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Dual stage PRO process: impact of the membrane materials of the process performance

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ABSTRACT

A dual stage pressure retarded osmosis (PRO) process was investigated for power generation using different types of membranes. Polyamide (PA) and cellulose triacetate (CTA) membranes were used in the first and second stage of the PRO process to improve the process performance due to the high water permeability of PA membranes. A comparison between dual stage PA–CTA and CTA–CTA membrane systems were carried out using seawater as a draw solution, while fresh water and wastewater effluent were the feed solution in the first and second stage of the process. The impact of draw solution flow rate on the process performance was evaluated. The performance of first and second stage of the PRO process increased by 11.5 and 28.6%, respectively, when the draw solution flow rate increased by a factor of 2.5. In return, there was a negligible increase in the total specific power consumption of the PRO process. In general, power consumption of the dual stage PRO process was as low as 0.3 kWh/m^3. Furthermore, the results showed that the performance of the dual stage PRO process increased with increasing seawater salinity from 32 to 50 g/L due to the higher net driving pressure across the membrane. Finally, power generation in the PA–CTA system was up to 33% higher than that in the CTA–CTA system.

Keywords: Pressure retarded osmosis; Dual stage PRO; Renewable energy; Polyamide membranes; Osmotic power plant

1. Introduction

Technologies for power generation from renewable sources has received a lot of attention for being able to reduce the environmental impact of fossil fuel burning for power generation in conventional power plants. One of the promising technologies of power generation is the pressure retarded osmosis (PRO). Initially, PRO was proposed by Sidney Loeb in the 1970s for power generation from a salinity gradient resource using a semipermeable membrane [1]. Theoretical and experimental works have shown promising results,

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which encouraged further works to investigate the process potential [2–6]. Unfortunately, there were not many successful examples of PRO applications for power generation in large commercial scales. As a matter of fact, there isn't any PRO power plant in operation around the world, despite the large number of work has been done in this field. There are, however, a number of small capacity pilot plants but data from these experimental works are scarce and not always available [6].

In the conventional PRO process, two solutions of different concentrations and osmotic pressure are pumped into the membrane module for fresh water extraction from solution of lower concentration. The high-concentration solution is pressurized before entering the PRO module and power generation takes place as the high-concentration solution is depressurized by a hydro turbine system [1]. For an economical PRO process, the power density (which is the power produced per unit area of membrane) should be between 3 and 5 W/m^2 [6,7]. A novel dual stage PRO has been recently proposed to improve the process performance [8]. In the dual stage PRO process, seawater or the high-concentration solution is pressurized and fed into the first stage of the PRO membrane, whereas the low-concentration solution is fed on the opposite side of the PRO membrane. The pressurized flow divides into two parts after leaving the PRO membrane, part one goes to a pressure exchanger to pressurize the draw solution and part two enters a second PRO membrane. In the second stage of the PRO process, fresh water permeates from the feed to the seawater making it more diluted. Then, the entire pressurized flow from the second stage goes to a hydro turbine system to convert the hydraulic energy to electricity. Dual stage PRO process has several advantages over the conventional PRO process, such as (Fig. 1) [8] (1) accepting different feed water quality in the different stages of the process (2) reducing the effect of feed salinity on the process performance (3) using different types of membrane at the different stages of the PRO process, and finally (4) increasing the amount of power generated from the PRO process. Besides, dual stage PRO process doesn't require a high-pressure pump or a pressure exchanger in the second stage of the process.

PRO membrane plays a key role in the process performance. Typical PRO membrane should have high water permeability, high ions rejection rate, and good mechanical stability [9]. Two types of membranes are typically used in the PRO process: cellulose triacetate (CTA) and polyamide (PA) membranes [5,8]. CTA membrane exhibits higher chlorine resistance, lower water flux, and lower hydrophobicity than the PA membrane [9,10] but PA membranes have higher water permeability than the CTA membranes [11]. Because of its high resistance to chlorination, CTA membranes are more suitable for PRO process, especially when wastewater effluent is used as the feed solution [11]. This is particularly important in case of insufficient amount of fresh water that is available as feed solution [8]. In case of using fresh water in the first stage and wastewater effluent in the second stage of the PRO process, PA and CTA membranes can be used in the first and second stages, respectively. This scenario is possible in regions with a limited amount of fresh water and wastewater effluent is used a supplement low salinity resource. The high water permeability and low chlorine resistance PA membrane is suitable for first stage seawater–fresh water salinity gradient, whereas the relatively high chlorine resistance CTA membrane is suitable for the second stage seawater–wastewater effluent salinity gradient. It has been experimentally demonstrated that using wastewater effluent in the PRO process feed increases membrane fouling propensity [12]. Therefore, it is desirable using CTA membrane of relatively high chlorine resistance with wastewater effluent feed.

In the current study, a dual stage PRO process has been proposed for power generation using fresh water and wastewater effluent as feed solution in the first and the second stage of the process, respectively, and seawater as the draw solution. PA and CTA membranes have been used in the first and the second stage of the PRO process, respectively. Using two different PRO membranes at the different stages of the dual stage PRO process has the advantage of reducing the membrane fouling and improving the PRO performance. The performances of the first and second stage of the PRO process were evaluated for a number of feed salinities ranging from 32 to 50 g/L. Furthermore, the impact of increasing seawater (draw solution) flow rate on the performance of PRO process was also evaluated. Finally, pre-developed software was used throughout this study to estimate the performance of the PRO process.

2. Methodology

Dual stage PRO process is a novel technique for enhancing the process performance using different salinity gradient resources. It is assumed, here, that fresh water and wastewater effluent were the feed solutions in the first and second stage, respectively, while seawater was the draw solution. The Total Dissolved Solids (TDS) of fresh water and wastewater was assumed about 200 mg/L [13], whereas seawater concentrations varied between 32 and 50 g/L. Two

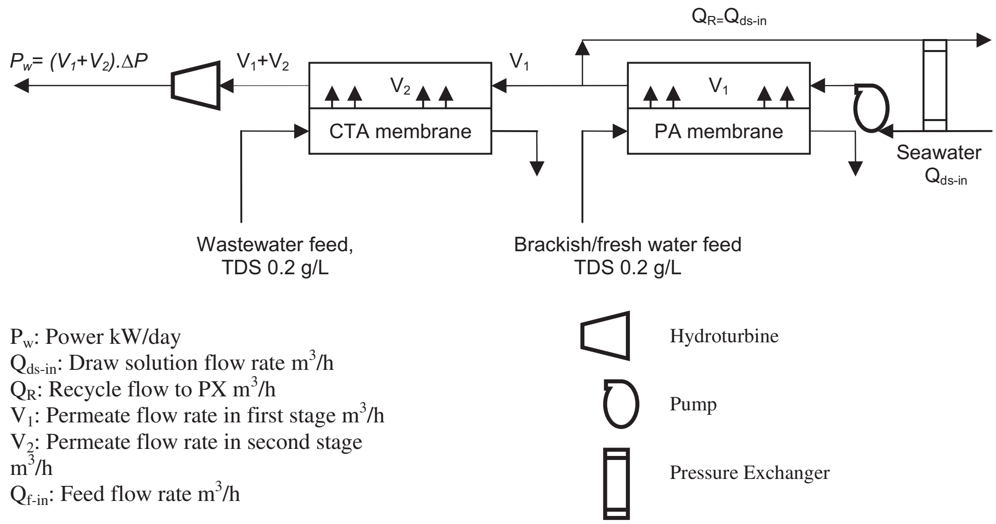

Fig. 1. Dual stage PRO process using PA and CTA membranes.

types of PRO membrane, i.e. PA and CTA, where used in the first and second stage of the PRO process, respectively, for higher process performance. The water permeability, A_w ($L/m^2\,h\,bar$) of the PA and CTA membranes was estimated from the following equation:

$$A_w = \frac{J_w}{(\Delta P - \Delta \pi)} \quad (1)$$

where J_w is water flux ($L/m^2\,h$), ΔP is the hydraulic pressure difference across the membrane (bar), and $\Delta\pi$ is the osmotic pressure difference across the membrane (bar). Water permeability is usually measured in reverse osmosis (RO) cell using a range of applied hydraulic pressures. In the current study, the water permeabilities of PA and CTA membranes were estimated about 1.13 and 0.79 $L/m^2\,h\,bar$, respectively [12,14]. Additionally, salt permeability coefficient, B (m/d), can be estimated from the following equation:

$$B = \frac{(1 - R_j)J_w}{R_j} \quad (2)$$

where R_j is the membrane rejection rate. Both CTA and PA membrane have high rejection rate to monovalent and divalent ions; 98 and 99.5%, respectively [14–16]. Power density, W (W/m^2), of the PA and CTA membranes in the first and second stage of the process, respectively, have been calculated from the following equation:

$$W = \Delta P \times J_w \quad (3)$$

According to Eq. (3), power density is a function of the hydraulic pressure difference across the membrane and membrane flux. Power density reaches a maximum amount at $\Delta P \leq \Delta\pi/2$. Therefore, for a given salinity gradient resource, the applied hydraulic pressure cannot be increased indefinitely to increase the power density. Alternatively, power density can be increased by using a high water permeability membrane such as PA membrane. These membranes have low fouling and chlorine, and hence were employed in the first stage of the PRO process where fresh water and seawater are the salinity gradient resource [9,17].

Generally, specific power consumption, Es (kWh/m^3), of the first and second stage of the PRO process was calculated using the following equation [18]:

$$\mathrm{Es} = \frac{(P_f \times Q_f) + (P_{ds} \times Q_{ds})}{\eta \times Q_p} \quad (4)$$

where P_f and P_{ds} are the feed and draw solution pressure (bar), Q_f and Q_{ds} are the feed and draw solution flow rate (m^3/h), η is pump efficiency, and Q_p is permeate flow rate (m^3/h). In the current study, the pump efficiency, η, was assumed 0.8. Furthermore, the specific power consumption of the PRO process was calculated for a number of draw to feed solution ratios, $Q_{ds\text{-}in}/Q_{f\text{-}in}$, ranging from 1 to 2.5. Finally, the total specific power consumption of the PRO process, Est, is equal to the summation of specific power consumption of the first and second stage, Es1 and Es2, respectively:

$$\mathrm{Est} = \mathrm{Es1} + \mathrm{Es2} \quad (5)$$

As illustrated in Fig. 1, the gross power generation (P_{wt}), which is the function of the total permeate flow (V_1+V_2) multiplied by the hydraulic pressure (ΔP), can be calculated from the following equation:

$$P_w = \Delta P \times (V_1 + V_2) \tag{6}$$

In the current study, it is assumed that there is no pressure loss in between stages one and two, and therefore, the hydraulic pressures of the first and second stage are equal. The permeate flow rates of the first and second stage of the dual PRO process are calculated from the following equations [8]:

$$Q_{p1} = Q_{pt}\frac{J_{w1}}{J_{wt}} \tag{7}$$

$$Q_{p2} = Q_{pt}\frac{J_{w2}}{J_{wt}} \tag{8}$$

In Eqs. (7) and (8), Q_{p1} and Q_{p2} are the first and second stage permeate flow rate (m^3/h), Q_{pt} is the total permeate flow rate (m^3/h), J_{w1} and J_{w2} are, respectively, the membrane flux of first and second stage, and J_{wt} is the total membrane flux. Eqs. (7) and (8) are used to calculate the membrane area, A_m (m^2) as in the following equations:

$$A_{m1} = \frac{Q_{p1}}{J_{w1}} \tag{9}$$

$$A_{m2} = \frac{Q_{p2}}{J_{w2}} \tag{10}$$

Seawater composition can be found in previous literature [19]. Finally, two scenarios of dual stage PRO process have been investigated in this study; in the first scenario, CTA membrane was used in the first and second stage of the PRO process, CTA–CTA system. While in the second scenario, PA and CTA membranes were used in the first and second stage, respectively, PA–CTA system. For simplicity, these systems were denoted to as a CTA and PA in the legends of figures throughout the paper. It should be mentioned here that one full-scale HTI CTA/PA spiral wound element, area is about 16.5 m^2, was used in the simulation process.

3. Results and discussion

3.1. Impact of applied hydraulic pressure

Fig. 2(a)–(f) illustrates the impact of hydraulic pressure on the performance of PRO process at different

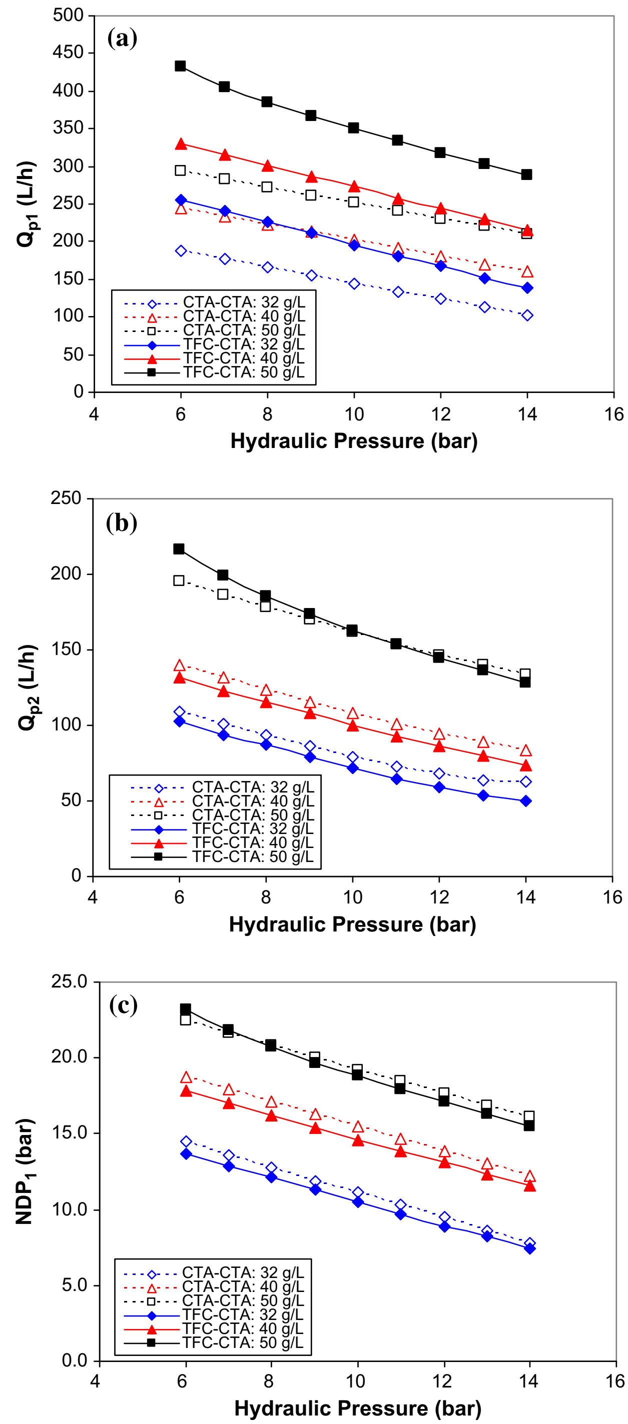

Fig. 2. Impact of hydraulic feed pressure on the PRO performance at different seawater salinities (a) Q_{p1} (b) Q_{p2} (c) NDP_1 (d) NDP_2 (e) $\%Re_1$ (f) $\%Re_2$.

seawater salinities, $Q_{ds\text{-}in}/Q_{f\text{-}in}$ ratio equals to 1. At any given hydraulic pressure, the permeate flow rate of the first stage increased with increasing the TDS of seawater from 32 to 50 g/L (Fig. 2(a)). This was

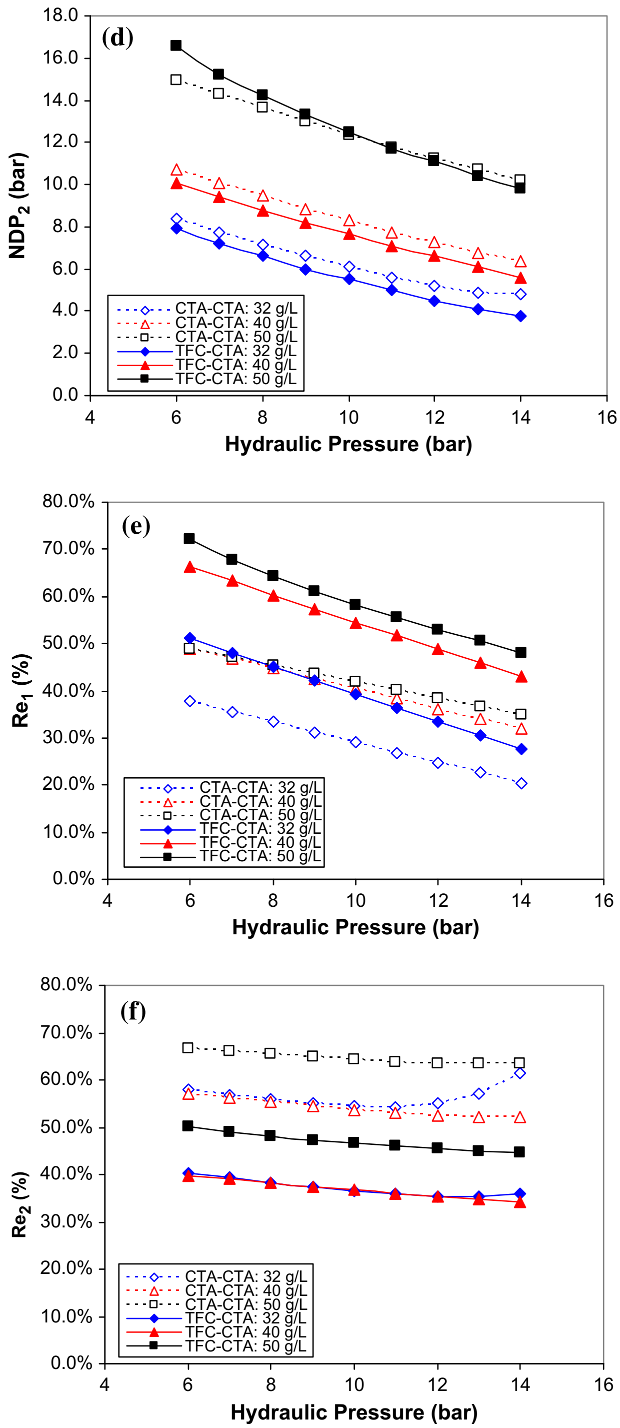

Fig. 2. (*Continued*).

attributed to the higher osmotic pressure of seawater at 50 g/L. However, the results show that the permeate flow rate decreased with increasing the hydraulic feed pressure. In the PRO process, the net driving pressure (NDP) across the membrane decreases with increasing the hydraulic pressure causing membrane flux decline. This is obviously shown in Fig. 2(c), a sharp drop in the NDP_1 occurred with increasing the hydraulic pressure. The results also showed that the permeate flow of the PA membrane was higher than that of the CTA membrane (Fig. 2(a)). As explained before, the water permeability coefficients of PA and CTA membranes were 1.13 and 0.79 L/m^2 h, respectively.

In general, membrane flux increases with increasing its water permeability coefficient which explains the higher permeate flow in case of the PA membranes. The permeate flow rate in the second stage of the PRO process is shown in Fig. 2(b). Practically, CTA membranes have higher resistance to the chlorination process, and hence are more suitable for wastewater effluent treatment [9]. Unlike the first stage of the PRO process, permeate flow rates of the CTA–CTA and PA–CTA are equal in the second stage of the PRO process (Fig. 2(b)). As a matter of fact, the permeate flow of the second stage was slightly higher in the CTA–CTA system than in the PA–CTA system. The reason for that was due to the higher NDP_2 in the CTA–CTA system than in the PA–CTA system. In the first stage of the PRO process, permeate flow rate was higher in the PA–CTA system than in CTA–CTA system, which resulted increased seawater dilution in the former system. This, in fact, affected the osmotic pressure of seawater to the second stage, which was lower in the PA–CTA system than in the CTA–CTA system (Fig. 2(d)). Yet, the total permeate flow rate was higher in the PA–CTA system than in the CTA–CTA system. The recovery rate of the PRO process is shown in Fig. 2(e) and (f). The higher the membrane recovery rate the lower membrane area is required for filtration. Apparently, the recovery rate of the first stage decreased with increasing the hydraulic feed pressure because of the lower NDP (Fig. 2(c)–(e)). Table 1 shows the osmotic pressure difference across the membrane in the first and second stage of the PRO process at difference feed hydraulic pressures. Obviously, the osmotic pressure difference was higher at higher seawater TDS and was also higher in the first stage than in the second stage. Higher the osmotic pressure across the membrane, higher the membrane flux. It also observed that the recovery rate increased with increasing the salinity of seawater as shown in Fig. 2(e). In the first stage of the PRO process, the difference in the recovery rate between 50 and 40 g/L salinities was significantly lower than that between 50 and 32 g/L salinities. In the PA–CTA system, for example, the difference in the recovery rate between 32 and 50 g/L salinities was 29%, while between 50 and 40 g/L was 8% only. This was due to the

Table 1
Osmotic pressure difference in the first and second stage of the dual PRO process

	32 g/L		40 g/L		50 g/L	
P (bar)	$\Delta\pi_1$	$\Delta\pi_2$	$\Delta\pi_1$	$\Delta\pi_2$	$\Delta\pi_1$	$\Delta\pi_2$
6	20.4	14.4	24.7	16.7	29.2	22.6
7	20.6	14.7	24.9	17.1	28.8	22.2
8	20.8	15.1	25.1	17.5	28.7	22.2
9	20.9	15.6	25.3	17.9	28.7	22.3
10	21.1	16.1	25.5	18.3	28.8	22.5
11	21.3	16.6	25.7	18.8	28.9	22.7
12	21.5	17.2	25.8	19.2	29.1	23.1
13	21.6	17.9	26.0	19.8	29.3	23.4
14	21.8	18.8	26.2	20.4	29.5	23.8

increased effect of concentration polarization at higher seawater salinity, which resulted in a lower NDP across the membrane [20]. In the second stage, there was a negligible difference in the recovery rate between 32 and 40 g/L for a wide range of hydraulic pressures.

This trend was noticed in both systems; i.e. PA–CTA and CTA–CTA. It should be mentioned, here, that the $\%Re_1$ is higher in the PA–CTA than in the CTA–CTA, while $\%Re_2$ was higher in the CTA–CTA than in the PA–CTA (Fig. 2(e) and (f)). Practically, the recovery rate is affected by the NDP and the membrane flux; the higher water permeability of PA membrane increased the $\%Re_1$ and the dilution of seawater. Consequently, the NDP_2 was lower in the PA–CTA system than in the CTA–CTA system.

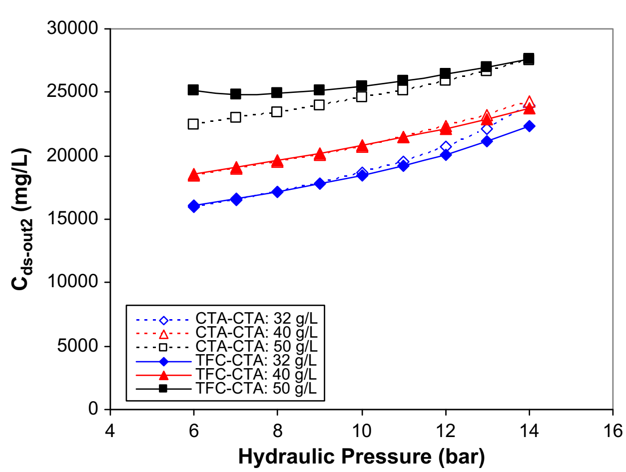

Fig. 3. Impact of hydraulic pressure on the concentration of seawater to discharge.

Finally, the TDS of draw solution leaving the second stage of the PRO process, $C_{\text{ds-out2}}$, is shown in Fig. 3. The concentration of draw solution increased with increasing the seawater salinity and the hydraulic feed pressure. Up to 50%, seawater dilution can be achieved at low hydraulic pressure; this phenomenon has been observed at all seawater salinities. The results also show a negligible difference in the $C_{\text{ds-out2}}$ between the CTA–CTA system and PA–CTA system. Usually, the brine waste from the PRO process is discharged to the sea. In arid and semi-arid areas, the diluted brine waste from the PRO process can be processed by suitable membrane or thermal desalination for fresh water extraction at low cost [20,21]. Furthermore, $C_{\text{ds-out2}}$ will affect the cost of draw solution regeneration if a custom-design osmotic agent is used as a draw solution.

3.2. Impact of draw solution flow rate

Previous works on PRO and Forward Osmosis (FO) have shown that the impact of increasing the feed flow rate on the process performance was negligible, particularly, at low feed concentrations [20]. However, changing the flow rate of draw solution showed a tangible impact on the performance PRO process [20–22]. Earlier studies showed that increasing the draw solution flow rate reduces the effect of concentration polarization at the membrane surface as a result of increasing the draw solution bulk concentration [20]. Fig. 4 shows the effect of draw solution flow rate on the performance of PRO process, using 32 g/L seawater salinity and CTA membrane in both stages. Apparently, membrane flux increased with increasing the $Q_{\text{ds-in}}/Q_{\text{f-in}}$ ratio from 1 to 2.5 (Fig. 2(a)). Increasing the flow rate of draw solution increased the concentration of bulk solution and the concentration of boundary layer at the membrane surface [23]. Consequently, the permeation of fresh water from the feed to the draw solution increased due to the higher NDP at across the PRO membrane (Fig. 4(b)). As it was expected, permeate flow and membrane flux was higher in the first stage than in the second stage of the PRO process, due to the higher NDP in the first stage (Fig. 4(b)). This observation holds true for a range of hydraulic feed pressures.

However, only part of the Q_{p1}, which is equal to V_1, goes to the second stage of the PRO process for further treatment (Fig. 1). It is important mentioning, here, that the membrane flux decreased as the hydraulic feed pressure increased from 7 to 10 bar due to the lower NDP across the PRO membrane (Fig. 4(b)). The impact of increasing the flow rate of draw solution on the PRO recovery rate is illustrated in Fig. 4(c). The

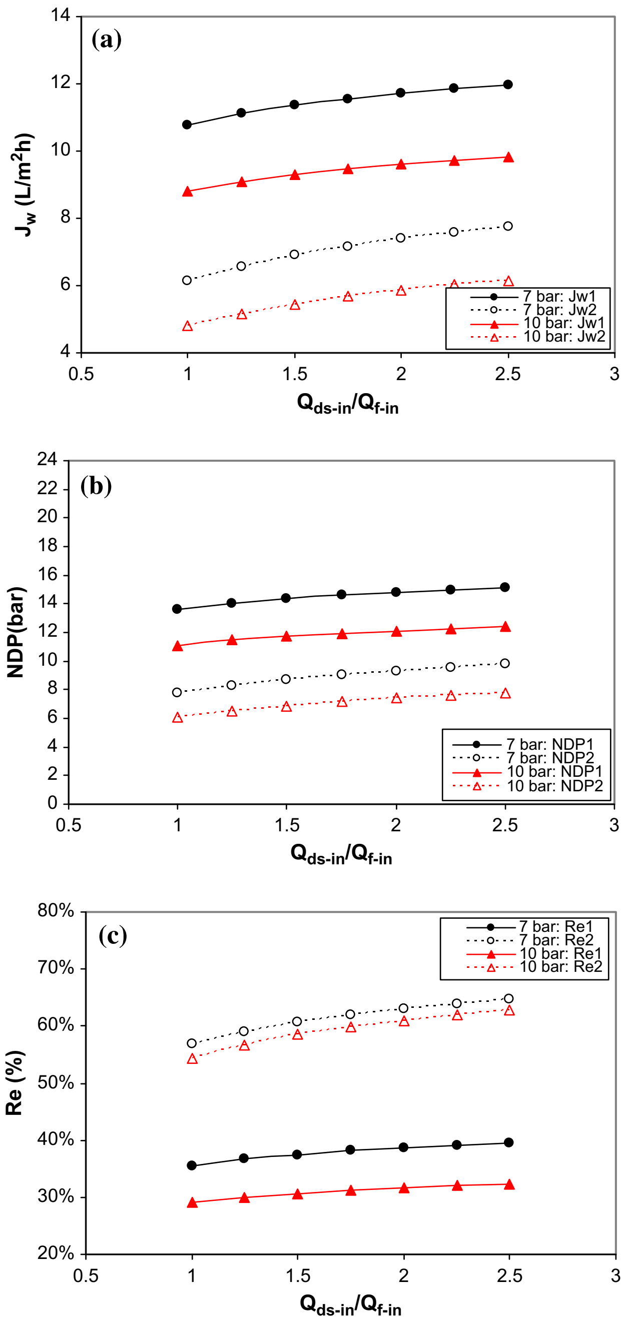

Fig. 4. Effect of draw solution flow rate on the performance of dual stage PRO process (a) membrane flux (b) NDP (c) recovery rate.

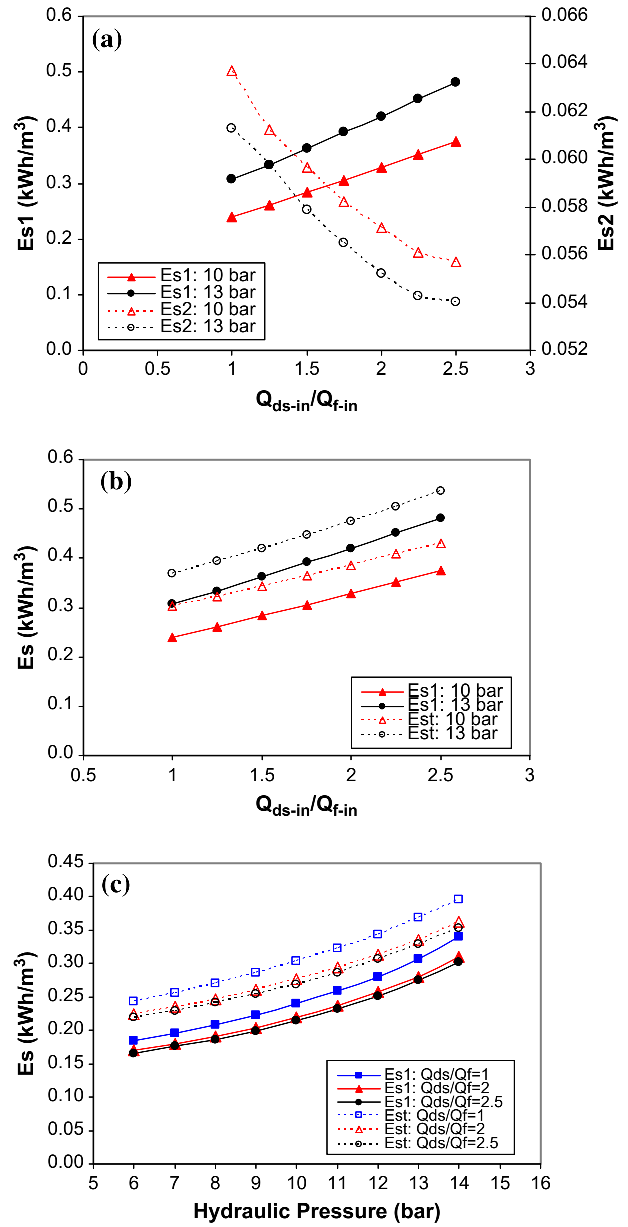

Fig. 5. Specific power consumption in the PRO process at different hydraulic pressures and draw solution flow rates.

results show that the recovery rate increased with increasing the flow rate of draw solution and decreased with increasing the hydraulic feed pressure. The latter observation was due to the lower NDP across the membrane at higher hydraulic pressure (Fig. 4(b)). Additionally, the recovery rate was higher in the second stage than in the first stage of the PRO process. However, the high percentage recovery rate of the second stage should not be confused with the fact that the membrane flux was lower in the second stage. As a matter of fact, the membrane flux was lower in the second stage because of the lower NDP across the membrane. Previous study demonstrated that the size of the first and second stage of the process will not be significantly different [8].

In general, increasing the draw solution flow rate enhances the membrane flux, but it contributes to higher power consumption. Eq. (4) was used to estimate the specific power consumption of the first and second stage of the PRO process, whereas the total specific power consumption was estimated from Eq. (5). It should be taken into account that the P_{ds} of the second stage is equal to zero. Fig. 5(a) shows the specific power consumption in the first stage, Es1, and second stage, Es2, of the PRO process. As expected, the specific power consumption in the first stage increased with increasing the $Q_{ds\text{-}in}/Q_{f\text{-}in}$ ratio. Power consumption increases with increasing the flow rate, Q_{ds}, as illustrated in Eq. (4), and hence the cost of power generation by PRO process will be slightly higher. For example, at 10 bar hydraulic pressure, the specific power consumption in the first stage of the PRO process increased by almost 56%, when the $Q_{ds\text{-}in}/Q_{f\text{-}in}$ ratio increased from 1 to 2.5. On contrary, specific power consumption in the second stage of the PRO process decreased as the $Q_{ds\text{-}in}/Q_{f\text{-}in}$ ratio gradually increased from 1 to 2.5 (Fig. 2(a)). At 10 bar hydraulic feed pressure, about 12% decrease in the second stage specific power consumption was achieved by increasing the $Q_{ds\text{-}in}/Q_{f\text{-}in}$ ratio from 1 to 2.5.

However, power consumption in the second stage of the PRO process is much lower than that in the first stage of the PRO process. Under testing condition of 10 bar hydraulic feed pressure and $Q_{ds\text{-}in}/Q_{f\text{-}in}$ ratio equals to 1, for example, the specific power consumption in the second stage of the PRO process was only 17% of that in the first stage (Fig. 5(a)). In general, the operation cost of second stage of the PRO process is infinitesimal compared to the first stage, which is in turn is much lower than that in hyperfiltration processes. Furthermore, Fig. 5(b) shows that the total specific power consumption, which is estimated from Eq. (5), increased with increasing the draw solution flow rate. The figure also shows that there is insignificant difference between Est and Es1, due to the subtle energy requirement for the operation of the second stage of the PRO process. In fact, this trend was observed for a number of feed hydraulic pressures.

Furthermore, the impact of increasing the draw solution flow rate on the power consumption of PRO process was evaluated at different hydraulic pressures (Fig. 5(c)). The results showed that the total specific power consumption increased with increasing the hydraulic pressure. As shown in Fig. 2(a), there is an inverse relationship between the applied hydraulic pressure and the permeate flow rate. The latter parameter decreased with increasing the hydraulic pressure and hence, resulted in higher specific power consumption according to the Eq. (4). As stated before, the difference between Est and Es1 was insignificant, which is an indicative of the fact that the contribution of the second stage of the PRO to the process total power consumption was negligible.

The results show that increasing the draw solution flow rate enhances the performance of PRO process at insignificant increase in the power consumption. Ignoring the cost of draw solution pretreatment (if required), it is therefore desirable operating the PRO process on high draw solution flow rates. The impact of draw solution flow rate on the membrane power density is shown in Fig. 6 using 32 g/L seawater salinity and osmotic pressure difference about 24 bar. At 10 bar feed hydraulic pressure, the power density of the first and second stage increased by 11.5 and 28.6%, respectively, with increasing the $Q_{ds\text{-}in}/Q_{f\text{-}in}$ ratio from 1 to 2.5. This resulted in increasing the total specific power consumption of the dual stage PRO process from 0.3 to 0.43 kWh/m^3. To evaluate the energy efficiency of the second stage of the PRO process, the ratio of second stage specific power consumption of the total specific power consumption, Es2/Est, was calculated. At 10 bar hydraulic pressure, the Es2/Est ratio was 21 and 13% at 1 and 2.5 $Q_{ds\text{-}in}/Q_{f\text{-}in}$ ratios, respectively. These results show two important facts; firstly, the specific power consumption of the second stage deceases with increasing the draw solution flow rate and secondly, the power consumption of the second stage is insignificant compared with the total power consumption of the dual stage PRO process. In nutshell, osmotically driven membrane processes, such as PRO and FO, are cheaper to operate than the pressure-driven membrane process like RO and Nanofiltration processes.

3.3. Power generation

The key output parameter to evaluate the performance of PRO process is power density or the power

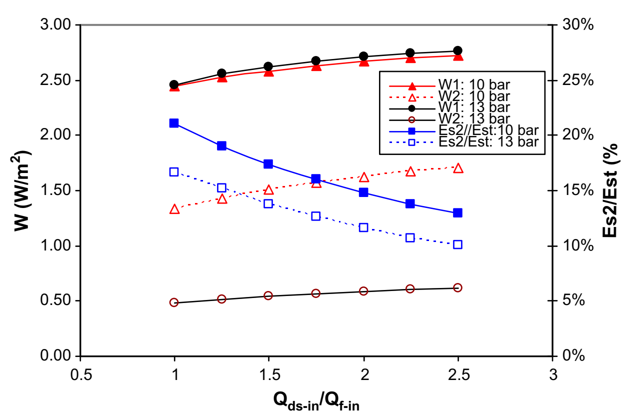

Fig. 6. Power density at different draw solution flow rates.

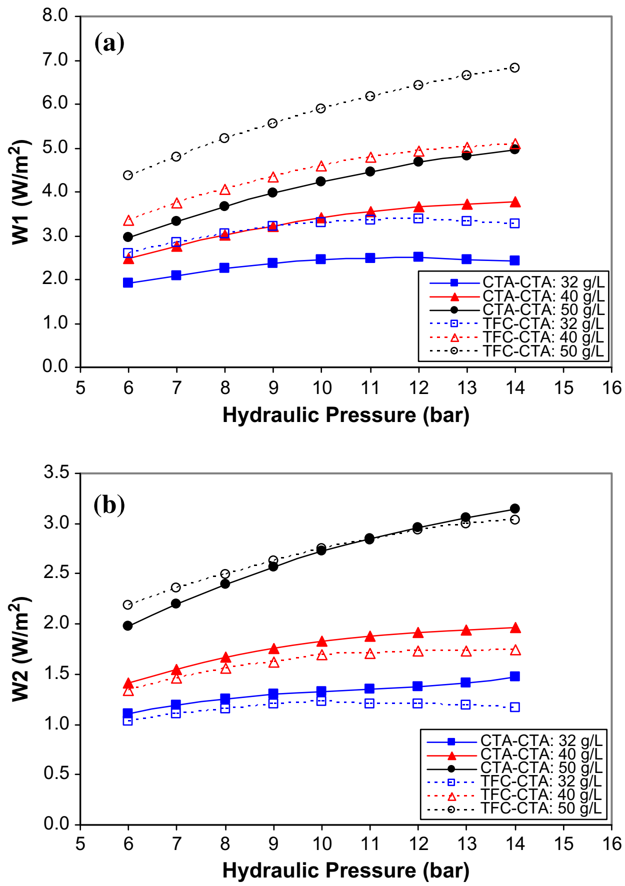

Fig. 7. Power density in the CTA–CTA and PA–CTA dual stage PRO process (a) stage 1 (b) stage 2.

generated per unit area of membrane. Power densities of the first and second stage of the PRO process were calculated and presented in Fig. 7(a) and (b), respectively. In the first stage of the PRO process, power density of the PA–CTA system was higher than that of the CTA–CTA system (Fig. 7(a)). This was due to the higher water permeability in the PA membrane than in the CTA membrane, which resulted in a higher power density in the PA–CTA system (Fig. 2(a)). The highest power density was achieved at 50 g/L seawater salinity, which indicates that power density increases with increasing the osmotic pressure of draw solution. Power density also increased with increasing the hydraulic pressure until it reached a maximum amount at $\Delta P = \Delta\pi/2$. This phenomenon is clearly manifested in Fig. 7(a). At 32 g/L seawater salinity and osmotic pressure difference about 24 bar, power density increased with increasing the hydraulic pressure from 6 to 12 bar, then started to decrease as the hydraulic pressure further increased up to 14 bar. Oppositely to the first stage of the PRO process, power density in the second stage was about 6% higher in the CTA–CTA system than in the PA–CTA system (Fig. 7(b)). It should be mentioned that the difference in the second stage power density between the CTA–CTA and PA–CTA systems was insignificant, especially at 50 g/L seawater salinity. Based on these results, dual stage PRO process performs better at high draw solution concentrations than at low concentrations. Furthermore, the first stage power density of PA–CTA system has met the threshold recommended for economical PRO process; 3 to 5 W/m^2. This holds for all seawater salinities tested in the current study. In case of CTA–CTA, the system performance was lower than the PA–CTA system. For instance, at 32 g/L seawater salinity, the first stage power density was 2.5 and 3.37 W/m^2, respectively, for CTA–CTA and PA–CTA system. However, the second stage power density was less than 3 W/m^2 at 32 and 40 g/L, but it was just over 3 W/m^2 at 50 g/L (Fig. 7(b)). This substantiates the findings that PRO process performs better at higher seawater salinities. It should be mentioned, here, that there is no significant difference in the second stage power density between the CTA–CTA and PA–CTA systems as shown in Fig. 7(b).

In the dual stage PRO process, the permeate flow from the second and first stage of the process are combined together and sent to a hydro turbine system for power generation (Fig. 1). The total power generation of the dual stage PRO process is calculated from Eq. (6) and illustrated in Fig. 8. The results show that the power generation was higher in the PA–CTA system than in the CTA–CTA system. Since same type of membrane was used in the second stage of the PRO process, the higher power generation in the PA–CTA system was mainly attributed to the high performance of the first stage, in which a PA membrane was used.

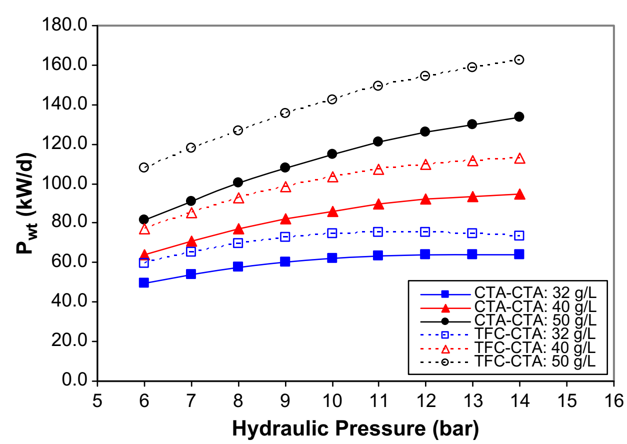

Fig. 8. Gross power generation by the CTA–CTA and PA–CTA systems at different seawater salinities.

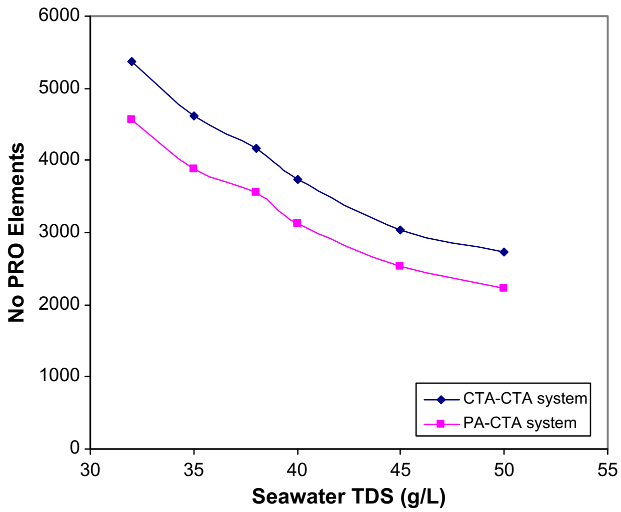

Fig. 9. Number of PRO in the CTA–CTA and PA–CTA systems at different seawater TDS.

Apparently, the difference in total power between the PA–CTA system and the CTA–CTA system increased with increasing the salinity of seawater water (Fig. 8). At 6 bar hydraulic feed pressure and 32 g/L seawater salinity, there was a 20% difference in P_{wt} between the PA–CTA system and the CTA–CTA system. However, the difference increased to 33% at 50 g/L seawater salinity.

In general, PA–CTA system is more efficient than the CTA–CTA system for power generation from dual stage PRO process. PA membranes enjoy higher water permeability than CTA membrane, and have wide application in water treatment and desalination. The amount of power generation increased with increasing the hydraulic pressure and reached a maximum amount at $\Delta P_w \geq \Delta\pi/2$. In addition, the results showed that the performance of the dual stage PRO process increased with increasing the concentration of draw solution. This was particularly noticeable in the PA–CTA system, which exhibited much higher performance than the CTA–CTA system at high seawater salinities. This means that the performance of dual stage PRO process would be higher in regions, where seawater salinity is very high like the Middle East. Additionally, a custom-design draw solution of high concentration can be used as a draw solution. In this case, a third stage treatment process is required for the regeneration of the draw solution.

3.4. Membrane elements

The sizing of CTA–CTA and PA–CTA plants is briefly investigated here. The required membrane areas of the first and second stage of the PRO process were estimated from Eqs. (7) and (8), respectively. Number of PRO elements of the first and second stage was then calculated from Eqs. (9) and (10), respectively. For a 30,000 m^3/d plant capacity, the number of required PRO elements of CTA–CTA and PA–CTA systems is shown in Table 2. The results show that PA–CTA system requires less membrane elements than the CTA–CTA system because of its higher total permeate flux. At 50 g/L seawater salinity, for example, the PA–CTA system required about 18.5% less membrane elements than the CTA–CTA system. The results also show that the number of PRO membrane elements decreased with increasing seawater salinity from 32 to 50 g/L because membrane flux was higher at higher seawater salinity (Fig. 9).

In general, the PA–CTA system is more efficient in term pf power density and power generation than the CTA–CTA system. PA membranes have high tolerance to a wide range of feed pH and can be used with fresh or brackish feed waters of low fouling concentration. The cost of PA and CTA membrane are almost the same, but PA–CTA system requires less membrane element which will reduce the plant cost. Additionally, the footprint of PA–CTA plant is less than that of the CTA–CTA system which decreases the plant capital cost.

Table 2
Number of required element in PRO plant, $Q_{\text{ds-in}}/Q_{\text{f-in}} = 1$

	CTA–CTA system			PA–CTA system		
SW (g/L)	Stage 1 (No. Elem.)	Stage 2 (No. Elem.)	Total (No. Elem.)	Stage 1 (No. Elem.)	Stage 2 (No. Elem.)	Total No. Elem.
32	2,689	2,689	5,379	2,281	2,281	4,562
35	2,311	2,311	4,622	1941	1941	3,882
38	2086	2086	4,172	1,776	1,776	3,551
40	1870	1870	3,741	1,563	1,563	3,125
45	1,519	1,519	3,038	1,265	1,265	2,530
50	1,366	1,366	2,732	1,112	1,112	2,224

4. Conclusion

The impact of membrane type of the dual stage PRO process was realized through increasing water permeability of the first stage of the PRO process, where the PA membrane was applied. The results showed that power density was higher in the PA–CTA system than in the CTA–CTA system. This outperformance was attributed to the higher water permeability of the PA membrane in the first stage of the PA–CTA system. In the PA–CTA system, power density of the first stage was over 3 W/m^2 for all seawater salinities. The performance of the second stage of the PRO process was almost equal in both systems; namely the PA–CTA and the CTA–CTA. Furthermore, the power density of the first and second stage of the PRO process increased by 11.5 and 28.6%, when the $Q_{\text{ds-in}}/Q_{\text{f-in}}$ ratio increased from 1 to 2.5. In return, this resulted in negligible increase of the total specific power consumption of the dual stage PRO specific. Therefore, it is desirable to apply high $Q_{\text{ds-in}}/Q_{\text{f-in}}$ ratios in the PRO process to enhance the process performance. However, this would increase the pretreatment cost of seawater draw solution. Finally, the results showed that the PA–CTA system requires less membrane elements than the CTA–CTA system. As a result, the cost of PA–CTA system would be cheaper than the CTA–CTA system. The footprint of PA–CTA plant would also be less than that of the CTA–CTA system.

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