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Specific energy requirement of direct contact membrane distillation

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\begin{abstract}

The study aims to provide a clear picture of the thermal energy requirements of Direct Contact Membrane Distillation (DCMD) system as function of different variables influencing the specific energy consumption. This includes membrane properties, operating conditions, recovery factor and the option of heat recovery from the permeate and retentate streams.

We simultaneously analyze the variation in specific energy demand and membrane surface area needed as a function of the membrane characteristics, operating conditions and recovery rate, taken as a design parameter. We observe that the specific energy demand of DCMD shows a relatively weak dependence on temperature polarization and membrane properties considered in the current study and a strong dependence on the recovery rate.

The advantages of using a heat exchanger very much depends on the recovery rate of the process.

\end{abstract}

1. Introduction

Membrane processes significantly contribute to modern desalination and wastewater treatment sectors. In addition to the conventional processes (ultrafiltration, microfiltration, nanofiltration, reverse osmosis, electrodialysis, etc.), relatively new processes with different separation potential, driving force and energy consumption are gaining interest. Membrane distillation (MD) is a prominent example of the latter. The process uses a vapor pressure difference, created across a microporous hydrophobic membrane mainly through a temperature difference, as a driving force. It offers the benefit of using low grade heat to induce the required driving force and can treat highly concentrated solutions such as reverse osmosis brines. Due to theoretical 100% rejection of all nonvolatile components, the product is of very high purity.

For a given feed, the performance of MD is mainly dependent upon membrane characteristics and process variables (operating conditions, module characteristics, applied configuration, etc.). High permeability, high liquid entry pressure (LEP), stable hydrophobic character and low thermal conductivity are the main requisites for MD membranes. In order to incorporate these features, membranes with different porosities, pore sizes, materials and hydrophobic characteristics have been fabricated, making it interesting to compare the process performance of these membranes with commonly used commercial membranes prepared for other separation purposes. Process design improvements in MD mainly focus on improving heat and mass transfer (Yang et al., 2011a; Ali et al., 2015; Phattaranawik et al., 2001) and energy recovery (Geng et al., 2014; Lin et al., 2014). Temperature polarization (TP), defined as the difference in bulk and membrane surface temperatures, plays an important role in governing heat and mass transport across the membrane (Ali et al., 2013). In DCMD, a temperature polarization is observed on the feed side (the temperature at the interface is lower than in the bulk because of the solvent vaporization and heat conduction through the membrane) and on the distillate side as well (the temperature at the interface is higher than in the bulk due to transfer of heat from feed side by condensation and heat conduction from the retentate compartment). Various experimental and theoretical approaches have been reported to quantify TP in MD (Ali et al., 2013; Gryta et al., 1997; Gryta and Tomaszewska, 1998; Phattaranawik et al., 2003; Tamburini et al., 2013), showing that the quantitative effect of temperature polarization on the overall process performance must be accounted for.

To better analyze the true commercial potential of MD, besides technical aspects, economical standing position of the process must
The objective of the current work was then to provide an analysis of the energy demand of DCMD including the recovery rate together with other common variables such as membrane characteristics, temperature polarization and option of using an energy recovery device.
<table>
<thead>
<tr>
<th>Configuration</th>
<th>Membrane characteristics</th>
<th>Operating conditions</th>
<th>Feed type</th>
<th>SEC (kWh/m³)</th>
<th>Plant capacity (m³/h)</th>
<th>Refs.</th>
</tr>
</thead>
<tbody>
<tr>
<td>DCMD</td>
<td>Spiral wound PTFE (SEP GmbH), Pore size 0.2 μ, porosity 80%</td>
<td>35–80 5–30</td>
<td>Radioactive solution</td>
<td>600–1600</td>
<td>0.05</td>
<td>Harasimowicz and Chmielewski (1999)</td>
</tr>
<tr>
<td>AGMD</td>
<td>PTFE, Pore size 0.2 μ</td>
<td>60–85</td>
<td>Seawater</td>
<td>140–200</td>
<td>0.2–20</td>
<td>Koschikowski et al. (2003)</td>
</tr>
<tr>
<td>AGMD</td>
<td></td>
<td>313–343</td>
<td>Brackish water</td>
<td>30.8</td>
<td></td>
<td>Bouguecha et al. (2005)</td>
</tr>
<tr>
<td>AGMD</td>
<td></td>
<td>–</td>
<td>Seawater</td>
<td>200–300</td>
<td>3.46–19</td>
<td>Banat et al. (2007)</td>
</tr>
<tr>
<td>DCMD in hybrid systems</td>
<td>PP modules from Microdyn Nadir, Pore size 0.2 μ, porosity 73%</td>
<td>–</td>
<td>Seawater</td>
<td>1.6–27.5</td>
<td>931 (overall)</td>
<td>Macedonio et al. (2007)</td>
</tr>
<tr>
<td>DCMD</td>
<td>Commercial membranes from Membrana with Pore size 0.2 μ and thickness 91 μ</td>
<td>39.8–59 13.4–14.4</td>
<td>Distilled water</td>
<td>3550–4580</td>
<td>–</td>
<td>Criscuoli et al. (2008)</td>
</tr>
<tr>
<td>VMD</td>
<td>PP, thickness 35 μ, Pore size 0.1 μ</td>
<td>15–22</td>
<td>Underground water</td>
<td>8100.8–9089.5</td>
<td>2.67–6.94</td>
<td>Wang et al. (2009)</td>
</tr>
<tr>
<td>PGMD</td>
<td>n.p</td>
<td>60–80</td>
<td>Seawater</td>
<td>200–360</td>
<td>0.007–0.02</td>
<td>Raluy et al. (2012)</td>
</tr>
<tr>
<td>AGMD</td>
<td>LDPE, thickness 76 μ, Pore size 0.3 m, porosity 85%, Am 7.4 m²</td>
<td>50–70</td>
<td>Tap water, synthetic seawater</td>
<td>~65 to ~ 127</td>
<td>0.0034–0.0094</td>
<td>Duong et al. (2016)</td>
</tr>
<tr>
<td>VMD</td>
<td>Flat sheet PP, thickness 400 μm, Pore size 0.1 μ, porosity 70%, Am 5 m²</td>
<td>80</td>
<td>Distilled water</td>
<td>130</td>
<td>–</td>
<td>Criscuoli et al. (2013)</td>
</tr>
<tr>
<td>DCMD</td>
<td>PVDF hollow fiber, thickness 240 μm</td>
<td>80 30</td>
<td>Simulated reverse osmosis brine</td>
<td>~130 to 1700</td>
<td>–</td>
<td>Guan et al. (2014)</td>
</tr>
<tr>
<td>DCMD</td>
<td>PTFE with PP support, mean pore size 0.5 ± 0.08, porosity 91 ± 0.5, active layer thickness 46 ± 1 μm, Am 0.67 m²</td>
<td>60 18–21</td>
<td>Wastewater</td>
<td>1500</td>
<td>3.85</td>
<td>Dow et al. (2016)</td>
</tr>
<tr>
<td>DCMD</td>
<td>Several commercial membranes with different characteristics</td>
<td>85 20</td>
<td>Seawater</td>
<td>~697 to 10,457</td>
<td>–</td>
<td>Ali et al. (2012)</td>
</tr>
</tbody>
</table>

VMD—vacuum membrane distillation.
AGMD—air gap membrane distillation.

* Mean feed temperature obtained through solar collectors.

b Calculated from GOR data.
The optimum operating conditions in terms of energy demand and membrane area requirements have also been identified. The effect of membrane characteristics has been assessed by considering two membranes i) a commercially available membrane (Accurel, Polypropylene) originally made for gas/liquid contacting purposes but widely used for MD experiments and ii) a lab made PVDF membrane specifically developed for MD applications and precisely described in the literature (Yang et al., 2011b). The Accurel membrane can be considered as a reference since it is commercially available, whereas the other one which has been developed specifically for membrane distillation has very good performances and can be considered as an achievable target for a company willing to develop new MD membranes. The effect of temperature polarization (and therefore of hydrodynamic conditions) has been incorporated by assuming a fixed temperature polarization (here of 5 °C) on both feed and permeate sides at different feed and permeate temperatures.

2. Materials and methods

2.1. Membrane characteristics

In this work, we consider two types of hollow fiber membranes: a commercial PP membrane from Membrana (Accurel™ PP S6/2) and a PVDF membrane developed on purpose by the group of Fane and Wang and described in Yang et al. (2011b). The characteristics of both membranes and module dimensions are given in Table 2. The Accurel™ membrane was not designed for MD but is commercially available and technically compatible with the requirements of MD (porous, hydrophobic polymer matrix). Surprisingly, the liquid entry pressure for the Accurel (140 kPa) is higher than the one claimed by the authors of the PVDF (138 kPa), whereas the nominal pore sizes are in an opposite ratio (0.55 versus 0.41 respectively). We assumed that this can be explained by the difference in contact angles between water and the material the two membranes are made of (polypropylene and PVDF) and by the prospective surface treatments these may have been submitted to.

2.2. Description of the simulated system

The system considered in the present study is shown diagrammatically in Figs. 1 and 2. We will assume that it processes 1 m³/h of feed flow and that it produces a flow $\varphi$ m³/h of distillate water, $\varphi$ then being the recovery rate for the system. The system is fitted with a hollow fiber membrane module. Heat can be supplied to the membrane module so as to compensate the heat losses through the membranes. When a heat exchanger is considered, two possible arrangements are considered, whether the heat exchanger is located on the distillate line (Fig. 1) or on the retentate line (Fig. 2). If the feed stream is available at ambient temperature, preheating is then necessary before entering the DCMD module at temperature $T_{F,a}$. On the other side, we assume that the distillate compartment, which needs to be kept at a low temperature, is circulated with a distillate stream entering the module (on the shell-side) at ambient temperature $T_{D_1}$, which is set for the purpose of our modeling at 20 °C. This stream is warmed during the process by the conductive and condensation heat fluxes to a temperature $T_{D, a}$. The outlet distillate temperature has to be lower than the inlet feed one, $T_{F,a}$. The power to maintain a low temperature in this compartment is often overlooked in technical studies. In the present work, we use the flow that need to be pumped through the distillate compartment as an indicator to reflect the need for cooling the system. In practice, the cooling of distillate stream could be achieved by using a heat pump, though in general quite an expensive option.

The detailed conditions considered for simulation are provided in Table 2.

2.2.1. Axial and radial temperature distribution in a DCMD module

Temperature and temperature differences across the membranes are key parameters in a DCMD module. We have developed a model to simulate the local temperature differences in the radial direction across the hollow fiber section as well as in the axial direction, so as to define in which operating conditions the temperature gradients in both directions should be accounted for and those where it can be neglected. The model is based upon application of simultaneous heat and mass balance on differential elements along the fiber to determine temperature and flux profiles. The details on the model derivation and its validation for PP membrane considered in the present study have been given elsewhere (Ali et al., 2016a,b).

2.2.2. Limiting conditions and hypothesis

In this part of our work, we consider flow conditions such that the feed and distillate cross flows are high enough for the temperatures to be constant all along the module length. The permeate temperature at the module inlet ($T_{D,i}$) is set at 20 °C while the permeate temperature at the module outlet is equal to the average temperature in the permeate compartment ($T_{D,a}$) assumed ideally mixed. On the feed side, a heating system is needed to keep the temperature at a set value. Here again, we assume that the feed compartment is well mixed along the fiber axis. Evaporation of the solvent on the feed side and condensation on the distillate side combined with conduction through the membrane can create a temperature gradient in the direction radial to the main flow (“temperature polarization”).

In order to incorporate a possible effect of temperature polarization to the present study, a preliminary evaluation of such temperature differences was made using a model developed by one of us (Ali et al., 2016a). The results reported in Appendix A show that in our conditions and for the membranes considered, this difference was never larger than 4 K either on the permeate and feed side. Therefore, in order evaluate the effect of temperature polarization far from ideal conditions, we assumed a difference between the bulk and the wall temperature is 5 °C on either side (Table 3):

$$T_{F,m} - T_{F,a} = -5 ^\circ C$$

$$T_{D,m} - T_{D,a} = +5 ^\circ C$$

2.3. Modelling procedure

2.3.1. Membrane module analysis

Flux in MD can be described with the following equation.

$$J_m = B (P_{F,m} - P_{D,m})$$  (1)

where $P_{F,m}$ and $P_{D,m}$ are vapor pressures at the membrane surface on feed and permeate sides, respectively. $B$ is the distillation coefficient of the membrane (i.e. membrane permeability). Calculation of $B$ depends mainly upon the membrane characteristics. $B$ can be calculated using a Knudsen diffusion model (Eq. (2)) if $r < 0.5 \lambda$, where $r$ and $\lambda$ represent
Table 2 – Main properties of the membranes considered in current study.

<table>
<thead>
<tr>
<th>Property</th>
<th>PP</th>
<th>PVDF</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fiber inner radius ($r_1$)</td>
<td>0.9/0.46 mm</td>
<td>1.3/0.735 mm</td>
</tr>
<tr>
<td>Fiber outer radius ($r_2$)</td>
<td>1.3/0.46 mm</td>
<td>1.3/0.735 mm</td>
</tr>
<tr>
<td>Membrane thickness ($h_m$)</td>
<td>0.4/0.275 mm</td>
<td>0.4/0.275 mm</td>
</tr>
<tr>
<td>Nominal pore size ($r$)</td>
<td>0.22/0.0641 μm</td>
<td>0.22/0.0641 μm</td>
</tr>
<tr>
<td>Maximum pore size</td>
<td>0.55/0.421 μm</td>
<td>0.55/0.421 μm</td>
</tr>
<tr>
<td>LEP values determined for the largest pores of membranes</td>
<td>140 (Macedonio et al., 2014)/138 kPa</td>
<td>140 (Macedonio et al., 2014)/138 kPa</td>
</tr>
<tr>
<td>Membrane thermal conductivity ($\gamma$)</td>
<td>0.08/0.05 W/K m</td>
<td>0.08/0.05 W/K m</td>
</tr>
<tr>
<td>Porosity ($\phi$)</td>
<td>73/85 %</td>
<td>73/85 %</td>
</tr>
<tr>
<td>Hollow fiber length applied in the simulation ($L$)</td>
<td>1.1 m</td>
<td>1.1 m</td>
</tr>
<tr>
<td>Membrane characteristic parameter ($B$)</td>
<td>$8.81 \times 10^{-8}$</td>
<td>$1.31 \times 10^{-7}$</td>
</tr>
</tbody>
</table>

1st and 2nd numbers in each pair represent PP and PVDF membrane, respectively.

Table 3 – Thermal conditions considered in current study.

<p>| DCMD system assuming no temperature polarization |
|-----------------------------------------------|----|------|</p>
<table>
<thead>
<tr>
<th>$T_{Da}$ (°C)</th>
<th>$T_{Dm}$ (°C)</th>
<th>$\Delta T$ (°C)</th>
<th>$T_{fa}$ (°C)</th>
<th>$T_{fm}$ (°C)</th>
<th>$T_{Di}$ (°C)</th>
<th>$T_{Fini}$ (°C)</th>
<th>$C_{P_Di}$ (J/K kg)</th>
<th>$C_{P_Do}$ (J/K kg)</th>
<th>$\rho_{Di}$ (kg/m$^3$)</th>
<th>$\rho_{Do}$ (kg/m$^3$)</th>
</tr>
</thead>
<tbody>
<tr>
<td>20</td>
<td>20</td>
<td>15</td>
<td>35</td>
<td>35</td>
<td>20</td>
<td>20</td>
<td>4164</td>
<td>4158</td>
<td>999</td>
<td>998</td>
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<td>995</td>
<td>993</td>
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<td>20</td>
<td>4144</td>
<td>989</td>
<td>989</td>
<td></td>
</tr>
</tbody>
</table>

<p>| DCMD system with the presence of temperature polarization |
|-----------------------------------------------|----|------|</p>
<table>
<thead>
<tr>
<th>$T_{Da}$ (°C)</th>
<th>$T_{Dm}$ (°C)</th>
<th>$\Delta T$ (°C)</th>
<th>$T_{fa}$ (°C)</th>
<th>$T_{fm}$ (°C)</th>
<th>$T_{Di}$ (°C)</th>
<th>$T_{Fini}$ (°C)</th>
<th>$C_{P_Di}$ (J/K kg)</th>
<th>$C_{P_Do}$ (J/K kg)</th>
<th>$\rho_{Di}$ (kg/m$^3$)</th>
<th>$\rho_{Do}$ (kg/m$^3$)</th>
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</tbody>
</table>
mean pore size and mean free path of water vapors, respectively (Khayet, 2011). \( \lambda \) can be calculated by using Eq. (2):

\[
\lambda = \frac{k_b T}{\sqrt{2 \pi P_{avg} \sigma^2}}
\]

(2)

where \( k_b \), \( P_{avg} \) and \( \sigma \) represents the Boltzmann constant, the mean pressure inside the membrane pore and the collision diameter (2.641 Å for water vapor), respectively.

\[
B_k = \frac{22}{9} \frac{\sqrt{\pi T}}{\tau^3}
\]

(3)

In the range, 0.5\( \lambda \) < \( r \) < 50\( \lambda \), \( B \) can be calculated using a Knudsen-molecular diffusion model (Khayet, 2011) (Eq. (4)):

\[
B = \left( \frac{3}{2} \frac{\lambda}{r} \left( \frac{\pi R T}{8 M} \right)^{1/2} + \frac{6}{e^2} \frac{\rho_d R T}{\rho_f M} \right)^{-1}
\]

(4)

Where \( PD \) = 1.895 \times 10^{-5} \times (T)^{0.72} \ Pa = 101325 Pa

\( B \) is slightly dependent on the experimental conditions. \( B \) was calculated in each case considered in our work by using the data provided in Table 2 and Eqs. (2)–(4).

The relationship between the water vapor pressure and the temperature is given by the Antoine equation (Banat et al., 2007):

\[
P_{Fm|Dm} = \exp \left( 23.2 - \frac{3816.44}{T_{Fm|Dm} + 273.15} - 46.13 \right)
\]

(5)

\( T_{Fm|Dm} \) [°C] and \( T_{Dm} \) [°C] are the feed and permeate temperatures at the membrane surface, respectively.

The water flux \( J_w \) can be obtained by combining Eqs. (1), (3)–(5) and the required membrane area \( S \) can be calculated as follows:

\[
S = \frac{Q_w \times 1000}{J_w}
\]

(6)

The total heat flux through the membrane, \( P_{tot} \), is governed by two mechanisms: 1) conduction across the membrane material and its gas filled pores (\( P_{con} \)), and 2) latent heat associated to the water vapor molecules (\( P_{v} \)). Therefore, the net heat flux transported through the membrane can be expressed as follows:

\[
P_{tot} = P_{con} + P_{v}
\]

(7)

The enthalpy balance on the feed compartment is:

\[
Q_{f} - \nu_f C_{pf} T_{f} + P_{internal} = \left( 1 - \nu_f \right) Q_{w} C_{pw} T_{w} + P_{v} + P_{tot}
\]

(8)

where \( P_{tot} \), \( P_{v} \) and \( P_{internal} \) are the total conductive heat flux, evaporative heat flux and power to be supplied to the membrane module in Watt, respectively. \( \nu_f \) and \( \nu_v \) refer to the feed and retentate densities in kg/m³. \( C_{pf} \) and \( C_{pw} \) denote the feed and retentate specific heats in J/kg K. \( Q_w \) is equal to 1 m³/h in the present simulation. \( \nu \) notifies the recovery rate.

Identically, the heat balance on the distillate side is given by Eq. (9):

\[
\left( Q_{D_{min}} C_{pD} \phi_{D} T_{Di} \right) + P_{tot} + P_{v} = \left( Q_{D_{min}} \right)
\]

(9)

\( T_{Di} \) and \( T_{Da} \) represent the distillate temperatures at the module inlet and outlet. \( C_{pD} \), \( C_{pD} \), \( \rho_{D} \), and \( \rho_{D} \) indicate the distillate specific heats and the distillate densities at the membrane inlet and outlet, respectively. The mass flux transferred through the membrane equals to \( \psi . Q_w \). The minimum distillate flow rate \( (Q_{D_{min}}) \) which allows maintaining \( \Delta T \) of \( T_{f} - T_{D} \) is calculated by solving Eq. (9).

The latent heat flux \( (P_v) \) used in Eq. (9) for dilute solutions is determined by:

\[
P_v = \psi . Q_w . \nu_f . H_v
\]

(10)

The heat of evaporation is transferred to the distillate by condensation. The heat of vaporization varies upon the evaporating water temperature. In the range of temperature 273–373 K, water latent heat of vaporization is expressed as below (Phattaranawik et al., 2001):

\[
H_v(T) = 1.7535 . T_{Fm} + 2042.4
\]

(11)

\( H_v \) and \( T_{Fm} \) are in kJ/kg and K, respectively. The evaporative heat flux transferred through the membrane per cubic meter of produced distillate, \( W_v \) [kWh/m³] is defined as:

\[
W_v = \frac{P_v}{\psi . Q_w}
\]

(12)

For one hollow fiber of length \( L \), internal radius \( r_1 \) and external radius \( r_2 \), the conductive heat flux can be calculated by (Cengel, 2002):

\[
P_{c|fiber} = \frac{2 \pi . \nu . \Delta T . L . (r_2 - r_1)}{8 \mu \ln \frac{r_2}{r_1}}
\]

(13)

where \( \gamma \) [W/m K] is the thermal conductivity of the membrane used. \( \Delta T \) is the temperature difference between feed and permeate sides. Various models have been considered to calculate \( \gamma \). In general the following expression is used (Raluay et al., 2012):

\[
\gamma = \epsilon . k_s + (1 - \epsilon) . k_p
\]

(14)

where \( k_p \) is the thermal conductivity of the material forming the membrane matrix and \( k_s \) is the thermal conductivity of the gas filling the membrane pores.

The membrane surface area of a fiber is determined as below:

\[
S_{fiber} = 2 \pi . r_1 . L
\]

(15)

Knowing the membrane area \( S \) from Eq. (6) and membrane area of a fiber \( S_{fiber} \), the number of hollow fibers required \( (N_{fiber}) \) is found by combining Eqs. (6) and (15):

\[
N_{fiber} = \frac{S}{S_{fiber}}
\]

(16)

Therefore, the total conductive heat flux \( P_{tot} \) in Watt) is obtained by combining Eqs. (13) and (16):

\[
P_{tot} = P_{c|fiber} . N_{fiber}
\]

(17)
The total conductive heat flux transferred through the membrane pores per cubic meter of distillate, \( W_{\text{ctot}} \) [kWh/m\(^3\)] is determined by:

\[
W_{\text{ctot}} = P_{\text{ctot}} \frac{1}{\varphi.Q_F}
\]

(A)

A DCMD module needs to be heated in order to compensate the heat loss by conduction and evaporation and to keep a preset average feed temperature (\( T_{F3} \)). For a given inlet feed temperature and a determined recovery rate, this energy requirement per m\(^3\) of distillate, \( W_{\text{internal}} \) [kWh/m\(^3\)] is defined as below:

\[
W_{\text{internal}} = \frac{(1 - \varphi).Q_F.\varphi.C_F.T_{F3} + P_{\text{ctot}} + P_a - Q_F.\varphi.C_F.T_{F3}}{\varphi.Q_F}
\]

(B)

Some preheating energy has to be supplied to the feed in order to reach the desired temperature at the module inlet (\( T_{F3} \)). For the DCMD system assuming no external heat exchanger, the external energy consumption per cubic meter of produce distillate is obtained by:

\[
W_{\text{external without exchanger}} = \frac{Q_F.\varphi.C_F.(T_{F3} - T_{Fini})}{\varphi.Q_F}
\]

(C)

An average outdoor temperature of 20 °C is assumed. In this case (without external heat exchanger), \( T_{Fini} = T_{F0} = 20°C \). The preheating energy supplied to the feed is changed when an external heat exchanger is added to the membrane distillation system. The preheating energy required in the presence of a heat exchanger is defined as below:

\[
W_{\text{external with exchanger}} = \frac{Q_F.\varphi.C_F.(T_{F3} - T_{F0})}{\varphi.Q_F}
\]

(D)

where \( T_{F0} \) denotes the cold fluid temperature at the outlet of the heat exchanger used.

The specific energy consumption per cubic meter distillate, \( E_p \) [kWh/m\(^3\)] is simply obtained by:

\[
E_p = W_{\text{external}} + W_{\text{internal}}
\]

(E)

2.3.2. Heat exchanger

Many descriptions of DCMD involve the use of heat exchangers in order to recover part of the spare heat available in the streams flowing out of the MD unit and discuss their interest. We therefore have considered this option and assumed that we could use it either on the retentate or on the distillate line. We have considered various surface areas for the heat exchangers adapted to the flow to be handled. The main design equations used for calculating the heat recovered through this device are presented in Appendix A.

3. Result and discussion

3.1. Identification of set conditions

A variation of temperature along and across the flow area is expected in MD due to heat and mass transfer. To identify the conditions under which the assumption of the variation in temperature along the module holds true, temperature profiles on feed and permeate sides were obtained under various temperature and hydrodynamic conditions. As expected, the temperature distribution along the fiber becomes more uniform with an increase in feed flow rate for any combination of feed and permeate temperatures. However, it was noted that for low feed temperatures, it is relatively easy to achieve the conditions under which the temperature distribution in axial direction can be approximated uniform. Bulk and surface profiles along the fiber corresponding to feed and permeate inlet temperatures of 308.15 K and 293.15 K, respectively are shown in Fig. 3(a) and for feed and permeate inlet temperatures of 333.15 K and 318.15 K in Fig. 3(b). Feed and permeate velocities are kept at 2.5 and 3 m/s respectively for these conditions. It can be noticed that the average temperature polarization both on feed and permeate sides remains less than 2 K under these conditions. In our calculation and in order to consider a worse case situation, the effect of temperature polarization was incorporated by considering its value of 5 K on each side.

3.2. Flux

The flux calculated for PP and PVDF membranes according to the model described in Section 2.3.1 is shown in Fig. 4. Knudsen-molecular and Knudsen diffusion model was applied for PP and PVDF membrane respectively in accordance with the conditions mentioned in Section 2.2.2. The water flux transferred through the membrane pores is proportional to the partial pressure difference and as expected, it is much lower whatever the recovery rate and membrane type, when considering a temperature polarization of 5°C (Eq. (1)). The increase in water flux at a higher average feed temperature is well known and is due to the increase in partial pressure difference (Eqs. (1) and (5)) even if the temperature difference between the two compartments is assumed constant at 15°C. The PVDF membrane would produce a 73% higher flux than the PP one.

In the case of a saline solution, the water vapor pressure would change with the salt concentration. The following correlation has been proposed to reflect the effect of salt concentration on water vapor pressure (Yun et al., 2006):

\[
P = P_0 (1 - x) 1 - 0.5x - 10x^2
\]

(F)

where \( P_0 \) is the vapor pressure of pure water and \( x \) is the mole fraction of NaCl in solution. In the case of a reverse osmosis brine (c.a. 75 g/L) concentrated by MD up to 110 g/L, i.e. \( \approx 2\) M in NaCl, the mole fraction of NaCl would be \( \approx 0.04 \). According to Eq. (23), the vapor pressure would be only 8% lower than for pure water. The impact on the actual mass flux (8%) is quite limited compared to the broad variation in sodium chloride concentration from 0 to 110 g/L.

3.3. Membrane area

The membrane area requirement increases linearly with the recovery rate and can be very large as soon as the recovery exceeds 10–20%. Whatever the membrane type, a high feed temperature (\( T_{F3} \)) is essential to keep the required membrane area to reasonable levels (Figs. 5 and 6). These values of membrane area are however huge and clearly indicate the need for process improvement in terms of less temperature polarization and the development of high performance membranes specifically developed for MD (low conductivity and high flux), as long as significant recovery rates are expected.
3.4. Minimum distillate flow rate

The minimum permeate flow rate \( (Q_{D_{\text{min}}}) \) which allows maintaining a given average temperature difference between the feed and permeate sides is determined by solving Eq. \( (9) \).

The minimum distillate flow rate (which is an indicator of the amount of energy that would be needed to cool the distillate compartment) increases with the recovery rate because of higher heat flux transferred through the membrane. It is much reduced (85–78% smaller) when the feed average temperature is increased from 50 °C to 80 °C, and this is directly linked to the smaller surface area needed at this higher temperature and to a latent heat which is only slightly changed. The part played by the temperature polarization is interesting since it appears that the system needs the same distillate flow rate whether...
the temperature polarization is considered or not. In fact, for a given recovery rate, the transfer of latent heat is almost the same whether polarization is important or not. As described in Eq. (13), the heat transported by conduction depends on the product of the difference in temperature by the membrane area. The effect of the temperature polarization is to decrease the former and increase the latter, both effects compensating each other here.

In any case, keeping the distillate compartment at a reasonably low temperature compared to the feed temperature would need unrealistic volumes of cool water (or their equivalent in electric power to operate a heat pump) unless the feed stream is maintained at quite a high temperature and, of course, the membrane used is a high performance one. Therefore, cooling the distillate compartment can be one of the factors limiting the recovery rate when operating a DCMD unit, though this aspect of the process is often overlooked in the literature.

### 3.5. Specific energy

#### 3.5.1. System without external heat exchanger

For a given average feed temperature \(T_{fa}\), the energy needed to keep the feed compartment at a desired temperature decreases with the increase in \(T_{fa}\) but increases slightly when in the presence of temperature polarization because of a higher conductive heat flux (Fig. 8). Working with a higher performance membrane, the system needs less internal energy due to the reduction in conductive flux.

The total specific energy (Eq. (22)) plotted as a function of recovery rate gives curves which are typically represented in Fig. 9.

The specific energy demand decreases with the increase in average feed temperature as expected, except at very low recovery rate. The horizontal lines represent \(W_{\text{int}}\), and by comparison to the dotted lines, one can appreciate the cost of preheating the feed to a given temperature as a function of the recovery rate. The specific energy decreases sharply with the recovery rate at low values of the latter because in this range, the energy to warm the whole stream from ambient temperature to the module inlet temperature is a significant fraction of the total energy requirement. For large recovery rates, this specific energy demand remains almost constant. The result shows that the temperature polarization for a given recovery rate has almost no impact on the overall energetic performance because to produce a given quantity of distillate (a given recovery rate), the increase in membrane surface nearly compensates the temperature difference in generating conductive heat flux. As expected, the DCMD system requires much less specific energy when a high performance membrane is used. This is due to the reduction in conductive heat flux. The results presented in Fig. 9 show that for an energy point of view, preheating the feed at a relatively high temperature provides a much better yield provided that the recovery rate exceeds 10% or so.

Figs. 7 and 10 illustrate that the system needs less minimum distillate flow rate and less membrane area (i.e. less investment cost) at low recovery rate. On another hand, the higher the recovery rate, the lower the specific energy requirement. The tradeoff obviously would depend on the cost of energy and materials and it is definitely case-specific. It is beyond the scope of the present study to set the exact operating point, but it is clear in Fig. 10 that this point drifts towards higher recovery rates for warmer feeds. As expected, the best operation condition is running the system at high average feed temperature.

#### 3.5.2. System with external heat exchanger

As in any thermal process, one searches to recycle heat fluxes by using heat exchangers to transfer heat from exiting streams back to feed flows. In membrane distillation, two options (at least) can be considered: one is to connect the retentate which exits from the module to the feed stream entering the module (case n’2 in the present work) and the other one using the distillate stream to pre-heat the feed (case n’1 in the present work) and this has been described in many papers before (Lin et al., 2014; Geng et al., 2014).

As can be seen in Figs. 11 and 12, based on the flow sheets shown in Figs. 1 and 2 respectively, the impact of using a heat exchanger decreases when the recovery rate increases. A comparison between Figs. 11 and 12 clearly shows the advantage of plugging the heat exchanger on the retentate line, which enters the heat exchanger at a higher temperature than when the heat exchanger is connected to the exit of the distillate line. The flow of retentate would obviously decrease as the recovery increases, but this should not impact the final energy demand since at high recovery rate, we observe a limited impact of the presence of a heat exchanger, since as commented about Fig. 9, the part played by the pre heating energy
Fig. 9 – Specific energy requirement as a function of recovery rate for DCMD systems assuming no temperature polarization and operating with Accurel membrane/PVDF one at 2 average feed temperatures when an external heat exchanger is not used (Q_F = 1 m³/h).

Green and red solid lines represent the internal energies at 35 °C and 80 °C, respectively. (For interpretation of the references to color in this figure legend, the reader is referred to the web version of this article.)

Fig. 10 – Specific energy consumption and membrane area as a function of recovery rate for DCMD assuming no temperature polarization and operating with Accurel membrane when an external heat exchanger is not used (Q_F = 1 m³/h).

The result shows that whatever T_F a, the DCMD system requires less specific energy especially at very low recovery rate (61% smaller at the recovery rate of 1%) by using the external heat exchanger N 2. Therefore, it would be better to recover the heat available in the retentate stream to preheat the feed instead of in the distillate one.

Overall, the specific energy remains quite high, compared to other water distillation processes such as multistage distillation (70 kWh/m³) or to reverse osmosis (3–7 kWh/m³). Even when abundant sources of hot feed is available, the implementation of a heat exchanger, especially on the retentate line may prove profitable only to allow an operation at lower recovery rates, while keeping the specific energy demand low.

4. Conclusion

We have considered DCMD using a hollow fiber module coupled with an external heat exchanger, an external heating system and an internal energy source in the membrane module. This system is simulated in order to evaluate the thermal energy requirement as a function of the recovery rate, i.e. the ratio between the amount of distilled water produced by the amount of feed processed. The influence of the temperature polarization on the energy consumption is considered.

When considering situations where the feed stream has to be heated, we find that the specific energy required to produce a given amount of distilled water decreases sharply when the recovery rate increases from 0 to c.a. 20%. This level of specific energy obviously decreases when the DCMD is operated at high feed temperature and with a high performance membrane, as this has been pointed out in previous studies. The cost of heating the feed from ambient temperature to a higher temperature is worth the energy spent, provided that the recovery rate is high enough.

Our results show that for a given recovery rate, the specific energy requirement of DCMD is only slightly dependent on “temperature polarization”. This can be explained as TP
increases the membrane surface area needed to produce a given amount of distilled water, and at the same time decreases the temperature gradient which is the driving force for heat conduction through the membranes. In the conditions in this simulation, both phenomena balance each other almost exactly.

We also found that, under the conditions considered in current study, a heat exchanger is more efficient if implemented on the retentate line better than on the distillate. The impact of using a heat exchanger is however limited to the range of recovery rates for which the energy required for pre heating the feed is significant compared to the total energy needed by the process, i.e. in the range between 0 and 20% in our conditions.

Running the same calculations but for two different membranes allowed to appreciate that when switching from a non-optimized membrane to an optimized one, the membrane surface area necessary for a specific duty can be almost halved, whereas the role of the membrane on the energy demand is important at low recovery rates, but fades away at high recovery rates (Fig. 9).

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Appendix A. Estimation of the temperature polarization.

We used the model developed by one of us and published in [XX] to compute the temperature difference between the bulk and surface of the membrane for a range of operating conditions. The data gathered in Table A1 show that the temperature difference between the bulk and the membrane surface (subscript m) either on the feed side (subscript f) or on the permeate side (subscript D) is in a range [0.06–3.83 K].

Table A1 – Temperature.

<table>
<thead>
<tr>
<th>Tf,in (K)</th>
<th>Tf,m (K)</th>
<th>Vf (m/s)</th>
<th>Vp (m/s)</th>
<th>Tm,m – Tm (K)*</th>
<th>Tf,m – Tf,m (K)*</th>
</tr>
</thead>
<tbody>
<tr>
<td>308</td>
<td>293</td>
<td>3.13</td>
<td>3.13</td>
<td>0.06/0.15</td>
<td>0.25/0.46</td>
</tr>
<tr>
<td>308</td>
<td>293</td>
<td>0.13</td>
<td>0.13</td>
<td>0.69/1.09</td>
<td>0.74/1.49</td>
</tr>
<tr>
<td>353</td>
<td>338</td>
<td>3.13</td>
<td>3.13</td>
<td>0.23/0.53</td>
<td>0.97/1.63</td>
</tr>
<tr>
<td>353</td>
<td>338</td>
<td>0.13</td>
<td>0.13</td>
<td>2.07/3.83</td>
<td>2.21/2.84</td>
</tr>
</tbody>
</table>

* 1st and 2nd value refers to PVDF and PP membranes, respectively.

Appendix B. Design of a heat exchanger.

The application of recuperative heat exchanger ensures the energy recovery from distillate and retentate leaving the module.

\[ E = \frac{1 - \exp[-NUT (1 - R)]}{1 - R \exp[-NUT (1 - R)]} \quad \text{For } R \neq 1 \]  
(24)

Or \[ E = \frac{NUT}{NUT + 1} \quad \text{For } R = 1 \]  
(25)

where \( R \) is the ratio of the heat capacities:

\[ R = \frac{C_{\text{min}}}{C_{\text{max}}} = \frac{\min (Q_f \cdot \rho_f \cdot C_p, (1 - \alpha) \cdot Q_r \cdot \rho_r \cdot C_p)}{\max (Q_f \cdot \rho_f \cdot C_p, (1 - \alpha) \cdot Q_r \cdot \rho_r \cdot C_p)} \]  
(26)

when the external heat exchanger N’1 is used,

\[ R = \frac{C_{\text{min}}}{C_{\text{max}}} = \frac{\min (Q_f \cdot \rho_f \cdot C_p, (1 - \alpha) \cdot Q_r \cdot \rho_r \cdot C_p)}{\max (Q_f \cdot \rho_f \cdot C_p, (1 - \alpha) \cdot Q_r \cdot \rho_r \cdot C_p)} \]  
(26-bis)

when the external heat exchanger N’2 is used.

The number of transfer units is defined by:

\[ \text{NUT} = \frac{K_S}{C_{\text{min}}} \]  
(27)

Knowing the efficiency of the heat exchanger, the total power exchanged (\( P_e \)) by two fluids in heat exchanger can be given by the relation:

\[ P_e = E \cdot C_{\text{min}} \cdot (T_{D,a} - T_{F,\text{fin}}) \]  
(28)

when the external heat exchanger N’1 is applied,

\[ P_e = E \cdot C_{\text{min}} \cdot (T_{D,a} - T_{F,\text{fin}}) \]  
(28-bis)

when the external heat exchanger N’2 is applied.

Also, when the external heat exchanger N’1 is applied,

\[ P_e = (\alpha \cdot Q_f + Q_{D,\text{min}}) \cdot \rho_D \cdot C_p \cdot (T_{D,a} - T_{h,\text{in}1}) = Q_f \cdot \rho_f \cdot C_p \cdot (T_{\text{co}1} - T_{\text{fin}1}) \]  
(29)

when the external heat exchanger N’2 is applied,

\[ P_e = (1 - \alpha) \cdot Q_r \cdot C_p \cdot (T_{D,a} - T_{h,\text{in}2}) = Q_r \cdot C_p \cdot (T_{\text{co}2} - T_{\text{fin}1}) \]  
(29-bis)

Introducing the known values, \( P_e \), \( C_p \), and the inlet temperatures in Eq. (29) or (29-bis), the outlet temperatures (\( T_{\text{co}1} \) and \( T_{\text{co}2} \)) can be obtained. \( T_{\text{co}} \) is then applied in Eq. (21).

References


