BREATHABLE MEMBRANE TECHNOLOGY FOR DRYING WASTEWATER TREATMENT PLANT SLUDGE

by

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ABSTRACT

Drying of wastewater treatment plant sludge enclosed in hydrophobic breathable membrane was investigated for the potential application of breathable membrane for dewatering sludge. Loss of moisture from the membrane-enclosed sludge was studied using a cylindrical drying module that was composed of two chambers separated by a breathable membrane sheet. The top compartment contained the sludge while the sweeping air was supplied to the bottom compartment. Mass loss from the sludge was examined under two sludge heating strategies: direct heating of drying module in an oven and heating sweeping air. Sludge temperatures and flow rate of sweeping air were varied under both conditions to determine the optimum operation conditions for sludge drying. Drying test with the oven heating of drying module showed that vapor flux through the membrane increased from 0.13 to 2.4 g/cm^2/hr as the oven temperature increased from 80 to 120 ^{O}C. Similarly, increasing the sweeping air temperature from 25 to 100 ^{0}C increased vapor flux from 0.05 to 0.19 g/cm^2/hr. Increasing flow rate of sweeping air also increased vapor flux, but changing sludge thickness did not affect the sludge drying rate. As expected, direct heating of sludge in the oven resulted in higher vapor flux than drying with heated sweeping air. This difference may be attributed to greater temperature gradient across the membrane between the feed and permeate side. However, heating of sweeping air may be more efficient way of raising the sludge temperature at the membrane interface while achieving comparable drying rates. Furthermore, drying tests suggest that effective drying of from membrane-enclosed sludge depends on whether or not sludge is in

contact with the membrane. COMSOL simulation program was used to predict vapor flux through breathable membrane using the same experimental conditions. The simulation model was calibrated with the experimentally derived model parameters. Computer simulation results closely matched the experimental results from various sweeping air temperatures, sweeping air flow rates, and sludge thickness. This model was used to estimate a full-scale dimension of membrane-based drying system for a typical sludge dewatering plant.

Chapter 1

INTRODUCTION

Before 1972, many cities in U.S. disposed solid and liquid wastes into the ocean and freshwaters without treatment (Elliot and O'Connor, 2007). Discharges of municipal and industrials wastewater without treatment to rivers, lakes, and oceans created serious environmental consequences. Therefore, in 1972, US Environmental Protection Agency (EPA) established yhe Clean Water Act to control water pollution through establishing discharge standards for point-source discharges.

Currently, domestic, commercial, and industrial wastewater are transported through network of sewers to wastewater treatment plants. Wastewater treatment plants typically consist of (1) screening and grit removal (2) primary treatment (3) secondary treatment and (4) tertiary treatment. Throughout these processes, soluble organic and inorganic matter are removed to reduce their detrimental impact on receiving waters. When wastewater treatment process is completed, a semi solid, nutrient rich byproduct called biosolids is produced.

In U.S.A., about 8 million tons of dried biosolid are produced each year, of which about 55% are disposed by land application, landfill cover, or composting (Peccia and Westerhoff, 2015). Biosolids are produced from primary and secondary wastewater treatment processes, and contain approximately 97% water. Wet biosolids are typically further dewatered to reduce the volume.

One of the biggest challenges of biosolids management is dewatering sludge. There are also several types of dewatering methods, which include mechanical processes such as vacuum filters, plate-and-frame filter presses, centrifuges, and belt filter presses (EPA, 1999), air drying, and thermal processes. One of major drawbacks of these processes is the use of high amounts of polymers to increase sludge thickening. The mechanical drying approaches typically require polymer addition or chemical conditioning with polyelectrolytes, Fe (III), Fe (II), lime, or Al(III) (Chen et al., 2006). These polymers are used in high amounts, which lead to high contamination levels in biosolids in addition to being extremely costly. In addition, both mechanical and thermal processes are energy-intensive processes. Air drying process does not require polymers but it requires a large land area. Additionally, these natural drying methods generate odors that may be nuisance to the communities adjacent to treatment plants (Gruter et al., 1990).

An alternative method for dewatering and drying the sludge may be an application of breathable hydrophobic membrane, which prevents the transport of water vapor but allows the diffusion of water vapor. Hydrophobic breathable membrane is made from polytetrafluorethylene (PTFE). The hydrophobic nature of membrane prevents water molecules from passing through; however, it allows water vapor molecules to transport from the feed side to permeate side (Saxena, 2017). This process is similar to membrane distillation process and the vapor transport is driven by vapor pressure gradients, typically caused by a temperature difference (Gryta, 2011). For example, when a warm process water (feed) and a cold distillate (permeate) are separated by a hydrophobic membrane, the vapor pressure gradient resulting from the temperature difference causes the vapor to diffuse through the membrane and condense as cold distillate in the permeate side. The membrane distillation process has been shown to be effective under relatively low temperature and low-pressure conditions (e.g., atmospheric pressure), thus more energy efficient than conventional distillation techniques or high-pressure membrane technologies (Chung et al., 2014). Marzooghi et al. (2017) recently showed that breathable membrane technique was able to reduce the moisture content of anaerobically-digested biosolids from 97% to 12-30% under moderate temperature gradients.

The overall goal of this research presented in this thesis is to evaluate the application of breathable membranes technology for dewatering and drying wastewater treatment plant biosolids. This thesis focuses on the effect of various operational parameters on vapor flux through the hydrophobic breathable membrane, thus to gain an insight into parameters that give the optimum vapor flux. The specific objectives of this thesis are as follows:

1- Examine the effect of temperature and sweeping air flow on vapor flux across the membrane under constant heating of the biosolids.
2- Evaluate the feasibility of heating the sweeping air for drying of membrane-enclosed biosolids. The effect of temperature, sweeping air flow rate, and sludge thickness on the vapor flux through the membrane were examined.
3- Develop a drying rate model using COMSOL Multiphysics program and validate the developed model with the experimental data.
4- Design a simple membrane-based sludge dewatering apparatus and simulate the effects of operational parameters on performance of the drying apparatus using COMSOL program.

Chapter 2

LITERATURE REVIEW

2.1 Water Distribution within Sludge

Sludge from municipal wastewater treatment plants consists mainly of microbial cells (both live and dead) and water (Chen et al., 2006). Water molecules associated with solid particles inside sludge can affect the water vapor pressure, rate of water evaporation, and enthalpy. To understand this concept, it is important to study the manner and degree of association of water molecules within solid molecules (Katsiris et al., 1987).

Figure 2.1 shows that water molecules inside sludge can be categorized into four parts: (1) free water, (2) interstitial water, (3) surface water, and (4) bound water. First, free water molecules are not attached to solid particles and they can be easily removed from sludge by gravitational settling. Second, interstitial water molecules are formed due to capillary forces and strong mechanical forces are needed to remove these water molecules. Third, surface water consists of water molecules attached to solid molecules by adsorption and adhesion. Fourth, bound water molecules are intracellular water and water molecules chemically bound to solids. (Tsang and Vesilind, 1990; Chenet al., 2006).

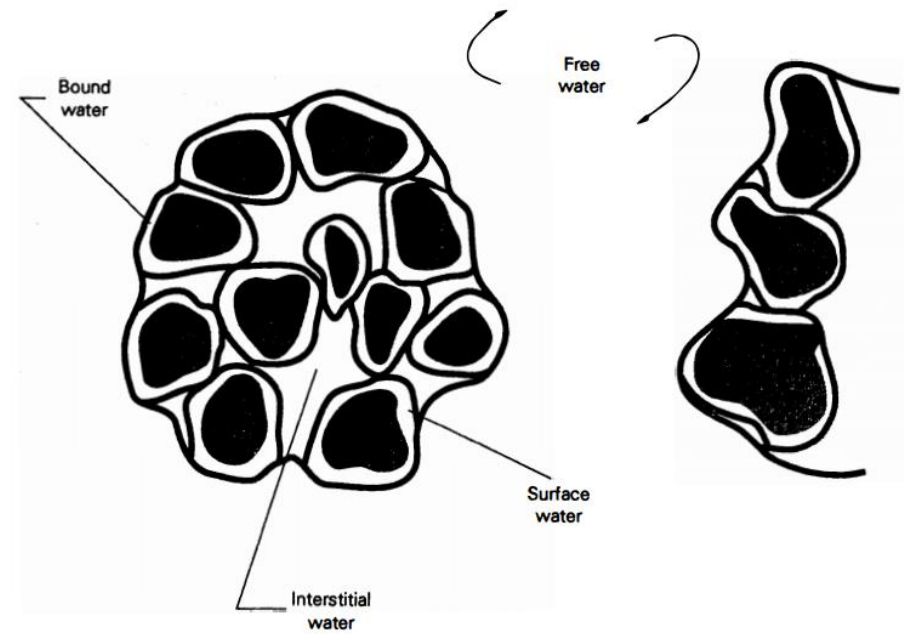


Figure 2.1 Schematic representation of kinds of water molecules within sludge (Tsang and Vesilind, 1990)

Both free water and interstitial water can be removed by traditional processes such as belt press or centrifuge. However, the remaining “surface” water is not removed by these conventional processes and requires evaporative processes to remove these strongly bound water molecules. In order to design and operate an optimum drying process for sludge drying, it is important to understand the factors that control the rate of evaporation. Factors include temperature of sludge, temperature of air flow, relative humidity, air velocity, and relative geometrical arrangement of solids within sludge. Internal factors include the chemical and physical characteristics of solid particles within sludge.

Sherwood (1936) and Coackley (1962) elucidate the general drying process based on three distinct phases: constant rate drying period, first falling rate period, and second falling rate period. Constant rate phase is the initial drying phase where evaporation of free water occurs. During the constant rate phase, the evaporation rate is not dependent on sludge properties such as moisture content, and the rate is limited by the diffusion of water from the interior of the biosolid to the surface. Thus, the drying rate will remain constant as long as the free water is present in the biosolids matrix and external factors, such relative humidity, ambient temperature, and the media diffusivity are constant. This process will reach falling rate period at the first critical moisture content, when the rate of moisture evaporation starts to decrease.

The first falling rate occurs after the evaporation of free water during constant rate period (Tsang and Vesilind, 1990). During this period, the sludge appears to be dry as free water is not present at the sludge-gas interphase. Moisture loss during this period is primarily due to the removal of interstitial water and surface water. The rate of falling rate period affected by air temperature and humidity, air velocity, or the feed temperature (Sherwood, 1936). In addition, resistance to moisture transfer within sludge matrix becomes important during this phase.

The second falling rate period occurs when the surface water is completely evaporated and the bound water is finally removed during this phase (Tsang and Vesilind, 1990). Thus, resistance to internal diffusion of water molecules is more important and the factors that affect the transport of water molecules within the solid matrix are affecting this period.

2.2 Mass Transfer through Breathable Membrane

Mass transport of water vapor molecules through a membrane is divided into three parts: 1) mass transfer within feed boundary layer, 2) mass transfer through the laminate, and 3) mass transfer within the boundary layer of permeate side (Srisurichan et al., 2006). Mass transfer within permeate side is neglected because the mole fraction of vapor molecules transported to permeate side is approximately equal to 1 (Srisurichan et al., 2006).

Film theory explains the transport in the boundary layers and Dusty-Gas model describes the mass transfer through the pores within the membrane (Srisurichan et al., 2006). The simplest form of Dusty-Gas model is:

$$J = C\ (P_f - P_p) \qquad 2\text{-}1$$

Where J is the flux through the membrane, C is membrane mass transfer coefficient ($L/m^2/hour$), which is approximately constant and dependent on the temperature, P_f is vapor pressure on the feed side, and P_p is vapor pressure on the permeate side (Schofield et al., 1987). Chiam and Sarbatly (2014), reported that

$$C \propto \frac{d^a \epsilon}{Tb} \qquad 2\text{-}2$$

where d is the mean pore diameter of the membrane (m), a is the exponential coefficient, ϵ is the membrane porosity, T is the conductive heat loss through the membrane, and b is the membrane thickness.

Furthermore, Dusty-Gas model shows that the transport of vapor molecules through the membrane is based on four mechanisms (Srisurichan et al., 2006): viscous flow, Knudsen diffusion, molecular diffusion, and surface diffusion. Mason and

Malinauskas (1983), Marzooghi et al. (2017), and Srisurichan et al. (2006) show that the formula traditionally used is molecular diffusion, Knudsen diffusion or both. Knudsen diffusion model is used when collision between the pore wall and molecules dominates the mass transfer through the membrane. The equation is:

$$N = \frac{1}{RT}\frac{2\varepsilon}{3\tau}\left(\frac{8RT}{\pi M_i}\right)^{\frac{1}{2}}\frac{(p_1 - p_2)}{\delta} \qquad 2\text{-}3$$

Where N is the molar flux (mol/m^2/hr), R is ideal gas constant (8.314 Pa m^3/mol K), T is temperature (K), ε is porosity of the membrane, τ is tortuosity, M_i is molecular weight of vapor molecules, P is the total pressure (Pa), δ is the membrane thickness (m), p_1 is the vapor pressure at feed membrane surface (Pa), p_2 is the vapor pressure at permeate side (Pa).

Molecular diffusion model is used when collision between molecular-molecular dominates the mass transfer through the membrane. The equation is:

$$N = \frac{PD_{ij}}{RT}\frac{\varepsilon}{\delta\tau}\frac{(p1 - p2)}{|P_a|_{ln}} \qquad 2\text{-}4$$

When both molecular diffusion and Knudsen diffusion happened frequently, the model used is the Knudsen-molecular diffusion transition model. The equation can be expressed as:

$$N = \frac{PD_{ij}}{RT}\frac{\varepsilon}{\delta\tau}\ln\left(\frac{P_a^2\frac{2r}{3}\left(\frac{8RT}{\pi M_i}\right)^{\frac{1}{2}} + PD_{ij}}{P_a^1\frac{2r}{3}\left(\frac{8RT}{\pi M_i}\right)^{\frac{1}{2}} + PD_{ij}}\right) \qquad 2\text{-}5$$

P_a^1 is the partial pressure of air at feed membrane surface, P_a^2 is the partial pressure of air at permeate side (Pa), D_{ij} is the diffusivity (m^2/s), and r is the nominal pore size (m).

Srisurichan et al. (2006) used all three mechanisms and compare them with experimental work. They validate that Knudsen-molecular diffusion model was the best for modeling the vapor transport through the membrane. Alkhudhiri et al. (2011) suggested that Knudsun-molecular diffusion may be applicable if pore size is sub-micron. Marzooghi et al. (2017) also suggested Knudsen-molecular diffusion model but the error in mass transfer coefficient would be less than 4% of Knudsen diffusion model.

Therefore, Marzooghi et al. (2017) used molecular-diffusion model, Stagnant film model, to describe the diffusion of water vapor through the laminated hydrophobic membrane:

$$J = \frac{P}{R(T_{ave} + 273.15)} \frac{D_{eff}}{\delta_l} \ln\left(\frac{P - P_{A1}}{P - P_{A2}}\right) \qquad 2\text{-}6$$

Where P is the atmospheric pressure, R is the ideal gas constant, T_{ave} is the average temperature across the membrane, P_{A1} is the vapor pressure on the feed side of the membrane, P_{A2} is the vapor pressure on air side of the membrane, δ_l is the membrane thickness, and D_{eff} is diffusivity of water vapor through the membrane. The value of D_{eff} can be determined using the following equation:

$$D_{eff} = D_{AB}\varepsilon/\tau \qquad 2\text{-}7$$

where ε is the porosity of the membrane, and τ is the tortuosity of the membrane. D_{AB} is the diffusivity of water vapor through air and can be expressed as follows (Gibson, 2000):

$$D_{AB} = (2.23 \times 10^{-5})\left(\frac{Tave + 273.15}{273.15}\right)^{1.75} \qquad 2\text{-}8$$

2.3 Heat Transfer through Hydrophobic Breathable Membrane

Vapor flux through the membrane was examined with direct heating of the sludge and heating the sweeping air. In both studies, the sludge was placed in the top compartment of the membrane module to maintain the sludge-membrane contact throughout the drying cycle. The constant flow of sweeping air was provided in the bottom compartment of the membrane module. The schematics of direct heating and sweeping air heating processes are shown in Figures 2.2 and 2.3, respectively.

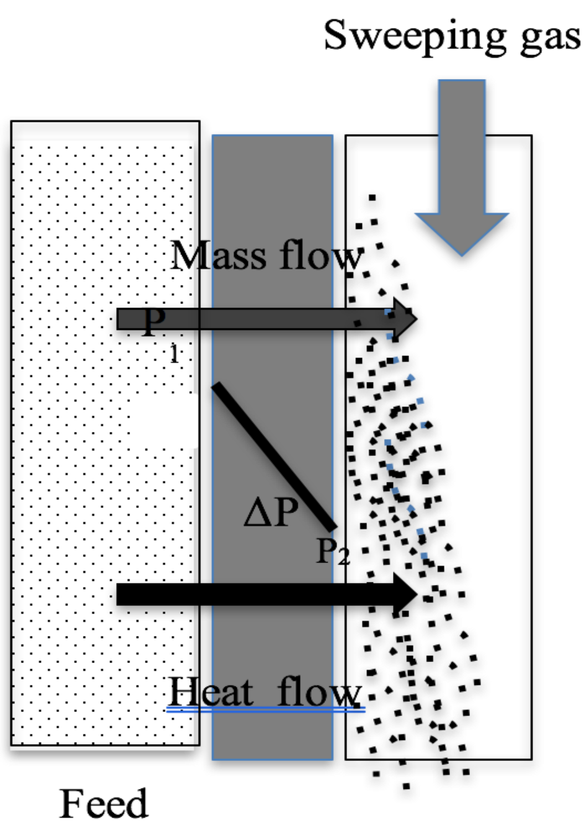


Figure 2.2 Schematic of sludge drying by direct heating of the membrane module

For drying the sludge by direct heating of the membrane module, there are two possible mechanisms (Schofield et al., 1987):

1- Heat conducts across the membrane from the feed side (sludge) to permeate side which can be expressed by the following equation:

$$Q_c = \left(\frac{k_m}{\delta}\right)(T_1 - T_2) \qquad 2\text{-}9$$

Where k_m is the effective thermal conductivity of the membrane.

2- Heat transfers from the feed side to permeate side due to the vapor flux, which can be expressed by the following conductive heat flux equation:

$$Q_v = C\frac{dP}{dT}\delta\ (T_1 - T_2) \qquad 2\text{-}10$$

Thus, the total amount of heat flux through the membrane is:

$$\mathrm{Q} = Q_c + Q_v \qquad 2\text{-}11$$

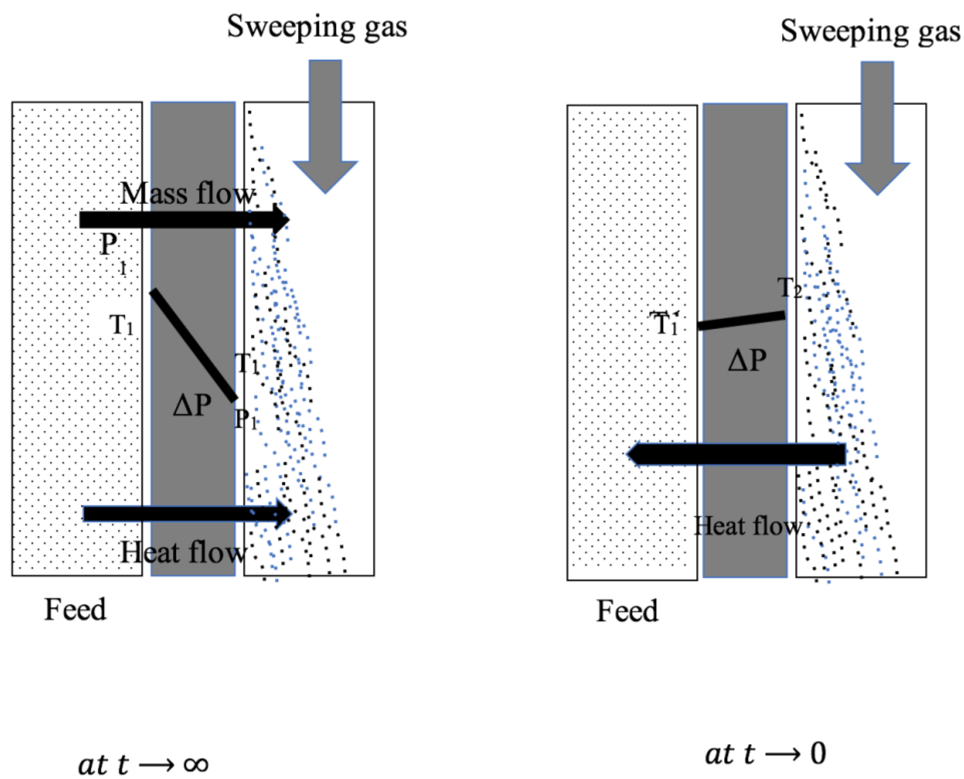


Figure 2.3 Schematic of sludge drying by sweeping air heating of the membrane module.

For heating the sweeping air, the theory of heat transfer applies to both heating the sludge and heating the sweeping air as below:

1- Initially, heat of the sweeping air conducts across the membrane from the permeate side to feed side, which can be expressed by the following equation (in this case T_2 is greater than T_1):

$$Q_c = \left(\frac{k_m}{\delta}\right)(T_1 - T_2) \qquad 2\text{-}12$$

Where k_m is the effective thermal conductivity of the membrane.

2- At t > zero, since the water molecules have higher heat capacity than air, heat moves from water to air, thus, T_1 becomes greater than T_2.

As time approaches infinity, heat transfer from the feed side to the permeate side occurs due to the vapor flux, which can be expressed by the following conductive heat flux equation:

$$Q_v = C\frac{dP}{dT}\delta\,(T_1 - T_2) \qquad 2\text{-}13$$

Thus, the total amount of heat flux through the membrane is:

$$Q = Q_c + Q_v \qquad 2\text{-}14$$

2.4 Modeling of Sludge Drying BY COMSOL Multiphysics Simulation Program

COMSOL is a multiphysics simulation program that can model complex processes. It is a unique-user friendly environment. COMSOL Multiphysics can provide many features, such as heat transfer, fluid flow, and equation based modeling in order to effectively analyze interaction between two or more physics domains at the same time. Drying by breathable, hydrophobic membrane is Multiphysics phenomena. It can be easily solved by COMSOL Multiphysics program which makes prototype for this process. It also accomplishes considerable savings in the development process. Therefore, in order to model this process, COMSOL Multiphysics was given the same inputs as those used in the experimental conditions.

The geometry of model drying system is shown in Figure 2.4. It is a two-dimensional system that contains the same configuration of membrane module used in the drying experiments.

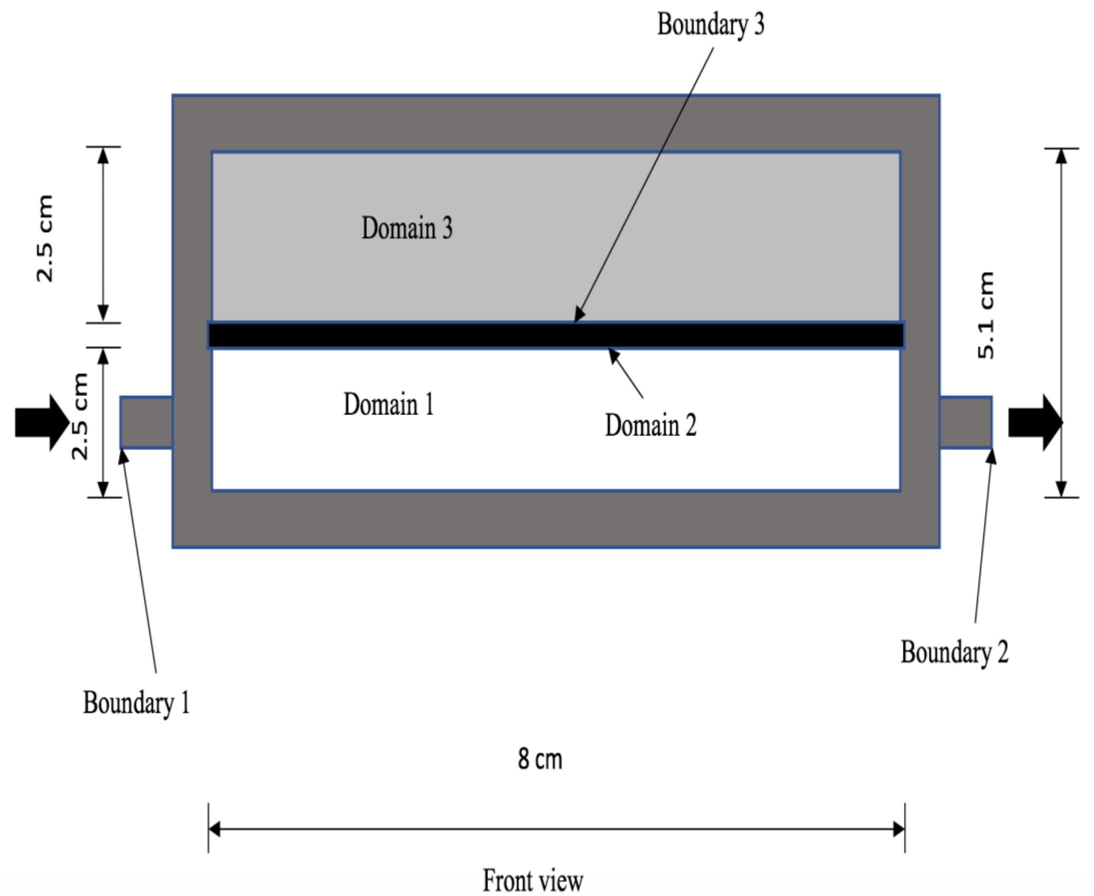


Figure 2.4 Schematic of membrane drying module for simulation by COMSOL program (Domain 1=air; Domain 2=membrane; Domain 3 = water).

Due to lack of default values of the membrane parameters, such as tortuosity, membrane diffusivity, and porosity in COMSOL, the program assumed the values of air for the membrane. Because of this, COMSOL uses diffusivity of vapor through the air, which can be calculated by the following equation (2-8) (Gibson, 2000):

$$D_{AB}(Tave) = (2.23 \times 10^{-5})\left(\frac{Tave + 273.15}{273.15}\right)^{1.75} \qquad 2\text{-}8$$

$$D_{eff} = D_{AB}\varepsilon/\tau \qquad 2\text{-}7$$

Where $D_{AB}(Tave)$ is the diffusivity of water vapor through the air (m^2/sec) and D_{eff} is the diffusivity of water vapor through the membrane (m^2/sec). $D_{AB}(Tave)$ is adjusted inside COMSOL by increasing the thickness of membrane in order to be close to D_{eff}. Three physical processes that were applied to the model system were heat transfer in fluids, transport of diluted species, and turbulent flow. Additionally, there was a multiphysics configuration that combined heat transfer of fluids and turbulent flow configuration. Below is the procedure and theory of each.

2.4.1 Heat Transfer of Fluids

The heat transfer inside the membrane module for all three domains was due to convective heat from turbulent flow of heated air through the membrane module. Thus, the governing equation that is used for heat transfer through the membrane module is expressed as:

$$d_z \rho C_p\left(\frac{\partial T}{\partial t} + \boldsymbol{u} \cdot \nabla \mathrm{T}\right) = \nabla \cdot (d_z\, k\, \nabla \mathrm{T}) + d_z \boldsymbol{Q} \qquad 2\text{-}15$$

Where ρ is the density, C_p is specific heat capacity, T is absolute temperature, u is the velocity vector, Q is heat source, k is fluid thermal conductivity, and d_z thickness of the membrane.

There were few boundaries assigned to the membrane module. The most important boundaries were inlet boundary, outlet boundary, and the boundary between sludge and membrane domains. Temperature of inlet air flow was an input, which varies from 25 to 100 ^{o}C (the same as the experimental conditions). Also, the air outlet of the membrane module was assigned as outflow boundary for the COMSOL simulations. The outflow governing equation is:

$$-\boldsymbol{n} \cdot (k\, \nabla \mathrm{T}) = 0 \qquad \text{2-16}$$

Where n is the normal vector of the boundary. Also, there are no heat transfer through the outside boundaries of the membrane module.

During the drying process, heat was transferred from the membrane to the boundary of the sludge. Thus, this boundary defined the latent heat source absorbed and released from the surface of water through the membrane sheet. The equation of this boundary is:

$$-\boldsymbol{n} \cdot \boldsymbol{q} = d_z\, Q_b \qquad \text{2-17}$$

$$Q_b = H_{vap} \times tds.ndflux_c \qquad \text{2-18}$$

where H_{vap} is the latent heat of vaporization.

2.4.2 Turbulent Flow

The air flow was modeled as turbulent flow because Reynold number of air flow in the laboratory studies ranged from slightly less than 2,000 to 4,000. Turbulent flow model used because turbulent effects needed to be considered. It was assumed that velocity and pressure field were independent of the air temperature and moisture content. In other words, turbulent flow was assigned to domain 1, which was air, and independent of domain 2 and domain 3. This procedure allowed us to determine the turbulent flow field and then use it as input for water vapor molecules transport and heat transfer. The governing equation of the turbulent flow is Navier-Stokes equation expressed as:

$$\left(\frac{\partial \boldsymbol{u}}{\partial t} + \boldsymbol{u} \cdot \nabla \mathbf{u}\right) = -\nabla \mathrm{p} + \nabla \cdot \left(\mu(\nabla \boldsymbol{u} + (\nabla \boldsymbol{u})^{T}) - \frac{2}{3}\mu(\nabla . \boldsymbol{u})\boldsymbol{I}\right) + \boldsymbol{F} \qquad \text{2-19}$$

Where u is the fluid velocity, p is the fluid pressure, I is the identity matrix, ρ is the fluid density, and μ is the fluid dynamic viscosity.

This equation is solved with the continuity equation and is expressed as:

$$\frac{\partial \rho}{\partial t} + \nabla \cdot (\rho \boldsymbol{u}) = 0 \qquad \text{2-20}$$

The inlet boundary was expressed as:

$$u = -u_0 \boldsymbol{n} \qquad \text{2-21}$$

$$\nabla G . \boldsymbol{n} = 0 \quad 2\text{-}22$$

Where u_0 is the inlet velocity. The experimental inlet was ranging from 5 to 20 liters per minute, thus the velocity was calculated by dividing the flow rate by the area of the inlet tube.

2.4.3 Transport of Diluted Species

To obtain the amount of water vapor from the membrane-sludge boundary layer, the transport of diluted species physics inside COMSOL features was used for the air and membrane domains. Applying transport of diluted species indicates that water evaporates at the membrane-water layer and then moves by diffusion through the air domain. The governing equation is expressed as:

$$\frac{\partial C_i}{\partial t} + \nabla \cdot (-D_i \nabla C_i) + \boldsymbol{u} \cdot \nabla C_i = R_i \quad 2\text{-}23$$

$$\boldsymbol{N}_i = -D_i \nabla C_i + \boldsymbol{u}\, C_i \quad 2\text{-}24$$

Where N_i is molar flux through the membrane, C_i is the concentration of vapor, u is field velocity, and R_i is the reaction, which is equal to zero when no reaction occurs. D_i is the diffusivity, which can be calculated by equation 2-8 (Gibson, 2000).

There are two important boundaries assigned for modeling the drying process in the membrane module: the inlet and the gap between the sludge and membrane. The equations for these two boundaries are:

$$C_i = C_{0,j} \quad 2\text{-}25$$

$$-n. -D_i \nabla C_i = 0 \quad 2\text{-}26$$

For the inlet boundary, $C_{0,j}$ is the initial value of relative humidity of moist air. These values were obtained from experimental measurements (typically 33% for most of the experiment). The equation of $C_{0,j}$ is expressed as:

$$C_{0,j} = \frac{ht.fluid1.psat(T)}{R_{constant} * T} = C_{sat} \qquad 2\text{-}27$$

Chapter 3

MATERIAL AND METHODS

3.1 Membrane

The three-layered eVent membrane laminate used for this study was purchased from CLARCOR Industrial Air (Overland Park, KS). This laminate contains a gas permeable expanded polytetrafluoroethylene (ePTFE) membrane that is hydrophobic. The membrane is supported by a hydrophobic fabric on the sludge side and a hydrophilic mesh-like backing fabric on the air side.

3.2 Sludge

The activated sludge samples were obtained from Wilmington Wastewater Treatment plant (Wilmington, Delaware, USA) and Elkton Wastewater Treatment Plant (Elkton, Maryland). The sludge samples were thickened by gravity for 24 hours and stored in a refrigerator at 4 ± 1 °C till used in the drying experiments.

3.2.1 Sludge Heating by Oven

Aluminum cylindrical drying cell was designed for sludge and deionized water drying as shown in Figure 3.1. Drying cell was composed of the feed and permeate chambers separated by the membrane sheet (membrane surface area = 5 x 10^{-3} m^2). The top compartment was filled with feed solution. After sludge or deionized water was loaded into the feed chamber, the assembled drying cell was placed in the oven for the water drying test. Ambient air was used as the sweeping gas and was supplied to the permeate chamber (bottom chamber) using the feed air pump. The air flowrate was controlled by a flowmeter (Dwyer RMA-26 with or RMA-23-SSV, Dwyer Instruments Inc., USA) with an accuracy of ± 4%. The temperature and relative humidity (RH) values of the sweeping air were measured at the inlet and the outlet of the membrane module (T_{in}, T_{out}, RH_{in}, and RH_{out}) using a humidity and temperature data logger (EL-USB-2, MicroDAQ.com, Ltd. USA) with an accuracy of ± 0.5 °C and ± 0.5% RH, respectively. Temperature values recorded at the center of the feed solution and locations close to the membrane surface on both chambers were recorded by Pt 100, K-type thermocouples connected to a thermometer (Omega RDXL4SD) with an accuracy/precision of ± 0.1 °C. The thermocouples were inserted in the center of the module vertically. To avoid frequent removal of the module from the oven for weighing during the drying period, the module was suspended from a precision electronic balance (XP4002S, Mettler Toledo, Switzerland) with a resolution of 0.01g. The mass data from the balance was automatically recorded by a computer at 1 min intervals.

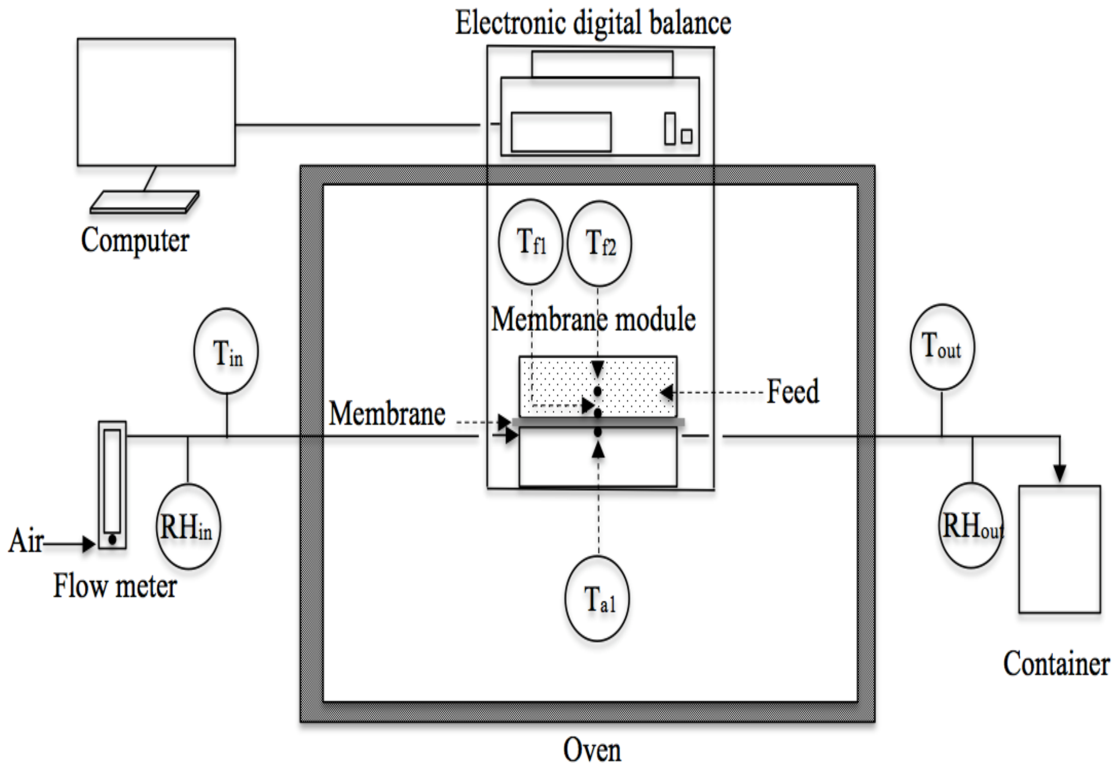
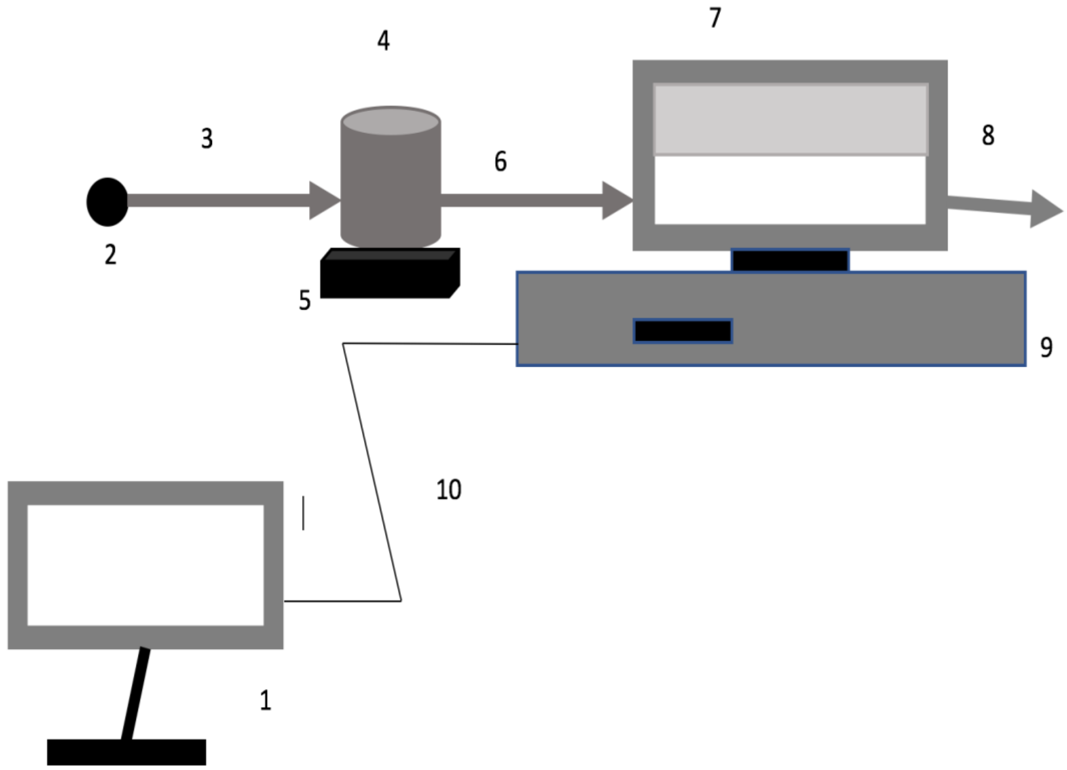


Figure 3.1 Schematic view of sludge drying apparatus when heating the sludge.

Prior to running the experiment, the preheated sludge ran for about two hours to acquire a steady state condition of mass and heat transfer. These experiments were conducted with distilled water and activated sludge at 80, 100 and 120 °C. The flow rate of the sweeping air through the membrane module varied from 2.5 to 30 L/min.

3.2.2 Sludge Drying by Heated Sweeping Air

Instead of heating the drying module in the oven, the drying experiments were conducted under an ambient condition with heated sweeping air. The same aluminum cylindrical drying cell (Figures 3.2 and 3.3) was used for the sweeping air-drying studies.



(1) Computer, (2) source of air, (3) cold air, (4) oil bath, (5) hot plate, (6) hot air, (7) membrane module, (8) disposal of air, (9) smart scale, (10) cable that connect the smart scale with computer.

Figure 3.2 Schematics of sludge drying apparatus to measure decreasing of sludge weighing in order to measure vapor flux through breathable hydrophobic membrane

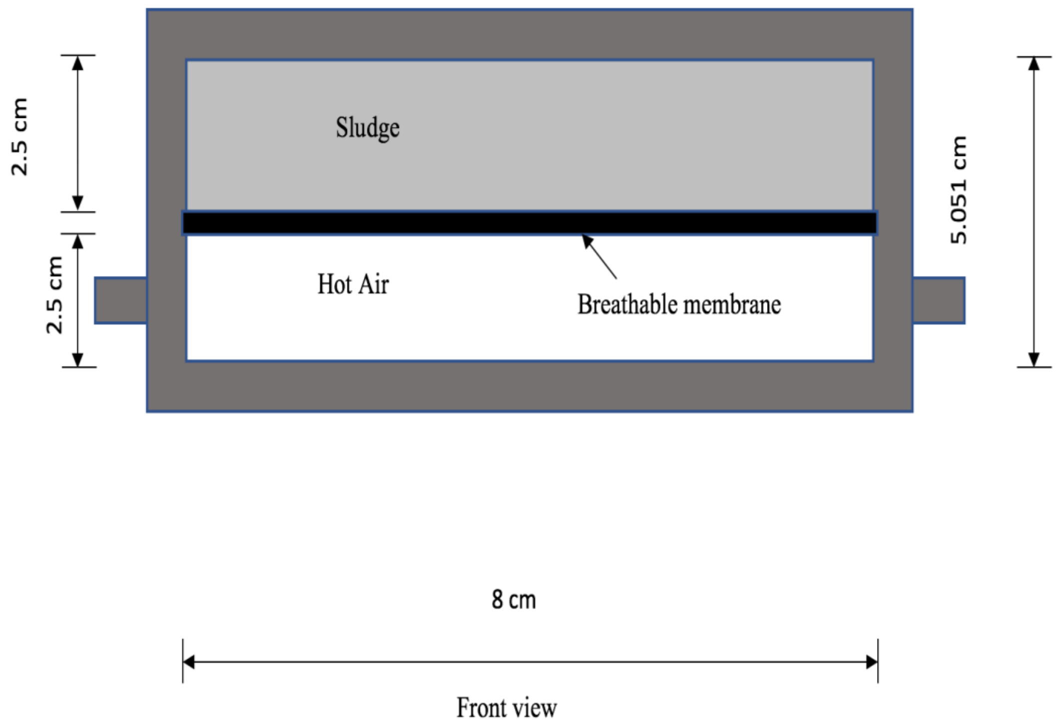


Figure 3.3 Schematic view of cylindrical sludge module. Sludge is on top, hot air on the bottom, and membrane in between sludge and membrane. Membrane-sludge interface is hydrophobic side and membrane-hot air interface is hydrophilic side

The sludge was placed in the top side of the module to keep the sludge in touch with the membrane. Heated air was supplied to the bottom side with a copper tubing. Inlet air stream was continuously heated by submerging the copper tube in the oil bath. The sludge module was placed on a precision electronic balance (XP4002S, Mettler Toledo, Switzerland) with a resolution of 0. 01g.The scale was connected to a computer to obtain mass data at 0.5 sec intervals. Temperatures of sweeping air were varied at 25, 50, 75, and 100 ℃ with flow rates of 5, 10, 20 L/min. Drying experiments were conducted with three different sludge volumes: 40, 80, 120 mL. Relative humidity and temperature of inlet air flow and outlet air flow was monitored using a humidity and temperature data logger (EL-USB-2, MicroDAQ.com, Ltd. USA) with an accuracy of ± 0.5 °C and ± 0.5% RH, respectively. Temperatures of sludge layer were measured at various depth with Pt 100, K-type thermocouples connected to a thermometer (Omega RDXL4SD). The thermocouples were inserted in the center of the module vertically.

Chapter 4

RESULTS AND DISCUSSION

4.1 Oven Heating of Drying Module

Oven heating of entire drying module with cooling air into the permeate side was tested to improve the vapor flux of free water due to the temperature and vapor pressure gradients. Figure 4.1 shows the effects of oven heating on the vapor flux of deionized water. The vapor flux was increased from 2.5 to 4.1 $kg/m^2/hr$ in the presence of membrane between the feed and permeate sides, but the membrane interfered the interaction between free water vapor and feed cooling air. When cooling air was directly supplied on the surface of feed water without membrane, the evaporation rate was substantially higher at about 10 $kg/m^2/hr$ and drying was completed within 4 hours. The effect of direct contact of water and membrane was tested by flipping over the drying module. The configuration allowed the surface of membrane to be in contact with water throughout the drying experiment. The vapor flux of the direct contact system was higher (about 6 $kg/m^2/hr$) than the non-contact system (Figure 4.1).

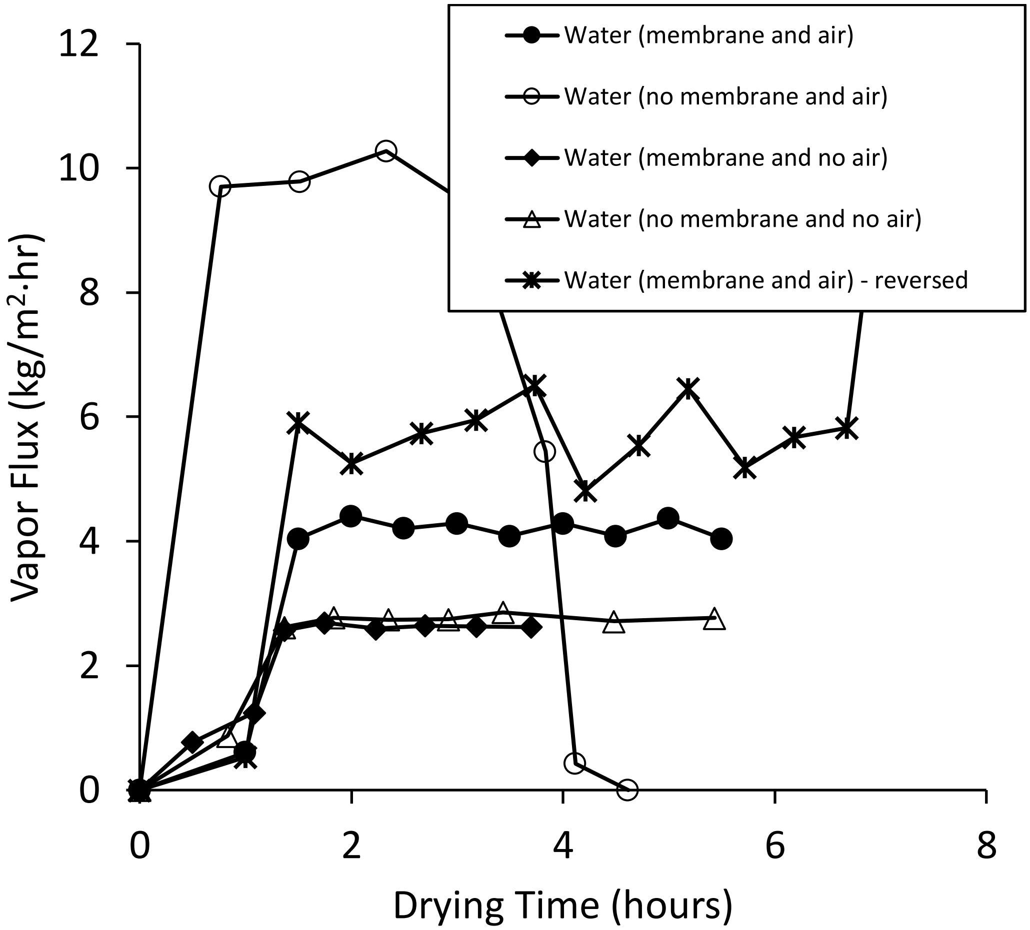


Figure 4.1 Deionized water drying test by oven heating of drying module (temperature of oven = 100°C)

For drying tests with sludge, three temperatures were compared: 80, 100, and 120 ^{0}C. Figure 4.2, the mass of sludge decreased because of losing water molecules by evaporation through breathable hydrophobic membrane. Figure 4.2 shows that vapor flux from sludge through the membrane increased from 0.13 g/cm^2/hr to 2.4 g/cm^2/hr with increasing heating temperatures from 80 to 120 ^{o}C. Due to oven heating the sludge directly, water molecules evaporated very quickly and the vapor pressure on the feed side increased.

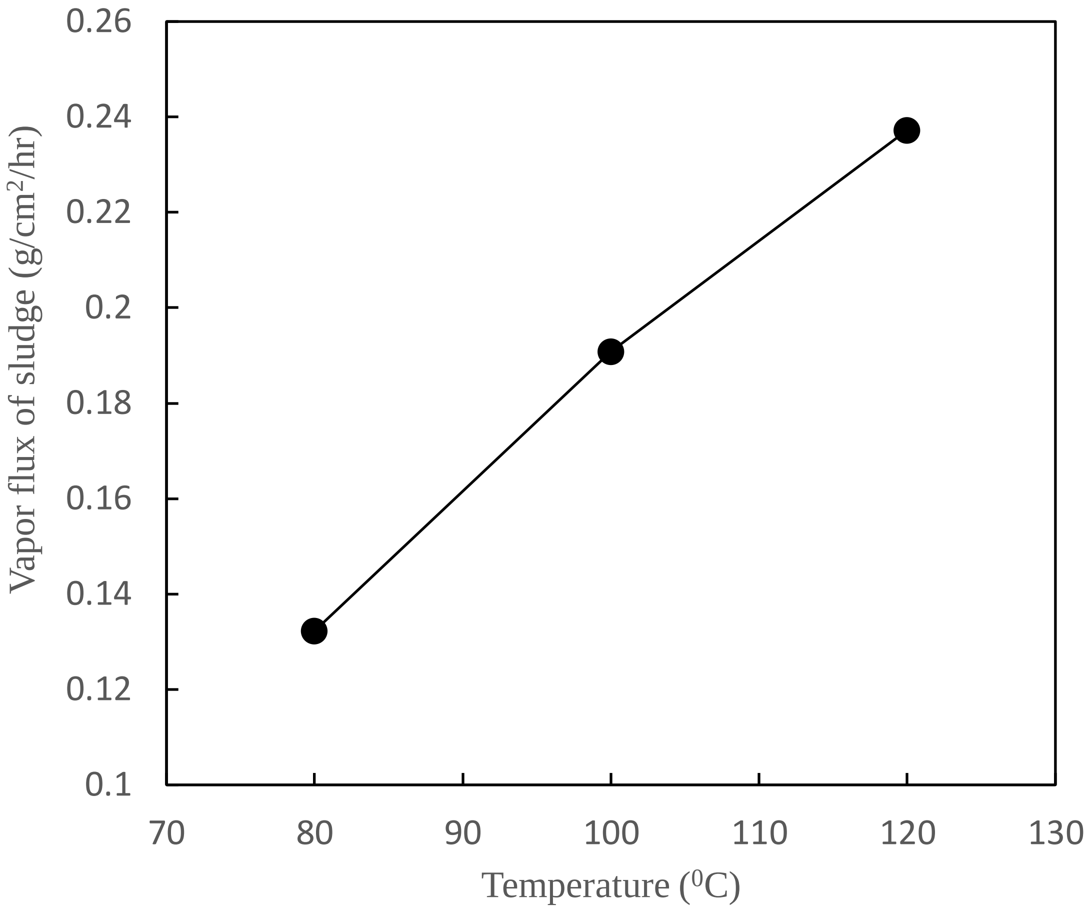


Figure 4.2 Effect of sludge heating on vapor flux through the membrane for a constant air flow rate of 5 L/min and sludge volume of 120 mL.

4.2 Sludge Drying by Heated Sweeping Air

To evaluate the sludge drying by heated sweeping air, four different air temperatures were studied: 25, 50, 75, and 100 ^{0}C. Mass losses from sludge-filled drying modules are presented in Figure 4.3. The mass of sludge in the membrane envelopes decreased rapidly and linearly for all tested temperatures. The completely dried sludge sample exhibited a wafer-like consistency throughout, indicating that thorough drying occurred consistently across the laminated breathable membrane (Figure 4.4).

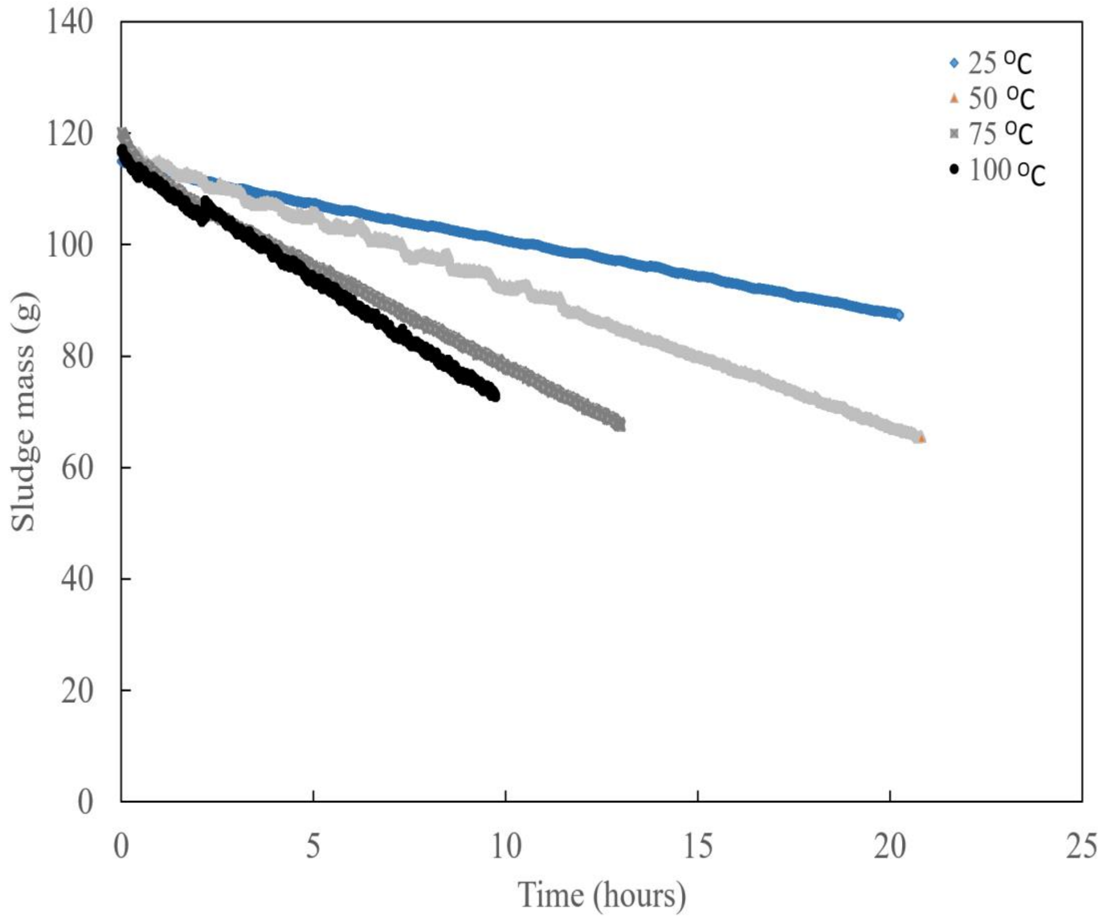


Figure 4.3 Effect of temperature on sludge drying by heated sweeping air (air flow rate = 5 L/min and sludge volume = 120 mL).

Figure 4.4 Dried sludge from the breathable membrane module experiment.

Figure 4.3 shows vapor flux through the membrane increased when the temperature of sweeping air is increased from 25 to 100 °C. It is hypothesized that the layer between sludge and the hydrophobic side of the membrane reaches 100% relative humidity and consequently high vapor pressure as the sludge temperature increases from sweeping air. In other words, due to the difference between relative humidity at the feed side and the permeate side, vapor pressure at the feed side is greater than vapor pressure at the permeate side. This observation can be explaining with the following empirical equation (Sherwood, 1936):

$$W = 0.027\, V^{0.8} (P_s - P_a) \qquad 4\text{-}1$$

Where W is rate of vaporization (kg/m^2/hr), V is the velocity of sweeping air over wet surface (m/sec), P_a is vapor pressure of water in the permeate side (mm Hg), and P_s is vapor pressure of water in the feed side (mm Hg).

Figure 4.5 compares the sludge drying between two heating methods: oven heating and sweep air heating. Following the initial rapid mass loss period, the mass of sludge in the drying module placed in the 75°C-oven decreased linearly from 102 g to 10 g in 13 hours. On the other hand, decrease in mass for the sludge under a constant flow of 75°C air decreased from 102 g to 52 g during the same drying period. This lower drying rate observed with sweeping air heating was expected as kinetic energy supplied to sludge-membrane layers. While kinetic energy supplied for heating the sludge module process, which heat is in contact with sludge, is higher than heating the sweeping air process, the results shows that vapor flux during heating the sludge is higher than heating sweeping air. In other words, this difference may be attributed to greater temperature gradient on sides of the membrane.

Although heating sludge process has high vapor flux through the membrane, heating the sweeping air is an efficient way of raising the sludge temperature at the sludge-membrane interface.

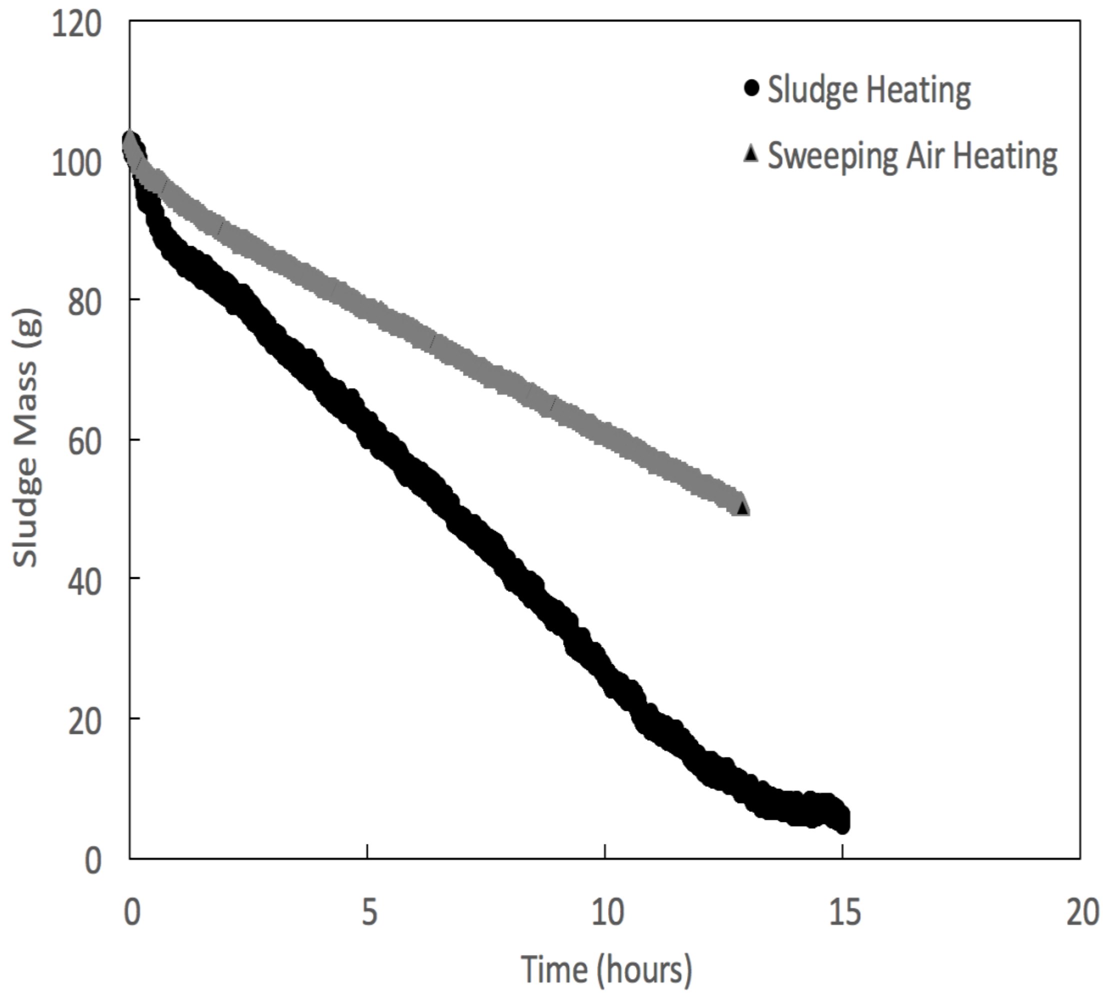


Figure 4.5 Effect of sludge heating and sweeping air heating on mass of sludge loss through the membrane (air flow rate = 5 L/min; temperature = 75 ^{0}C; sludge volume = 120 mL.

Figure 4.6, air flow in the membrane module carries water vapor out of the membrane module. If the air flow rate is relatively high, the relative humidity inside the air compartment should approach the relative humidity of inlet air. Since the relative humidity of the feed side would be 100%, a constant humidity gradient between the sludge side and air side, can be maintained with this rapid withdrawal of humidified air. Moisture flux across the membrane was calculated for the linear region of each drying curve by dividing the initial 8-h drying rate by the total membrane surface area (50 cm^2). For the air flow rate of 5 L/min, a sludge drying rate of 0.03 $g/cm^2/hr$. When the air flow was increased to 10 and 20 L/min, sludge samples dried at 0.04 and 0,05 $g/cm^2/hr$, respectively.

Figures 4.7, 4.8, and 4.9 shows that vapor flux through the breathable membrane is strongly influenceD by both temperature and flow rate. In order to determine the more important parameter that effects the sludge drying rate, Leonard et al. (2006) conducted the multilinear regression for both temperature and flow rate:

$$y = a_1 + a_2 T + a_3 \mathrm{V} \qquad 4\text{-}1$$

Where T is the temperature and V is the air velocity. Based on the regression analysis, Leonard et al. (2006) concluded that the temperature has more significant effect on vapor flux than air velocity.

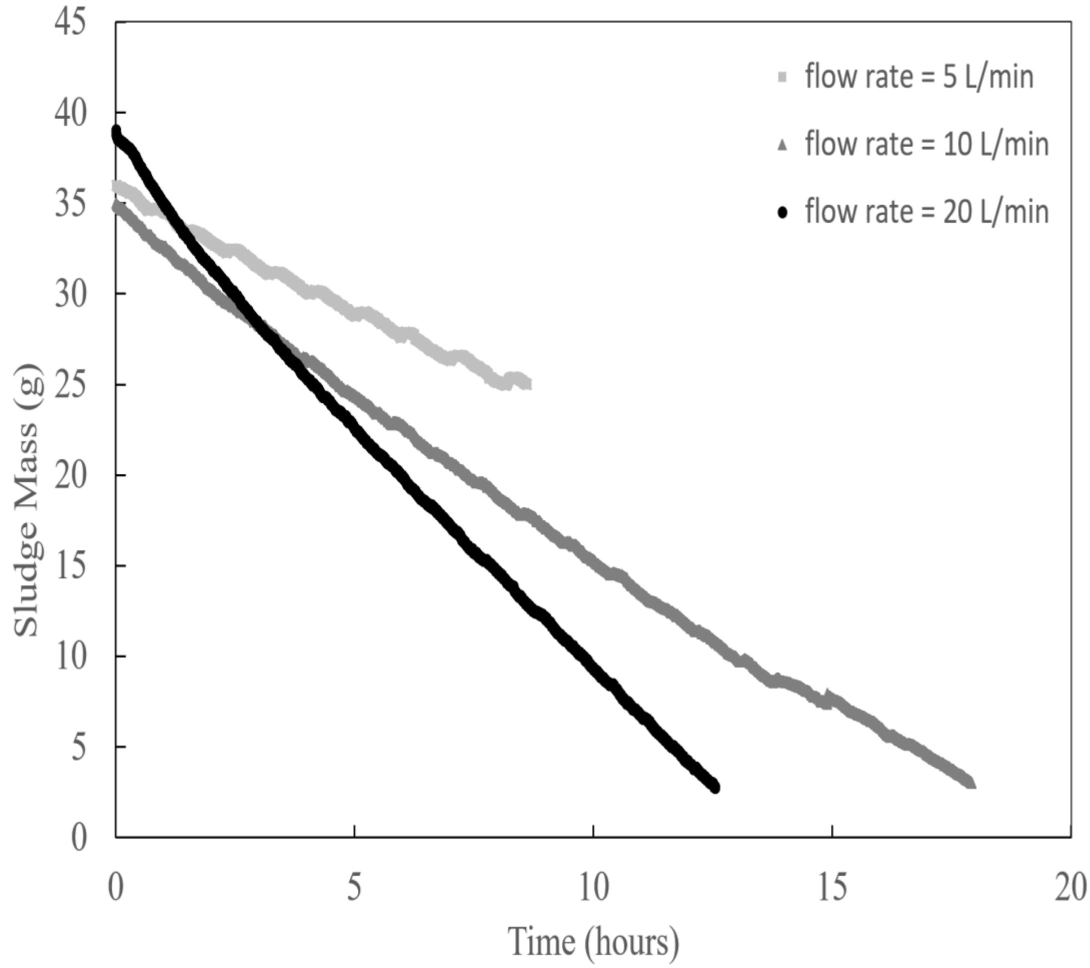


Figure 4.6 Effect of sweeping air flow rate on sludge drying (temperature = 25 ^{0}C and sludge volume = 40 mL).

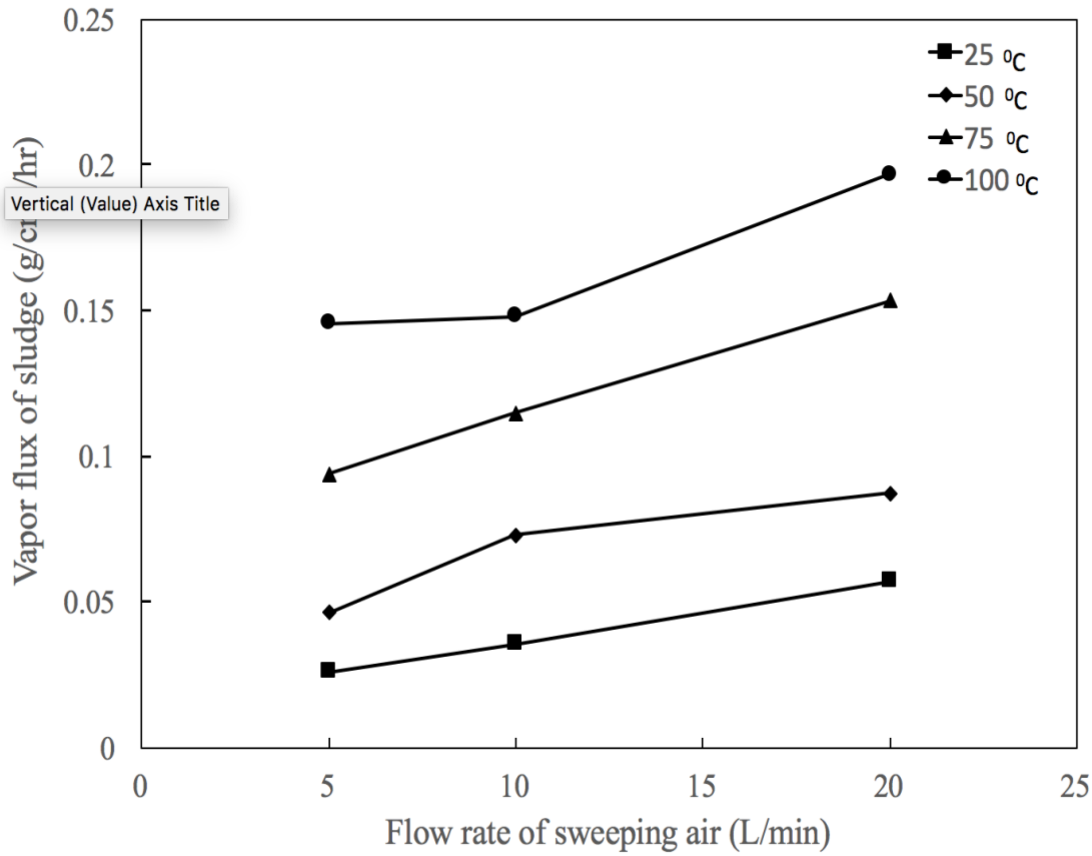


Figure 4.7 Effect of sweeping air temperature on vapor flux through the membrane (sludge volume = 40 mL).

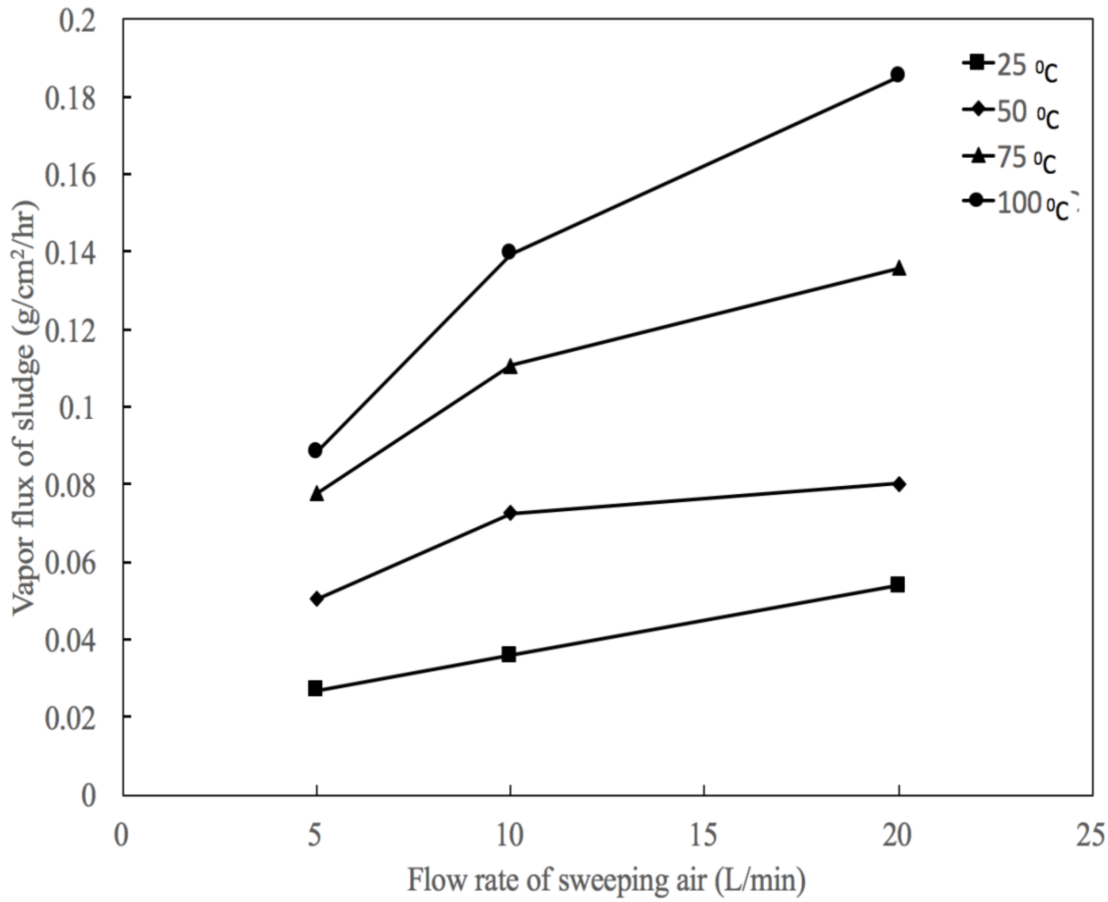


Figure 4.8 Effect of sweeping air temperature on vapor flux through the membrane (sludge volume = 120 mL).

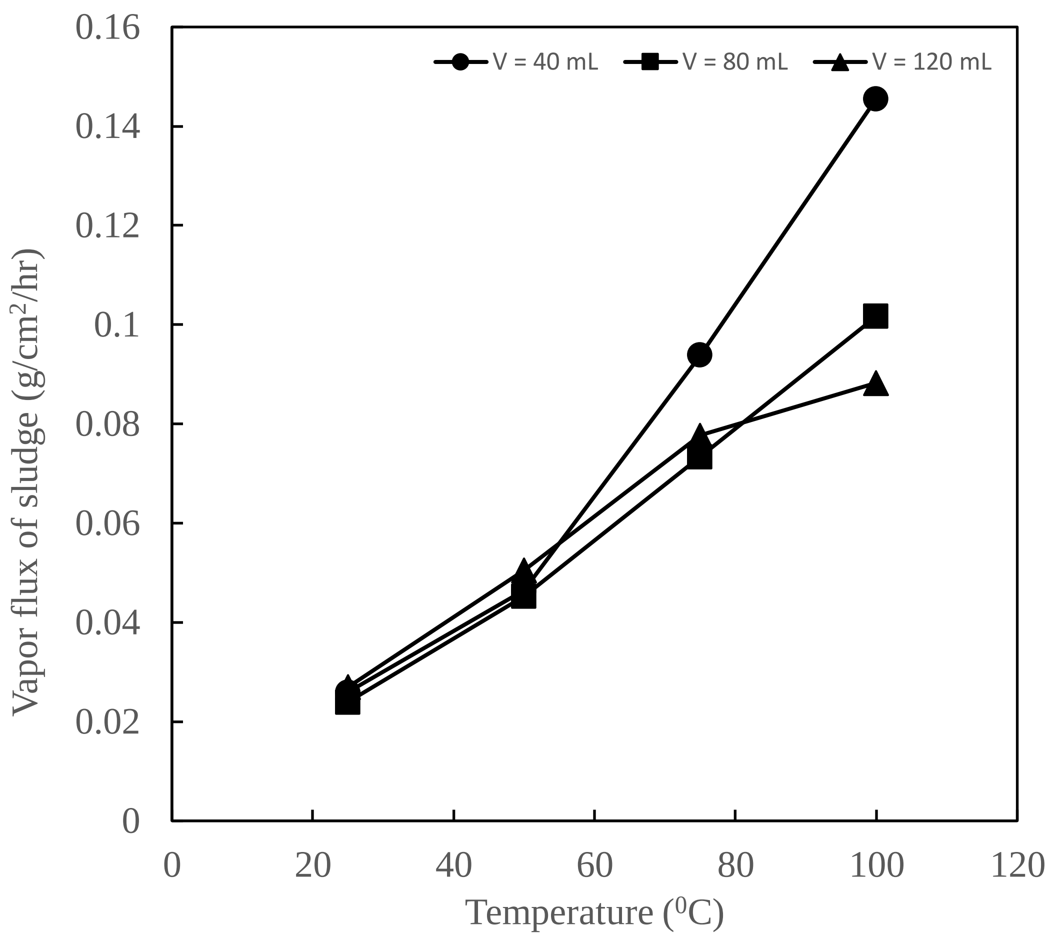


Figure 4.9 Effect of sludge volume and sweeping air temperature on vapor flux through the membrane (air flow rate = 5 L/min).

4.3 Effect of Diffusion Distance

Sweeping air drying experiments were typically conducted with the sludge in the top compartment and the sweeping gas supplied to the bottom chamber to ensure the contact between the sludge and membrane is maintained throughout the drying cycle. However, in full-scale application of membrane drying technology, it is expected that the sludge will not be in contact with the membrane surface throughout the drying cycle due to the decrease in bulk sludge volume. In order to determine the effect of air gap between the sludge and membrane on sludge drying, a drying experiment was conducted with sludge in the bottom side and the air flow on the top compartment. Figure 4.10 compares the difference in drying rates between the sludge on top of the membrane and sludge in bottom of the membrane. When the sludge was in touch with the membrane throughout the drying cycle (i.e., sludge on top), sludge mass decreased from 115 g to 50 g in 24 hours. When the sludge was placed in the bottom compartment, the sludge mass reached 65 g in 24 hr. The vapor flux values calculated from the linear region of each drying curve were 0.05 $g/cm^2/hr$ for sludge on top and 0.0326 $g/cm^2/hr$ for sludge in bottom.

Coackley (1962) and Sherwood (1936) indicated that if the feed side was not in contact with the permeate side, air gap generated between the feed surface and permeate surface or air sweeping surface can decrease moisture transport due to two factors. The first was heat transfer. More specifically, air gap caused additional heat transfer resistance between sweeping air and the sludge. Thus, as air gap increased, heat transfer from the sweeping air to sludge would be slower. The second factor was the diffusion distance. When water molecules evaporate, they diffuse from water surface to overlying air layer and then to sweeping air. Thus, the presence of air layer caused longer diffusion distance, thus resulted in slower drying rates.

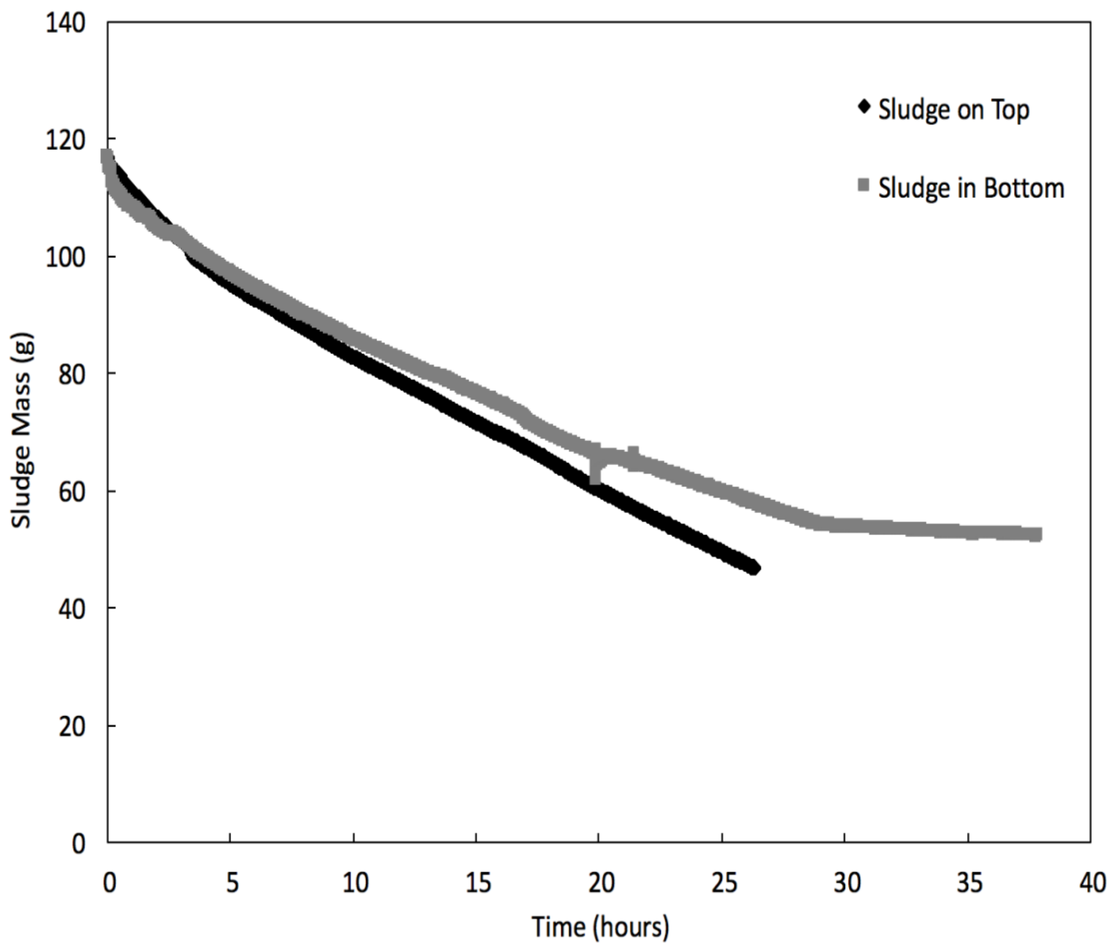


Figure 4.10 Effect of diffusion distance on sludge drying rate (air temperature = 25 ^{O}C; air flow rate = 20 L/min; sludge volume = 120 mL).

4.4 Validation of COMSOL Results

4.4.1 Temperature of Sweeping Air

A series of model simulation was performed using COMSOL program to evaluate the effect of temperature and air flow rate on the sludge drying. Figure 4.11 depicts the effect of sweeping air temperature on vapor flux. As air temperature increased from 25^{o}C to 100^{o}C, vapor flux through the membrane was increased from 0.045 to 0.2 g/cm^2/hr. In addition, Figure 4.11 shows that the simulation results closely matched experimental results, indicating that the parameter values and assumption included in COMSOL model were appropriate.

However, vapor flux values obtained from the model simulation deviated from the experimental values for the test conducted at an air flow rate of 5 L/min (Figure 4.12). The vapor flux values for 75^{o}C and 100 ^{o}C were 0.07 and 0.08 g/cm^2/hr, respectively, while the model predicted values were 0.08 and 0.13 g/cm^2/hr for the 75 ^{o}C and 100 ^{o}C air flows. This difference between the experimental results and model simulation results may be attributed to greater heat loss at low air flow rate than higher flow rate because of low thermal capacity of air. Since the drying module was not insulated, slower air stream may be subjected to greater heat loss due to longer retention time in the air compartment.

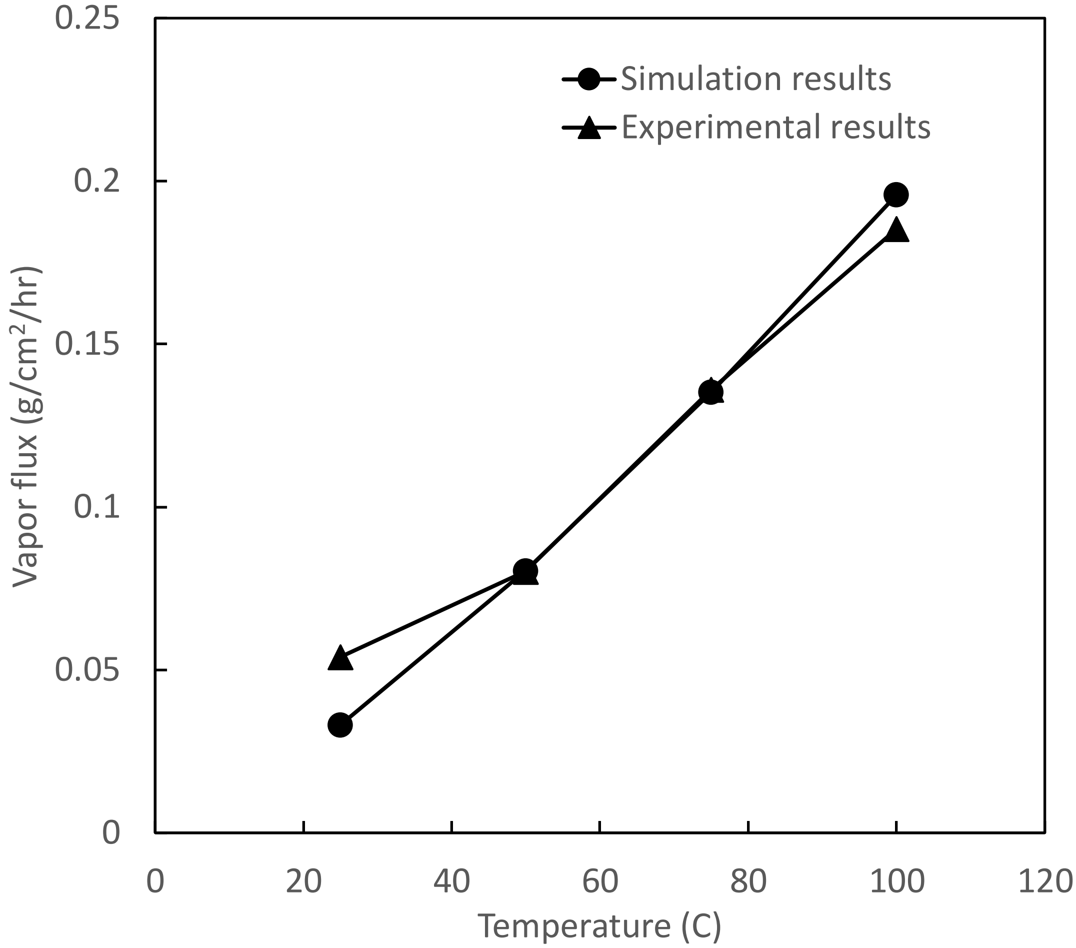


Figure 4.11 Comparison of experimental results and COMSOL simulation results on the effect of sweeping air temperature on vapor flux through the membrane (air flow rate = 20 L/min).

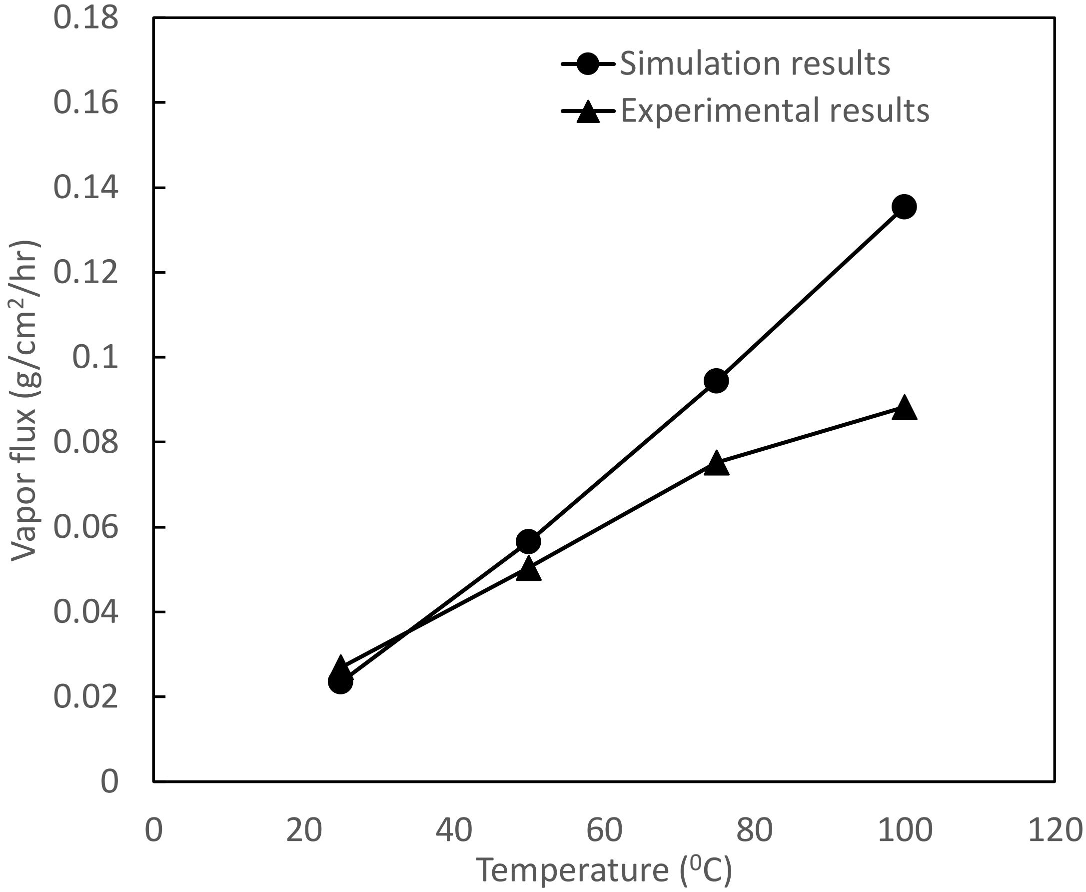


Figure 4.12 Comparison of experimental results and COMSOL simulation results on the effect of sweeping air temperature on vapor flux through the membrane (air flow rate = 5 L/min).

4.4.2 Effect of Air Flow Rates

Figure 4.13 shows that the model simulation results closely matched the experimental results for the effect of air flow rates on sludge drying. However, Figure 4.14 shows that when the sweeping air was heated to 100^{o}C the simulated vapor flux does not agree with the experimental vapor flux. This difference increases as the air flow rate decreases (Figure 4.14). This difference may be attributed to greater heat loss at low air flow rate than higher flow rate because of low thermal capacity of air. Since the drying module was not insulated, slower air stream may be subjected to greater heat loss due to longer retention time in the air compartment.

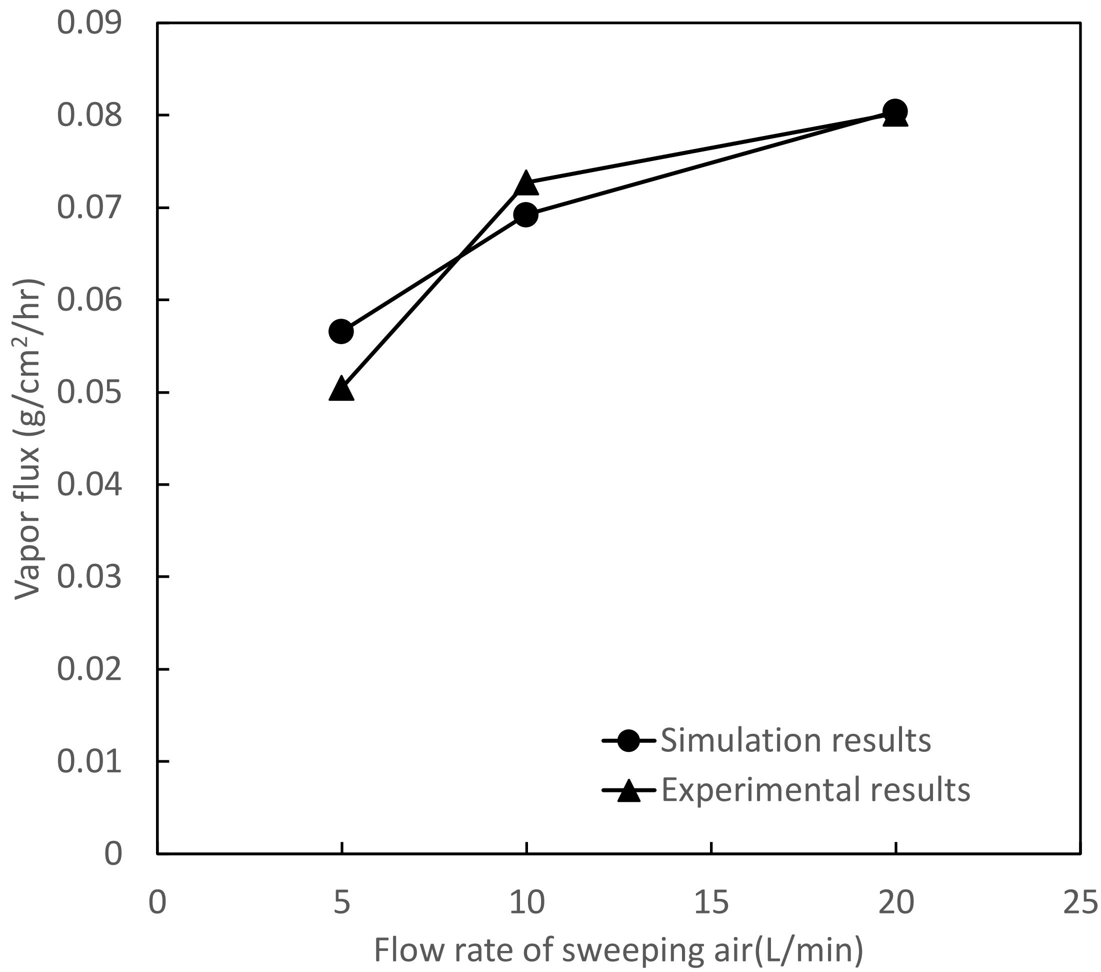


Figure 4.13 Comparison of experimental results and COMSOL simulation results on the effect of air flow rate on vapor flux through the membrane (temperature = 50 ^{0}C).

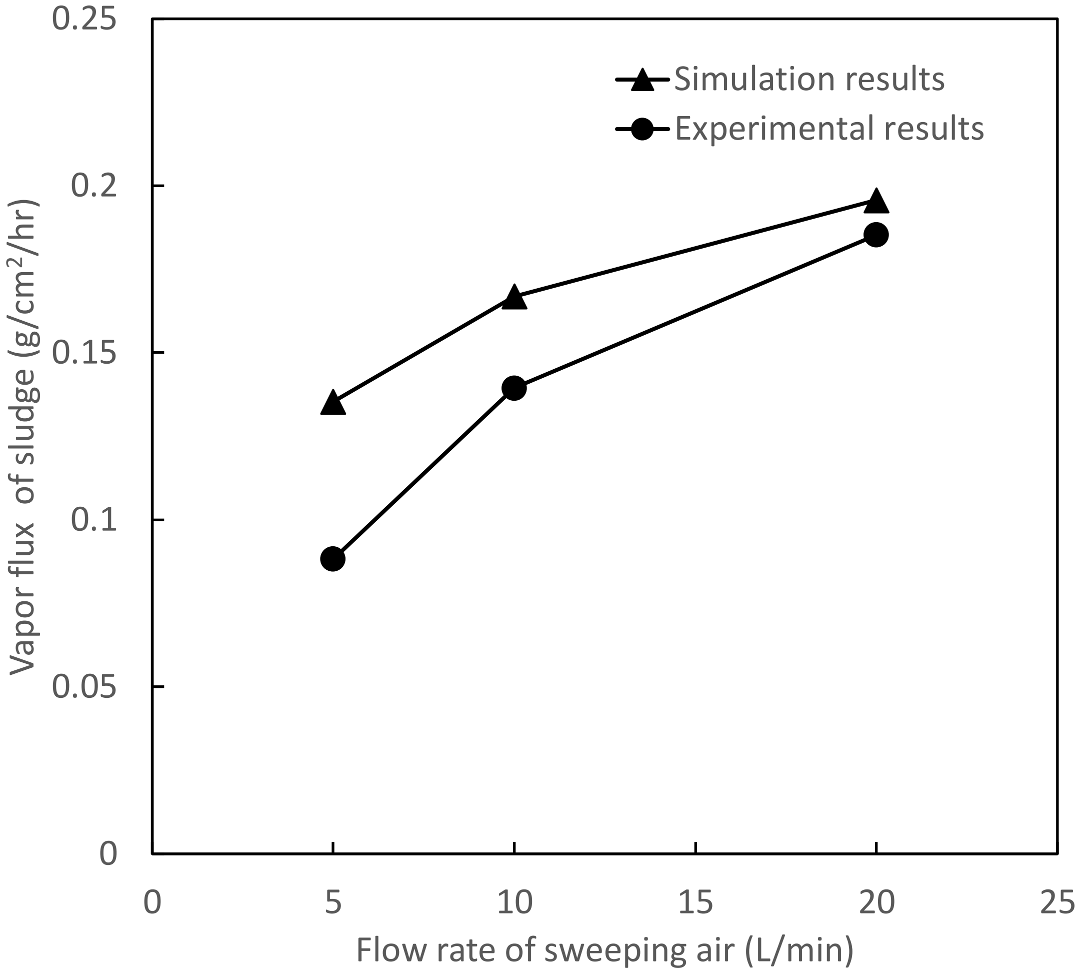
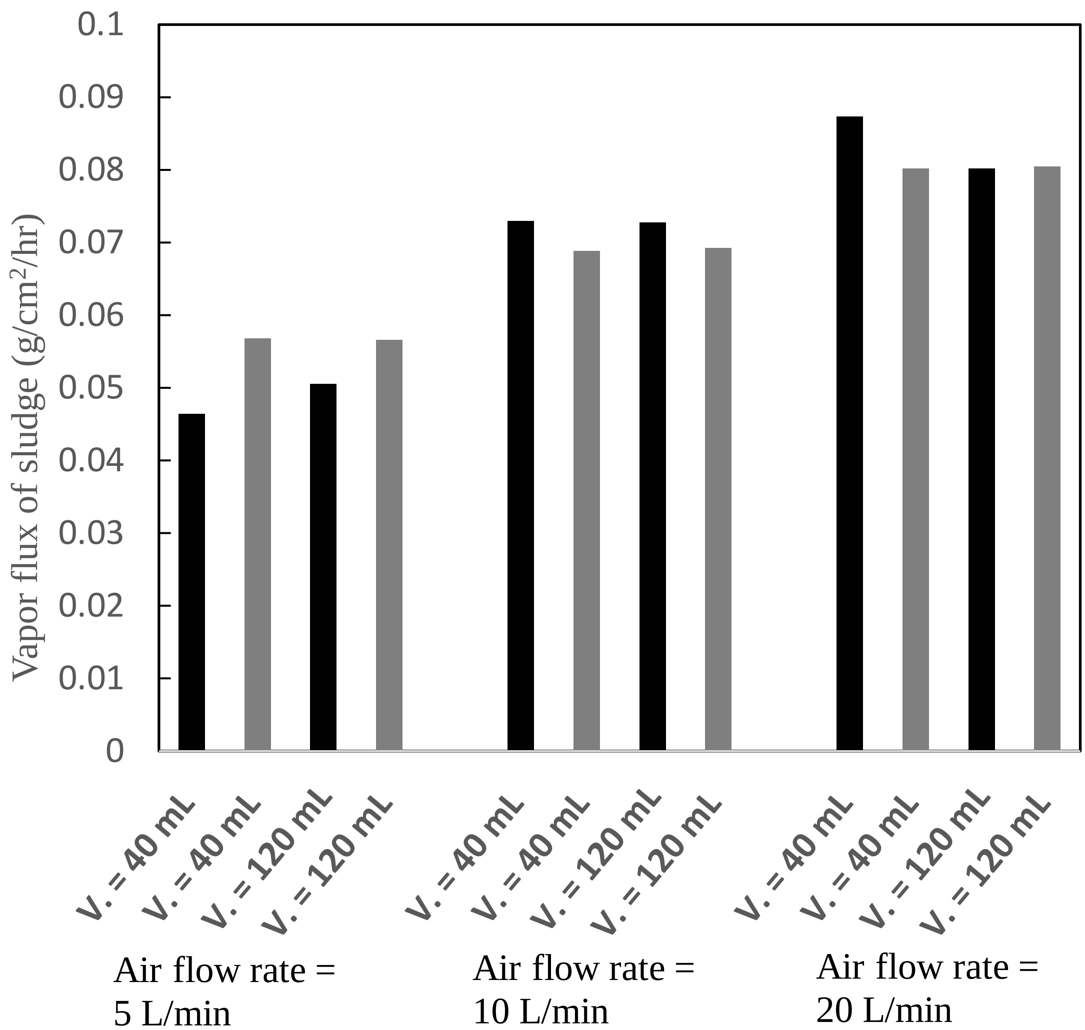


Figure 4.14 Comparison of experimental results and COMSOL simulation results on the effect air flow rate on vapor flux through the membrane (temperature = 100 ^{0}C).

4.4.3 Thickness of Sludge Inside the Membrane Module

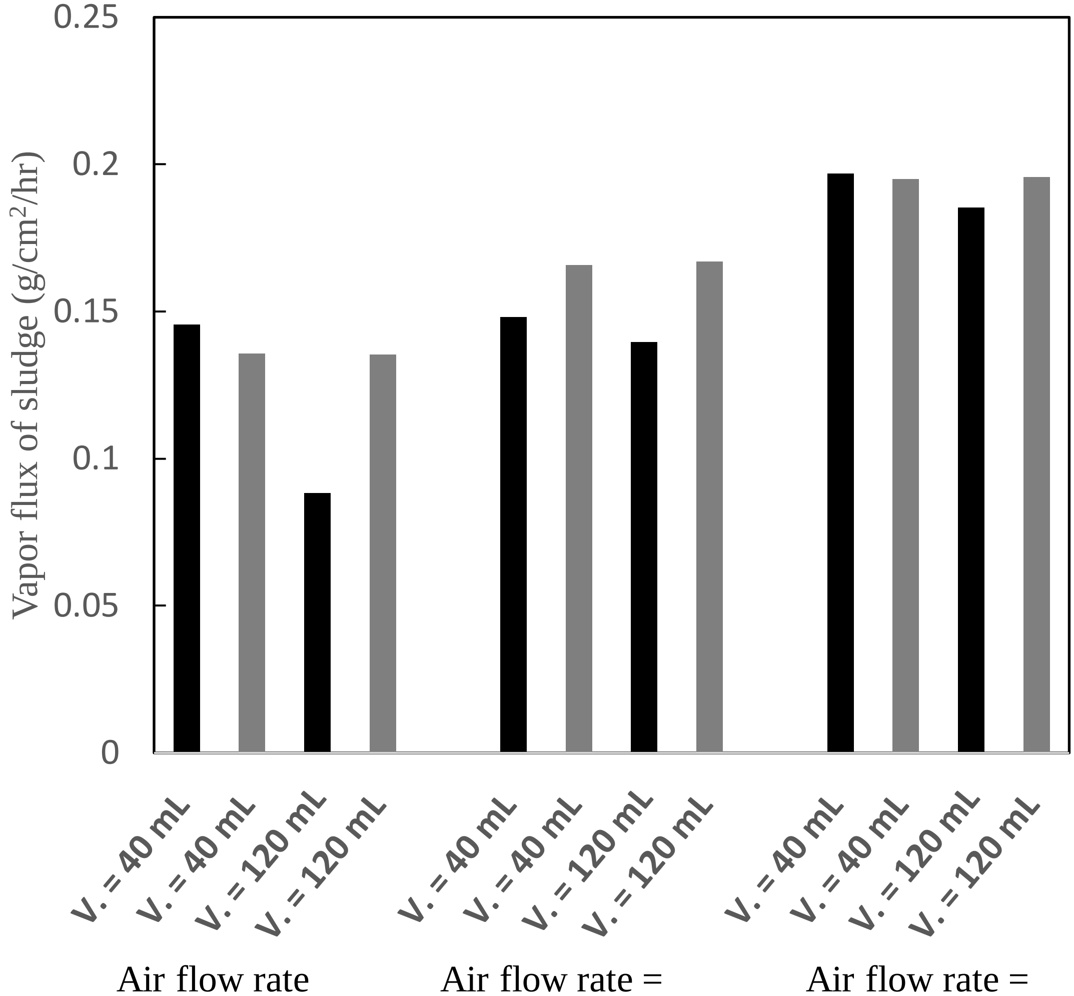
Similar values of vapor flux observed with different volumes of sludge inside the sludge compartment of the drying module (Figures 4.15 and 4.16). These results suggest that different thicknesses of sludge did not affect the vapor flux through the breathable membrane. Figures 4.15 and 4.16 also show that the simulation results closely agreed with the experimental results, confirming that the vapor flux across membrane is independent of sludge thickness



V. = sludge volume inside the membrane module,

Experimental results, Simulated results

Figure 4.15 Effect of different sludge volumes on vapor flux through the membrane at various air flow rates (temperature = 50 ^oC).



V. = sludge volume inside the membrane module,

Experimental results, Simulated results

Figure 4.16 Effect of different sludge volumes on vapor flux through the membrane at various air flow rates (temperature = 100 ºC).

4.5 Full Scale Design

In order to demonstrate the application of breathable membrane technology on scaled-up basis, a full-scale design was proposed for a typical sludge dewatering plant. It is assumed that the influent total suspended solids (TSS) concentration is 200 mg/L and effluent TSS is 20 mg/L. The average wastewater flow rate of the system is assumed to be 1 million gallons per day. Based on these information, the mass of wet sludge produced was approximately estimated to be 7000 kg/day. A hydrophobic breathable membrane system was designed to dewater the wastewater sludge 60% moisture content. The combustion of the sludge can be self-sustained in an incinerator without supplemental heat if sludge moisture contents are less than 70% (EPA, 2003).

The values of vapor flux obtained experimentally are tabulated in Table 4-1. These values are the same as the model-simulated vapor flux. From these flux values, required surface area was obtained. The dimension of a full-scale sludge dewatering unit equipped with sheets of breathable membrane is 3 meters long, 3 meters wide, and 3 meters height, which is easily constructed. This unit has multiple layers of membrane. Each membrane is in contact with sludge on one side and the other side is in contact with sweeping air. The estimated numbers of membrane sheets needed to achieve sludge moisture content of 60% are presented in Table 4-1.

For sweeping air flow rate of 5 L/min with a temperature of 25 ^{o}C, the number of membrane sheets needed are 56 sheets to achieve 60% moisture content. If the flow rate was increased to 20 L/min and the temperature of sweeping air increased to 50 ^{o}C, the number of sheets would decrease to 17 to achieve the same 60% moisture content.

Flow rate of sweeping air	Temperature	Vapor flux through the membrane	Calculated Surface area	Number of sheets
L/min	^0C	kg/ m^2/hr	m^2	
5	25	0.235	495	56
5	50	0.566	206	23
5	75	0.944	123	14
5	100	1.35	86	10
10	25	0.287	406	46
10	50	0.693	168	19
10	75	1.16	100	12
10	100	1.67	69	8
20	25	0.331	352	40
20	50	0.804	145	17
20	75	1.35	86	10
20	100	1.96	59	7

Table 4.1 Calculation of number of membrane sheets based on tabulated experimental and model-simulated vapor flux.

Chapter 5

CONCLUSION

Drying test with the oven heating of drying module showed that vapor flux through the membrane increased from 0.13 to 2.4 $g/cm^2/hr$ as the oven temperature increased from 80 to 120 ^OC. Similarly, increasing the sweeping air temperature from 25 to 100 ^0C increased vapor flux from 0.05 to 0.19 $g/cm^2/hr$. Increasing flow rate of sweeping air also increased vapor flux, but changing sludge thickness did not affect the sludge drying rate. As expected, direct heating of sludge in the oven resulted in higher vapor flux than drying with heated sweeping air. This difference may be attributed to greater temperature gradient across the membrane between the feed and permeate side. However, heating of sweeping air may be more efficient way of raising the sludge temperature at the membrane interface while achieving comparable drying rates. Furthermore, drying tests suggest that effective drying of from membrane-enclosed sludge depends on whether or not sludge is in contact with the membrane.

COMSOL simulation program was used to predict vapor flux through breathable membrane using the same experimental conditions. The simulation model was calibrated with the experimentally derived model parameters. Computer simulation results closely matched the experimental results from various sweeping air temperatures, flow rate sweeping air flow rates, and sludge thickness. This model was used to estimate a full-scale dimension of membrane-based drying system for a typical sludge dewatering plant.

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