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# Viability of integrating forward osmosis (FO) as pretreatment for existing MSF desalting unit

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## ABSTRACT

The feed to all Recirculation-multi-stage flash (R-MSF) desalting units in Qatar is pretreated with high temperature additive, which limits its top brine temperature (TBT) to 110˚C. The daily capacity of these R-MSF units is about one Million cubic meters ($Mm^3/d$). These units should continue their operation through their life time, (20–30 years). The capacity and performance of these units can be enhanced if Forward osmosis (FO) membrane system is used as pretreatment. The FO membranes can remove the scale constituents in the feed water and allows rising the TBT up to 135˚C, and thus increases the capacity of these units. The viability of using FO as feed water pretreatment to an existing operating R-MSF unit is discussed in this paper. A suggested arrangement to use FO as MSF feed pretreatment is presented. Part of the cooling seawater leaving the heat rejection is directed to the FO pretreatment unit as feed solution (FS). The flashing brine leaving the last stage with the maximum brine salt concentration (about 1.5 times of the seawater feed) is directed to the FO unit as draw solution (DS) that absorbs an amount permeate water (D) from the FS while the $Ca^{2+}$, $CO_3^-$, $Mg^{2+}$, and $SO_4^{2-}$ are rejected. The diluted DS enters the last stage condenser of the heat recovery section. This arrangement allows to increase the TBT and thus the unit distillate capacity. The potential of calcium sulfate deposite index inside MSF condenser tubes is calculated at different TBTs and different FO recovery ratio. The calculations show that Skillman index in reference MSF (operates at TBT = 110˚C) plant is greater than one which indicates the potential of scale deposit formation; however, antiscalant is used to disperse the crystalized scale. The simulation results showed that the potential of scale deposit is decreased as the FO recovery ratio increases due to the increase in divalent ions removal. The MSF can operate at TBT = 135˚C safely without calcium sulfate scale at FO recovery ratio of 40%. On the other hand, the increase in the FO recovery ratio, will reduce the osmotic pressure difference across the membrane which requires higher membrane surface area. The impact of feed salinity is investigated, and the results showed that lowering the feed salinity gives

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better performance of FO unit. The cost of the FO membranes is a major factor in applying the suggested use of the FO as pretreatment for the R-MSF unit. The decision to apply this method depends on the availability and reasonable cost of the FO membranes.

*Keywords:* Desalination; MSF; FO; Scale formation

## 1. Introduction

The Gulf Co-operation Countries (GCC) are using the main seawater thermal desalination systems known as Recirculation-multi-stage flash type (R-MSF) and Multi-effect-Thermal vapor compression (ME-TVC) systems. These systems are much more energy intensive and produce Desalted seawater (DW) at much higher cost compared with the world mostly used Seawater reverse osmosis (SWRO) system. However, the R-MSF and ME-TVC are still the main desalting methods used in Qatar, and all GCC, and their operation will continue for many years to come to the end of their expected life (20–30 years).

The capacity of the R-MSF units in Qatar is about one million cubic meters per day (Mm$^3$/d), with total installed capacity of 1.092 Mm$^3$/d, and 0.976 Mm$^3$/d on line, Table 1a. The online capacity of ME-TVC is 0.318 Mm$^3$/d, Table 1b. Similarly, the capacity of the R-MSF plant in GCC is 17.433 Mm$^3$/d, including 6.16 Mm$^3$/d in SA, 6.647 Mm$^3$/d in UAE, and 2.288 Mm$^3$/d in Kuwait [1].

The R-MSF system is the predominantly used desalting method in the GCC since its start in 1960 in Kuwait. It is known by its high reliability and maturity in design, operation, material selection, and maintenance; besides having the highest unit capacity of all desalting systems. The unit capacity reached 20-million imperial gallons per day (MIGD) in Ras AlKhair, SA. The main disadvantage of the MSF process is its high-consumed specific energy, about 265 MJ/m$^3$ of thermal energy, provided in the form of heating steam at 100–130℃ temperature range, and about 4 kWh/m$^3$ pumping energy. The Top Brine Temperature (TBT) in the MSF system is limited by the feed water pretreatment method used mainly to avoid calcium sulfate ($CaSO_4$), calcium carbonate ($CaCO_3$), and magnesium hydroxide Mg $(OH)_2$ scales deposition. The TBT is limited by 110℃ for high-temperature additives pretreatments methods.

The capacity of an R-MSF unit is affected mainly by the TBT, and other factors, but to less extent, such that the number of stages $n$, and specific heat transfer area, $A_h/D$. The number of stages $n$ is limited by the flashing temperature range of $(\mathrm{TBT} - T_n)$, and the temperature difference per stage ($\Delta T$), where $T_n$ is the brine temperature in the last stage, and

Table 1a
Capacity of Qatar's installed MSF units [1]

| Location | Capacity, m$^3$/d | Commission date | Plant's status | |
|---|---|---|---|---|
| Ras Laffan | 272,760 | 2006 | Online | CPDP |
| Ras Laffan | 182,000 | 2004 | Online | CPDP |
| Ras Abu Fontas | 150,000 | 1997 | Online | CPDP |
| Ras Abu Fontas | 136,380 | 2008 | Online | CPDP |
| Ras Abu Fontas | 90,920 | 1983 | Presumed online | CPDP |
| Ras Abu Fontas | 90,000 | 1978 | Presumed offline | CPDP |
| Ras Abu Fontas | 45,460 | 1992 | Presumed online | CPDP |
| Ras Abu Fontas | 45,000 | 1977 | Presumed offline | CPDP |
| Doha | 22,800 | 1973 | Presumed offline | Stand alone |
| Doha | 22,800 | 1978 | Presumed offline | Stand alone |
| Abu Aboud | 9,520 | 1968 | (Decommissioned) | Stand alone |
| Abu Aboud | 6,800 | 1963 | (Decommissioned) | Stand alone |
| Qatar | 5,440 | 1979 | Presumed offline | Stand alone |
| Umm Said | 2,900 | 1980 | Presumed online | Stand alone |
| Ras Laffan | 2,880 | 1998 | Online | Stand alone |
| Qatar | 2,640 | 1996 | Online | Stand alone |
| Umm Said | 2,500 | 1977 | Presumed offline | Stand alone |
| Doha | 1,362 | 1959 | Presumed offline | Stand alone |
| Doha | 227 | 1968 | Presumed offline | Stand alone |
| Doha | 100 | 1975 | Presumed offline | Stand alone |

Table 1b
Capacity of Qatar's installed thermal vapor desalting units [1]

| Location | Capacity, $m^3/d$ | Commissioned date | Plant's status | |
|---|---|---|---|---|
| Ras Laffan C | 286,400 | 2011 | Online | DPDP |
| Dukhan | 9,160 | 1994 | Presumed online | Stand-alone |
| Ras Laffan Pearl Gas to Liquids | 7,200 | 2008 | Online | Stand-alone |
| Mesaieed Industrial City MED | 5,760 | 2012 | Online | Stand-alone |
| Ras Laffan | 3,500 | 1995 | Presumed online | Stand-alone |
| Umm Said | 2,040 | 1996 | Online | Stand-alone |
| Qatar | 1,248 | 2002 | Online | Stand-alone |
| Ras Laffan | 1,200 | 1996 | Online | Stand-alone |
| PMP | 1,167 | 2013 | Online | |
| Umm Bab | 600 | 1996 | Online | Stand-alone |
| Umm Bab | 150 | 1993 | Presumed online | Stand-alone |
| Total Capacity | 318,425 | | | |

$\Delta T = (\text{TBT} - T_n)/n$. The minimum $\Delta T$ is about 2°C after considering the temperature losses in the stages, including the boiling point elevation (BPE).

Typical performance ratio (PR), or gain ratio (GR) in all MSF plants in SA is 7.8–8.7, and all Kuwait plants have a PR number of 8. The GR is defined as the distillate output ($D$) divided by the heating steam ($S$). Meanwhile, the PR is defined by $(Q/2{,}330)/D$, where $Q$ is the heat added to the thermal unit. Thus, the only way to increase the capacity of distillate output $D$ of any MSF unit, is to raise the TBT, or $(\text{TBT} - T_n)$.

To enable raising the TBT of MSF units, Nanofiltration (NF) was suggested as pretreatment method to remove some (or all) of the scale constituents such as sulfate, calcium, carbonate, and magnesium [2–5]. This method is technically feasible, but was not economically justified. The NF was also suggested to remove the scale constituents (totally or partially) to raise the recovery ratio (permeate to feed) in SWRO system.

Another method to remove the scale constituents is to use Forward osmosis (FO) as pretreatment method. The FO is a membrane process, similar to Reverse osmosis (RO) that can permeate almost pure water from saline water (e.g. seawater), called feed solution (FS) to a draw solution (DS) through semipermeable membrane due to the higher water chemical potential of the FS, $\mu_F$, compared to that of the DS, $\mu_D$. The DS (on one side of the FO membrane) has higher osmotic pressure $\pi_D$ than that of the FS (on the other side), $\pi_F$. The FO driving force is the water chemical potential ($\Delta\mu_w = \mu_F - \mu_D$) difference between the FS and the DS. The result of $\mu_F > \mu_D$ induces net water flow from the FS to the DS without applying pressure on the saline water such as in RO system.

The FO was also suggested as feed pretreatment for seawater reverse osmosis (SWRO) process before entering the SWRO membranes [6]. The water permeated through the FO membrane from the FS (e.g. seawater) to the DS can be separated from the diluted DS as water (product) plus concentrated DS either by membrane methods, or by thermal method [7].

Altae et al. [8,9], were the first to suggest the use of the FO process for pretreating the seawater (SW) feed to the thermal desalination plant by removing its scale constituent ions. They investigated the concept of hybrid of FO-MSF and FO-MED processes and showed the feasible application of FO as pretreatment method for thermal desalination processes under the TBT = 111°C for MSF and TBT = 65°C for MED. Simulation results showed the success of FO to reduce the scale ion in the SW feed to the MSF desalination plant [8]. The scale potential and scale precipitation thickness of $CaCO_3$ is estimated, theoretically, in MED system at a TBT = 65°C and under FO recovery ratio varied from 20 to 32%. The results showed the potential reduction in the scale thickness at higher FO recovery ratio [9].

The effectiveness of FO pretreatment process in removing divalent ions from feed SW solution to MSF at elevated temperatures was simulated [9], and showed that the water and salt fluxes across the FO membrane increased by increasing the SW salinity. However, for given SW salinity, the water and salt flux across the FO membrane decreased by increasing the FO recovery rate. It was found that the concentration of Ca, Mg, and $SO_4$ ions increased by increasing the operating temperature in the thermal plant, but decreased by increasing the recovery rate of the FO pretreatment [9].

The effects of temperature level and temperature difference across the FO membranes on the FO performance were investigated [10]. The results indicate an average increase in water flux up to 1.2% for every degree of temperature rise from 25 to 35˚C, and up to 2.3% for every degree of temperature rise from 25 to 45˚C. Providing a temperature difference by raising the temperature of the DS also enhances the water flux significantly, although it was lower than FO process operated at isothermal conditions. However, elevating only the temperature of F solution does not significantly improve the water flux although it was higher than the FO process operated at 25˚C.

Thus, in the present work a novel configuration of integrating the FO to MSF is presented to study the effect of temperature variation. A real operating MSF of 7 MIGD unit is chosen as a case study to show how it can be retrofitted in order to use FO as pretreatment method. Parameters affecting on the osmotic pressure difference of FO are presented.

## 2. Reference R-MSF unit to be retrofitted with FO as pretreatment

A Doha West R-MSF, as shown in Fig. 1, unit operating in Kuwait is chosen here as a reference unit to show how it can be retrofitted to include FO as pretreatment. This unit has 24 stages, 3 in HJS, and 21 in HRS. It can be operated at TBT = 90.5˚C to produce 6 MIGD, and at TBT = 110˚C to produce 7.2 MIGD. The actual operating data show that when the unit is operated with TBT = 110˚C, the total distillate (D) is 402 kg/s, and recirculation stream ($R$) flow rate is 3951.18 kg/s, and thus $R/D = 9.84$.

Fig. 1 shows that the recirculation stream $R$, at temperature $T_n$, enters to the condenser of heat recovery section (HRS) last stage (stage number $n - j$), and is successively heated as it flows in the HRS condenser tube bundles from the stages ($n - j$) to the first stage, and leaves this stage at temperature $t_1$. The $R$ stream enters the brine heater (BH), and leaves after being heated from $t_1$ to $T_0$ (TBT) by condensing the heating supply steam ($S$). The stream $R$ at pressure $P_0$ and $T_0$, called now the flashing brine, enters the bottom of the first stage, kept at pressure $P_1 < P_0$ and saturation temperature $T_{v1} < T_0$. The flashing brine temperature $T_0$ is spontaneously decreased in the stage to $T_1$ by flashing part of the flashing stream to attain equilibrium. It is noticed that, the vapor temperature, $T_{v1} = T_1 - \text{BPE} - \text{NEA}$, where BPE is the boiling point elevation and NEA is the non-equilibrium allowance. The flashed part of $R$ in the first stage is this stage distillate, and is equal to $D_1$.

The suggested arrangement to use FO as MSF feed pretreatment is shown in Fig. 2. Cooling seawater $M_c$ entering and leaving the heat rejection section (HJS) are kept the same as in the reference plant, using same SW cooling water pump. The feed seawater (part of $M_c$) having the SW salinity $X_f$ is directed to the new FO pretreatment unit (fully or partially) as feed solution (FS). The flashing brine ended in the last stage ($R - D$) with the maximum brine salt concentration $X_b$ (about 1.5 times $X_f$ or 70 g/L) is directed to the FO unit as draw solution (DS) that absorbs an amount $D$ from the FS entering with seawater salt concentration $X_f$. Due to the concentration difference (and thus the osmotic pressure difference), water at rate of $D$ is permeated from the FS to the DS, while the $Ca^{2+}$, $CO_3^-$, $Mg^{2+}$, and $SO_4^{2-}$ are rejected. The DS is diluted from $X_b$ to $X_r$ (salt concentration of the recirculation stream) that enters the HRS last stage condenser. While the capacity of the recirculation stream $R$ pump would be the same as the original unit, their delivering outlet pressure may need to increase to account for $R$ stream increasing TBT and thus pressure as it enters the first stage. This can be easily achieved by

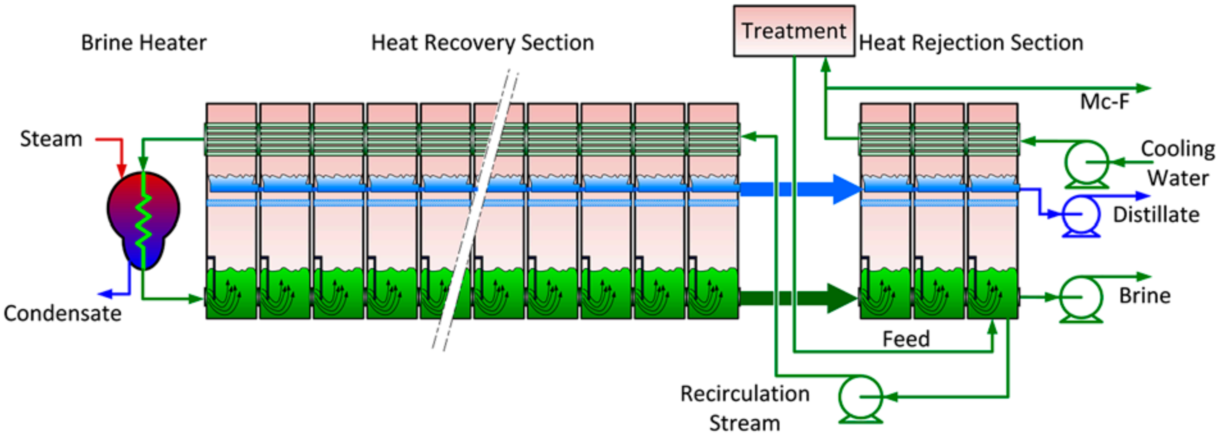

Fig. 1. Schematic diagram of recirculation-multi-stage flash (R-MSF) desalting unit.

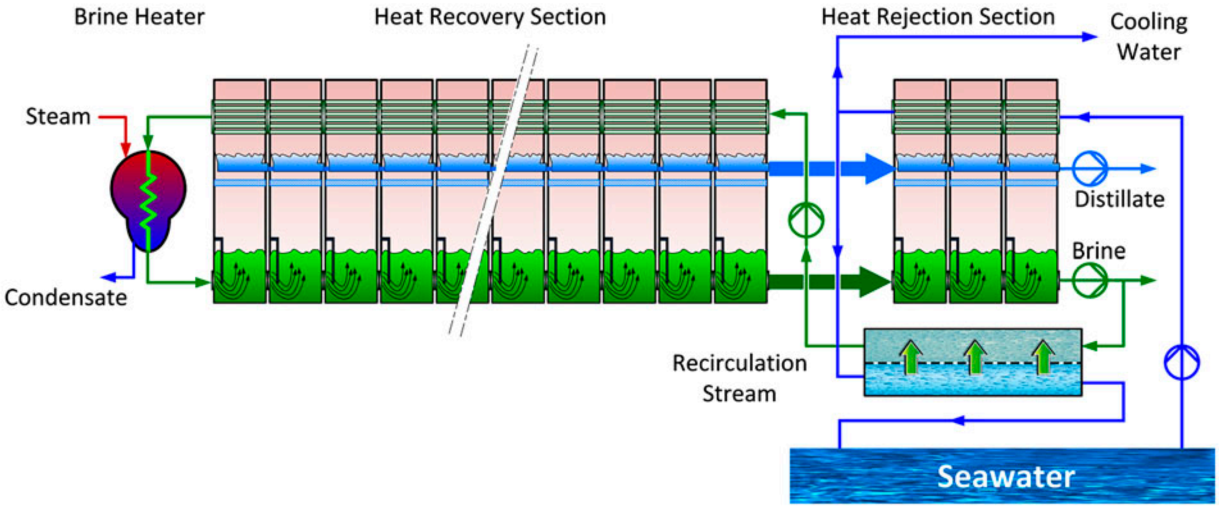

Fig. 2. The suggested arrangement of added FO as pretreatment to the reference MSF unit.

enlarging the motor driving the *R*-pump. Similarly, the seawater cooling pump would have the same capacity of the original unit, but with higher delivering pressure to overcome the pressure drop encountered in the FO feed solution side.

The feature of the suggested arrangement is utilizing the FO membrane at isothermal temperature, higher than that of SW temperature. The cooling SW leaves the HJS at $T_n \cong 40^\circ\text{C}$ would have higher water flux through the FO membranes than that suggested by Ref. [8] that use incoming cooling SW at 35˚C as FS to the FO system. This arrangement will not be affected with winter operation while seawater drop to below 20˚C.

## 3. FO membranes specifications and availability

Commercial FO membrane manufacturers include HTI, Oasys Water, Porifera, and Aquaporin. Large established RO membrane manufacturers, Hydranautics appears to have taken interest in FO through its collaboration with StatKraft, a company interested in generating power from osmotic pressure. Also, Toray, another RO membranes manufacturer announced it is developing high flux osmotic membranes too. HTI manufactures flat sheet FO membranes using cellulose acetates as well as a thin film composite (TFC) membrane of polysulphone and other polymer chemistry, and produces each membrane with various support structures to meet application requirements. HTI fabricates these membranes into various sizes of spiral membrane elements, and then engineers, fabricates, and delivers integrated FO system equipment.

The FO process permeability coefficient, as given by FO manufacturing company HTI, has a water permeability coefficient 1.125 $\text{L/m}^2\,\text{h bar}$ and NaCl rejection rate higher than 95%. Other data on three types of FO membranes are given in Table 2. Coday [11], shows through experimental work that ion such as $SO_4$ rejection in three types FO membranes is almost 100%.

A rough estimation for may be obtained by assuming that the osmotic pressure for each 1,000 ppm of Total Dissolved Solids (TDS) is equal to about 11 psi (or 0.77 bar). An example to show how to calculate the concentrations and osmotic pressure difference across the FO unit is shown in Fig. 3. Using simplified mode as shown in the appendix Eqs. (15–17), and assuming FO has 0.25 recovery ratio, and MSF has a $R/D = 6.5$ (for TBT close to 140˚C). The feed solution salt concentration would be equal to 56 g/L and the osmotic pressure would be 43.12 bar.

For 10 bar average osmotic pressure difference, and for $D = 402$ L/s (1,447,200 L/h), the required FO membrane areas is 128,640 $\text{m}^2$. The cost of the FO membranes per unit area is the main factor that decides the ratio of feed solution to the FO cell to that feed solution of the MSF unit, and economic viability of using the FO as pretreatment for the MSF.

## 4. Results and discussion

At different FO recovery ratio, the reduction in the $Ca^+$ ions in the MSF feed is calculated compared to the reference MSF process which operated without FO process. The potential of $CaSO_4$ scale formation in the

Table 2
Membrane physical and chemical properties [11]

| | Unit | CTA | TFC1 | TFC2 |
|---|---|---|---|---|
| Pure water permeability (A) | L/m$^2$-h-bar | 0.55 | 4.72 | 1.63 |
| Salt permeability (B) | m/s | $4.8 \times 10^{-8}$ | $1.2 \times 10^{-7}$ | $8.3 \times 10^{-8}$ |
| Structural parameter | µm | 463 | 365 | 690 |
| Zeta potential, active layer | mV[a] | −34.9 | −42.5 | −38.6 |
| Contact angle | ° | 63.7 ± 6.8 | 67.8 ± 11.8 | 27.7 ± 10.4 |
| Average water flux | L/m$^2$/h | 9.8 ± 0.1 | 27.1 ± 3.01 | 12.4 ± 1.0 |
| Average reverse NaCl flux | mmol/m$^2$/h[b] | 81.4 ± 9.6 | 217 ± 50 | 214 ± 55 |

[a] At pH 7.0.

[b] Virgin membranes at 20°C, 1 M NaCl draw solution, and deionized feed water over three month period.

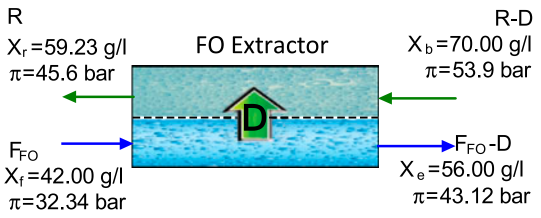

Fig. 3. Osmotic pressure at the inlet and exit of the FO cell for $R/D = 8$ and $F/D = 4$.

MSF feed after dilution is estimated at different TBT (115–135°C) using Skillman index [12].

Skillman et al. [12] developed a simple sulfate solubility index for estimating $CaSO_4$ deposit inside the MSF condenser tubes. Sulfate scale potential is ratio between the actual concentration, $[i]_{\text{actual}}$, of either calcium (Ca) or sulfate ($SO_4$) and its theoretical or equilibrium concentration whichever is the limiting species:

$$\text{SI} = \frac{[i]_{\text{actual}}}{\left(\left(\sqrt{x^2 - 4K_{\text{sp}}}\right) - x\right).10^3} \tag{1}$$

where $x$ is the absolute value of the excess common-ion concentration of calcium and sulfate ions:

$$x = \left|2.5\left[\text{Ca}^{2+}\right] - 1.04\left[\text{SO}_4^{2-}\right]\right|.10^5 \tag{2}$$

The solubility product constant ($K_{\text{sp}}$) can be determined from the equation of solubility, in g/L, established by Skillman [12]:

$$K_{\text{sp}} = 1000 \times e^{\left(-171.9773 - 0.077993*(T) + \frac{2903.293}{T} + 71.6*\text{LOG}(\text{T})\right)} \tag{3}$$

where $T$ is the absolute temperature of the feed water. The SI is calculated for the recycled brine (diluted draw solution) at the brine heater exit (TBT).

Fig. 4 shows the Skillman Index (SI) at different TBTs and variation of FO recovery ratio. The Skillman index of $CaSO_4$ solubility in case of reference MSF plant without FO operates at TBT = 111°C is calculated as 1.33. Since the calculated SI of traditional MSF is greater than one, the precipitate of $CaSO_4$ can occurs. However, in practical MSF plant antiscalant is used to disperse the crystalized scale. The SI = 1.33 is considered as reference of comparison and the value above 1.33 indicates scale formation while the lower value indicate for safe operation. As shown in Fig. 1, the Skillman index at different TBTs decreases as the FO recovery ratio increases. This is due to the increase in the removal of divalent ions. Fig. 5 shows the increase in Skillman index as the TBT increases. The MSF can operate safely without scale problems at TBT = 135°C and 40% FO recovery ratio, at TBT = 130°C and 35% FO recovery ratio, and at TBT = 125°C and 30% FO recovery ratio, and at TBT = 120°C and 25% FO recovery ratio, and finally at TBT = 115°C and 20% FO recovery ratio of 20%. The use of the simplified model presented in the appendix, Table 3 shows that $R/D$ would range from 10.5 to 6.5 as the TBT increasing from 110 to 140°C by applying the FO as pretreatment.

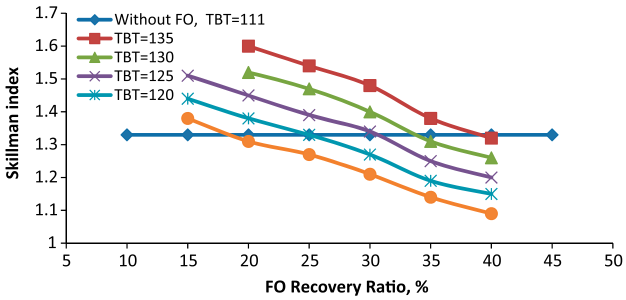

Fig. 4. Influence of FO on sulfate scale potential in BR-MSF plant.

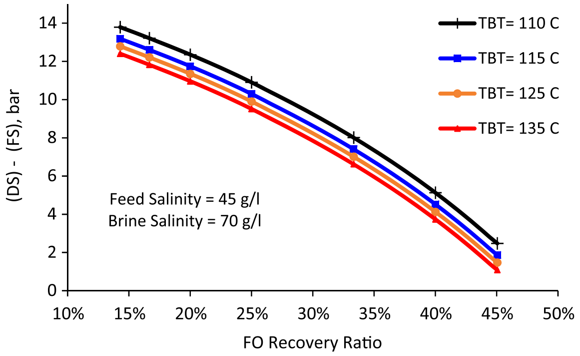

Fig. 5. FO average osmotic pressure difference variation with recovery ratio.

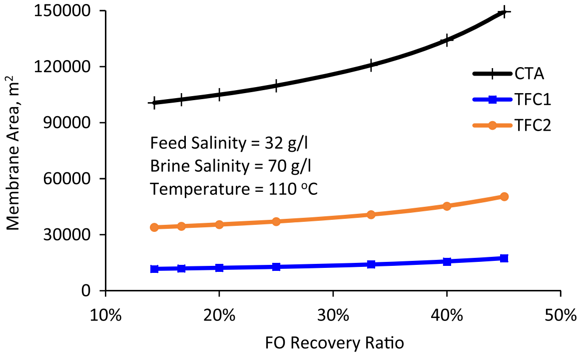

Fig. 6. FO membrane area vs. recovery ratio for membranes of Table 2.

The chemical potential difference across the FO membrane (and thus the osmotic pressure difference) represents the motive for water transfer across the FO membrane. Increasing this difference for specific permeated flow would reduce the required FO membranes area. In the present suggestion, the DS of the FO unit is the brine rejected from the MSF unit having (R-D) flow rate at the inlet and $R$ at the outlet. The range of the $R/D$ ratio is 8.48 at 115℃ to 6.8 at 135℃, when the FO is applied as MSF unit pretreatment. Meanwhile, the DS is part of $M_c$ leaving the HJS. The ratio $M_c/D$ is the range of 5–7 and is function of SW temperature. The FO recovery ratio is the permeated flow $D$ divided by the FS at the inlet and can vary from 15 to 40%. Fig. 5 shows the effect of FO recovery on the difference between the average osmotic pressure for inlet SW salinity of 45 g/L. The osmotic pressure is calculated at different values for $R/D$, while the maximum brine salinity is fixed at 70 g/L. As the FO recovery increases, the osmotic pressure difference decreases due to significant dilution of FO membrane at higher recovery ratio. At high FO recovery ratio, more permeates cross the FO membrane. Consequently, the average concentration at the DS side will decrease while the average concentration in the FS side will increase. As a result, the difference between osmotic pressure on both sides will decrease.

Also, at a fixed recovery ratio, as the TBT increase, the osmotic pressure difference decreases. This is due to increase in permeate flow rate compared to brine recycle ($R$), which dilutes the DS, and reduces its average osmotic pressure. So working at high TBT = 135℃ is technically possible, but it requires higher FO recovery ratio in order to avoid scale deposit formation. However, higher recovery ratio requires larger membrane surface area as shown in Fig. 6.

Fig. 6 shows the variation of the membrane surface area of three types of FO membrane at different FO recovery ratio. The TFC1 gives the lowest membrane area due to high water permeability ($A$ = 4.7 lmh/bar) as shown in Table 2. The CTA membrane gives the higher membrane area due to lower permeability ($A$ = 0.55 lmh/bar).

Fig. 7 shows the effect of FO recovery ratio at different TBT's. At fixed recovery ratio, the membrane area increases as the TBT increases due to lower osmotic pressure difference.

Table 3
Relationship between TBT and $R/D$

| $T_0$, ℃ | $T_n$, ℃ | $\Delta T$, ℃ | $t_1$, ℃ | $T_0 - t_1$, ℃ | $R/D$ | $Q/D$, MJ/$m^3$ | $D$, MIGD |
|---|---|---|---|---|---|---|---|
| 90 | 40 | 2.08 | 83.75 | 6.25 | 12.47 | 304.1 | 6.0 |
| 110 | 40 | 2.50 | 92.50 | 7.50 | 10.48 | 306.5 | 7.2 |
| 115 | 40 | 3.13 | 105.63 | 9.38 | 8.48 | 310.2 | 8.9 |
| 120 | 40 | 3.33 | 110.00 | 10.00 | 7.98 | 311.4 | 9.4 |
| 125 | 40 | 3.54 | 114.38 | 10.63 | 7.54 | 312.6 | 10.0 |
| 130 | 40 | 3.75 | 118.75 | 11.25 | 7.15 | 313.8 | 10.5 |
| 135 | 40 | 3.96 | 123.13 | 11.88 | 6.80 | 315.0 | 11.1 |

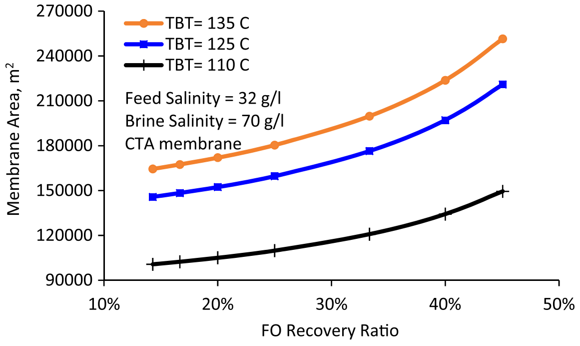

Fig. 7. FO membrane area variation at different TBTs.

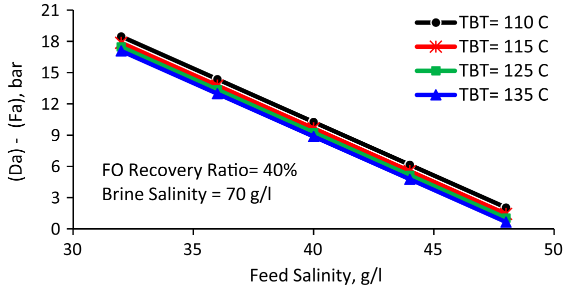

Fig. 8. Relationship between average osmotic pressure difference and feed salinity.

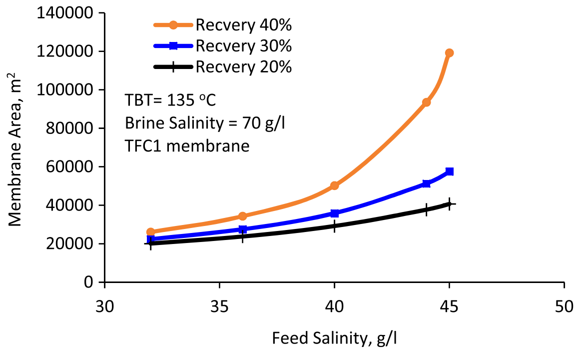

Fig. 9. Membrane area vs. seawater feed salinity variation.

The seawater salinity varies in Qatar's East Coast from 42 g/L in winter to 45 g/L in summer. The seawater salinity West of Qatar can easily reach 57 g/L, which is quite high and reduces the osmotic pressure difference across the FO membranes. For this reason it is not suitable to use FO as pretreatment there, as indicated by Fig. 4. Fig. 8 also shows the effect of increaseing the FS salinity on the average osmotic pressure difference at different values of TBT. Increasing FS salinity decreases the average osmotic pressure difference, consequently, and thus reduces the permeate flow across the FO membrane. In case of high feed salinty, the recirculated draw solution will be slightly diluted which will reduce the potential of increasing TBT of MSF.

The impact of the SW feed salinity on the FO membrane area is illustrated in Fig. 9. At certain recovery ratio, the FO membrane area increases according to the increase in FS salinity, while fixing the DS salinity at 70 g/L. This is due to lower difference of the concetration across the FO membrane acocrdingly lower osmotic pressure difference. Fig. 9 shows also, at certain FS salinity, the membrane surface area increases as the recovery ratio increses.

The results concludes the feasibility of increasing the MSF unit capacity using FO system as its pretreatment system. The MSF unit cpacity increase is realized by increasing its TBT. In case of high SW salinty, the FO recovery ratio would be reduced. The compromise among the unit capacity increase and the addinonal capex due to FO membrane deployment has to be reliazed.

## 5. Conclusion

A suggested arrangement to increase the TBT of MSf unit, and thus increase its output capacity using FO membrane system as pretreatment is studied. The potential of calcium sulfate scale deposit inside the MSF condenser tubes is calculated at different TBT's and different FO recovery ratio. The Skillman index in reference MSF (operates at TBT = 110˚C) plant is greater than one which indicates the potential of scale deposit formation; however, antiscalant is used to disperse the crystalized scale. The simulation results showed that the Skillman index decreases as the FO recovery ratio increases due to the increase in removal of divalent ions. The MSF can operate at TBT = 135˚C safely without scale (Calcium Sulfate) at FO recovery ratio of 40%. On the other hand, the increase in the FO recovery ratio, will reduce the osmotic pressure difference across the membrane which requires higher membrane surface area. The impact of feed SW salinity is investigated, and the results showed that low-feed salinity requires less FO membrane area. The cost of the FO membranes is a major factor in applying the suggestion of using the FO as pretreatment for the R-MSF unit. A compromise between the unit capacity increase and the addinonal capital cost due to FO membrane deployment has to be exploited according to the membrane development and availaibility in commercial scale.

## Nomenclature

$C$ — specific heat capacity, kJ/kg°C
$D$ — distillate flow rate, kg/s or MIGD
$F$ — feed
$L$ — latent heat, kJ/kg
$M$ — flow rate, kg/s
$Q$ — heat transfer, W
$R$ — stream flow rate enters the brine heater or leaving FO membrane from draw solution side, kg/s
$S$ — heating steam flow rate, kg/s SW seawater
$T$ — temperature, °C
$T_s$ — heating steam temperature, °C
$T_0$ — top brine temperature, °C
$X$ — stream salt concentration, ppm
$\mu$ — chemical potential
$\pi$ — osmotic pressure, bar
DS — draw solution
TBT — top brine temperature

### *Subscripts*

$c$ — cooling water
$D$ — draw solution
Da — average of draw solution side
DE — draw solution exit
$F$ — feed
$f$ — feed
Fa — average of feed solution side
FE — feed exit
$l$ — lost
$n$ — last stage or number of stages
$w$ — water

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## Appendix A: The MSF process

The flashing process from the brine in the first stage gives:

$$D_1 = \; R\;C\;(T_0 - \;T_1)/L \tag{A1}$$

The flashing brine leaves the first stage at $T_1$ enters the second stage kept at saturation temperature $T_{v2} < T_1$. Its temperature is spontaneously decreased to $T_2$ by evaporating $D_2$ from the flashing brine $(R - D_1)$, and the distillate evaporated and condensed in the second stage (product of this stage $D_2$) is:

$$D_2 \; = (R - D_1)(T_1 - T_2)/L \tag{A2}$$

The process is repeated, say to stage $i$, and $D_i$ is equal to:

$$D_i = \; \left(R - \sum D_{(1-(i-1))}\right)(T_{i-1} - \;T_i)/L \tag{A3}$$

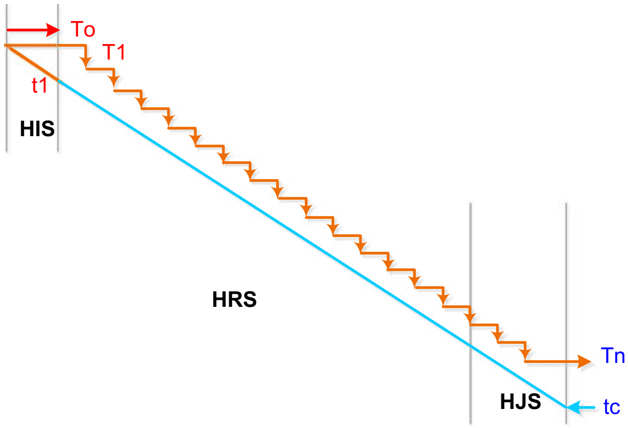

Fig. 1A. Temperature profile in the (R-MSF) desalting unit.

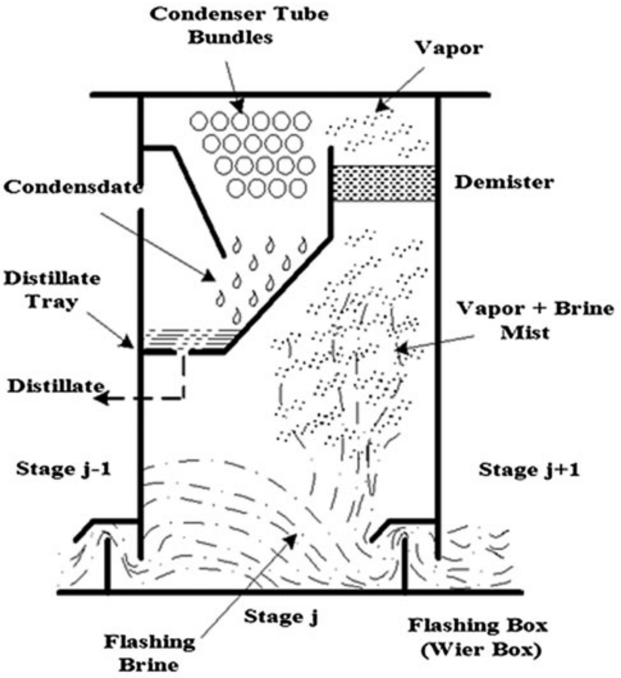

Fig. 2A. Schematic diagram of single stage.

The process is repeated to the last stage $n$, where the flashing brine flow rate becomes equal to $(R - D)$, at $T_n$. The flashed vapor (or distillate output) $D_i$ at a stage ($i$) is moved upward in the stage through demister to the condenser bundle located in the upper volume of stage. The vapor $D_i$ meets with the flashed vapor from the accumulated distillate moving from stage $(i - 1)$ to stage $i$ and both are condensed and heat the stream $R$ flowing inside the HRS condenser tubes (or cooling seawater in the HJS). The cooling water stream $M_c$ enters the condenser tubes of the last HJS stage $n$ at seawater temperature $t_c$, and leaves at stage $(n - j + 1)$ at temperature $T_n$. The leaving $M_c$ is divided to $(M_c - F)$ which is rejected back to sea, and feed $F$ is to be treated before joining as a part of the stream $R$. The treated $F$ enters the last stage n, and is mixed with finished flashing brine $(R - D)$, and the mixed stream $(R - D + F)$ leaves the last stage by recirculation pump. Part of the exit stream equal to $(F - D)$ is rejected back to sea by throttling without using pump and the balance $R$ enters the last HRS stage condenser. This arrangement avoids the use of blow-down pump, and feed delivery pump to the last stage. However, it wastes chemicals as part of pretreated feed is rejected to sea, before being used for evaporation.

The mass of salt entered with the flashing brine in the first stage $RX_r$ is equal that ended with the flashing brine in the last stage $(R - D)\ X_b$, where $X_r$ and $X_b$ are the salt concentrations of the recirculation flow $R$ at its entry to the first stage and of the flashing brine at the last stage $n$. The ratio of $R/D$ in terms of $X_r$ and $X_b$ can be calculated in both cases of including $F$ with $R$ as follows:

$$RX_r = (R - D)X_b \tag{A4}$$

$$\frac{X_r}{X_b} = \frac{(R - D)}{R} = 1 - \frac{D}{R} \tag{A5}$$

$$\frac{R}{D} = \frac{X_b}{(X_b - X_r)} \tag{A6}$$

The feed to distillate ratio $F/D$ can be determined in terms of the feed salt concentration $X_f$ (depends on locality) and the maximum allowable brine salt concentration $X_b$ (design parameter usually taken as less than 70,000 ppm) as follows: when the feed $F$ is added to the brine in the last stage before rejecting the blow-down, the salt balance gives:

$$(R - D)X_b + FX_f = (R - D - F)X_r \tag{A7}$$

By applying Eq. (A3) into Eq. (A6) and rearrange it, then it gives:

$$FX_f = (F - D)X_r \tag{A8}$$

$$\frac{F}{D} = \frac{X_r}{(X_r - X_f)} = \frac{X_b(1 - D/R)}{X_b(1 - D/R) - X_f} \tag{A9}$$

$$\frac{F}{D} = \frac{X_b}{[X_b - X_f R/(R - D)]} \tag{A10}$$

An energy balance of the desalting plant as whole gives:

$$\begin{aligned} R\ C\ (T_0 - t_1) + M_c C\ t_c = {} & (M_c - F)\ C\ T_n \\ & + (F - D)C\ T_n + D\ C\ T_{vn} \\ & + Q_l \end{aligned} \tag{A11}$$

The terms $RC(T_0 - t_1)$ represents the energy added to the BH and $M_c\ Ct_c$ is the energy conveyed to the system by the incoming cooling seawater $M_c$ at $t_c$. The term $(M_c - F)\ C\ T_n$ is the convective heat rejected by the partially rejected cooling seawater ($M_c - F$) *at* $T_n$ from the HJS. The term $(F - D)CT_n$ is the convective heat rejected by the brine blow-down $(F - D)$ *at* $T_n$ from the last stage to sea. The term $D\ C\ T_{vn}$ is the heat carried out by the product water $D$ *at* $T_{vn}$ from the last stage. The term $Q_l$ represents the heat loss from the plant. Since $T_{vn} = T_n - \delta T_L \approx T_n$, then:

$$R\,C\,(T_0 - t_1) = M_c\,C\,(T_n - t_c) + Q_l \tag{A12}$$

The relation between the $R/D$ ratio and the operating temperatures $T_0$, and $T_n$ is determined as follows. The flashing brine suffers successive temperature drops from its entry to the first stage at flow rate $R$ and temperature $T_0$ to its exit from the last stage ($n$ stage) at flow rate $(R - D)$ and temperature $T_n$. In stage $i$, the flashing brine temperature drop $\Delta T_i = T_{i-1} - T_i$, consequently, its heat loss $R_iC\Delta T_i$ is used to evaporate an amount of vapor $D_i$ in that stage such that:

$$R_i C\, D\, T_i = D_i L \tag{A13}$$

$R_i$ is the flashing brine flow rate to the stage $i$ and $D_i$ is the vapor generated from $R_i$ in that stage. The heat loss from the average flashing brine flow rate: $\bar{R} = R - D/2$ (between its flow rates $R$ and $R - D$ at the first stage inlet and last stage outlet) is equal to:

$$\sum_{i=1}^{n} R_i\ C\ \Delta T_i = \bar{R}\ C(T_0 - T_n) \tag{A14}$$

This heat is used to generate vapor $D$ in all stages, i.e. $(R - D/2)\ C(T_0 - T_n) = DL$, and

$$R/D = 0.5 + L/C(T_0 - T_n) \tag{A15}$$

The heat supplied to the BH ($Q_i$) is used to heat the recirculation $R$ from $t_1$ to $T_0$, or

$$Q_i = SLs = RC(T_0 - t_1) \tag{A16}$$

Using the expression of $R/D$, the gain ratio (GR) is given by:

$$D/S = \left(1 - \frac{D}{2R}\right)\frac{Ls}{L}\frac{(T_0 - T_n)}{(T_0 - t_1)} \tag{A17}$$

The existing MSF unit performance can be checked by calculating $R/D$;

$$\frac{R}{D} = 0.5 + \frac{L}{C(T_0 - T_n)}$$
$$= 0.5 + 2316/[3.89\ (110 - 43.55)] \approx 9.5$$

$$\frac{D}{S} = \frac{Ls}{L}\left(1 - \frac{D}{2R}\right)\frac{(T_0 - T_n)}{(T_0 - t_1)}$$
$$\approx \left(1 - \frac{1}{2 \times 9.84}\right)\left(\frac{(110 - 43.55)}{(110 - 102)}\right) \approx 7.9$$

It is noticed here that the GR = $D/S$ = 7.9 is less than the designed GR = 8 because the relatively high $T_n$ (43.55°C) as a result of high seawater temperature ($t_c$ = 32.22°C) in summer.

**Appendix B: The FO process**

The general equation describing water transport through the FO membrane, and pressure retarded osmosis (PRO) and RO is:

$$J_w = A(\sigma\,\Delta\pi - \Delta P) \tag{A18}$$

The term $J_w$ is the water flux across the membrane, $A$ is the water permeability constant of the membrane and is proportional to the water diffusion coefficient in the membrane $D_w$ and is inversely proportional to the membrane thickness ($t$), $\sigma$ presents the reflection coefficient, $\Delta\pi$ is differential osmotic pressure, $\Delta P$ is the applied hydraulic pressure, and the difference $(\sigma\,\Delta\pi - \Delta P)$ represents the effective driving force. The reflection coefficient $\sigma$ represents salt water coupling and can have values from 0 to 1. When $\sigma = 0$ the membrane shows no salt rejection, when $\sigma = 1$ the membrane shows total salt rejection. For RO membranes with high salt rejection, $\sigma$ approaches 1.

The relation of the draw and feed solutions at the inlet and exit from the FO cell can be expressed by:

$$(R/D{-}1)X_b = R/D\ X_r \tag{A19}$$

where $X_b$ and $X_r$ are the salt concentrations of the draw solution at the inlet and exit of the FO cell, respectively. The relation salinity of the feed solution at the inlet and exit of the FO cell can be expressed by:

$$X_f F_{FO}/D = X_{fe}(F_{FO}/D - 1) \tag{A20}$$

where $F_{FO}$ is the feed solution to the FO cell, $X_f$ is the seawater salinity, and $X_{fe}$ is the exit FO cell feed solution salinity.