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Techno-economic analysis of hybrid high performance MSF desalination plant with NF membrane

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ABSTRACT

Techno-economic analysis of a newly developed high performance multistage flash configuration with deaeration and brine mix (MSF-DM) is presented. The techno-economical analysis includes also the use of nanofiltration (NF) as a pretreatment method for MSF to increase its top brine temperature (TBT) to 130˚C. A mathematical model of NF membrane is developed and verified using Visual Design and Simulation program for typical operating NF unit (Umlluj, KSA). The techno-economic analysis of integrating NF pretreatment for the existing multistage flash-brine recirculation (MSF-BR) and newly developed MSF-DM configurations is performed. Integration of NF system to existing desalination plant (NF-MSF) and treatment of only 30% of make-up enable to increase the TBT up to 130˚C, the production can be increased to 19%. The cost analysis showed the unit product cost is 5.4% higher than that conventional MSF (at 110˚C) due to the additional capital cost of NF system. Integrating NF system to new configuration (NF-MSF-DM) desalination plant at the TBT = 130˚C, the gain output ratio could be as high as 16, i.e. double the convention MSF-BR. The new NF-MSF-DM configuration significantly reduces the unit's input thermal energy to suit the use of (the relatively expensive) solar energy as a desalination plant driver. On the other hand, the levelized water cost of NF-MSF-DM (at TBT = 130˚C) is 14% lower than conventional MSF (at 110˚C) at the current oil price 104 $/bbl.

Keywords: Desalination; MSF; NF; Cost analysis

1. Introduction

Multistage flash (MSF) technology has proven to be a mature thermal technology for large-scale capacity production with high quality desalinated water especially for high severe feedwater quality. The development of the MSF system resulted mainly from the accumulated experience obtained from the operation of these plants in GCC and MENA countries.

The first MSF of 0.5 MGD per unit evaporator was built in 1957 in Kuwait using the once-through MSF-OT configuration by the Westinghouse Company [1]. The design was modified according to the recommendation of the client, the Ministry of Electricity and water in Kuwait engineers for reliable operation. For some time the market was dominated with the once-

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through (MSF-OT) due to its simplicity and high thermodynamic efficiency. However, due to high oxygen and CO_2 gas liberation in addition to large amount of feedwater to be pretreated forced the market to shift to the brine recirculation configuration (MSF-BR). The first 19-stage 1 MGD MSF-BR plant was built by Weir Company in 1959 in Kuwait [1]. The developed specifications led to more reliable, easy to operate and maintain, and longer life units. Recently, the MSF evaporator production capacity was increased dramatically over years to reach 20 MGD in UAE and designs of 25–30 MIGD are available. The disadvantage of MSF-BR system is the higher brine concentration, which increases the potential for scale deposits on the heat transfer surfaces and the liquid boiling point elevation, thus penalizing the coefficient of heat transfer and the available condensing temperature difference, respectively.

Increasing the MSF unit production (for both new designs and operating units) can be carried out either by: (i) increasing the recirculating brine flow rate or (ii) increasing the flashing range. Increasing the recirculating brine flow rate is limited, however, by the available pumps capacity and the chamber load (flashing brine flow velocity). Increasing the flashing range (TBT-BBT) can be carried out by increasing top brine temperature (TBT), with hard scale solution, or reducing bottom brine temperature (BBT), lower heat sink temperature (naturally in fall/winter/spring or utilizing deep intake or cooling towers). Increasing TBT is the addressed parameter in this paper.

At high TBT, scale deposits of high seawater brine concentration present a real problem in MSF plants as it directly affects the heat transfer rates on the heating surface. The main scale forming constituents are calcium (Ca^{++}), magnesium (Mg^{++}), bicarbonate (HCO_3^-), and sulfate (SO_4^{2-}) ions. On heating, bicarbonate decomposes into carbonate CO_3^- which with Ca^{++} forms calcium carbonate ($CaCO_3$) that precipitates on the heat transfer surface (if saturation limits exceeded). At high temperature, magnesium hydroxide (MgOH) will also be formed. At higher temperature, >120˚C, non alkaline calcium sulfate ($CaSO_4$) precipitates if saturation limits also exceed, due to the inverse solubility. Formation of alkaline scale ($CaCO_3$ and MgOH) can be controlled by lowering pH (acid additives) or by antiscalant. Non alkaline (hard) scale (as $CaSO_4$) is only controlled, nowadays, by limiting TBT below 120˚C.

Increasing TBT with hard scale solution can be carried out by: (i) introducing high temperature antiscalant and (ii) Reducing hard scale ions to avoid it reaching saturation conditions. The first is not available yet, while the second is considered, in literature [2–7], through the use of nanofiltration (NF) membrane system for make-up feedwater pretreatment.

The application of NF in seawater desalination has gained significant attention in the desalination industry due to selective removal of divalent ions. SWCC, R and D team [2–4,8] carried out extensive experiments on MSF test pilot unit with NF as pretreatment. NF pressure was 24 bar and its recovery ratio ranged from 60 to 65%. The total concentration of the sulfate and calcium ions of the brine recycle at TBT of 130˚C and with a make-up entirely formed from NF permeate were below their solubility limits. This result indicated the possibility to operate the MSF plant safely and without any scaling problem at TBT at or higher than 130˚C. However, many questions about the addition of capital cost which might weigh the savings in operational cost still need a clear answer.

Awerbuch [5,6] tried partial pretreatment of MSF (25% of make-up) by NF. Earlier, test on evaporator of Layyah MSF plant (UAE) was successful to increase TBT from 105 to 110˚C and the production from 1,044.4 t/h (design value at TBT = 110˚C) to 1,253 t/h (19.9%). The maximum production was 1,260 t/h (20.6%) at 117˚C with product conductivity of 454 μS/cm^2. However, based on personal communication, the NF plant was shut down due to operational problem in the pretreatment section.

The NF originally is applied to reject electrolytes and obtain ultrapure water with high volume flux at low operating pressure, because most membranes have either positive or negative charge due to their compositions [7]. The NF membrane possesses molecular weight cutoff of about hundreds to a few thousands, which is intermediate between reverse osmosis (RO) membranes and ultrafiltration. The pore radii and fixed charge density of practical membranes were evaluated from permeation experiments of different neutral solutes of sodium chloride. The pore radii of these NF membranes were estimated to range from 0.4 to 0.8 nm [7].

The aim of this work is to present techno-economic analysis of a newly developed high performance MSF configuration with de-aeration and brine mix (MSF-DM). The techno-economical analysis includes the use of NF as a pretreatment method for MSF to increase its TBT to 130˚C. A mathematical model of NF membrane is developed and verified using Visual Design And Simulation (VDS) program for typical operating NF unit. The techno-economic analysis of adding NF pretreatment for the existing MSF-BR and newly developed MSF-DM configurations is performed.

2. Description of newly MSF-DM configuration

A modified MSF-DM configuration has been proposed as shown in Fig. 1. In this MSF-DM configuration, the heat rejection section is removed and the bottom part of the deaerator is utilized as a mixer where part of the last stage brine is mixed with deaerated make-up. The new configuration is half-way between brine recirculation MSF-BR and once-through MSF-OT will benefit from both techniques and overcome the limitation encountered through operation. The GOR of the MSF-DM configuration at the TBT = 110 could be as high as 12.

The MSF-DM design configuration is targeting high MSF GOR to be adopted by solar energy application (High GOR is also needed as the cost of energy is increasing). Since the capital cost of solar energy systems is expensive, it will be cost effective to develop high performance MSF to reduce the CAPEX of the solar energy systems. High performance MSF system requires a combination of more evaporating stages; more heat transfer surface area sequentially increases the MSF CAPEX. The increase of the MSF CAPEX could be balanced by the reduction of MSF OPEX and accordingly the CAPEX reduction of the solar energy system will be the main contribution to the developed system.

3. VDS for NF system

Fig. 2 illustrates the input and output parameters used for the mass and energy balance equations, Appendix A of the NF membrane [9–12].

The Umm-Lujj NF-RO plant [4] is considered as a case study to verify the mathematical model of the NF membrane as well as estimate the permeate con-

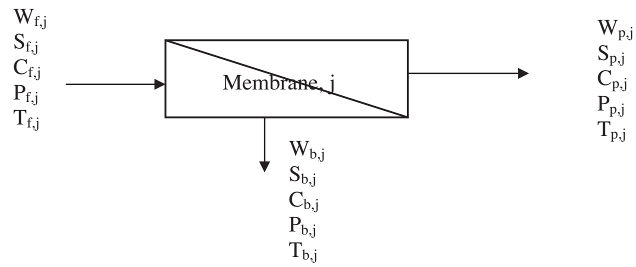

Fig. 2. Schematic diagram of the NF membrane streams.

stants A and the solute constant B. This plant consists of 27 pressure vessels and six NF elements per vessel. The feed characteristics are 360 m^3/h, temperature 32°C, and the salinity is 45.46 g/l. The applied feed pressure is 25 bar. The data of Umm-Lujj, shown in Table 1, are used as input data of VDS software as shown in Fig. 3.

The VDS simulates the Umm-Lujj plant of NF to estimate the permeate production and the exact value of the membrane constants A and B. After several runs, the membrane water permeability A of the considered NF membrane is determined as follows:

$$A = 5.8 \times 10^{-9}\ \mathrm{m^3/m^2\ s\ kPa} \tag{1}$$

The membrane salt permeability coefficient B is estimated as follows:

$$A = 9 \times 10^{-8} \tag{2}$$

Using the estimated values A and B, the VDS results are compared against the typical plant as shown in

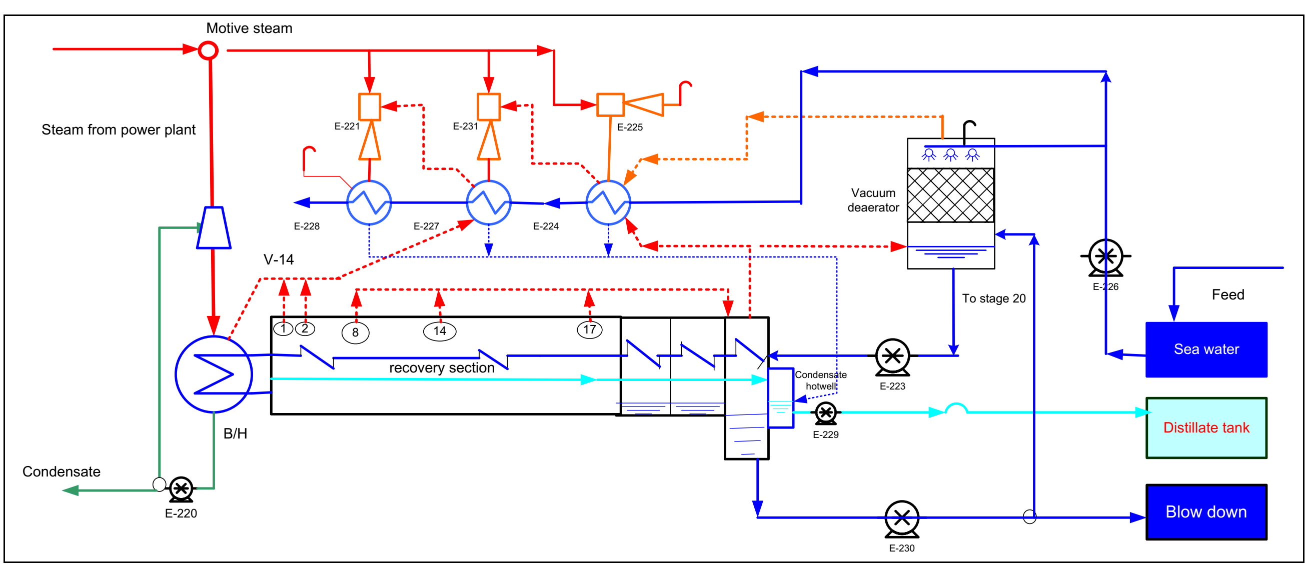

Fig. 1. The interface of the new MSF-DM for desalination plant.

Table 1
Comparison between VDS and Umm-Lujj typical results

Variable	VSP results	Umm-Lujj SWCC Exp.	% Error
Feed flow rate, m^3/h^*	360	360	–
Feed salinity, TDS, g/l*	45.46	45.46	–
Stages no.*	1	1	–
No. of pressure vessels*	27	27	–
Feed temperature, ℃*	32	32	–
Fouling factor*	0.95	NA	–
Feed pressure, bar*	25	25	–
Elements no. per vessel*	6	6	–
Permeate flow rate, m^3/h	245	234	4.7
Recovery ratio	0.68	0.65	4.6
Permeate salinity, TDS, mg/l	29.11	28.26	3

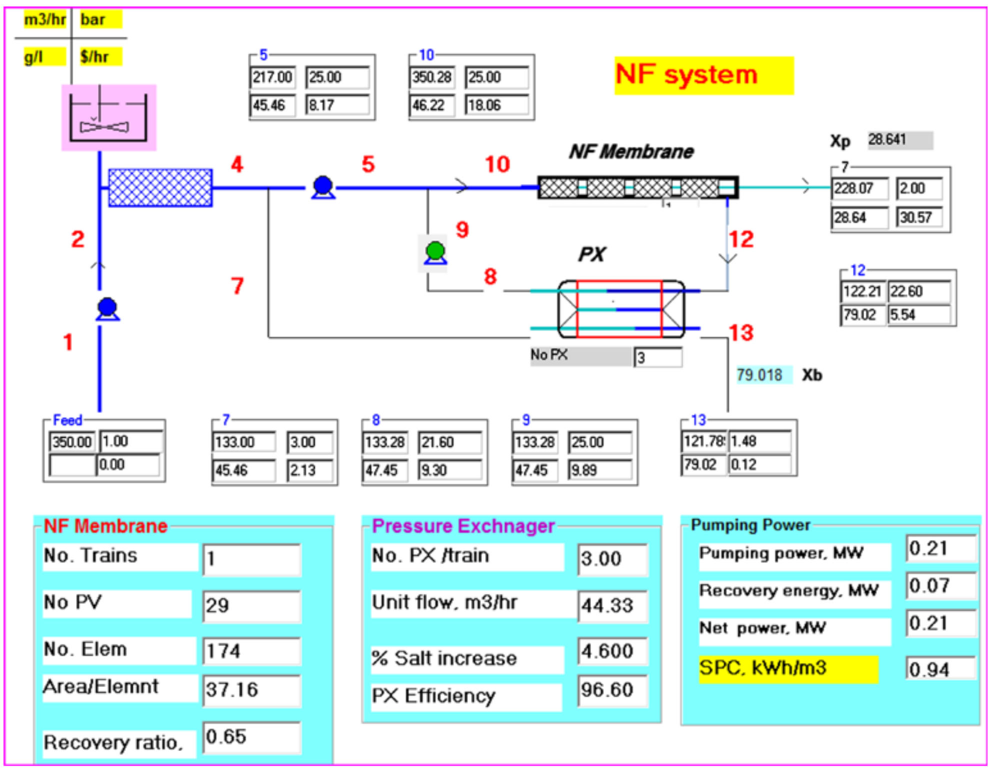

Fig. 3. VDS interface of NF system with pressure exchanger.

Table 1. The comparison results show a good agreement between the VDS results and the typical real plant.

4. Process design and techno-economical program (VDS)

The flexible and powerful tool ''Visual Design and Simulation program (VDS)'' is used to perform process and techno-economical calculations. VDS was developed for the design and simulation of different types and configurations of the desalination processes [9–12]. Typical desalination processes are simulated to show the wide scope and high capability of the developed package. The description of the VDS software, how to access and handle the package are presented in references [9–12]. In this work, the scope of VDS program will be extended to develop and build up NF system and new MSF configuration model. The NF system mathematical model will be verified using typical NF plant data.

The VDS performs process design calculations by specifying the heating steam operating conditions (pressure and temperature), the target capacity by evaporator (distillate rate per hour), TBT, feed seawater conditions (temperature and salinity), make-up flow rate, brine recirculation salinity, blow down and reject brine temperature, and also some design parameters such as the number of stages, tube length, diameters, material type, price of tube, and shell material used in the evaporator manufacturing. Using VDS, all process stream characteristics are determined (mass, temperature, pressure, entropy, and rated cost) and also the heat transfer surface area (number of tubes), evaporator size, internal dimensions, and pumps are sized. So, a detailed CAPEX analysis is performed and estimated. The VDS calculates the heating steam consumption rate, the consumed chemicals (anti scales, anti foam, and chlorination) as well as the pumping power (OPEX items). The detailed costs of the evaporator materials, pumps, valves, and controls are also estimated. The price of electricity and heating steam is estimated and calculated as illustrated in [14]. The total evaporator cost is levelized for plant life cycle of 20 years. So the final tariff of water unit cost is obtained. The annual investment cost (fixed capital cost depreciation rate per year) of each component in the desalination plant is calculated according to the following relation:

$$\text{Annual investment} = \text{CAPEX} \times \frac{i \times (1+i)^n}{(1+i)^n - 1} \tag{3}$$

Using an interest rate, $i = 7\%$ and the amortization year, $n = 20$ years, the operation and maintenance costs are calculated by multiplying the equipment purchased cost by a factor of the equipment cost index. The hourly cost ($/h) of desalination plant is calculated as follows:

$$\text{hourly} - \text{CAPEX} = \frac{\text{Total annual investment}}{365 \times 24 \times 0.9} \tag{4}$$

Similarly, the hourly OPEX is calculated as follows:

$$\text{hourly} - \text{OPEX} = \text{LP steam} + \text{Electricity} + \text{Chemiclas} \tag{5}$$

Then the unit product cost of the desalted water is calculated as follows:

$$\text{Unit product cost, } \$/\text{m}^3 = \frac{\text{hourly\_CAPEX} + \text{hourly\_OPEX}}{\text{hourly\_Product}} \tag{6}$$

5. Results and discussion

5.1. Techno-economic analysis of NF system

The VDS program is used to size NF system to produce 226 m^3/h which represents the one-third of make-up required for 1 MIGD MSF. As shown in Fig. 3, the required number of pressure vessels is 29 with 174 membrane elements. The calculated system recovery ratio is 65%. The high pressure pump is assigned by 25 bars. Three units of pressure exchangers are used to recover electrical energy of 0.07 MW as shown in Fig. 3. Each unit capacity is 44 m^3/h and the percentage of salt increase is only 4.6%. The net pumping power required is 0.21 MW, and the specific power consumption is 0.94 kWh/m^3.

Table 2 shows the CAPEX cost analysis of NF system which produces 226 m^3/h. The direct cost of purchased equipments (membrane section, filters, pumps, valves, and piping) is included. The indirect costs of buildings structure, engineering, and project development are also included. The intake cost is not included and is assumed to be burden to MSF CAPEX cost. The levelized cost is calculated (based on the 7% interest rate and 15 year life span) as 0.0775 $/$m^3$ of NF permeate as shown in column three of Table 2.

Table 3 shows the operational cost of NF system which includes labor, O&M, NF membrane replacement, electricity, and chemicals. The analysis showed the cost of electricity which represents the highest part of the total OPEX; the specific operational cost is 0.0566 $/$m^3$ of NF permeate. From both Tables 2 and 3, the calculated unit permeate cost is 0.134 $/$m^3$.

Table 2
CAPEX cost analysis of NF system

Items	$	$/$m^3$
Direct cost		
PV, pass 1	26,100.00	0.001510285
Element, NF	87,000.00	0.005034285
Pumps	118,026.20	0.006829626
PX/turbine	115,359.61	0.006675323
Piping and valves	188,939.88	0.010933071
Filters	249,015.74	0.014409381
Others, building, start up	258,803.85	0.014975773
Subtotal	1,043,245.28	0.060367744
Indirect cost		
Engineering design	131,455.62	0.007606724
Financial	164,319.52	0.009508405
sub total	295,775.14	0.017115129
TCI	1,339,020.42	0.077482874

Table 3
OPEX cost analysis of NF system

Items	Cost, \$/year	Cost, \$/m^3
Labor	21,909.27	0.011546899
Maintenance	16,431.95	0.008660173
Insurance	6,572.78	0.003464069
Replacement of membrane	1,305.00	0.000687777
Electricity	34,740.04	0.018309132
Chemicals	26,401.62	0.013914513
Total	107,360.66	0.056582563

5.2. Techno-economic analysis of MSF-BR-NF

Fig. 4 shows the interface of existing 5,000 m^3/day MSF-BR desalination plant at TBT = 110°C [13]. The evaporator consists of 20 stages, 17 stages for heat recovery section and 3 stages for heat rejection section. The extracted steam from power side is directed to the brine heater as a heat source. Seawater flows through tubes of the heat rejection section condensers as a coolant. Part of this coolant outlet is used as make-up and the remaining coolant is rejected back to the sea. The make-up is directed to the deaerator and pretreatment chemicals are added, then, is mixed with a portion of the last stage brine. The circulation pump circulates the diluted mixed brine to the condensers of the heat recovery section. The tube material used in this plant is CuNi 90/10 for brine heater and heat recovery section and CuNi 70/30 for heat rejection section. The evaporator length is 29 m, width is 7 m, and the height is 2.5 m. The design conditions are 27°C for seawater and the brine velocity inside the tube is 2 m/s. The working pressure of the deaerator is 0.055 bar which is lower than the saturation temperature of the make-up of 38°C.

Fig. 5 shows the interface MSF-BR with NF system which allows to increase TBT to 130°C. The NF system treats one-third of the make-up. The feed of NF system is extracted from the cooling reject stream (48 g/l) as shown in Fig. 5. The NF permeate is mixed with the remaining make-up and directed to the deaerator. The mixed make-up of low salinity of 43 g/l (15% less) flows to the last stage of heat rejection section. Due to increase of TBT from 110 to 130°C, the distillate production increases by 19%. No increase of the GOR since the heating steam increased by the order of 19%.

Table 4 shows that the CAPEX of NF-MSF-BR system is 65.5% higher than the conventional MSF.

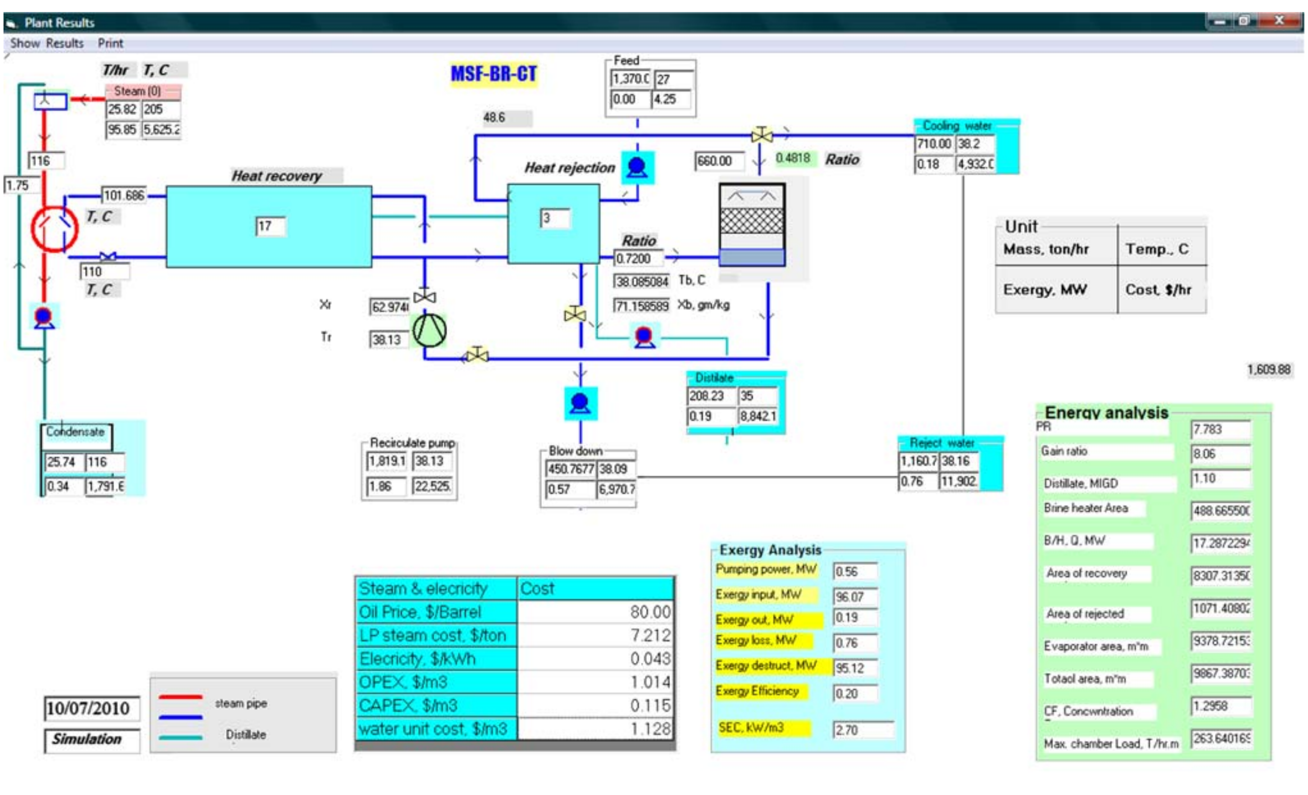

Fig. 4. Interface of existing MSF-BR desalination plant at TBT = 110°C.

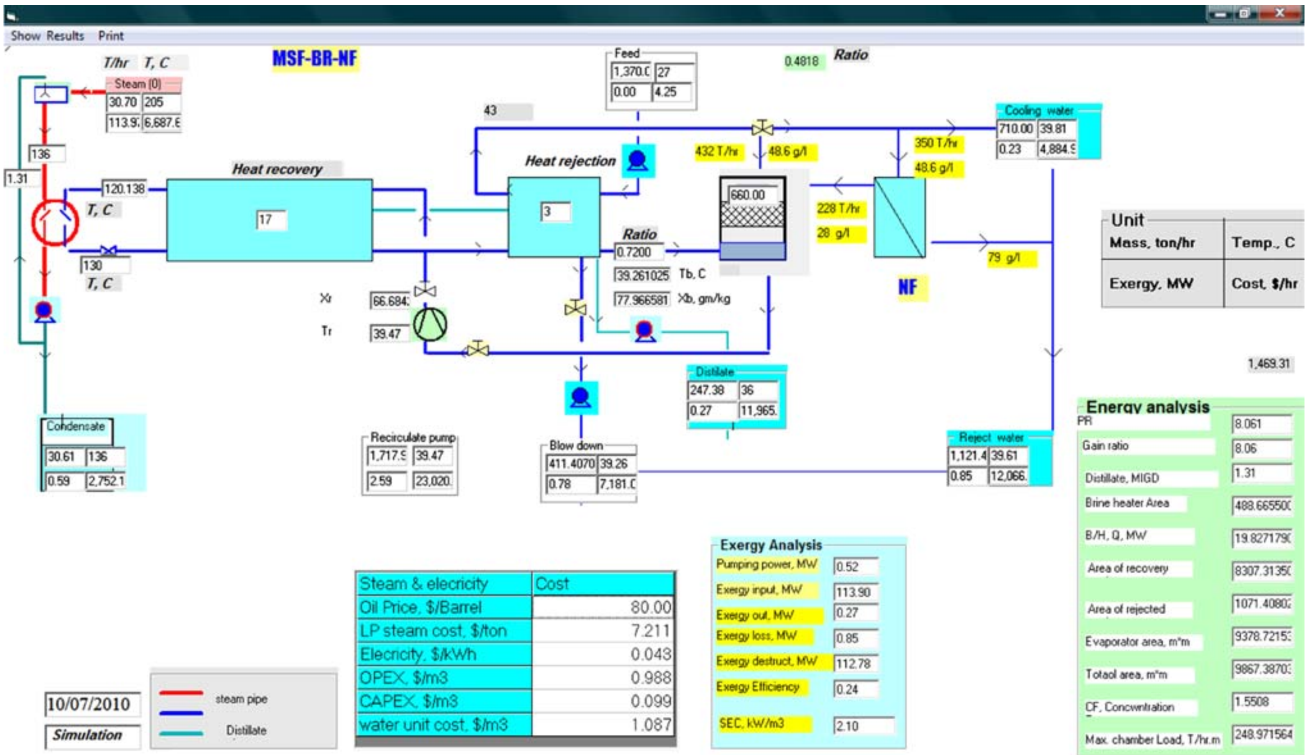

Fig. 5. Interface NF-MSF-BR desalination plant.

Table 5 shows the operating cost of NF-MSF-BR system which increased by 22.4% over the conventional MSF.

Table 6 shows the levelized CAPEX cost of MSF-BR at TBT = 130˚C is 16% lower than the conventional MSF at TBT = 110˚C, this is due to increase the distillate production by 19%. Also due to increase the productivity, the specific OPEX reduced by 2.5%. However due to adding the NF system, the levelized OPEX of NF-MSF at TBT = 130˚C is 2.65% higher while the specific CAPEX of NF-MSF is 28.7% higher than conventional.

The unit product cost of NF-MSF is 5.4% higher than that of MSF plant. The analysis of CAPEX and OPEX results shows that the OPEX cost has significant effect on the total unit water cost. This concludes that integrating NF system to an existing MSF plant (just to increase the production) is not enough to reduce the unit product cost.

Fig. 6 shows the percentage difference between NF-MSF and conventional MSF decreases as the oil price increases. This result leads us to think about high GOR MSF configuration may change this situation.

5.3. NF with the newly developed MSF-DM

Fig. 7 shows the configuration of NF with the newly developed deaeration brine mix NF-MSF-DM

Table 4
CAPEX analysis of MSF (TBT = 110˚C) and NF-MSF-BR(TBT = 130˚C)

CAPEX, 1 MIGD	Conventional	NF-MSF	% Diff.
Item	Cost, US$	Cost, US$	
Evaporator	1,040,551.34	1,040,551.344	–
Pumps	306,223.35	306,223.350	–
Pipes, valves, I&C	302666.77	302,666.770	–
Intake	394,560	394,560.000	–
NF system	–	1,339,020.42	
Total	2,044,001.46	3,383,021.88	65.5

Table 5
OPEX analysis of MSF and NF-NF-MSF-BR

Items	MSF-BR	MSF-BR-NF	% Diff.
LP steam cost	186.23	221.4	18.89
Electricity	23.8	23.8	–
Chemicals	0.48	0.48	–
Total	210.92	244.21	15.78
NF (OPEX)		13.61,753,678	
Total	210.92	257.8,275,368	22.24

Table 6
Levelized cost of MSF and NF-MSF-BR

	MSF-BR (TBT = 110℃)	MSF-BR-NF (TBT = 130℃)	% Diff.
Interest rate	0.07	0.07	
Life span	20.00	20.00	
Amortization factor	0.09	0.09	
Annual investment	192,468.70	192468.70	
Hourly production	208.07	247.21	18.81%
Hourly investment	24.41	24.41	
Specific CAPEX, MSF	0.12	0.10	−15.83%
Specific OPEX, MSF	1.01	0.99	−2.50%
NF			
NF, specific OPEX	–	0.05	
NF, specific CAPEX	–	0.05	
MSF-NF, specific CAPEX	0.12	0.15	28.68%
MSF-NF, specific OPEX	1.01	1.04	2.65%
Total water unit cost	1.13	1.19	5.35%

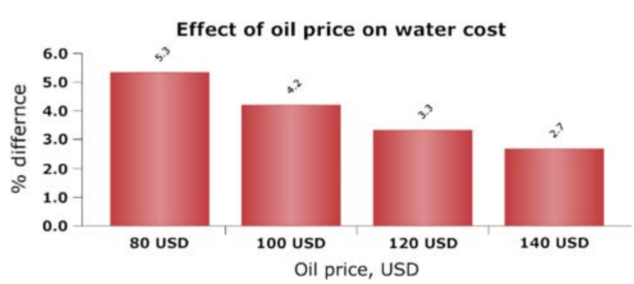

Fig. 6. Effect of oil price on the water price.

system to reduce the operational cost (OPEX). NF enables to increase the TBT = 130℃ while the MSF-DM enables to increase the GOR.

Table 7 shows the process calculation of MSF-DM at TBT = 130℃ compared with conventional MSF at TBT = 110℃. The GOR of MSF-DM-NF is twice the conventional MSF, however, the heat transfer area increased by 72% for the same capacity.

As shown in Fig. 7, the stage number of MSF-DM increased to 35 which is 75% higher than convention MSF-BR. The 61.5% of the last stage brine is mixed with the deaerated make-up flow of 675 m^3/h. The make-up is diluted from 48 to 43 g/l using NF system permeate of TDS = 28 g/l. This mixture is directed to the MSF condensers at 32.8℃ which is 15% lower than conventional MSF (38℃). This lower temperature of coolant enhanced the heat transfer process (condensation). However, the reducing cooling water reduces the logarithmic mean temperature difference across condenser compared with that of conventional. This explains why the heating surface area of MSF-DM increased by 72%. One feature of increasing heat transfer area of heat recovery section is reducing the temperature difference across the brine heater which sequentially increases brine heater surface area as well. Increasing heat transfer area of heat recovery section increases the recovered energy which minimizes the external source of heating. Reducing source of heating

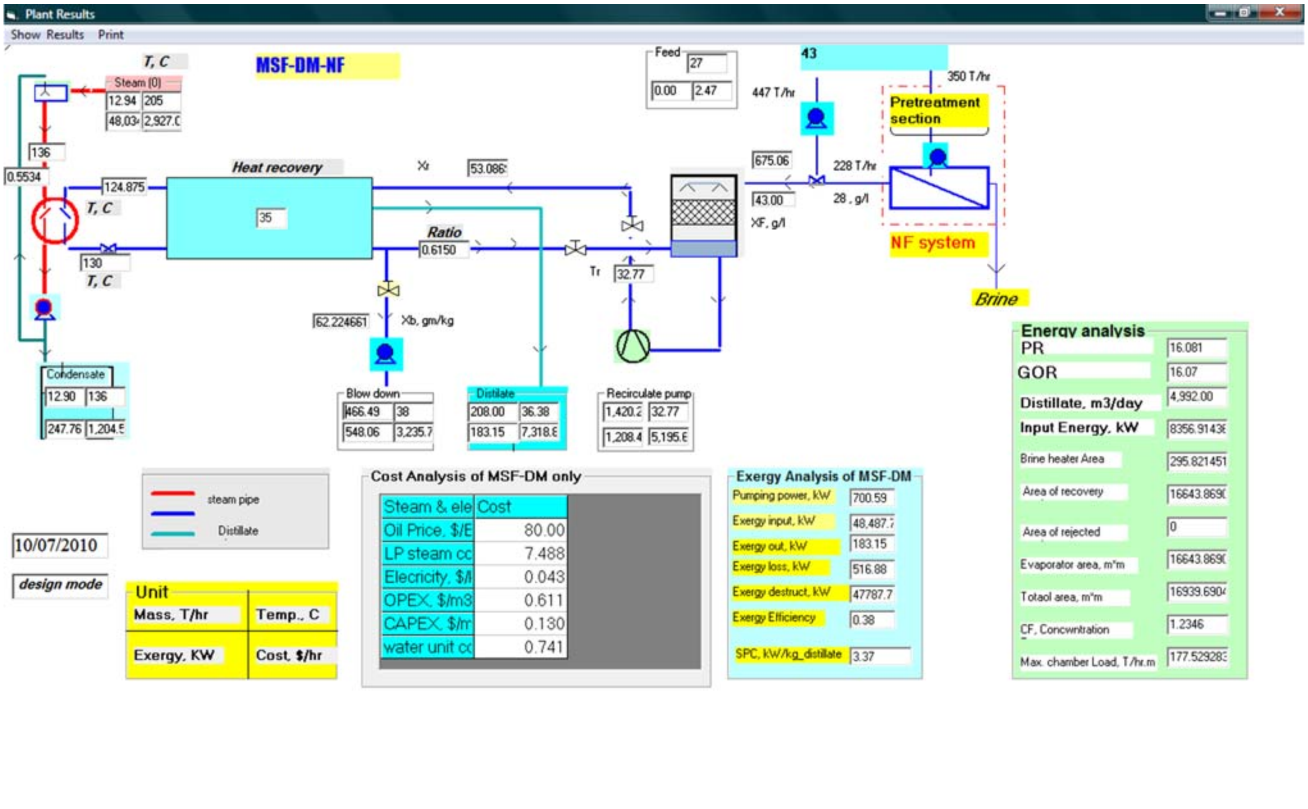

Fig. 7. NF-MSF-DM configuration.

Table 7
Process calculation of MSF and MSF-DM

	MSF-BR	NF-MSF-DM	% Diff.
Ton/h	208	208	0
TBT, C	110	130	18
Seawater flow rate, m^3/h	1,370	797	−42
Make-up, m^3/h	660	675	2
Sea salinity, g/l	48.60	48.6	0
Recycle ratio	0.72	0.615	−15
Recycle salinity, g/l	62.90	60	−5
Blow down salinity, g/l	70.00	70	0
No. stages	20.00	35	75
Heat transfer area, m^2	9,868	16,940	72
Tube length, m	7.45	8	7
Tube diameter, m	19.05	17	−11
GOR	8	16	100
Velocity, m/s	1.98	1.91	−4
SPC, KWh/m^3	2.70	3.42	27
Evaporator length, m	29.30	70.9	142
Evaporator height	2.50	2.44	−2
Evaporator width, m	7.45	8	7

(steam) for fixed capacity will increase the GOR. The process calculations show that the GOR is 100% higher than that of MSF-BR, see Table 7. This means that the steam consumption is reduced by 100% as well.

Table 7 and Fig. 7 show that the intake seawater of MSF-DM is 42% lower than that of MSF-BR, which reduces the seawater supply pump capacity as well as will reduce the intake civil work. One feature of MSF-DM is that make-up has the same value of the conventional which leads to the same chemical cost of treatment and same manufacturing cost of deaeration.

Table 7 shows that the specific power consumption of MSF-DM is 27% higher than MSF-BR. This is because the increase of the friction loss due to the increase of the stage number by 75%. The evaporator length is increased by 142% in case of MSF-DM; the evaporator width decreased by 2% while height increased by 7% as shown in Table 7.

The purchased equipment cost (PEC) of these components is estimated based on the recent market prices. In case of the scarce data about the real installation cost of the desalination plant, the PEC of the individual components could be calculated based on cost relations. These relations of estimating the capital and operating cost of the components such as pumps, valves, piping, and instrumentation are presented in Ref. [13].

Detailed cost breakdown is shown in Table 8. The evaporator (shell & tubes, de-airator) cost of MSF-DM is 47% higher due to the increase of heat surface area by 72% as shown in Table 7. The evaporator manufacturing including labor cost of MSF-DM is 52% higher than that of MSF-BR. The cost of pumps, piping, valves, and I&C control of MSF-DM is lower than that of the conventional system; this is due to the removal the heat rejection section. The cost analysis shows that the intake construction cost of MSF-DM is 42% lower than the conventional; this is due the lower of the sea water flow rate by the same order. So the increase of MSF-DM evaporator cost is partially compensated by the cost reduction of auxiliaries and intake cost. The total capital cost (CAPEX) of the proposed configuration MSF-DM is 6% higher than the conventional MSF-BR. However, the total CAPEX cost of NF-MSF-DM system is 71% higher than convention MSF. The increase of CAPEX is mainly contributed to additional NF system.

Table 9 shows OPEX items of both conventional MSF-BR and MSF-DM configurations. The cost of the steam and electricity is calculated based on average 80 $/barrel of oil price and the recent purchased cost of power generation cycle [14]. The cost of low pressure steam price directed to the desalination plant and the steam utilized for power generation is allocated based on exergy analysis [14]. Using levelization method through 20 years and 7%, the specific cost of low pressure steam is calculated as 7.5 $/m^3 of steam and the cost of generated electricity is 0.043 $/kWh.

The OPEX cost analysis, Table 9, shows that the NF-MSF-DM is 33% lower than the conventional MSF. The reduction in OPEX contributed to the reduction of the heating steam cost due higher GOR.

As shown in Table 9, the low pressure steam cost in MSF-DM configuration is 48% lower than that of the conventional MSF-BR, even the steam consumption of MSF-DM is 100% lower than that consumed by conventional MSF-BR. This is mainly due to different steam cost invoked from power side since the temperature of heat steam temperature is higher in MSF-DM (TBT = 130). The electricity cost of the MSF-DM is 27% higher than that of the conventional MSF-BR due to higher pumping power by the same order. The chemicals cost increased by only 2% higher than conventional MSF-BR. This is mainly due to increase of make-up to be treated. The total OPEX items of the proposed configuration MSF-DM is 33% lower than that of the conventional MSF-BR which is mainly due to low amount of steam consumption.

The levelized cost of capital purchased components and operating invested (Chemicals, steam, electricity, and O&M) to produce water are calculated as shown in Table 10. The specific OPEX of the MSF-DM is 34% lower than that of the conventional MSF-BR. The specific CAPEX of the MSF-DM is 12% higher than that of the conventional MSF-BR. However, the sum of the total cost invested using the MSF-DM is 34% lower than that of the conventional MSF-BR.

Table 8
CAPEX analysis of MSF and NF-MSF-DM configurations

Items	MSF-BR (TBT = 110°C)	NF-MSF-DM (TBT = 130°C)	% Diff.
Evaporator	1,066,100.45	1,570,420	47
Pumps	306,223.35	246,014	−20
Piping, valves, I&C	302,666.77	179,202	−41
Intake	394,560.00	194,416	−51
Total	2,069,550.57	2,190,052	6
NF	–	1,339,020.42	100
Total	2,069,550.57	3,529,072.713	71

Table 9
OPEX analysis of MSF and MSF-DM-NF

	MSF-BR	NF-MSF-DM	% Diff.
LP steam cost	186.23	96.9	−48
Electricity	23.8	30.23	27
Chemicals	0.48	0.49	2
Total, MSF	210.92	127.63	−39
NF		12.9	
Total, MSF-NF	210.92	140.53	−33

Table 10
Levelized cost of MSF and NF-MSF-DM configurations

Levelization cost, $/m3	MSF-BR (TBT = 110˚C)	NF-MSF-DM (TBT = 130˚C)	% Diff.
Interest rate	0.07	0.07	–
Life span	20	20	–
Amortization factor	0.094392926	0.094392926	–
Annual investment	192,468.7044	215,680.4949	12.06
Hourly production	208.07	208	−0.03
Hourly investment	24.41257032	27.35673451	12.06
Specific CAPEK	0.117328641	0.131522762	12.10
Specific OPEX	1.014038462	0.613605769	−39.49
Total	1.131367102	0.745128531	−34.14
NF			
Specific OPEX, NF	–	0.062023194	–
Specific CAPEX, NF	–	0.08493315	–
Total specific CAPEX	0.117328641	0.216455912	84.49
Total specific OPE	1.014038462	0.807151725	−20.40
	1.131367102	1.023607638	−9.52

Due adding NF system, the specific OPEX of NF-MSF-DM is 20% lower that of conventional while the specific CAPEX increased by 84% as shown in Table 10. The total unit product cost of NF-MSF-DM is 9.5% lower than conventional MSF-BR.

Fig. 8 shows that the percentage safe in water unit cost due to use hybrid NF-MSF-DM increases with increase in oil price. At current oil price of 100 $/bbl, the unit water cost of NF-MSF-DM is 14% lower than that of the conventional MSF.

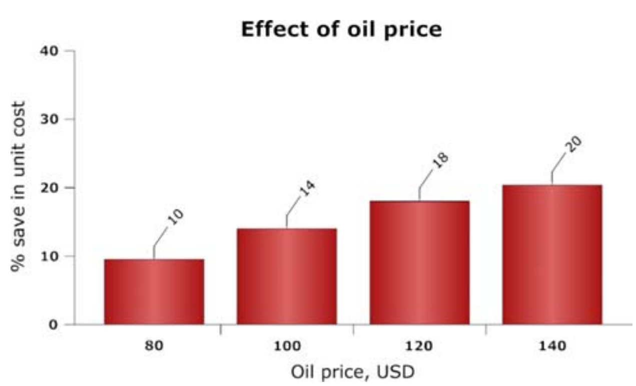

Fig. 8. Effect of oil price on the water unit cost.

6. Conclusions

- The cost analysis of stand-alone NF system without including intake cost showed that the NF unit permeate cost is 0.134 $/m^3.
- Adding NF system to an existing MSF plant, the production increases by 19% due to increase in TBT up to 130˚C. However, the unit product cost increases by 5% due to integration of NF CAPEX.
- At the current oil price of 100 $/bbl, the newly developed NF-MSF-DM configuration increases GOR to 16 and reduces the water production cost by 14% lower than conventional MSF.
- The newly developed NF-MSF-DM is recommended especially in adopting solar energy as

heating source instead of highly priced barrel oil nowadays. In this configuration, the input heating source will be reduced to the half by utilizing high performance MSF system. This clue with will reduce the CAPEX of solar system.

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Appendix A. NF mathematical model

Fig. A1 illustrates the input and output parameters used for the mass and energy balance equations of the NF:

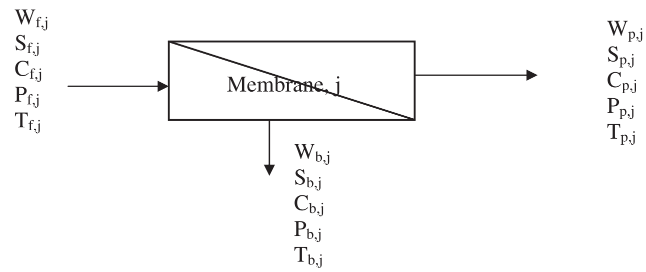

Fig. A1. Schematic diagram of the NF membrane streams.

Mass balance is written as follows:

$$W_{\mathrm{f},j} = W_{\mathrm{p},j} + W_{\mathrm{b},j} \tag{A1}$$

$$S_{\mathrm{f},j} = S_{\mathrm{p},j} + S_{\mathrm{b},j} \tag{A2}$$

The following relation defines the rate of water passage through a semipermeable membrane [6,7,9,10]

$$W_{\mathrm{p},j} = (\Delta P_j - \sigma\Delta\pi_j) \times K_\mathrm{w} \times A_j \times \mathrm{TCF} \times \mathrm{FF} \times \rho_{\mathrm{p},j} \tag{A3}$$

$$\Delta P_j = \bar{P}_j - P_{p,j} \tag{A4}$$

$$\Delta\pi_j = \bar{\pi}_j - \pi_{p,j} \tag{A5}$$

$$\bar{P}_j = 0.5(P_{\mathrm{f},j} + P_{\mathrm{b},j}) \tag{A6}$$

Since the seawater salt concentrations ratio is almost constant, an approximation for π value in kPa can be given as:

$$\pi = 6.895 \times \frac{38.5 \times C_{\mathrm{fbNaCl}} \times (T + 273)}{1,000 + C_{\mathrm{fbNaCl}}} \tag{A7}$$

$$C_{\mathrm{fbNaCl}} = 0.934348 \times C_{\mathrm{fb}} - 0.54169 \tag{A8}$$

The rate of salt flow through the membrane is defined as:

$$S_{\mathrm{p},j} = (C_{\mathrm{m},j} - C_{\mathrm{p},j}) \times K_\mathrm{s} \times A_j \times \mathrm{TCF} + (1 - \sigma) \times J_{\mathrm{v},j} \times \bar{C} \times K_\mathrm{s} \times A_j \times \mathrm{TCF} \tag{A9}$$

$$J_{\mathrm{v},j} = (\Delta P_j - \sigma\Delta\pi_j) \times K_\mathrm{w} \times \mathrm{FF} \times \mathrm{TCF}\ (\mathrm{m/s}) \tag{A10}$$

where the temperature factor correction (TCF) is calculated from the following equations [10]:

$$\mathrm{TCF} = \mathrm{e}^{8.859 \times \frac{T-25}{T+273}}, \quad \text{for } T \geq 25\,^\circ\mathrm{C} \tag{A11}$$

$$\mathrm{TCF} = \mathrm{e}^{11.678 \times \frac{T-25}{T+273}}, \quad \text{for } T \leq 25\,^\circ\mathrm{C} \tag{A12}$$

$$C_{p,j} = S_{p,j} \times \rho_{p,j}/(S_{p,j} + W_{p,j}) \tag{A13}$$

A material balance within the mass transfer boundary layer near the membrane wall between the solute carried to the membrane by convection and the solute carried away by diffusion yields an expression that quantifies concentration polarization [7].

A material balance within the mass transfer boundary layer near the membrane wall between the solute carried to the membrane by convection and the solute carried away by diffusion yields an expression that quantifies concentration polarization:

$$\varphi = \frac{C_m - C_p}{C_b - C_p} = e^{J_w/k} \tag{A14}$$