Assessing the performance of solar thermal driven membrane distillation for seawater desalination by computer simulation

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Disciplines
Engineering | Science and Technology Studies

Publication Details

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This journal article is available at Research Online: https://ro.uow.edu.au/eispapers1/581
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Revised Manuscript Submitted to

Journal of Membrane Science

July 2017

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Abstract: A computer model was developed to simulate the performance of an integrated solar thermal driven direct contact membrane distillation (DCMD) system for seawater desalination using recorded weather data. The results highlight the importance of simulating the DCMD process together with the energy source. Indeed, when considered in isolation from the thermal energy source, increasing water cross flow velocities in the feed and distillate channels results in an increase in water flux and thermal efficiency of the DCMD module. By contrast, when coupling the DCMD module with the solar thermal collector, increasing water cross flow velocities reduces both the process water flux and thermal efficiency. This is because of the limited supply of solar thermal at any given time, and hence the feed temperature decreases when cross flow velocities increase. Thus, any benefits in the reduction of temperature polarisation due to increasing cross flow velocities are overwhelmed by the effects of feed temperature decrease on water flux and thermal efficiency. Results from our simulation also demonstrate the viability of the solar thermal driven DCMD process for small-scale seawater desalination applications. Distillate production is dependent on the availability of solar radiation during the day; nevertheless, a small system with a 7.2 m² spiral-wound DCMD module and a 22.6 m² flat plate solar thermal collector can produce over 140 kg of distillate each day under real weather conditions. This is equivalent to a daily distillate production rate of 19.7 kg per m² of membrane or 6.3 kg per m² of solar thermal collector.

Keywords: membrane distillation (MD); seawater desalination; solar thermal energy; simulation; process optimisation.
1. Introduction

Membrane distillation (MD) has significant potential for small-scale solar thermal seawater desalination in remote coastal areas. In the MD process, a microporous hydrophobic membrane is used to facilitate the transport of water vapour while retaining liquid water and hence all non-volatile substances and dissolved salts; therefore, ultrapure water can be obtained from seawater MD desalination [1, 2]. Unlike pressure-driven membrane desalination processes such as reverse osmosis (RO), MD utilises a vapour pressure gradient induced by a temperature difference across the membrane as the driving force for water transfer. Thus, water flux in MD is not affected by the feed water osmotic pressure [3, 4].

MD is arguably the most suitable platform for small-scale and off-the-grid seawater desalination applications [5-9]. MD is less susceptible to membrane fouling than RO given the absence of a high hydraulic pressure and the discontinuity of the liquid phase across the membrane. As a result, MD can be operated without feed water pre-treatment, making it an ideal process for small and stand-alone seawater desalination applications. Furthermore, MD systems can be made from inexpensive and noncorrosive plastic materials to reduce process investment and operational costs. Finally, MD can be operated at low feed temperature [10-12]; thus, low-grade heat sources such as solar thermal energy can be utilised to meet MD energy demands.

MD can be applied in various configurations including direct contact membrane distillation (DCMD), air gap membrane distillation (AGMD), permeate gap membrane distillation (PGMD), vacuum membrane distillation (VMD), and sweeping gas membrane distillation (SGMD). Amongst these configurations, DCMD has the simplest arrangement and thus is probably best suited for small-scale seawater desalination applications [13, 14]. In DCMD, the hot feed and the cold distillate are in direct contact with the membrane, allowing for efficient heat and mass transfer to and from the feed and distillate membrane surfaces, and thus facilitating high water flux [14, 15]. However, the direct contact configuration also facilitates the heat conduction through the membrane, thus rendering DCMD less thermally efficient than other configurations.

Numerous experimental studies have been conducted to investigate the feasibility of DCMD using low-grade heat sources such as waste heat and solar thermal energy for desalination applications. Dow et al. [14] reported a sustainable operation of a pilot DCMD system over 3 months using waste heat from a thermal power station. They obtained an average water flux of 3 kg/(m²·h) and thermal efficiency of 0.42 [14]. The low thermal efficiency obtained by Dow et al. [14] was deemed acceptable given the freely available waste heat from the power plant. Suarez et al. [16] examined a lab-scale desalination process using DCMD with low-temperature heat
extracted from a salt-gradient solar pond. Their results revealed a limited thermal efficiency, with only about 35% of the heat extracted from the solar pond was utilised for the water transfer across the membrane [16]. In another study, Shim et al. [17] evaluated a DCMD system coupled with a solar thermal collector for seawater desalination application. The solar thermal collector could provide more than 77% heating energy of the DCMD process. However, because of the low process thermal efficiency, a large area of solar thermal collector (i.e. 4.7 m²) was required to power 0.06 m² DCMD membrane module [17]. Solar thermal collectors have also been utilised to power AGMD and PGMD systems for small-scale seawater desalination application [18-20].

Computer-aided simulations can play an important role in optimising module design and process operation for a solar thermal driven MD system. Simulations of complex processes during MD operation including heat and mass transfer, temperature and concentration polarisation effects, and thermal efficiency have been reported in the literature [21-25]. Computer software packages such as MATLAB, ANSYS Fluent, Aspen Custom Modeller, COMSOL Multiphysics, and Microsoft Excel have been successfully employed to perform MD simulations [21, 23, 24, 26-28]. However, in contrast to the many previous simulation studies on MD membrane module and system performance, very few works have attempted to simulate the integrated solar thermal driven MD process [29]. Unlike previous studies with a constant supply of energy, solar thermal is only available during the day and there is a significant temporal variation in the amount of thermal energy generated per square meter of collector [29, 30].

This study reported a comprehensive simulation of an integrated solar thermal driven DCMD system for small-scale seawater desalination application. First, important performance parameters including water flux, temperature polarisation effect, and thermal efficiency along the DCMD module channels under two flow modes and various operating conditions were simulated and examined. Subsequently, a solar thermal driven DCMD process, in which a solar thermal collector was coupled with the DCMD module, was simulated to demonstrate the intermittent performance of the process. The influences of operating conditions, most importantly the water cross flow velocities, on process water flux and thermal efficiency were elucidated to optimise the solar thermal seawater DCMD process. Thermal efficiency and distillate production rate of the process were analysed to elucidate the feasibility of DCMD for solar thermal driven seawater desalination. The TRNSYS, which is a transient system simulation tool [31], was used to simulate the availability of solar thermal for the MD process.
2. Theories

In MD, the transfer of mass and heat through the membrane occurs simultaneously. Water (mass) transfer is driven by the difference in partial water vapour pressure across the membrane. At the same time, heat is transferred via conduction and in the form of latent heat.

The mass transfer of water is described as:

\[ J = C_m(P_{m.f} - P_{m.d}) \]  

(1)

where \( J \) is water flux (kg/(m\(^2\)h)), \( C_m \) is the membrane mass transfer coefficient (kg/(m\(^2\)h-Pa)), and \( P_{m.f} \) and \( P_{m.d} \) are the water vapour pressure (Pa) at the feed and distillate membrane surfaces, respectively. The water vapour pressure at the membrane surfaces can be calculated as [2, 13]:

\[ P = x_{\text{water}}(1 - 0.5x_{\text{salt}} - 10x_{\text{salt}}^2)P_0 \]  

(2)

where \( x_{\text{water}} \) and \( x_{\text{salt}} \) are the molar fraction of water and salt in the feed and distillate solution, and \( P_0 \) is the vapour pressure of pure water, which can be calculated using the Antoine equation [2, 13]:

\[ P_0 = \exp(23.1964 - \frac{3816.44}{T - 46.13}) \]  

(3)

where \( T \) is the temperature (K) of the feed and distillate at the membrane surface.

The membrane mass transfer coefficient (\( C_m \)) is dependent on membrane properties and process operating conditions. For seawater desalination by DCMD, \( C_m \) can be described as [13, 17, 32, 33]:

\[ C_m = \left[ \frac{3\tau\delta}{2\varepsilon r} \left( \frac{\pi RT}{8M} \right)^{1/2} + \frac{\tau\delta P_a RT}{\varepsilon PD M} \right]^{-1} \]  

(4)

where \( \delta, \varepsilon, \tau, \) and \( r \) are the membrane thickness (m), porosity (dimensionless), pore tortuosity (dimensionless), and pore radius (m), respectively, \( M \) is the molecular weight of water (kg/mol), \( R \) is the gas constant (i.e. 8.314 J/(mol-K)), \( T \) is the mean water vapour temperature (K) inside the membrane pore, \( P \) and \( P_a \) are the total pressure and the air partial pressure (Pa) inside the membrane pore, and \( D \) is the water diffusion coefficient (m\(^2\)/s). The multiplication of \( P \) and \( D \) is a function of temperature, and can be calculated as:

\[ PD = 1.895 \times 10^{-5}T^{2.072} \]  

(5)
The heat flux ($Q$) through the membrane is a combination of conduction heat and latent heat, and can be described as:

$$Q = \frac{k_m}{\delta} \left( T_{m,f} - T_{m,d} \right) + J \Delta H_v$$  \hspace{1cm} (6)

where $Q$ is in kJ/(m$^2$·h), $k_m$ is the membrane thermal conductivity (W/(m·K)), $\Delta H_v$ is the latent heat of evaporation (kJ/kg), and $T_{m,f}$ and $T_{m,d}$ are the temperatures of the feed and distillate streams at the membrane surfaces, respectively. The membrane thermal conductivity is the volume-average of polymer thermal conductivity ($k_s$) and gas thermal conductivity ($k_g$) as follows [34]:

$$k_m = \left[ \frac{\varepsilon}{k_g} + \frac{1-\varepsilon}{k_s} \right]^{-1}$$  \hspace{1cm} (7)

The procedure to calculate $k_s$ and $k_g$ is available from Alkhudhiri et al. [13]. $\Delta H_v$ is calculated using the mean water vapour temperature inside the membrane pore as:

$$\Delta H_v = 1.7535T + 2024.3$$  \hspace{1cm} (8)

The temperature of the feed and distillate streams at the membrane surfaces (i.e. $T_{m,f}$ and $T_{m,d}$) can be determined based on the bulk stream temperatures (i.e. $T_{b,f}$ and $T_{b,d}$) by iteration as follows [35, 36]:

$$T_{m,f} = \frac{T_{b,f} h_f + h_m \left( T_{b,d} + T_{b,f} \frac{h_f}{h_d} \right) - J \Delta H_v}{h_f \left( 1 + \frac{h_m}{h_d} \right) + h_m}$$  \hspace{1cm} (9)

$$T_{m,d} = \frac{T_{b,d} h_d + h_m \left( T_{b,f} + T_{b,d} \frac{h_d}{h_f} \right) + J \Delta H_v}{h_d \left( 1 + \frac{h_m}{h_f} \right) + h_m}$$  \hspace{1cm} (10)

where $h_f$ and $h_d$ are the heat transfer coefficients in the feed and distillate thermal boundary layers respectively, and $h_m$ is the heat transfer coefficient across the membrane. The calculation of the heat transfer coefficients in the boundary layers ($h_f$ and $h_d$) involves the Nusselt number ($Nu$):

$$h_{f,d} = \frac{Nu_{f,d} k}{d_h}$$  \hspace{1cm} (11)
where \( k \) is the thermal conductivity (W/(m·K)) of the feed and distillate streams, and \( d_h \) is the hydraulic diameter of the feed and distillate channels. For a flat sheet DCMD module at laminar flow, \( Nu \) can be calculated as [36, 37]:

\[
Nu = 0.097 Re^{0.73} Pr_b^{0.13} \left( \frac{Pr_b}{Pr_m} \right)^{0.25}
\]  

(12)

where \( Re \) is the Reynolds number, \( Pr \) is the Prandtl number, and the subscripts \( b \) and \( m \) represent bulk fluid and membrane surface, respectively. The calculations of \( Re \) and \( Pr \) are as follows:

\[
Re = \frac{\rho u d_h}{\mu}
\]  

(13)

\[
Pr = \frac{\mu C_p}{k}
\]  

(14)

where \( \rho \), \( \mu \), and \( C_p \) are density (kg/m\(^3\)), dynamic viscosity (kg/(m·s)), and specific heat capacity (kJ/(kg·K)) of the feed and distillate streams, and \( u \) is the fluid cross flow velocity (m/s).

During the DCMD process, temperature polarisation effect renders the temperature difference between the hot and cold membrane surfaces smaller than that between the bulk feed and distillate streams, thus reducing the actual process driving force. The temperature polarisation coefficient (\( \Phi \)) is calculated as:

\[
\Phi = \frac{T_{m.f} - T_{m.d}}{T_{b.f} - T_{b.d}}
\]  

(15)

In the DCMD process, temperature and concentration polarisation can occur simultaneously. Because the influence of concentration polarisation on water flux is insignificant compared to that of temperature polarisation during the desalination of seawater by DCMD [4, 13, 32], concentration polarisation is excluded in this study.

Thermal efficiency (\( \Pi \)) is an important parameter to evaluate the performance of DCMD. \( \Pi \) is the ratio of the latent heat over the total heat flux through the membrane and can be calculated as:

\[
\Pi = \frac{J \Delta H_v}{J \Delta H_v + \frac{k_m}{\delta} (T_{m.f} - T_{m.d})}
\]  

(16)
3. Computer software package and process simulation

3.1. TRNSYS simulation package

The Transient System Simulation (TRNSYS) software package version 17.2 was used to simulate an integrated solar thermal driven DCMD system. Accredited built-in components including a flat plate solar thermal collector (Type 539), a water pump (Type 3b), an online plotter (Type 65c), and weather data reading and processing (Type 15-3) in the TRNSYS library were used without any modification. Other components including a spiral-wound DCMD module, a feed tank, and controllers were composed by the authors. The solar thermal collector (i.e. composed of 10 flat panels in series) had an effective surface area \( A \) of 22.6 m\(^2\), intercept efficiency \( a_0 \) of 0.687, the 1\(^{\text{st}}\) order and the 2\(^{\text{nd}}\) order efficiency coefficients \( a_1 \) and \( a_2 \) of 1.505 and 0.0111, respectively.

A spiral-wound DCMD module from AquaStill (Sittard, The Netherlands) was chosen for the simulation. The DCMD module had 6 feed and distillate parallel channels, each of which had height, width, and length of 2 mm and 0.4 m, and 1.5 m, respectively [38]. The total membrane surface area was 7.2 m\(^2\) [38]. According to the manufacturer, the water circulation flowrates of this module were from 600 to 1500 L/h (equivalent to 3.5 to 8.7 cm/s in cross flow velocities) [38]. The initial water volume of the feed tank was set at 3 m\(^3\). Hourly recorded weather data from 1\(^{\text{st}}\) to 5\(^{\text{th}}\) January 1991 for a coastal location (-33.87° N; 151.20° E) in New South Wales Australia were arbitrarily selected as solar radiation input for the simulation.

3.2. Process simulation

The flow directions and descriptions of heat and mass transfer in the DCMD membrane module under co-current and counter-current flow are illustrated in Fig. 1. For both flow modes, the mass flux of water \( (J_i) \) through an incremental membrane area \( i \) is calculated using Eqs. (1-5) with \( T_{m,f,i} \) and \( T_{m,d,i} \) determined via the mathematical iterative models (Eqs. 9 & 10). Then, the mass flow of water from the feed to the distillate \( (d_{m,d}) \) can be calculated as:

\[
d_{m,i} = J_i W dx
\]

where \( dm_i \) is in kg/h, \( W \) is the width of the flow channels (m), and \( dx \) is the incremental membrane length (m). The heat flux through the membrane area \( W dx \) is:

\[
dQ_i = \left( J_i \Delta H_e + \frac{k_m}{\delta} (T_{m,f} - T_{m,d}) \right) W dx
\]
Fig. 1. Schematic diagram of two incremental membrane areas along the DCMD module under the (A) co-current and (B) counter-current flow mode.

From mass and heat balance, the mass flow \((m_f)\), bulk temperature \((T_{b,f})\), and salinity \((S_f)\) of the feed stream at the area \(i+1\) can be determined as follows:

\[
m_{f,i+1} = m_{f,i} - dm_i
\]  
(19)

\[
S_{f,i+1} = \frac{m_{f,i}S_{f,i}}{m_{f,i+1}}
\]  
(20)

\[
T_{b,f,i+1} = \frac{m_{f,i}C_P f_iT_{b,f,i} - dQ_i}{m_{f,i+1}C_P f_i}
\]  
(21)

The mass flow \((m_d)\) and the bulk temperature \((T_{b,d})\) of the distillate stream at the area \(i+1\) under co-current flow are calculated as:

\[
m_{d,i+1} = m_{d,i} + dm_i
\]  
(22)

\[
T_{b,d,i+1} = \frac{m_{d,i}C_P d_iT_{b,d,i} + dQ_i}{m_{d,i+1}C_P d_i}
\]  
(23)

For counter-current flow, these parameters are determined as follows:

\[
m_{d,i+1} = m_{d,i} - dm_i
\]  
(24)
\[ T_{b,d,i+1} = \frac{m_{d,i} Cp_{d,i} T_{b,d,i} - dQ_i}{m_{d,i+1} Cp_{d,i}} \] (25)

For simplicity, it is assumed that: (1) the system is at steady state; (2) the feed and the distillate streams only flow in the \( x \) direction; (3) salt rejection is 100\%; and (4) there is no heat loss to the environment.

The calculation algorithms of the DCMD process are provided in Fig. 2 and Fig. 3 for co-current and counter-current, respectively. The calculation starts from the feed inlet end (i.e. \( x_0 = 0 \)) and finishes at the feed outlet end (i.e. \( x_n = L \)) of the DCMD module. For co-current flow, the initial parameters of the feed and distillate streams are readily available. On the other hand, for counter-current flow, initial guesses of the mass flow and temperature of the distillate at the outlet (i.e. \( m_{d,\text{out}} \) and \( T_{d,\text{out}} \)) are required (Fig. 3).
In the solar thermal driven DCMD system under brine-recycled operation, seawater is circulated through the solar thermal collector to the DCMD module and back to the feed tank (Fig. 4). Within the DCMD module, distillate is collected and the feed seawater is concentrated.
Thus, the feed water mass ($m_{tank}$), salinity ($S_{tank}$), and temperature ($T_{tank}$) in the feed tank change with operating time. Their values at two consecutive simulation steps (i.e. $j$ and $j+1$) are calculated as:

$$S_{tank,j+1} = \frac{S_{tank,j} m_{tank,j}}{m_{tank,j} - F_{MD, dist}}$$  \hspace{1cm} (26)$$

$$T_{tank,j+1} = \frac{F_{MD, brine} T_{MD, brine} + (m_{tank,j} - F_{MD, feed}) T_{tank,j}}{m_{tank,j} - F_{MD, dist}}$$  \hspace{1cm} (27)$$

$$m_{tank,j+1} = m_{tank,j} - F_{MD, dist}$$  \hspace{1cm} (28)$$

where $F_{MD, dist}$ is the distillate production rate (kg/h) of the DCMD module, $F_{MD, brine}$ and $T_{MD, brine}$ are the mass flow (kg/h) and temperature (°C) of the brine leaving the DCMD module respectively, and $F_{MD, feed}$ is the mass flow of the feed water into the DCMD module.

Under the single-pass operation, the concentrated brine leaving the DCMD membrane module is discharged from the system (Fig. 4); thus, the feed water temperature and salinity remain unchanged.

During the solar thermal driven DCMD process, heat loss to the environment from the solar thermal collector, the membrane module, the feed tank, and piping can occur [30]. Heat loss from the solar thermal collector is embedded in its efficiency coefficients (e.g. $a1$ and $a2$). For simplicity, it is assumed that the system can be fully insulated and heat loss from other system components is negligible.
4. Results and discussions

4.1. Mass and heat transfer within the DCMD module

Co-current and counter-current flow result in two distinctive temperature profiles along the channel length (Fig. 5). In the co-current flow, the feed temperature decreases while the distillate temperature increases, resulting in a continuous decrease in the temperature difference ($\Delta T$) between the feed and distillate channels along the membrane module. In contrast, in the counter-current flow, the feed and distillate temperatures decrease at a similar rate; thus, $\Delta T$ is relatively constant along the channel length (Fig. 5).

![Fig. 5. Water flux, bulk feed temperature ($T_{f,b}$), and bulk distillate temperature ($T_{d,b}$) along the DCMD module channels under (A) co-current and (B) counter-current flow. Operating conditions: feed inlet temperature, $T_{f,in}$ = 60 °C; distillate inlet temperature, $T_{d,in}$ = 25 °C; feed and distillate cross flow velocities of 5.2 cm/s (i.e. equivalent to process water circulation rates of 900 L/h); feed inlet salinity, $S_{f,in}$ = 35,000 ppm.](image)

The difference in $\Delta T$ behaviour along the channel length results in two distinctive water flux profiles under the two flow modes (Fig. 5). In the co-current flow, water flux decreases rapidly along the channel length as a result of the reduction in $\Delta T$ between the feed and distillate streams (Fig. 5A). Due to the exponential relationship between water vapour pressure and temperature (Eq. 3), water flux decreases at a faster rate than the decline in $\Delta T$ in the co-current flow. Indeed, between the module inlet and outlet, water flux decreases by 20 times whereas $\Delta T$ only decreases by 12 times (Fig. 5A). Under the counter-current flow mode, a gradual decrease in water flux is observed despite a constant $\Delta T$ along the channel length (Fig. 5B). This is because the water
vapour partial pressure difference between the feed and distillate channels depends not only on \( \Delta T \) but also the temperature of the feed.

According to Eq. 9 and Eq. 10, temperature polarisation is reduced when the water flux decreases. Thus, for both flow modes, temperature polarisation coefficient (\( \Phi \)) increases along the channel length (Fig. 6A). It is noteworthy that \( \Phi \) values from this (Fig. 6A) and other modelling investigations [24, 39] are lower than those obtained in experimental studies using lab-scale DCMD modules (i.e. \( \Phi >0.8 \)) [40-42]. This is because lab-scale experimental studies are often based on operating parameters (e.g. high feed/distillate cross flow velocities and small membrane surface area) that are not realistic for pilot or full-scale operations.

**Fig. 6.** Simulated temperature polarisation coefficient (\( \Phi \)) and thermal efficiency (\( \Pi \)) along the DCMD module channels under co-current and counter-current flow. Operating conditions: feed inlet temperature, \( T_{f,\text{in}} \) = 60 °C; distillate inlet temperature, \( T_{d,\text{in}} \) = 25 °C; feed and distillate cross flow velocities of 5.2 cm/s (process water circulation rates of 900 L/h); feed inlet salinity, \( S_{f,\text{in}} \) = 35,000 ppm.

Thermal efficiency (\( \Pi \)) decreases along the channel length under both co-current and counter-current flow modes (Fig. 6B). At the feed inlet, \( \Pi \) is 0.52 and 0.60 under co-current and counter-current flow, respectively. At the feed outlet, \( \Pi \) decreases to 0.43 for co-current flow and to 0.38 for counter-current flow. It has been established that operating the DCMD process at a higher feed temperature increases the process thermal efficiency [43-45]. Thus, the decrease in \( \Pi \) can be attributed to the declined feed temperature along the channel length (Fig. 5).

### 4.2. Performance of the DCMD module

The relationships between water flux and thermal efficiency with feed temperature at the module level are demonstrated in Fig. 7. These results are consistent with the data presented in the previous section. It is interesting to note that under the two flow modes, the DCMD module
exhibits similar thermal efficiency but different water flux, particularly at high feed inlet temperatures (Fig. 7). The divergence in the module water flux between the two flow modes at high feed inlet temperatures is attributed to the $\Delta T$ variation along the module channels (Fig. 5). Under counter-current flow, at a high feed inlet temperature, a large $\Delta T$ value can be obtained and maintained along the membrane module, resulting in a high module water flux. On the other hand, under co-current flow, a high feed inlet temperature results in a large $\Delta T$ only at the module entrance, and $\Delta T$ decreases rapidly along the membrane module. As a result, the overall module water flux under co-current is lower than that under counter-current flow (Fig. 7).

**Fig. 7.** Influences of feed inlet temperature on the water flux and thermal efficiency of the DCMD module under co-current and counter-current flow. Other operating conditions: distillate inlet temperature, $T_{d,in} = 25 ^\circ C$; feed and distillate cross flow velocities of 5.2 cm/s (process water circulation rates of 900 L/h); feed inlet salinity, $S_{f,in} = 35,000$ ppm. The DCMD module has 1.5 m long channels.

Increasing water cross flow velocities in the feed and distillate channels leads to an increase in $\Delta T$, thus improving both the module water flux and thermal efficiency (Fig. 8). Indeed, the simulation results show that increasing water cross flow velocities from 3.5 to 8.7 cm/s doubles the average $\Delta T$ under both co-current and counter-current flow modes. Moreover, increasing the cross flow velocities in the feed and distillate channels can also reduce the temperature polarisation effect within the DCMD module. As the cross flow velocities of the feed and distillate channels increase from 3.5 to 8.7 cm/s, the average temperature polarisation coefficient
(Φ) of the DCMD module at the feed inlet temperature of 60 °C increases from 0.49 to 0.64 under both co-current and counter-current flow modes.

The influences of water cross flow velocities on water flux and thermal efficiency of the DCMD membrane module are much smaller than those of feed inlet temperature. Increasing water cross flow velocities by 2.5 times (i.e. from 3.5 to 8.7 cm/s) results in an increase of the same magnitude in the module water flux and a slight improvement in the module thermal efficiency (Fig. 8). On the other hand, increasing feed inlet temperature from 30 to 80 °C exponentially raises the module water flux, and markedly increases the module thermal efficiency (Fig. 7). These results have an important implication on the optimisation of a solar thermal driven DCMD system. Indeed, when the supply of solar thermal is limited, increased feed water cross flow velocities can significantly reduce the DCMD feed inlet temperature. Thus, in an integrated system, increasing the cross flow velocities might even be counter-productive. The influence of water cross flow velocities on water flux and thermal efficiency of the solar thermal driven DCMD process will be discussed in the section 4.3.

The simulation results reported here provide two important implications towards process optimisation of a solar thermal driven DCMD system for seawater desalination. First, under the
operating conditions investigated in this study, the DCMD module exhibits similar thermal efficiency under co-current and counter-current flow modes; however, counter-current flow results in a higher water flux. Thus, counter-current flow is preferable to co-current flow when the supply of thermal energy is limited. Second, the hot brine (i.e. at 40 °C for the feed inlet temperature of 60 °C under counter-current flow) leaving the membrane module has a considerable amount of sensible heat. Indeed, Duong et al. [46] have experimentally demonstrated the benefits of brine recycling in the DCMD process with respects to thermal efficiency. Thus, by simply recycling the hot brine to the module, this sensible heat can be recovered to enhance the overall thermal performance of the DCMD process. The next section analyses the performance of a solar thermal driven DCMD process with brine recycling under real weather conditions.

4.3. Performance of a solar thermal driven DCMD process

A solar thermal driven DCMD system consisting of a 7.2 m² membrane module under counter-current flow and 22.6 m² of flat plate solar thermal collector was simulated using TRNSYS. The results show a highly intermittent nature of the solar thermal driven DCMD process. The process achieves maximum water flux and thermal efficiency when the feed inlet temperature reaches the highest value at peak solar radiation (Fig. 9). On the hottest day of the operation (i.e. the 2nd day), the system under brine-recycled operation obtains the maximum water flux of 2.7 kg/(m²·h) at peak feed inlet temperature of 52 °C (i.e. coinciding with the peak total solar radiation of 1100 W/m²), and produces 142 kg of distillate for the day. This daily distillate production rate is equivalent to 19.7 kg/day per 1 m² of DCMD membrane and 6.3 kg/day per 1 m² of solar thermal collector. The overall thermal efficiency of the system varies from 0 (in early morning and late afternoon) to about 0.5 (at peak solar radiation).

Results from our simulation are comparable to the data from several previous pilot studies (Table 1). Compared to these pilot scale data, the simulated solar thermal driven DCMD system achieves a higher daily distillate production rate per 1 m² of membrane. On the other hand, in terms of the collector area, the simulated daily distillate production rate is lower when comparing to the data experimentally obtained from pilot operation (Table 1). These differences are attributed to the higher water flux but lower thermal efficiency of DCMD as compared to AGMD and PGMD. It is noteworthy that no previous pilot scale solar thermal driven DCMD studies are available for a direct comparison.
Fig. 9. Variations in (A) the total solar radiation, and (B) the feed inlet temperature ($T_{f,in}$), water flux, and thermal efficiency of the solar thermal driven DCMD process during a brine-recycled operation. Operating conditions: $T_{d,in} = 25$ °C, water cross flow velocities in DCMD channels of 5.2 cm/s (i.e. equivalent to process water circulation rates of 900 L/h). The simulation data are for the 1st to the 5th of January 1991.
Table 1. Comparisons between the simulated DCMD system and the data from pilot operations of solar thermal driven MD systems in the literature.

<table>
<thead>
<tr>
<th>Configurations</th>
<th>Simulated data from this study</th>
<th>Pilot data from the literature</th>
</tr>
</thead>
<tbody>
<tr>
<td>Membrane area (m²)</td>
<td>7.2</td>
<td>10  -  8</td>
</tr>
<tr>
<td>Thermal collector area (m²)</td>
<td>22.6</td>
<td>5.73  5.73  5.9</td>
</tr>
<tr>
<td>Daily distillate production (kg/d)</td>
<td>142</td>
<td>&lt;120  64.17  81</td>
</tr>
<tr>
<td>Distillate production/membrane area (kg/(m²·d) membrane)</td>
<td>19.7</td>
<td>&lt;12.0  -  10.13</td>
</tr>
<tr>
<td>Distillate production/collector area (kg/(m²·d) collector)</td>
<td>6.3</td>
<td>&lt;20.9  11.2  13.7</td>
</tr>
</tbody>
</table>

Our simulation data highlight the need to consider MD operation together with the source of thermal energy. Unlike the results in section 4.2 when the DCMD module is considered independently of the energy source, in an integrated system, increasing the DCMD water cross flow velocities from 3.5 to 8.7 cm/s decreases the maximum water flux and thermal efficiency at peaked hours from 3.2 to 2.5 kg/(m²·h) and from 0.48 to 0.41, respectively (Fig. 10). This is because the feed inlet temperature is no longer independent from the feed cross flow velocity. Increasing feed cross flow velocity shortens the residence time of the feed within the thermal collector, thus leading to a decrease in DCMD feed inlet temperature (Fig. 10). Operating the solar thermal DCMD process at low cross flow velocities is beneficial with respects to water flux and thermal efficiency. However, the seawater DCMD process at low cross flow velocities might be strongly affected by concentration polarisation effect and membrane scaling at high process water recoveries [46, 47]. Thus, further simulations of the solar thermal driven seawater DCMD process that take account of concentration polarisation effect and membrane scaling are recommended.
**Fig. 10.** Influence of water cross flow velocities on the feed inlet temperature, water flux, and thermal efficiency of the solar thermal driven DCMD process at peak solar radiation. Operating conditions: $T_{\text{d,in}} = 25 \, ^\circ\text{C}$. The simulation data are for the 2\textsuperscript{nd} of January 1991.

The simulation data also confirm the benefits of brine recycling during the solar thermal driven seawater DCMD desalination process (Fig. 11). Returning the warm brine to the feed tank in lieu of discharging it from the process recovers the sensible heat of the brine [30, 46], thus elevating the temperature of the feed water in the feed tank. The warmer is the feed water in the feed tank, the hotter feed water can be obtained at the outlet of the solar thermal collector. As demonstrated in Fig. 11A, during a single-pass operation at water cross flow velocities of 5.2 cm/s, the peak collector outlet temperature is 40 °C (i.e. with seawater feed temperature of 25 °C), whereas the brine-recycled operation achieves the peak collector outlet temperature of 52 °C. As a result, the peak water flux of the brine-recycled solar thermal driven DCMD process significantly increases to 2.7 kg/(m\textsuperscript{2}\cdot h) as compared to 1.2 kg/(m\textsuperscript{2}\cdot h) obtained in a single-pass process (Fig. 11B). Accordingly, the peak process thermal efficiency also increases from 0.35 to 0.45 when the operation mode is switched from single-pass to brine-recycled (Fig. 11C).
Fig. 11. (A) thermal collector outlet temperature, (B) water flux, and (C) thermal efficiency of the solar thermal driven DCMD process during the daytime at two operation modes: single-pass and brine-recycled. Operating conditions: $T_{d,in} = 25 \degree C$, water cross flow velocities in DCMD
channels of 5.2 cm/s (i.e. equivalent to process water circulation rates of 900 L/h). The simulation
data are for the 2nd of January 1991.

In addition to the water cross flow velocity and operation mode, other design factors such as
sizes of the MD feed tank, the solar thermal collector, and the MD membrane module as well as
solar radiation condition can exert strong influences on water flux and thermal efficiency of the
solar thermal DCMD system. Reducing the MD feed tank volume leads to an increase in the feed
water temperature (Eq. 27) and the collector outlet temperature, thus increasing the system water
flux and thermal efficiency [29]. Increasing the solar thermal collector area can improve the
system performance but requires a higher capital expenditure. It is noted that simulation results
from this study are from optimistic solar radiation condition. On a typical winter day (i.e. the 1st
of July), under brine-recycled operation, the simulated peak water flux and thermal efficiency are
1.4 kg/(m²·h) and 0.24, respectively.

5. Conclusions

Computer simulations were conducted to assess the performance of an integrated solar
thermal driven seawater DCMD system. The results show that DCMD operation under co-current
and counter-current flow results in two distinctive profiles in water flux, temperature polarisation
effect, and thermal efficiency along the channel length. Although, under the two flow modes,
similar thermal efficiency was obtained, the counter-current flow mode is preferable over co-
current with respects to water flux, particularly at high feed operating temperatures. More
importantly, the simulated data highlight the need to consider the DCMD process together with
the source of thermal energy. When considered in isolation from the supply of thermal energy,
increasing water cross flow velocities inside the module channels increases the overall water flux
and thermal efficiency of the DCMD membrane module. However, when the membrane module
is coupled with the solar thermal collector, increasing the feed and distillate cross flow velocities
reduces the water flux and thermal efficiency. Using previously recorded weather data, it is
shown that, each day a solar thermal driven DCMD system with brine recycling can produce 19.7
kg of clean water per m² of membrane or 6.3 kg of clean water per m² solar thermal collector.

Nomenclature

\[ C_m \] Membrane mass transfer coefficient, kg/(m²·h·Pa)
\[ C_p \] Specific heat capacity, kJ/(kg·K)
\[ D \] Water diffusion coefficient, m²/s
\[ dm \] Water mass flow through an incremental membrane area, kg/h
- $d_h$: Hydraulic diameter of the module channels, m
- $dQ$: The heat flux through an incremental membrane area, kJ/h
- $dx$: The length of an incremental membrane area, m
- $F_{MD,\text{brine}}$: The MD brine mass flow, kg/h
- $F_{MD,\text{feed}}$: The feed stream mass flow, kg/h
- $F_{MD,\text{dist}}$: Distillate production rate, kg/h
- $h_d$: Mass transfer coefficient of the distillate stream, W/m$^2$·K
- $h_m$: Mass transfer coefficient of the membrane, W/m$^2$·K
- $h_f$: Mass transfer coefficient of the feed stream, W/m$^2$·K
- $i$: Incremental membrane area index
- $J$: Water flux, kg/(m$^2$·h)
- $j$: Simulation step
- $k_g$: Gas thermal conductivity, W/(m·K)
- $k_m$: Membrane thermal conductivity, W/(m·K)
- $k_s$: Membrane polymer thermal conductivity, W/(m·K)
- $L$: Membrane module channel length, m
- $M$: Molecular weight of water, kg/mol
- $m_d$: Distillate mass flow, kg/h
- $m_f$: Feed mass flow, kg/h
- $m_{\text{tank}}$: The mass of water in the MD feed tank, kg
- $Nu$: Nusselt number
- $P$: Total pressure inside membrane pores, Pa
- $P_a$: Partial pressure of air inside membrane pores, Pa
- $P_{m,f}$: Water vapour pressure at the feed membrane surface, Pa
- $P_{m,d}$: Water vapour pressure at the distillate membrane surface, Pa
- $P_0$: Water vapour pressure of pure water, Pa
- $Pr$: Prandtl number
- $Pr_b$: Prandtl number in the fluid
- $Pr_m$: Prandtl number at the membrane surface
- $Q$: Heat flux through the membrane, kJ/(m$^2$·h)
- $R$: Universal gas constant, J/(mol·K)
- $Re$: Reynolds number
- $r$: Membrane pore radius, m
- $S_{\text{f,in}}$: Feed inlet salinity, ppm
- $S_{\text{f,out}}$: Feed outlet salinity, ppm
- $S_{\text{tank}}$: Salinity of water in the MD feed tank, ppm
- $T$: Mean water vapour temperature inside the membrane pores, K
- $T_{b,d}$: Temperature in the bulk distillate stream, K
- $T_{b,f}$: Temperature in the bulk feed stream, K
- $T_{m,d}$: Temperature at the distillate membrane surface, K
- $T_{m,f}$: Temperature at the feed membrane surface, K
- $u$: Fluid cross flow velocity, m/s
- $W$: Flow channel width, m
- $x_{\text{salt}}$: Molar fraction of salt
- $x_{\text{water}}$: Molar fraction of water

Greek symbols
Acknowledgments

The Vietnam International Education Development (VIED), under the Ministry of Education and Training (MoET), and the University of Wollongong (UOW) are acknowledged for PhD scholarship support to Hung C. Duong.

References


