Desalination and Water Treatment
www.deswater.com
1944-3994 / 1944-3986 © 2011 Desalination Publications. All rights reserved.
doi: 10.5004/dwt.2011.2373

31 (2011) 144–150
July

Researching and modelling the dependence of MD flux on membrane dimension for scale-up purpose

Jianhua Zhang[a], Jun-De Li[b], Stephen Gray[a]*

[a]*Institute of Sustainability and Innovation, Victoria University, PO Box 14428, Melbourne, Victoria 8001, Australia*
Tel. +61 (3) 99198097; Fax +61 (3) 99197696; email: stephen.gray@vu.edu.au
[b]*School of Engineering and Science, Victoria University, PO Box 14428, Melbourne, Victoria 8001*

Received 3 August 2010: Accepted in revised form 23 March 2011

ABSTRACT

Membrane distillation has great advantages in treating high-concentration brine water, and desalinating salty water in areas where low grade heat is available. In this paper, we develop a mathematical model to predict the effect of process design (eg. temperature, flowrates, module length) on flux for direct contact membrane distillation (DCMD). As with traditional engineering practice, basic data such as mass transfer coefficients and heat transfer coefficients are obtained directly from experimental results. Results for both co-current and counter flow DCMD are presented and comparisons are made to verify these models using experimental results conducted at different temperatures.

Keywords: Modelling; Direct contact membrane distillation

1. Introduction

Membrane distillation (MD) is a membrane-based separation process, for which the driving force of separation is a vapour pressure difference across the membrane. The vapour pressure difference arises because of either a temperature difference across the membrane or a reduced vapour pressure on one side of the membrane. Waste heat as low as 40°C may be used to drive the process.

Smolders and Franken [1] have stated that the MD process should have the characteristics listed below:

- The membrane should be porous
- The membrane should not be wetted by the process liquids,
- No capillary condensation should take place inside the pores of the membranes,
- Only vapour should be transported through the pores of the membrane,
- The membrane must not alter the vapour equilibrium of different components in the process liquids, and
- For each component, the driving force of the membrane operation is a partial pressure gradient in the vapour phase.

Fig. 1 illustrates four configurations of MD system based on the nature of the permeate side of the membrane [2]:

(A)Direct contact membrane distillation (DCMD), in which the membrane is in direct contact with liquid phases only. This is the simplest configuration. It is best suited for applications such as desalination and concentration of aqueous solutions [3–5]

(B) Air gap membrane distillation (AGMD), in which an air gap is interposed between the membrane and a condensation surface (cooling plate). AGMD is the most general configuration and is used when energy efficiency is a key criteria [6]

* Corresponding author.

Presented at EuroMed 2010 — Desalination for Clean Water and Energy: Cooperation among Mediterranean Countries of Europe and MENA Region, 3–7 October 2010, David Intercontinental Hotel, Tel Aviv, Israel. Organized by the European Desalination Society.

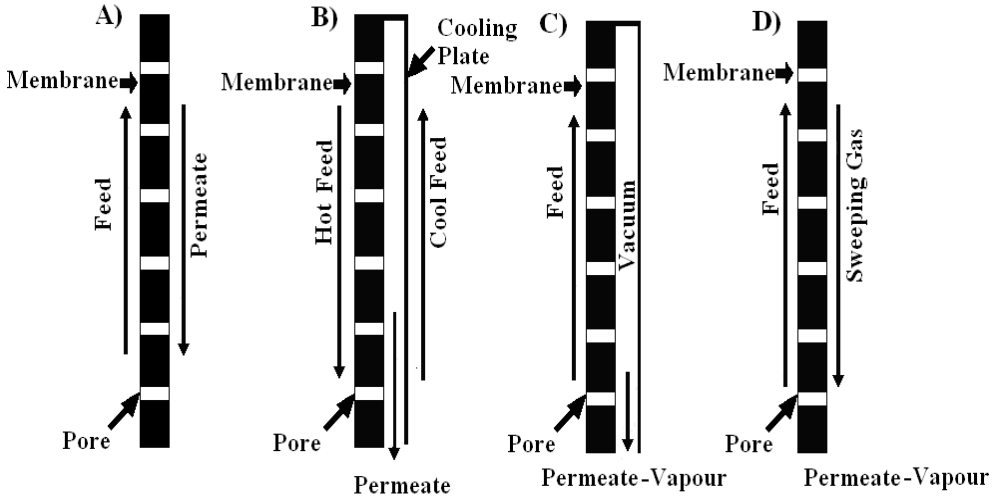


Fig. 1. Different MD configurations.

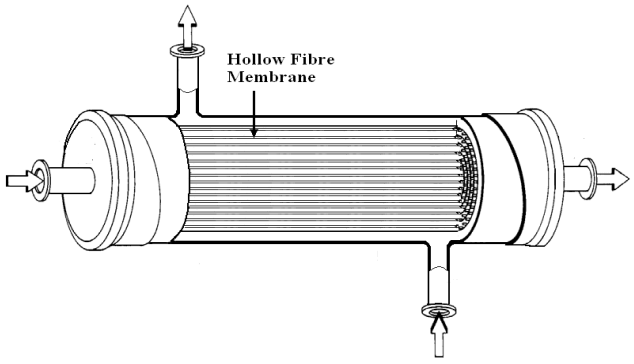
(C) Vacuum membrane distillation (VMD), in which the vapour phase is transported across the membrane under reduced pressure and condensed, if needed, in a separate device. It is suitable for application in removing volatiles from aqueous solutions [7,8], and
(D) Sweeping gas membrane distillation (SGMD), in which a stripping gas is used as a carrier for the produced vapour. It is mainly employed in removing volatiles from aqueous solutions [9–11].

1.1. Configurations of MD modules

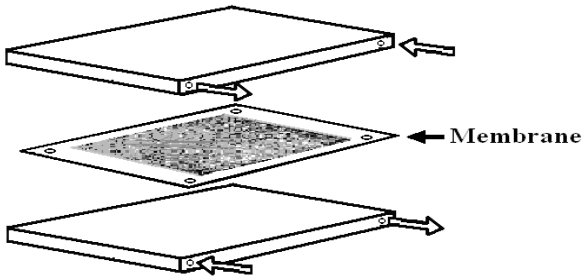
There are two types of MD module configurations shown in Fig. 2; a tubular module and a flat-sheet module. Both have been widely studied and employed in pilot plants. The flat-sheet module is versatile and can be easily set up in laboratory, while the tubular module has large specific area and is more attractive to commercial use [12].

1.2. Advantages and existing problems of MD

In comparison with other desalination methods, MD has distinct advantages such as: a nearly complete rejection of non-volatile components, a low operating pressure that is not related to feed concentration as is the case for reverse osmosis (RO), a simple structure and operation, much larger pores than other desalination membranes that are not as sensitive as to fouling, a small vapour space, and low operating temperatures (40–80°C) [13]. These characteristics make it a promising technique for desalination where low grade heat is available, such as in industrial sites. Furthermore, high salinity wastewater, such as the concentrate of RO processes, is difficult to treat by RO because of their high osmotic pressure. MD could be used after RO to recover additional water from the concentrate streams.



(a) Tubular module for hollow fibre



(b) Plate and frame module for flat sheet membrane

Fig. 2. Two different types of MD modules.

MD was introduced in the late 1960s [14,15], but it was not commercially applied at that time because membranes with adequate characteristics were not available and the economics of using MD were not attractive [16]. With the far-from-optimal membrane and system, the temperature polarization coefficient was estimated by Schofield et al. to be 0.32 [17]. In the 1980s, as new membranes were developed and a better understanding of MD was achieved,

research into MD boomed again [18–20]. However, MD for commercial applications is still in its early stages and further work is needed to develop a commercial system. Also many results from MD experiments were obtained using microfiltration membranes, which are poor materials for use in MD processes and are therefore unlikely to yield results of interest for commercial processes.

MD modelling has mainly focused on the overall mass transfer and heat transfer processes related to membrane properties, i.e. the porosity, pore size, etc. [6,21,22], which are important for membrane design but are less so for process design. Although a model based on engineering measurable parameters is very important for scaling-up MD, there are very few articles in literature focused on this.

The aims of this paper are to develop mathematical models to predict the effect of various process parameters on fluxes for direct contact membrane distillation (DCMD) and to verify these models using experimental results, and to provide guidance in scaling up MD processes.

2. Simulation and experiment

2.1. Theoretical analysis of one-dimension model for DCMD

For a given MD system, it can be expected that the flux (J) depends on many parameters and can be generally written as

$$J = f(L, Q_H, Q_C, T_H, T_C, \alpha, U) \tag{1}$$

or

$$J = \alpha \Delta P \tag{2}$$

Here, L is the length of the exposed membrane surface in the module; Q_C and Q_H are the mass flow rates for the cold side and hot side, respectively; T_C and T_H are temperatures at the cold side and hot side, respectively; α is the mass transfer coefficient of permeate across the membrane; U is the heat transfer coefficient; and ΔP is the vapour pressure difference between the feed and permeate.

Assumptions in developing the models include:

- no heat loss through the module to the atmosphere,
- specific heat of vaporisation and condensation does not change with concentration,
- with a given membrane at a given flowrate, both α and U are constant,
- in balancing the heat transfer, the sensible heat carried by the permeate can be neglected
- there is no temperature gradient across the membrane perpendicular to the flow direction
- in balancing the mass transfer, the mass of the permeate can be neglected

Fig. 3 shows the energy transfer process in a co-current DCMD in a flat sheet module.

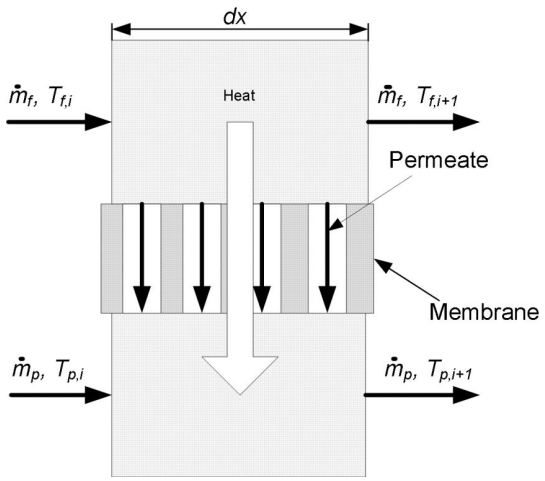


Fig. 3. Schematic diagram for analysing the balance of heat transfer.

The temperature change in the hot side can be derived as

$$\begin{aligned}
C_v dm dT_H &= -\left(H_v J dA dt + U dA (T_H - T_C) dt\right) \\
C_v \rho H d A d T_H &= -\left(H_v J d A d t + U d A (T_H - T_C) dt\right) \\
C_v \rho H d T_H &= -\left(H_v J dt + U (T_H - T_C) dt\right) \\
\frac{C_v \rho H W V_H d T_H}{W} &= -\left(H_v J dt + U (T_H - T_C) dt\right) V_H \\
\frac{C_v Q_H d T_H}{W} &= -\left(H_v J + U (T_H - T_C)\right) dx \\
d T_H &= -\frac{W\left(H_v J + U (T_H - T_C)\right) dx}{C_v Q_H}
\end{aligned} \tag{3}$$

Similarly, the temperature change on the permeate (cold) side can be written as

$$d T_C = \frac{W\left(H_v J + U (T_H - T_C)\right) dx}{C_v Q_c} \tag{4}$$

From Eqs. (3)–(4), it can be obtained

$$d T_H / d T_C = Q_C / Q_H \tag{5}$$

Because

$$J = \alpha (P_H - P_C) \tag{6}$$

thus,

$$d T_H = -\frac{W\left(H_v \alpha (P_H - P_C) + U (T_H - T_C)\right) dx}{C_v Q_H} \tag{7}$$

Here, C_v is the specific heat of water, H_v is the vaporisation latent heat of water, W and H are the width and height of the module respectively, t is the time, ρ is the density of water, V_H is the linear velocity of the hot water, and P_H and P_C are vapour pressures at T_H and T_C, respectively,

which can be calculated by the Antoine equation (given in Eq. (8) later).

$$P = \exp\left(23.1964 - 3816.44/(T + 227.02)\right) \quad (8)$$

The above equations can be solved numerically. For co-current flows, the numerical procedure is shown in Fig. 4.

For counter flows, an iteration process is required and the process is shown in Fig. 5.

2.2. Experimental apparatus and procedure

A schematic diagram of the DCMD process employed in this experiment is shown in Fig. 6. Flowrates of the feed and permeate were controlled with two peristaltic pumps. A heater and a chiller were used to set temperatures of the feed and permeate.

The membrane has nominal pore size of 1 μm, contact angle of 126º±5, an effective area of 0.014 m² and an effective length of 0.0145 m. Two spacers were arranged in the module on both sides of the membrane to reduce polarisations of concentration and temperature.

Four thermocouples were arranged at the inlets and outlets of the permeate and feed to measure their temperatures. A balance was used to measure the weight of the permeate. A conductivity meter was installed in the permeate stream to monitor the salt rejection of membrane. The feed solution was made by dissolving 10 g of sodium chloride in 1 litre of deionised water. Deionised water was used on the permeate side as coolant.

The initial feed temperature and permeate temperature were 60°C and 20°C, respectively. The flowrates of both the hot feed and permeate were kept equal and varied in range of 300–700 mL/min.

The thermal transfer coefficients and mass transfer coefficients used in modelling were calculated from the experimental results using Eq. (1) and Eq. (2).

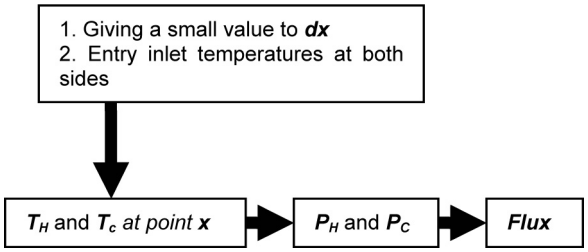


Fig. 4. Schematic diagram for numerical flux calculation of co-current flow.

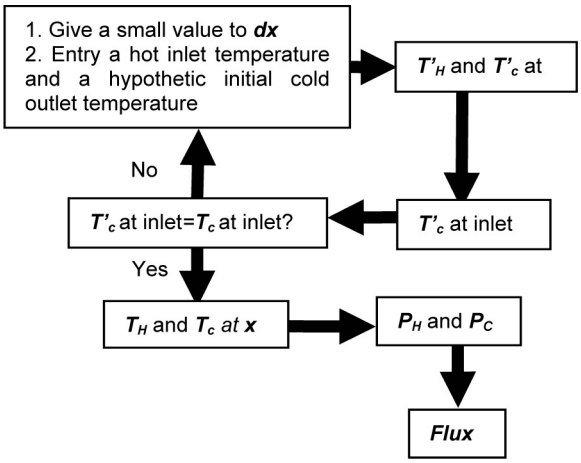


Fig. 5. Schematic diagram for numerical flux calculation of counter flow.

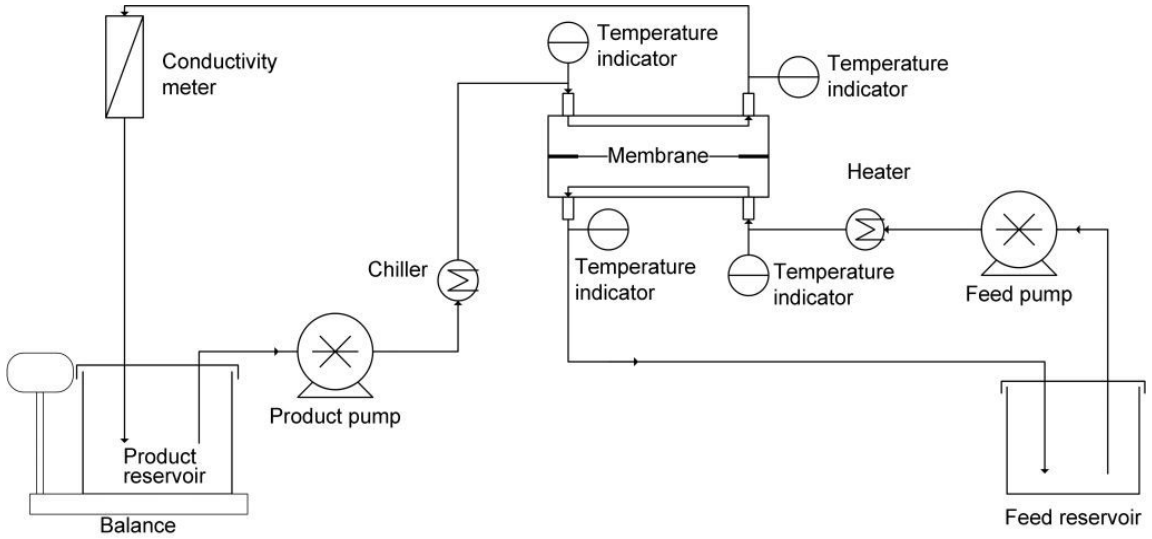


Fig. 6. A Schematic diagram of the DCMD process.

3. Results and discussion

3.1. Model parameters and comparison at different temperatures

The model was used to calculate the heat and mass transfer coefficients from counter flow DCMD experiments performed at different flowrates. Fig. 7 shows the experimental data and the calculated flux using the heat and mass transfer coefficients derived from each data point. The errors between the model and the experimental results are small, being in the range of 2.6–6.4%. Such agreement is anticipated given the heat and mass transfer coefficients were obtained from the same data.

To verify the model, the heat and mass transfer coefficients derived from the data shown in Fig. 7 were used to predict the flux at different temperatures. Fig. 8 shows predicted and experimentally derived flux results for counter flow DCMD conducted at different hot feed temperatures and a fixed flowrate of 600 mL/min. The largest errors were no more than 15% and appeared at 40°C and 80°C respectively, indicating there is reasonable agreement between the model and experimental results for the hot feed temperatures of interest. This provides some confidence that the temperature profiles predicted from the model will be accurate.

However, variations in the flowrate will produce different turbulence regimes and are expected to significantly alter the mass transfer and heat transfer coefficients. Therefore, the flowrate specific heat and mass transfer coefficients derived from the data in Fig. 7 were used in subsequent predictions.

3.2. Mathematical modelling of membrane distillation

3.2.1. Co-current flow modelling results

Fig. 9 shows the modelling predictions for the temperature distribution along the length of the module for co-current flow membrane distillation. These results predict greater temperature differences as the flowrates increase, because the increased flowrates reduce the residence time of the permeate and the feed in the module. Hence, at the higher flowrates, the turbulence, mean temperature differences are increased, which leads to higher fluxes as predicted and observed in Fig. 7.

Fig. 10 shows the predicted relationship between flux and membrane length. The fluxes increase with flowrate, but the flux differences become smaller at higher flowrates. A similar asymptotic trend of permeate flux with increasing feed flow rates has been reported previously [11,23], and this provides further support for the reliability of the model predictions. The model also predicts that the fluxes reduce as the module becomes longer. The reason can be found in Fig. 9, where the mean temperature difference drops as the length of the membrane increases.

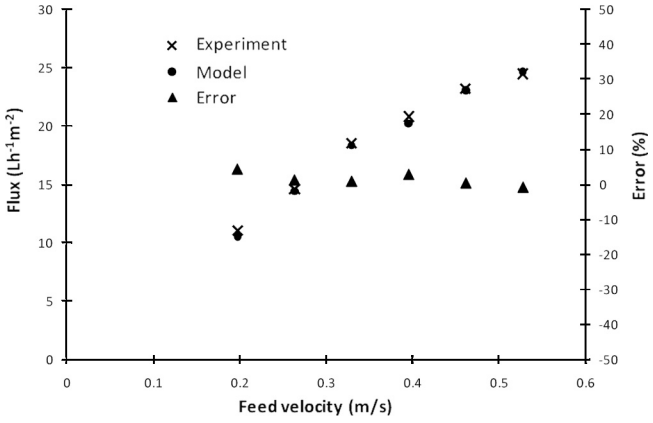


Fig. 7. Comparison between experimental and modelling results at different flowrates for counter-current.

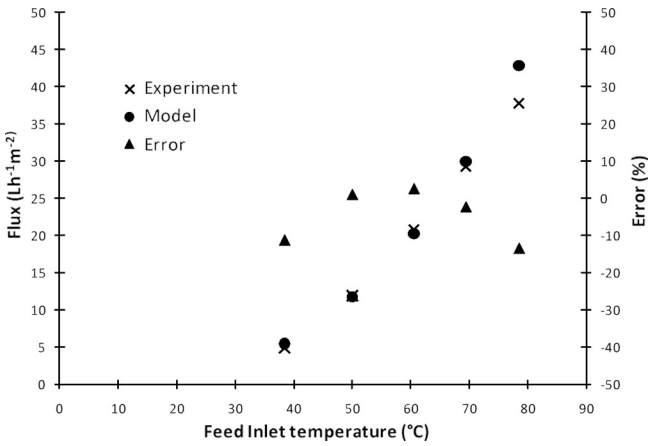


Fig. 8. Comparison between experimental and modelling results at different temperatures for counter-current.

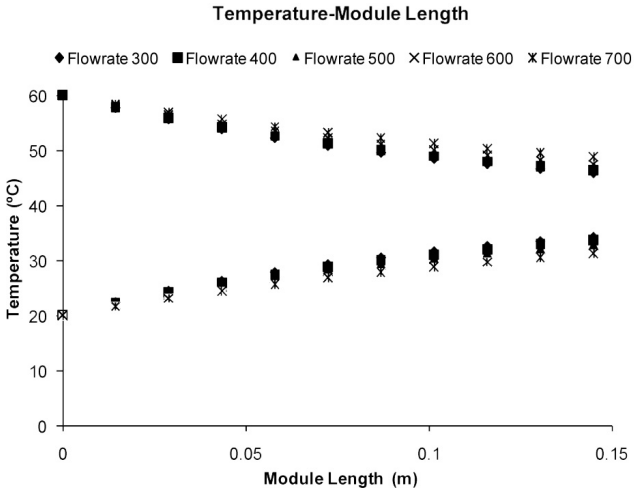


Fig. 9. Temperature distributions in membrane length direction for co-current flows.

3.2.2. Counter flow modelling results

The temperature distributions along the membrane are shown in Fig. 11. Similar to the co-current model, the temperature differences are increased at faster flowrates, because of the reduced residence time in the module. At a certain flowrate, the temperature distribution profiles of permeate side and feed side are parallel to each other, which is different from the curves of the co-current flow model in which the curves approach each other.

The modelled relationships between flux and membrane length are shown in Fig. 12. A similar trend to the co-current model is predicted, except the decrease in flux with increasing length (ie. the slope) is smaller than those of the co-current model.

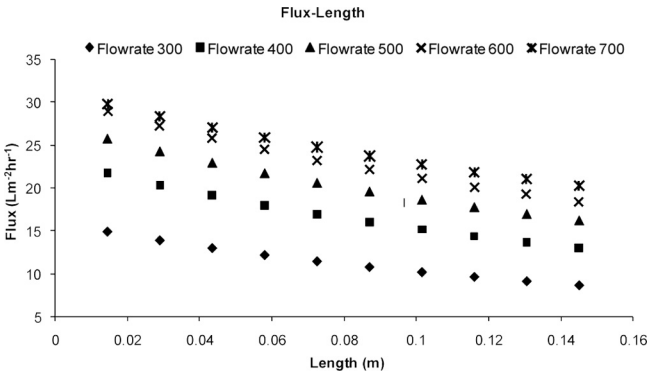


Fig. 10. Relation between flux and membrane length for co-current flows.

4. Conclusions

A preliminary mathematical model for MD process in desalination based on heat and mass balance has been developed to predict process performance with variations in parameters such as temperature and module length. The model predictions also showed reasonable agreement with experimental results, when experimentally derived heat and mass transfer coefficients were used.

This work also suggests that reporting of heat and mass transfer coefficients for characterisation of MD membranes rather than flux, would provide more generalised characterisation of the membranes, and allow easier comparison membrane performance.

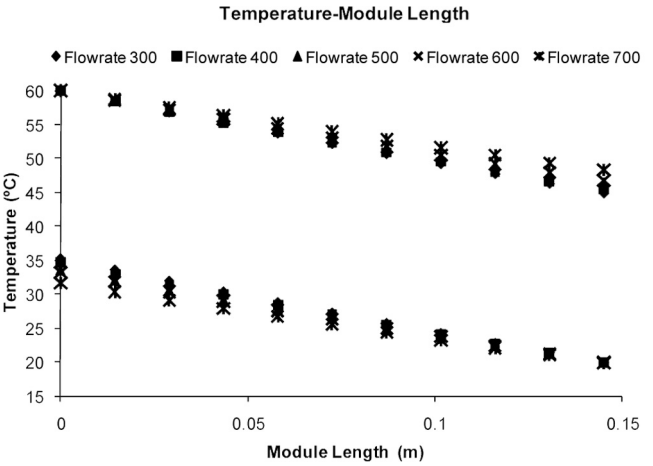


Fig. 11. Temperature distributions in membrane length direction for counter flow.

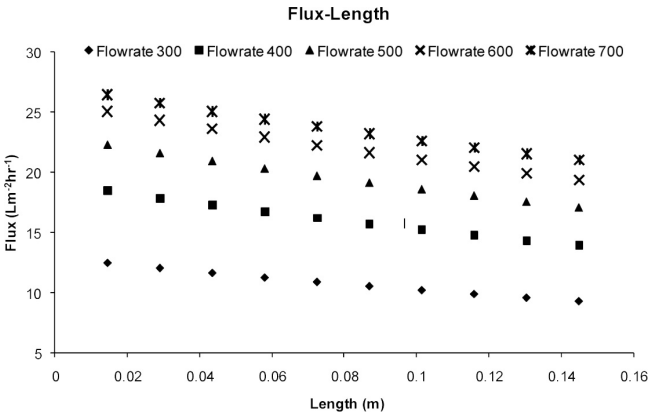


Fig. 12. Relation between flux and membrane length for counter flow.

References

[1] K. Smolders and A.C.M. Franken, Terminology for membrane distillation, Desalination, 72(1989) 249–262.
[2] A.M. Alklaibi and N. Lior, Membrane-distillation desalination: Status and potential, Desalination, 171 (2005) 111–131.
[3] L. Martinez-Diez and F.J. Florido-Diaz, Theoretical and experimental studies on desalination using membrane distillation, Desalination, 139 (2001) 373–379.
[4] K.W. Lawson and D.R. Lloyd, Membrane distillation, J. Membr. Sci., 124 (1997) 1–25.
[5] J. Phattaranawik and R. Jiraratananon, Direct contact membrane distillation: effect of mass transfer on heat transfer, J. Membr. Sci., 188 (2001) 137–143.
[6] A.S. Josson, R. Wimmerstedt and A.C. Harrysson, Membrane distillation — a theoretical study of evaporation through microporous membranes, Desalination, 56 (1985) 237–249.
[7] S. Bandini, C. Gostoli and G.C. Sarti, Separation efficiency in vacuum membrane distillation, J. Membr. Sci., 73 (1992) 217–229.
[8] G.C. Sarti, C. Gostoli and S. Bandini, Extraction of organic components from aqueous streams by vacuum membrane distillation, J. Membr. Sci., 80 (1993) 21–33.
[9] C.A. Rivier, M.C. Garcia-Payo, I.W. Marison and U. von Stockar, Separation of binary mixtures by thermostatic sweeping gas membrane distillation: I. Theory and simulations, J. Membr. Sci., 201 (2002) 1–16.
[10] L. Basini, G. D'Angelo, M. Gobbi, G.C. Sarti and C. Gostoli, A desalination process through sweeping gas membrane distillation, Desalination, 64 (1987) 245–257.

[11] M.C. Garcia-Payo, C.A. Rivier, I.W. Marison and U. von Stockar, Separation of binary mixtures by thermostatic sweeping gas membrane distillation: II. Experimental results with aqueous formic acid solutions, J. Membr. Sci., 198 (2002) 197–210.
[12] E. Curcio and E. Drioli, Membrane distillation and related operations: A review, Separ. Purif. Rev., 34 (2005) 35–86.
[13] M. Tomaszewska, Membrane distillation — Examples of applications in technology and environmental protection, Environ. Studies, 9 (2000) 27–36.
[14] P.K. Weyl, Recovery of Demineralized Water from Saline Waters, US Patent 3340186, 1967.
[15] M.E. Findley, Vaporization through porous membranes, Ind. Eng. Chem. Proc. Des. Dev., 6 (1967) 226–230.
[16] W.T. Hanbury and T. Hodgkiess, Membrane distillation — an assessment, Desalination, 56 (1985) 287–297.
[17] R.W. Schofield, A.G. Fane and C.J.D. Fell, Heat and mass transfer in membrane distillation, J. Membr. Sci., 33 (1987) 299–313.
[18] L. Carlsson, The new generation in sea water desalination SU membrane distillation system, Desalination, 45 (1983) 221–222.
[19] S.I. Andersson, N. Kjellander and B. Rodesj, Design and field tests of a new membrane distillation desalination process, Desalination, 56 (1985) 345–354.
[20] T.J.v.G. Klaus Schneider, Membrandestillation, Chem. Ing. Technik, 56 (1984) 514–521.
[21] J. Phattaranawik, R. Jiraratananon and A.G. Fane, Effect of pore size distribution and air flux on mass transport in direct contact membrane distillation, J. Membr. Sci., 215 (2003) 75–85.
[22] J. Phattaranawik, R. Jiraratananon and A.G. Fane, Heat transport and membrane distillation coefficients in direct contact membrane distillation, J. Membr. Sci., 212 (2003) 177–193.
[23] M. Matheswaran, T.O. Kwon, J.W. Kim and I. Moon, Factors affecting flux and water separation performance in air gap membrane distillation, J. Ind. Eng. Chem., 13 (2007) 965–970.