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Research Paper

Preliminary Evaluation for Vacuum Membrane Distillation (VMD) Energy Requirement

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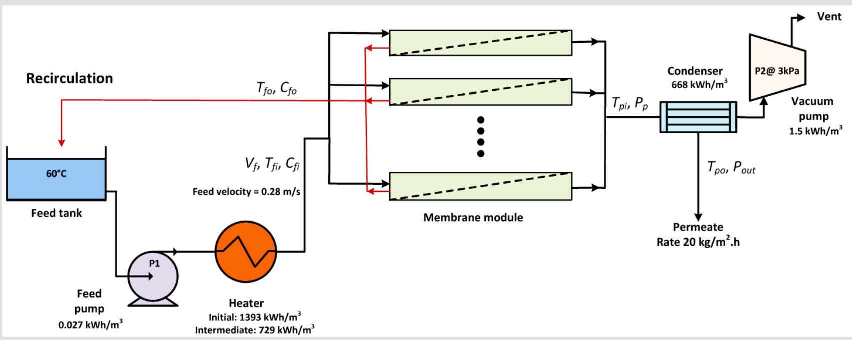
GRAPHICAL ABSTRACT



HIGHLIGHTS

- VMD energy required was modelled using experimental results and theoretical data.
- The trend between both data is generally consistent with each other.
- It is generally more energy efficient to run VMD in recirculation mode.
- Thermal energy contributes the most to the total energy required.
- To reduce operating cost, waste heat and heat recovery option should be considered.

ABSTRACT

The energy requirement of vacuum membrane distillation (VMD) with or without recirculation was modelled using both experimental results and theoretical data. The trends are generally consistent between the theoretical and experimental data. Thermal energy contributes the most to the total energy required for the VMD process. To lower the thermal energy cost, waste heat resource and heat recovery of latent heat from the permeate vapour are needed. The electrical energy consumption for VMD is slightly higher than brackish water reverse osmosis (RO) but lower than sea water RO. It is generally more energy efficient to operate the VMD in recirculation mode than single pass mode. Process engineering modelling results indicate that VMD may not be able to compete with RO directly but could be used as a complimentary process to RO, such as for brine concentrate treatment.

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1. Introduction

Increasing population growth and global warming has created greater disparities between the supplies and demands of fresh water sources. Seawater and brackish water desalination technologies have been used to overcome water scarcity issues by providing reliable fresh water [1]. Major desalination technologies include Reverse Osmosis (RO), Electrodialysis Reversal (EDR), Multi-stage Flash (MSF), Multiple Effect Distillation (MED) and Vapour Compression (VC). Each has its advantages and disadvantages and the choice of which technology to use is highly dependent on the requirement at hand, the use of different energy sources and restrictions faced at the specific site. Karagiannis and Soldatos [2] conducted an extensive literature review on water desalination cost for different desalination technologies. Cost estimates seem to be very much site specific and the water production cost ranges from installation to installation because the water cost depends upon many factors including the desalination method, the level of feed water salinity, the energy source and the capacity of the desalination plant. Thermal methods such as MSF and MED are generally adopted in Gulf countries and only financially viable in the larger scale

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seawater desalination plants with high capital cost [2]. Their energy consumption is generally high regardless of the level of salt concentration and it is therefore not a viable option for brackish water desalination [3]. VC is used mainly for small systems with production around 1000 m^3/day [2]. When low cost thermal energy such as waste heat is available, these thermal processes could have operating cost advantages. For EDR, the major energy requirement is the direct current used to separate the ionic substances in the membranes stack and approximately 1 kWh electrical energy is required to extract 1 kg of salt [3]. Because the power consumption of EDR is directly proportional to the feedwater salinity, it is mostly suitable for brackish feedwaters. In the last two decades, with advances in membrane materials and improvement in energy recovery, RO technology has improved considerably and more RO plants are being constructed throughout the world [3]. RO accounts for >65% of total world desalination capacity and distillation (mainly MSF) accounts for about 30% [4].

Membrane distillation (MD) is a membrane-based thermal separation process [5, 6]. Although a membrane is involved in MD, the driving force is quite different from other membrane processes, being the vapour pressure difference across the membrane which drives the mass transfer through a membrane, rather than an applied absolute pressure difference, a concentration gradient or an electrical potential gradient. In MD, hydrophobic membranes (pore size approximately in the range of 0.1-1 µm) [5, 7] are in direct contact with the aqueous feed solutions and are employed as a barrier between the feed and the product water. MD has 100% theoretical rejection of non-volatile components and can utilize low grade heat sources of 40-80 °C to achieve the vapour pressure difference. It is a well-known process for concentrate treatment at low temperature, because MD is not significantly affected by concentration polarization as are nanofiltration (NF) and RO [5].

Compared to RO, MD does not require a high pressure feed, can tolerate complete dry out of the membrane, and can process very high salinity brines. Compared to other large thermal processes such as Multiple Stage Flash (MSF), it is easily scalable [8]. In addition, MD can be conveniently integrated with conventional RO processes to increase the recovery ratio of desalted water and/or improve the energy efficiency of the system [9], to reduce the footprint of evaporation ponds or even substitute for the evaporation ponds in processing RO concentrates. The possibility of using plastic equipment also reduces or avoids corrosion problems. Therefore, MD is a potential alternative for applications such as desalination utilizing low grade heat, concentration of thermally sensitive solutions and the treatment of wastewater of high-salt concentrations [10].

In comparison to other thermal desalination technology (i.e. MSF), the path length of the vapour phase in MD is approximately the membrane thickness (~100 µm), which is much shorter. It is potentially a commercial desalination technique if it can be combined with solar energy, geothermal energy or waste heat available in power stations or chemical plants. However, if low cost thermal energy is not available or in low supply, as a thermal distillation process, MD is also an energy intensive technique. Hence, a significant improvement of Gain Output Ratio (GOR) is required for effective production of fresh water. The economics of thermal processes with the trade-off between thermal efficiency and plant capital cost is well described in [11]. A high GOR is not always economically viable because of the added plant capital required to recycle heat. Careful thought should be put towards the cost and abundance of the thermal energy in deciding the best MD configuration and GOR.

Of the four major configurations developed for the MD process, vacuum membrane distillation (VMD) is the least studied with only about 8% of published MD references that focused on VMD [12]. In VMD, the permeate vapour does not condense in the module chamber, instead it is drawn out of the MD module by the vacuum and condenses in an external condenser. Heat conduction through the membrane in VMD is negligible in general due to the insulating nature of the vacuum on the permeate side. Thus, the thermal efficiency of the VMD is higher than direct contact MD.

In our previous studies [13,14], we developed a model to simulate hollow fiber VMD performance. The theoretical predictions were assessed experimentally to gain an understanding of the effect of various operating parameters, such as module length, feed velocity, feed temperature and vacuum pressure, on VMD performance. This paper aims to extend our previous studies to evaluate the energy requirement of VMD with or without recirculation using both theoretical results and experimental results obtained previously.

2. Experimental

2.1. Process flow diagram

Two modes of operation for VMD were considered for process engineering modelling: single pass and recirculation depending on whether the reject stream from the membrane module is discharged (single pass) or recirculated back to the feed tank (recirculation). The schematic process flow diagram for VMD in either recirculation or single pass mode is shown in Figure 1.

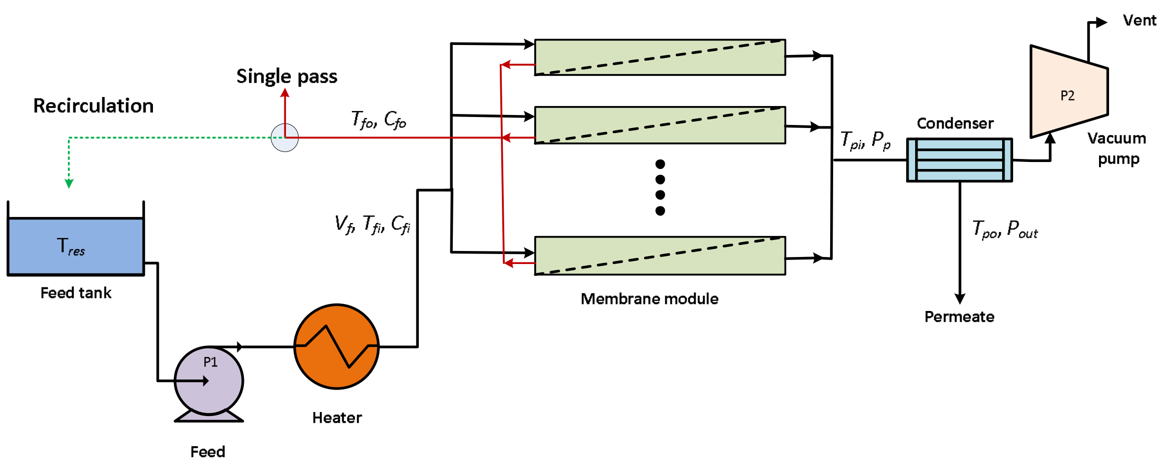

Fig. 1. Schematic of VMD in recirculation/single pass mode.

2.2. VMD testing and modelling

A recirculating VMD configuration similar to Figure 1 was used in the experimental study. The hollow fiber membranes with 40% packing density and module length of 0.5 m were used in this study. The detailed specification of the applied membrane and module configuration used has been described in detail in our previous work [9]. The feed flowed through the lumen side of the membrane. It was circulated by a pump and heated to the set temperature by a heater before entering the lumen side of the hollow fiber. The permeate was collected on the shell side of the module which was subjected to negative pressure controlled by a vacuum pump. Temperatures and pressures of feed inlet, feed outlet and module shell (permeate side) were all monitored. The flow rate of the feed stream was recorded by a flow meter and was controlled by a flow control valve. The water vapour was condensed in a heat exchanger using 3.6 ºC chilled water. Salt rejection was monitored by a conductivity meter, and was greater than 99% in all experiments.

VMD performance modelling was developed based on the membrane properties achieved through a gas permeation test [14]. In the VMD model, the sensible heat loss (<3% of latent heat) through the module wall was neglected.

2.3. Energy estimation

The energy estimation assessed on the major components for VMD are

(i) the feed heating, E_1, (ii) the feed circulation, E_2, (iii) the vacuum pump, E_3, and (iv) permeate cooling/condensation, E_4 (see Figure 2). Feed heating (E_1) and permeate cooling/condensation energy (E_4) were classified as the thermal energy, whereas the electrical power associated with the feed pump (E_2) and vacuum pump (E_4) were classified as the electrical energy. Both single pass and recirculation modes were included in this assessment.

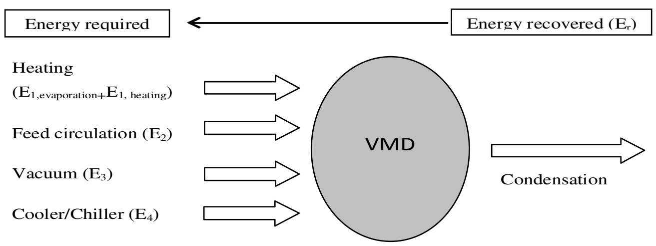

Fig. 2. Energy requirement and recovery for VMD.

The overall energy requirement is the summation of all contributions:

$$E_{Total}=E_1+E_2+E_3+E_4 \tag{1}$$

If heat recovery is included in the process, it is carried out by capturing the latent heat from the outlet permeate stream so E_r can be subtracted from E_{Total} (watt):

$$E_{Total}=E_1+E_2+E_3+E_4-E_r \tag{2}$$

The energy required to heat the feed (E_1) contributes most to the overall energy requirement in any MD configuration. There are two ways of estimating E_1; one is based on the operating conditions of the VMD (feed flow rate and inlet and outlet temperatures). For a single pass operation, the following equation is used:

$$E_1 = \dot{m}_f C_{pf}(T_{fi} - T_{Res}) \tag{3}$$

where $\dot{m}_f$ is the mass flow rate of the feed (kg/s), Cp_f is its heat capacity (J/kg K), T_{Res} and T_{fi} are the temperatures of the feed reservoir and feed inlet (K), respectively. When the feed stream is recirculated, a one-off heating (E_{init}) to increase the feed reservoir to the desired temperature from the initial temperature is required. Once the temperature of the reservoir reaches T_{fi}, a makeup stream with an additional heat will need to be accounted for, the makeup stream will have the same mass flow rate with permeate flow rate m_p, and temperature T_{Res}, hence creating the second term in equation 5.

$$E_{init}(J) = m_{Res}\, C_{pf}(T_{fi} - T_{Res}) \tag{4}$$

$$E_1 = \dot{m}_f C_{pf}\left(T_{fi} - T_{fo}\right) + m_p C_{pf}(T_{fo} - T_{Res}) \tag{5}$$

m_{Res} is the mass flow rate of the feed from reservoir to the membrane module (kg/s), m_p is the mass flow rate of the permeate flux (kg/s), T_{fo} is the temperature of the feed outlet (K).

The feed stream circulation to the membrane module is normally induced by a feed pump, P_1 [15]:

$$E_2 = \frac{\Delta P_f V_f}{\varepsilon_{p1}} \tag{6}$$

where V_f is the volumetric flow rate of feed, ε_{p1} is the pump efficiency whereas ΔP is the pressure drop due to friction determined by [16]:

$$\Delta P = f \frac{L}{D_H} \rho \frac{v^2}{2} \tag{7}$$

where f is the Darcy friction factor, L is the channel length, D_H is the hydraulic diameter, ρ is the density, and v is the linear velocity of the feed or cooling water stream. For a stream velocity in the laminar region (Re<2100), the following correlation is applied:

$$f = \frac{64}{Re} \tag{8}$$

with the R_e defined as:

$$Re = \frac{v\rho D_H}{\mu} \tag{9}$$

where ρ is the fluid density, μ is the fluid viscosity, v is the fluid linear velocity, C_p is the liquid heat capacity evaluated at bulk temperatures. The hydraulic diameter is calculated from the geometry of the flow channel.

For turbulent flows (Re>2100), the pressure drop is also affected by the roughness of the surface:

$$\sum f = f(Re, \frac{\varepsilon}{D}) + \sum(\mathbf{e_v}) \tag{10}$$

where the first term on the right hand side refers to the friction loss due to the material of the piping or tubing and can be estimated from the Moody diagram based on the knowledge of the Reynolds number and the roughness of the pipe characterize by ε/D. For common materials such as PVC or silicone a smooth surface can be assumed (ε/D~0). The second term on the right hand side of the equation (e_v) represents the minor loss due to the disturbances in the flow channel and common values for the minor loss factors can be found in [17].

In VMD, a vacuum pump is required to start the system and remove non-condensable gases from the module. At the steady state, the vacuum required could generally be achieved by condensation of the permeate. This means that the power required for the vacuum pump at steady state will be quite low as the condenser will do most of the work for maintaining the vacuum as the permeate is the water vapour which is condensable. Non-condensable vapour mainly includes air and carbon dioxide dissolved in the feed stream and air leakage from the vacuum system. The required electrical energy (power consumption) for the vacuum pump P_2 (see Figure 1) can be estimated based on the principles of adiabatic vapour expansion and contraction and related to the flow rate of non-condensable gases at steady state by the following equation[18]:

$$E_3 = \frac{m_{nc} R\, T_p}{MW \cdot \varepsilon_{p2}} \frac{\varphi}{\varphi - 1}\left[\left(\frac{p_{out}}{p_p}\right)^{(\varphi-1)/\gamma\varphi} - 1\right] \tag{11}$$

where m_{nc} is the mass flow rate of the non-condensable (kg/s), ε_{p2} is the vacuum pump efficiency, R is the universal gas constant (8.314 J/mol•K), T_p is the permeate side temperature, MW is the molecular weight of air, P_{out} is the vacuum pump exit pressure (normally atmospheric pressure), P_p is the vacuum pump inlet pressure, and φ is the adiabatic expansion coefficient defined as:

$$\varphi = \frac{C_{pp}}{C_{pv}} = \frac{C_{pp}}{C_{pp} - R_{air}} \tag{12}$$

where C_{pp} is the heat capacity of air at constant pressure, C_{pv} is the heat capacity of air at constant volume, and R_{air} is the gas constant for air (0.287 kJ/kg•K).

Apart from the energy required to condense water vapour, additional sensible heat needs to be removed to lower the temperature of water vapour. The sensible heat released in the condenser is comprised of two parts: desuperheating from T_{pi} to condensation temperature T_{pc} and subcooling from condensation temperature to T_{po}, thus the energy required to cool and condense water vapour (E_4) is calculated as:

$$E_4 = m_p \lambda + m_p \int_{T_{pi}}^{T_{pc}} C_{p,g}\, dT + m_p \int_{T_{pc}}^{T_{po}} C_{p,l}\, dT \tag{13}$$

where T_{pi} , T_{pc} and T_{po} are the permeate inlet, condensation and outlet temperatures and $C_{p,g}$ and $C_{p,l}$ are the heat capacities of water vapour and liquid, respectively.

2.4. Modelling basis

The required energy consumption by VMD was calculated using Equations 1 to 13. Theoretical and experimental results obtained in our previous study [13, 14] have been used as the basis for this process modelling.

For engineering modelling, the following assumptions have been made for estimating the energy consumptions:

- Feed reservoir temperature: 20°C
- Reservoir size: 30 times the permeate production
- Pump efficiency: 80%
- Energy required for pretreatment is beyond the boundary condition for this VMD process.
- Heat loss/exchange to/from surroundings and between equipment is neglected

3. Results & Discussion

3.1. Breakdown of energy consumption

The energy required for the VMD process is divided into thermal and electrical energy. The thermal energy consists of two parts, namely the heating component and the cooling component. The heating component can be further broken down into the initial heating for raising the reservoir temperature to the desired temperature, and intermediate reheating of the recycled feed stream to compensate for heat losses and maintaining the desired feed inlet temperature during circulation. In single pass mode, only heating of the feed stream from the temperature of feed reservoir to the desired feed inlet temperature is considered. In recirculation mode, initial heating of the feed stream is required to start up the system. At the steady state, the thermal heating mainly includes intermediate reheating to compensate for heat losses and maintain the desired feed inlet temperature. The electrical energy consists of electrical power required for the feed recirculation pump and the vacuum pump.

Figures 3 and 4 show a typical breakdown of the energy components for a VMD system under single pass and recirculation mode, respectively. In both cases, thermal energy is the most energy intensive component. In single pass mode, the required thermal energy for feed heating and permeate cooling/condensation is 2307 kWh/m^3 and 668 kWh/m^3, respectively. The heating supplied to the feed stream is much greater than the energy required for permeate cooling/condensation, indicating a very low thermal efficiency of the system as most of the heat in the feed stream will not be utilized to evaporate water through the membrane and is lost with the reject stream in single pass mode.

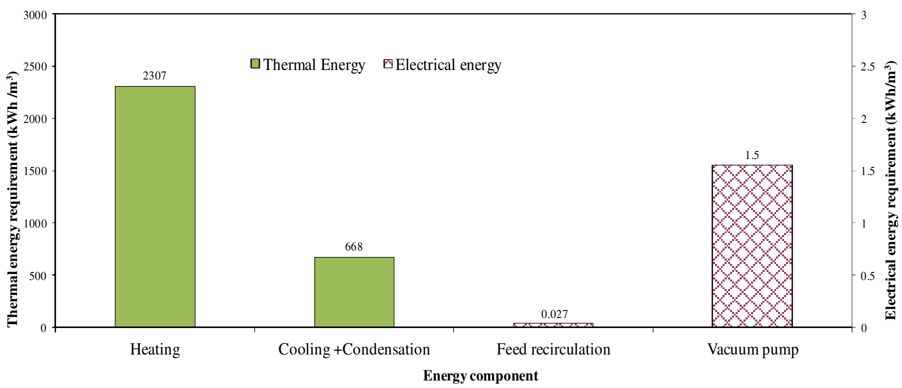

Fig. 3. Breakdown of energy components for a single pass system (feed temperature 60°C, permeate pressure 3.0 kPa, linear feed velocity 0.28 m/s and water flux 20.0 kg/m^2.h).

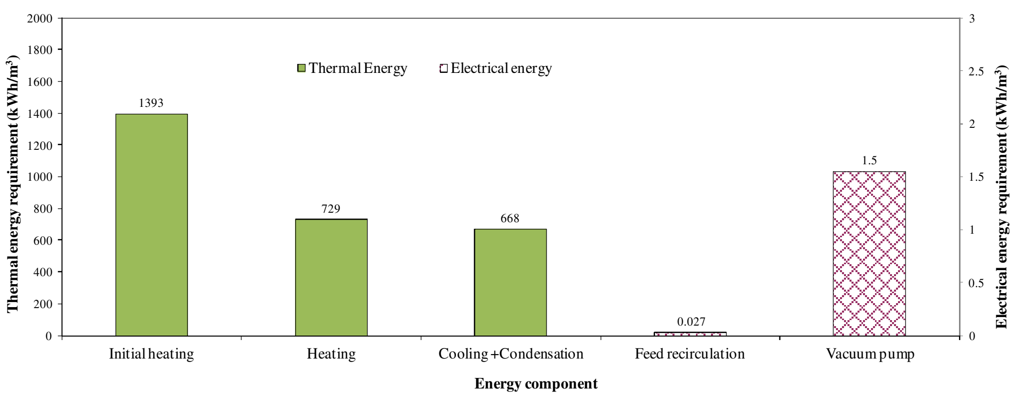

Fig. 4. Breakdown of energy components for a recirculated system (feed temperature 60°C, permeate pressure 3.0 kPa, linear feed velocity 0.28 m/s and water flux 20.0 kg/m^2.h).

For a similar system in recirculation mode, a one off initial heating energy of 1393 kWh/m^3 is required to heat up the reservoir from room temperature to the desired feed temperature of 60 °C. It is worth mentioning that this initial heating energy depends heavily on the size and initial temperature of the reservoir. The higher initial temperature of the reservoir, the lower the initial heating energy; the larger the reservoir size, the higher the initial heating. At steady state, only 729 kWh/m^3 is required to maintain a stable feed temperature.

Changing the operation mode has no effect on the electrical energy component as the vacuum pump power consumption is only related to the pump inlet pressure controlled by the cooling temperature at a given production capacity whilst the circulation pump power consumption is only affected by the change of feed flow rate and/or pressure drop. In both cases, the power consumption for the vacuum pump is significantly higher (1.5 kWh/m^3) than that of feed circulation pump (0.027 kWh/m^3).

3.2. Effect of operating conditions on energy consumption

3.2.1. Effect of feed temperature

Figure 5 shows the effect of feed inlet temperature on thermal energy required in recirculation mode by using both theoretical data and experimental results. Generally, total thermal energy displays a linear increment with feed temperature. This was mainly due to increased initial heating at higher feed temperatures. At lower temperatures (<50°C), the experimental results fit quite well with the theoretical data. However, at higher temperatures (50-75 °C), experimental results show a much higher thermal energy demand. This is understandable, because as feed temperature is increased, extra heat is required to account for heat loss to the surroundings as opposed to the theoretical model where negligible heat loss was assumed. Intermediate heating is relatively constant using theoretical data but increases more prominently with feed temperature experimentally. This discrepancy is due to the negligible heat loss assumed in the theoretical model. On the other hand, in experimental conditions, the heat loss became more severe when operating VMD at higher feed temperature as a result of a larger temperature difference between the feed and the surrounding environment.

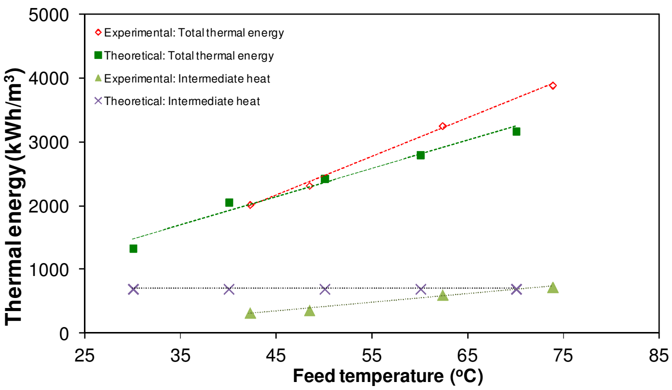

Fig. 5. Thermal energy requirement at various feed temperatures (theoretical: linear feed velocity= 0.28 m/s, permeate pressure= 3.0 kPa; experimental: linear feed velocity=0.81-0.94 m/s, permeate pressure= 2.2-4.0 kPa).

Figure 6 shows the effect of feed inlet temperature on electrical energy required by using both the theoretical data and experimental results. Both data show a downward electrical energy trend with increasing feed inlet temperature. However, the theoretical model shows a lower electrical energy requirement and also lower and steady decrement compared to the experimental model which shows a larger decrease. These discrepancies are mainly due to the differences in vacuum pressure used between the models; 3.0 kPa in the theoretical model whereas it is 2.2-4.0 kPa in the experimental model. The constant vacuum pressure used in the theoretical model gives a relative steady trend. When the feed temperature is increased, both the solubility of non-condensable gases was dissolved and the viscosity in the feed stream decreased, and consequently reduced the power required by the vacuum pump and recirculation pump, respectively. As a result, there is a slight reduction in total electrical energy required as inlet temperature is increased.

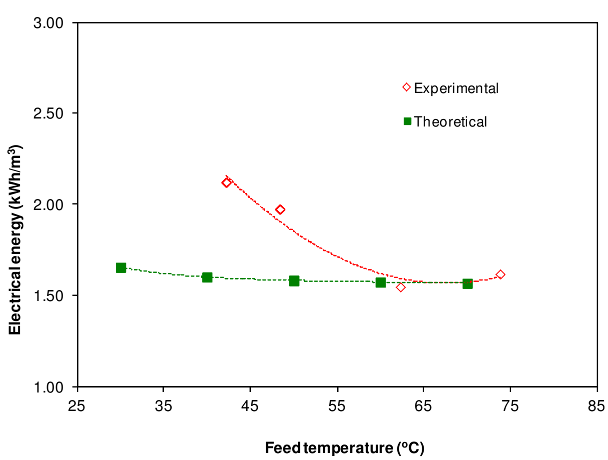

Fig. 6. Electrical energy requirement at various feed temperatures (theoretical: linear feed velocity= 0.28 m/s, permeate pressure= 3.0 kPa; experimental: linear feed velocity=0.81-0.94 m/s, permeate pressure= 2.2-4.0 kPa).

3.2.2. Effect of feed velocity

Figure 7 shows the effect of changing feed velocity on intermediate heating and total thermal energy required for both theoretical data and experimental results. In general, results using theoretical data show a relatively constant thermal energy requirement with increasing feed velocity whereas experimental results show an increasing trend in thermal energy with feed velocity. The total thermal energy requirements using experimental results are higher than those using theoretical data. This could be due to the heat loss to the membrane module and surroundings with increasing feed velocity as a result of higher average temperature in the module. On the other hand, the heat loss to the surrounding is neglected in the theoretical model. In addition, the efficiency of the heating device is anticipated to be lower for small scale laboratory systems. Therefore, higher thermal energy requirement is expected using experimental results. For intermediate heating, experimental results fit well with the theoretical data and have similar magnitude.

Figure 8 shows the effect of changing feed velocity on required electrical energy in recirculation mode. Changing the feed velocity has no impact on electrical power required for the vacuum pump but significantly affects that required for the feed recirculation pump. This is because the pressure drop in the module and along the connecting pipes becomes higher as the feed velocity increases. As a result, a higher work load for the recirculation pump and consequently, higher electrical power consumption is required (Equation 6). In this scenario, increasing feed velocity to higher than 1.7 m/s has resulted in a sharper increase of recirculation pump energy as the pressure drop is related to the square of the feed velocity (Equation 7) and increased more significantly. In addition, previous studies have found that the feed velocity only has a small influence on flux [14]. Hence, running the membrane module at higher feed velocity is not recommended as it is not beneficial to either the flux or the electrical energy requirement.

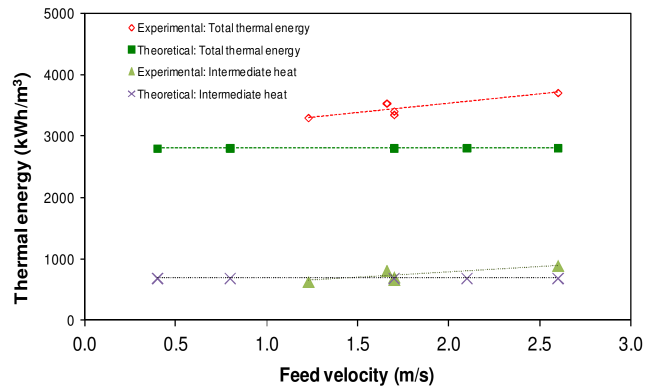

Fig. 7. Thermal energy requirement at various feed velocities (theoretical: feed temperature= 60°C, permeate pressure= 2.0 kPa; experimental: feed temperature=60-62°C, permeate pressure= 2.7-3.6 kPa).

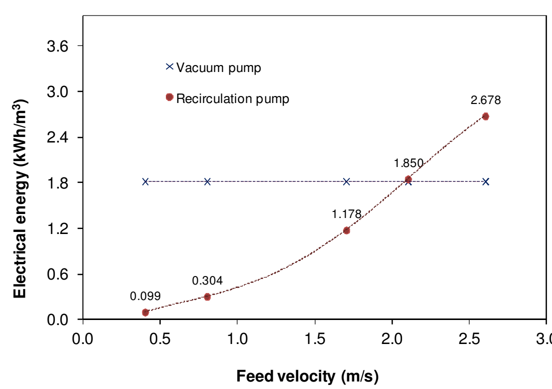

Fig. 8. Breakdown of electrical energy requirement at various feed velocities (feed temperature= 60°C, permeate pressure= 2.0 kPa).

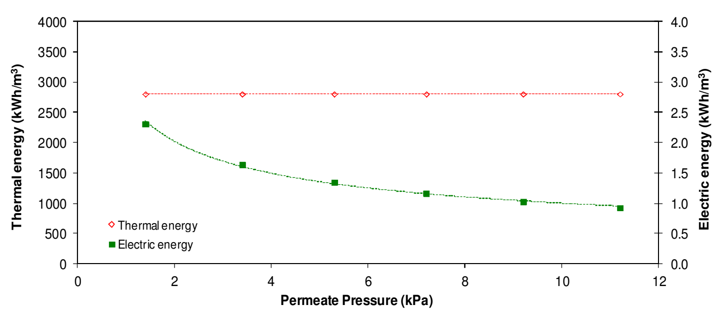

Fig. 9. Thermal and electrical energy requirement at various permeate pressures (feed temperature= 60°C, linear feed velocity=0.28 m/s).

3.2.3. Effect of permeate pressure change

Figure 9 shows the trend of thermal and electrical energies with the permeate pressure in recirculation mode. The thermal energies (including initial heating, intermediate reheating and permeate cooling/condensation) remain unchanged and the electrical energy increases with decreasing permeate pressure. At a fixed production capacity, thermal energy is only related to the feed inlet and permeate temperatures which are normally constant. Therefore, the required thermal energy remains constant. On the other hand, more work needs to be done by the vacuum pump to attain the desired vacuum pressure as the power consumption of the vacuum pump is directly related to the vacuum pump inlet pressure (Equation 11). Although it is advantageous to operate the vacuum pump at higher permeate pressure (i.e. low vacuum) to reduce the energy required, it is worth mentioning that an optimum permeate pressure needs to be chosen as the water flux decreases significantly at higher permeate pressure due to the lower driving force across the membrane.

3.3. Heat recovery and waste heat option

Thermal energy is one of the driving forces for permeation of vapour through the membrane; and a higher flux can usually be attained when more thermal energy is introduced to the feed stream. Generally, better thermal efficiency can be achieved at higher temperatures. However, running membrane distillation processes at higher temperature also means higher operating costs, and sometimes the benefits in the increment of the flux will not offset its additional expense. Therefore, measures have been sought constantly to increase the economic value of the process. The latent heat from the condensation of the permeate stream represents potential heat energy that could be recovered in the process. Options to the lower cost of the thermal energy required for VMD include 1) using a free waste heat or low grade heat source that is readily available in most of the medium to big scale power plants, 2) recovering the latent heat of condensation gained in the permeate condenser.

Figure 10 compares the effective thermal energy required for a VMD process coming from a direct electrical power source at different feed temperatures with the option of using free waste heat and/or recovering 90% of latent heat. It is obvious that the effective thermal energy from the direct heating source can be reduced by 5 and 30-fold when running the process with the options of i) free waste heat but no heat recovery and ii) free waste heat and 90% latent heat recovery, respectively at the feed temperature of 70 °C. It has been reported that the average energy consumption was 2.2-3.0 kWh/m^3 for sea water RO, 0.7-1 kWh/m^3 for brackish water RO and 1.2 kWh/m^3 for industrial effluents [19].

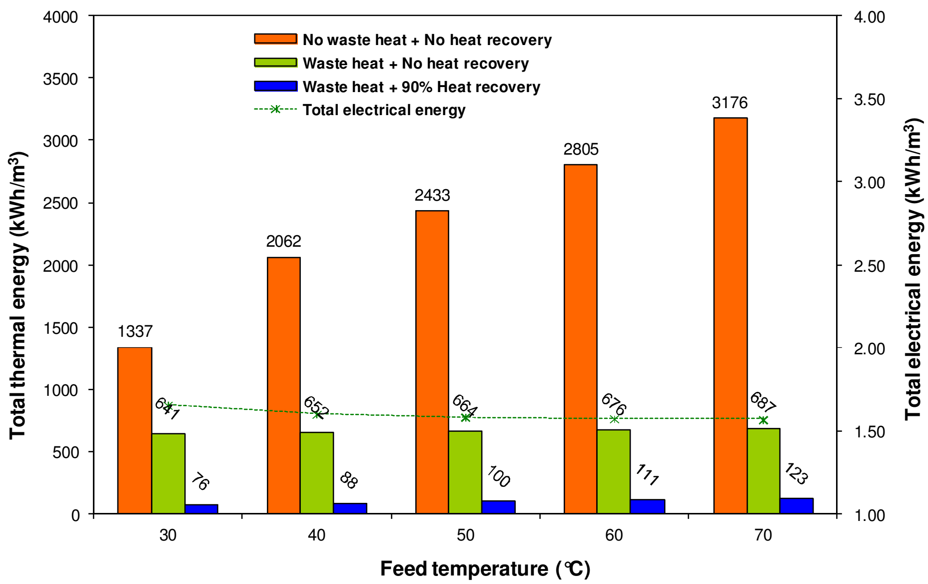

Fig. 10. Effective thermal and electrical energy requirement with/without heat recovery and alternative heat source (feed velocity 0.28 m/s, permeate pressure 3.0 kPa).

Based on our process engineering modelling result, the electrical energy consumption for VMD is slightly higher than brackish water RO but lower than that for SWRO. This is while VMD requires additional thermal energy as opposed to SWRO, which requires no thermal energy to operate. It is also worth mentioning that when compared with RO and other conventional thermal desalination processes (MED, MVC, MSF), none of them can be used for a near saturation point like in MD process. Combining these together, these results indicate that VMD may not be able to compete with RO directly, but could be used as a complimentary technology for RO such as brine concentrate treatment. Because of the limitation from osmotic pressure, a high water recovery is not attainable in RO processes. Consequently, large volumes of brines are discharged into the sea and the flow rate produced (permeate) is limited. In this regard, VMD could be used as a complementary process to RO to further concentrate RO brines and increase the overall water recovery of the process. This will significantly reduce the area required for evaporation ponds which are generally used for brine management for inland areas.

Similar findings have been previously reported. Sirkar and Song [20] built and operated a pilot DCMD plant successfully in Connecticut, US for a period of 3 months. The effective membrane surface area was up to 6.6 m^2 which yielded a 3.8 LPM permeate production rate. A detailed cost calculation was conducted for the production of 3.8 MLPD of permeate. They found that the cost of water by the DCMD process is competitive with RO if hot brine is available from existing low grade heat sources. If the cost of the concentrated brine disposal is taken into account for inland desalination applications, the economics of MD-based desalination will be even better. Their findings confirmed our conclusion that running VMD could be an economical and viable option when free or low grade waste heat sources are readily available for the feed stream.

Operating conditions such as a highly permeable membrane, high feed temperature, low permeate pressure and a turbulent fluid regime will achieve high water fluxes even for a very high salt concentration with high water recovery at reasonable energy consumption. Coupling VMD to waste heat or renewable energy such as solar energy will make this process more energy efficient and economically viable.

In addition, VMD has some significant advantages over other processes and heat and mass transfer across the membrane is the basic mass transportation and energy transformation process. Since the VMD process can concentrate and separate high concentration saline solution in a way similar to the desorption or regeneration process in the absorption refrigeration system and the concentrated solution can also act as an energy storage mode due to the transformation of heat energy into solution chemical energy [21], it could also be potentially applied to both the refrigeration system and solution chemical storage system involving energy transformation processes.

4. Conclusions

The energy requirement of VMD with or without recirculation was evaluated using both experimental results and theoretical data obtained in our previous studies. Most of the trends are consistent between theory and experimental data. It is generally more energy efficient to operate the VMD in recirculation mode than single pass mode. Single pass mode operation will only have an advantage at low feed temperatures where the initial heating is not required and a waste heat source is not available.

In both single pass and recirculation mode, thermal energy requirement is significant and contributes the most to the total energy required for VMD which makes the VMD an energy intensive process. To lower the cost accrued from thermal energy requirement, a free low grade waste heat resource and heat recovery of latent heat from the permeate vapour are needed. The latent heat of condensation of the permeate stream indicates that the potential heat energy could be recovered in the process. With the option of using free waste heat and recovering 90% of latent heat of condensation from the permeate stream, the effective thermal energy requirement from the direct heating source could be reduced significantly, <76 kWh/m^3 at a feed temperature of 30 °C.

In addition to the thermal energy requirement, the electrical energy required for VMD is about 1.5 kWh/m^3 in both single pass and recirculation mode at a feed temperature of 60 °C and permeate pressure of 3 kPa, which is slightly higher than brackish water RO but lower than sea water RO. These results indicate that VMD will not be able to compete with RO directly but could be used as a complimentary process to RO such as brine concentrate treatment to maximize the water recovery and minimize the brine discharge. Moreover, it could also be potentially applied to both the refrigeration system and solution chemical storage system involving energy transformation processes.

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