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Technoeconomic analysis of tri hybrid reverse osmosis-forward osmosis-multi stage flash desalination process

Abdelnasser Mabrouk[a,b,*], Muammer Koc[b], Ahmed Abdala[c]

[a]*Qatar Environment and Energy Research Institute, Hamad Bin Khalifa University, Doha, Qatar,*
[b]*College of Sciences and Engineering, Hamad Bin Khalifa University, Doha, Qatar, email: aaboukhlewa@hbku.edu.qa (A. Mabrouk), mkoc@hbku.edu.qa (M. Koc)*
[c]*Chemical Engineering Program, Texas A&M University at Qatar, Doha, Qatar, email: ahmed.abdala@qatar.tamu.edu*

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ABSTRACT

A new tri-hybrid reverse osmosis-forward osmosis-multi stage flash (RO-FO-MSF) desalination process is developed to increase the overall plant recovery ratio and reduce energy consumption. The FO membrane is employed as pretreatment to the MSF section to allow the increase of the top brine temperature and to obtain maximum allowable recovery from the RO brine reject. The product blend of the MSF distillate and the single pass RO permeate enables controlling the boron concentration below the acceptable limit. The current recovery ratio limitation of the existing MSF and RO plant is analysed. A process simulator is developed to carry out mass balance, heat balance, and electrical and thermal energy consumption. The total energy consumption of the tri hybrid process is compared with that of commercial MSF and RO desalination plants. The simulation results show that the RO-FO-MSF process recovery ratio is 30% higher than that of the standalone RO and MSF processes. The specific total energy consumption, (electrical plus equivalent thermal) of the tri hybrid process is 65% lower than that of MSF, but 20% higher than RO. The lifecycle cost analysis showed that the unit water cost of the tri hybrid process is 20% lower than that of the RO and 40% lower than that of the MSF standalone processes. This water cost saving is attributed to the use of the MSF cooling rejected as feed to the RO section and using the RO brine as feed solution to FO, which decreases the capital cost of the civil work and construction of the intake/outfall facilities. Moreover, the product blend of the MSF distillate and the RO permeate enables employing a single pass RO and eliminate the need for a second pass and thus reducing the capital investment.

Keywords: Desalination; Multi stage flash; Reverse osmosis; Forward osmosis

1. Introduction

Thermal desalination has established a stronghold in co-generation power-desalting plants in the Gulf Cooperation Council (GCC) countries (Saudi Arabia, United Arab Emirates, Kuwait, Qatar, Bahrain, and Oman) where large amount of desalting water are needed and energy cost are considerably low. On the other hand, seawater reverse osmosis (SWRO) has lower energy consumption due to the recent developments in membrane materials and energy recovery technology [1]. Nevertheless, thermal desalination plants is still used in the GCC countries due to its better reliability compared to RO due to the challenging Gulf seawater conditions (high salinity, high temperature, high turbidity and red tides that have forced shut down of RO plants) [2]. Realizing the benefits and the challenges of the thermal and membrane technologies, process designers are investigating hybrid (thermal–membrane) configurations [2]. The synergy of the present commercial hybrid thermal-membrane desalting plants is limited to using common intake/outfall facilities, and mixing the desalinated water of each system, while running independently at the same site [2].

Seawater in hot climate areas such as the GCC countries usually has high salinity and high boron content (7 mg/L).

*Corresponding author.

Although during the hot season, the SWRO rejection of all ionic species is relatively high and ranges between 99.35% to 99.95%, while the boron rejection is usually between 75 and 90% [3]. Therefore, to reduce the permeate boron concentration below the required 0.5 mg/L limit; two-pass RO system is typically used. Nir and Lhav [4] investigated the effect of the RO recovery ratio on the boron permeate concentration by coupling membrane transport and chemical equilibrium models. Their simulation results indicated that boron permeate concentration is pH dependent and increases with the recovery ratio. Moreover, the effect of different RO process configurations on the boron permeates concentration and the water unit costs were also investigated by Taniguchi et al. [5]. These configurations include two stages RO with alkali dosing before or after the second stage RO, permeate adsorption as post treatment, and a combination of alkali dosing and adsorption post-treatment. The single stage RO process cannot reduce the boron concentration below the required 0.5 mg/L limit. On the other hand, two stage RO configuration with alkali dosing, adsorption, or both can reduce the boron permeate concentration below 0.5 mg/L, however, the water unit cost is increased. The solution to reduce the boron concentration does not only increase the water unit cost, but also increase the environmental impact as it requires use of chemical and/or disposal of spent adsorbents.

Analysis of different power and hybrid MSF-RO systems [6] showed that employing higher capacity of SWRO in a hybrid configuration would decrease the power/water ratio, which means reducing the electrical power exported to the grid from the cogeneration plant and increasing the desalted water production. For example, if the SWRO capacity share in the hybrid process is raised to 60%, the equivalent energy is reduced to 9.5 kWh/m^3 from the 16 kWh/m^3 for standalone MSF [6].

In another study, Elsayed et al. examined the performance of a commercial MSF/RO hybrid system of 300 m^3/d nominal capacity SWRO plant [7]. In this hybrid system, the feed to the SWRO plant was taken directly from the cooling seawater leaving the MSF heat rejection section. As a result, the SWRO permeate output increased on average by about 2.2% per degree Celsius increase in the feed seawater temperature. The specific energy consumption of MSF/RO hybrid operation can be reduced by up to 8% compared to the stand-alone SWRO process [7].

Recently, Hilal et al. carried a technoeconomic analysis of different hybrid MSF/RO configurations in comparison to stand-alone brine recycle MSF and two passes SWRO processes [8]. The results showed that the hybrid RO/MSF plant has lower specific capital cost and higher water recovery compared to the stand-alone systems. This cost saving is attributed to the smaller intake, the use of single-stage RO process, the longer membrane life, and the lower salt rejection. Moreover, the reduction in the steam cost allows the MSF process to compete with hybrid RO/MSF plants. These results highlight the advantage of coupling MSF plants and steam power plants if the exhaust steam is cheap heat source.

A number of commercial desalination plants in Saudi Arabia (Jeddah, Al-Jubail, and Yanbu) and UAE (Fujairah) are currently adopting a simple, but not fully hybrid desalination processes where MSF and RO plants operate entirely independent, but have common intake and outfall facilities and the RO water product is blended with the MSF product. However, using the preheated seawater leaving the MSF heat rejection, as feed to the RO plant has not been applied yet [9].

Forward osmosis (FO) is an emerging separation process in which pure water permeates from seawater through the FO membranes to the other side containing relatively higher concentration of draw solution. The pure water is then recovered from that draw solution, which requires additional energy consumption. Therefore, extensive research work on hybridization of FO with a thermal-membrane process to lower its energy consumption and develop more efficient draw solutions has recently been the subject of numerous research publications, which were reviewed by Darwish et al. [10].

Spiral wound forward osmosis membrane modules with different spacer designs (corrugated spacer [CS] and medium spacer [MS]) were investigated for the fertilizer-drawn forward osmosis (FO) desalination of brackish groundwater (BGW) at a pilot-scale level [11]. The draw solution (DS) concentrations using ammonium sulfate ($(NH_4)_2SO_4$, SOA) on the performance of two membrane modules played a significant role in FO membrane process. CS module performed slightly better than MS module during all experiments due to probably enhanced mass transfer and lower fouling propensity associated with the CS. Besides, CS spacer provides larger channel space that can accommodate larger volumes of DS, and hence, could maintain higher DS concentration. However, the extent of dilution for the CS module is slightly lower [11].

The structural features of a spiral-wound forward-osmosis (SW FO) membrane module via a pilot-scale was experimentally analysed [12]. The pilot test employed 4040 SW FO module with a lower draw flow rate than feed flow rate. Because of the structural features of the SW FO module, the draw solution flowed inside of the membrane envelope under a considerable pressure in order to overcome the flow resistance. A water flux equation based on a temperature-correction factor (TCF) was proposed to predict the water flux at a given temperature [12].

Hybrid (MSF-FO) and (MED-FO) where FO is used as a pre-treatment for existing MSF or MED plants were also suggested [13,14]. The aim of this hybridization is to reduce the concentration of divalent ions, which cause $CaSO_4$ hard-scale deposition. Removal of these divalent ions from the MSF makeup water enables increasing the top brine temperature (TBT) above 110°C and thus enhances the plant performance and productivity. Moreover, the MSF-FO hybrid design does not require external draw solution which reduces the FO cost associated with operating cost (OPEX). However, due to the low osmotic pressure of the MSF brine, the required FO membrane area is relatively high leading to high capital cost (CAPEX). Therefore, the trade-off between the increase in CAPEX and the reduction of OPEX needs to be evaluated.

The typical high level of sparingly soluble hardness ions in Arabian Gulf seawater has great effect on limiting the process recovery (fresh water production). Increasing the recovery ratio beyond the hardness ions solubility limits leads to scale formation on the heat exchanger tubes of thermal desalination plants or on the RO membrane surface, which both decline the plant performance. This situation motivates us to search for novel hybridization of thermal and membrane processes that increase the process recovery ratio while retaining the plant operation performance.

In this article, a novel tri-hybrid reverse osmosis – forward osmosis – multi stage flash (RO-FO-MSF) desalination process is developed to increase the overall plant recovery ratio, hence reduce the specific energy consumption.

2. Description of the Tri hybrid RO-FO-MSF Process

The process flow diagram of our proposed 50 MIGD tri hybrid RO-FO-MSF desalination process is presented in Fig. 1. The process can be described as follows: seawater feed (1) enters the heat rejection section (HJS) of the MSF as seawater cooling. The exit cooling seawater from the MSF HJS (2) is chemically treated and then directed to the RO section using high pressure pump. The RO brine reject (3) is directed to the feed side of the FO section. The MSF brine reject from the last stage (5) is directed to the concentrated side of the FO membrane. Due to the osmotic pressure difference between the highly concentrated brine (MSF reject) and the feed side (RO reject), pure water transfers from the feed side (RO reject) to the MSF brine side. The diluted brine (6) is circulated back to the MSF evaporator. Since the role of FO is to selectively retain the divalent ions from the feed side and allow pure water transport to the concentrated side. This allows increasing the TBT above the current limit of 110°C to up to 130°C. The TBT increase consequently increases the MSF unit distillate production and recovery ratio. Blending the MSF distillate (9) and the RO permeate (10) allows the use of single pass RO, excluding the second pass, which reduces the capital cost. Moreover, the MSF and RO product blending will dilute the RO permeate with the MSF distillate and thus reduces the Boron concentration to below the 0.5 mg/L limit. The RO brine has potential pressure energy and residual chemicals, which is used to assist the FO membrane process.

3. Methodology

Analysis of the current limitation in the process recovery ratio of standalone MSF and RO processes is first discussed by simulating two existing desalination plants. 42.5 MIGD desalinated water capacity RO plant located in AL-Gubrah, Oman and 4 × 15 MIGD desalinated water capacity MSF plant located in Rass Laffan, Qatar. The maximum recovery ratio from the feed seawater using the proposed tri hybrid process is then examined. The mechanical energy equivalent to thermal energy (heating steam) consumption of the MSF process in addition to the electrical energy consumption is calculated as function of the recovery ratio. Finally, the energy consumption of the tri hybrid configuration is computed as function of the recovery ratio using our previously developed and verified Visual Simulation Program (VSP) [15] and compared to the energy consumption of the standalone MSF and RO. The VSP program is developed for design and simulation of different types and configurations for both thermal and membrane desalination processes.

The capital and operational cost of the desalination plant is calculated using recent bidding of commercial desalination projects [1,24]. The VSP simulator also calculates the heating steam consumption rate in the case of the MSF desalination plant. The consumed chemicals (anti-scales, anti-foam, and chlorination) as well as the electrical power are calculated for each configuration, i.e., MSF, RO and RO-FO-MSF. The price of electricity and heating steam for MSF are estimated based on the analysis of power side [21]. The capital cost of the MSF, RO and tri hybrid configuration is levelized along the plant life cycle of 20 years with 7% interest rate to obtain the tariff water unit cost. To evaluate the cost effectiveness of the tri hybrid RO-FO-MSF, it is necessary to compare the annual values of capital and operation costs. The capital cost of the MSF desalination plant includes the evaporator, electrometrical, Instrumentations, electricity work, intake/outfall and potable tanks [23]. The capital cost of RO includes the membrane section, pretreatment, mechanical equipment, instrumentation, electrical

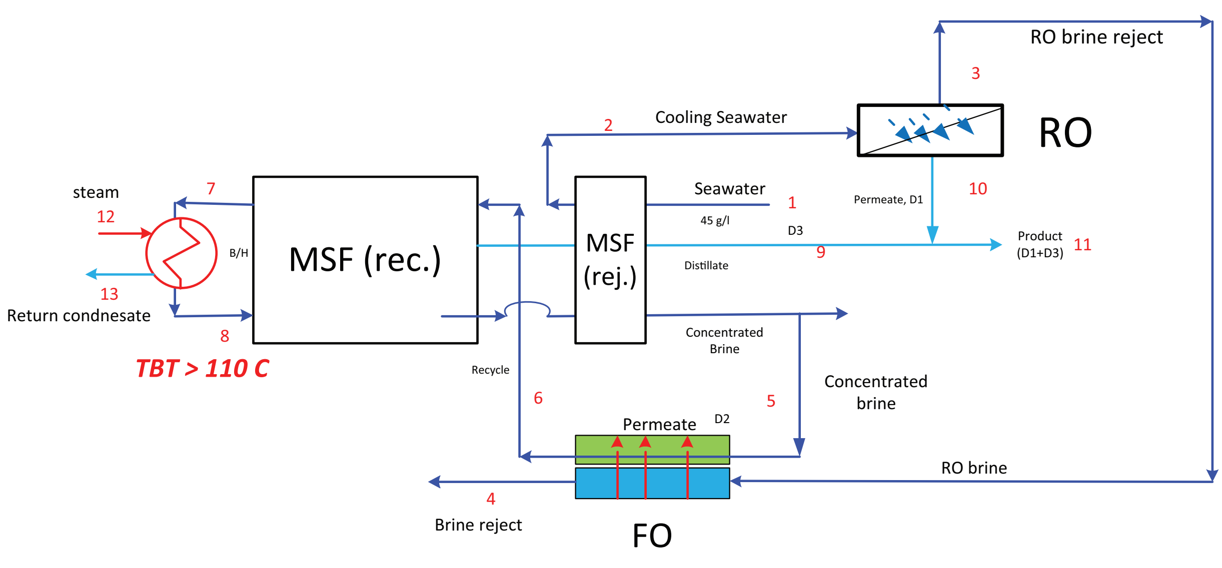


Fig. 1. Process flow diagram of the novel tri hybrid RO-FO-MSF.

work, intake/outfall constructions and potable tanks. The engineering, the project development, the insurance and taxes are included and added to the direct capital cost.

The annual investment cost is calculated using the total cost investment (TCI) according to the following relation:

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

Using an interest rate of, i = 7 % and the amortization year, n = 20 years for SWRO while n = 25 is used for MSF. The hourly levelized investment cost ($/h) of desalination plant is calculated (for plant availability = 0.95) as follows:

$$Capex = \frac{Annual\ investment}{365\times 24\times 0.95}, \$/h \tag{2}$$

The levelization annual cost of OPEX items (steam, electricity and chemicals) is calculated, (as the present value of steam, electricity and chemical costs are calculated) from the process flow diagram and relevant price of each component as follows:

$$\begin{aligned} Opex &= electricity + Steam(MSF) \\ &+ chemcials + labour + replacemnt, \$/h \end{aligned} \tag{3}$$

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

$$Water\ cost = \frac{Capex + Opex}{production\ rate}, \$/m^3 \tag{4}$$

3.1. Simulation of RO system

The mass balance of pure water and salt passage around one membrane element is given as follows:

$$W_{f,j-1} = W_{p,j} + W_{b,j} \tag{5}$$

$$S_{f,j-1} = S_{p,j} + S_{b,j} \tag{6}$$

The water permeate through a semi permeable membrane can be calculated as [16]:

$$W_{p,j} = (\Delta P - \Delta\pi)\times K_w \times A \times TCF \times FF \tag{7}$$

where K_w *i*s the water permeability coefficient, Δp denotes the osmotic pressures of feed solution, DP denotes the hydraulic pressure difference across the membrane, A denotes the membrane surface area, *FF* denotes fouling factor, and *TCF* denotes temperature correction factor. Under the commercial operating conditions of Fujairah RO plant The membrane water permeability K_w of the membrane SW membrane type is determined as 1.5×10^{-9} (m^3/ m^2/s/kPa), However for the brackish water membrane is determined as 1.05×10^{-8} (m^3/ m^2/s/kPa) .

The rate of salt flow through the membrane is defined as

$$S_p = (C_m - C_p)\times K_s \times A \times TCF \tag{8}$$

where, K_s is the salt permeability coefficient, C_m denotes the concentration at the membrane wall, C_p denotes the permeate concentration. Under the operating conditions of the commercial Fujairah RO plant, The salt passage coefficient, K_s of the brackish water membrane type is determine as $K_s = 1.6 \times 10^{-8}$ (m^3/m^2.s) however its value is 1.1×10^{-7} for the brackish water membrane which used in the second pass.

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 (*j*) [5] and [16]:

$$\varphi = \exp\left(\frac{J_w}{k}\right) = \left(\frac{C_m - C_p}{C_b - C_p}\right) \tag{9}$$

where $J_w = \dfrac{W_p}{A}$ is the mass transfer coefficient (k) that can be estimated using appropriate empirical mass transfer expression [5] and [16].

The VSP simulated the RO system at different recovery ratio from (15 to 30 %) by varying the high pressure pump. The seawater feed conditions (flow rate, salinity, temperature, pressure) and the number of pressure vessel, number membrane elements, membrane area, the membrane characteristic, are specified as input to the VSP simulator. The system model equations of the RO process will be solved. The permeate condition, the brine condition, the pumping power and recovery ratio are calculated.

3.2. Simulation of FO system

A process model of FO membrane is developed and presented in [17]. The draw solution (NaCl) is directed to the substrate side while the Feed solution (brine of RO) is directed to the active layer side. The concentration draw solution is adjusted at a higher value than the incoming feed solution concentration. Due to concentration difference, permeate water passes through the solution from the lower concentrated side (Feed solution) to the higher concentration side (Draw solution).

Knowing the specifications of the FO membrane (permeability and salt passage) and membrane area, the water and salt flux (J_w and J_s) are calculated.

The reverse solute is considered and the water flux and solute flux are presents as follows [18]:

$$J_w = k_w\left(\frac{\pi_D \exp\left(-\frac{J_w S}{D}\right) - \pi_F \exp\left(\frac{J_w}{k}\right)}{1 + \frac{k_s}{J_w}\left[\exp\left(\frac{J_w}{k}\right) - \exp\left(-\frac{J_w S}{D}\right)\right]}\right) \tag{10}$$

$$J_s = k_s\left(\frac{C_D \exp\left(-\frac{J_w S}{D}\right) - C_F \exp\left(\frac{J_w}{k}\right)}{1 + \frac{k_s}{J_w}\left[\exp\left(\frac{J_w}{k}\right) - \exp\left(-\frac{J_w S}{D}\right)\right]}\right) \tag{11}$$

where k_W *i*s the water permeability coefficient, p_D and p_F denotes the osmotic pressures of the draw and feed solution respectively.

where, k_s is the salt permeability coefficient, the support layer structure, s, is defined as $s=\frac{t_s\tau}{\varepsilon}$, with t_s is the support thickness, τ is tortuosity, and ε its porosity.

Knowing J_W, J_s, and the input flow rates, by substituting in Eqs. (12)–(17), the outlet stream flow rate and composition are calculated.

Permeation water flow rate:

$$W_p = J_w \times A \times \rho \tag{12}$$

Salt passage flow rate:

$$S_p = J_s \times A \tag{13}$$

Pure water balance at the draw solution side:

$$W_{DS,out} - W_{DS,in} = W_p \tag{14}$$

Salt balance at draw solution side:

$$S_{DS,out} - S_{DS,in} = S_p \tag{15}$$

Pure water balance at feed solution side:

$$W_{FS,in} - W_{FS,out} = W_p \tag{16}$$

Salt balances on the feed solution side:

$$S_{FS,in} - S_{FS,out} = S_p \tag{17}$$

3.3. Simulation of MSF system

The VSP for solving MSF model is previously verified and published by the authors in [15]. Process design calculations are performed by specifying the heating steam operating conditions (pressure, temperature), top brine temperature (TBT), sea water conditions (temperature, and salinity), make-up flow rate, brine recirculation salinity, blow down and reject brine temperature. The design parameters such as the number of stages, tube length, diameters, number of tubes, material type, price of tubes and shell material used in evaporator manufacturing are specified. Using VSP, all process streams are determined (mass, temperature, pressure, and rated cost), the distillate flow rate, evaporator size, internal dimensions and pumps are sized. Nevertheless, the VSP simulator calculates the heating steam consumption rate, the consumed chemicals (anti scales, antifoam, and chlorination) as well as the pumping power.

In this article, the VSP simulator is revised to incorporate FO model and investigate the integration of the FO process with the MSF process. The brine of the last stage of MSF is directed to FO (as draw solution), while the RO brine is directed to the FO (as feed solution). Due to the concentration difference, the permeate water crosses the FO membrane from the feed solution to the draw solution. As such permeate dilute the draw solution, and recirculate back to the heat recovery section of MSF to get heat. Since the diluted brine is divalent free, the TBT would be increased above 110°C. The VSP simulator is then used to investigate the integration system at different recovery ratio of the RO process at fixed TBT of 130°C.

4. Results and discussion

4.1. Technical analysis

As we have demonstrated, at high seawater salinity of 45 g/l, see Table 1, which is typical of GCC seawater, the process recovery ratio of a standalone RO and MSF is limited at the range of 30–35% see in the Appendix A. This limitation is due to the boron concentration constrain in the RO and due to scale deposition in the MSF tubes. This low recovery ratio, motivated us to propose a new hybrid configuration that overcomes the low recovery ratio limitation. The new hybrid configuration would abstract maximum utilization of the seawater feed and utilizes the potential pressure and chemical energy in the RO brine blowdown. Moreover, residual chemicals in the RO brine such as antiscalant should also be useful to reduce scale in the MSF evaporator tubes.

The proposed Tri hybrid RO-FO-MSF configuration is simulated in three sequential steps. In the first step, the single pass RO is simulated separately. In the second step, the FO mathematical model is incorporated in the VSP simulator and the FO section is simulated separately, where the RO brine is used as FO feed-side and the MSF brine (concentrated NaCl) is used as the draw solution. In the third step, the permeate of FO is used to dilute the MSF brine before being recirculated to the heat recovery section.

The boundary conditions of the simulation calculation are summarized as below:

- Seawater temperature and salinity are 30°C and 45 g/L
- RO recovery ratio varies between 15 and 35%
- FO recovery ratio varies between 35 and 15%
- Spiral-wound FO (SW FO) with water permeability coefficient is specified as 1 LMH/bar,
- MSF evaporator size is varied while the specific heat transfer area is fixed,
- The MSF evaporator chamber load is maintained below 1600 t/m/h,
- TBT is fixed at 130°C,
- MSF condensers in-tube velocity = 1.4 to 2.6 m/s,
- The Temperature of the MSF brine reject to the RO should not exceed 40°C,
- Temperature of the brine recycle to FO should not exceed 40°C,

Table 1
Typical composition of Arabian Gulf seawater

Cations	ppm	Anions	ppm	Element	ppm
K^+	500	Cl^-	25,000	Boron	5.6
Na^+	1,4000	SO_4^{-2}	3,500		
Mg^{+2}	1,660	HCO_3^-	180	TDS	45,000
Ca^{+2}	540	CO_3^-	10		

- Salinity of the MSF brine recycle before directed to FO is varied from 85 to 115 g/l
- The feed solution exits from FO should not exceed 80 g/l (at maximum FO recovery ratio),

The simulation is performed at specified seawater feed flow rate of 24,000 m^3/h and salinity of 45 g/L to single pass RO system and the recovery ratio is calculated as 35%. The RO permeate flow rate of 8085 m^3/h (42 MIGD with salinity of 0.456 g/L) will be blended with the MSF distillate. The calculated salinity of the brine, which is directed to the FO feed side, is 70 g/L. The residual chemicals in the brine in addition to the available pressure (3 bar) would assist the FO process. The pressure exchanger assists the RO high pressure pump and recovers 46% of the brine energy as shown in Fig. 2.

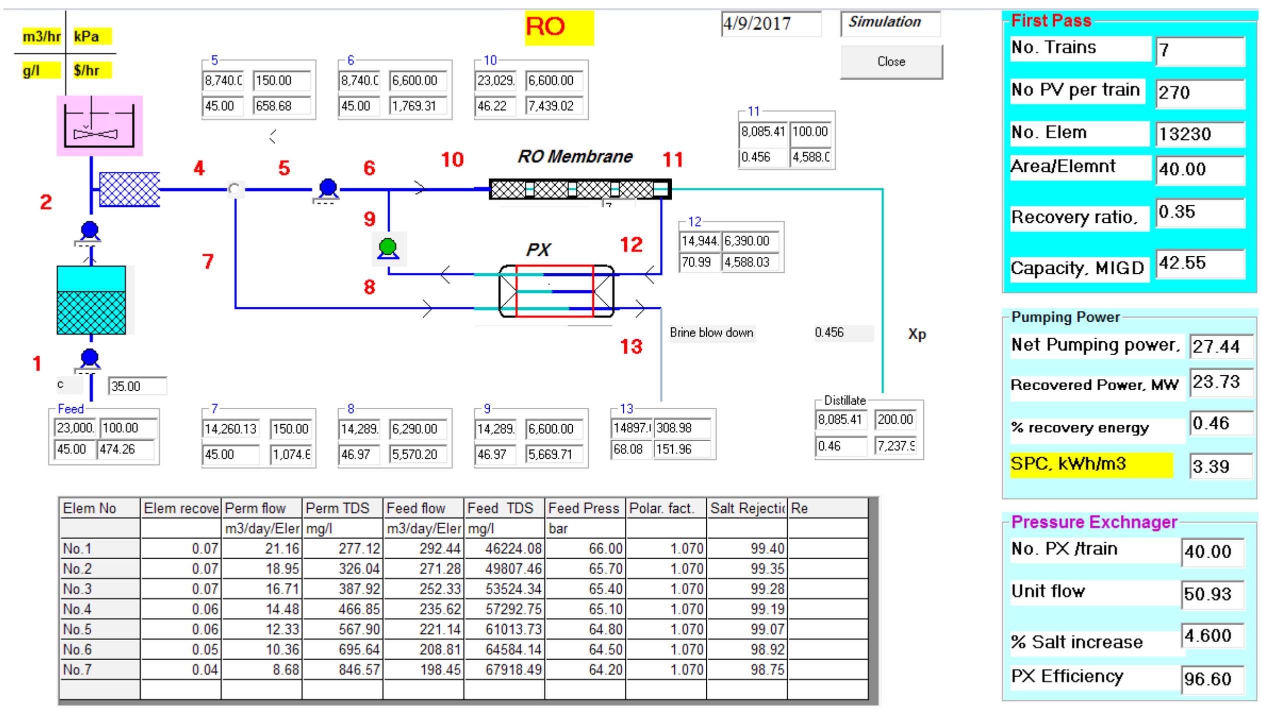
As shown in Fig. 3, the concentrated draw solution (NaCl) from the last MSF stage (15,000 t/h and 142 g/L of 114 bar osmotic pressure) is directed to the draw solution side of the FO membrane. In a counter flow operation, the RO brine (15,000 t/h and 68 g/L) is directed to the feed side of the FO membranes.

It is worth noting, that the current available FO module area is limited to 9–11 m^2 [11] and [12]. The packing density of the spiral-wound FO (SW FO) modules is about half of that of spiral wound RO 8-inch modules. This because in forward osmosis processes there must be a cross flow of solutions on either side of each individual membrane layer. This requirement increases the total thickness of spacers between membrane layers and subsequently decreases the packing efficiency. Nevertheless, in the present study, the SW FO membrane area per element is proposed 40 m^2 rely on the noticeable FO market growing stepwise. As such, SW FO technology challenges would be resolved in near future either through research and development to reduce the spacer thickness or using wider pressure vessel e.g. 12–16 inch to pack 40 m^2 flat sheet.

The water permeability is assumed as 1.0 LMH/bar [19,20]. The required number of FO membranes is calculated as 1148, which is relatively lower than that of BWRO (2450 elements) used in the second pass of the Fujairah RO desalination plant. The calculated FO recovery ratio is 15% (2262 t/h). It is proposed to employ four FO elements in each vessel, corresponding, the number of vessels is calculated as 287 to obtain the designed permeate and recovery ratio.

Fig. 4 shows the simulation results of the integrated MSF- FO membrane. The brine of the RO system is directed to the feed side of the FO membrane. The FO role is to retain the divalent ions in the feed side and only allows pure water to pass. The concentrated draw solution from the last MSF stage is diluted, allowing increasing the TBT from 110°C to 130°C. The increase of TBT enables increasing the number of MSF stages to 25 instead of 16 stages, which increases the GOR from 8 to 10 (25% increase). The MSF system recovered 98.8% of the permeated water through FO membrane (2200 t/h). The specific electrical energy consumption is 2.6 kWh/m^3 and the mechanical energy equivalent to the thermal energy is 10.9 kWh/m^3. Thus, the total energy consumption of the FO-MSF is 13.5 kWh/m^3 as shown in Fig. 4

The production (desalinated water) of the tri configuration at different recovery ratio of the RO section is calculated and presented in Fig. 8a. The total production of the tri hybrid configuration (RO-FO-MSF) is kept constant at 10,000 m^3/h, regardless the recovery ratio of the RO system. This implies that as the production of the RO section increases, the production of the FO-MSF decreases as constrained by the available osmotic pressure difference.



Elem No	Elem recove	Perm flow	Perm TDS	Feed flow	Feed TDS	Feed Press	Polar. fact.	Salt Rejectio	Re
		m3/day/Eler	mg/l	m3/day/Eler	mg/l	bar			
No.1	0.07	21.16	277.12	292.44	46224.08	66.00	1.070	99.40	
No.2	0.07	18.95	326.04	271.28	49807.46	65.70	1.070	99.35	
No.3	0.07	16.71	387.92	252.33	53524.34	65.40	1.070	99.28	
No.4	0.06	14.48	466.85	235.62	57292.75	65.10	1.070	99.19	
No.5	0.06	12.33	567.90	221.14	61013.73	64.80	1.070	99.07	
No.6	0.05	10.36	695.64	208.81	64584.14	64.50	1.070	98.92	
No.7	0.04	8.68	846.57	198.45	67918.49	64.20	1.070	98.75	

Fig. 2. Simulation results of single pass RO section of the tri hybrid process.

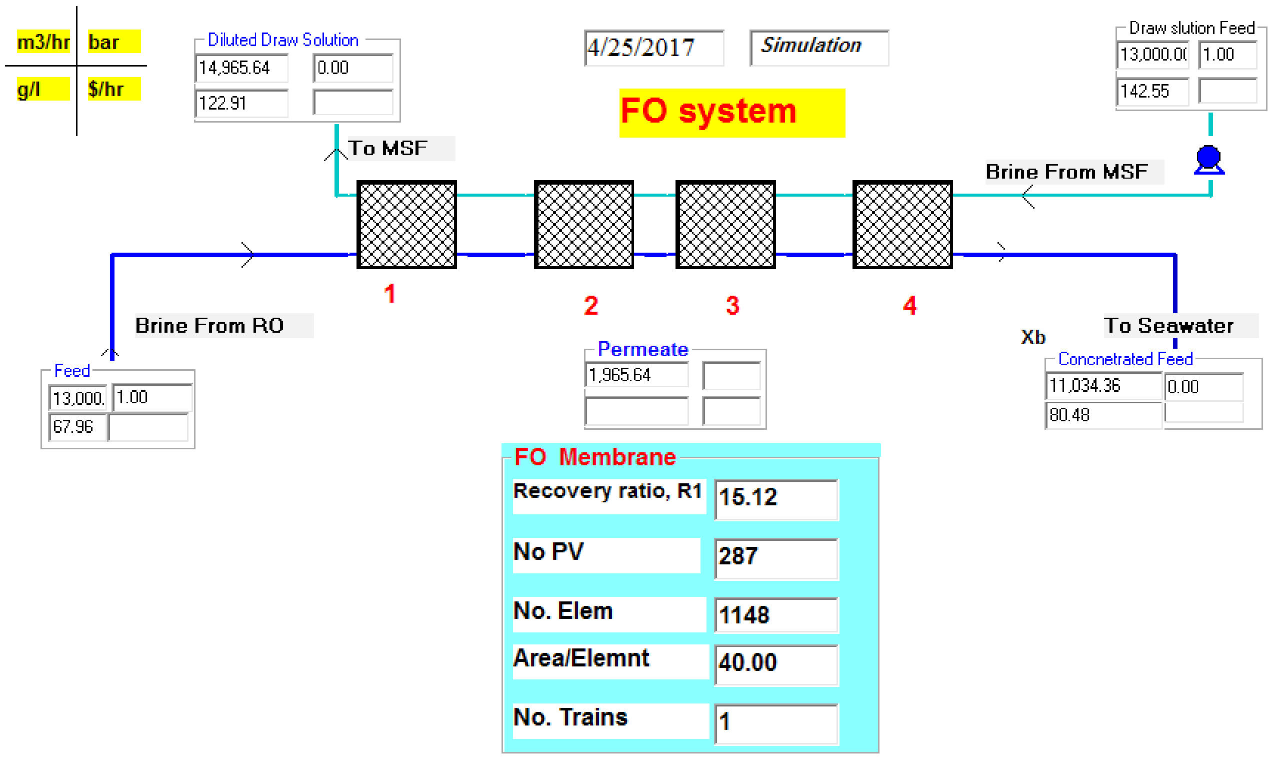


Fig. 3. VSP interface of FO membrane process.

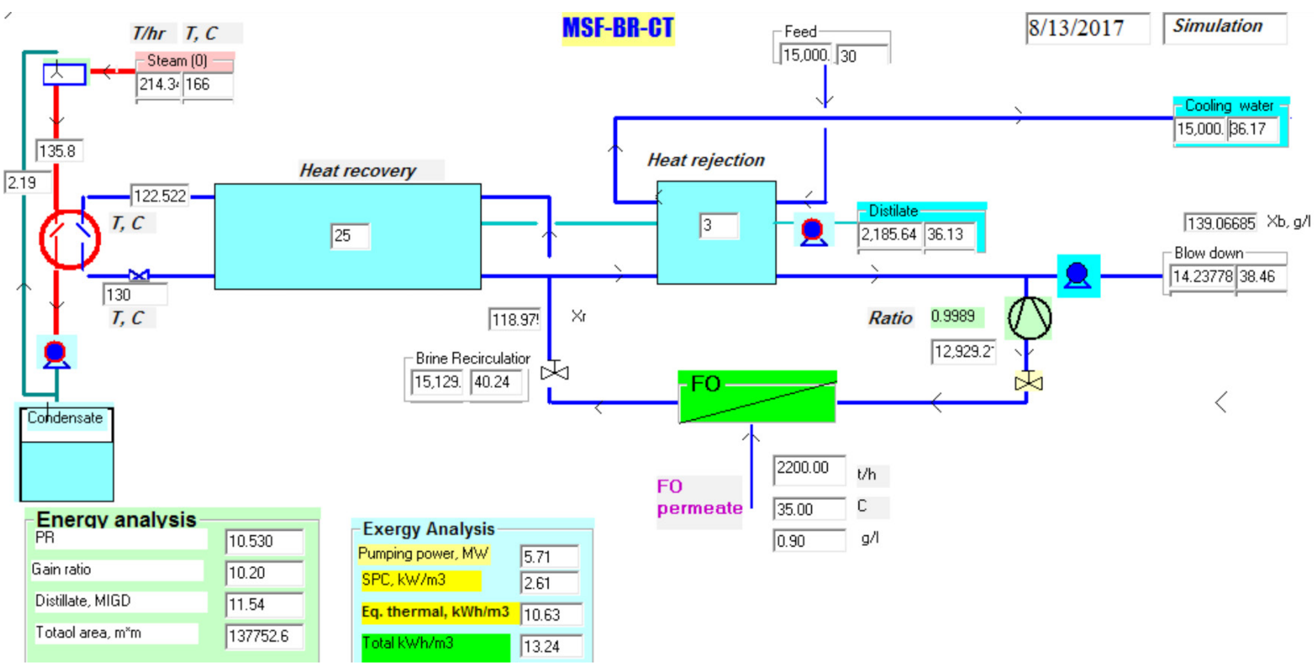


Fig. 4. The interface of the VSP for the integrated RO-FO-MED desalination plant.

The simulation is performed at varying the recovery ratio of the RO system from 15 to 35% and the recovery ratio of the FO system is calculated accordingly and presented in Fig. 5b. The results show that as the recovery ratio of the RO system increases, the recovery ratio of the FO membrane decreases and vice versa as shown in Fig. 5b. This is attributed to the increase in the RO brine salinity as at RO recovery ratio increases, thus, decreasing the driving force across the FO membrane leading to a decrease in the FO recovery ratio.

The overall process recovery ratio of the proposed tri hybrid RO-FO-MSF configuration is the resultant value of the recovery ratio of the single pass RO, FO and MSF.

The recovery ratio of a single pass RO, R_1:

$$R_1 = \frac{D_1}{F} \tag{18}$$

The recovery ratio of the FO system, R_2 is calculated as:

$$R_2 = \frac{D_2}{F - D_1} \tag{19}$$

The recovery ratio of the MSF, R_3 system is calculated as:

$$R_3 = \frac{D_3}{D_2} \tag{20}$$

The overall recovery ratio is calculated as:

$$R = \frac{D_1 + D_3}{F} = \frac{D_1}{F} + \frac{D_3}{F} = R_1 + \frac{D_3}{F} \tag{21}$$

Substituting from Eq. (23) into Eq. (24)

$$R = R_1 + \frac{D_2 R_3}{F} \tag{22}$$

Substitute from Eq. (22) into Eq. (25)

$$R = R_1 + \frac{(F - D_1) R_2 R_3}{F}$$

$$R = R_1 + R_2 R_3 \left(1 - \frac{D_1}{F}\right)$$

$$R = R_1 + R_2 R_3 (1 - R_1) \tag{23}$$

Fig. 5a shows the overall tri-hybrid process recovery ratio is maintained at 45% as regardless of the change in recovery ratio of the RO and FO sections. When the seawater feed salinity is 45 g/l and the allowable brine concentrate of the RO system is 69, thus, the maximum recovery ratio of the RO system, R_1 = 35% as calculated from Eq. (18). Having set the brine salinity exit from the FO section at 80 g/l, the maximum recovery ratio of the FO section, R_2 = 15% as calculated from Eq. (19). By using Eq. (20) and referring to Fig. 4 the recovery ratio of the hybrid MSF-FO is calculated as 99.3 %. By using Eq. (23), the overall recovery ratio of tri-hybrid RO-FO-MSF is calculated as 45%. The same overall recovery ratio of 45%, can also be obtained at 15% RO recovery ratio and corresponding 35% FO recovery ratio. Eq. (23) can be applied at any recovery ratio values, which is in agreement with the simulation results presented in Fig. 5b.

Fig. 5b shows that the specific energy consumption decreases as the RO section recovery ratio increases mainly due to the decrease in RO section power consumption at higher recovery ratio. Fig. 5b shows that the lower specific total energy consumption = 5.7 kWh/m³ at RO recovery ratio 35% and corresponding FO recovery ratio of 15%.

Fig. 6 shows the specific membrane surfaces are of RO section decrease as the RO section recovery ratio increase this is attributed to the increase the production while the area is kept constant. The specific heat transfer area of the MSF is kept constant at 65 m²/(m³/h) due to the fixed temperature range across the MSF (130°C – 40°C = 90°C) and the fixed MSF recovery ratio of 98.8%. The RO brine flow rate decreases with the RO recovery ratio decreasing the FO-MSF production. Fig. 9 also shows that the FO specific membrane area decreases with the increase of the RO recovery ratio due to the decrease of the FO recovery ratio with increasing the RO recovery ratio. Figs. 5b and 6 clearly indicate the tri hybrid process would be more economical when operating at the maximum RO recovery ratio of 35% and corresponding while FO recovery ratio of 15%.

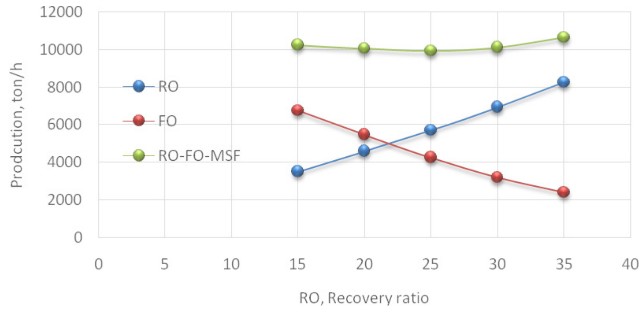


Fig. 5a. Production of tri hybrid RO-FO-MSF.

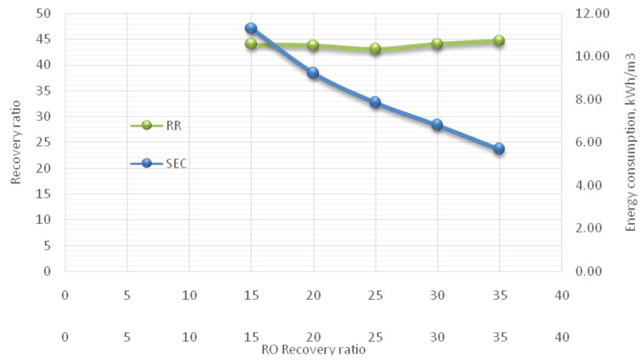


Fig. 5b. Recovery ratio of tri hybrid RO-FO-MSF.

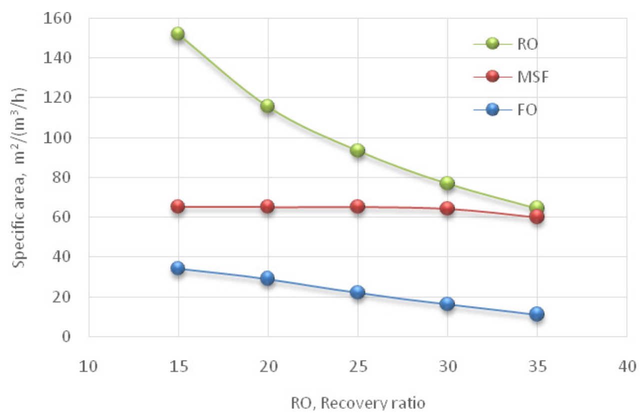


Fig. 6. Specific heat transfer and membrane area.

Table 2 provides a comparison between the tri hybrid RO-FO-MSF (RO recovery ratio of 35% and correspond-

Table 2
Technical and cost analysis of MSF, RO and hybrid RO-FO-MSF processes

	MSF (ref.)	RO (ref.)	RO-FO-MSF (35% RO + 15% FO)
Capacity, MIGD (t/h)	42.5 (8075)	42.5 (8075)	56 (10,640)
SEC, kWh/m^3	17	4.7	5.7
Feed flow rate, t/h	24,000	24,000	24,000
Intake flow rate, t/h	72,000	24,000	24,000
Outfall flow rate, t/h	63,925	15,925	13,360
Outfall salinity, g/L	51	70	80
Recovery ratio	34	34	45 (+30%)
Heat transfer area, m^2	468,363	–	137,752
Membrane area, m^2	–	670,200	RO(529,200) + FO(45,920)

ingly the FO recovery ratio of 15%) and the standalone RO (42 MIGD) and MSF (3×14 MIGD). The recovery ratio of the tri hybrid RO-FO-MSF configuration is 30% higher than both RO and MSF standalone processes because directing the RO brine to the FO feed side while the draw solution side in the MSF is maintained at higher concentration. Due to the concentration difference between draw water side and feed water side, the pure water transferred from the low concentration (feed side) to the high NaCl concentration side (draw solution side). This means FO extracts the maximum allowable water from the RO brine. The extracted pure water will be recovered in the MSF evaporator. Therefore, the tri hybrid configuration extract maximum pure water within the process limit of RO and implementing new operating conditions for MSF, which overcome its limitation by circulating brine with scale free ions. Table 2 shows that the specific total energy consumption, including the equivalent thermal energy of the tri hybrid RO-FO-MSF is 65% lower than the stand alone MSF and the recovery ratio is 30% higher. Moreover, the specific energy consumption of the tri hybrid RO-FO-MSF is only 20% higher than RO.

4.2. Cost analysis

The purpose of the cost analysis is to analyse the economic performance of the MSF, SWRO and the proposed RO-FO-MSF configuration. The lifecycle cost of the water production for the proposed tri hybrid configuration (RO-FO-MSF) is calculated and compared with referenced MSF and RO desalination plants. The cost parameters are based on either cost data from similar recent project or available market recent material costs [1, 23–25].

4.2.1. Cost analysis of Standalone MSF

Capital cost estimation was carried using the VSP simulator and the results are provided in Table 3 and graphically presented in Fig. 7a. The breakdown of the capital cost associated with MSF desalination plant consisting of the three evaporators with total capacity of 42.5 MIGD indicates the evaporator represents the highest capital cost item at 27% of the total capital investment (TCI). The evaporator cost includes the tubes, tube sheet, tube support, shell and manufacturing cost associated with the brine heater, 19 stages and de-aerator. The Titanium tubes are specified for tubes with average price of 25,000 $/ton [24]. The cost of the mechanical components, which includes the circulation pumps, represents

Table 3
Cost analysis of MSF desalination plants (3 × 14 MIGD)

Capital cost breakdown		
	Item	US $
1	Evaporators	58,426,433
2	Mechanical parts	21,272,808
3	Electrical parts	17,936,299
4	Civil work &building	25,921,136
5	Intake/outfall	20,736,000
6	Potable Tanks	22,505,744
7	Indirect cost	52,229,155
	TCI	219,027,576
Operational cost break down		
	Item	$/h
1	LP steam cost	4118
2	Electricity	1064
3	Chemicals	41
4	Spare parts	331
5	Labor	596
6	Insurance	331
	Total	6482
Levelized cost analysis		
1	Interest rate	0.07
2	Life span	25.00
3	Amortization	0.086
4	Yearly cost, $/year	6,264,956
5	Availability	0.97
6	Hourly rate, $/h	737.30
7	CAPEX, $/m^3	0.27
8	Opex, $/m^3	0.800
9	Total unit water cost, $/m^3	1.08

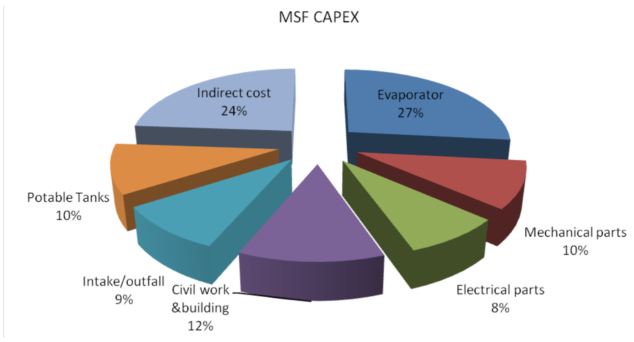


Fig. 7.a. Capital cost breakdown of 3 × 14 MIGD MSF desalination plant.

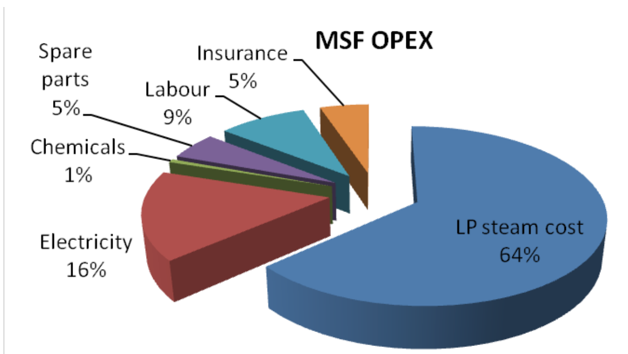


Fig. 7.b. Operation cost break down of 3 × 14 MIGD MSF desalination plant.

10% of the TCI and the cost of the electrical components corresponding to the running pumps, instrumentation and control represents 8% of the TCI. The civil work and building and site preparation represents 12% of the TCI while the intake/outfall construction and facilities represent 9% and the portable storage tank represents 10% of the TCI. The indirect cost includes the engineering, project development and tax and financing accounts to 24% of the TCI. The operating costs are calculated using the VSP simulator assuming oil price of 16 $/barrel, cost of the low pressure steam directed to the MSF is 3.0 $/ton steam, and electricity cost is 0.03 $/kWh and the results are provided in Table 3 and presented graphically in Fig. 7b. The heating steam is priced based on the fuel allocation cost and common equipment among electrical power side and desalination side using exergy method [23]. The breakdown of the operating costs associated with the 42.5 MIGD MSF desalination plant indicates the cost of the heating steam represents 64% of the total operating cost, the electricity costs represent 16% of the total operating cost, and the chemicals, labour, spare part and insurance accounts for 1%, 9%, 5% and 5% of the total operating costs, respectively. The water unit cost of the standalone MSF is calculated by levelized capital and operating cost as shown in Table 3 assuming 7% interest rate, 25 years plant life. Accordingly, the specific capital cost = 0.274 $/$m^3$ and the OPEX = 0.8 $/$m^3$. Therefore, the total water unit cost of the standalone MSF is 1.074 $/$m^3$.

4.2.2. Cost analysis of standalone RO

Al Guabrah 42.5 MIGD RO desalination of 42.5 MIGD capacity located in Sultanate of Oman recently commissioned in 2015 is taken as a case study [1]. As with many plants in the Gulf region, protection against red tides and oil spills is of significant concern for feed water quality and the first line of protection against such events is the subsurface intake. The 1.65 km long intake installed solely for this plant takes feed water through a passive screen filter at a depth of 10 m. The construction required careful planning as significant dredging was required for the installation of the new intake, including around an existing and operational brine outfall and conveyance pipeline from the neighbouring desalination plant. The second line of defence is the pre-treatment system; a dissolved air floatation (DAF) system followed by anthracite sand dual media filtration is installed prior to the RO system [1]. The plant employs only 7 operating RO skids, with an extra 8th skid available as a backup for maintenance or emergency situations. The RO plant design requires two-pass system to achieve the stringent boron rejection limits of <0.5 mg/L imposed by the Omani government. The procurement model is based on BOO (20 years) and the capital cost invested = $300 million [1].

The VSP simulator performed cost breakdown, which is verified against the recent commercial plant of Al-Gubrah as shown in Table 4 and Fig. 8a. The costs associated with the pretreatment section represent 10% of the TCI while the RO modules represent only 5% of the TCI. The costs associated with the mechanical components represent 18% of the TCI and the electrical components accounts for 16% of the TCI. The civil work represents 12% while the intake/outfall construction represents 13% of the TCI and the indirect costs including the engineering, R&D, financial are 19% of the TCI.

The operating cost break down was calculated assuming electricity price of 0.03 $/kWh for standalone RO as the same value used for MSF and the results are provided in Table 7 and Fig. 8b. The electricity cost represents the major OPEX component with 27% of the total OPEX while chemicals, labour, and membrane replacement cost accounts for 34, 15, and 16% of the OPEX. The water unit cost generated by RO is calculated by levelized capital and operating cost and presented in Table 4. Using 7% interest rate and 20 years for life plant, the specific capital cost = 0.42 $/$m^3$ and the OPEX = 0.47 $/$m^3$ such that the total water unit cost of the standalone RO plant is 0.89 $/$m^3$ as shown in Table 4.

4.2.3. Cost analysis of the RO-FO-MSF tri hybrid

Table 5 shows the capital investment of the tri hybrid FO-MSF-RO process (RO recovery ratio of 35% and correspondingly FO recovery ratio is 15%). In the present study, the cost of FO membrane is assumed 40% higher than the relevance brackish water reverse osmosis BWRO element to consider the manufacturing process and packing densities. The cost of FO pressure vessel, is assumed to be equal the cost of the pressure vessel of BWRO even the FO pressure vessel is designed for 3 bar, however, BWRO PV is designed for 20 bar. The capital cost of the FO-MSF section represents 15% of the TCI and that of the RO section represents the

Table 4
Cost analysis of RO desalination plant

Capital cost breakdown			
No.	Items	USD	%
1	Pretreatment	28,810,922	10
2	RO modules	16,616,000	5
3	Mechanical parts	54,956,788	18
4	Electrical parts	47,057,839	16
5	Civil & building	36,877,980	12
6	Intake/outfall	38,414,563	13
7	Potable tank	22,346,527	7
8	Indirect cost	57,429,771	19
	TCI	302,510,389	100
Operational cost breakdown			
1	Item	$/h	
2	Parts	346	
3	Chemicals	958	
4	Labour	623	
5	Replacement	459	
6	Electricity	1,042	
7	Insurance	346	
	Total	3,775	
Life cycle cost analysis			
1	Interest rate	0.07	
2	Life time	20	
3	Amortization	0.094392926	
4	Year payment, $/year	28,554,840.71	
5	Hourly payment, $/h	3431.247382	
6	CAPEX, $/m^3	0.42	
7	OPEX, $/m^3	0.47	
8	Total, $/m^3	0.89	

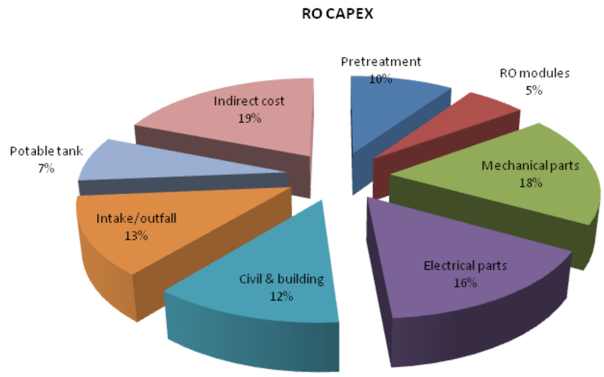


Fig. 8a. Capital cost break down of RO desalination plant (Al Gubrah, Oman).

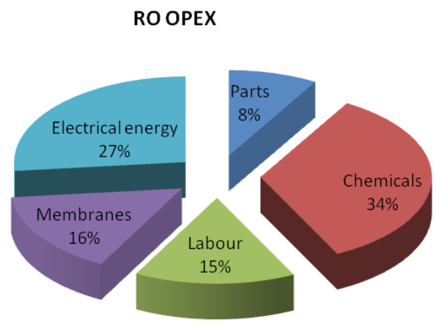


Fig. 8b. Operating cost breakdown of RO desalination plant (Al Gubrah, Oman).

Table 5
Cost breakdown of Tri hybrid FO-MSF-RO configuration

Capital cost, $	
FO-MSF	43,780,777
RO-single pass	239,535,176
Capital investment, TCI	283,315,953
Life cycle cost	
Interest rate	0.07
Life time	20
CAPEX, $/m^3	0.30
Operating cost	
RO, $/h	3,554
FO-MSF, $/h	1,119
OPEX, $/m^3	0.41
Unit water cost, $/m^3	0.71

remaining 85% of the TCI. This does not come as surprise as the RO section produces 80% of the overall water production. Using 7% interest rate and 20 years life cycle, the levelized capital cost = 0.3 $/m^3 and the OPEX = 0.41 $/m^3. Hence, the water unit cost of the RO-FO-MSF tri hybrid is 0.71 $/m^3.

4.2.4. Cost comparison

As shown in Fig. 9, the specific capital cost is 0.28, 0.43 and 0.3 $/m^3 for MSF, RO and the tri hybrid ROWFO-MSF, respectively. The specific capital cost of the tri hybrid is lower than RO but slightly higher than that of the MSF. This is explained by the higher recovery ratio of the tri hybrid process (30% higher) and the reduction of the civil work and construction of intake/outfall by up to 50% the tri hybrid process only requires intake for the RO section. Moreover, the product blend of the MSF distillate and RO permeate eliminates the need for a second RO pass and allows the use of a single pass RO, which reduces the capital investment cost. Fig. 9 shows that the operating cost (Electricity, heating steam, chemicals and labour) is 0.8, 0.47 and 0.41 $/m^3 for MSF, RO and the tri hybrid RO-FO-MSF, respectively. The OPEX of the tri hybrid is lower than MSF due to the lower heating steam consumption of the tri hybrid process. The OPEX of the tri hybrid is also lower than that of RO even there is heating steam used for driving the MSF

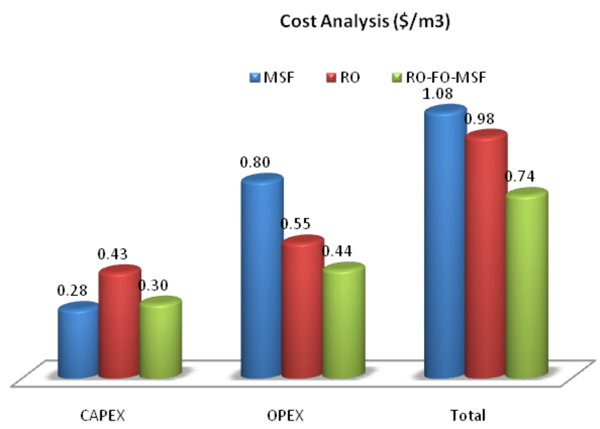


Fig. 9. Specific unit water cost for MSF, RO and RO-FO-MSF.

section because of the higher production (recovery ratio) for the same feed.

As shown in Fig. 9 the total specific water unit cost is 1.08, 0.89 and 0.71 $\$/m^3$ for the MSF, RO, and the tri hybrid process, respectively. The water unit cost of the tri hybrid is 40 % lower than that of the standalone MSF and 20% lower than that of the standalone RO plant.

5. Conclusions

A novel tri hybrid RO-FO-MSF desalination process is proposed, modelled and simulated. The new configuration provides a solution to the process limited recovery ratio of the existing RO and MSF desalination processes, in particular where seawater salinity and boron content are relatively high. Well-verified Visual Design and Simulation software is used to prototype and perform process simulation and carry technoeconomic analysis of the tri hybrid system. The technoeconomic analysis shows the superiority of the tri hybrid RO-FO-MSF over standalone RO and MSF especially in the GCC countries where high salinity and boron concentration characterize the seawater feed leading to reduction of the water unit cost by 40 and 20% over standalone MSF and RO, respectively, and requires 6.5% less capital investment than standalone RO at the same production rate.

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Appendix A

Process recovery ratio limitation of SWRO

In order to show the process recovery limitation, commercial RO desalination plant (Al-Gubrah, Oman) of two-pass configuration as shown in Fig. A1 is simulated. The first pass consists of 7 trains; each train contains 270 vessels. Each vessel contains 7 SWRO membrane elements. The second pass is used to reduce the permeate salinity and decrease the boron effluent. In the second pass, the BWRO membranes are arranged in two stages. The first stage of consists of 7 trains each contains 30 vessels and the second stage consists of 7 trains each contains 20 vessels. The permeate blend stream which has bypassed the second pass is set to zero to reduce the permeate salinity up to a comparable level of the MSF product. The process simulation shows that, at seawater salinity of 45 g/L, the first pass recovery ratio is calculated as 0.38. The recovery ratio of the second pass is 0.91. This arrangement reveals with permeate salinity of 27 ppm, which is almost the same range of thermal desalination product salinity. Under these operating conditions, the specific electric energy consumption is calculated as 4.75 kWh/m^3 and the overall RO plant recovery ratio is 0.33 which explained by applying mass balance around plant streams, as shown in Fig. A.2:

The recovery ratio in the first RO pass is calculated as:

$$R_1 = \frac{D_1}{F} \textit{ This lead to } D_1 = FR_1 \tag{A.1}$$

The recovery ratio of the first stage in the second pass:

$R_2 = \dfrac{D_2}{D_1}$, by substitution from (A.1), which leads to:

$$D_2 = R_2 R_1 F \tag{A.2}$$

The recovery ratio of the second stage in the second pass:

$$R_3 = \frac{D_3}{B_2} \;\&\; B_2 = D_1 - D_2$$

$$D_3 = R_3 B_2 = R_3(D_1 - D_2) = R_3(FR_1 - FR_1R_2) = FR_1R_3(1 - R_2) \tag{A.3}$$

Applying mass balance around mixer, as shown in Fig. A.2, the feed to the first pass is the sum of the seawater feed and the brine recycle of the second stage of the first pass:

$$F = F_0 + B_3 = F_0 + (B_2 - D_3) = F_0 + (D_1 - D_2 - D_3) \tag{A.4}$$

$$F_0 = F - (D_1 - D_2 - D_3) = F\big(1 - (R_1 - R_2R_1 - R_1R_3(1 - R_2))\big) \tag{A.5}$$

The seawater feed as a function of the subsystem recovery ratio is calculated as:

$$F_0 = F\big(1 - (R_1 - R_2R_1 - R_1R_3(1 - R_2))\big) \tag{A.6}$$

The overall system recovery (R) is calculated as:

$$R = \frac{D_2 + D_3}{F_0} = \frac{R_2R_1F + FR_1R_3(1 - R_2)}{F\big(1 - (R_1 - R_2R_1 - R_1R_3(1 - R_2))\big)} = \frac{R_2R_1 + R_1R_3(1 - R_2)}{1 - R_1 + \big(R_2R_1 + R_1R_3(1 - R_2)\big)} \tag{A.7}$$

Using the simulation results presented in Fig. A.1, by substitution $R_1 = 0.38$, $R_2 = 0.53$ and $R_3 = 0.38$, into Eq. (A.1), so that the overall recovery ratio is calculated as 0.3 as presented as follows:

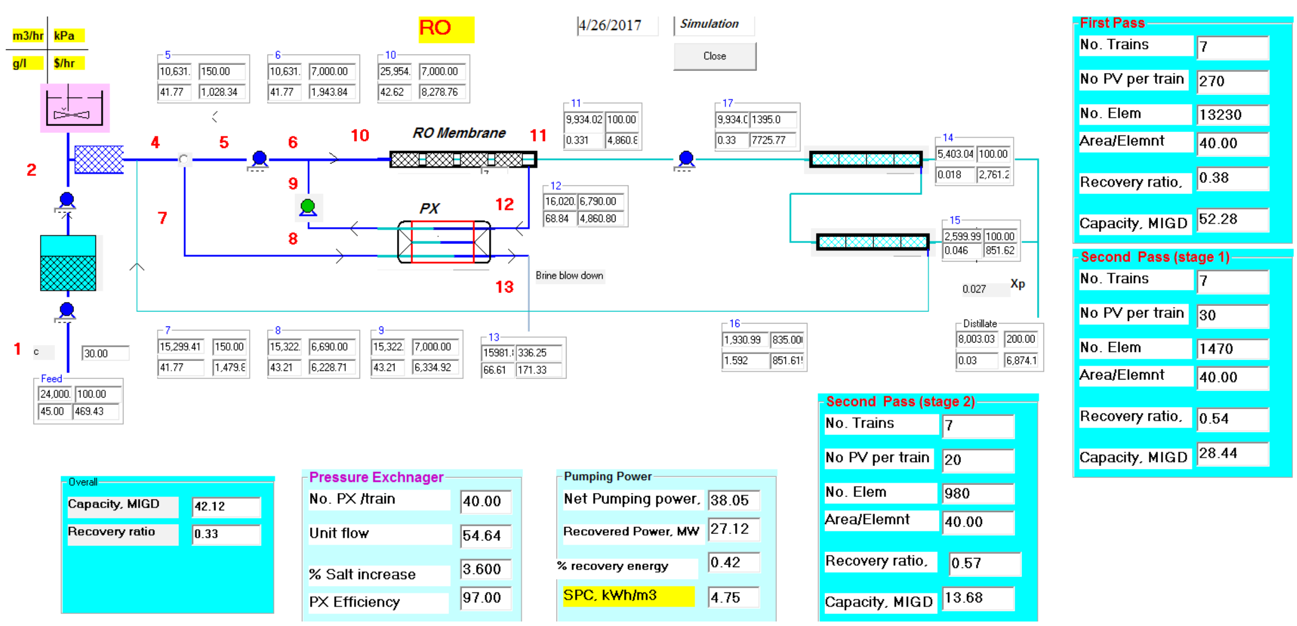

Fig. A.1. Interface of the VSP software of RO plant (Al-Gubrah, Oman).

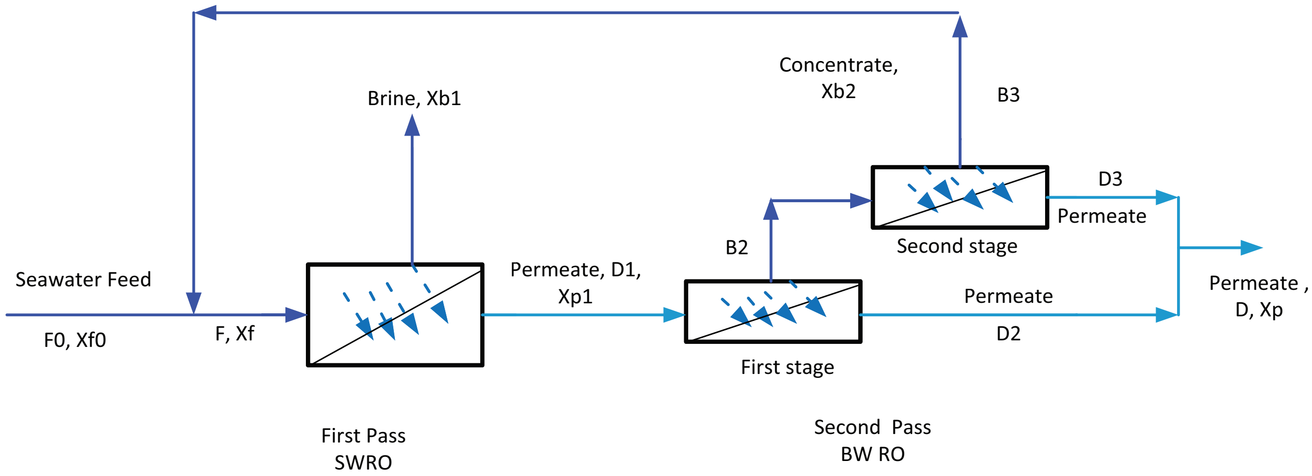

Fig. A.2. Process flow diagram of two passes RO.

$$R = \frac{0.38 \times 0.53 + 0.38 \times 0.38\left(1 - 0.53\right)}{1 - 0.38 + \left(0.38 \times 0.53 + 0.38 \times 0.38\left(1 - 0.53\right)\right)} = 0.3$$

The analytical calculation is in agreement with the simulation results and shows that the process recovery ratio of RO under the GCC is limited to 0.3.

Process recovery ratio limitation of MSF

In the current operating MSF plants, and to allow for a safety margin of the sulphate scale deposition limits, the TBT is limited to 110°C [21]. The sulphate scale deposits, once their solubility limits are exceeded, have a direct influence on the efficiency of heat transfer across the tube heat exchanger which consequently affecting the process performance [22].

The Brine Recycle (MSF-BR) evaporator of 15 MIGD (68,400 m³/day) distillate capacity (Rass Laffan, Qatar) is considered as a reference plant for the present work calculation as shown in Fig. A3. The heat recovery section consists of 16 stages while the heat rejection section consists of three stages. The gain output ratio (GOR) is calculated as 8 (kg distillate/kg steam). The specific electricity consumption is 4.2 kWh/m³, while the equivalent thermal energy is 13.5 kWh/m³ thus the total energy consumption is 17.5 kWh/m³ [23]. The makeup feed is specified to be 9,100 m³/h which is about 3 times the unit production. The recovery ratio becomes 0.317 in order to avoid scaling deposition on the tube's surface. This can be explained as follows:

The salt balance around the MSF evaporator yields the following equation:

$$M \times x_f = B \times x_b \tag{A.8}$$

where M is the makeup to the MSF and B is the brine blow down. x_f and x_b denote the salinity of the feed and the brine, respectively.

The overall mass balance:

$$M = \left(B + D\right) \tag{A.9}$$

By substituting B from Eq. (20) into Eq. (19), the process recovery ratio is calculated as follows:

$$R = \frac{D}{M} = 1 - \frac{x_f}{x_b} \tag{A.10}$$

Mass balance around the brine mixer yield:

$$M_r x_r = M x_f + (M_r - D - B) x_b \tag{A.11}$$

Rearranging Eq. (A.11), we get:

$$\frac{M}{M_r} = \frac{x_b - x_r}{x_b - x_f} \tag{A12}$$

we can get:

$$\frac{D}{M_r} = 1 - \frac{x_r}{x_b} \tag{A13}$$

In order to avoid scale deposit, the recycle salinity, x_r should not exceed 60 g/l, which is the practical allowable concentration at TBT = 110°C [22]. The heat balance around the heat recovery section gives:

$$D \times \lambda = \left(Mr\right) \times C_p \times \left(TBT - Tn\right) \tag{A14}$$

where λ is the latent heat (2326 kJ/kg), Tn is the temperature of the brine at the last stage (42.6°C), and C_p is the specific heat (~4 kJ/kg °C), thus for $\frac{D}{M_r} \cong 0.11$. The maximum allowable brine salinity, at seawater feed salinity of 45 g/l, the maximum recovery ratio, $\frac{D}{M} = 0.33$ which is close to the simulation results of 31.7 as shown in Fig.A3.

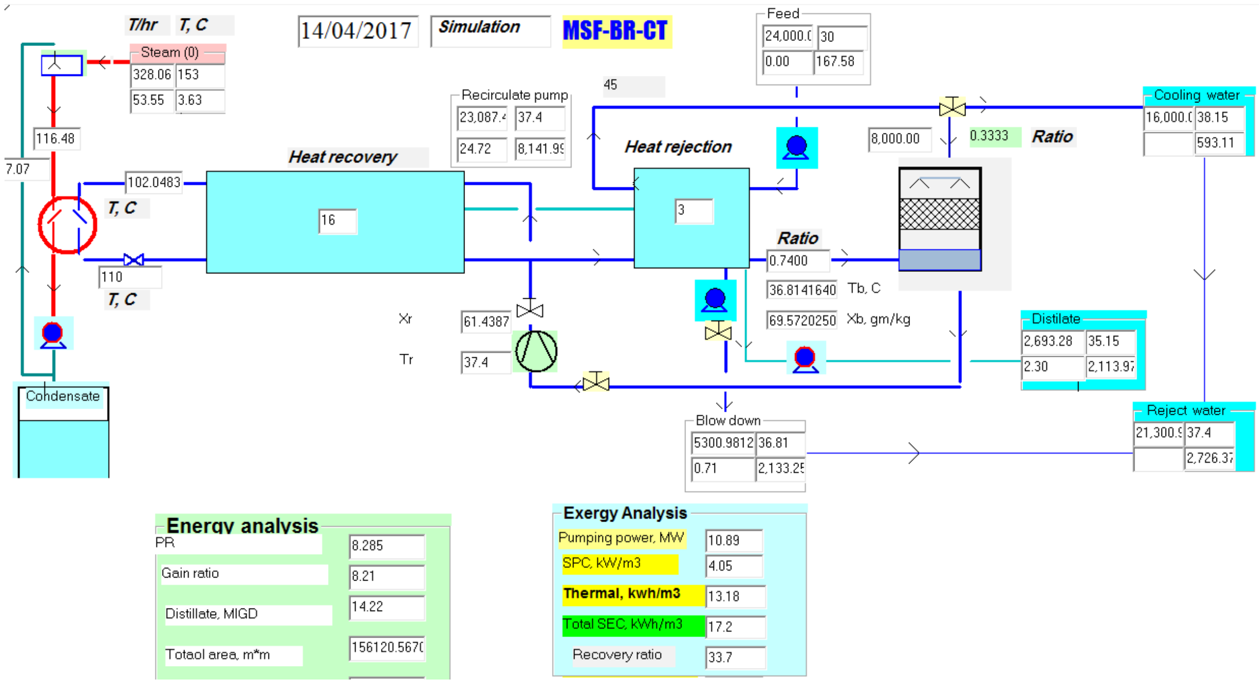

Fig. A.3. The VSP interface of the Rass Laffan (Qatar) MSF-BR unit.