Integration of Multi Effect Evaporation and Membrane Distillation Desalination Processes for Enhanced Performance and Recovery Ratios

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Abstract

This work focuses on the integration of multiple effect evaporation (MEE) and membrane distillation (MD) processes. The MEE brine is treated by the MD allowing at the same time a preheating of the MEE feed water. Different structures of integration are assessed theoretically. Simulation models based on the heat and mass transfer balances are developed for the parallel feed (PF), parallel cross (PCF), forward feed (FF)-MEE, and direct contact MD (DCMD) configurations. The main performance parameters such as the performance ratio, recovery ratio, and specific energy consumption are investigated via simulation of various MEE-MD configurations and operating parameters. The results show that hybrid MEE and MD structures outperform the standalone MEE systems. The performance ratio increased by more than 25\% when hybrid structure is used as compared to standalone MEE. The results revealed also that the hybrid parallel PF-MEE-MD configuration yields the best performance compared to other hybrid configurations. It also outperforms the standalone PCF-MEE which is widely adopted and used in practice. Besides, the performance of the hybrid configurations is especially sensitive to the feed temperature’s variation in both MEE and MD units. Increasing MD feed temperatures has a positive impact on the recovery ratios of the combined systems.

Keywords: Desalination, Multiple Effect Evaporation, Membrane Distillation, Hybrid system, Performance ratio, Recovery ratio.
1. Introduction

Thermal-based desalination processes, mainly multi-stage flash (MSF) and multiple effect evaporation (MEE), are still dominating the desalination market in the Arabic Gulf Cooperation Council (GCC) countries, including the Kingdom of Saudi Arabia (KSA) because they have proven to be robust, given that Gulf seawater quality has high salinity and biological activity, for meeting specific water quality requirements [1]. However, thermal processes are known for their high energy consumption and low water recovery [2]. More recently, several seawater reverse osmosis (SWRO) plants have been installed in the Arabian Gulf but they are facing several operational challenges due to the low performance of the pretreatment process especially during red tide events where some SWRO plants have been forced to shut down for several weeks [3]. Therefore, in order to enhance the performance of the conventional desalination processes, several attempts have been made such as developing new types of RO membranes, better utilization of the supplied energy and appropriate management of the discharged brine [4]. Integrating two or more desalination processes has shown interests among the academic and industrial desalination communities [5,6,7]. Helal [5] discussed several hybrid structures integrating two or more desalination processes and outlined the main advantages and limitations of each structure. Hamed [6] presented some commercially available hybrid desalination plants and provided an overview on nano-filtration NF/MSF and NF/RO/MSF hybrid structures. Zak and Mistos [8] proposed several hybrid thermal-thermal desalination structures and assessed them in terms of performance ratio (PR), recovery ratio (RR) and specific heat transfer area (sA). The results of their simulations showed that the MSF-MEE system exhibited larger PR and RR values as compared to the standard MSF-once through (OT) system for the same operating conditions and the same number of stages and effects.

There exist several hybrid conventional desalination plants integrating thermal (MSF or MEE) and reverse osmosis (RO) processes. The main examples are the AlFujairah-2 (United Arab Emirates) and Ras Alkhair-1 (KSA) plants. The AlFujairah-2 plant with a total capacity of 591,000 m$^3$/day combines MEE-TVC and RO [9], while Ras Alkhair plant integrates MSF and RO. The total capacity of the Ras Alkhair plant reaches about 1.036 million of m$^3$/day with the MSF’s capacity being 727,203 m$^3$/day and RO capacity being 309,061 m$^3$/day [10].
The following advantages could be achieved when hybrid desalination processes combining thermal and membrane technologies are used [5,6,8]:

- The hybrid configurations provide more flexibility in operation.
- Larger capacities can be reached.
- Lesser specific energy consumption and lesser gas emissions.
- Smaller intake and outfalls structures: lower environmental impacts (brine discharge).
- Better water and power matching for dual purpose power and desalination plants.

This work focuses on investigating the use of the MD process to treat thermal desalination brines, mainly from the MEE plants to reduce the temperature of the brine discharge to the acceptable levels and enhance the overall process performance. We propose the integration of the MEE with the MD because the former is by far the most economical and efficient thermal-based mature process and the latter is a promising emerging technology which is able to treat desalination brines without a significant decrease in its performance [7,11-13].

Ghaffour et al. [7] presented a critical review on membrane distillation (MD) hybrids for water production and highlighted the growing interest gained by hybridization in desalination for energy saving and process improvement. Minier-Matar et al. [13] evaluated the application of MD to desalinate concentrated brines from thermal plants. Two MD pilot units, namely vacuum multi-effect MD (unit A) and single effect air gap MD (unit B) were selected for the field evaluations. Both units were operated in parallel to treat brines (TDS = 70 g/kg) from the local MSF facility. The results have shown that the MD is a feasible technology to produce potable water from concentrated brines. Besides, unit A was found to be able to desalinate the brines with a stable permeate flux of 6.2 LMH and high quality product (99.9 % rejection). Gonzalez-Bravo et al. [14] proposed an optimization approach for the selection of existing and emerging desalination processes based on their heating and cooling requirements. Various desalination techniques have been considered in this study including MEE, MSF and RO. The results showed that the integration of MD with MEE or MSF has attractive economic and environmental benefits compared to MD combined with other desalination technologies, such as Reverse Osmosis (RO) and Vapor Compression (VC). In addition, the MEE-MD hybridization appears to have higher total annual profit compared to MSF-MD integration.
The use of MD integrated with other processes particularly RO has been investigated in several studies [12, 15-22]. Criscuoli and Drioli [16] conducted an energy and exergy analysis of an integrated system coupling RO, MD and NF modules where the MD unit operated by utilizing RO brine as a feed while the NF unit was used for the RO feed pretreatment. The authors concluded that such integrated systems represent an attractive alternative to RO and to conventional thermal desalination processes. El-Zanati and El-Khatib [18] proposed an integrated system composed of NF, RO and VMD processes. The authors observed that the overall system’s recovery during seawater treatment was increased from 30% - 35% to more than 76% when using standalone RO and the integrated system, respectively. Pangarkar et al. [19] presented a review on combining RO and MD processes for desalinating groundwater. The use of MD as a process for crystallization of RO brines has been suggested by Ji et al. [20]. The authors reported a water recovery ratio of as high as 90%. In order to overcome the low MD recovery ratio (RR), Ali et al. [21] investigated various design configurations of MD-RO hybrid systems including brine recycling and cascading arrangements. The authors also performed the cost analysis and found that brine recycling considerably improved the recovery rate to up 40% at an energy cost of 0.9 $/m$^3$. To increase the energy efficiency of the MD process, Swaminathan et al. [22] proposed to combine it with a mechanical vapor compression (MVC) unit by replacing the preheater heat exchangers by the MD system. The results showed that the use of the hybrid MVC-MD system could decrease the cost of the produced water by 6% as compared to a standalone MVC system under the same operating conditions. Another interesting hybrid system proposed by Kim et al. [23] consisted of integrating a vacuum MD system with adsorption desalination (AD) cycle where the low vacuum pressure naturally created by adsorbents during the adsorption cycle drives the VMD without the need for a vacuum pump. The authors reported that water RR and water production increased by 23% and 21%, respectively as compared to a standalone VMD unit by using an external vacuum pump to drive the process.

It is worth to mention that the studies on MEE-MD process are rare and most of them are conducted theoretically. De Andres et al. [24] conducted one of the earlier studies on the MEE-MD integration. The authors investigated experimentally the performance of a combined single-effect evaporator (SEE) and a DCMD module. The hot brine rejected from the SEE was used to feed the MD module. This integration permitted an increase of the distillate production rate by 7.5% and
an improvement of the energy efficiency of about 10%. Mabrouk et al. [25] proposed an experimental evaluation of a hybrid two effects MEE–MD driven by solar energy. The brine of each MEE effect is used to feed a MD module. Only preliminary results showing overall the benefits of the hybrid system over the standalone units have been presented. Bamufleh et al. [26] proposed and applied an optimization approach to the design of MEE-MD integrated system coupled to an industrial facility. The MEE brine was further treated by the MD. Several parameters including the preheating temperature of the brine supplied to the MD, the membrane area and the extent of the distillate water were optimized. A case study with four scenarios illustrated the applicability and the usefulness of the developed optimization method. The case study was concerned with the design of a MEE plant and an associated MD module as a brine treatment system having a combined capacity of 3850 m$^3$/day from a seawater with 35 g/kg salinity. Besides, a cost analysis giving the minimum combined cost of the MEE–MD coupled system was conducted. Recently, Farsi and Dincer [27] evaluated thermodynamically a geothermal based poly-generation system producing simultaneously distilled water, cooling, power and hydrogen. The MD, connected to the absorber of the cooling system, was used to produce additional fresh water while the MEE brine is rejected to the surrounding. If was found in particular that for the desalination system, the condenser in MEE and membrane module in DCMD are the most exergy destructive elements.

In summary and based on the above literature review, the following points can be drawn:

- The MD process can be considered as a feasible technology to treat desalination brines.
- The available literature on the integration of MEE or MSF with the MD process is scarce. Therefore, additional studies on various aspects of MD hybrids including energy management and efficiency, long-term effect of membrane fouling, economics and environmental evaluations are required.
- The integration of MEE and MD process appears to be very attractive compared to other combinations of MD and other conventional processes [7]. However, very few studies have focused on the performance evaluation of MEE-MD hybrid systems.

This work investigates the performance of various integrated MEE and MD configurations by using theoretical models based on mass and energy balances. The proposed hybrid structures
incorporated the MD in a simple, flexible but effective way to treat the MEE brines and preheat their feed waters. A sensitivity analysis with respect to sea water conditions and MEE effect number for the various hybrid structures has been carried out. In particular, three main hybrid configurations were proposed and compared to the conventional standalone MEE structures on the basis of important performance parameters including recovery ratio (RR), performance ratio (PR), and specific area (sA).

This work, which has as main objective of developing a frame for MEE and MD integration for seawater and brine desalination, is presented in several sections, as follows. The main conventional MEE configurations are first presented and described. The mathematical formulation of the MEE, MD and hybrid systems is then given. Systematic validation and comparison tests with available results from literature for all developed models are presented and discussed. The generated results from various simulations are then analyzed and discussed.

2. Multi Effect Evaporators Configurations

Figures 1 shows the standard MEE with different feed arrangements including forward feed (FF), parallel feed (PF) and parallel-cross feed (PCF). Besides standard MEE incorporating the feed preheaters is depicted in Figure 1d and denoted PCF-pre. It is believed that the distillate production of PCF-pre increases due to the elevation of the feed temperature $T_f$. The benefits of using the preheaters in the MEE process have been discussed in [28-30].

The main idea of this study is to propose configurations and assess their respective performances by using MD modules to obtain benefits from the exited MEE streams. This would essentially result in producing additional fresh water and reducing the volume and temperature of the rejected brine. Figures 2-4 represents hybrid MEE-MD structures. One can observe from these figures that two MD units are proposed for each structure. The first MD unit (MD1) is essentially fed by the MEE brine but also by an amount of seawater that is preheated in the MEE condenser. The excess heat of the distillate is also recovered. The second MD unit (MD2) uses the cooling water heat gained from the MEE condenser.

Figures 2-4 demonstrate the MEE-MD hybrid systems denoted as PCF-Hyb, PF-Hyb and FPCF-Hyb. In these systems, instead of rejecting the concentrated brine to the environment, it is fed to an array of MD units. These structures have dual objectives. First, the reject brine is used to
produce additional distillate in the MD array. Secondly, the application of MD will increase the temperature of the MEE feed; however, the increment of such increase will depend on the thermal energy of the MD feed itself. The number of MD units comprising the MD array equals to the ratio of the total seawater feed \((F_t)\) to the maximum allowable feed rate to a single MD module. In our study, the maximum MD feed rate was equal to 600 L/h due to specification of the pilot-plant MD module. In addition, to avoid possible damage of MD module, equal feed rate for the hot and cold streams was assumed. Therefore, the hot feed which is mainly the reject brine, was mixed with a makeup stream (called “recycle” and denoted by \(R_c\)) to compensate for the difference between the total MEE feed flow rate and the brine flow rate to be fed to the MD module. The latter varies upon the type of the hybrid configuration. For example, in the PCF-Hyb system, the effluent brine of the last effect is used while only the effluent brine of the first effect is used in the FPCF-Hyb system. In the PF-Hyb system, the combination of reject brine for all effects is used. The purpose of studying these three configurations is to find which brine stream would maximize the temperature of the MD hot feed \((T_{hf})\), hence achieving the best thermal efficiency of the whole system. The latter has a direct impact on the MD performance and on the MEE feed temperature, \(T_f\). Note that each brine stream to be used as MD feed has different temperature and mass flow rate. The hybrid MEE system is different than that of the standard system in a way that MEE feed is pre-heated in the MD compartment instead of the down condenser. The seawater which is fed to the down condenser is partially used as a makeup to the MD hot feed stream. The average temperature of the combined MEE distillates is higher than that of the makeup stream. Hence, its thermal energy is utilized to heat up the makeup stream in heat exchange HE1. The MD system comprises of an array of the equal size parallel single MD units with a 10 m\(^2\) of the effective membrane area. The hot and cold feed streams of the MD system are divided into equal parallel feed streams and each stream satisfies the inlet feed limitation of a single MD unit.

3. Mathematical Models

3.1. Mathematical Model of the MEE System

The mathematical modelling for the standard MEE system is well established in the literature [28-32]. The MEE model developed in this work is obtained based on the following assumptions:

- Temperature difference across each effect is assumed to be the same.
- The distillate is salt free.
• The system is operated at a steady state.
• Equal specific heat capacity for the feed, brine and distillate streams.
• Equal feed flow rate in each effect.
• Physical properties of water and the heat transfer coefficient vary with temperature and salinity.

The mass, salt and energy balance and heat transfer equation for the first effect is common for all configurations and given by:

\[ F_1 = B_1 + D_1 \]  
\[ X_f F_1 = X_1 B_1 \]  
\[ M_s \lambda_s = D_1 \lambda_1 + F_1 c_p (T_{b_1} - T_f) \]  
\[ M_s \lambda_s = U_1 A_1 (T_s - T_{b_1}) \]

\[ F_1, B_1 \text{ and } D_1 \text{ denote the feed, brine, and distillate flow rate of the first effect respectively while } X_f, T_f \text{ and } X_1 \text{ denote the salinity and temperature of the feed and first effect. } M_s, T_s \text{ and } \lambda_s \text{ represent the mass flow rate, temperature, and latent heat of the steam respectively. } T_{b_1}, U_1 \text{ and } A_1 \text{ stand for brine temperature, heat transfer coefficient and surface area of the first effect.} \]

For any effect \( i \) between 2 and \( n \), the last eqs. (1-4) are modified based on the feed and brine flow arrangements as follows:

For FF:

\[ B_{i-1} = B_i + D_i \]  
\[ X_{i-1} B_{i-1} = X_i B_i \]  
\[ D_{i-1} \lambda_{i-1} = D_i \lambda_i - B_{i-1} c_p (T_{b_{i-1}} - T_{b_i}) \]  
\[ D_{i-1} \lambda_{i-1} = U_i A_i (T_{v_{i-1}} - T_{b_i}) \]

For PF:

\[ F_i = B_i + D_i \]  
\[ X_f F_i = X_i B_i \]  
\[ D_{i-1} \lambda_{i-1} = D_i \lambda_i + F_i c_p (T_{b_i} - T_f) \]  
\[ D_{i-1} \lambda_{i-1} = U_i A_i (T_{v_{i-1}} - T_{b_i}) \]
For PCF:

\[ F_i + B_{i-1} = B_i + D_i \]  
\[ X_i F_i + X_{i-1} B_{i-1} = X_i B_i \]  
\[ D_{i-1} \lambda_{i-1} = D_i \lambda_i + F_i C_p (T_{bi} - T_f) - B_{i-1} C_p (T_{bi-1} - T_{bi}) \]  
\[ D_{i-1} \lambda_{i-1} = U_i A_i (T_{vi} - T_{bi}) \]

In equations (5-16), \( F_i, B_i, D_i \) denote the feed, brine, and distillate flow rate of the \( i \)th effect. \( X_i \) denotes the salinity in ppm of the \( i \)th effect. \( T_{bi} \) and \( T_{vi} \) represent the brine temperature, and vapor temperature of \( i \)th effect respectively. \( \lambda_i, U_i \) and \( A_i \) stand for latent heat, heat transfer coefficient and surface area of the \( i \)th effect.

In the above equations (1-16), the latent heat of vaporization \( (\lambda) \), heat capacity \( (C_p) \) and overall heat transfer coefficient \( (U) \) are calculated from the following correlations [28,32]:

\[ \lambda(T) = 2589.583 + 0.9156T - 4.834 \times 10^{-2}T^2 \]  
\[ C_p(T,X) = a + bT + cT^2 + dT^3 \]

Where \( a, b, c \) and \( d \) are coefficients function of the salinity \( (S) \) (expressed in kg/m\(^3\)), defined as:

\[ S = X/1000 \]  
\[ a = 4206.8 - 6.6197S + 1.2288 \times 10^{-2}S^2 \]  
\[ b = -1.1262 + 5.4178 \times 10^{-2}S - 2.2719 \times 10^{-4}S^2 \]  
\[ c = 1.2026 \times 10^{-2} - 5.3566 \times 10^{-4}S + 1.8906 \times 10^{-6}S^2 \]  
\[ d = 6.8777 \times 10^{-7} + 1.517 \times 10^{-6}S - 4.4268 \times 10^{-9}S^2 \]  
\[ U(T) = 1.9695 + 1.2057 \times 10^{-2}T - 8.5989 \times 10^{-5}T^2 + 2.5651 \times 10^{-7}T^3 \]

In eqs. (1-12), the latent heat of vaporization of the corresponding effect is computed at the saturated vapor temperature \( (T_v) \) and the overall heat transfer coefficient \( (U) \) is calculated at the brine temperature \( (T_{bi}) \). This means that solving the mass and energy balances requires the prior
knowledge of the brine and vapor temperatures in each effect. To simplify calculations, the brine temperature in each effect is computed using the following approximate distribution [28]:

The total temperature drop (\(\Delta T_t\)):

\[
\Delta T_t = T_s - (n - 1)\Delta T_i - T_{bn}
\]

(25)

Where \(n\) is the number of effects and \(T_{bn}\) is the brine temperature of the last effect. The temperature drop per effect is expressed as:

\[
\Delta T_i = \frac{\Delta T_t}{U_i \sum_{i=1}^{n} U_i}
\]

(26)

The brine temperature in each effect is given by:

\[
T_{b1} = T_s - \Delta T_1
\]

(27)

\[
T_{bi} = T_{bi-1} - \Delta T_i; \quad i > 1
\]

(28)

In eq. (25), \(\Delta T_t\) is the temperature drop due to thermodynamic losses; it is set to 2 °C for all stages. However, since \(U\) is a function of the brine temperature itself, eqs. (26-28) must be solved iteratively until the \(T_b\) in each effect reaches a constant value. Having determined the brine temperature, the vapor and condensate temperatures in each effect can be calculated as follows:

\[
T_{vi} = T_{bi} - BPR_i
\]

(29)

\[
T_{ci} = T_{vi} - \Delta T_c
\]

(30)

In Eq. (30), \(T_c\) is the distillate temperature and \(\Delta T_c\) accounts for temperature losses due to pressure losses and friction; it is set to 0.5 °C. The boiling point rise (BPR) can be computed from [33]:

\[
BPR = (0.0825431 + 1.883 \times 10^{-4}T + 4.02 \times 10^{-6}T^2)(10^{-4}X)
\]

\[
+ (-7.625 \times 10^{-4} + 9.02 \times 10^{-5}T - 5.2 \times 10^{-7}T^2)(10^{-4}X)^2
\]

\[
+ (1.522 \times 10^{-4} - 3 \times 10^{-6}T - 3 \times 10^{-8}T^2)(10^{-4}X)^3
\]

(31)
The down condenser energy balance is given as follows:

\[ D_n \lambda_n (T_{vn}) = (F_t + M_{cw}) C_p (T_f - T_{cw}) = U_c A_c LMDT_c \]  \( (32) \)

where:

\[ LMDT_c = \frac{T_f - T_{cw}}{\ln \left( \frac{T_{cn} - T_{cw}}{T_{cn} - T_f} \right)} \]  \( (33) \)

\[ U_c = 1.7194 + 3.2063 \times 10^{-3} T + 1.5971 \times 10^{-5} T^2 - 1.9918 \times 10^{-7} T^3 \]  \( (34) \)

\( F_t, M_{cw}, T_{cw} \) represent the total feed flow rate, mass rate and temperature of the cooling water, respectively. In Eqn, (32), the subscript \( n \) and \( c \) refer to the last effect and down condenser, respectively.

In the FF and PCF schemes, portion of the brine entering the subsequent effect flashes due to temperature difference. Brine flashing occurs in the second effect and onwards. The amount of flashing brine \( (d_{bi}) \) can be computed as follows:

\[ d_{bi} = \frac{B_{i-1} C_p (T_{bi-1} - T_i')}{\lambda_i (T_i')} \]  \( (35) \)

where:

\[ T_i' = T_{bi} + NEA_i \]  \( (36) \)

The non-equilibrium allowance (NEA) for each effect can be calculated using the correlation [34]:

\[ NEA_i = \frac{33 (T_{bi-1} - T_{bi})^{0.33}}{T_{vi}} \]  \( (37) \)

Some industrial MEE units contain flashing boxes where the distillate of the second effect flashes due to non-equilibrium allowance and the formed vapor is fed to the condenser side of the third
effect along with the main vapor formed by boiling in the second effect. The procedure is repeated
in all subsequent effects. The flashed vapor \(d_{vi}\) can be computed as follows [32]:

\[
d_{vi} = \frac{D_{i-1} \cdot \rho(T_{cl-1} - T'_i)}{\lambda_i(T''_i)}
\]  

(38)

where:

\[
T''_i = T_{vi} + NEA'_i
\]  

(39)

The flashed vapor \(d_{vi}\) can be computed as follows [32]:

\[
d_{vi} = \frac{D_{i-1} \cdot \rho(T_{cl-1} - T'_i)}{\lambda_i(T''_i)}
\]  

(38)

The non-equilibrium allowance (NEA') can be calculated using the following correlation [34]:

\[
NEA'_i = \frac{0.33(T_{cl-1} - T_{vi})}{T_{vi}}
\]  

(40)

It should be noted that although the distillate flashing occurs in the second effect and onwards, it
starts affecting the system from the third effect and onwards.

When brine and distillate flashing are considered, eqs. (7, 11 and 15) are modified as follows:

\[
D_{i-1} \lambda_{i-1} + d_{bi-1} \lambda_{i-1} + d_{vi-1} \lambda_{i-1} = D_i \lambda_i - B_{i-1} \cdot \rho(T_{bi-1} - T_{bi})
\]  

(41)

\[
D_{i-1} \lambda_{i-1} + d_{bi-1} \lambda_{i-1} + d_{vi-1} \lambda_{i-1} = D_i \lambda_i + F_i \cdot \rho(T_{bi} - T_f)
\]  

(42)

\[
D_{i-1} \lambda_{i-1} + d_{bi-1} \lambda_{i-1} + d_{vi-1} \lambda_{i-1} = D_i \lambda_i + F_i \cdot \rho(T_{bi} - T_f) - B_{i-1} \cdot \rho(T_{bi-1} - T_{bi})
\]  

(43)

The total distillate water production \(D_t\) of the MEE is thus:

\[
D_t = \sum_{i=1}^{n} D_i
\]  

(44)

The key performance indices for the MEE system are as follows:
The PR:

\[ PR = \frac{D_t}{M_s} \]  (45)

Where \( D_t \) is the total distillate.

The RR:

\[ RR = \frac{D_t}{F_t} \]  (46)

The specific heat transfer area (sA):

\[ sA_t = \frac{A_c + \sum_{i=1}^{n} A_i}{D_t} \]  (47)

The specific condenser energy consumption (sQc):

\[ sQ_c = \frac{(F_t + M_{cw})C_p(T_f - T_{cw})}{D_t} \]  (48a)

The amount of cooling water per unit of fresh water product in the MEE and the hybrid plants is an another important quantity to be evaluated. It is expressed as:

\[ sM_{cw} = \frac{M_{cw}}{D_t} \]  (48b)

The salt rejection ratios for the MEE (CR\textsubscript{MED}) and the hybrid system (CR\textsubscript{hyb}) are defined as follows:

\[ CR_{MED} = \frac{X_n}{X_f}, \quad CR_{hyb} = \frac{C_n}{X_f} \]  (49)

Where \( X_n \) is the water salinity in the last effect and \( C_n \) is the water salinity exiting the MD unit.

3.2. Mathematical Model of the MD system

The mathematical model of a single MD unit has been previously developed and validated in the direct contact (DCMD) pilot plant (SolarSpring, Germany) by Ali and Orfi [35]. It is briefly summarized in Appendix A. The dimensions of the DCMD module were as follows: 14 m (channel length), 0.7 m (channel width), 0.002 m (channel height). A hydrophobic Polytetrafluoroethylene
(PTFE) membrane with pore sizes of 0.2 μm and thickness of 80 μm was used in the entire series of experiments. The total effective membrane’s surface area was 10 m².

3.3. The Hybrid system

The hybrid system was modelled by slightly modifying the equations previously used to model MEE and MD processes. For hybrid system, the energy balance around the down condenser becomes:

$$D_n \lambda_n = (R_c + M_{cw}) C_p (T_{cw_0} - T_{cw}) = U_c A_c LMD T_c$$  \hspace{1cm} (50)

$$LMD T_c = \frac{T_{cw_0} - T_{cw}}{\ln \left( \frac{T_{cn} - T_{cw}}{T_{cn} - T_{cw_0}} \right)}$$  \hspace{1cm} (51)

In general, the flow rate of the makeup stream $R_c$ (figures 2-4) is set equal to total MEE distillate mass rate ($D_t$) for the PF and PCF configurations and total MEE distillate mass rate plus brine mass rate of the last effect ($D_t + B_n$) for the FPCF configuration (Figure 4). The temperature of the cooling water exiting condenser ($T_{cw_0}$) is set to $T_{vn} - 4$, hence eqs. (50-51) are used to compute the cooling water mass rate ($M_{cw}$) and condenser heat transfer area ($A_c$). The heat exchanger HE1, which is assumed to have an effectiveness of 80%, is used to heat up the makeup stream according to:

$$T_{cw_m} = T_{cw_0} + 0.8 (D_t - D_n) C_p (T_{ca} - T_{cw_0})/(R_c C_p)$$  \hspace{1cm} (52)

where $T_{cw_m}$ is the temperature exiting HE1 in the hybrid structure and $T_{ca}$ is the average temperature of the distillate collected from the various effects.

The temperature of the combined brine and makeup, (i.e. the MD hot feed), $T_{hf}$, is calculated using the following equation assuming a reference temperature of 25 °C:

$$T_{hf} = 25 + \left[ R_c C_p (T_{cw_m} - 25) + B_r C_p (T_{br} - 25) \right]/F_t C_p$$  \hspace{1cm} (53)
where \( B_r \) is the brine mass rate, i.e. \( B_n, B_1 \) and \( \sum B_i \) for PCF, FPCF and PF configurations, respectively. Similarly, \( T_{br} \) is the temperature of \( B_r \), i.e. \( T_{bn}, T_{b1} \) and average brine temperature for PCF, FPCF and PF configurations, respectively. The total number of parallel MD units (\( N \)) is estimated as:

\[
N = \left( \frac{F_t}{\rho} \right) / Q_{max}
\]

(54)

\( Q_{max} \) is the maximum allowable flow rate to be fed to the MD system.

The performance of a hybrid system is controlled by various design and operating conditions including the sea water parameters (salinity, temperature and flow rate), number of MEE effects (\( n \)), steam temperature (\( T_s \)) and mass flow rate (\( F_t \)). Moreover, the system performance is expressed in terms of various key indicators including the RR, the PR, the distillate flow rate (\( D_{tot} \)), the specific cooling water mass flow rate (\( sM_{cw} \)), the specific condenser heat consumption (\( sQ_c \)) and the heat transfer surface areas (\( sA_t \)).

4. Model Solution and Simulation

The solution algorithms for the stand-alone MEE and MD systems are shown in Figure B.1 and B.2 in Appendix B while that for the hybrid system is shown in Figure 5. In order to solve the mathematical model for the MEE system (i.e., eqs. (1-40)), the design parameters need to be fully specified and used as inputs to the model. The specified input parameters are the steam rate and temperature (\( T_s \) and \( M_s \), respectively), cooling water temperature (\( T_{cw} \)), seawater feed temperature (\( T_f \)), seawater salinity (\( X_f \)), minimum brine temperature (\( T_{bn} \)) and the total number of effects (\( n \)).

Next, either the salinity of each effect [33] or the feed flow rate of each effect [32] should be specified. In the first scenario, the salinity profile is approximated by a linear distribution and the model computes the feed, the brine and distillate mass rates and the heat transfer area for each effect. In the second scenario, the total feed rate can be equally distributed over the entire effects [32] or the feed rate of each effect is kept as the ratio of the total feed to the total formed distillate in order to keep the salinity of each effect less than the maximum reject value of \( X_{bn} \) [29]. Note that the assumption of linear salinity distribution is not realistic. Moreover, keeping the feed rate for each effect as a ratio of the distillate product cannot be implemented easily in practice. Therefore, in this work we follow the equally distributed feed rate (i.e., second scenario).
Given the MD unit characteristics, only the feed operating conditions are required to solve the mathematical model (Figure B.2). The known feed conditions are therefore the volumetric flow rates and temperatures of the hot and cold streams. The unknowns, i.e. \( T_{\text{cout}} \), \( T_{\text{hme}} \), \( T_{\text{cm}} \) are determined by the two nested loops. In the external loop, an initial value for \( T_{\text{cout}} \) is set and in the inner loop, the initial value of the membrane interface temperatures is set. Having set the initial values, the mass and energy transfer equations are solved to determine new values of the membrane interface temperature. If the new values match the older values, the inner loop stops, otherwise the iterations continue. When the inner loop converges, the energy balance is solved to estimate new value of \( T_{\text{cout}} \). When the newly estimated \( T_{\text{cout}} \) matches the older value, the outer loop stops, otherwise the iteration continues. Figure 5 represents a combination of MEE and MD algorithms to formulate the hybrid system algorithm. In this case, the MEE feed temperature is initialized. Next, the MEE algorithm is solved to determine the feed flow rate (\( F_t \)) and its temperature (\( T_{\text{hf}} \)) to the MD system. Afterwards, the MD algorithm is solved to determine a new value for the MEE feed temperature (\( T_f \)). If the new value matches the older one, the procedure is terminated, otherwise the iteration is resumed.

5. Results and Discussion

5.1. MEE Model Validation

The model developed in this paper is validated by using operation data provided by Darwish and Abdulrahim [29] for the FF and PCF MEE configurations. In both cases, brine flashing due to non-equilibrium allowance is considered and the results of model validation are listed in Table 1. As seen in Table 1, the output data calculated by our model are in a good agreement with the corresponding referenced values. The observed minor differences were due to the fact that the benchmarked data were based on constant average values for physical and design parameters such as heat capacity, latent heat, heat transfer coefficient, and boiling point raise, as well as equal temperature drop across the evaporators. To further assess the model’s validity, its output was compared with that achieved by using operational data from Shen et al. [31] study. The operating conditions of this study for the FF and PCF MEE configurations are shown in Table 2 and the comparison results are shown in Figures 6 and 7. In the reference, brine flashing as well as inter-effect flash boxes are used to enhance the overall performance. Moreover, these flow arrangements are tested over different number of effects and steam temperature values. Figure 6 compares the
variation of four performance indices namely the performance ratio (PR), concentration ratio (CR), specific heat transfer area ($sA_t$) and specific condenser capacity ($sQ_c$) with increasing number of effects (evaporators, $n$). As seen in Figure 6, the model output results provide acceptable outcome for both feed configurations FF and PF as compared to the results of Shen et al. [31], especially for the PR and CR parameters (Figures 6a and 6d). However, minor offset is manifested in the $sA_t$ and $sQ_c$ (Figures 6b and 6c). This discrepancy could be explained as follows. In their study, Shen et al. [31] used constant and equal heat transfer area for all effects while our model utilized the heat transfer equations to calculate the necessary heat transfer area based on the pre-specified temperature drop across all effects. The down condenser load is directly proportional to the vapor amount generated in the last effect which decreases as number of effects grows. The $sQ_c$ profiles of both FF and PF flow arrangements obtained in Shen et al. [31] decreased at slower rate than those obtained by our model because different correlation was used to estimate the heat transfer coefficient. Moreover, the reference study utilized a fixed heat transfer area for all evaporators. As a result, our module produced slightly lower $sQ_c$ and $sA_t$ values at any number of effects as compared to those in Shen et al. [31].

We next investigated the effect of increasing steam temperature ($T_s$) on the PR, $sA_t$, $sQ_c$ and CR indices and the results are shown in Figure 7. Similarly, the model’s outcome is still satisfactory when compared to the benchmark data from Shen et al. [31] study with the minor deviation in the $sQ_c$ values (Figure 7c) which could be attributed to the same reasons as discussed in previous paragraph. Although the $sA_t$ profiles exceptionally match the benchmarked data at high steam temperatures, a sudden $sA_t$ growth is observed at the lowest steam temperature of 60 °C. The enlarged heat transfer area at low steam temperature is also reported by El-Dessouky et al. [33].

5.2. Performance of the Hybrid Configurations

Having validated the MEE model, we further examined the performance of different configurations including those of the hybrid system. The results of these simulations are shown in the grey highlighted rows of Table 3. We considered the operating conditions of Shen et al. [31] as given in Table 2. The first eight design parameters shown in Table 3 were fixed to achieve the fair comparison of different system configurations. Additionally, the flash boxes for distillate evaporation in all tested configurations were not taken into account. By comparing three flow
arrangements (FF MEE, PF MEE and PCF MEE) with respect to PR as the key performance index, it is clearly seen that PCF MEE which had the highest PR value of 5.01 over-performed FF-MEE and PCF-MEE designs (PRs of 2.94 and 4.67, respectively). Further improvement in the PCF performance is observed when preheaters are incorporated to heat the feed streams of each effect. In this case, the performance ratio of PCF-pre MEE is increased to 5.31 which is 6% higher than that obtained in case of the PCF MEE configuration.

In the next stage of our simulations, the effectiveness of the hybrid system is tested for the three configurations shown in Figures 1 (FF MEE, PF MEE and PCF MEE, respectively). In hybrid systems, the MEE feed temperature \( T_f \) varies because it is controlled by the MD process. Thus, the PF-Hyb which has the highest MEE feed temperature \( T_f = 45.2 \degree C \) delivered the highest PR of 6.33. Note that the performance of these hybrid configurations depends heavily on the feed temperature \( T_{hf} \) of the supporting MD units. The high \( T_{hf} \) would not only improve the RR of the associated MD system but also increases the feed temperature \( T_f \) in the MEE process. As known, increasing the latter has a solid and positive impact on the MEE performance [28]. \( T_{hf} \) is directly and partially influenced by the temperature of the brine stream \( T_{bm} \) which is used as a portion of the MD feed, and by the amount of the recycled seawater \( R_c \) which forms the second portion of the MD feed stream, i.e. makeup stream. For this reason, the PF-Hyb possess the highest \( T_{hf} \) of 51.8 \degree C as given in Table 3. Although FPCF-Hyb has the highest brine temperature \( T_{bm} \) which is basically the top brine temperature \( T_{bT} \), it has the smallest flow rate because it equals the brine stream exiting the first stage. Therefore, the temperature of the combined stream is quenched by the large recycle which accounts for 92% of the total MD feed. On the other hand, the PCF-Hyb has almost similar brine mass rate as compared to that of the PF-Hyb, as both have almost equal RR (0.44 and 0.41, respectively). However, the brine temperature \( T_{bm} \) of the PCF-Hyb is much lower than that of other processes as shown in Table 3 leading to less temperature value \( T_{hf} \) for the combined streams. Nevertheless, the hybrid systems outperform the standard configurations in terms of PR (6.33, 5.84 and 5.65 for PF-Hyb, PCF-Hyb and FPCF-hyb systems, respectively). According to Table 3, PF-Hyb outperforms all other hybrid systems in terms of PR, RR and sA_t, hence we will focus our further analysis on this particular system. For example, the use of the hybrid PF-Hyb structure could lead to 19% and 26% PR enhancements as compared to the PCF-pre-MEE and conventional standalone PCF-MEE configurations. This high PR value of 6.33 can
be achieved by standard MEE when 10 effects are used, and flash boxes are incorporated, as shown previously in Figure 6.

As mentioned earlier, $T_{hf}$ and consequently $T_f$ play an important role in the overall performance of the hybrid system. Hence, increasing the feed temperatures by adjusting the minimum brine temperature can enhance the overall system’s performance. Figures 8 and 9 demonstrate the effect of the bottom brine temperature ($T_{bn}$) on various temperature profiles and key performance parameters of the hybrid PF-Hyb system, respectively. As shown in Figure 8, the MD and MEE feed temperatures ($T_{hf}$ and $T_f$, respectively) linearly increase with the $T_{bn}$ increase from 35 °C to 55 °C due to the same reasons as discussed in the previous paragraph. Moreover, the seawater temperature exiting the down condenser ($T_{cwo}$), the recycle water temperature ($T_{cwm}$) and the average temperature of the combined distillate of all effects except the last one ($T_D$) also increase when $T_{bn}$ was raised to 55 °C. Intuitively, all these temperatures will increase with the $T_{bn}$ increase. Usually, $T_{cwo}$ is moderately low compared to $T_D$. Hence the combined distillate stream is used to heat up the recycle seawater resulting in a higher temperature value as shown by the increasing trend of $T_{cwm}$. It should be noted though, the enhanced performance of the hybrid system is also attained at smaller total specific heat transfer area ($sA_t$) which has a good reflection on the capital cost.

The increase in $T_{bn}$ from 35 °C to 55 °C enhanced the distillate production rate (D) for both MD and MEE processes and total PF-Hyb system (Figure 9a). Therefore, the PR and RR ratios achieved in the integrated PF-MEE-MD system would also increase and can reach 18% by both at a maximum tested $T_{bn}$ of 55 °C (Figure 9b, c). Besides, the achieved PR value is higher than what can be achieved by 12 stand-alone multi-effect evaporators as shown in Figure 6a. Moreover, as seen in Figures 9b, the PR increased to a more extent with the $T_{bn}$ increase in the case of hybrid PF-MEE-MD as compared to that achieved with a single MEE. It is also important to note that the increase in PR and RR values was obtained by modelling 1957 MD units assuming each MD unit could handle a feed flow rate of up to 600 L/h (based on the pilot plant MD system installed at the KSU). Figure 9f depicts that the hybridization slightly reduces the amount of the MEE required cooling water. In fact, the cooling water flow rate per unit of produced distillate ($sM_{cw}$) swiftly decreases from 16.7 and 15.7 and reached 2.5 and 2.0 at a $T_{bn}$ of 55 °C for the hybrid PF-MEE-
MD and MEE systems, respectively. On other side, the total area per unit of distillate for the MEE process \( (A_t) \) increases sharply for high \( T_{bn} \) values (between 50 °C and 55 °C) as shown in Figure 9d. Note the total surface area of MEE is dominated by the evaporators’ area as the condenser surface area is relatively minor and marginally increases with \( T_{bn} \). However, the enhanced PR and RR are obtained at the expenses of large number of MD units and consequently high specific surface area as shown in Figure 9d. The required specific surface area of the MD systems is the combination of the total surface area of MD1 and MD2 systems (see Figures 2-4). The total surface area of the MD system is the product of the total number of MD units by the unit size. Therefore, according to eq. (54), the total required surface area for the MD1 system is proportional to the MD1 system feed rate \( (F_t) \), hence it does not vary with the \( T_{bn} \) because \( F_t \) itself is invariant with \( T_{bn} \) as well. On the other hand, the total number of MD2 system units and, hence, the associated total surface area decreases with \( T_{bn} \) increase because the MD2 feed \( (M_{cw}) \) decreases with \( T_{bn} \) increase. It should be noted that MD2 system is activated at \( T_{bn} = 40 \) °C and beyond because it is utilized only when its feed temperature \( (T_{cwo}) \) is above 35 °C. This situation created the sudden jump in the combined specific area of MD systems as shown in Figure 9d. A salt reject ratio (CR) is also shown in Figure 9e. The CR for the standalone MEE system decreases slightly with the \( T_{bn} \) increase. Note CR of the PF configuration is simply the ratio of the feed to the brine flow rates of the last effect. This ratio decreases with \( T_{bn} \) due to the reduction in the evaporation rates. On the other hand, the overall CR increases with \( T_{bn} \) increase. As \( T_{bn} \) increases, the performance of MD1 improves (Figure 9a) thereby producing more salty brine. Hence, the ratio of the salinity of the brine discharged from MD1 to the feed salinity increases.

Another important parameter affecting the performance of the distillation processes is the cooling water temperature \( (T_{cw}) \) which in real conditions is a function of the plant location. Figure 10 highlights the impact of \( T_{cw} \) on key performance parameters of the PF-MEE-Hyb system. As shown by Figures 2, 3 and 4, hybridization isolates the effect of cooling water from the MEE system. As a result, the MEE distillate rate and consonantly its corresponding PR and RR, the salt rejection and consequently the required surface area are invariant with changes in the cooling water temperature (Fig10a-e). Interestingly, the performance of the hybrid system improves in terms of RR and PR because of the contribution of MD2. As cooling water temperature rises, the driving force decreases at the down condenser which incurs additional cooling water flow rate (Figure 10f).
to absorb the latent heat of the last effect distillate. The augmented cooling water rate and increased condenser-outlet temperature improve the distillate production rate of MD2 which is reflected on the overall system. By increasing the cooling water temperature from 15 °C to 35 °C, the RR and PR of the hydride system increase by 10% each. This achievement is of course at the expense of increase of the surface area of the MD units. In fact, for the given $T_{cw}$ range, the specific surface area of MD should be doubled.

Figure 11 depicts the effect of the feed salinity on the performance of the PF-MEE-MD hybrid. Such an effect is expected to be limited due to the weak relations between the feed salinity and the system design and operating parameters. This effect can be attributed to the variation in the boiling point elevation and heat capacity with composition. In addition, the MD systems are unaffected by the feed composition which is well known for such separation technologies [11]. This weak impact of the feed salinity is observed in the various profiles presented in Figure 11 of the distillate flow rates ($D_{MD}$, $D_{MEE}$, $D_{tot}$) and the associated RR, PR and CR. The variation of the specific area (sA) with the feed salinity shows however a different trend. The minor variation of the specific area for the MEE system is attributed to the variation of the physical and thermodynamics properties with the feed water salinity inside the effects. The slight variation in the specific area of the MD system is due to the minor escalation of the cooling water rate which demanded additional MD units in the MD2 system. The little changes in the cooling water rate is the result of the weak growth in the MEE distillate rate.

On the other side, the impact of the feed flow rate ($F_t$) on the performance parameters of the hybrid system appears to be pronounced, as shown in Figure 12. Increasing the $F_t$ adversely influences the MEE distillate production because it quenches the mixture leading to less evaporation (Figure 12a). On the other hand, the pure water production in the MD system, which is dominated by the MD1, increases because of the augmented throughput due to using parallel MD structure to maximize the throughput capacity. Nevertheless, the net distillate production of the hybrid process marginally decreases. Accordingly, the corresponding overall RR, PR and CR diminishes with $F_t$ (Figures 12b, c and e). However, as it is also seen, the RR, PR and CR reduce swiftly because the distillate production decreases while the feed flow rate increase simultaneously making their ratio to degrade progressively. The degraded evaporation especially in the last effect has its impact on
the down condenser performance by demanding less amount of cooling water due to less condensation energy requirement (Figure 12f). The required specific surface area (sA) for the MD system reduces with feed flow rate as shown in Figure 12d. Note the number of required MD1 units rises because the MD1 feed (F_t) increases while the number of required MD2 units decreases as the MD2 feed (M_cw) declines. Hence, the total necessary number of MD units grow slightly. However, the combined distillate production of MD1 and DM2 systems grow rapidly causing sA depletion. On the other hand, as shown in Figure 12d, the desired sA for the MEE system follows a concave trend with the F_t increase. At low F_t, both D_tot (Figure 12a) and sA (Figure 12d) decline making their corresponding ratio to drop, but as F_t increases, D_tot decreases at faster rate causing the specific surface area to increase again.

6. Conclusions

This work analyses theoretically the performance of various configurations integrating MEE and MD processes. Instead of rejecting the concentrated brine exiting from the MEE plant to the environment, it is fed to an array of MD units. This structure has dual objectives. First, the reject brine is used to produce additional distillate in the MD array and reduces brine discharge volumes. Secondly, the MD system will increase the temperature of the MEE feed. Mathematical models for three MEE configurations and a DCMD process are derived based on heat and mass balance for each system component. Corresponding numerical models have been also developed and validated. The simulation results which are presented in terms of various performance parameters of the desalination systems including RR, PR and sA, are discussed and analyzed. The conducted modelling revealed significant benefits of the hybrid MEE-MD processes over the standalone MEE process. The main concluding remarks of the present work can be summarized as follows:

- The integration of the MEE and MD systems can be achieved under numerous configurations. We proposed structures incorporating the MD systems in simple and flexible way to treat the MEE brines and preheat MEE feed water.

- Based on various key performance indicators, the investigated hybrid MEE and MD structures outperform the standalone MEE systems. Thus, PR can increase by more than 20% when MEE is coupled with MD. This corresponds to a single MEE structure with an additional 5 effects.
- Among the standalone MEE configurations, the PCF process outperforms the PF and FF processes. In case of hybrid MEE-MD structures, PF-MEE process is characterized by the highest PR.

- The performance of the hybrid configurations depends on several design and operating parameters including the feed temperatures of the MEE units and supporting MD system that of the \( T_f \) and \( T_{fh} \), respectively. Increasing \( T_f \) and \( T_{hf} \) has a positive impact on the PR and RR values.

- The effect of cooling water temperature \( (T_{cw}) \), feed salinity \( (X_f) \) and MEE feed flow rate \( (F_t) \) was also analyzed. The impact of the feed salinity in particular appears to be weak. High \( F_t \) should also be avoided because it degrades the RR and PR. However, high cooling water temperature is found beneficial as it improves the MD performance, and consequently the overall system’s performance.

The present work will be extended to include exergy and economic studies of the hybrid MEE-MD as well as the incorporation of thermal vapor compression (TVC).

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Nomenclature

$A_s$ MD membrane Surface area, m$^2$
$A_c$ Condenser heat transfer area, m$^2$
$A_{ev}$ Total evaporator surface area, m$^2$
$A_t$ Total surface area for MEE process, m$^2$
$A_{MD}$ Total surface area for MD system, m$^2$
$B_i$ Brine mass rate, kg/s
$B_i$ Brine mass rate of $i^{th}$ effect, kg/s
$B_1$ Brine mass rate of the first effect, kg/s
$B_n$ Brine mass rate of the last effect, kg/s
$BPR$ Boiling point rise, °C
$C_m$ Membrane Permeability coefficient, kg/m$^2$sPa
$C_k$ Knudsen mass flux coefficient, kg/m$^2$sPa
$C_d$ Moléculaire diffusion mass flux coefficient, kg/m$^2$sPa
$C_T$ Transition mass flux coefficient, kg/m$^2$sPa
$C_p$ Heat capacity, J/kg°C
$CR$ Salt concentration ratio
$D$ Distillate mass rate, kg/s
$D_{tot}$ Total MEE distillate mass rate, kg/s
$D_{MD}$ Total distillate rate for the MD system, kg/s
$d_{bi}$ Flashing brine rate, kg/s
$d_{vi}$ Flashing distillate rate, kg/s
$d_h$ Hydraulic diameter of MD membrane, m
$F$ Feed mass rate, kg/s
$F_t$ Total feed flow rate, kg/s
$H_v$ Latent heat of vaporization of water, J/kg
$h_f$ Hot Feed heat transfer coefficient, W/m$^2$K
$h_p$ Cold Feed heat transfer coefficient, W/m$^2$K
$h_m$ Membrane heat transfer coefficient, W/m$^2$K
$J$ Mass flux, kg/m$^2$h
$k_m$ Membrane conductivity, W/mK
$k$ Water thermal conductivity, W/mK
$LMTD$ Logarithmic mean temperature difference
$l_m$ Mesh size, m
$M_w$ Molecular weight of water, g/mole
$M_{cw}$ Cooling water mass flow rate, kg/s
$m_{hi}$ Hot water inlet, m$^3$/h
$m_{ho}$ Hot water outlet flow rate, m$^3$/h
$m_c$ Permeate flow rate, m$^3$/h
$m_w$ Distillate flow rate, m$^3$/h
$M_s$ Steam mass rate, kg/s
$NEA$ Non-equilibrium allowance, °C
$Nu$ Nusselt number
$n$ Number of effects
$N$ Number of MD units
$P_1$ Vapor pressure at feed membrane surface, Pa
$P_2$ Vapor pressure at permeate membrane surface, Pa
$PD$ Membrane pressure multiplied by diffusivity, Pam$^2$/s
$P_a$ Entrapped air pressure, Pa
Performance ratio
Prandtl number
Heat transfer rate at MD feed sections, W/m²
Heat transfer rate at MD permeate sections, W/m²
Heat of evaporation and conduction, W/m²
Overall heat flux, W/m²
Maximum allowable MD volumetric feed rate, m³/s
MD membrane pore size, m
Ideal gas constant, J/moleK
Makeup water flow rate for the MD system, kg/s
Reynold number
Recovery ratio of MEE
Recovery ratio of hybrid system
Specific heat transfer area, m²/(kg/s)
Specific condenser energy consumption, kJ/kg
Specific cooling water mass rate, kg/kg
Water salinity in kg/m³
Brine temperature, °C
Minimum brine temperature, °C
Top brine temperature, °C
Brine temperature to be fed to MD, °C
Distillate temperature for MEE, °C
Permeate (cold) bulk temperature for MD, °C
Cooling water temperature entering condenser, °C
Cooling water temperature exiting condenser, °C
Average MEE distillate temperature, °C
Recycle (makeup) stream temperature after heating, °C
Average temperature of the all combined MEE distillate except the last one, °C
MEE feed temperature, °C
MD hot feed temperature, °C
Feed (hot) bulk temperature, °C
Membrane interface temperature, feed side, °C
Membrane interface temperature, permeate side, °C
Reference temperature, °C
Steam temperature, °C
Saturated vapor temperature, °C
Temperature drop due to thermodynamic losses, °C
Temperature drop due to pressure losses and friction, °C
Temperature polarization coefficient
Overall heat transfer coefficient, W/m²K
Condenser heat transfer coefficient, W/m²K
Salt concentration, ppm
Salt concentration (salinity) of the feed, ppm
Salt concentration (salinity) in the last effect (ppm)
MD membrane porosity
Termination criteria
MD membrane tortuosity
MD membrane thickness, mm
Density, kg/m³
Latent heat, kJ/kg
References


Table 1: Model validation data.

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<th>Parameter</th>
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<th>PCF Model [29]</th>
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<td>n</td>
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**Input data:**

- \(T_s\) [°C]: 98, 98, 70.6, 70.6
- \(T_{bn}\) [°C]: 39, 39, 36, 36
- \(T_{cw}\) [°C]: 29, 29, 28, 28
- \(T_f\) [°C]: 34, 34, 32.3, 32.3
- \(M_s\) [kg/s]: 1.86, 1.86, 15.8, 15.8
- \(F_i\) [kg/s]: 12.5, 12.5, 157.8, 157.8
- \(X_f\) [ppm]: 42,000, 42,000, 46,000, 46,000

**Output data:**

- \(X_n\) [ppm]: 73,738, 70,000, 67,253, 69,000
- PR: 2.89, 2.69, 3.16, 3.34
- \(M_d\) [kg/s]: 5.38, 5, 49.9, 52.6
- \(T_bT\) [°C]: 90.15, 90.2, 63.7, 64
- \(A_t\) [m²]: 463.0, 507.0, NA, NA

Table 2: Operating conditions according to Shen et al. [31]

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<td>RR</td>
<td>0.33</td>
</tr>
<tr>
<td>$T_{bn}$ [°C]</td>
<td>NA</td>
</tr>
<tr>
<td>$T_{bf}$ [°C]</td>
<td>NA</td>
</tr>
<tr>
<td>Re [%]</td>
<td>NA</td>
</tr>
<tr>
<td>RRO</td>
<td>NA</td>
</tr>
</tbody>
</table>

Figure 1: Schematics of standard MEE configurations

Table 3: Performance indicators of standard and hybrid configurations.
Figure 2: PCF hybrid (PCF-hyb) system.

Figure 3: PF hybrid (PF-hyb) system.
Figure 4: First PCF hybrid (FPCF-hyb) system.

Figure 5: Solution algorithm for the hybrid system.
Figure 6: Model validation over different number of effects (n) with respect to PR (a), sA_t (b), sQ_c (c) and CR (d) using Shen et al. [31] operating conditions at T_s = 70 °C.

Figure 7: Model validation over different values for steam temperature (T_s) using Shen et al. [31] data at n = 10.
Figure 8: Effect of minimum brine temperature ($T_{bn}$) on the temperature profiles for the PF-Hyb configuration at $n = 7$, $T_s = 70 \, ^\circ C$ and operating conditions at $T_{cw} = 25 \, ^\circ C$, $X_f = 32,000 \, \text{ppm}$, $F_t = 326 \, \text{kg/s}$.

Figure 9: Effect of minimum brine temperature ($T_{bn}$) on the PF-Hyb configuration performance at $n = 7$ and $T_s = 70 \, ^\circ C$. Operating conditions: $T_{cw} = 25 \, ^\circ C$, $X_f = 32,000 \, \text{ppm}$, $F_t = 326 \, \text{kg/s}$.
Figure 10: Effect of cooling water temperature ($T_{cw}$) on the PF-Hyb configuration performance at $n = 7$ and $T_s = 70 \, ^\circ C$. Operating conditions: $T_{bn} = 50 \, ^\circ C$, $X_f = 32,000 \, ppm$, $F_i = 326 \, kg/s$.

Figure 11: Effect of feed salinity ($X_f$) on the PF-Hyb configuration performance at $n = 7$ and $T_s = 70 \, ^\circ C$. Operating parameters: $T_{bn} = 50 \, ^\circ C$, $T_{cw} = 25 \, ^\circ C$, $X_f = 326 \, kg/s$. 

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Appendix A: DCMD desalination unit model

The mass flux \( J \) of the vapor transfer through the membrane pores is given by the following equation:

\[
J = C_m (P_1 - P_2) \frac{k g}{m^2 s}
\]  
(A.1)

where: \( P_1 \) and \( P_2 \) are the vapor pressures of water vapor estimated at the membrane surface temperatures \( T_{hm} \) and \( T_{cm} \), respectively.

The vapor pressures in Pa is estimated using the Antoine equation (Fard et al. [36] and Khayet [37]):

\[
P_1 = \exp \left( 23.238 - \frac{3841}{T_{hm} - 45} \right) (1 - C_s)(1 - 0.5C_s - 10C_s^2)
\]  
(A.2)

\[
P_2 = \exp \left( 23.238 - \frac{3841}{T_{cm} - 45} \right)
\]  
(A.3)
where: \( C_s \) is the water salinity (%); \( C_m \) is the MD membrane permeability coefficient calculated from three correlations depending on the type of mass transfer regime (Knudsen, Molecular diffusion or Knudsen-Molecular diffusion transition):

**Knudsen flow mechanism:**

\[
C_m^k = \frac{2\varepsilon r}{3\tau_\delta \left(\pi R T\right)} \left(\frac{8M_w}{\pi R T}\right)^{1/2}
\]  
(A.4)

**Molecular diffusion mechanism:**

\[
C_m^D = \frac{\varepsilon PDM_w}{\tau_\delta P_a R T}
\]  
(A.5)

**Knudsen-Molecular diffusion transition mechanism:**

\[
C_m^C = \left[\frac{3\tau_\delta \left(\pi R T\right)}{2\varepsilon r \left(8M_w\right)}\right]^{1/2} + \frac{\tau_\delta P_a R T}{PDM_w}
\]  
(A.6)

These different regimes depend on the wall collision theory of water molecules, and each regime dominates at a specific range of values for the mean free path of a water molecule. The heat transfer process occurs in three steps:

i. Convection from the feed bulk to the vapor–liquid interface at the membrane surface:

\[
q_f = h_f(T_h - T_{hm})
\]  
(A.7)

ii. Convection from the vapor–liquid interface at the membrane surface to the permeate side:

\[
q_p = h_p(T_{cm} - T_c)
\]  
(A.8)
where: \( h_f \) and \( h_p \) denote the heat transfer coefficients on the feed and cold stream sides, respectively.

The heat transfer coefficients can be estimated from the following Nusselt correlations (Da costa et al. [38], Phattaranawik et al. [39]):

\[
\text{Nu}_i = \frac{h_i d_h}{k_i} = 0.664 Re_i^{0.5} Pr_i^{0.33} \left( \frac{d_h}{l_m} \right)^{0.5} \tag{A.9}
\]

Where \( d_h \) is the hydraulic diameter for the spacer-filled channel, \( k \) is the thermal conductivity of water flow, \( Re \) and \( Pr \) are Reynold’s and Prandtl number of feed flow respectively. \( l_m \) is the mesh size and the index \( i \) denotes either permeate side or hot side.

iii. Evaporation and conduction through the microporous membrane:

\[
q_m = J H_v + h_m (T_{hm} - T_{cm}) \tag{A.10}
\]

where: \( H_v \) is the latent heat of water vaporization which can be estimated using eq. (A.11) [36]:

\[
H_v(T) = 1850.7 + 2.8273 T - 1.6 \times 10^{-3} T^2 \tag{A.11}
\]

and \( h_m \) is the conductive heat transfer coefficient defined as follows:

\[
h_m = \frac{k_m}{\delta} \tag{A.12}
\]

where: \( k_m \) and \( \delta \) denote the membrane thermal conductivity and its thickness, respectively.

The total heat flux across the membrane is directly proportional to the bulk temperature gradient and can be expressed as follows:

\[
Q = U(T_h - T_c) \tag{A.13}
\]
For the countercurrent flow, the bulk temperatures are taken as $T_h = T_{h\text{in}}$ and $T_c = T_{c\text{out}}$.

The overall heat transfer coefficient is given by:

$$U = \left[ \frac{1}{h_f} + \frac{1}{h_m + \frac{f h_m}{f c_m - f c_m}} + \frac{1}{h_p} \right]^{-1}$$  \hspace{1cm} (A.14)

Under the steady-state operation, the heat transfer in the three individual parts of the system reaches equilibrium as follows:

$$q_f = q_m = q_p$$  \hspace{1cm} (A.15)

Considering the macroscopic scale of the DCMD unit (Fig. (A.1)), the heat balance around the permeate side is given by eq. (A.16) [40]:

$$UA_s(T_h - T_c) = m_c \rho C_p(T_{c\text{out}} - T_{c\text{in}})$$  \hspace{1cm} (A.16)

where: $m_c$ and $C_p$ denote the volume flow rate and specific heat at constant pressure, respectively. Eq. (A.16) is used to compute the permeate exit temperature, $T_{c\text{out}}$.

Similarly, assuming a constant density and heat capacity, the mass and heat balance around the feed side are given by:

$$UA_s(T_h - T_c) = m_{h\text{in}} \rho C_p(T_{h\text{out}} - T_{\text{ref}}) - m_{h\text{out}} \rho C_p(T_{h\text{in}} - T_{\text{ref}})$$  \hspace{1cm} (A.17)

$$m_{h\text{in}} - m_{h\text{out}} = m_w$$  \hspace{1cm} (A.18)

Eq. (A.17) is used for computing $T_{h\text{out}}$. It should be noted that eqs. (A.16) and (A.17) are based on the ideal case when the heat losses are negligible.

The definitions of various variables, the numerical values of the physical and design parameters in the eqs. (A.1-A.18), and additional supporting correlations are provided by Safavi and Mahmmadi [41] and Chen and Ho [42].
Figure A.1: Schematic of the typical DCMD unit.

Appendix B: Organigrams

Figure B.1: Solution algorithm for the MEE system.
Figure B.2: Solution algorithm for the MD system.