



Energy efficient membrane distillation through localized heating

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ABSTRACT

Membrane-based desalination technologies are crucial for generating clean water, but are energy intensive. Localizing heat generation at the liquid-membrane interface could reduce energy consumption while maintaining sufficient mass flux. Here, a locally heated membrane distillation approach is analytically and experimentally evaluated. The measured evaporated mass flux matches the analytical limits considering diffusive and advective mass transfer across a heated porous membrane into stagnant air. Experimentally, thermally-stable silver membranes demonstrate similar mass flux compared to conventional polymer based membrane at the same surface temperature. At high membrane temperatures ($\sim 80^\circ\text{C}$), the locally heated membrane distillation demonstrates good efficiency with up to 75% reduction of energy compared to direct contact membrane distillation and airgap membrane distillation. Efficiency can be further improved with better thermal design of the supporting structure and permeate heat recovery.

1. Introduction

Membrane based water purification accounts for more than 50% of installed water plant due to its simplicity and relatively low energy cost [1]. Among membrane technologies, Reverse Osmosis (RO) is the most mature technology. Nearly, 75 million liters of fresh water are produced daily using RO at a flux of $10.2\text{ L}/(\text{m}^2\text{ h})$ [2,1]. Despite those high-performance metrics, RO has its own challenges. Separation efficiency is affected by concentration polarization resulted from selective species transfer across membrane. Also, membrane fouling is a major challenge for production yield and reliability. A single stage RO configuration commonly operates in a range of 40–50% feed recovery rate having a water quality of 200–500 ppm [2–4]. Thus, researchers are striving to tackle those challenges. Among the promising alternative technologies is Membrane Distillation (MD), which uses a non-isothermal membrane for separation [5]. The temperature gradient provides a driving mechanism for fluid transport due to different vapor pressures at different temperatures. MD membranes only allow volatile vapor molecules to pass through hydrophobic pores while preventing feed fluid from penetrating through membrane [6]. As a result of the absence of liquid entrainment in the MD method, species such as ions, colloids, and macromolecules, which are unable to evaporate and diffuse across the membrane, are completely rejected.

In contrast to conventional distillation that relies on high velocity vapor to provide ultimate vapor liquid contact, MD utilizes hydrophobic microporous media to maintain vapor-liquid contact. This allows MD to operate at low temperature without compromising the

vapor-liquid contact, which permits the use of low grade waste heat such as solar heat. Additionally, membranes in MD play a relatively small role in separation: they maintain the vapor-liquid interface, but do not filter contaminations from feed. This contrasts with filtration processes where the pores are sized smaller than contaminants. Over time, for pressure driven filtration processes, the pores get clogged affecting flux. On the other hand, membranes for MD operate with larger pores sizes compared to RO, which makes the membranes less prone to clogging [7,8]. Even though fouling could affect wetting properties of MD membranes, previous studies have reported that RO flux declines significantly more than MD flux in the presence of organic foulant and low concentration organic foulant [9]. Thus, the MD technique is particularly advantageous for high contamination concentrations.

Despite the benefits, commercialization of MD technology is constrained by high energy consumption. Energy consumption for MD is affected by the membrane material and design configuration. Many configurations have been designed to optimize performance by varying the downstream phase or pressure conditions. For example, permeating the vapor into air, similar to the air gap MD configuration, improves thermal efficiency since thermal resistance to the air side of the membrane is higher [10]. Yet, the required energy consumption per permeate flux is still high at $57,600\text{--}122,400\text{ kJ}/\text{kg}$ [11], measured by the electrical power consumed in heating and recirculating feed. Cascade configurations have been also attempted to recover energy on the permeate side. Indeed, Multi-stage MD (MSMD) has been reported to improve thermal efficiency by recovering and reusing of vapor latent heat [8,12,13]. Nonetheless, one of the major factors for energy

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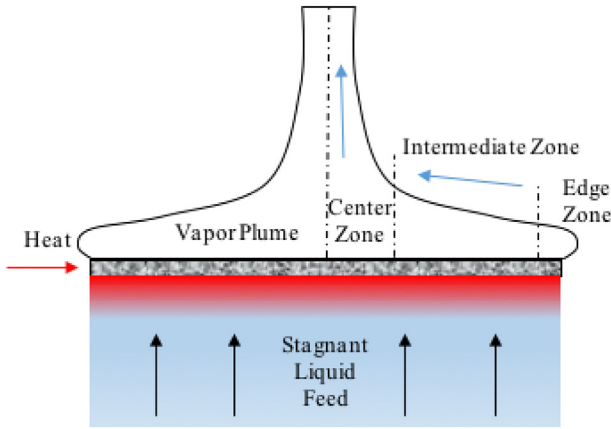


Fig. 1. Schematic diagram of the locally heated membrane distillation configuration. The membrane is heated from the edges with an electrical heater (not shown), rather than heating the feed fluid as with conventional membrane distillation approaches. The relatively stagnant feed flow reduces the energy required to maintain the liquid-vapor surface at the desired temperature. A hydrophobic, microporous membrane helps to ensure that the liquid-vapor interface remains at membrane surface where heat is supplied. Only vapor passes through the membrane to a lower vapor pressure zone away from heated region and can be collected with a condenser.

consumption is heating the entire circulating feed flow. Thus, the use of direct membrane heating has been analyzed to improve thermal efficiency [14].

This paper evaluates energy efficiency improvements of direct membrane heating method through controlled MD configuration, locally heated membrane distillation (LHMD), that accurately measures energy requirement per unit mass of distilled fluid as shown in Fig. 1. Specifically, heat is localized at the region of the fluid in close contact with the membrane using an electric heater. Compared to heating the bulk feed water, localized heating can reduce energy consumption to reach same temperature near the membrane due to the reduced effective thermal mass and elimination of the feed re-circulation power consumption. This configuration has the potential to be applied directly on naturally occurring contaminated or seawater surfaces including seawater ponds for localized clean water production. Additionally, membrane structural reliability increases due to reduction of feed circulation erosion and hydrodynamic pressure. Localized heating at the membrane interface allows this membrane distillation method to be used in miniature and modular designs to fit wide range of evaporation application beyond desalination alone. Further, the temperature gradient away from membrane could induce cool crystallization away from membrane allowing for salt collection stage in continuous operation.

This work first develops an analytical model for thermal and fluid transport in the LHMD system, then demonstrates a lab-scale version of the same. In particular, we estimate analytical mass flux limits for free surface natural convection and introduce the mass flow resistance of membrane, analogous to thermal flow resistance, to account for flow through the membrane. Then, the predicted mass flux is experimentally validated including measurements of energy consumption. Ultimately, this work demonstrates a significant energy consumption reduction of MD processes. Lab scale experimental validation demonstrates that in this first prototype design, 25% of the input electrical power is used for evaporating stagnant fluid at 80 °C, and this efficiency can be significantly improved through better insulating the fixture.

2. Theoretical limits for desalination (mass flux)

In the LHMD process, the transfer of water vapor through the member involves several thermo-fluid phenomena. Liquid is in contact with the bottom surface of the heated membrane. Evaporation occurs at

the liquid-vapor interface at the bottom of the porous membrane. Water vapor flows through the membrane and into the air above the membrane.

To predict the mass flux, we first consider the top surface of the heated membrane. At this surface, the temperature dependent evaporative mass flux is analytically calculated by analogy to natural convection above circular disk. Then, we superimpose the porous media mass resistance effect on free surface natural evaporation. Analogous to a thermal flux being proportional to a temperature gradient, the mass flux is proportional to water vapor pressure gradient. In this analytical model, we use the Boussinesq approximation: variable air density is only considered in the buoyancy terms. The flow is also assumed to be axisymmetric. Thus, the velocities in vapor phase region depend on the cylindrical coordinates (r and z): $u = u\vec{r} + w\vec{z}$. Using these assumptions and noting that gage pressure is defined as $\varpi = P - P_{stat}$ where P_{stat} is the hydrostatic pressure such that $(-\rho_{\infty}g - \frac{\partial P_{stat}}{\partial z} = 0)$, the steady non-dimensional Navier-Stokes and mass transport equations become

$$\frac{1}{\tilde{r}} \frac{\partial(\tilde{r}\tilde{u})}{\partial\tilde{r}} + \frac{\partial\tilde{w}}{\partial\tilde{z}} = 0, \quad (1)$$

$$\left(\tilde{u} \frac{\partial\tilde{u}}{\partial\tilde{r}} + \tilde{w} \frac{\partial\tilde{u}}{\partial\tilde{z}}\right) = -\frac{\partial\tilde{\varpi}}{\partial\tilde{r}} + \left(\frac{\partial^2\tilde{u}}{\partial\tilde{r}^2} + \frac{1}{\tilde{r}} \frac{\partial\tilde{u}}{\partial\tilde{r}} - \frac{\tilde{u}}{\tilde{r}^2} + \frac{\partial^2\tilde{u}}{\partial\tilde{z}^2}\right), \quad (2)$$

$$\left(\tilde{u} \frac{\partial\tilde{w}}{\partial\tilde{r}} + \tilde{w} \frac{\partial\tilde{w}}{\partial\tilde{z}}\right) = -\frac{\partial\tilde{\varpi}}{\partial\tilde{z}} + Gr \tilde{c} + \left(\frac{\partial^2\tilde{w}}{\partial\tilde{r}^2} + \frac{1}{\tilde{r}} \frac{\partial\tilde{w}}{\partial\tilde{r}} + \frac{\partial^2\tilde{w}}{\partial\tilde{z}^2}\right), \quad (3)$$

$$\tilde{u} \frac{\partial\tilde{c}}{\partial\tilde{r}} + \tilde{w} \frac{\partial\tilde{c}}{\partial\tilde{z}} = \frac{1}{Sc} \left[\frac{1}{\tilde{r}} \frac{\partial}{\partial\tilde{r}} \left(\tilde{r} \frac{\partial\tilde{c}}{\partial\tilde{r}} \right) + \frac{\partial^2\tilde{c}}{\partial\tilde{z}^2} \right], \quad (4)$$

where $Gr = \frac{|\rho_s - \rho_{\infty}| g R^3}{\rho_{\infty} \nu^2}$ is the Grashof number, ρ is vapor density, ν is the kinematic viscosity, g is acceleration due to gravity, R is the radius of the membrane, $Sc = \nu/D$ is the Schmidt number, D is mass diffusivity, $\tilde{r} = r/R$, $\tilde{u} = Ru/\nu$, $\tilde{w} = R w/\nu$, $\tilde{\varpi} = (R^2\varpi)/(\rho_{\infty}\nu^2)$, and non-dimensionalized vapor concentration is $\tilde{c} = (c - c_{\infty})/(c_s - c_{\infty})$. Subscripts s and ∞ correspond to saturation conditions and ambient conditions respectively. Here, the saturated vapor concentration at liquid vapor interface, c_s , and the ambient air vapor concentration, c_{∞} , are used as boundary conditions. Using those governing equations and Fick's law, the total mass flux is calculated from an integral over the surface of the membrane:

$$Q = \int -D \frac{\partial c(r, z)}{\partial z} \Big|_{z=0} dS, \quad (5)$$

where D is the diffusivity of the vapor phase into air.

2.1. Diffusion limits

To estimate the lower limit of the evaporated mass flux, we consider the case of the Grashof number to be much smaller than 1. This condition means that the mass flux is purely from diffusion, with no driving buoyancy force of the vapor. In estimating diffusion limits, we also assume that vapor diffusivity, D , is constant. Consequently, uncertainties in the diffusion limits increase at high temperatures. It is important to bear in mind that these limits are developed for free surface natural evaporation in the absence on mass transport resistance caused by membrane. However, in Section 2.3, membrane effects are superimposed onto free surface natural convection limits. Hence, Eqs. (1)–(4) reduce to

$$\frac{1}{\tilde{r}} \frac{\partial}{\partial\tilde{r}} \left(\tilde{r} \frac{\partial\tilde{c}}{\partial\tilde{r}} \right) + \frac{\partial^2\tilde{c}}{\partial\tilde{z}^2} = 0. \quad (6)$$

Eq. (6) can be solved by separation of variables in oblate spheroidal coordinates (k, θ) as in Ref. [15], such that the non-dimensional concentration, \tilde{c} , the local diffusive mass flux, j_{diff} , and the total diffusive mass flux, Q_{diff} , are

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