# Recovery of Desalination Brine Concentrate Using Waste Heat and Membrane Evaporation

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## ABSTRACT

The desalination processes produces highly concentrated brine effluent which is costly to dispose of for inland desalination facilities. The current study uses a membrane evaporation process to reduce brine volumes and recover treated brines. Membrane evaporation was evaluated at the bench scale for different brine flow rates, airflow rates and air temperatures, extent of fouling and possible means to control fouling. The evaporation rate was found to increase with airflow rate and air temperature and decrease with brine flow rate. Model predictions, derived based upon literature values for heat and mass transfer correlations, agree well with the measured evaporation rate. The economics of the proposed process appear favorable because energy demands can be neglected when using waste heat. Fouling was observed in experiments for brines with and without ferrous iron. Cleaning of the membranes by flushing the fibers with brine at higher flow rates was not able to alleviate fouling. Fouling could be alleviated when the pH of the brine solution was lowered for the durations of the studies conducted.

Keywords: membranes, desalination brine, wasteheat

# **1 INTRODUCTION**

The current study focuses on developing an alternative technique to reduce brine concentrate volumes. As described by Johnson et al. [1], membrane evaporation processes act similar to heat exchangers where a cold fluid can be pumped inside the hollow membranes while exposing the outer surface of the membrane to hotter fluids flowing in a counter direction. In an application for evaporating RO brine concentrate, the brine concentrate is the cold fluid that enters inside hollow fiber lumens while exposing outer surface to low grade heat. During this process, heat is exchanged and some of the water molecules from brine concentrate convert into vapor and escape through the membrane micro pores while the humidity of the air increases if heated air is used as the low grade heat source. If desired, the water vapor can be collected as condensate when air leaving the membrane module cools. Low grade heat sources can be used because this membrane process does not require high temperature differences to carryout mass transfer operations. An example of low grade heat is compressed air at wastewater treatment facilities. Aeration is the most energy-intensive operation in

wastewater treatment, amounting to 45–75% of plant energy costs [2]. During aeration, the compression of the air causes an increase in air temperature (often modeled as adiabatic). This waste heat can be utilized to drive heat and mass transfer operations for evaporating brine concentrate from inside the membrane. As an example, the activated sludge water recycling center for the city of San Antonio TX treats 303 MLD of wastewater in a two-stage aeration mode by providing 1,400 MLD of air with a rated discharge pressure of 186 kPa and a temperature of 345 K. In addition to aeration waste heat from wastewater treatment plants, there are other industries which generate waste heat at large scale [3], one such source is thermal and nuclear power plants. Power plants which produce electricity by burning either fossil fuels or radioactive elements lose energy as waste heat. The efficiency of the coal power plants in the US ranges from 30% to 40%. The waste heat is either dumped into the surrounding air or is cooled using freshwater from nearby water bodies. Utilizing waste heat to assist in brine disposal requires co-location of the water treatment facilities. The colocation or at least a conveyance connection concept utilized here is not entirely new as it is currently used for coastal power plants and seawater desalination plants..

#### **2 THEORY**

The rate of water evaporation through the membrane is modeled as resistances coupled in series through the membrane and gas boundary layer similar to as descried by [1]. In the present analysis, a counter-current parallel flow membrane contactor configuration is used and the lumen side water mass balance is provided in equation 1 for a single hollow fiber.

$$\frac{dM_{water}}{dz} = K\pi d_o(\rho_{air} - \rho_{sat}) \tag{1}$$

The accompanied change in shell side water density in air due to all membrane fibers (N) in the module is provided in equation 2 with  $Q_{air}$  being the gas flowrate.

$$\frac{d\rho_{air}}{dz} = \frac{\partial M_{water}}{\partial z} * \frac{N}{Q_{air}}$$
(2)

As the water evaporates, the water inside the membrane is cooled and heat is extracted from the air with the resulting change in lumen side water temperature as provided in equation 3:

$$= \frac{\frac{dT_{water}}{dz}}{\frac{\partial M_{water}}{\partial z} * h_v + H\pi d_o (T_{air} - T_{water})}{M_{water} C_{p-water} T_{water}}$$
(3)

The accompanied change is shell side air temperature due to all membrane fibers is provided in equation 4.

$$\frac{dT_{air}}{dz} = \frac{HN\pi d_o(T_{air} - T_{water})}{Q_{air}C_{p-air}\rho_{air}}$$
(4)

In equations 1-4, z is the location along the length of the hollow fiber, N is the number of hollow fiber membranes within the module,  $Q_{air}$  is the air volumetric gas flow rate passing through the shell of the membrane module  $(m^3 s^{-1})$ ,  $h_{v}$  is the latent heat of vaporization of water which can be defined as energy required to convert from saturated liquid phase to vapor phase at constant temperature,  $T_{air}$  and  $T_{water}$ vary with distance within the membrane shell and lumen respectively.  $C_{p-water}$  is the specific heat of water, and  $\rho$  is the mass per volume density of air within the membrane shell. Correlations are available in literature for predicting mass (K) and heat (H) transfer coefficients. When combined, equations 1-4 can be utilized to size membrane systems for evaporating brine solutions given known values for inlet gas humidity and gas and water temperatures. Additional details for modeling analysis can be found in [1]

## **3 EXPERIMENTS**

Membrane modules were made by packing a required number of hollow fiber membranes in a pipe of suitable diameter. In this study, a micro porous hydrophobic hollow membrane fabric (CELGARD) made of polypropylene manufactured by 3M (St. Paul, MN) was used. A weaved sheet of 60 fibers was rolled and inserted into stainless steel tubing of 6.35 mm diameter with length of 0.35 m with both ends fitted with two T-fittings and sealed with epoxy. A constructed module is shown in Figure 1. The module was attached to system of flow meters, thermo couples and pressure gauges and weighing balances for continuous monitoring of temperatures pressures, water flow and airflow along the membrane similar to as depicted in figure 2.



Figure 1. Left: Membrane module sealed end for brine water connections. Right: Module with tee fittings enabling countercurrent brine to air flow operating configurations.

Air was sourced from a laboratory fume hood and flowed between  $4 - 30 \text{ L min}^{-1}$  into an oven housing a coiled copper tubing heat exchanger to produce heated air between 296 -353K. .DI waster was fed to the system at a rate of 0.3 -3 mL min<sup>-1</sup> for model validation (solving equations (1)-(4)). Synthetic brine was fed to the system to evaluate fouling and had a conductivity of 14,900 µmho/cm. Additional details for the apparatus are described by Johnson et al.[1].



Figure 2. Bench-scale membrane evaporator

# **4 RESULTS & DISCUSSION**

Model validation results were obtained by Johnson et. al. [1] and shown in Figure 3 for fraction of water evaporated over the operating conditions used. Overall the model slightly under predicts the amount of water evaporated (up to nearly 60%) in the bench scale studies but appears suitable for making conservative performance estimates for full scale systems. Full scale system estimates were made for evaporating brine using aeration wasteheat with an air temperature of 345K, an airflow rate of 1400  $10^6$  L d<sup>-1</sup> at relative humidity of 20%, 40% and 60% and brine flowrates up to 4.2  $10^6$  L d<sup>-1</sup>. Scaled membrane module dimensions were: module diameter of 119 cm, module length of 30.5 cm with a membrane packing of 25%.



Figure 3 Parity plot between measured and modeled values, adapted from [1]



Figure 4. Water evaporated fraction and rate for full scale conditions at relative humidity (RH) of 20, 40 and 60%.

Predictions for the full scale conditions for brine fraction evaporated and amount of brine evaporated are shown in Figure 4 for various brine flow rates and the aeration operating conditions described previously. As relative humidity (RH) of the air decreases, fraction evaporated increases. At a typical RH of 60% representative of San Antonio Texas, the fraction (0.001) and amount of brine evaporated (4,600 L d<sup>-1</sup>) a are negligible when operated a full scale brine flow conditions of 4.20E+06 L d<sup>-1</sup>. At a RH of 40%, representative of low humidity days in San Antonio TX and typical for Phoenix AZ, the amount of evaporation for a full scale system would be ~  $8,200 \text{ L} \text{ d}^{-1}$  In these scenarios, the membranes act as heat exchangers, warming the brine 3 deg C with minimal evaporation. If the brine flow rate were reduced to 50,000 L d<sup>-1</sup>, approximately 40% of the brine (20,000 L d<sup>-1</sup>) could be evaporated at a high RH of 60%. More evaporation occurs as the water is heated by nearly 16 deg C. It is important to note that the process is not limited by the water content of the air in the contactor as the relative humidity of the air leaving the module for these conditions is only 34%. However, because this air is warmer still than the water in the membrane pores, the water vapor pressure in the heated air approaches the saturated water vapor pressure in the colder membrane pores and thus additional evaporation becomes minimal unless warmer water is used when entering the membrane module.



Figure 5 – Fouling control in laboratory modules when operated at low pH, adapted from [1].

In this analysis it is assumed the membranes remain unfouled. Prior conducted fouling results are shown in Figure 5 where it was demonstrated that when operating at low pH of 4 the system was able to maintain a constant rate of evaporation [1]. The main precipitate in this analysis was assumed to be CaCO<sub>3</sub> and pH control appears adequate to control fouling.

Brine fraction evaporated can be recovered as the air cools either to ambient conditions at 100% relative humidity or by chilling the air to provide near 0% relative humidity. Condensate produced along with amount of water evaporated in Figure 4 for a brine flow rate of 50,000 L d<sup>-1</sup> with1,400 MLD of air at RH = 60% are shown in Figure 6. Because the incoming air contains significant amount of water vapor, compression in the blowers causes condensation after the air cools and the expected amount of condensate produced (when no chillers are used) slightly exceeds the amount of water evaporated as shown in Figure 6. If a chiller were utilized, nearly 87% more condensate could be recovered but at additional expense associated with cooling the air.



Collection 20,000 L d-1 of brine ( representative of ambient conditions without a chiller) and given an estimated cost of membrane at approximately \$15 m<sup>-2</sup>, a required full scale membrane area of 1,120 m<sup>2</sup> and assumed life time of 3- 5 years [4], equates to an unamortized membrane treatment cost of \$0.46 - \$0.77 m<sup>-3</sup> of brine evaporated. Brine disposal through deep well injection is estimated by the San Antonio utility to cost \$0.68 m<sup>-3</sup> of injected brine and based on the cost of for membrane, membrane evaporation appears competitive with deep well injection. It is important however to note for the cost incurred that the brine volume requiring disposal is reduced while high quality permeate is produced. Thus if water production costs are considered, comparisons are more favorable. For RO, desalination costs are reported to be  $0.81 \text{ m}^{-3}$  [5] with 76 % of the costs ( $0.62 \text{ m}^{-3}$ ) associated with membranes, power, residual waste disposal, labor and 24% (\$0.19 m<sup>-3</sup>) attributed to chemicals, minor equipment and maintenance and other [6] Thus, the total cost of disposal by deep well injection and water production by RO sums to \$1.49 m<sup>-3</sup> whereas it is \$0.65- 0.96 m<sup>-3</sup> for brine evaporation with condensate recovery. In this comparison, it is assumed the membrane evaporation process would share similar chemicals, minor equipment and maintenance costs as RO at \$0.19 m-3 additional to the cost of membrane.

## 5 SUMMARY

Model predictions, derived based upon literature values for heat and mass transfer correlations were used for predicting brine concentrate evaporated fraction and condensate collected. The volume of water treated was low but the process economics appear favorable because energy demands were neglected when using waste heat.

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