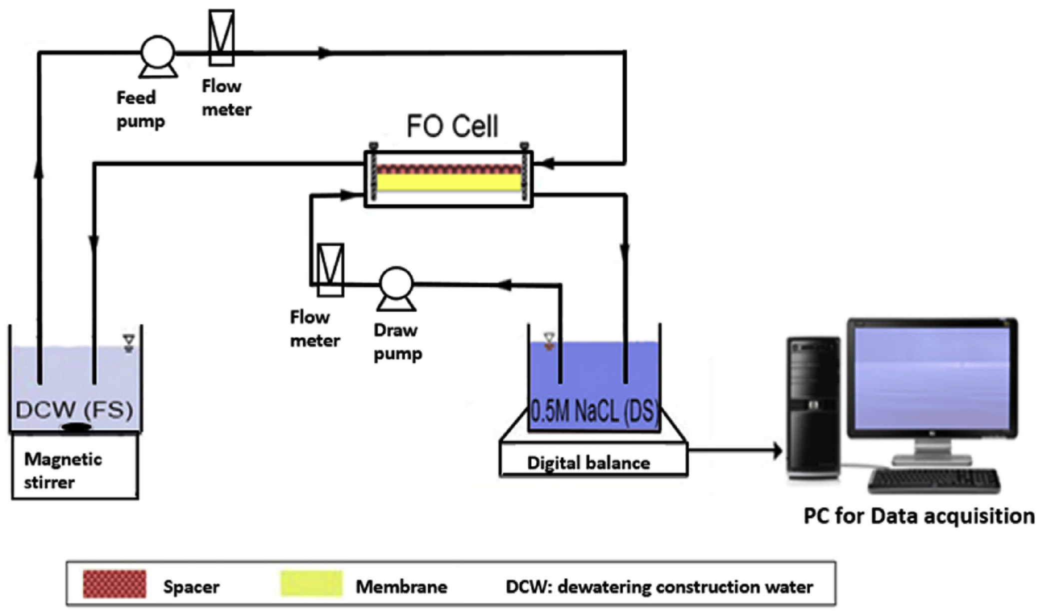

Fig. 1. Schematic diagram for the FO system used.

membrane was cut to fit the FO cell with dimensions of 11.5 × 5.75 cm and a thickness of 0.5 mm. At the beginning of each experiment the membrane was washed with distilled water for 30 min to remove any additives on the membrane active surface. To study the effect of the addition of spacer on process performance, a polymeric One Sepa CF high fouling mesh spacer, with dimensions of 8 × 3.5 cm and a thickness of 1 mm, was added in the feed solution side and in contact with the CTA membrane.

2.4. Experimental procedure

To study the impact of different parameters and experimental conditions on the FO process, the flow rate of feed and draw solutions were set to 0.8, 1.5, 2.2, 2.9 L min^{-1}. Each experiment was run for almost 1000 min on a recycle mode (i.e. feed and draw solutions were recycled in the FO unit until the end of the experiment). The impact of the existence of a spacer on the membrane's support layer was investigated at each flow rate. Three types of feed solutions were used: untreated DCW, settled DCW and multimedia filtered DCW. Then finally, different membrane orientations were studied; active layer facing draw solution (AL- DS) and active layer facing feed solution (AL-FS), using the multimedia filtered DCW. Quality tests for the DCW were performed prior to the experiments to determine the characteristics of the DCW in terms of turbidity and total suspended solids (TSS).

2.5. Analytical procedure

The conductivities of both draw and feed solutions were measured using the Agilent C5111 conductivity meter. The turbidity of the DCW was measured using 2100 P Turbidimeter supplied by Hach with nephelometric turbidity units (NTU). The TSS of the DCW was measured by filtering 50 ml sample using a 1.5 μm pore size TSS Glass Fiber Filter followed by gravimetric analysis after drying the filtrate at 105 °C for 2-h. TSS was then calculated in ppm.

To investigate the effectiveness of the FO process at different experimental conditions, the permeate flux was measured by recording the weight of the draw solution on a digital EW-11017-04 Ohaus Ranger™ Scale over time at a 15 min interval, the data is recorded on a computer with a data acquisition software. Each run was repeated at least twice and the data points were averaged to obtain the reported results. The error bars were calculated using the following equation:

$$\sqrt{\frac{1}{N}\sum_{1}^{N}(X_i-\overline{X})^2} \quad ; 2 \le N \le 3 \tag{1}$$

Where X is the measured data, $\overline{X}$ is the average value of the measured data and N is the number of replicates.

Microscopic observation for the membrane surface using Scanning Electron Microscope (SEM) was carried at different experimental conditions.

2.6. Filter setup

In order to enhance the quality of dewatering construction water, a multimedia filter was utilized to filter the sample before being placed as a feed solution in the FO unit. Fig. 2 shows the lab-scale multimedia filter manufactured by Atico in India. It consists of two tanks; one contains the DCW, and the other contains clean water to be used in the backwash process. The filtration media consists of 4 different layers with varying granular sizes; Activated carbon (anthracite) (0.8–1.6 mm), coarse sand (0.71–1.18 mm), fine sand (0.4–0.8 mm) and gravel. The height of the activated carbon is 10 cm, coarse sand is 25 cm, fine sand is 25 cm and gravel is 5 cm. The filter can operate in two modes, the normal run mode and the backwash mode. The flow meter controls the discharge going into the system with values up to 5 L min^{-1} the valves can control the water flow direction for the normal run mode and the backwash mode. A Kirloskar Wonder 3 domestic pump was used to pump water into the system. Before running the multimedia filter, the system was backwashed using tap water for 20 min with a 0.5 L min^{-1} flow rate. Then the turbid water valve was opened to permit a constant flow to the filter at a 0.5 L min^{-1} flow rate. After the filtration stage, water is collected from the effluent sampling port to be used in the FO channel.

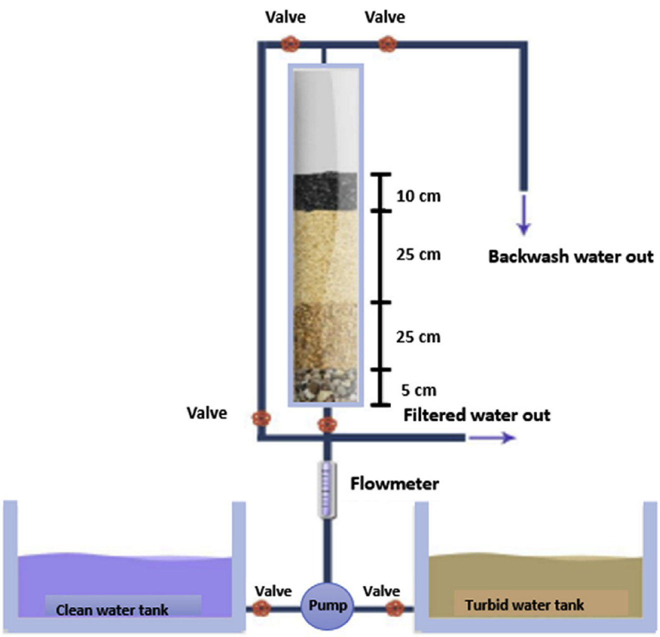

Fig. 2. Multimedia sand filter unit.

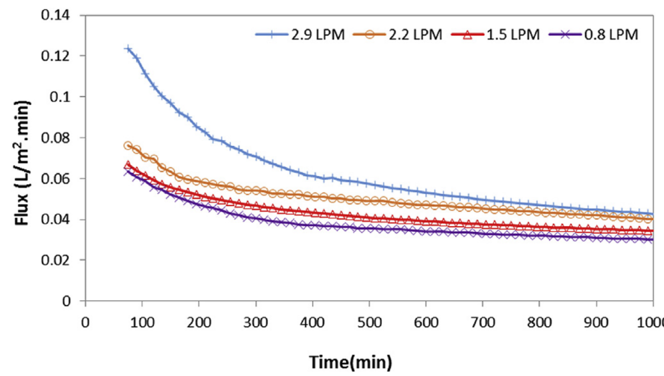

Fig. 3. Membrane Flux under different FS and DS flow rates, operation condition used are; 35,000 ppm seawater DS, DCW as FS.

3. Results and discussion

Several experimental runs were carried out based on different operational conditions by changing: the flow rate of FS and DS, adding and removing spacer, the type of feed solution treatment (without treatment, settling and multimedia filtration) and finally membrane orientation (AL-FS) and (AL-DS). The following section discusses the impact of changing these parameters on the membrane permeate flux.

3.1. Effect of feed solution and draw solution flow rate on membrane flux

Draw solution and feed solution flow rates effect on the membrane flux were evaluated using 35,000 ppm seawater draw solution and around 2000 ppm dewatering construction water as the feed solution. The membrane active layer was facing the draw solution (AL-DS) and no spacer was used. The flow rates were set to 0.8, 1.5, 2.2, 2.9 L min^{-1}. All experiments were performed at room temperature. The flow rate of FS was equal to the flow rate of DS at all times to eliminate any pressure effects on the membrane. Moreover, each experiment was run for almost 1000 min. The permeate fluxes across the membrane in the FO cell using different FS flow rates are summarized in Fig. 3.

The membrane flux is found to decrease over time and approach steady state towards the end of the runs. Moreover, the membrane flux decreased with the decrease in the flow rate of the feed and draw solution. From Fig. 3 the highest membrane flux was obtained at a FS flow rate of 2.9 L min^{-1} where the flux was initially 7.44 L m^{-2}.h^{-1} then dropped to 2.58 L m^{-2}.h^{-1} after 1000 min of operation. At the lowest flow rate of 0.8 L min^{-1} the membrane flux went down from being 3.84 L m^{-2}.h^{-1} at the beginning of the experiment to 1.8 L m^{-2}.h^{-1}. The reduction in permeate flux with the duration of the experiment and the influence of different flow rates on the membrane flux can be ascribed to the existence of two concentration polarization (CP) effects, namely, dilutive external concentration polarization (DECP), occurring at the active layer of the membrane facing the DS, and concentrative internal concentration polarization (CICP), occurring at the porous support layer of the membrane facing the FS. These effects, being investigated in previous studies (Gray et al., 2006; McCutcheon and Elimelech, 2006) are found to play a major role in reducing the osmotic pressure difference across the membrane and hence the driving force for the filtration process. Increasing the feed and the draw solution flow rates caused more turbulent flow that diminished the boundary layer thickness and increased the mass transfer coefficient, consequently, reducing the extent of CP (Devia et al., 2015; Jung et al., 2011).

Fig. 4 illustrates how the increase in the flow velocity causes an increase in concentration of the draw solution at the membrane active layer reducing the DECP effect. Simultaneously, the concentration of the FS at the porous support layer of the membrane has also decreased reducing the CICP effects, hereafter, increasing the net osmotic pressure difference ($\Delta\pi$) and the flux of the FO membrane.

From Fig. 5, it can be seen that the minimum flux reduction percentage was 47.3% obtained at a flow rate of 2.2 L min^{-1}. However, at a higher flow rate of 2.9 L min^{-1} the flux reduction was 67.7%. In this case, the high water flux at the beginning of the experiment accelerated the accumulation of fouling particles inside of the porous support layer and reduced the permeate flow (Tang et al., 2010). Experimental work showed that increasing water flux promotes higher membrane fouling in forward osmosis treatment of wastewater (Lutchmiah et al., 2014). Additionally, it caused a severe dilution effect on the draw solution that reduced the osmotic driving force and caused the water flux to decrease rapidly (Zhang et al., 2014). The results demonstrated that increasing the flow rate would not necessarily enhance the membrane performance and that a flow rate of 2.2 L min^{-1} was the optimum flow rate before severe colloidal fouling occurred.

3.2. Effect of placing spacer on the feed solution side

In order to decrease the occurred colloidal fouling a spacer has been placed on the membrane support layer and its impact on membrane flux was studied. Experimental runs were conducted using different flow rates for the same samples of FS and DS and results were compared with the FO runs without a spacer. The membrane active layer was facing the DS, while the membrane support layer was facing the FS. Results are summarized in Fig. 6.

The highest membrane flux for the FO process incorporating a spacer was found at a flow rate of 2.9 L min^{-1} with an initial value of 0.053 L m^{-2}.min^{-1} then decreased to 0.034 L m^{-2}.min^{-1} after 1000 min operational time. Same decreasing trends were obtained for the water fluxes at flow rates of 2.2, 1.5 and 0.8 L min^{-1}

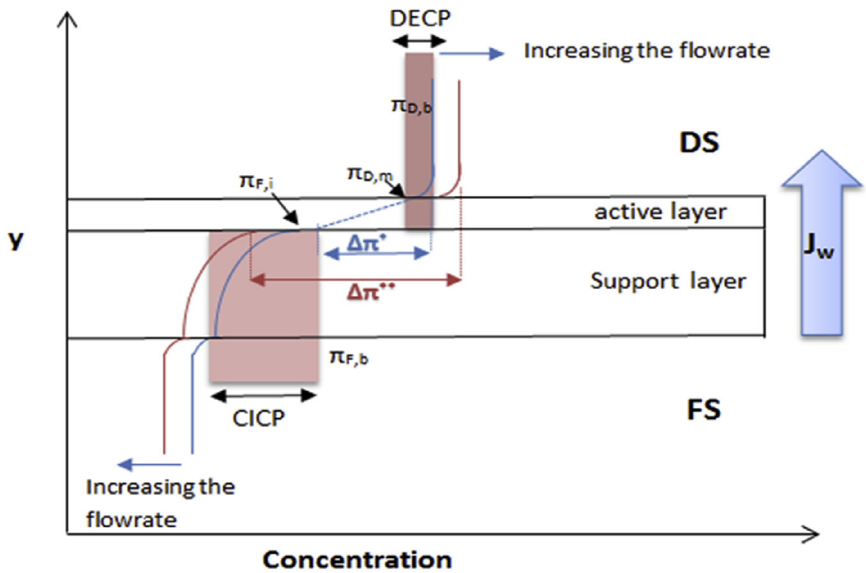

Fig. 4. Illustration of the coupled CICP and DECP effects on the net osmotic pressure difference ($\Delta\pi$) in the asymmetric membrane with the dense active layer against the draw solution. . $\pi_{D,b}$ is the bulk draw solution osmotic pressure and $\pi_{D,m}$ is the membrane surface osmotic pressure at the permeate side. $\pi_{F,b}$ is the bulk osmotic pressure of the feed and $\pi_{F,i}$ is the effective osmotic pressure of the feed. $\Delta\pi^*$ is the net osmotic pressure difference at low FS flow rate and $\Delta\pi^{**}$ is the net osmotic pressure difference at high FS flow rate. Note: it is assumed that no ECP occurs at the porous support layer.

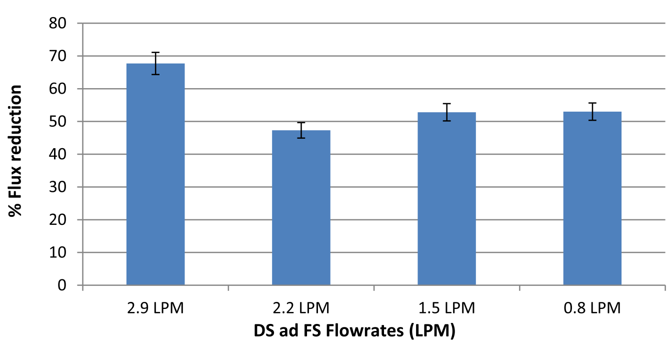

Fig. 5. Flux reduction percentage under different FS and DS Flow rate, with no spacer.

highlighting the impact of CP on the FO process. Initially, the higher water flux was in the following order $2.9 > 2.2 > 1.5 > 0.8\ \mathrm{L\,min^{-1}}$ flow rates. Increasing feed and draw solution flow rates resulted in

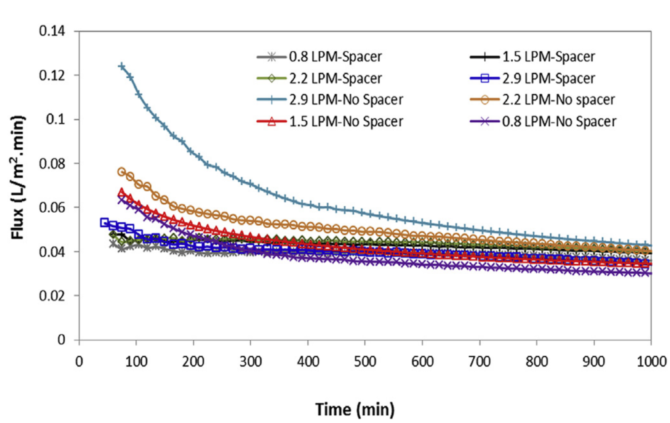

Fig. 6. Membrane flux under different FS and DS flow rates, operation condition used are; 35,000 ppm seawater DS, DCW as FS, with adding spacer in the feed solution side and without.

decreasing the effect of ECP and to less extent ICP as demonstrated in previous studies (Hawari et al., 2016) and hence improved water flux. The membrane flux for the $2.9\ \mathrm{L\,min^{-1}}$ flow rate, with and without spacers, decreased sharply in comparison to the other flow rates due to the high initial flux that encouraged the deposition of colloidal particles on the membrane and also due to the increased dilution effect on the draw solution side. More importantly, a significant difference in the initial permeate flux can be seen between the runs including a spacer and those not; with the highest water flux obtained in the FO experiments without the use of a spacer. This suggests that the spacer did not have a positive impact on the FO process; although it is known to promote turbulence of the inlet feed solution that may reduce the effect of concentration polarization, namely, CICP (Zhang et al., 2014). Yet, this process enhancement technique was negligible especially at high flow rates due to membrane fouling induced by conductive colloidal particles transport to the membrane porous media on the feed side and accumulation of solute, which penetrated by the reverse solute diffusion. These findings are in accordance with a previous study by Park and Kim (2013) in which they numerically scrutinized the impact of spacers on concentration polarization, and used a concentration polarization index (CPI), that is proportional to the extent of the CP and inversely proportional to the membrane permeate flux, to quantitatively measure the degree of CP. In their results, the authors established that the spacers near the membrane surface aggravate the CP effect due to the solutes, which penetrate by reverse diffusion, getting captured by the spacers adjacent to the membrane. The orientation of the membrane, AL-DS in the present case, further increases the ICP values, as the diffused solute tends to get hindered by the porous support layer.

Nevertheless, the percentage flux reduction was low for all flow rates when a spacer was used. This suggests that high initial water flux increases membrane fouling due to i) more intense CP at the feed and draw solution and ii) the conductive drag flow of fouling matters towards the membrane surface on the feed side. This phenomenon is significant in the FO tests without spacers because of the higher water flux, which caused sharp flux reduction. Initial water flux was higher in the FO tests without spacers than with

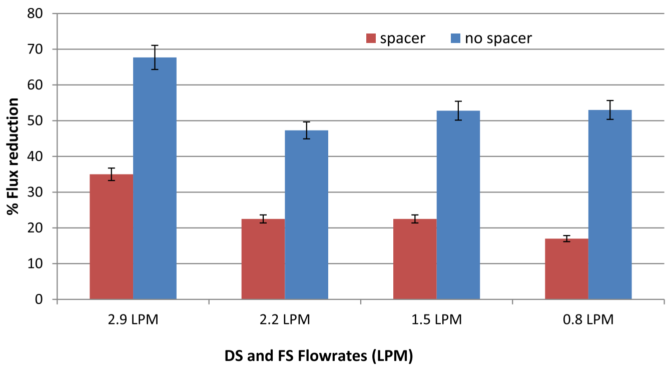

Fig. 7. Flux reduction percentage under different FS and DS flow rate, with and without spacer.

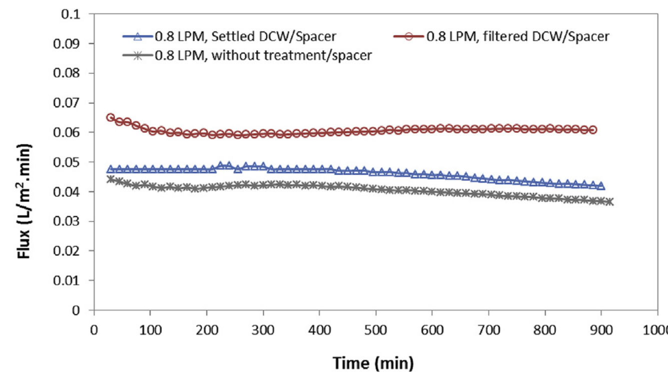

Fig. 9. Membrane flux for untreated, pre-settled and multimedia filtered DCW, operational conditions used are; 35,000 ppm seawater DS, DCW as FS, a spacer added in the feed solution side.

spacers, which induced more sever CP and membrane fouling. It should be mentioned that at a very low flow rate of 0.8 L min^{-1}, the addition of a spacer seemed to have improved the permeate flux and made it more consistent as the percentage reductions in permeate flux (17%), shown in Fig. 7, was much lower than without the use of a spacer (53%). This is due to the low membrane cross velocity resulting in low turbulence. Therefore, the addition of a spacer caused an enhancement in the inlet flow turbulence that urged the suspended solids to spread from the support layer into the bulk feed solution, consequently, elevating CICP effects and reducing the membrane fouling propensity (Zhang et al., 2014). Unfortunately, water flux at 0.8 L min^{-1} feed flow rate is much lower than that at 1.2–2.9 L min^{-1} feed flow rates and hence it is impractical. SEM images off the membranes in Fig. 8 further support these findings.

3.3. Effect of feed solution pretreatment on membrane flux: settling and multimedia filtration

In order to enhance the performance of the FO process two different pretreatment processes were evaluated; settling and multimedia filtration.

Initially, a settling tank was employed for the removal of suspended solids from the feed solution; this approach was selected because it is inexpensive and DCW is often stored in a storage tank after extraction. In settling tanks, the settleable solids tend to settle out at the bottom of the tank by gravitational sedimentation. The initial DCW turbidity was 300 NTU and TSS concentrations was 325 ppm, from which a turbidity value of 26 NTU and a TSS value of 45 ppm were attained after a 1-h settling time. After settling the DCW was collected and used in the FO process.

Similar to previous runs, the same draw solution and membrane orientation were used (AL-DS), the FS and DS flow rates were set to 0.8 L min^{-1}. A spacer was also used in the feed side of the membrane as it was found to enhance the FO process at low flow rates. All runs were performed at room temperature.

As shown in Fig. 9, pre-settling the FS had a slight positive effect on the permeate flux compared to the case without feed pretreatment, the water flux started at 0.048 L m^{-2}.min^{-1} and dropped to 0.042 L m^{-2}.min^{-1} after 900 min operation, with a flux reduction of 12.5% (Fig. 10). For the case of untreated FS, the permeate flux dropped from 0.044 L m^{-2}.min^{-1} to 0.036 L m^{-2}.min^{-1}, with a flux reduction of 17%. Due to the nature of the pretreatment method used and its duration, only the large suspended solid particles with sizes greater than 100 µm can be considered as settleable and removed from the feed solution, while the smaller particles remain suspended in the DCW (Metcalf and Eddy, 2003). These small particles are difficult to monitor and are often the ones that cause fouling to the membrane. Hence explains why the pre-settling of the FS did not show much improvement on the FO process performance.

Secondly, DCW was filtered through the multimedia filtration unit shown in Fig. 2 then the effluent with a turbidity of 24 NTU, was collected and used in the FO unit as the FS with a flow rate of 0.8 L min^{-1}. The rest of the experimental setting was similar to the settling method.

From Fig. 9 it could be seen that the best case scenario was obtained when the multimedia filtration method was employed, the permeate flux decreased from 0.065 L m^{-2}.min^{-1} to 0.06 L m^{-2}.min^{-1} and it remained fairly stable until the end of the experimental run. The multimedia filtration method gave the

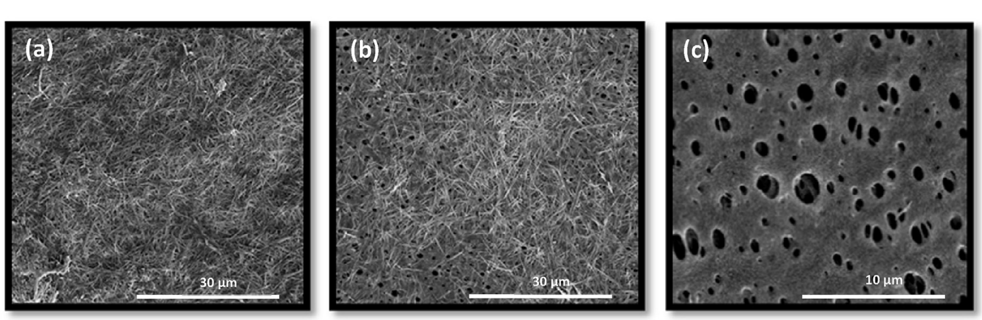

Fig. 8. SEM image for the CTA membrane after the FO process at a DCW flow rate of 0.8 L min^{-1}, a) without spacer (highly fouled surface), b) with spacer in the feed channel (reduced membrane fouling) c) clean membrane. (a) and (b) at magnification of 5,000X and (c) at magnification of 10,000X.

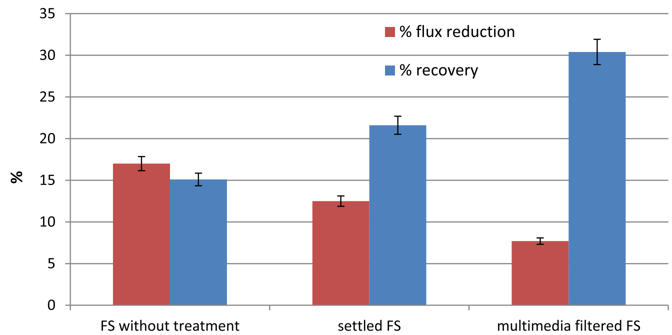

Fig. 10. Percentage flux reduction and percentage recovery for untreated, pre-settled and multimedia filtered DCW, operational conditions used are; 0.8 L min^{-1} FS flow rate, 35,000 ppm seawater DS, a spacer added in the feed solution side.

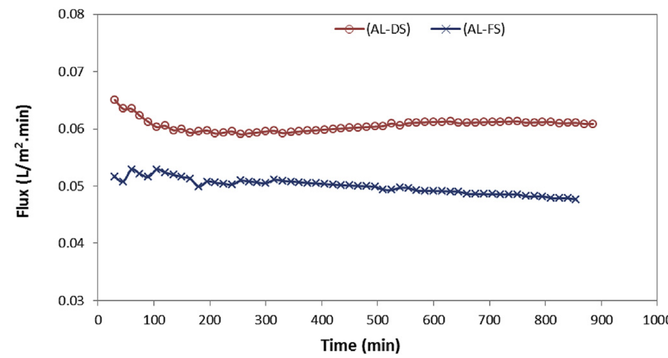

Fig. 12. Membrane flux of filtered DCW FS in (AL-DS) and (AL-FS) operational modes using 0.8 L min^{-1} of FS and DS flow rates.

lowest membrane flux reduction of 7.7% and the highest recovery rate of 30.4% as depicted in Fig. 10. The suspended solids presented in the original DCW with TSS value of 325 ppm got trapped in the media filter beds and resulted in an effluent with 24 NTU turbidity and 21 ppm TSS. These low values indicate that pre-filtration of the DCW feed removed most of the suspended solids which were responsible for membrane fouling and resulted in an elevated and a more consistent membrane flux. Moreover, the use of spacer in AL-DS operation mode with pre-filtered feed solution at low flow rate promoted mixing of feed solution and reduced the concentration polarization effects.

Membrane fouling often causes a noticeable decrease in the permeate flux as time proceeds. However, this was not the case when the multimedia filtered feed was used in which the permeate flux attained a steady value of 0.06 L m^{-2}.min^{-1} after only 100 min runtime. Fig. 11 reveals the differences between the SEM images of the CTA membranes following the FO process; run with pre-settled FS and multimedia filtered FS. It could be seen that the accumulation of fouling particles on the membrane surface was lower and more uniform when multimedia filtration was used compared to the case of pre-settled DCW. This further denotes that the pre-treatment process using multimedia filtration reduced membrane fouling making it a promising application in the FO process.

3.4. Effect of membrane orientation

One of the main important factors within any FO system is the membrane orientation. Generally, FO membranes tend to be asymmetric in nature (Qasim et al., 2015; Zhao et al., 2012), they are composed of an active dense layer and a porous support layer that provides mechanical support to the membrane (Wang et al., 2010). This structure leads to two distinct membrane orientations; active layer facing the feed side (AL-FS), or active layer facing the draw side (AL-DS), these orientations are unique to FO processes because of the presence of two different solutions with different concentrations on each side of the membrane.

Fig. 12 shows the effect of membrane orientation on flux performance using the same experimental conditions as in previous experiments; seawater draw solution, filtered DCW feed solution with a flow rate of 0.8 L min^{-1} for both FS and DS, and a spacer been inserted in the membrane feed side. It can be revealed that the permeate flux was almost stable in both runs and that the flux was 1.3 times higher for the case in which the membrane active layer was facing the draw solution. Despite the same net osmotic pressure difference being applied between the bulk FS and the bulk DS in the two membrane orientation modes, the resulting permeate flux was different due to the existence of different types of concentration polarization effects causing a reduction in the effective osmotic pressure. In the AL-DS membrane orientation mode, the incoming water flux moves from FS to the DS, diluting the DS at the active layer and increasing the concentration on the support layer facing the feed side causing concentrative internal CP (CICP). While, in the AL-FS membrane orientation mode, the DS dilution effect occurs within the support porous layer of the membrane and results in dilutive internal CP (DICP). Many studies agreed that DICP found in the AL-FS mode affects the membrane flux more severely than the AL-DS mode due to a high concentration of DS facing the porous support layer (Cornelissen et al., 2008; Gray et al., 2006; Zhang et al., 2014), which was the case in the current study. Accordingly, AL-DS operation mode can be considered more appropriate for the current application. Practically, DS-AL mode induces higher osmotic pressure across the membrane and generates more osmosis flux than FS-AL mode. Internal dilutive CP is more effective in reducing osmosis flux and should be considered in

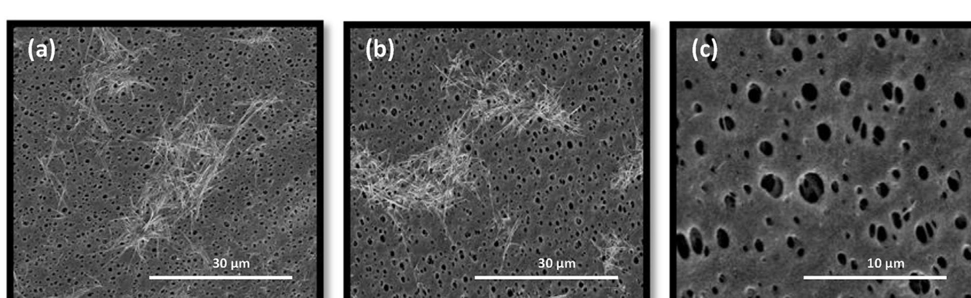

Fig. 11. SEM image for the CTA membrane after the FO process at a FS flow rate of 0.8 L min^{-1}, (a) pre-settled FS, (b) multimedia filtered FS, (c) clean membrane. (a) and (b) at magnification of 5,000X and (c) at magnification of 10,000X.

Table 2
Total cost of FO treatment of DCW for a 50 m^3/d plant capacity. The energy cost is 0.03 USD/kWh in Qatar, FO recovery rate is 30%, and one FO module. The actual membrane area required for DS-AL is 579 m^2 and for FS-AL is 698 m^2.

FO flux (m^3/m^2d)	FO cost (USD/mod)	FO mode	Es for SW pret (kWh/m)^3	Es for FO (kWh/m^3)	Power cost (USD/kWh)	Es for SW&DCW pret (USD/d)	Es for FO cost (USD/d)	FO module cost (USD/d)	Total cost (USD/d)
0.0864	12000	DS-AL	0.3	0.23	0.03	0.9	0.34	6.58	7.82
0.072	12000	FS-AL	0.3	0.27	0.03	0.9	0.402	6.58	7.88

FO processes. The results here suggest that DS-AL mode combined with multimedia pretreatment, probably, is the best option for DCW treatment by FO process. A moderate flow rate (2.2 L min^{-1}) of feed and draw solution should also be considered in the treatment process.

3.5. Cost analysis

Treatment cost is a crucial issue when it comes to technology development and application. Implementation of the FO technology for DCW treatment would increase the capital and operation costs of the site management and this cost should be estimated to determine the cost-effectiveness of the treatment process. Total cost of the FO process consists of the capital and energy costs based on an assumption of 50 m^3/d FO plant for DCW treatment. Commercial FO membranes come in different materials and configurations such as spiral wound FO (HTI, USA), flat sheet FO (Porifera, USA) and hollow fiber FO (Toyobo, Japan). The latter membrane has high packing density and comes in large modules, up to 700 m^2 of membrane area per module, with a reasonable price of USD 12,000 per module including the housing unit. Specific power consumption, *Es*, in the FO process can be determined from the following equation (Altaee et al., 2017):

$$E_s = \frac{P_f * Q_f}{\theta * Q_p} \qquad (2)$$

where, P_f is the feed pressure, Q_f is the feed flow rate, η is the pump efficiency (it is assumed to be 0.8), and Q_p is the permeate flow rate. The energy cost in Qatar is about 3 cent/kWh (0.03 USD/kWh). Seawater pretreatment for the FO membrane, using a conventional sand filtration, requires 0.3 kWh/m^3 (Straub et al., 2016). Based on the experimental results of this study (Fig. 12), water flux of the FO operating on DS-AL mode is 0.06 L/m^2.h (0.0864 m^3/m^2.d), FS and DS flow rate is 0.8 L/min (0.048 m^3/h), multimedia filtration pretreatment of SW and DCW, and the FO recovery rate is 30.4%. With 30.4% recovery rate, seawater will be diluted by 30% assuming the feed and draw solutions flow rates are equal. This will also reduce the energy consumption in the downstream RO process, which is responsible for 82% of the energy consumption in the FO-RO system (Altaee et al., 2017). For FO operating on FS-AL mode, water flux is 0.05 L/m^2.h (0.07 m^3/m^2.d) while FS and DS flow rates are similar to those in the DS-AL mode. Table 2 shows that the total cost for DCW treatment is 7.82 USD/d for DS-AL mode and 7.88 USD/d for FS-AL mode. The design criteria of DS-Al and FS-AL operating modes require only one Toyobo FO module of 700 m^2 of an active membrane area, although the actual membrane area required for DS-AL mode was 17% lower than that for the FS-AL mode. The total cost for 50 m^3/d includes the cost of FO membrane, pretreatment and FO treatment. Cost breakdown for DS-AL shows that 84% of the total cost was due the membrane area, 12% due to the pretreatment energy cost and 4% due to the FO energy cost (Fig. 13A). The corresponding values for FS-AL mode were 84% for the membrane area, 11% for pretreatment energy cost and 5% for FO energy cost (Fig. 13B). Changing the operating mode of the FO process from DS-AL to FS-AL increased the percentage of energy cost of the FO from 4% at DS-AL to 5% at FS-AL. It should be mentioned that the required FO membrane area was 579 m^2 for the DS-AL mode and 694 m^2 for the FS-AL mode. Treatment cost is very reasonable especially that the price of power in Qatar is the cheapest worldwide. Nevertheless, the total cost of FO treatment of DCW would still be reasonable in countries where energy prices are more expensive (normally 25 to 35 cent/kWh) because of the low energy requirements for FO treatment; 0.53 kWh/m^3. New research on FO membrane fabrication and development may further reduce the cost of membrane and improves water flux through reducing the effects of concentration polarization (Su et al., 2010; Zhang et al., 2010). This will make FO process more economic and viable technology in the near future.

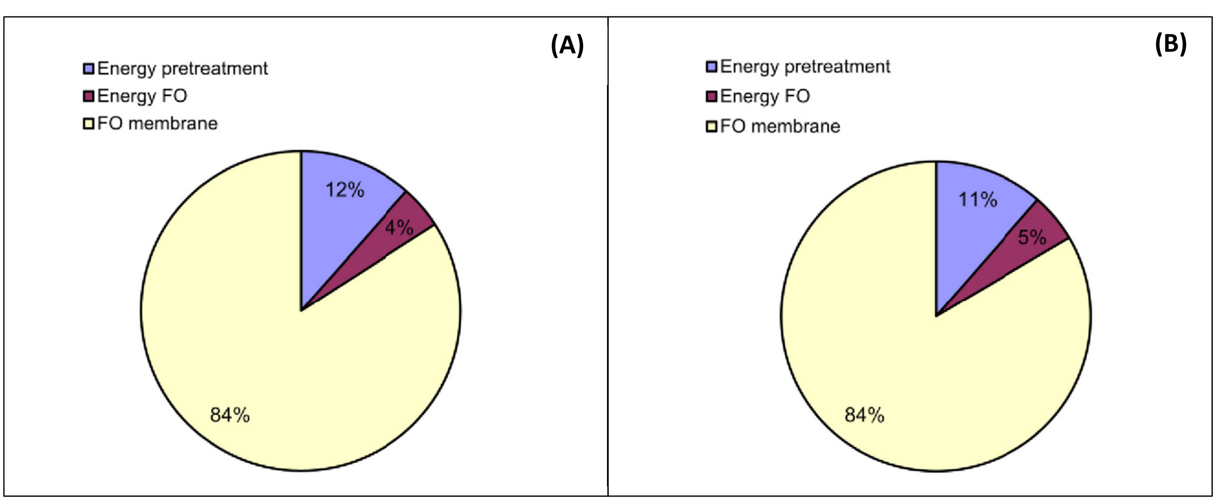

Fig. 13. Cost breakdown FO treatment of DCW A) DS-AL B) FS-AL.

4. Conclusion

In this study dewatering construction water was used as a feed solution and sea water was used as a draw solution in an integrated forward osmosis system. Experimental work showed that sand filtration of FS was more effective than column settlement as a satisfactory pretreatment for the FO process. Increasing feed solution resulted in an increase of the osmosis flux but increased membrane fouling especially for FS without pretreatment or with column settlement pretreatment only. Spacer placement did not improve water flux of the FO without FS pretreatment but higher water flux occurred when spacer is used in conjunction with FS column settlement. The impact of: flow rates of the FS and the DS, placement of spacer in the membrane feed side, pretreatment of the feed solution and orientation of the membrane, on process performance was investigated, and the following conclusions were drawn:

- Increasing the FS and DS flow rates enhanced the permeate flux but accelerated the accumulation of fouling particles at the membrane porous support layer facing the feed side, causing the water flux to decrease rapidly.
- The addition of a spacer on the feed side and in contact with the membrane's support layer was found to be effective at low FS flow rates, with a 17% flux reduction at 0.08 L min^{-1} as opposed to 53% when no spacer was used.
- Multimedia filtration pretreatment improved the quality of the feed solution to the FO process, reduced membrane fouling and resulted in a 64.3% enhancement in permeate flux using 0.8 L min^{-1} feed flow rate.
- The study recommends using the AL-DS membrane orientation mode with a spacer placed in the porous support layer and multimedia filtered feed solution at low flow rates to reduce the effects of membrane fouling in the seawater desalination process.

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