Experimental and Theoretical Approach of a Multi-Stage Membrane Distillation System

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Abstract
A multi-stage membrane distillation unit was designed and tested in various operating conditions such as the feed flow rate, the water flow rate and the temperature at the hot and cold stream respectively and the applied vacuum pressure. The development of a mathematical model assisted in the effective design of the unit and validating the simulation results of the model.

Keywords
Multi-stage membrane distillation; experimental setup; mathematical model

INTRODUCTION

The rapid growth of population, particularly in the developed countries, the desire to improve the living standards and the concern for the environment (protection of natural water resources) have resulted, in the last years, in the increasing demand and supply of sufficient quantity and good quality drinking water (Khawaji, 2008; Sharon, 2015). The common thermal desalination technologies (multi-stage flash distillation, multiple stage evaporation, vapor compression) and the separation processes (distillation, reverse osmosis), even if they are commercially available, they present significant energy consumption and high costs. Membrane distillation (MD) is a relatively new and promising technology for the desalination of brackish and seawater (Lawson, 1997).

Membrane Distillation (MD) is a thermal membrane separation process that involves the transport of water vapor molecules from a hot aqueous solution through a microporous hydrophobic membrane, due to the vapor pressure difference created by the temperature difference between the two sides of the membrane. This is an attractive alternative solution to the conventional desalination techniques, due to the production of high quality distillate and the possibility of operating at low temperatures and pressures. The capability of utilizing solar thermal energy or waste heat from other processes, makes MD an energy and cost efficient and environmental friendly process (Al-Obaidani, 2008).

Although the MD technology is known for several decades, its exploitation in practice remained very limited until recently, where there has been an renaissance of research interest for applications in desalination, mainly due to developments in the field of materials and in particular the membranes (Drioli, 2014). The investigation presented is a part of a wider research action with aim to elaborate technologically reliable and economically viable desalination MD solutions. The specific objective is the design of a pilot plant with the possibility of integration of innovative membranes and energy design optimization solutions, in terms of improving water recovery rate, gained output ratio (GOR) and the performance ratio (PR).

The effective design of a desalination plant requires, except the research experience from previous installations, the experimental investigation of the effect of the operating parameters of the system to various conditions of use and the development of an appropriate mathematical model with aim of maximizing the productivity and the energy optimization of the process. Furthermore, the experimental results of the desalination unit will be used to validate the theoretical model.

MATHEMATICAL MODEL DEVELOPMENT

For the thermohydraulic optimization of multi-stage membrane distillation systems, an one dimensional (1-D) mathematical model was developed for an air gap membrane distillation system (AGMD) with a flat sheet membrane. The model is based on mass and energy balances and combines the vacuum membrane distillation (VMD) and the multi-stage distillation concept. The system was treated as three different sections, which are related to each other through simultaneous mass and heat transfer. As illustrated in Fig. 1, the sections, in order from left to right, are: i) the evaporator, ii) a stage, and iii) the condenser.

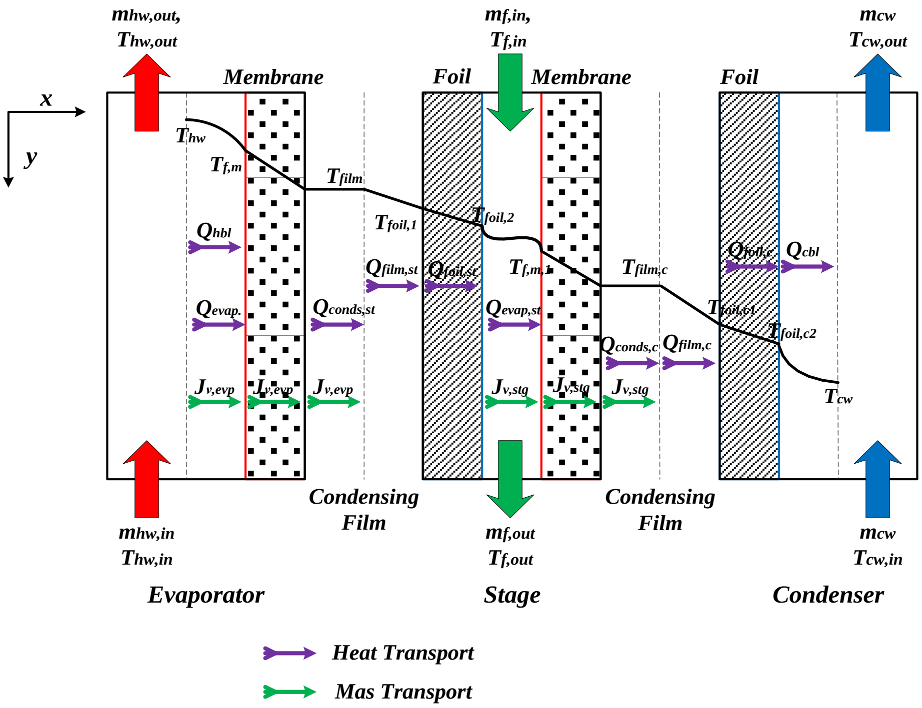

Figure 1. Schematic illustration of mass and heat transfer at a multi-stage system.

For simulation with reasonable computational expense, the model was simplified according to the following assumptions: i) the system is operating under steady state conditions, ii) there is no heat loss to the ambient environment, iii) the hot water stream and the cold water stream flow in the x-direction only, iv) there are no mass and heat effects in the z-direction, v) the boundary layers in each stream are fully hydrodynamically and thermally developed, vi) there is a linear temperature distribution across the membrane, vii) only water vapor molecules are transported through the membrane pores, viii) constant water vapor temperature at the vapor-gap region (space between the membrane and the condensing film surface), ix) the concentration polarization affects (negatively) the process, x) the physicochemical properties (viscosity, density, etc.) of the fluid (fresh water and saline water) in each stream are depended on the temperature and the salinity.

The mass balances at the evaporator and the stage can be described as follows:

Evaporator

- For the hot water stream: $m_{hw,in} = m_{hw,out} + J_{v,evp} dA_{evp}$ (1)

Stage

- For the feed saline solution: $m_{f,in}^{sw} = m_{f,out}^{sw} + J_{v,stg} dA_{stg}$ (2)

- For the salt: $$m_{f,in}^{sw} x_{f,in}^{salt} = m_{f,out}^{sw} x_{f,out}^{salt} \quad (3)$$

where $m_{hw,in}$, $m_{hw,out}$, $m_{f,in}^{sw}$, $m_{f,out}^{sw}$ (kg/sec) are the mass flow rates of the hot water stream and the feed saline water at the entrance and the exit of the evaporator and the stage, $J_{v,evp}$, $J_{v,stg}$ (kg/m^2.sec) is the water vapor mass flux that passes through the membrane at the evaporator and the stage, A_{evp}, A_{stg} (m^2) is the membrane surface area at the evaporator and the stage and $x_{f,in}^{salt}$, $x_{f,out}^{salt}$ are the salt mass fractions at the entrance and the exit of the stage respectively.

The energy balances at each section can be written as:

Evaporator: $$Q_{hw} dA_{evp} = m_{hw,in} C_{p,hw} T_{hw,in} - m_{hw,out} C_{p,hw} T_{hw,out} \quad (4)$$

Stage: $$Q_{foil,st} dA_{stg} = m_{f,out}^{sw} C_{p,sw} T_{f,out} - m_{f,in}^{sw} C_{p,sw} T_{f,in} + Q_{evap,\mathrm{st}} dA_{stg} \quad (5)$$

Condenser: $$Q_{cw} dA_{cond} = m_{cw} C_{p,cw} \left(T_{cw,out} - T_{cw,in}\right) \quad (6)$$

where Q_{hw}, $Q_{foil,st}$, Q_{cw} (W/m^2) is the heat flux at the hot water stream, at the foil of the stage and at the cold water stream, $Q_{evap,st}$ (W/m^2) is the latent heat of evaporation at the membrane surface at the stage, $C_{p,hw}$, $C_{p,sw}$, $C_{p,cw}$ (J/kg.K) is the specific heat capacity of the hot water stream, the saline water and the cold water stream, A_{cond} (m^2) is the area of condenser, m_{cw} (kg/sec) is the mass flow rate of the cold water stream and $T_{hw,in}$, $T_{hw,out}$, $T_{f,in}$, $T_{f,out}$, $T_{cw,in}$, $T_{cw,out}$ (K) are the temperatures of the hot water stream, the feed saline water and the cold water stream at the entrance and the exit of each section correspondingly.

Mass Transfer

In the MD desalination system, the mass transfer occurs in the feed boundary layer in the stage (concentration polarization) and across the porous membrane.

With the presence of NaCl in the feed solution, an additional resistance is formed adjacent to the membrane surface. A mass balance across the feed side boundary layer, described by the film theory, yields a relationship between the water vapor mass flux, J_v, the solute mass transfer coefficient, $k_{f,s}$, and the solute mole fraction, x_f^{salt} (Martinize-Diez, 1998):

$$x_{f,mem}^{salt} = x_f^{salt} exp\left(\frac{J_v}{\rho_f k_{f,s}}\right) \quad (7)$$

where $x_{f,mem}^{salt}$ is the solute mole fraction at the feed – membrane interface, ρ_f (kg/m^3) is the density of the feed saline solution. The solute mass transfer coefficient, $k_{f,s}$ (m/sec), can be estimated from Sherwood number (Sh) via existing mass transfer correlations (Khayet, 2011).

The mass transfer through the gas-phase membrane pores is typically described by the Darcy's law, whereby the water vapor mass flux is proportional to the partial pressure difference of water vapor across the membrane:

$$J_v = C_m \Delta P_i = C_m \left(P_{f,m} - P_{film}\right) \quad (8)$$

where J_v (kg/m^2.sec) is the water vapor mass flux that permeates the membrane, C_m (kg/m^2.sec.Pa) is the membrane mass transfer coefficient and $P_{f,m}$, P_{film} (Pa) are the partial pressures of water vapor at the feed – membrane interface and at the condensing film surface respectively.

The membrane mass transfer coefficient, C_m, is primarily a function of membrane structural properties (porosity, tortuosity, thickness and pore size), the physical and chemical properties of the vapour passing through the membrane (molecular weight, diffusivity) and the operating conditions (temperature and pressure). The C_m depends on the dominating mass transport mechanism within the membrane pores. The controlling diffusion mechanism in the membrane is related with the collisions between diffusing molecules and/or with the molecules – pore wall collisions and can be determined by the ratio the ratio of the mean free path (λ_i) of the transported molecule to the pore size (d_p) of the membrane (Knudsen number). The mean free path of water vapor ranges between 0.28 μm and 0.49 μm in the temperature range of 85 – 65 ^oC, close to the nominal membrane pore size of the current system, d_p ~ 0.2 μm. Also, low vacuum pressure is applied to the system. Therefore, it is assumed that the controlling mass transport mechanism within the membrane is a combination of Knudsen diffusion – Poiseuille flow. The membrane mass transfer coefficient is defined as:

$$C_m = \left(\frac{2}{3}\frac{\varepsilon r_p}{\tau\delta_m}\left(\frac{8M_i}{\pi RT_m}\right)^{\frac{1}{2}}\right) + \left(\frac{1}{8\mu_i}\frac{\varepsilon r_p^2}{\tau\delta_m}\frac{M_iP_m}{RT_m}\right) \qquad (9)$$

where $\varepsilon, r_p, \tau, \delta_m$, is the porosity, the pore size (m), the tortuosity, and the thickness (m) of the membrane and M_i is the permeating specie molecular weight (kg/mol), T_m, P_m is the average temperature and pressure at the membrane.

Heat Transfer

Evaporator

The evaporator is built from membranes and is optimized to create steam from fresh water as driving thermal energy source for the process. The heated fresh water is evaporated at the feed – membrane interface and the produced water vapor passes through the membrane and flows to the first stage, where it is condensed. The total heat is transferred from the hot water stream through the hot water boundary layer to the feed – membrane interface by convection (Eq.10). The transferred heat at the membrane surface is consumed only by the latent heat of evaporation (Q_{evap}), assuming that the heat conduction through the membrane can be neglected (Eq.11).

$$Q_{hbl} = h_{hbl}(T_{hw} - T_{f,m}) \qquad (10)$$

$$Q_{evap} = J_{v,\mathrm{evp}}\Delta H_v \qquad (11)$$

where Q_{hbl} (W/m^2) is the heat flux across the hot water boundary layer, h_{hbl} $(W/m^2.K)$ is the heat transfer coefficient at the hot water boundary layer (Khayet, 2011), T_{hw} and $T_{f,m}$ (K) is the average temperature at the hot water stream and the temperature at the feed – membrane interface respectively and ΔH_v (J/kg) is the latent heat of evaporation.

Stage

A stage is comprised from alternating foils and membranes, starting with a foil. The generated water vapor from one stage (or the evaporator) flows through a channel to the next stage and is completely condensed at the cold surface of the impermeable foil (Eq.12). As pure water vapor is condensed on the foil, a condensing film is formed at the top of the foil and flows downward under the influence of gravity. The latent heat of evaporation in the form of condensation is transferred through the condensing film and the foil by conduction (Eq.13, 14) and it is transmitted to the feed channel, where it heats up the feed saline stream.

$$Q_{evap} = Q_{conds,st} = J_{v,evp}\Delta H_v \qquad (12)$$

$$Q_{film,st} = \frac{k_{film}}{\delta_{film}}\left(T_{film} - T_{foil,1}\right) \quad , \qquad (13)$$

$$Q_{foil,st} = \frac{k_{foil}}{\delta_{foil}}\left(T_{foil,1} - T_{foil,2}\right) \qquad (14)$$

where k_{film} (W/m.K), δ_{film} (m), k_{foil} (W/m.K), δ_{foil} (m) is the thermal conductivity and thickness of the condensing film (Incropera, 2007) and the foil and T_{film}, $T_{foil,1}$, $T_{foil,2}$ (K) are temperatures at the surface of the condensing film, at the vapor-gap – foil interface and at the foil – feed saline stream interface respectively.

The conducted heat through the foil is transferred to the feed channel by convection at the saturation temperature, assuming that the heating of the feed saline solution is fast, since the feed channel width is very small and the thermal boundary layer at the foil – feed saline solution and feed – membrane interface is fully developed.

$$Q_{fconv} = h_f(T_{foil,2} - T_{sat}) \tag{15}$$

where $T_{sat}(= T_{f,m,1} + BPE)$ is the saturation temperature of the hot feed saline solution.

In the feed channel the feed saline solution is initially pre-heated from the ambient temperature to its boiling point, T_{sat}, and then is partially evaporated at the membrane surface (instead of Equation 5).

$$Q_{foil,st}dA_{eff} = m_{f,in}^{sw}C_{p,sw}\left(T_{sat} - T_{f,in}\right) + Q_{evap,st}dA_{eff} \tag{16}$$

At the feed – membrane interface the water is partially evaporated and new water vapor is produced. The heat is transferred from the feed channel to the feed – membrane interface of the stage and it is only consumed by the latent heat of vaporization ($Q_{evap,\mathrm{st}}$, W/m^2), assuming the heat conduction through the membrane can be neglected.

$$Q_{evap,\mathrm{st}} = J_{v,stg}\Delta H_v \tag{17}$$

Condenser

The condenser is consisted of foils, where the produced water vapor from the last stage is completely condensed (Equation 18). The latent heat of condensation is transferred through the condensing film and the foil by conduction (Eq. 19, 20). In the boundary layer of the cold water stream the heat is transported by convection (Eq. 21).

$$Q_{evap,st} = Q_{conds,c} = J_{v,stg}\Delta H_v \tag{18}$$

$$Q_{film,c} = \frac{k_{film}}{\delta_{film,c}}\left(T_{film,c} - T_{foil,c1}\right) \tag{19}$$

$$Q_{foil,c} = \frac{k_{foil}}{\delta_{foil}}\left(T_{foil,c1} - T_{foil,c2}\right) \tag{20}$$

$$Q_{cbl} = h_{cbl}\left(T_{foil,c2} - T_{cw}\right) \tag{21}$$

where $\delta_{film,c}$ (m) is the condensing film thickness at the foil of the condenser, h_{cbl} (W/m^2.K) is the heat transfer coefficient at the cold water boundary layer and $T_{film,c}$, $T_{foil,c1}$, $T_{foil,c2}$ (K) are the temperatures at the condensing film surface, at the vapor-gap – foil interface, at the foil – cold water stream interface and T_{cw} (K) is the average temperature at the cold water stream.

Based on the above mass and energy balances and the heat transfer relations, a set of non-linear algebraic equations comes out for the temperatures $T_{hw}, T_{f,m}$ at the evaporator, $T_{film}, T_{foil,1}$, $T_{foil,2}$, T_{sat}, $T_{f,m,1}$ at the stage and $T_{film,c}, T_{foil,c1}$, $T_{foil,c2}$, T_{cw} at the condenser. For solving these equations, algorithms were developed in Matlab v.8.2, which were solved by the iterative method.

EXPERIMENTAL SETUP

The mathematical model that was developed earlier, contributed to the effective design of an experimental multi-stage membrane distillation unit. This MD unit used to validate the theoretical model and tested to several operating parameters, such as the feed flow rate at the stage, the heating and cooling water temperature, the heating and cooling water flow rate and the level of the applied vacuum pressure. The experimental unit consists of 1) the main distillation unit (membranes, foils, steam production and condenser), 2) heating, cooling, feed, brine and distillate tanks, 3) sensors and transducers for temperature, flow rate, pressure and water electrical conductivity, 4) power and control cabinet and finally 5) the vacuum cycle. This unit promises to lower the specific thermal energy consumption (STEC) and maximize the Gain Output Ratio (GOR) mainly due to the application of multi-stage and vacuum design.

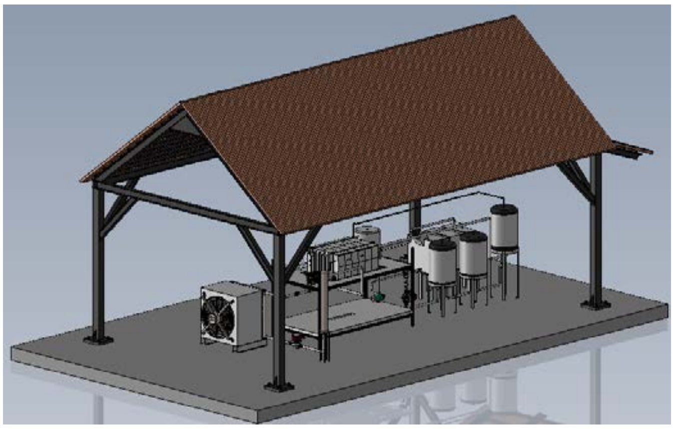

Figure 2. Schematic illustration of the experimental multi-stage MD unit.

RESULTS – DISCUSSION

In Figure 1a and 1b is illustrated the effect of feed flow rate on the distillate flux, the recovery ratio, the gained output ratio (GOR) and the specific thermal energy consumption (STEC) respectively.

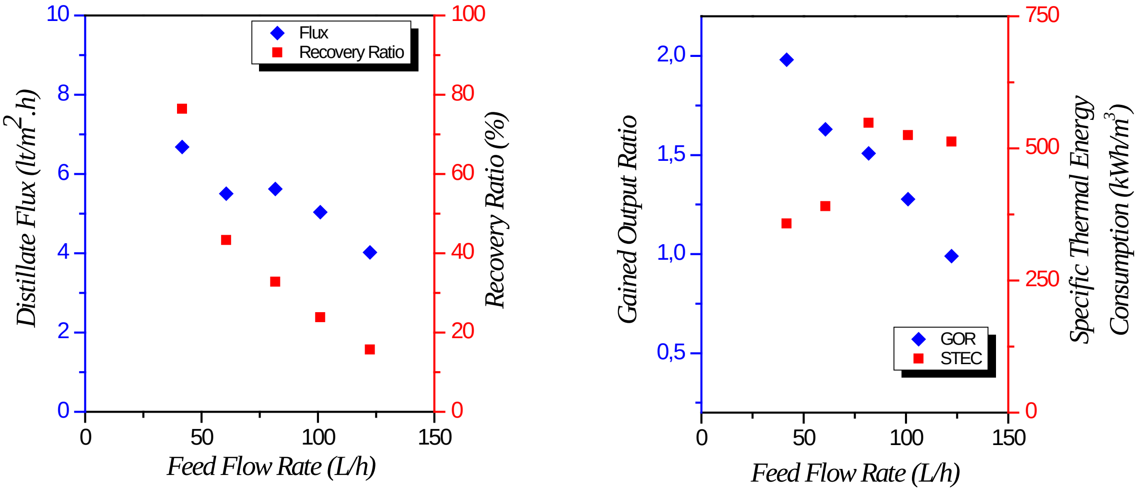

Figure 1. Feed flow rate as a function of a) water vapor flux and the recovery ratio, and b) gained output ratio (GOR) and the specific thermal energy consumption (STEC)

The vapor flux as well as the recovery ratio increases with the decrease of the feed flow rate, due to the fact that at lower flow rates the solution has the sufficient time (higher residence time) to be heated, evaporated and pass through the membrane. The STEC also decreases with the feed flow rate, whereas the GOR presents an opposite behavior and reaches a plateau at higher values of flow rate. The model predictions present low deviations, in the range of 1% to 15%, from the experimental data for pure water, whereas for saline water the deviations were in the range of 5% – 22%. This can be attributed to the inaccurate estimation of the mass and heat transfer coefficients, due to the fact the presence of the solutes in high concentration alters the fluid dynamics and influences the prediction of the mass and heat transfer.

CONCLUSIONS

An experimental multi-stage membrane distillation unit was designed and tested to several operating parameters. At low values of feed flow rate the production of pure water increases, but the thermal performance of the system decreases. The experimental data were used to validate an already developed mathematical model. This model is based on mass and energy balances, using empirical mass and heat transfer correlations. The theoretical predictions were in a good agreement with the experimental results.

REFERENCES

A.D. Khawaji, I. Kutubkhanah, J.M. Wie, (2008) Advances in seawater desalination technologies, Desalination, 221, 47 – 69.

H. Sharon, K.S. Reddy, (2015) A review of solar energy driven desalination technologies, Renew. Sustain. Energ. Rev., 41, 1080 – 1118.

K.W. Lawson, D.R. Lloyd, (1997) Membrane Distillation, J. Membr. Sci., 124, 1 – 25.

S. Al-Obaidani, E. Curcio, F. Macedonio, G. di Profio, H. Al-Hinai, E. Drioli, (2008) Potential of membrane distillation in seawater desalination: Thermal efficiency, sensitivity study and cost estimation, J. Membr. Sci., 323, 85 – 98.

Drioli, E., Ali, A., Macedonio, F., (2014) Membrane Distillation: Recent developments and perspectives, Desalination, 356, 56 – 84.

L. Martinize-Diez and M.I. Vazquez-Gonzalez, (1998) Effects of Polarization on Mass Transport through Hydrophobic Porous Membranes, Ind. Eng. Chem. Res., 37, 4128 – 4135.

M. Khayet, (2011) Membranes and theoretical modeling of membrane distillation: A review, Advances in Colloid and Interface Science, 164, 56 – 88.

F. Incropera, D. Dewitt, T. Bergman, A. Lavine, (2007) Fundamentals of heat and mass transfer, J. Wiley & Sons, Sixth Edition.

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