INSTIGATING BUOYANCY DRIVEN CONVECTION TO IMPROVE MEMBRANE DISTILLATION PERFORMANCE

by

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ABSTRACT

Membrane distillation is a thermally driven desalination process that is capable of treating complicated wastewaters with high concentrations of dissolved solids. However, temperature polarization and concentration polarization reduce its distillate production and impede its industrial adoption. Surprisingly, no prior work has considered that temperature and concentration polarization both increase the feed density near the membrane. In this exploratory study, we explore whether these differences in local feed density can instigate buoyancy driven convection in the feed flow and mitigate polarization. We also explore whether we can strengthen this convection by actively heating the feed channel in a direct contact membrane distillation (DCMD) system. For that purpose, we develop several actively heated prototypes and perform a series of experiments to explore how active heating affects distillate production. The prototypes are tested in multiple orientations to determine if buoyancy driven convection is truly taking place. The tests are also performed at multiple feed flowrates and feed concentrations to explore how temperature and concentration polarization affect the development of buoyancy driven convection.

Overall, our experiments suggest that, with the cell oriented properly, buoyancy driven convection can be harnessed to significantly improve distillate production. For all flowrates tested, we observed a linear increase in distillate flux with increasing wall heating. The impact of wall heating also increases as the flowrate decreases.
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Membrane distillation (MD) is a thermally driven desalination process that is capable of treating complicated wastewaters with high concentrations of dissolved solids. MD operates by flowing warm feed water over a porous hydrophobic membrane, as illustrated in Figure 1.1. The hydrophobicity of the membrane prevents the liquid feed from entering the pores of the membrane; however, a difference in partial vapor pressure across the membrane causes water to evaporate from the feed side of the membrane and travel through the membrane pores [1, 2]. Nonvolatile solutes remain in the feed. There are several different approaches to MD, which differ in how the transmembrane partial vapor pressure difference is generated, and how the water vapor is condensed and collected. This study focuses on Direct Contact Membrane Distillation (DCMD). This approach to MD flows cool liquid distillate water on the opposite side of the membrane from the feed side, as sketched in Figure 1.1. The temperature difference ($\Delta T_m$ in Figure 1.1) between the feed and distillate generates the difference in vapor pressures. The water vapor condenses in the cool distillate and is carried away in the distillate flow.

Our interest in DCMD stems from the rapid growth in reverse osmosis (RO) systems used to desalinate seawater, brackish water, and wastewater from municipalities, industry, and agriculture [3–5]. RO is a pressure-driven desalination process in which the feed is pressurized (up to roughly 80 bar) and forced through a semi-permeable membrane that blocks dissolved salts. The high feed pressure is required to overcome the feed osmotic pressure. Using this approach, conventional RO systems can concentrate feeds to roughly 70 g/L of total dissolved solids (TDS) [6]. This produces a recovery rate of about 50% for
seawater and up to 90% for brackish water. The remaining waste brine generated by RO poses environmental challenges due to its high salt concentration and presence of residual pretreatment chemicals, cleaning chemicals, and heavy metals from corrosion [7]. In seawater desalination, waste brine is often discharged to the sea, where it adversely affects water quality and marine life [7–9]. For inland desalination, waste brine is often discharged to surface waters, sewers, wells, or land surfaces, where it contaminates water resources and soil [7, 8, 10]. Brine can also be discharged to evaporation ponds; however, these consume considerable area as they must be shallow to maximize evaporation [8].

MD has several characteristics that make it an attractive technology for recovering additional water from high concentration brines, and minimizing the volume of brine discharged. First, while RO is a pressure driven process, MD is an evaporative process whose thermodynamics are less sensitive to osmotic pressure [2, 11]. This allows MD to concentrate feeds beyond 350 g/L TDS [12]. Other existing thermal desalination processes like multi-stage flash, multi-effect distillation, and vapor compression can also concentrate waste brines, but they have much larger physical footprints and capital costs compared to MD [13, 14]. MD can also operate at relatively low feed temperatures (40 - 90 °C), which can be provided by renewable sources or low quality waste heat from nearby industrial processes [11, 15]. MD systems have been successfully operated using solar, geothermal, and industrial heat [16–22].

Two main issues impeding the industrial adoption of MD are temperature polarization and concentration polarization [23, 24]. Temperature polarization refers to the formation of temperature boundary layers along the feed and distillate surfaces of the membrane, as demonstrated by the temperature profile sketched in Figure 1.1. These boundary layers form due to heat transfer through the membrane via conduction and water evaporating along the feed side of the membrane and condensing on the distillate side. Temperature polarization reduces the transmembrane temperature difference, $\Delta T_m$, which reduces permeate production. Concentration polarization refers to the accumulation of dissolved salts in a concentration boundary layer along the feed side of the membrane. This occurs because salts are advected to the membrane where they are blocked, while water vapor is removed. Concentration polarization is represented in Figure 1.1 by the black dots along the membrane. The higher solute concentration at the membrane reduces the feed partial vapor pressure, but the reduction is typically small compared to that caused by temperature polarization. Concentration polarization is nevertheless a major concern because if the solute concentration at the membrane becomes saturated, salts can precipitate onto the membrane. This precipitation, called mineral scaling, blocks membrane pores and reduces permeate production. Mineral scaling can also reduce the hydrophobicity of the membrane and cause pore wetting [11, 25]. Pore wetting refers to the intrusion of liquid water into the pores of the membrane. Liquid feed can flow through wetted pores, and carry solutes into the permeate, which reduces salt rejection [26].
Surprisingly, no prior work has considered that temperature and concentration polarization both increase the feed density near the membrane. By flipping the orientation in Figure 1.1, so that the feed channel is on the bottom and gravity is pointing down, buoyancy driven convections can be triggered such that plumes of cool, high concentration, feed sink away from the membrane, as sketched in Figure 1.2(a). We hypothesize that this brings low concentration, high temperature, feed to the membrane, mitigating both temperature and concentration polarization. This hypothesis is supported by recent computational fluid dynamics (CFD) simulations performed in our group. The simulations, presented in Chapter 2, show that buoyancy driven convection does occur for select operating conditions. The convection increases local permeate production, but then quenches itself downstream as energy is lost to evaporation and transmembrane conduction. We subsequently hypothesize that buoyancy driven convections can be strengthened and sustained by heating the channel plate opposite the membrane to generate rising plumes of warm fluid, as sketched in Figure 1.2(b). These rising plumes of warm fluid could replenish heat lost to evaporation and conduction, and also interrupt the concentration and temperature boundary layers that form along the membrane surface. The plate can be heated using resistive Joule heating or by flowing a heat transfer fluid from a renewable or waste heat source, as sketched in Figure 1.2(b). Each method has unique benefits and drawbacks. Using a heated fluid facilitates the use of renewable heat sources, like solar or geothermal, without needing additional infrastructure, such as solar panels, to convert the heat to electricity. However, additional pumps would be required to circulate the heating fluid, and the additional flow channels for the heating fluid would reduce how tightly the membranes could be packed together. The heat provided would also likely be uneven. In contrast, resistive heating elements would require additional electrical input, but simplify the design. Thin resistive heaters are already available “off the shelf” that
allow membranes to be packed efficiently without additional pumping costs. Electrical heaters may also provide more even heating and control over how much heat is provided inside the cell.

**Objectives:**
Thus motivated, this exploratory study has the following objectives:

1. Explore the practical challenges of heating feed within a DCMD cell using resistive heating or a secondary working fluid.

2. Narrow down to one or two designs for a more in depth experimental study.

3. Experimentally confirm that the cell orientation and operating conditions can induce buoyancy driven convection in DCMD. To accomplish this, we perform a parametric study in which we systematically vary the cell orientation, the feed and distillate flowrates, and the wall heating. The performance of the cell is evaluated in terms of the average distillate flux and a heat balance that estimates what percentage of the wall heating successfully contributed to increasing distillate production.

4. Evaluate the potential for wall heating and buoyancy driven convection to improve the performance of DCMD.

5. Identify promising avenues for future work.

**Organization:**
The remaining thesis is organized as follows. Chapter 2 presents a review of relevant literature. Chapter 3 describes our prototyping process, system and flow cell information, experimental procedure, and how we determine how active heating is being utilized. Chapter 4 presents our results for our experiments with multiple prototypes. Finally, we summarize our findings in Chapter 5.
CHAPTER 2
BACKGROUND

2.1 Motivating Case Study: Produced Water

As a motivating case study, we consider produced water generated by oil and gas extraction. Produced water refers to naturally occurring water that is brought to the surface with oil and gas [27, 28]. Produced water is the largest waste stream in the oil and gas industry [29]. Worldwide, the ratio of produced water to oil is about 3:1, while the average in the U.S. is estimated to be closer to 7:1 [29]. Produced water can contain a vast array of chemical compounds including salts, oil and grease, heavy metals, radioactive materials, and chemical additives from drilling operations, but the types and concentrations of impurities vary greatly between sites. In Colorado, the quality of produced water ranges from a few grams per liter of TDS to over 140 g/L TDS [30]. Produced water is primarily disposed in subsurface salt water aquifers, many of which previously contained hydrocarbon deposits, via injection wells [31–33]. This subsurface disposal is problematic because it is linked to an increase in seismic activity [34–36]. This disposal method can also be costly and contaminate drinking water, which has led to increasing regulatory pressures [30, 37]. Though RO can concentrate feeds to 70 g/L, the volumes of produced water are so large, that the remaining waste brine remains a serious challenge. In many cases, however, produced waters have initial concentrations above 70 g/L, such that conventional RO cannot be used at all.

2.2 Development and types of MD

Review articles date the origins of MD to the 1960s when patents were filed by Bruce Bodell and Peter Weyl [1, 38]. To our knowledge, the first papers on MD were published in the late 1960s [1, 38–40]. These papers compared different membrane materials in DCMD. Due to the low flux rates achieved with MD at the time, interest in the process quickly declined. In the early 1980s, interest in MD resurged due to improvements in membrane manufacturing techniques [1, 38].

The four common types of MD, demonstrated in Figure 2.1, are direct contact membrane distillation (DCMD), sweep gas membrane distillation (SGMD), air gap membrane distillation (AGMD), and vacuum membrane distillation (VMD). The methods differ in how the partial vapor pressure is controlled in the distillate channel and how the water vapor is condensed. DCMD, sketched in panel (a), flows cool liquid distillate in direct contact with the membrane. Vapor condenses in the liquid distillate and flows out of the channel. DCMD is the most widely tested and simplest type of MD to operate because it requires the least equipment [1, 41]. The main drawback of DCMD is that conductive heat losses through the membrane
Figure 2.1 Configurations for (a) DCMD, (b) VMD, (c) AGMD, and (d) SGMD. In each configuration, the liquid feed is represented by the red arrow on the left and the permeate channel is on the right. Adapted from [1].

Tend to be high because the hot and cold flows are in direct contact with the membrane [12, 41]. DCMD also has temperature polarization on both sides of the membrane due to heat flow into the liquid distillate via conduction and thermal energy released from water vapor condensing.

VMD, sketched in Figure 2.1(b), uses a vacuum pump to maintain a low partial vapor pressure in the permeate channel and draw the water vapor out of the flow cell and into a condenser. The major advantage of VMD is that the low thermal conductivity of the rarefied vapor in the distillate channel minimizes conductive losses [1, 23]. Unfortunately, the use of a vacuum pump increases the energy demand and complexity of the system. AGMD, sketched in Figure 2.1(c), attempts to reduce the conductive heat losses associated with DCMD by maintaining a layer of stagnant air on the distillate side of the membrane. The water vapor passes through the stagnant air and condenses on a cool surface and flows out of the channel to be collected. The air gap acts as an insulator that reduces conductive heat losses, but it also increases resistance to mass transfer as the vapor must travel through the stagnant air to reach the condensing plate. This causes AGMD to typically have lower flux values than other modes of MD [1, 23]. SGMD, sketched in Figure 2.1(d), flows an inert gas through the permeate channel to sweep the water
vapor out of the cell and into a condenser. The sweep gas reduces conductive heat losses and enhances vapor transfer in comparison to AGMD. The major disadvantage of SGMD is that the concentration of water vapor in the sweep gas is small. A large condenser is consequently required to minimize loss of water vapor to the environment [1, 23].

2.3 Alternative methods of treating high concentration brines

Historically, high concentration brines have been treated using thermal distillations processes such as multistage flash and single/multiple-effect distillation [42]. Single-effect distillation, sketched in Figure 2.2(a), places feed water in an evaporator tank with a steam heat exchanger. The exchanger heats the feed, and the pressure in the evaporator tank is lowered to increase the evaporation rate. The water vapor then flows to a condensing chamber and is collected. In multi-effect distillation, sketched in Figure 2.2(b), multiple evaporation chambers are placed in series. The first chamber uses steam to boil feed water. The steam produced in the first evaporator then flows to a heat exchanger that doubles as a condenser in a second evaporation chamber. As the steam moves through the heat exchanger, the steam condenses and the heat lost to the feed causes it to evaporate. The steam produced in the second chamber is subsequently used in a similar manner in another evaporation chamber. This process can then continue through more evaporation chambers in series. Each evaporation chamber in the series is kept at a lower pressure than the previous chamber to increase the evaporation rate. Multistage flash, sketched in
Figure 2.2(d), works similarly to multi-effect distillation, except that the feed is not reheated in each stage. Instead, the feed water is initially heated before passing through multiple chambers with decreasing pressures. As the feed moves from one chamber to the next, the pressure drops and causes more water to evaporate [42]. These thermal distillation methods are well known, well studied, and have been used in desalination plants around the world. However, they take up more space and have higher predicted capital costs than MD [13, 14].

To overcome the limitations of conventional RO, high pressure RO (HPRO) systems have recently been proposed that operate at pressures around 200 bar to concentrate feeds to 250 g/L TDS. The major challenge to HPRO is that it requires the development of new RO cells and membranes that can withstand high operating pressures. This development is an ongoing field of research [43–45]. Instead of increasing the operating pressure, several recently proposed RO methods treat high concentration brines by flowing a saline solution on the permeate side of the membrane to reduce the transmembrane osmotic pressure [6, 46, 47]. Cascading osmotically mediated RO (COMRO) uses multiple RO stages in series. Figure 2.3(a) shows an N+1 stage COMRO system. Low pressure feed enters the system through the first RO stage on the permeate side of the membrane. It similarly flows on the permeate side of the membranes for each downstream stage until it reaches the final N+1 stage. The final stage uses conventional RO to desalinate the feed and produce the final permeate. The waste brine from this final stage is then brought back through each of the previous RO stages as the feed. Overall, the initial feed is diluted as it flows through each stage.
downstream through the permeate channels of the first N stages. The waste brine from the final stage is then further concentrated as it flows back through the feed channels of the N upstream stages. In the process, the transmembrane osmotic pressure in the first N stages is reduced, allowing COMRO to treat high concentration brines [46]. In osmotically assisted RO (OARO), sketched in Figure 2.3(b), high pressure feed enters the first stage on the feed side of the membrane, while a saline solution flows through the permeate channel to reduce the transmembrane osmotic pressure. This requires the saline sweep solution to have a lower concentration than the feed solution. The saline sweep solution flows in a cycle between the permeate channel of the first stage and the feed channel of a second stage. The second stage once again uses a saline sweep flow on the permeate side to reduce the transmembrane osmotic pressure. This process continues until reaching a final RO stage that produces the final permeate [46, 47].

The major drawback to COMRO and OARO is that conventional spiral wound RO modules are not designed to flow feed or saline solutions on the permeate side of the membrane. Low salt rejection RO (LSRRO), sketched in Figure 2.3(c), overcomes this limitation using low salt rejection membranes. The feed initially passes through a conventional RO stage with a high salt rejection membrane. The waste brine from the first stage then passes through a second stage that contains a low-rejection membrane. The leakage of salt through the membrane reduces the transmembrane osmotic pressure. This allows the second stage to further concentrate the feed. Because the permeate produced from this stage is saline, it is pumped back to the feed of the first stage to be polished. The waste brine from the second stage is pumped to a third stage with a lower-rejection membrane. This process continues in each stage of the system, with permeate being pumped to the feed of the previous stage, and waste brine being pumped to the feed of the next cell.

2.4 Previous attempts to actively heat in MD modules

Previous attempts to actively heat feed as it flows through an MD cell focused on solar heating and resistive Joule heating. Chen and Ho [48] heated a DCMD system by exposing the feed channel plate to direct sunlight. They achieved a 16% increase in flux compared to when no solar heat was added. Similar designs have been tested in AGMD, SGMD, and VMD [19, 20, 49, 50]. Another option for using solar energy is to coat the membrane in a solar absorbing material. The goal of heating the membrane directly is to mitigate temperature polarization. This method was first used by Summers and Lienhard [20] with an AGMD setup. Summers and Lienhard concluded that heating the membrane surface directly was more efficient than an absorbing plate. In recent years, there has been a surge in interest in direct membrane heating, and new photothermal membranes have been developed in a number of studies [51–55]. The main challenge with these solar powered designs is their large physical footprint. Because they require direct
sunlight, the membranes must be spread over a large area, and cannot be stacked.

Resistive heating has also been used to heat membrane surfaces directly. A major challenge with this approach is the corrosion of the heating elements [56]. In recent years, carbon nanotubes (CNTs) have shown promise for use as resistive heating elements in membranes. Dudchenko et al. [57] were the first to successfully use CNTs to heat a membrane for MD. They were able to prevent the CNT structure from degrading by using high frequency AC currents to provide the necessary energy input to the CNT elements. Since then, heated membranes using CNTs have been the focus of several studies [58–61].

A recent study by Wang et al. [62] developed a composite membrane with a thermally conductive layer made of an aluminum mesh. To avoid corrosion, the membrane was operated in a VMD system with the mesh on the distillate side. It should be noted that this method did not use Joule heating. The mesh is simply attached to a heat source, and heat conduction transfers heat into the cell. Using these designs, Wang et al. [62] successfully provided enough heat to drive permeate production. However, a subsequent study by Lou et al. [63] showed that the heating was unevenly distributed, and increased concentration polarization considerably. This method is unfortunately not applicable to DCMD, because the heated mesh would be in direct contact with the distillate.

2.5 Spacers and their impact on MD performance

Membrane filtration systems, such as RO and MD, densely pack membranes to maintain a small physical footprint. When sheet membranes are used, spacers are placed in the flow channels to support and separate the membranes. Commercial spacers, shown in Figure 2.4, are polypropylene meshes about 1 mm thick [64]. In addition to providing support, the geometry of a spacer strongly affects the fluid flow and solute transport. Figure 2.5 shows the results of a CFD simulation performed by Lou et al. [65] to understand the effects of spacers on polarization in DCMD systems. They found that spacers create
vortical flows that strongly impact temperature and concentration polarization.

2.6 Motivating CFD Simulations

Here we briefly present results of CFD simulations performed by Dr. Federico Municchi, a post doctoral fellow in Dr. Tilton’s group at the Colorado School of Mines. Details of the numerical methods are provided in Federico et al. [66]. Here we present only the governing equations and several key results to motivate our study. The simulations consider a 2D feed channel in the VMD system sketched in Figure 2.6. The channel has a length \( L = 15 \) cm and height \( H = 3 \) mm. There is a flat-sheet membrane at \( y = H \) and an impermeable wall at \( y = 0 \). A feed solution of DI water and NaCl enters at \( x = 0 \) with a mean velocity \( U_{\text{in}} \), a uniform NaCl concentration \( C_{\text{in}} \), and a uniform temperature \( T_{\text{in}} \). Concentrate exits at \( x = L \). As feed flows downstream, water vapor exits the domain through the membrane. We assume that a vacuum
pump maintains a constant vapor pressure $p_{vac} = 5000$ Pa (absolute pressure) on the vapor side of the membrane. We assume that the gravitational vector $\mathbf{g}$ points in the positive (stable) or negative (unstable) $y$-direction.

### 2.6.1 Governing equations

The feed mixture velocity $\mathbf{u} = [u \ v \ w]$ and pressure $p$ are governed by the conservation of mass and momentum equations,

$$\frac{\partial \rho}{\partial t} + \nabla \cdot (\rho \mathbf{u}) = 0,$$

$$\frac{\partial (\rho \mathbf{u})}{\partial t} + \nabla \cdot (\rho \mathbf{u} \mathbf{u}) = -\nabla p + \rho \mathbf{g} + \nabla \cdot \mathbf{\tau},$$

where $\rho$ is the feed density, and the viscous stress tensor is given by

$$\mathbf{\tau} = \mu \left[ \nabla \mathbf{u} + (\nabla \mathbf{u})^T \right] + \lambda (\nabla \cdot \mathbf{u}) \mathbf{I},$$

where $\mu$ is the dynamic viscosity, $\lambda = -2\mu/3$ is the second viscosity coefficient, $\mathbf{I}$ is the identity matrix, and $(\nabla \mathbf{u})^T$ is the gradient transpose of $\mathbf{u}$. Heat transport is governed by the thermal energy equation [67],

$$\frac{\partial (\rho h)}{\partial t} + \nabla \cdot (\rho u h) + \frac{DK}{Dt} - \frac{\partial p}{\partial t} = \nabla \cdot \left( \kappa \frac{\nabla h}{c_p} \right) + \rho (g \cdot \mathbf{u}),$$

where $h$, $c_p$, and $\kappa$ are the mixture’s specific enthalpy, specific heat capacity, and thermal conductivity, respectively, $DK/Dt = \partial K/\partial t + \nabla \cdot (\mathbf{u}K)$, and $K = \rho (\mathbf{u} \cdot \mathbf{u})/2$. The transport of dissolved NaCl is governed by an advection-diffusion equation,

$$\frac{\partial c}{\partial t} + \nabla \cdot (\mathbf{u} c) = \nabla \cdot (\rho \Gamma \nabla \xi),$$

where $c$ is the NaCl concentration in units of mass of solute per liter of solution (g/l), $\xi$ is the NaCl mass fraction expressed as mass of solute per mass of solution, and $\Gamma$ is the mass diffusivity of NaCl. The variation of all thermophysical properties ($\mu$, $\rho$, $c_p$, $\kappa$, $\Gamma$) with temperature and concentration are expressed using polynomials determined by fitting to data from thermodynamic software (OLI [68]).

Vapor transport through the membrane is approximated using the model of Schofield et al. [2], which assumes the transmembrane vapor mass flux $j_v$ is proportional to the transmembrane difference in partial vapor pressure,

$$j_v = -B [p_m(x,t) - p_{vac}],$$

where $p_m$ is the partial vapor pressure of the solution on the feed side of the membrane, and $B$ is the vapor permeability of the membrane. We set $B = 1.8676 \times 10^{-6}$ kg/m$^2$ s Pa, which models a polypropylene membrane studied in Lou et al. [69]. As in most prior literature, we apply the no-slip condition ($u = 0$) to
the velocity component tangential to the membrane.

The dependence of the partial vapor pressure \( p_m \) on temperature and concentration is expressed as in Schofield et al. [2],

\[
p_m = a_w p_{sat}, \quad p_{sat} = \exp \left( 23.238 - \frac{3841}{T_m - 318} \right), \tag{2.6}
\]

where \( p_{sat} \) is the saturation pressure and is computed using the Antoine equation, \( T_m \) is local the temperature at the membrane \( (y = H) \), and \( a_w \) is an activity coefficient given by

\[
a_w = 1 - 0.03112 c \bigg|_{y=H} - 0.001482 c^2 \bigg|_{y=H}. \tag{2.7}
\]

Heat leaves the domain through the membrane due to advection by the vapor and conduction through the vapor and membrane material. In VMD systems, the contribution due to conduction is usually negligible. Conservation of energy principles then produce the following boundary condition for the temperature at the membrane,

\[
-\kappa \frac{\partial T}{\partial y} \bigg|_{y=H} = j_v \lambda_v, \quad \lambda_v = 3177800 - 2464.4 T \bigg|_{y=H}, \tag{2.8}
\]

where \( \lambda_v \) is the latent heat of evaporation of pure water. We also assume that NaCl does not leave the channel through the membrane. This requires the following condition applied to the mass fraction \( \xi \),

\[
\left( j_v \xi - \rho_f \frac{\partial \xi}{\partial y} \right) \bigg|_{y=H} = 0. \tag{2.9}
\]

We assume the feed enters the domain at \( x = 0 \) with uniform temperature \( T_{in} \), and uniform concentration \( C_{in} \). We also impose an inlet parabolic flow profile with mean velocity \( U_{in} \). At the outlet, we impose the following Neumann conditions,

\[
\frac{\partial T}{\partial x} \bigg|_{x=L} = 0, \quad \frac{\partial c}{\partial x} \bigg|_{x=L} = 0, \quad \frac{\partial u}{\partial x} \bigg|_{x=L} = \frac{\partial v}{\partial x} \bigg|_{x=L} = 0. \tag{2.10}
\]

At the lower plate, we apply no-slip and no-penetration boundary conditions to the velocity, and a zero salt-flux condition to \( c \)

\[
u \bigg|_{y=0} = v \bigg|_{y=0} = 0, \quad \frac{\partial c}{\partial y} \bigg|_{y=0} = 0. \tag{2.11}
\]

We apply a desired heat flux \( q \) at the plate through the condition,

\[-\left( \kappa \frac{\partial T}{\partial y} \right) \bigg|_{y=0} = q. \tag{2.12}
\]

### 2.6.2 Simulation results without wall heating \((q = 0)\)

To demonstrate buoyancy driven convection, we set the inlet temperature and concentration to \( T_{in} = 80 \) °C and \( c_{in} = 70 \) g/l, respectively, which are typical of a VMD system treating waste brine from an RO
Figure 2.7 Instantaneous snapshots of the concentration $c$ for the unstable configuration in the ranges $x \in [0, 2.5]$ cm (a), $x \in [2.5, 5]$ cm (b), $x \in [5, 7.5]$ cm (c), and $x \in [10.5, 13]$ cm (d). The gray rectangle represents the membrane, and the black lines indicate the boundary layer. Gravity is pointing downwards, and there is no wall heating.

We set the mean inlet velocity to produce an inlet Reynolds number $Re = 75$, where $Re = \rho_{in}U_{in}H/\mu_{in}$ is defined using the inlet density $\rho_{in}$ and viscosity $\mu_{in}$. Simulations were then performed for both the stable and unstable orientation of gravity. The wall heating was set to zero, $q = 0$. Figure 2.7 shows the instantaneous concentration field at time $t = 300$ s for different downstream sections (see figure caption) when gravity points in the negative $y$-direction (the unstable orientation). We identify the concentration boundary layer by plotting the black iso-lines $c = 1.01 c_{in}$. Panel (a) shows that for the initial length $0 \leq x \leq 2.5$ cm, the boundary layer initially grows without any discernible buoyancy-driven convection. In panel (b), however, a sinusoidal perturbation becomes evident in the iso-line around $x \approx 3$ cm. Over the downstream section $3 \leq x \leq 4$ cm, the perturbation grows and solute rich plumes become evident for $x > 4.25$ cm, particularly in panel (d), which shows the downstream section $10.5 \leq x \leq 13$ cm. The absence of black lines in panel (d) means that $c > 1.01c_{in}$ everywhere.

Figure 2.8 shows the corresponding instantaneous temperature field at $t = 300$ s. We identify the thermal boundary layer by plotting the black iso-line $T = 0.99 T_{in}$. As in Figure 2.7, a sinusoidal
Figure 2.8 Instantaneous snapshots of the temperature $T$ for the unstable configuration in the ranges $x \in [0, 2.5]$ cm (a), $x \in [2.5, 5]$ cm (b), $x \in [5, 7.5]$ cm (c), and $x \in [10.5, 13]$ cm (d). The gray rectangle represents the membrane, and the black lines indicate the boundary layer. Gravity is pointing downwards, and there is no wall heating.

Perturbation becomes evident around $x \approx 3$ cm. However, because the thermal diffusivity is orders-of-magnitude larger than the mass diffusivity, energy mixing occurs more rapidly than solute mixing, and the iso-line $T = 0.99T_{in}$ disappears around $x \approx 5$ cm. Figure 2.8(d) shows a substantial mixing and cooling of the feed in the downstream section $x \in [10.5, 13]$ cm.

To compare the results of stable and unstable cases, we compute the local time-averaged permeate flux

$$\langle j_v \rangle (x) = \frac{1}{\Delta t_s} \int_{t_0}^{t_0 + \Delta t_s} j_v(x, t) \, dt,$$

where $\Delta t_s$ and $t_0$ are the sampling interval and start time, respectively. The red line in Figure 2.9(a) shows the downstream variation of $\langle j_v \rangle$ for the simulation shown previously in Figure 2.7. For comparison, the blue line shows results for the stable orientation. For the stable case, $\langle j_v \rangle$ decreases monotonically with downstream distance over the full channel length. The unstable case shows a similar decrease in permeate production up to $x = 3.6$ cm, after which $\langle j_v \rangle$ begins increasing rapidly. Hereinafter, we refer to the
Figure 2.9 (a) The time averaged flux for the stable and unstable configurations when Re = 75. (b) The time averaged flux in the unstable orientation when Re = 50 (dashed line), 75 (solid line), and 100 (dash-dotted line). There is no wall heating in either panel.

The x-location where $\langle j_v \rangle$ begins increasing as the onset distance $x_{on}$. Downstream of the onset distance, $\langle j_v \rangle$ reaches a local maximum and subsequently begins decreasing. This decrease likely occurs because the increase in permeate production downstream of $x_{on}$ also increases energy lost to evaporation, so that the instability quenches.

To investigate the impact of the inlet Reynolds number, we perform simulations with Re = 50, 75, and 100 for the unstable orientation. Figure 2.9(b) shows the resulting time-averaged vapor fluxes $\langle j_v \rangle$. We see that decreasing the Reynolds number promotes instability by pushing $x_{on}$ upstream and increasing the local maximum $j_{max}$. Though decreasing the Reynolds number increases vapor production in the upstream half of the channel, it also decreases vapor production in the downstream half of the channel due to earlier quenching.

**2.6.3 Simulations with wall heating**

Figure 2.10 shows the impact of wall heating ($q$) on the time-averaged vapor flux $\langle j_v \rangle$ when Re = 150 in a VMD system in the unstable configuration. When $q = 0$, there is little instability in the channel. However, as the wall heating is increased to 10, 20, and 40 kW/m$^2$, instability moves upstream. Moreover, for 20 and 40 kW/m$^2$, instability never shuts down, maintaining strong vapor flux downstream.

**2.7 Proposed Scope**

Balancing the merits of each approach to MD, we identified DCMD and VMD as two promising candidates for active wall heating. DCMD is promising due to its popularity and simplicity. Our lab is also
Figure 2.10 Impact of wall heating on the time averaged vapor flux when $Re = 100$.

experienced in the design and operation of DCMD systems, and had an existing system available for use. The main drawback to DCMD is the conductive heat loss through the membrane. We expect these losses to increase with active heating, and limit the maximum length of a heated DCMD cell due to heating of the distillate. This issue could be solved by actively cooling the distillate channel; however, the system would become much more complex. These issues make VMD a promising alternative, due to its low conductive heat losses. The main drawback to VMD is its complexity. Furthermore, our lab has little experience with VMD and the issues associated with keeping a system airtight, collecting and measuring the permeate flux, and obtaining repeatable results. Consequently, for this exploratory study, we consider DCMD, and leave VMD to future work. Because the physics of the vortices created by spacers are not fully understood, we also primarily performed experiments without a spacer in the feed channel. Not including a spacer in the feed channel also allows us to more easily see the effects of buoyancy driven convection since there is no additional mixing in the feed from the spacer geometry. Finally, while we do explore designs that can use a heated fluid to provide active heating, our experimentation focused on using resistive Joule heating so that we could avoid the additional complexity that an additional fluid loop would add to the system.
CHAPTER 3
METHODOLOGY

3.1 Heated Cell Prototypes

Figure 3.1 Sketch (not to scale) of a standard DCMD cell that was used to test prototypes. (a-c) shows how the cell is assembled with two gaskets, two acrylic plates, a spacer in the distillate channel, and a membrane sheet. (d) shows how the feed and distillate inlets are setup.

We explored several methods of distributing heat within the feed channel of a DCMD cell. Here we summarize those attempts, and motivate the final designs for which we present results in Chapter 4. Each method was initially tested in an existing DCMD cell, sketched in Figure 3.1. The feed and distillate channels are formed by cutting the width $W_c = 75$ mm and length $L_c = 250$ mm from two rubber gaskets (panel (a)). The gaskets are placed on opposite sides of a membrane coupon (panel (b)) and a spacer is placed on the distillate side. The gaskets, membrane, and spacer are then sealed between two clear acrylic plates using bolts (panel (c)). As the bolts are tightened, tension is applied to the membrane tabs to remove wrinkles from the membrane. The bolts are tightened sufficiently to ensure no leakage and that the channel heights are set by the spacer height. When no feed spacer is used in our experiments, we must ensure that the channel height remains constant between tests. For that purpose, we printed an “empty” spacer that only had fillaments going along the perimeter of the cell. This is discussed later in this section and in Figure 3.6.
Prototype A: As a quick proof of concept, we began by simply packing a folded nichrome wire into the feed channel of the flow cell. Figure 3.2(a) shows a photograph in which the nichrome wire is visible through the clear acrylic plate. This design was briefly tested with DI water, but quickly proved inefficient and inconsistent.

Prototype B: We then wove the nichrome through a PTFE coated fiberglass mesh to provide support to the wires and increase repeatability, as seen in Figure 3.2(b). The fiberglass mesh increased the packing density of the wires and made it easier to fit the wire in the feed channel. While testing this design, we noticed that the nichrome wire corroded. We consequently coated the wires and mesh with rubber sealant spray. Unfortunately, when an electrical current was passed through the wires, the coating separated. Overall, we also found it nearly impossible to consistently fabricate identical heated meshes and obtain repeatable results. Moreover, these early designs heat the feed using an immersed wire, rather than through the feed plate.
Prototype C: We subsequently explored whether nichrome wires could be glued directly to the feed channel plate. The first attempt glued nichrome wires to a sheet of PTFE. The wires were straightened with a device lent to us by Dr. Craig Brice that stretched the wire between two aluminum blocks with pins, pictured in Figure 3.3(a-b). The wire was then glued to the PTFE sheet with epoxy and sealed with silicone to prevent corrosion. We believe that this design shows promise, but we leave it to future study because the manufacturing challenges are significant, and remain beyond our scope. This approach also provides too many parameters that we would need to vary when testing, including the distance between wires (labeled w in Figure 3.3(c-d)), the orientation of the wire (longitudinal, as in Figure 3.3(c), or transverse, as in Figure 3.3(d)), and the wire diameter (labeled D in Figure 3.3(e)). This design opens the potential for optimal combinations of these parameters, such that distillate production is maximized while heat input is minimized. We leave this to future study.

Prototype D: Though we limited our study to testing with resistive heating, we briefly considered one approach to heating with a secondary heat transfer fluid. This design involved 3D printing the feed side plate of a standard DCMD cell out of stainless steel. The plate would be printed with an integrated channel which would carry a heat transfer fluid. A CAD drawing of a cross section for one of these proposed plates can be seen in Figure 3.4(a). The surface of the cell could also be textured, such as with fins, to improve heat transfer and feed mixing. Though the 3D printer can print complicated geometries,
we focused on designs that could potentially be mass produced using low cost extrusion. We believe that this design also shows promise, but leave it to future study because the manufacturing challenges are beyond our scope. As with prototype C, where wires were glued to a PTFE sheet, we would want to test with multiple different geometries. 3D printing these cells out of stainless steel is also expensive and time consuming. A printed visualization piece can be seen in Figure 3.4(b).

**Prototype E:** Our next design, pictured in Figure 3.5, used a flat silicone surface heater purchased from Omega Engineering. The heater was 3 in wide and 8 in long. It had a maximum power density of 10 W/in², and came with an adhesive on one side, which was used to attach the heater to the wall of the feed channel. Two holes were drilled through the acrylic plate to allow the heater leads to exit the cell. The holes were sealed with red RTV silicone gasket maker. The heater itself was not rated for aqueous

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**Figure 3.4** (a) CAD drawing of cross section of proposed 3D printed feed plate. (b) A small visualization piece which shows a cutout of a printed cell.

**Figure 3.5** Flow cell design with an Omega flat silicone surface heater and an empty spacer.
environments due to gaps where the leads enter the heater. We filled these gaps with clear silicone sealant to waterproof the heater. Most of the testing that we did with this design did not include the use of a feed spacer. In order to set the height of the feed channel and keep it constant between experiments, we used an “empty” spacer which only had filaments along the perimeter of the feed channel. A CAD drawing of the empty spacer can be seen in Figure 3.6. Support for the membrane is provided by the longitudinal filaments and the taller areas at either end of the spacer, which were set to a height that was 1.5 mm taller than the surface heater. The design allowed the cell to be easily loaded and unloaded. We performed a series of tests in which we adjusted the power supplied to the silicone heater under multiple feed and distillate flowrates. These results are presented in Chapter 4.

**Prototype F:** Though promising, placing the surface heater in direct contact with the feed had two main drawbacks. First, the heater had a wavy surface that likely generated uncontrolled mixing that inhibited buoyancy driven convection. It also caused large error bars in our measurements. Second, the heater would only last several weeks before burning out, which likely occurred due to leakage through the leads. We also realized that the DCMD cell should be longer to allow more distance for buoyancy driven convection to develop, and the feed channel should be wider to reduce the feed velocity. The main goal of our final design, shown in Figure 3.7, was to make the feed channel wall a flat flush surface, and to make the cell longer and wider. To make the feed channel wall flush, we made a new flow cell that had a cavity milled into the feed side plate. A 4 in wide and 12 in long surface heater was then placed inside the cavity, and covered with a metal plate. The metal plate was sealed at both ends, near the inlet and outlet, to prevent feed fluid from entering the cavity and contacting the heater. Thin silicone sheets were also placed between the heater and the acrylic cell to make sure that the heater was making direct contact with the metal plate. A k-type...
Figure 3.7 Pictures of prototype F showing (a) the assembled cell viewed from the feed side, (b) the feed channel in the cell, (c) the feed side plate without the gasket, (d) the heater inside the feed side plate, and (e) the feed side plate interior channel with silicone sheets and thermocouple.
thermocouple was also placed in the silicone sheets to monitor the internal temperature of the acrylic plate.

304 stainless steel was initially used for the metal plate, but it quickly corroded when tested with high concentration NaCl feed solution. Aluminum 5052 was then tested, but it also began to corrode, albeit slower than the 304 stainless steel. 316 stainless steel was finally selected because it did not corrode. The 316 stainless steel was sealed at either end with GE advanced silicone specialty metal silicone caulk and Nashua waterproofing foil tape, because the standard silicone sealant that we had been using previously did not bond to the steel. While the total area of the feed channel was 4 in x 13.75 in, the total active membrane area was reduced to 4 in x 13 in by sealing the ends of the membrane’s surface, over the inlet and outlet troughs, with silicone. This was done so that only the membrane area opposite the metal plate could produce permeate. Small strips of industrial spacer with a thickness of 2 mm were placed near the feed inlet and outlet, and held in place by the same sealant used to seal the metal plate to the cell. These spacer strips served to set the feed channel height and to provide some noise to the feed flow at the inlet. We wanted to mix the flow slightly because our simulations showed that buoyancy driven instability developed more easily if the inlet flow had some noise to it, instead of being fully developed. The same industrial spacer material was also used in the distillate channel to set the channel height and enhance mixing.

3.2 Flow Control and Instrumentation

The flow cell was operated in a continuous closed-loop system sketched in Figure 3.8. The feed loop is depicted on the left side of the figure and the distillate loop is depicted on the right. Feed solution exits the base of a 10 L storage tank and is pressurized by a recirculation pump before passing through a heat exchanger (Brazetek BT-STX-85) that maintains the desired inlet feed temperature by circulating a heated glycol solution. The feed then passes through the flow cell and returns to the tank through a hole in the tank’s upper side wall. This drops the returning feed into the stored feed below to enhance mixing. The tank is sealed with a lid to prevent evaporation to the surrounding environment. Inside the cell, heat is provided by a flat silicone surface heater with a maximum power density of 10 W/in$^2$ (OMEGA SRFGA). A 120 V 3 A adjustable DC power supply (Kungber SPS1203B) was used to power the heater in prototype E. A larger 500 W adjustable AC power supply (YaeCCC TDGC-0.5KM) was used to power the heater in our experiments with prototype F because the previous power supply did not go to a high enough wattage for the bigger heater, and because the heater was rated for AC voltage. Both power supplies were run through a ground-fault circuit interrupter so that the power would shut off if water got inside the heaters. The feed is circulated using a centrifugal pump (Micropump CA series powered by an EagleDrive DEMSE motor). Two temperature probes (Electronic Innovations EI1034) and pressure transducers (OMEGA
PX309-030G5V) monitor the feed temperature and pressure entering and exiting the flow cell. The measured inlet temperature is used to automatically turn the heat exchanger on and off to ensure a constant inlet temperature. The feed exiting the cell also passes through a conductivity probe (Sensorex TCSMA Blind Toroidal Conductivity Transmitter) before returning to the feed.

The distillate loop shown on the right side of Figure 3.8 operates similar to the feed loop. Distilled water exits a 10 L storage tank and passes through a recirculation pump (Micropump CA series centrifugal pump powered by an I-Drive IMS motor) and a heat exchanger (Brazetek BT-STX-85) that uses chilled water to maintain the desired inlet temperature to the flow cell. A digital flow meter on the feed and distillate loop is used to automatically adjust the pump power to maintain the desired flow rate. The temperature and pressure of the distillate entering and exiting the cell are monitored by temperature probes and pressure transducers (identical models as on the feed side). The measured distillate inlet temperature is used to automatically adjust the flow rate of chilled water through the heat exchanger via a solenoid valve. A conductivity probe (Eutech alpha-COND500) monitors the distillate leaving the flow cell to detect whether feed liquid is leaking through the membrane due to pore wetting or membrane damage. Finally, the distillate returns to the distillate tank through a port near the lower side wall to minimally disturb the water within. This reduces noise measured by a pressure transducer (OMEGA PX309-002GV) at the base of the distillate tank. This pressure measurement is used to compute the volume of water within the tank.

The distillate tank is placed above the feed tank, and the base of the distillate tank is fitted with a solenoid valve that allows a desired volume of distillate to return to the feed tank. This allows us to return the feed to the same salt concentration after each cycle is complete. The change in distillate volume over
time is also used to measured the vapor flux through the membrane. The data collection and operating conditions in all experiments were controlled using Labview software and a multichannel DAQ. The components of the distillate and feed loop are connected by vinyl tubing. The vinyl tubing that connects the heat exchangers to the flow cell inlets are wrapped in foam insulation to reduce heat losses.

All experiments used a CLARCOR QP952 membrane with a porosity between 0.70–0.85 and a nominal pore size of 0.45 µm. The membrane thickness varied between 150 µm–300 µm and featured an active layer of elongated polytetrafluorethylene (ePTFE) and a support layer of non-woven polyester (PES). This membrane was chosen because its support layer reduces membrane warping and increases repeatability of results. The flow cell is operated in a cocurrent configuration to reduce membrane warping due to opposing shear stresses on the feed and distillate membrane surfaces. Further membrane details can be found in the experimental study by Vanneste et al. [70].

3.3 Experimental Procedure

In all experiments, the feed tank was filled with 5L of either DI water or a solution of DI water and 70 g/L NaCl. This concentration is well below the solubility limit of NaCl (≈ 350 g/L), and is the typical concentration limit for conventional RO. The distillate tank was filled with 2L of DI water for each experiment.

For the experiments using prototype E, the feed and distillate flowrates were set to be equal. The tested flowrates were 0.5, 0.4, 0.3, and 0.2 L/min, which produce average inlet velocities of 7.29, 5.83, 4.37, and 2.92 cm/s respectively when divided by the cross-sectional area of the feed channel. The feed and distillate temperatures were maintained at 60 ± 3 °C and 28 ± 1 °C. To minimize any unsteady membrane flapping, the inlet feed pressure was kept slightly higher than the distillate inlet pressure so that the membrane would be pushed against the spacer placed in the distillate channel. Each experiment began by running the system for an hour with the surface heater turned off to allow the system to reach steady state. The system was then allowed to run for another hour with the heater turned off to get a baseline flux value. The flux value is calculated using the volume of water in the distillate tank, which is tracked over time by the system. The distillate volume data is imported into MATLAB, where it is then smoothed using the “movmean” function. A linear best fit is then applied to the smoothed data, and the slope of the best fit is divided by the membrane active area to obtain the flux. Examples of raw and smoothed distillate volume data are shown in Figure 3.9. Once the baseline data is acquired, the permeate is drained back to the feed tank, until 2L of permeate remains in the distillate tank. The surface heater is then turned on, adjusted to deliver 75W of power, and the system is run until the distillate temperature returns to steady state, typically 15-30 minutes. Once a steady temperature is achieved, the distillate is once again drained.
to 2L, and the system is run for an hour to measure flux. This process is then repeated with the heater set to 150 W and then 240 W. As stated earlier, while testing prototype E, we had issues with the surface heater burning out. This was likely caused by the intrusion of water into the heater, but it could have also been caused by too high of an internal heater temperature. To reduce the risk of burning out more heaters, after completing our experiments at 0.5 L/min, the maximum heating value was reduced from 240 W to 235 W for the remaining tested flowrates. Each flowrate was run with the feed on top, referred to as the stable orientation (Figure 3.10(a)), and on bottom, referred to as the unstable orientation (Figure 3.10(b)). This changes the direction of gravity with respect to the cell, and determines if buoyancy driven convection occurred.

For the set of experiments using prototype F, the distillate flowrate was set to 1.8 L/min, which produces an average inlet velocity of 14.8 cm/s. The tested feed flowrates were 1.4, 1.0, 0.8, 0.6, 0.4, and
0.2 L/min, which produce average inlet velocities of 11.5, 8.20, 6.56, 4.92, 3.28, and 1.64 cm/s respectively, and varies the Reynolds number between $128 \leq Re \leq 893$. The Reynolds’s numbers are calculated using equation 3.1

$$Re = \frac{\rho_{in} U D_h}{\mu_{in}},$$

where $\rho_{in}$ is the inlet feed density, $\mu_{in}$ is the inlet feed dynamic viscosity, $U$ is the mean inlet velocity, and $D_h$ is the hydraulic diameter of the feed channel. The feed and distillate temperatures were maintained at $60 \pm 3 \degree C$ and $30 \pm 1 \degree C$. The feed inlet pressure was again set slightly higher than the distillate inlet pressure to stabilize the membrane. These experiments were run using the same procedure used for the experiments with prototype E. The system is allowed to run for an hour to get to steady state, then it is run for an hour without additional heat provided to the cell to get baseline data. In the unstable orientation, the system runs through three one hour cycles with power levels increasing to 120 W, 240 W, and 360 W. In the stable orientation, the system only runs through two additional cycles with the heater set to 120 W and 240 W, due to the cell getting too hot at 360 W.

All experiments were run in triplicate to measure repeatability. Each of the three runs for all sets of tests were run with different membrane coupons. This was done because the properties of the membrane material are not uniform across the entire membrane area. Because of this, different membrane coupons will produce slightly different flux values. To conserve membrane material, each membrane coupon was used for as many runs as possible, but the same membrane coupon was never used for multiple runs of the same experiment.

After each experiment, the system was flushed with DI water until the feed conductivity probe read zero conductivity for a full hour. Between rounds of experiments, when it was time to switch the membrane, the flow cell prototypes were removed from the system and cleaned by hand with DI water and dish soap. During this time, a standard DCMD cell was hooked up to the system so that the system could be deep cleaned. For that, DI water is first flushed through the system for an hour. Next, the DI was drained and replaced with a NaOH solution with a pH of 11.5 or greater. The system is run for an hour and then flushed with DI water. A HCl solution with a pH of 2.5 or less is then run through the system for an hour. Finally, the system was flushed with DI again until the conductivity probe read zero conductivity and was allowed to run for another hour.

### 3.4 Heat Analysis

To estimate the percentage of supplied heat that successfully contributed to distillate production, we performed an energy balance. We start by assuming that the system is at steady state and use the first law
of thermodynamics, which can be written for a control volume as
\[
\frac{dE_{cv}}{dt} = 0 = \dot{Q}_{cv} - \dot{W}_{cv} + \dot{m}_i \left( u_i + p_i v_i + \frac{V_i^2}{2} + g z_i \right) - \dot{m}_e \left( u_e + p_e v_e + \frac{V_e^2}{2} + g z_e \right)
\] (3.2)
where the terms \(\dot{Q}_{cv}\) and \(\dot{W}_{cv}\) represent the net rate of energy transfer by heat and work respectively, \(\dot{m}\) is the fluid mass flow rate, \(u\) is the fluid internal energy, \(p\) is the pressure, \(V\) is the fluid velocity, and \(z\) is the height of the cell. Next, we begin eliminating terms. The kinetic and potential energy terms can be eliminated because the height of the cell does not change between the inlets and outlets and the difference between the inlet and outlet velocities is negligible. We also eliminate the pressure term because the pressure drop across the length of the cell is very small. Finally, we assume that no mechanical work is done on the system. This results in the energy equation reducing to
\[
\dot{Q}_{cv} + \dot{m} \left( c_p \Delta T \right) = 0
\] (3.3)
The inlet and outlet mass flowrates are also constant so \(\dot{m}_i = \dot{m}_e = \dot{m}\), and \(u\) can be calculated using \(u = c_p T\) with a constant heat capacity. This further reduces the energy equation to
\[
\dot{Q}_{cv} + \dot{m} \left( c_p \Delta T \right) = 0
\] (3.4)
where \(c_p\) is the heat capacity and \(\Delta T\) is the change in temperature. Figure 3.11 shows that the energy flows entering and leaving the system are the feed and distillate flows, environmental losses \((Q_{env})\), and \(P_{in}\) supplied by the surface heater. Plugging these terms into the energy balance gives
\[
P_{in} - Q_{env} + \dot{m}_i \left( c_{p,f} \Delta T_f \right) + \dot{m}_d \left( c_{p,d} \Delta T_d \right) = 0
\] (3.5)
We can then solve for the environmental losses by using the known heater, feed, and distillate values
\[
Q_{env} = P_{in} + \dot{m}_f \left( c_{p,f} \Delta T_f \right) + \dot{m}_d \left( c_{p,d} \Delta T_d \right).
\] (3.6)
Next, we calculate the heat contributing to distillate flux by multiplying the mass flux across the membrane, \( J \), by the heat of evaporation, \( \lambda \)

\[
Q_{evap} = J \rho m \lambda. \tag{3.7}
\]

Finally, with all other values known, we assume that environmental losses occur only on the feed side and solve for conductive losses

\[
\dot{m}_f (c_{p,f} \Delta T_f) = P_{in} - Q_{env} - Q_{evap} - Q_{cond}. \tag{3.8}
\]

To determine how much heat from the surface heater is going towards \( Q_{env}, Q_{evap}, Q_{cond} \), and how much is leaving the cell in the feed (\( Q_{feed} \)), we compare each value when the heater is on to the same values when the heater is off. For example, if \( P_{in} \) were set to 240 W, then we would calculate the heating values as

\[
Q_{env,total} = Q_{env, 240W} - Q_{env, 0W} \tag{3.9}
\]
\[
Q_{evap,total} = Q_{evap, 240W} - Q_{evap, 0W} \tag{3.10}
\]
\[
Q_{cond,total} = Q_{cond, 240W} - Q_{cond, 0W} \tag{3.11}
\]
\[
Q_{feed} = \dot{m}_f (c_{p,f, 240W} \Delta T_{f, 240W} - c_{p,f, 0W} \Delta T_{f, 0W}) \tag{3.12}
\]

These values can then be divided by \( P_{in} \) to convert them into percentages.
4.1 Results for prototype E with heater in direct contact with feed

We start by presenting results for prototype E, in which the heater is in direct contact with the feed. We began by testing prototype D using the feed and distillate flowrates $\dot{V}_{f_{in}} = \dot{V}_{d_{in}} = 0.5$ L/min. This produces the inlet Reynold’s number $Re = 472$. Experiments were repeated using a feed of DI water and a solution of DI water and 70 g/L NaCl. When the feed is DI water, there is no concentration polarization, and density variations only arise due to temperature polarization. When the feed is a NaCl solution, density variations arise due to both concentration and temperature polarization, and we hypothesize that the potential for buoyancy driven convection increases. We also repeated the experiments without a feed spacer and with an industrial feed spacer. We hypothesize that when the industrial feed spacer is present in the feed channel, the flow is dominated by vortical flow structures, such that buoyancy plays a small role in the flow regime. Nevertheless, the vortical structures alone may efficiently transport heat from the plate to the membrane surface. In each experiment, the heater power was set to $P_{in} = 0, 75, 150, \text{and } 240$ W.

![Figure 4.1](image-url) Variation of the measured distillate flux $J$ with input wall heating $P_{in}$ when $\dot{V}_{f_{in}} = \dot{V}_{d_{in}} = 0.5$ L/min and DI water is used as the feed. Panel (a) shows results without a feed spacer. Panel (b) shows results with a feed spacer. In both panels, the empty blue circles show measured data for the stable orientation, and the solid red stars show data for the unstable orientation. The solid lines show linear best fits.

Figure 4.1(a) shows the resulting distillate flux $J$ versus the heating power $P_{in}$ with a feed of DI water and no feed spacer. The symbols show the measured data for the stable (blue circles) and unstable (red
Figure 4.2 Drawing of bubbles that form in the feed channel. (a) depicts bubble formation without a feed spacer, while (b) depicts bubble formation with a feed spacer.

The lines show linear best fits, computed using MATLAB’s “polyfit” function. Overall, we see that the distillate flux increases linearly with $P_{in}$ for both orientations. Counterintuitively, we found that the stable orientation produced slightly more distillate flux than the unstable orientation. We believe that this occurred because when feed passes through the heater, the increase in temperature causes dissolved gasses to be released from the fluid, which then collect as bubbles. In the unstable orientation, the buoyant air bubbles rest against the membrane, as sketched in Figure 4.2(a), where they impede distillate production. The slopes of the linear best fits were 0.0139 LMH/W for the stable orientation and 0.0134 LMH/W for the unstable orientation. The increase in $J$ between $P_{in} = 0$ and 240 W was 3.38 LMH for the stable orientation and 3.32 LMH for the unstable orientation. These correspond to a relative increase of 25.0% and 25.6% respectively. Lastly, we note that the error bars in Figure 4.1(a) are rather large, and on the same order-of-magnitude as the increase in $J$ between $P_{in} = 0$ and 240 W.

Figure 4.1(b) shows $J$ versus $P_{in}$ using a feed of DI water and a spacer in the feed channel. We again observe that both orientations show a linear increase in $J$ with $P_{in}$. The slopes of the linear best fits were 0.0147 LMH/W for the stable orientation and 0.0164 LMH/W for the unstable orientation, which are slightly greater than those observed without the feed spacer. The increase in $J$ between the lowest and highest values of $P_{in}$ were 3.67 LMH (25.8%) for the stable orientation and 4.13 LMH (31.1%) for the unstable orientation. These values are again slightly better than those observed without the feed spacer. The data again showed that the stable orientation performed better than the unstable orientation; however, the difference in performances was slightly greater than when no feed spacer was used. We hypothesize that this can again be explained by air bubbles forming in the system. By observing the feed channel through the clear acrylic plate, we observed that air bubbles tended to accumulate where the feed spacer met the membrane surface, as sketched in Figure 4.2(b). In both orientations, the spacers appeared to exacerbate the trapping of air bubbles within the cell. This would further reduce the membrane’s active area in the unstable orientation. It is also possible that the orientations are performing very similarly, and
Figure 4.3 Variation of the measured distillate flux \( J \) with input wall heating \( P_{in} \) when \( \dot{V}_{in}^F = \dot{V}_{in}^D = 0.5 \text{L/min} \) and DI water with 70 g/L NaCl is used as the feed. Panel (a) shows results without a feed spacer. Panel (b) shows results with a feed spacer. In both panels, the empty blue circles show measured data for the stable orientation, and the solid red stars show data for the unstable orientation. The solid lines show linear best fits.

that the stable orientation just happened to average slightly higher. Overall, both orientations performed slightly better when a feed spacer was used. We hypothesize that the vortical flow structures generated by the feed spacer provided additional mixing in the feed. The difference, however, was surprisingly small.

Figure 4.3 shows results from when the experiments in Figure 4.1 were repeated using a feed with 70 g/L NaCl. Panel (a) shows \( J \) versus \( P_{in} \) without a feed spacer, while panel (b) shows results with a feed spacer. Looking at both sets of data, we observe that the distillate flux continues to increase linearly with \( P_{in} \), and is slightly increased with the use of a feed spacer. Now, however, we observe no impact from the cell orientation. We also note that the distillate fluxes are consistently around 3 LMH less than those obtained in Figure 4.1 for a feed of DI water. This occurs due to a decrease in partial vapor pressure as concentration increases. Overall, when no feed spacer was used, the average flux started at 10.07 LMH and increased by 3.71 LMH (37%). When a feed spacer was used, the average flux started 10.62 LMH and increased by 3.77 LMH (36%). The percent change in flux values is higher than those observed in Figure 4.1, but this is just due to the initial flux values being lower, which inflates the percent change.

None of the data shown in Figure 4.1 or Figure 4.3 showed any evidence of buoyancy driven convection. Motivated by our CFD in Chapter 2, which predicted that buoyancy driven convection preferentially occurs at low Reynold’s numbers, we then repeated our experiments using the feed and distillate flowrates \( \dot{V}_{in}^F = \dot{V}_{in}^D = 0.4, 0.3, \) and \( 0.2 \text{L/min} \). For brevity, we consider only a feed solution of 70 g/L NaCl, without a feed spacer. These conditions were chosen to maximize the opportunity for buoyancy driven convection.
Figure 4.4 Variation of the me (a) 0.4 L/min with average inlet Re = 345, stable $R^2 = 0.947$, and unstable $R^2 = 0.969$. (b) 0.3 L/min with average inlet Re = 268, stable $R^2 = 0.979$, and unstable $R^2 = 0.982$. (c) 0.2 L/min with average inlet Re = 182, stable $R^2 = 0.956$, and unstable $R^2 = 0.952$.

Figure 4.4 shows $J$ versus $P_{in}$ for (a) 0.4 L/min, (b) 0.3 L/min, and (c) 0.2 L/min, while Table 4.1 lists the starting and ending flux for each flowrate and orientation with an empty feed spacer, along with the change in flux, percent change in flux, and the linear best fit slope. From Figure 4.4, it is clear that the orientation of the cell still had no impact on flux. As the flowrate was lowered, the flux at every tested $P_{in}$ dropped, but the change in flux between $P_{in} = 0$ and 235 W remained around 3 to 4 LMH, and showed no
Table 4.1 Starting and ending flux values for each tested flowrate with a feed of DI water with 70 g/L NaCl and no feed spacer along with difference in flux, % difference in flux, and slope of linear best fit.

<table>
<thead>
<tr>
<th>Flowrate (L/min)</th>
<th>Orientation</th>
<th>Re</th>
<th>Start Flux</th>
<th>End Flux</th>
<th>Difference</th>
<th>% Difference</th>
<th>Slope</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.5</td>
<td>Stable</td>
<td>425</td>
<td>10.3</td>
<td>14.1</td>
<td>3.83</td>
<td>37.3</td>
<td>0.0155</td>
</tr>
<tr>
<td></td>
<td>Unstable</td>
<td>425</td>
<td>9.87</td>
<td>13.5</td>
<td>3.58</td>
<td>36.3</td>
<td>0.0147</td>
</tr>
<tr>
<td>0.4</td>
<td>Stable</td>
<td>340</td>
<td>8.27</td>
<td>11.2</td>
<td>2.90</td>
<td>35.1</td>
<td>0.0118</td>
</tr>
<tr>
<td></td>
<td>Unstable</td>
<td>340</td>
<td>8.42</td>
<td>11.3</td>
<td>2.85</td>
<td>33.9</td>
<td>0.0125</td>
</tr>
<tr>
<td>0.3</td>
<td>Stable</td>
<td>255</td>
<td>5.97</td>
<td>8.61</td>
<td>2.64</td>
<td>44.2</td>
<td>0.0112</td>
</tr>
<tr>
<td></td>
<td>Unstable</td>
<td>255</td>
<td>5.53</td>
<td>8.63</td>
<td>3.10</td>
<td>56.1</td>
<td>0.0134</td>
</tr>
<tr>
<td>0.2</td>
<td>Stable</td>
<td>170</td>
<td>3.08</td>
<td>7.10</td>
<td>4.02</td>
<td>131</td>
<td>0.0162</td>
</tr>
<tr>
<td></td>
<td>Unstable</td>
<td>170</td>
<td>4.56</td>
<td>7.75</td>
<td>3.19</td>
<td>70.0</td>
<td>0.0143</td>
</tr>
</tbody>
</table>

obvious trend. Overall, these results show that while actively heating the feed plate increased flux, there is no evidence of buoyancy driven convection. Instead, there is likely other mixing going on due to the geometry of the heater in the flow.

Figure 4.5 Both plots show data collected at each tested flowrate, in the unstable orientation, at the maximum value of $P_{in}$, with a feed solution of DI water and 70 g/L NaCl, and no feed spacer is used. (a) plots the percentage of heat from the surface heater going towards distillate flux, environmental losses, conduction through the membrane, and leaving the cell outlet with the feed, versus the feed flowrate when the NaCl feed solution is used, in the unstable orientation, without a feed spacer, at the maximum tested value of $P_{in}$. This plot shows that the percentage of heat that leaves the cell with the feed increases as the flowrate increases. There are no obvious trends in the environmental or conductive heat losses.
losses, or in the heat contributing to distillate flux. Panel (b) plots the average feed and distillate temperatures at the cell outlet, under the same testing conditions as panel (a). This data shows that both the feed and distillate outlet temperatures decrease as the flowrate increases. This makes sense because as the feed and distillate flowrates increase, the same amount of heat from the heater is delivered to a greater mass flow, so the temperatures do not increase as much. Panel (b) could be showing that the heater is able to replace the heat lost from the feed at low flowrates, but increased temperature polarization in the distillate channel could also be affecting these results.

4.2 Results with prototype F

With the previous set of tests, we expected the unstable orientation to produce more distillate flux than the stable orientation. We believe that several issues with prototype E inhibited buoyancy driven convection. First, the heater had an uneven surface that likely induced mixing. Prototype F consequently places the heater behind a metal plate to provide a smooth channel surface. This also solved the issue of heaters burning out. Second, the small channel width may have produced too large of a mean velocity, and associated Reynold's number. The feed channel in prototype F was therefore made wider. Third, the small channel length may have been insufficient to permit the development of convection. Our heat analysis in Figure 4.5(a) also suggests that up to 62.5% of the supplied heat was carried out of the cell by the feed. Prototype F was made longer so that there is more room for buoyancy driven instability to develop and to allow more time for heat to reach the membrane. Most importantly, we realized that there was a flaw in our initial experimental conditions that likely inhibited convection. By reducing the distillate flowrate with the feed flowrate, we were encouraging temperature polarization in the distillate channel. Our tests with prototype F consequently used the maximum distillate flowrate permitted by the system, which was kept constant to minimize temperature polarization in the distillate channel.

In this section, we present our results obtained with prototype F, which were far more successful than prototype E. We begin in section 4.2.1 by showing results using a feed solution of 70 g/L NaCl with no feed spacer. These show convincing evidence that buoyancy driven convection is taking place. Then, in section 4.2.2, we show results using a feed solution of DI water, with no feed spacer, to investigate the effects of concentration polarization on buoyancy driven convection. Next, in section 4.2.3, we show results using a feed solution of 70 g/L NaCl, with a feed spacer, to investigate the effects of vortical flow structures that the spacer creates. Finally, in section 4.2.4, we present results using a feed solution of 70 g/L NaCl, with no feed spacer, and with the feed inlet temperature reduced from 60 °C to 50 °C. In all cases, the distillate flowrate was set to $V_{d_{in}} = 1.8$ L/min, the feed flowrate was varied between $0.2 \leq V_{f_{in}} \leq 1.4$ L/min, and the heating power between $0 \leq P_{in} \leq 360$ W. The distillate flowrate was set as high as the system would allow.
in order to minimize temperature polarization in the distillate flow. A distillate flowrate of 2.0 L/min was originally tested, but the distillate chiller could not maintain the desired distillate temperature.

### 4.2.1 Results with a feed solution of 70 g/L NaCl with no feed spacer

Figure 4.6 shows our results for $J$ versus $P_{in}$ when treating a feed solution of 70 g/L NaCl without a feed spacer. Panel (a) shows results for the unstable orientation, while panel (b) shows results for the stable orientation. Results are shown for the feed flowrates $\dot{V}_{in} = 0.2, 0.4, 0.6, 0.8, 1.0, \text{ and } 1.4 \text{ L/min}$. In the unstable orientation, we consider the heating powers $P_{in} = 0, 120, 240, \text{ and } 360 \text{ W}$. In the stable orientation, however, we only consider the heating powers $P_{in} = 0, 120, \text{ and } 240 \text{ W}$, because the internal feed plate temperature exceeded $100 \degree C$ when $P_{in} = 360 \text{ W}$. Comparing panels (a) and (b), it is now clear that the unstable orientation produces greater flux at every flowrate and heating level. Table 4.2 compares all the flux values when $P_{in} = 0 \text{ W } (J_0)$ and $360 \text{ W } (J_{360})$. $J_{360}$ for the stable data is extrapolated using the linear best fits. For every flowrate, the unstable orientation had greater $J_0$ and $J_{360}$ values, a larger change in flux ($\Delta J = J_{360} - J_0$) and percent change in flux ($\% \Delta J$), and steeper linear best fit slopes than the stable orientation. This likely indicates that buoyancy driven convection occurred, because the only difference between the experiments is the orientation of gravity with respect to the flow cell. In Figure 4.6(a), we observe that for the unstable orientation, the distillate flux decreases as the feed flowrate is lowered. However, the slopes of the linear best fits increase with decreasing feed flowrates. This suggests that increasing the power of the active heating further could eventually cause flux values for lower flowrates
Table 4.2 Flux values at 0 W and 360 W for each tested flowrate and orientation along with difference in flux, % difference in flux, and slope of linear best fit.

<table>
<thead>
<tr>
<th>Flowrate</th>
<th>Orientation</th>
<th>Re</th>
<th>$J_0$</th>
<th>$J_{360}$</th>
<th>$\Delta J$</th>
<th>$% \Delta J$</th>
<th>Slope</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.4 L/min</td>
<td>Stable</td>
<td>893</td>
<td>6.85</td>
<td>7.57</td>
<td>0.72</td>
<td>10.5</td>
<td>0.0021</td>
</tr>
<tr>
<td></td>
<td>Unstable</td>
<td>893</td>
<td>8.78</td>
<td>10.6</td>
<td>1.82</td>
<td>20.7</td>
<td>0.0048</td>
</tr>
<tr>
<td>1.0 L/min</td>
<td>Stable</td>
<td>638</td>
<td>5.99</td>
<td>6.35</td>
<td>0.36</td>
<td>6.01</td>
<td>0.0011</td>
</tr>
<tr>
<td></td>
<td>Unstable</td>
<td>638</td>
<td>8.13</td>
<td>10.1</td>
<td>1.97</td>
<td>24.2</td>
<td>0.0058</td>
</tr>
<tr>
<td>0.8 L/min</td>
<td>Stable</td>
<td>510</td>
<td>5.48</td>
<td>5.68</td>
<td>0.20</td>
<td>3.65</td>
<td>0.0006</td>
</tr>
<tr>
<td></td>
<td>Unstable</td>
<td>510</td>
<td>6.95</td>
<td>9.48</td>
<td>2.53</td>
<td>36.4</td>
<td>0.0071</td>
</tr>
<tr>
<td>0.6 L/min</td>
<td>Stable</td>
<td>383</td>
<td>4.31</td>
<td>5.05</td>
<td>0.74</td>
<td>17.2</td>
<td>0.0021</td>
</tr>
<tr>
<td></td>
<td>Unstable</td>
<td>383</td>
<td>6.58</td>
<td>9.33</td>
<td>2.75</td>
<td>41.8</td>
<td>0.0076</td>
</tr>
<tr>
<td>0.4 L/min</td>
<td>Stable</td>
<td>255</td>
<td>3.42</td>
<td>4.26</td>
<td>0.84</td>
<td>24.6</td>
<td>0.0023</td>
</tr>
<tr>
<td></td>
<td>Unstable</td>
<td>255</td>
<td>5.59</td>
<td>8.66</td>
<td>3.07</td>
<td>54.9</td>
<td>0.0086</td>
</tr>
<tr>
<td>0.2 L/min</td>
<td>Stable</td>
<td>128</td>
<td>2.12</td>
<td>3.88</td>
<td>1.76</td>
<td>83.0</td>
<td>0.0049</td>
</tr>
<tr>
<td></td>
<td>Unstable</td>
<td>128</td>
<td>3.19</td>
<td>7.56</td>
<td>4.37</td>
<td>137</td>
<td>0.0120</td>
</tr>
</tbody>
</table>

to surpass those of higher flowrates. Extrapolating the linear best fit data for $\dot{V}_{in}^f = 0.2$ and 1.4 L/min in Figure 4.6(a), we predict that the flux produced for $\dot{V}_{in}^f = 0.2$ L/min would surpass that produced by $\dot{V}_{in}^f = 1.4$ L/min when $P_{in} > 794$ W. This high value, however, is likely impractical.

Figure 4.7 Plots of $\Delta J$ versus $V_{in}^f$. (a) shows the unstable orientation with a log power fit of $2.11 V_{in}^{0.453}$. (b) shows the stable orientation using extrapolated data for $J_{360}$.

Figure 4.7(a) shows the variation in flux increase $\Delta J$ with the feed flow rate $\dot{V}_{in}^f$. We see that $\Delta J$ decreases monotonically with increasing $\dot{V}_{in}^f$. A log power rule with the equation $2.11 V_{in}^{0.453}$ is fit to the data, with a maximum error of less than 8%. We offer two potential explanations for this behavior. First, increasing $\dot{V}_{in}^f$ decreases the residence time that feed spends in the cell. This reduces the time that a plume has to advect heat to the membrane. Second, our CFD simulations show that increasing the flowrate
Figure 4.8 Sketch of temperature boundary layer formation in the feed channel

pushes the onset of convection downstream.

Figure 4.7(b) shows the corresponding variation of $\Delta J$ with $V_{in}^f$ for the stable orientation. The results suggest that $\Delta J$ initially decreases with increasing $V_{in}^f$, reaching a minimum at $V_{in}^f = 0.8$ L/min. For $V_{in}^f > 0.8$ L/min however, $\Delta J$ begins to increase with $V_{in}^f$. This could potentially be explained as follows. In the absence of buoyancy driven convection, or any other sort of mixing, heat is transported from the plate to the membrane through a thermal boundary layer whose depth, $\delta$, increases with downstream distance, as sketched in Figure 4.8. There is consequently an inlet region where $\delta << h$ and the membrane is unaffected by the active heating. Increasing the feed flowrate decreases the boundary layer thickness, $\delta$, such that less heat reaches the membrane. The increase in $\Delta J$ above $V_{in}^f = 0.8$ L/min could be explained by the appearance and growth of secondary flows, vortical structures, or even turbulence. These could be generated by the flow entering the feed channel through the slot in the plate.

Figure 4.9 shows the variation of the distillate flux $J$ with feed flowrate. The blue lines show results for the stable orientation, while the red lines show results for the unstable orientation. Results are shown for $P_{in} = 0, 120, 240,$ and $360$ W. The data suggests that the flux in the unstable orientation may plateau with increasing $V_{in}^f$. The flow meter in the feed flow loop was rated for low flowrates, which limited us to a maximum $V_{in}^f$ of $1.4$ L/min. In future study, higher values of $V_{in}^f$ should be tested to see how this trend continues.

Figure 4.10(a) shows the percentage of the wall heating that contributed to distillate production. The percentage is shown as a function of the feed flow rate $V_{in}^f$. Results are shown for the stable (blue lines) and unstable (red lines) orientations. As expected, a greater percentage of the provided heat goes towards evaporation in the unstable orientation. The percentage in the unstable orientation also tends to decreases as the flowrate increases, varying from roughly $25\%$ at $0.2$ L/min, to around $12\%$ at $1.4$ L/min.

Figure 4.10(b) shows the corresponding results for the percentage of heat exiting the feed outlet. For the
stable orientation, most of the supplied heat is carried out of the cell by the feed. The percentage increases with $V_{in}^f$ from roughly 63% at 0.2 L/min to above 90% at 1.4 L/min. The percentage of heat leaving with the feed in the unstable orientation started around 27% at 0.2 L/min and generally increased with the flowrate until it began to approach similar values as the stable orientation at 1.4 L/min. The large percentage of heat leaving with the feed at higher flowrates could potentially be captured in a secondary heat recovery step or recycled by running the system in a closed circuit loop.

Figure 4.9 Plot of how flux is affected by flowrate at each tested power level.

Figure 4.10 Plots of (a) the percentage of supplied heat contributing towards distillate production and (b) the percentage of supplied heat leaving the cell in the feed. In both plots, the unstable orientation is in red and the stable orientation is in blue.
Figure 4.11 Plots comparing the flux values with and without 70 g/L NaCl in the feed for (a) 0.2 L/min and (b) 1.4 L/min with linear best fits.

4.2.2 Results with no salinity

To help us understand the compounding effects of temperature and concentration polarization on buoyancy driven convection, we ran a limited set of experiments using DI water as the feed without a feed spacer. We considered the feed flowrates 0.2 L/min and 1.4 L/min. Figure 4.11 shows the variation of $J$ with $P$ in the stable (blue) and unstable (red) orientations for a feed solution of 70 g/L NaCl (circles) and DI water (stars). Panel (a) shows results when $V_{f\text{in}} = 0.2$ L/min. Panel (b) shows results when $V_{f\text{in}} = 1.4$ L/min. At both flowrates and orientations, the feed of DI water produced greater distillate flux than the NaCl feed solution. Looking at the unstable orientation data, it is clear that the change in flux as active heating is increased is not strongly affected by the feed salinity at either flowrate. At a flowrate of 0.2 L/min, the change in flux ($\Delta J$) 0.16 LMH less when NaCl is used in the feed, while the change in flux at 1.4 L/min is almost completely unaffected by the presence of NaCl. We found these results counter-intuitive. On one hand, we expected the distillate flux to decrease with the addition of NaCl to the feed, because NaCl decreases the feed partial vapor pressure. However, we expected the NaCl to promote convection because concentration polarization would increase the feed density near the membrane. The results in Figure 4.11 suggest that temperature polarization is the primary driving force behind buoyancy driven instability.
4.2.3 Results with an industrial feed spacer

To explore how the vortical structures generated by spacers affect buoyancy-driven convection, we ran another limited set of experiments with a feed spacer at the feed flowrates $V_{f}^{in} = 0.2$ L/min and 1.0 L/min. We originally intended to test at 1.4 L/min instead of 1.0 L/min, however the distillate chiller could not maintain the desired distillate temperature. Figure 4.12 shows the variation of $J$ with $P_{in}$ in the stable (blue) and unstable (red) orientations without a feed spacer (circles) and with a feed spacer (stars).

Panel (a) shows results when $V_{f}^{in} = 0.2$ L/min. At 0.2 L/min, the distillate flux in the unstable orientation was relatively unaffected by the presence of a spacer in the feed channel. However, the spacer did improve the distillate flux in the stable orientation. This suggests that at low flowrates, buoyancy-driven instability can mix the feed just as well as a spacer could. It’s also possible that lowering the flowrate even further could make buoyancy-driven instability more efficient at mixing than a spacer. This is because lowering the flowrate should reduce the mixing caused by the spacer, while buoyancy-driven instability grows stronger at low flowrates. The stable data shows that the feed spacer does improve the cell’s performance in the stable orientation. This is expected because buoyancy-driven convection does not occur in the stable orientation. We believe that the difference in performance between the stable and unstable orientations when a feed spacer is used can be explained by the low flowrate. Since the feed has a low Reynold’s number at this flowrate, the spacer is likely unable to provide as much mixing as it would at higher flowrates. As a result, it is possible that when the cell is in the stable orientation, some of the hotter feed close to the channel wall remains close to the wall due to its lower density. Conversely, when the cell is
in the unstable orientation, the feed close to the channel wall can move up towards the membrane due to its lower density, where it then meets the bulk flow and is mixed by the spacer.

Panel (b) shows results when $V_{\text{in}} = 1.0 \text{ L/min}$. At 1.0 L/min, the cell’s performance is unaffected by the orientation when a feed spacer is used. The presence of a feed spacer also results in higher flux values than when no spacer is used in both orientations. This provides further evidence that, as the flowrate increases, the mixing caused by the spacer becomes more efficient, while buoyancy driven instability becomes less efficient. These results also strengthen our belief that the orientation of the first prototype did not have an effect on flux because mixing was being caused in the feed channel by the heater geometry.

4.2.4 Results with a lower inlet feed temperature

The final set of tests that we ran reduced the inlet feed temperature to 50 °C. We hypothesized that lowering the feed temperature would cause active heating to have a greater impact on flux. Figure 4.13 shows the variation of $J$ with $P_{\text{in}}$ in the stable (blue) and unstable (red) orientations with inlet feed temperatures of 60 °C (circles) and 50 °C (stars). (a) shows results when $V_{\text{in}} = 0.2 \text{ L/min}$. Panel (b) shows results when $V_{\text{in}} = 1.4 \text{ L/min}$. Overall, decreasing the feed inlet temperature reduced flux values and slightly lowered the flux gain from active heating. At 0.2 L/min, the change in flux, $\Delta J$, dropped by about 0.49 LMH for the stable orientation and 1.03 LMH for the unstable orientation compared to the cases run at 60 °C. At 1.4 L/min, the change in flux dropped by about 0.11 LMH for the stable orientation and 0.49 LMH for the unstable. We hypothesize that this occurred because as the temperature difference
between the feed and distillate inlets decreased, temperature polarization also decreased. This would cause the difference in local feed temperatures, and therefore local densities, near the membrane and the channel wall to decrease, which could weaken buoyancy driven instability. The lower flux values would also reduce concentration polarization, which could also weaken buoyancy driven instability. To get a better idea of how feed temperature affects active heating performance, it would be beneficial to redo these tests with a constant difference in distillate and feed inlet temperatures. For example, testing with feed and distillate temperatures of 50 °C and 20 °C or 70 °C and 40 °C.
Summary:

In this study, we developed six different potential prototypes to provide active heating in a DCMD flow cell. We began by making heated spacer designs using nichrome wires as heating elements. We quickly abandoned these designs because the nichrome began to corrode when in direct contact with the feed. We consequently focused on using off-the-shelf surface heaters which were purchased from Omega Engineering. We found that when the heater was glued to the feed plate and in direct contact with the feed, the surface of the heater produced uncontrolled mixing and large error bars. The surface heater also had a tendency to burn out after a few weeks of usage. With this design, we were unable to see any evidence that buoyancy driven convection occurred. Our final prototype improved upon the previous design by placing the surface heater behind a metal plate, so that it was no longer in direct contact with the feed. We also changed our experimental procedure so that the inlet distillate flowrate was kept constant at 1.8 L/min to minimize distillate temperature polarization. This prototype, referred to as prototype F, produced promising results and was the focus of our remaining experimentation.

In our experiments with prototype F, we saw clear evidence of buoyancy driven convection. Compared to the stable orientation, the unstable orientation produced greater flux at every flowrate and heating level. The unstable orientation also had a larger change in flux ($\Delta J = J_{360} - J_0$), percent change in flux ($\% \Delta J$), and steeper linear best fit slopes than the stable orientation. Overall, these results showed that the flux increased with flowrate, while the change in flux, $\Delta J$, decreased at higher flowrates. This meant that a greater percentage of the heat provided by the surface heater is able to contribute to distillate flux at lower flowrates. At a feed flowrate of 0.2 L/min, around 25% of the supplied heat went towards distillate flux, while only 12% contributed to distillate flux at 1.4 L/min.

Future Work:

During our prototyping phase, we identified two potential approaches to wall heating that deserve future study. The first approach involved gluing nichrome wires to a sheet of PTFE, which could then be attached to the feed plate. This design can be easily customized to adjust the orientation, spacing, and diameter of the wires. The wires can also provide texture which could enhance mixing. This design showed promise, but there were too many parameters that we would need to vary when testing. This design opens the potential for optimal combinations of these parameters, such that distillate production is maximized while heat input is minimized.
The second proposed approach involved 3D printing the feed side plate out of stainless steel. The plate would be printed with an integrated channel that carries a heat transfer fluid to provide active heating. The surface of the cell could also be textured, such as with fins, to improve heat transfer and mixing. Using a heat transfer fluid, instead of resistive Joule heating, would facilitate the use of renewable heat sources, like solar or geothermal, without needing additional infrastructure, such as solar panels, to convert the heat to electricity. However, the additional pumps that would be required to circulate the heating fluid, and the additional flow channels for the heating fluid, would reduce how tightly the membranes could be packed together and would make the system more complex. Additionally, 3D printing these cells out of stainless steel is expensive and time consuming.

Our tests with prototype F showed that lower feed flowrates produced lower distillate flux, but more efficiently used the heat provided by the surface heater. In future study, the total energy demand of the system should be measured so that we can calculate the specific energy consumption and determine the total efficiency of the system. Due to the limitations of our system, the maximum feed flowrate that we used was 1.4 L/min. It would also be beneficial to test at higher flowrates in future studies to see how trends in distillate flux continue. Furthermore, this design should be tested in a counter-current configuration, so that its performance can be compared to our co-current data.

All of the experiments that we ran used a feed of either DI water or a solution of DI water with 70 g/L NaCl, neither of which could cause membrane scaling. Mineral scaling is a major issue in membrane distillation and is one of the factors that bars it from industrial adoption. Future experiments should investigate how active heating and buoyancy driven convection affect mineral scaling. Two commonly tested minerals that have a tendency to scale are gypsum and silica. It would be best to perform scaling experiments with each of these compounds.

In future study, prototype F should be tested in VMD so that its performance can be compared to our DCMD data. VMD is a promising candidate because of its low conductive losses. We focused on DCMD in this study due to its simplicity, and because our lab had ample experience in the design and operation of DCMD systems.

Finally, a longer version of prototype F should be manufactured and tested. The length of DCMD cells are typically limited due to temperature polarization in the feed and distillate flows. As the transmembrane temperature difference decreases down the length of the cell, so too does the local distillate flux. Because of this, we expect the benefits of active heating to increase with the cell length.
REFERENCES


