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# Numerical study on multi-stage vacuum membrane distillation with economic evaluation

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## HIGHLIGHTS

- This study aims to present an economic analysis of suggested MVMD system configurations.
- The suggested system configurations are series type, parallel type, and mixed type.
- 4-types of MVMD configurations are suggested to reduce the water product cost.
- Water product cost can be as low as \$0.52/m$^3$.

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## ABSTRACT

The productivity and specific heat energy consumption of vacuum membrane distillation (VMD) can be enhanced by the appropriate module length because conductive heat transfer through the membrane is negligible. Therefore, this paper suggests the use of a multi-stage vacuum membrane distillation (MVMD) system. The suggested stages in the MVMD system can be arranged in series or in parallel and/or mixed. The MVMD system is simulated using a one dimensional in-house code with energy and momentum balance and heat and mass transfer equations. The capital cost, maintenance cost, spares cost, and operation cost are considered to evaluate the water product cost. Comprehensive analysis of productivity, the water product cost, and the membrane wetting problem is carried out to find the best configuration in the systems studied. As a result, the mixed MVMD system (which has 20 stages) has the highest productivity (3.79 m$^3$/day), lowest water product cost (\$1.16/m$^3$), and lowest maximum transmembrane pressure difference (93.8 kPa) in the studied configuration. If the MVMD system uses the waste heat source, the water product cost can be reduced to \$0.52/m$^3$. This value shows the feasibility of using the MVMD system compared with the standard water product cost of \$0.63/m$^3$ for reverse osmosis.

## 1. Introduction

Vacuum membrane distillation (VMD) is a thermally driven process to separate pure water from brine water [1]. In comparison with previously developed distillation processes, such as multi-stage flash (MSF), multiple effect distillation (MED), and reverse osmosis, the advantages of the VMD method are the low requirement of plant space, low operation temperature and hydrostatic pressure, low influence of NaCl concentration, low heat transfer with the feed and permeate sides, no need for permeate side cooling, low mass transfer resistance, and low heat loss [2–4,7,24,26]. VMD processes can be applied in various parts of the industrial process, such as concentrating aqueous solutions, removing volatile organic compounds (VOCs) from contaminated water, and treating wastewater from industrial outputs [28]. However, VMD processes have critical performance disadvantages, such as high energy consumption for heating brine water [3]. Therefore, the MVMD is necessary to increase the productivity and decrease the water product cost. The productivity of VMD can be enhanced by having a longer module because high productivity and reduced specific energy consumption can be achieved due to the negligible conduction heat transfer between the feed and vacuum permeate side. However, a maintenance problem exists if a longer module is used in the VMD system. Therefore, the proposed MVMD system with a short module attached to each stage is required to minimize the membrane maintenance problem.

Fig. 1 shows the schematic of the VMD process [32]. The hollow fiber type VMD module has three layers: the shell feed side, the membrane

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layer, and the tube vacuum permeate side. Membrane distillation (MD) is a selective membrane separation process driven by a vapor pressure gradient from the inlet to the outlet of the module. Water evaporates on the membrane surface when the hot brine solution flows over it. The water vapor diffuses through the membrane.

Many previous studies highlight that the feed and permeate side inlet temperature and inlet velocity, module shape and process design are crucial factors for the water product cost from MD desalination. If the residence time is insufficient, then the temperature difference increases from the inlet to the outlet of the module, which adversely affects the productivity of distilled water. Also, if an inadequate inlet brine temperature flows, a low productivity or high specific energy consumption through the MD process can be incurred by a low inlet feed temperature, and these are especially severe at lower temperatures. Mengual et al. [4] presented and verified the heat and mass transfer model in VMD hollow fiber modules under many well-known, empirical heat transfer correlations developed for non-porous and rigid heat exchangers, and the experimental results were compared. Wirth and Cabassud [5] demonstrated the experimental results of two VMD hollow fiber module configurations (inside/out and outside/in). Banat et al. [6] noted that the mass flux of distilled water is highly sensitive to the feed temperature, especially at high vacuum pressure values. The mass flux is more sensitive to the vacuum pressure at low feed temperature levels than at high ones.

Many researchers have reported cost estimates for membrane distillation plants. Al-Obaidani et al. [7] carried out exergy analyses, sensitivity studies, and economic evaluations to assess the feasibility of the DCMD process. For DCMD with heat recovery, the estimated water product cost was $\$1.17/\text{m}^3$ using an MD system (24,000 m$^3$/day). Banat and Jwaied [8] estimated the cost of potable water produced by the compact unit ($\$15/\text{m}^3$) and large unit ($\$18/\text{m}^3$). Ali et al. [9] presented that under optimized conditions, the water production cost can be less than $\$1/\text{m}^3$ using the MD system (5 m$^3$/day), neglecting the cost of waste heat input to the system. Liu and Martin [10] estimated the cost of product water ($\$1.13/\text{m}^3$) using the MD system with a power plant (36.6 m$^3$/day). Meindersma et al. [11] estimated the cost of product water ($\$0.26/\text{m}^3$) using an air gap membrane distillation module (105,000 m$^3$/day). Sarbatly and Chiam [12] estimated the geothermal energy involved in desalination by vacuum membrane distillation, and found that the distilled water can save at least $\$0.72/\text{m}^3$ when geothermal energy is used. In addition, many researchers studied the reduction of water product cost with low-grade waste heat or alternative energy resources such as wind, solar, and geothermal energies [13–20].

This study aims to present an economic analysis of the suggested MVMD system configurations, and find the best MVMD system configuration among the studied systems using a commercial module (MD020CP2N). The suggested system configurations are series type, parallel type, and mixed type. The productivity analysis was conducted in MVMD systems under the following operating parameters: inlet feed temperature of 55 °C to 70 °C, inlet initial feed velocity of 1.2 to 2.4 m/s, inlet feed mass fraction of 0.025, and inlet permeate vacuum level of 4 kPa. The water product cost and membrane wetting problem analysis are conducted under the following operating parameters: inlet feed temperature of 70 °C, inlet initial feed velocity of 2.4 m/s, inlet feed mass fraction of 0.025, and inlet permeate vacuum level of 4 kPa, because VMD performs better at the high inlet temperature and high initial feed velocity. The capital cost, maintenance cost, spares cost, and operation cost are considered to evaluate the water product cost. The productivity, water product cost, and membrane wetting problem are simultaneously analyzed to find the best configuration in the studied systems, and to assess the feasibility of MVMD as a desalination process through numerical studies.

**Nomenclature**

| | |
|---|---|
| $A$ | Membrane area ratio for heat transfer through the fiber outside, fiber wall, or fiber inside [−] |
| $A_s$ | Heat exchanger surface area |
| $B'$ | Membrane distillation coefficient [kg/m$^2$s pa] |
| $C_p$ | Specific heat [J/mol K] |
| $d_h$ | Hydraulic diameter [m] |
| $d_i$ | Inside diameter of fiber [m] |
| $d_o$ | Outside diameter of fiber [m] |
| $d_s$ | Inside diameter of shell [m] |
| $h$ | Convective heat transfer coefficient [W/m$^2$K] |
| $\Delta H$ | Enthalpy of water evaporation [J/kg] |
| $J$ | Mass vapor flux [kg/m$^2$h] |
| $k$ | Thermal conductivity coefficient [W/mK] |
| $L$ | Module length [m] |
| $\dot{m}$ | Mass flow rate [kg/s] |
| $M$ | Molecular weight [g/mol] |
| $N$ | Number of fibers [−] |
| $P$ | Pressure [Pa] |
| $\Delta P_m^F$ | Transmembrane pressure difference [Pa] |
| $Q$ | Heat energy [W] |
| $r$ | Pore radius [μm] |
| $R$ | Ideal gas constant [J/mol K] |
| $T$ | Temperature [K] |
| $w$ | Mass fraction in liquid phase [−] |
| $x$ | Molar fraction in liquid phase [−] |
| $v$ | Velocity [m/s] |
| $V$ | Molar volume [m$^3$/mol] |
| $z$ | Axial coordinate for hollow fiber [m] |

*Dimensionless numbers*

| | |
|---|---|
| $Nu$ | Nusselt number [−] |
| $Pr$ | Prandtl number [−] |
| Re | Reynolds number [−] |

*Greek letters*

| | |
|---|---|
| $\alpha$ | Membrane surface area based on fiber inside diameter per unit length per fiber layer [m] |
| $\delta$ | Membrane thickness [m] |
| $\varepsilon$ | Membrane porosity [m] |
| $\mu$ | Kinematic viscosity [Pas] |
| $\rho$ | Density [kg/m$^3$] |
| $\phi$ | Packing density [%] |
| $\tau$ | Tortuosity [−] |
| $\eta$ | Pump efficiency |

*Subscripts*

| | |
|---|---|
| b | Bulk |
| m | Membrane |
| mean | Average value |
| s | Salt |
| w | Water |
| circulation | Circulation pump |
| heat | Heat |

*Superscripts*

| | |
|---|---|
| F | Feed side |
| P | Vacuum permeate side |
| W | Water |