Desalination and Water Treatment
www.deswater.com

doi: 10.1080/19443994.2016.1152648

57 (2016) 24662–24673
November

Evaluation of economic feasibility of reverse osmosis and membrane distillation hybrid system for desalination

Yong-Jun Choi[a], Sangho Lee[a,*], Jaewuk Koo[b], Seung-Hyun Kim[c]

[a]*School of Civil and Environmental Engineering, Kookmin University, Jeongneung-Dong, Seongbuk-Gu, Seoul 136-702, Korea, Tel. +82 2 910 5060; email: choiyj1041@gmail.com (Y.-J. Choi), Tel. +82 2 910 4529; Fax: +82 2 910 4939; email: sanghlee@kookmin.ac.kr (S. Lee)*
[b]*Department of Construction Environmental Research, Korea Institute of Construction Technology, 2311 Daehwa-Dong, Ilasan-gu, Kyonggi-do, Republic of Korea, Tel. +82 31 910 0755; email: koojaewuk@naver.com*
[c]*School of Civil Engineering, Kyungnam University, Woryeong-dong, Masanhappo-gu, Changwon-si, Gyeongsangnam-do, 631-701, Republic of Korea, Tel. +82 2 3700 0798; email: shkim@kyungnam.ac.kr*

Received 23 October 2015; Accepted 8 January 2016

ABSTRACT

This study seeks to evaluate the economic feasibility of membrane distillation (MD) and reverse osmosis (RO)-MD hybrid system for seawater desalination. A theoretical cost model was applied to analyze the effects of flux, recovery, membrane properties, and energy price on RO, MD, and RO-MD hybrid system. The simulation results showed that MD stand-alone system and RO-MD hybrid system can be cost-competitive compared with RO systems when the recovery and flux of MD system are higher than those of RO system and the steam cost is relatively cheap. It is also revealed that the water costs of RO-MD hybrid system and RO system are same under similar operating conditions, but the water cost of MD stand-alone system is higher. The effect of thermal energy cost on water cost for MD and RO-MD systems was also analyzed. Based on these results, guidelines for an analysis of economic feasibility of MD and RO-MD were suggested.

Keywords: Desalination; Cost; Model; Reverse osmosis; Membrane distillation; Hybrid system

1. Introduction

Only about 0.5% of the overall global water is available as fresh water, while seawater accounts for about 97% of them. Approximately 41% of the world population live in the arid regions, thus fresh water shortage is becoming a worldwide problem [1,2]. Accordingly, many countries have pointed out alternative sources of fresh water. Among them, seawater desalination has been proved to be a reliable and economically sustainable water source [3]. Over the past few decades, a number of technologies have been developed, including thermal distillation (multi-stage flash distillation; MSF, multi-effect distillation; MED, mechanical vapor pressure compression distillation;

*Corresponding author.

Presented at 2015 Academic Workshop for Desalination Technologyheld in the Institute for Far Eastern Studies Seoul, Korea, 23 October 2015

MVC), membrane separation (reverse osmosis; RO, nanofiltration; NF), freezing and electrodialysis [4]. Of particular interest is reverse osmosis (RO), which becomes a dominant technology [5,6]. In 2009, over 15,000 desalination plants were in operation worldwide with approximately half of them being RO plants [7]. Nevertheless, RO technology also has drawbacks such as high electricity consumption and low recovery ratio of product water [8].

Recently, membrane distillation (MD) has drawn attention as a novel technology to overcome the drawbacks of RO technology. MD is a separation process using a vapor pressure, which results from the temperature difference between feed and permeate water [9]. The hydrophobic microporous membrane facilitates the transport of water vapor through its pores, while maintaining vapor–liquid interfaces at the pore entrance, but it does not participate in the actual separation process. MD has several advantages compared to RO and other desalination processes for the treatment of saline water and wastewater [10–12]. Because water is transported through the membrane only in a vapor phase, MD can offer complete rejection of all non-volatile constituents in the feed solution; thus, almost 100% rejection of ions, dissolved non-volatile organics, colloids, and pathogenic micro-organisms can be achieved via the MD process. But more importantly, due to the discontinuity of the liquid phase across the membrane, water flux in MD is not influenced by the osmotic pressure gradient across the membrane. Consequently, the greatest potential of MD can be realized through the treatment of highly saline solutions [10].

Nevertheless, MD is still in its early stage in terms of commercial applications [13]. Therefore, further work is still required to bring MD technologies into practice, including development of new MD membranes, design of optimization systems and development of heat exchange systems. In this context, this study aims to evaluate the economic feasibility of the MD and RO-MD hybrid system and to propose a guideline by which MD process might be more price-competitive in the field. To do this, theoretical analysis was carried out using a simple cost model to investigate the effects of flux, recovery, membrane, and energy price on the cost of water production by different desalination technologies.

2. Cost model

In order to analyze the effects of major parameters such as flux, recovery, membrane, and energy cost on RO, MD, and RO-MD hybrid system, a set of cost functions were used [14–19]. Based on these cost function, a theoretical model was developed to evaluate the economics of RO, MD, and RO-MD hybrid system.

2.1. RO cost model

The RO system is made of five major parts: The seawater intake and pre-treatment process, high-pressure pump, booster pump, RO membrane module and energy recovery device. The capital and operating costs of the intake and pretreatment are expressed as follows [17–19].

$$\mathrm{CC_{IP}}\ (\$) = 997(Q_{\mathrm{IP}}\ (\mathrm{m^3/d}))^{0.8} \tag{1}$$

$$\mathrm{OC_{IP}}\ (\$/\mathrm{d}) = \frac{0.028P_{\mathrm{IP}}\ (\mathrm{bar})\ Q_{\mathrm{IP}}\ (\mathrm{m^3/d})\ D_{\mathrm{EP}}\ (\$/\mathrm{kWh})}{\eta_{\mathrm{IT\_S}}} \times \mathrm{PLF} \tag{2}$$

where CC and OC denote the capital cost and operating cost, respectively. The subscript, IP denotes the intake and pre-treatment. Q_{IP} is the intake and pretreatment flow rate, P_{IP} is the pressure of intake and pretreatment, D_{EP} is the unit electricity price, η_{IP} is the efficiency of the intake and pretreatment pump and PLF is the plant load factor.

The capital and operating costs of high-pressure pump, booster pump, and energy recovery device are expressed as follows [17–19].

$$\mathrm{CC_{HP}}\ (\$) = \frac{Q_{\mathrm{f}}\ (\mathrm{m^3/d})(393000 + 10710P_{\mathrm{f,in}}\ (\mathrm{bar}))}{30000} \tag{3}$$

$$\mathrm{OC_{HP}}\ (\$/\mathrm{d}) = \frac{0.028P_{\mathrm{f,in}}\ (\mathrm{bar})\ Q_{\mathrm{f}}\ (\mathrm{m^3/d})\ D_{\mathrm{EP}}\ (\$/\mathrm{kWh})}{\eta_{\mathrm{P\_HP}}} \times \mathrm{PLF} \tag{4}$$

$$\mathrm{CC_{BP}}\ (\$) = \frac{(Q_{\mathrm{f}}\ (\mathrm{m^3/d}) - Q_{\mathrm{p}}\ (\mathrm{m^3/d}))(393000 + 10710P_{\mathrm{f,in}}\ (\mathrm{bar}))}{60000} \tag{5}$$

$$\mathrm{OC_{BP}}\ (\$/\mathrm{d}) = \frac{0.028((P_{\mathrm{f,in}} - P_{\mathrm{f,out}})\ (\mathrm{bar}))\ \eta_{\mathrm{ERD}}\ (Q_{\mathrm{f}} - Q_{\mathrm{p}})\ (\mathrm{m^3/d})\ D_{\mathrm{EP}}}{\eta_{\mathrm{P\_BP}}} \times \mathrm{PLF} \tag{6}$$

$$\mathrm{CC_{ERD}}\ (\$) = \frac{(Q_\mathrm{f}\ (\mathrm{m^3/d}) - Q_\mathrm{p}\ (\mathrm{m^3/d}))(393000 + 10710 P_\mathrm{f,in}\ (\mathrm{bar}))}{40000} \tag{7}$$

where subscripts, HP, BP, and ERD denote the high-pressure pump, booster pump, and energy recovery device, respectively. Q_f is the feed flow rate, Q_p is the permeate flow rate, $P_\mathrm{f,in}$ is the inlet pressure of RO vessel, and $P_\mathrm{f,out}$ is the outlet pressure of RO vessel.

Assuming that the capital cost of the membrane module is linear to the membrane area, the annualized capital cost of the membrane is calculated by the following [20]:

$$\mathrm{CC_{Mem}}\ (\$) = \mathrm{Area_{Mem}}\ C_\mathrm{mem}\ (\$/\mathrm{m^2}) \tag{8}$$

where subscripts, Mem denote the membrane, $\mathrm{Area_m}$ is the total membrane area, and C_Mem is the unit membrane cost.

The total capital cost is composed of the direct capital cost and the indirect capital cost. The direct capital cost is the sum of the cost for plant equipment and the cost for site development, which is set at 20% of [17]. The indirect capital cost is set at 30% of the direct capital cost [17]. The annualized capital cost is calculated using Chauvel [21]. The total and annual capital costs of RO process are expressed as follows [17–19,21]:

$$\mathrm{CC_{Equipment}}\ (\$) = \mathrm{CC_{IP}} + \mathrm{CC_{HP}} + \mathrm{CC_{BP}} + \mathrm{CC_{ERD}} + \mathrm{CC_{Mem\_RO}}\ \mathrm{CC_{Mem}} \tag{9}$$

$$\mathrm{CC_{Site}}\ (\$) = \mathrm{CC_{Equipment}} \times 0.2 \tag{10}$$

$$\mathrm{DCC_{RO}}\ (\$) = \mathrm{CC_{Equipment}} + \mathrm{CC_{Site}} \tag{11}$$

$$\mathrm{ICC_{RO}}\ (\$) = \mathrm{DCC_{RO}} \times 0.3 \tag{12}$$

$$\mathrm{TCC_{RO}}\ (\$) = \mathrm{DCC_{RO}} + \mathrm{ICC_{RO}} \tag{13}$$

$$\mathrm{ACC_{RO}}\ (\$/\mathrm{y}) = \mathrm{TCC_{RO}} \frac{i(1+i)^n}{(1+i)^n - 1} \tag{14}$$

where $\mathrm{DCC_{RO}}$ is the direct capital cost, $\mathrm{ICC_{RO}}$ is the indirect capital cost, $\mathrm{TCC_{RO}}$ is the total capital cost, $\mathrm{ACC_{RO}}$ is the annual capital cost, i is the interest rate, and n is the plant lifetime.

The annual operating cost is composed of the annual power cost, annual membrane replacement cost and other cost (e.g. labor, chemicals, and maintenance). The annual operating costs of RO process are expressed as follows [17–19].

$$\mathrm{OC_{Power}}\ (\$/\mathrm{y}) = (\mathrm{OC_{IP}} + \mathrm{OC_{HP}} + \mathrm{OC_{BP}}) \times 365 \tag{15}$$

$$\mathrm{OC_{MR}}\ (\$/\mathrm{y}) = \mathrm{CC_{Mem}} \times 0.2 \tag{16}$$

$$\mathrm{OC_{etc}}\ (\$/\mathrm{y}) = \mathrm{AOC_{RO}} \times 0.3 \tag{17}$$

$$\mathrm{AOC_{RO}}\ (\$/\mathrm{y}) = \mathrm{OC_{power}} + \mathrm{OC_{MR}} + \mathrm{OC_{etc}} \tag{18}$$

where $\mathrm{OC_{Power}}$ is the annual power cost, $\mathrm{OC_{MR}}$ is the annual membrane replacement cost, and $\mathrm{OC_{etc}}$ is the other cost.

Finally, the water cost of RO process is calculated as follows:

$$\mathrm{WC_{RO}}\ (\$/\mathrm{m^3}) = (\mathrm{ACC_{RO}} + \mathrm{AOC_{RO}})/(365 \times Q_\mathrm{p} \times \mathrm{PLF}) \tag{19}$$

2.2. MD cost model

MD process has four different configurations, including direct contact MD, air gap MD, sweeping gas MD, and vacuum MD (VMD). Among them, the VMD configuration was selected. In VMD, vacuum is applied in the permeate side of the membrane module by means of vacuum pump. As vacuum is applied on the permeate side, the distillate production rate increased and thermal energy loss is reduced [10].

The VMD system is made of five major parts: the seawater intake and pre-treatment process, vacuum pressure pump, feed pump, VMD membrane module, and heat exchanger. The capital and operating costs of the intake and pretreatment are expressed as follows [17–19].

$$\mathrm{CC_{IP}}\ (\$) = 997(Q_\mathrm{IP}\ (\mathrm{m^3/d}))^{0.8} \tag{20}$$

$$\mathrm{OC_{IP}}\ (\$/\mathrm{d}) = \frac{0.028 P_\mathrm{IP}\ (\mathrm{bar})\ Q_\mathrm{IP}\ (\mathrm{m^3/d})\ D_\mathrm{EP}\ (\$/\mathrm{kWh})}{\eta_\mathrm{IT\_S}} \times \mathrm{PLF} \tag{21}$$

The capital and operating costs of vacuum pressure pump, feed pump, and heat exchanger are expressed as follows [17–19,22].

$$\mathrm{CC_{VP}}\ (\$) = \frac{Q_\mathrm{f}\ (\mathrm{m^3/d})(393000 + 10710))}{12000 P_\mathrm{VP}\ (\mathrm{mbar})} \tag{22}$$

$$\mathrm{OC_{VP}}\ (\$/\mathrm{d}) = 1.0\left[\frac{\mathrm{kWh}}{m^3}\right] Q_\mathrm{p}\ (\mathrm{m^3/d})\ D_\mathrm{EP}\ (\$/\mathrm{kWh}) \times \mathrm{PLF} \tag{23}$$

$$\mathrm{CC_{FP}}\ (\$) = 4.78 \times 10^{-6} \times Q_\mathrm{f}\ (\mathrm{m^3/d}) \times 120000 \tag{24}$$

$$\mathrm{OC_{FP}}\ (\$/\mathrm{d}) = \frac{0.028(P_\mathrm{f,in}\ (\mathrm{bar}))\ (Q_\mathrm{f}\ (\mathrm{m^3/d}))\ D_\mathrm{EP}}{\eta_\mathrm{FP}} \times \mathrm{PLF} \tag{25}$$

$$\mathrm{CC_{HX}}\ (\$) = 1000 \times \mathrm{Area_{HX}} \tag{26}$$

where subscripts, VP, FP, HX denote the vacuum pressure pump, feed pump, and heat exchanger, P_VP is the vacuum pressure of VMD vessel and $\mathrm{Area_{HX}}$ is the total heat-exchanger area.

Assuming that the capital cost of the membrane module is linear to the membrane area, the annualized capital cost of the membrane is calculated by the following same as RO membrane module [20]:

$$\mathrm{CC_{Mem}}\ (\$) = \mathrm{Area_{Mem}}\ C_\mathrm{mem}\ (\$/\mathrm{m^2}) \tag{27}$$

The total and annual capital costs of VMD system are expressed as follows:

$$\mathrm{CC_{Equipment}}\ (\$) = \mathrm{CC_{IP}} + \mathrm{CC_{VP}} + \mathrm{CC_{FP}} + \mathrm{CC_{HX}} + \mathrm{CC_{Mem}} \tag{28}$$

$$\mathrm{CC_{Site}}\ (\$) = \mathrm{CC_{Equipment}} \times 0.2 \tag{29}$$

$$\mathrm{DCC_{MD}}\ (\$) = \mathrm{CC_{Equipment}} + \mathrm{CC_{Site}} \tag{30}$$

$$\mathrm{ICC_{MD}}\ (\$) = \mathrm{DCC_{RO}} \times 0.3 \tag{31}$$

$$\mathrm{TCC_{MD}}\ (\$) = \mathrm{DCC_{MD}} + \mathrm{ICC_{MD}} \tag{32}$$

$$\mathrm{ACC_{MD}}\ (\$/\mathrm{y}) = \mathrm{TCC_{MD}}\frac{i(1+i)^n}{(1+i)^n - 1} \tag{33}$$

The annual operating costs of VMD system are expressed as follows:

$$\mathrm{OC_{Steam}}\ (\$/\mathrm{d}) = (\mathrm{Steam}\ (\mathrm{kg/day}) \times C_\mathrm{Steam}\ (\$/\mathrm{kg})) \tag{34}$$

$$\mathrm{OC_{Power}}\ (\$/\mathrm{y}) = (\mathrm{OC_{IP}} + \mathrm{OC_{VP}} + \mathrm{OC_{FP}} + \mathrm{OC_{Steam}}) \times 365 \tag{35}$$

$$\mathrm{OC_{MR}}\ (\$/\mathrm{y}) = \mathrm{CC_{Mem}} \times 0.2 \tag{36}$$

$$\mathrm{OC_{etc}}\ (\$/\mathrm{y}) = \mathrm{AOC_{MD}} \times 0.3 \tag{37}$$

$$\mathrm{AOC_{MD}}\ (\$/\mathrm{y}) = \mathrm{OC_{power}} + \mathrm{OC_{MR}} + \mathrm{OC_{etc}} \tag{38}$$

where $\mathrm{OC_{Steam}}$ is the daily steam cost and C_Steam is the unit steam cost.

Finally, the water cost of VMD system is as follows:

$$\mathrm{WC_{MD}}\ (\$/\mathrm{m^3}) = (\mathrm{ACC_{MD}} + \mathrm{AOC_{MD}})/(365 \times Q_\mathrm{p} \times \mathrm{PLF}) \tag{39}$$

3. Results and discussion

The economics of seawater desalination were theoretically evaluated for the following three systems:

(1) RO stand-alone system (water production capacity of 50,000 $\mathrm{m^3/d}$).
(2) MD stand-alone system (water production capacity of 50,000 $\mathrm{m^3/d}$).
(3) RO-MD hybrid system (water production capacity of 50,000 $\mathrm{m^3/d}$).

In the RO-MD hybrid system, the RO brine is treated by the MD system to increase overall system recovery ratio. Model parameters and operating conditions used in this analysis are summarized in Table 1.

Table 2 shows calculation results for a RO plant using the parameters and operating conditions in Table 1. For this calculation, the permeate flux, feed pressure, and recovery were set to 12 LMH, 55.0 bar, and 40%, respectively, as shown in Table 1. The recovery and pressure of intake and pretreatment were set to 90% and 5.0 bar, respectively.

As a result of the calculations, the water cost of RO system was found to be 0.75 $\$/\mathrm{m^3}$. The cost of the RO plant construction was 43% from the total water cost shown as Fig. 1. The cost of energy, membrane replacement, labor and chemical, filters, and miscellaneous items(C.F.M) of the RO plant construction was 27, 13, 11, and 6%, respectively, from the total water cost.

The two methods were used to performed economic evaluation for 50,000 $\mathrm{m^3/d}$ MD and RO-MD hybrid desalination systems. The recovery, flux, steam cost, and membrane cost of MD system were determined as most important factors to influence to water cost. In the first method, the recovery and flux of MD

Table 1
Process parameters and operating conditions

Parameter	Value	Parameter	Value
RO		*MD*	
Membrane area	40 m^2	Membrane area	20 m^2
Permeate flux	12 LMH	Permeate flux	10–30 LMH
Feed pressure	55.0 bar	Feed pressure	1.0 bar
Recovery	40%	Recovery	20–60%
Feed TDS	35,000 mg/L	Feed TDS	35,000 mg/L
Efficiency		*Efficiency*	
Intake & pretreatment pump	80%	Intake & pretreatment pump	80%
High-pressure pump	75%	Vacuum pump	–
Booster pump	80%	Feed pump	80%
Energy recovery device	90%	Heat exchanger	90%
Cost		*Cost*	
Electricity bill	0.08 $/kWh	Electricity bill	0.08 $/kWh
Membrane cost	50 $/$m^2$	Membrane cost	30–100 $/$m^2$
Plant load factor	0.91	Steam cost	0–5 $/ton
		Plant load factor	0.91
Interest rate	0.08	Interest rate	0.08
Plant life	20 years	Plant life	20 years

Table 2
Calculation results for a 50,000 m^3/d RO system

Energy consumption	
Intake and pretreatment	1,004 kW
High-pressure pump	4,242 kW
Booster pump	468 kW
Specific energy	2.74 kWh/m^3
Capital cost	
Intake and pretreatment	10,222,097 $
High-pressure pump	11,784,600 $
Booster pump	3,074,683 $
Energy recovery device	3,375,325 $
Membrane	8,694,000 $
Site	7,430,140 $
Direct capital cost	44,580,845 $
Indirect capital cost	13,374,253 $
Capital cost	57,955,099 $
Annual capital cost	5,902,854 $/year
Operating cost	
Power cost	3,604,557 $/year
Membrane replacement	1,738,800 $/year
Chemicals, filters, and miscellaneous items	763,337 $/year
Labor	1,526,673 $/year
Annual operating cot	7,633,400 $/year
Water cost	
Water cost	0.75 $/$m^3$

system were adjusted, while the other factors including steam cost and membrane cost of MD system were fixed. The recovery changed from 20 to 60% and the permeate flux changed from 10 to 30 LMH. The steam cost and membrane cost was set to 1 $/ton and 50 $/$m^2$, respectively. The other model parameters and operating conditions were listed in Table 1.

On the other hand, in the second method, the steam cost and membrane cost were adjusted and the other factors including recovery and permeate flux were fixed. The steam cost changed from 1 to 5 $/ton and the membrane cost changed from 30 to 100 $/$m^2$. The recovery and permeate flux was set to 40% and 12 LMH, respectively, same as RO system. In both cases, the permeate flux was not coupled with operating temperature in MD system and depended on the membrane properties.

In MD system, flux and recovery are two main operating parameters that determine the process performance. Accordingly, the effect of these parameters on water cost of MD was investigated as shown in Fig. 2. The water cost of MD system was estimated to vary from 0.6 to 1.5 $/$m^3$, by the changes in the permeate flux and recovery. The water cost of MD system decreases with increasing recovery and permeate flux due to decreasing the requirement of large capacity intake and pretreatment, thermal energy, and the required membrane area. Especially, decreasing recovery from 60 to 20% significantly affects the water cost.

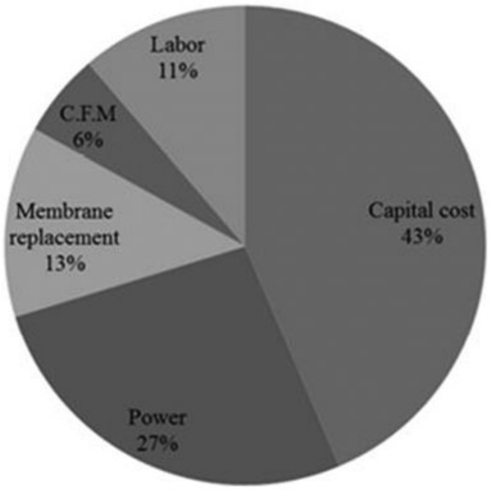

Fig. 1. Percent of water cost of 50,000 m^3/d RO plant.

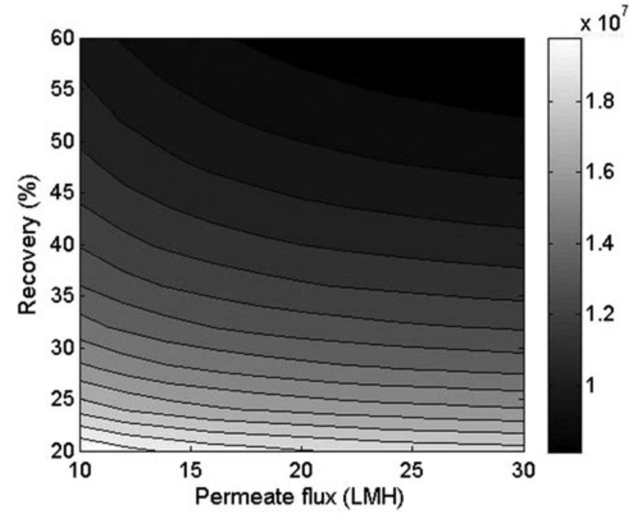

Fig. 2. Contours of water cost ($/$m^3$) of MD system at different permeate flux and recovery.

Considering the fact that the water cost of RO system was calculated as 0.75 $/$m^3$, the permeate flux and recovery should exceed 12 LMH and 45%, respectively, to make the water cost of MD system cheaper than that of RO system.

In addition to water cost, the annual capital cost was considered at different permeate flux and recovery in Figs. 3 and 4. The overall trend of the annual capital cost and operating cost contours are same as that of water cost. The recovery is inversely proportional to capital and operating costs of intake and pretreatment process, vacuum pressure pump, feed pump, and steam cost. The permeate flux is also

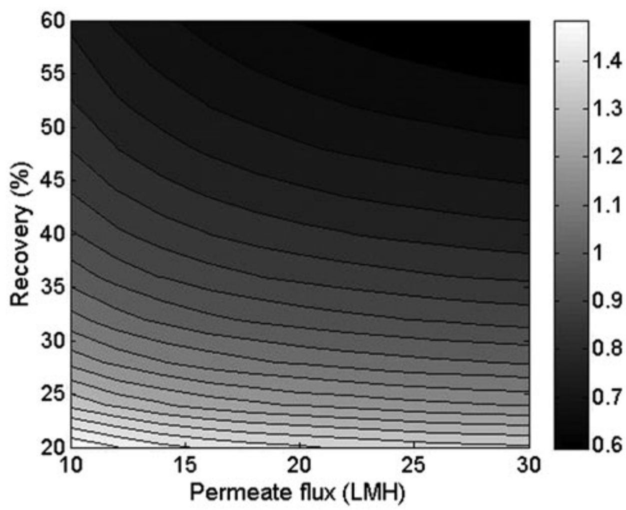

Fig. 3. Contours of annual capital cost ($/year) of MD system at different permeate flux and recovery.

inversely proportional to membrane cost and membrane replacement cost.

At the same recovery (40%) and permeate flux (12 LMH) conditions to RO, the capital cost of MD system decreased down to approximately 24% since the sum of the capital cost of high-pressure pump, booster pump, and ERD are higher than the sum of the capital cost of vacuum pressure pump, feed pump, and heat exchanger under these simulation conditions. However, the operating cost increased to approximately 60%, since the energy cost of MD system is much higher than that of RO process.

Fig. 5 illustrates contours of gained output ratio (GOR) of MD system at different permeate flux and recovery. The GOR is a measure of how much thermal energy is consumed in a desalination system [22]. The simplest definition is as follows; amount of product water/amount of used stream. Typically, the value of GOR ranges from 1.0 to 10. Higher GOR systems consume less energy, therefore, have lower operating costs. The GOR of MD system was estimated to range from 2.0 to 5.5, by changing the permeate flux and recovery. In addition, the GOR of MD system increases with increasing the recovery of the product water. According to the simulation results, the GOR was found to be related to the recovery.

The contours of the power cost of MD plant at different permeate flux and recovery are shown in Fig. 6. The power cost of MD plant significantly decreases by increasing recovery from 20 to 60%. In fact, the power cost of MD system is calculated from the sum of energy consumption of intake and pretreatment, vacuum pump, feed pump, and heat energy (steam).

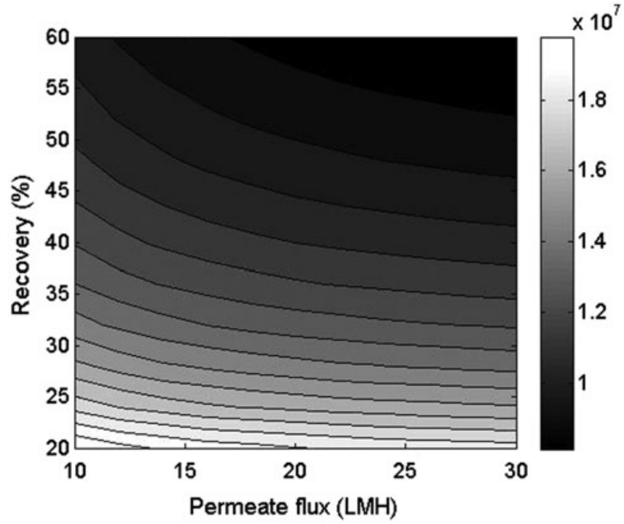

Fig. 4. Contours of annual operating cost (\$/year) of MD system at different permeate flux and recovery.

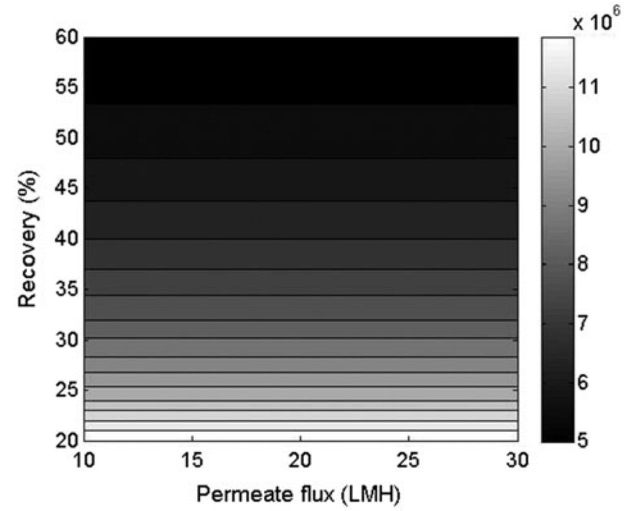

Fig. 6. Contours of power cost (\$/year) of MD system at different permeate flux and recovery.

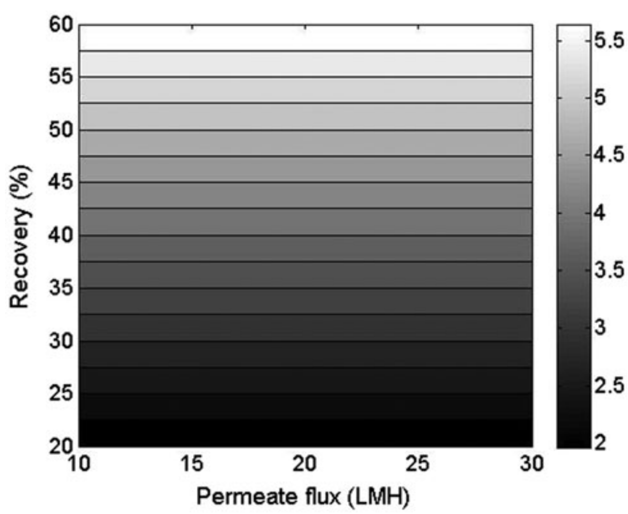

Fig. 5. Contours of GOR of MD system at different permeate flux and recovery.

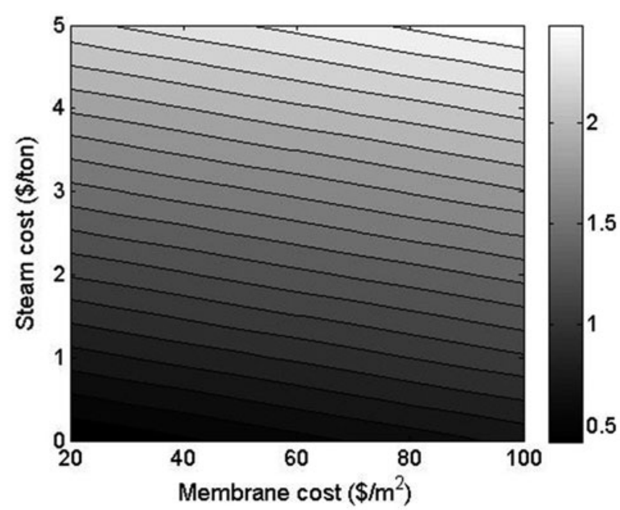

Fig. 7. Contours of water cost (\$/m^3) of MD system at different membrane and steam cost.

In this study, operating pressure of intake and pretreatment and feed pump were set to 5.0 and 1.0 bar, respectively. In addition, the energy consumption of vacuum pump was assumed to 1.0 kWh/m^3. Accordingly, the power cost of MD system decreases considerably with increasing recovery since the amount of feed water to be heated is reduced.

In VMD system, thermal energy is provided in the form of steam, which has the latent heat. Accordingly, the steam cost is one of the important factor affecting the water cost of MD system. Fig. 7 shows contours of water cost of MD plant at different membrane and steam cost. As expected, the water cost of MD system is sensitive to the steam cost. The water cost of MD plant was found to range from 0.4 to 2.4 \$/m^3, with the changes in the steam cost and the membrane cost. Based on the simulation results, the membrane and steam cost should not exceed 0.5 \$/ton 70 \$/m^2, respectively, to make the MD system more cost-competitive than the RO system,

The dependence of annual capital cost and annual operating cost of MD system on membrane and steam

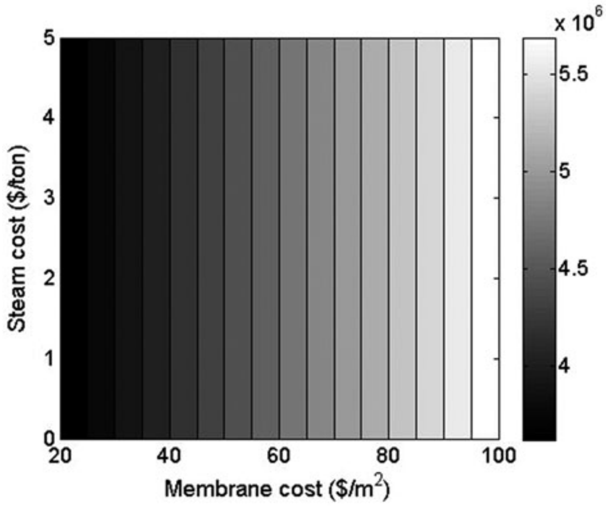

Fig. 8. Contours of annual capital cost ($/year) of MD system at different membrane and steam cost.

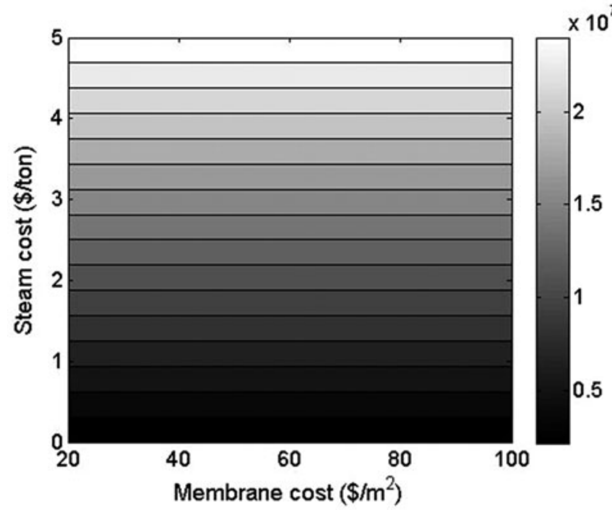

Fig. 10. Contours of annual power cost ($/year) of MD system at different membrane and steam cost.

cost is demonstrated as contour plots in Figs. 8 and 9. The annual capital cost is directly proportional to membrane cost, but the steam cost has no direct effect on the capital cost. The trend of the operating cost contours is same as water cost contours. Annual capital cost is inversely proportional to membrane and steam cost. Under the same conditions, the operating cost is higher than capital cost due to consumption of significant amount of thermal energy.

The contours of power cost of MD plant as a function of membrane and steam cost are shown in Fig. 10. As mentioned above, the power cost is considered to be the sum of energy consumption of intake and pretreatment, vacuum pump, feed pump, and heat energy (steam). The power cost is not affected by the membrane cost. The power cost of MD system increases significantly with increasing steam cost. The power cost of MD system increases from 2,174,000 to 25,429,000 $/year with an increase in steam cost from 0 to 5 $/ton.

The water cost of RO-MD hybrid system is compared with that of the stand-alone MD system. The

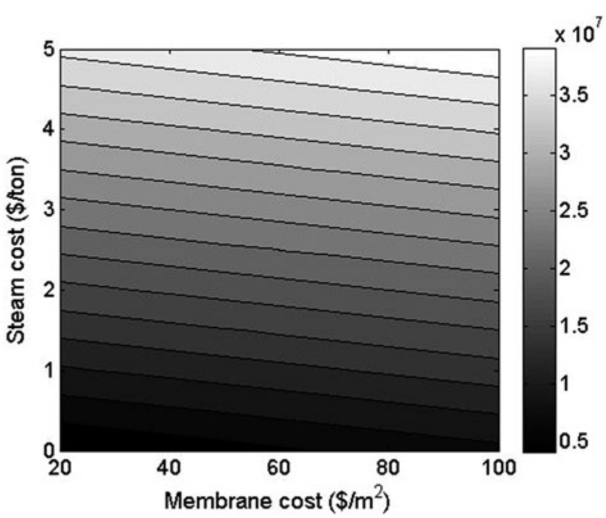

Fig. 9. Contours of annual operating cost ($/year) of MD system at different membrane and steam cost.

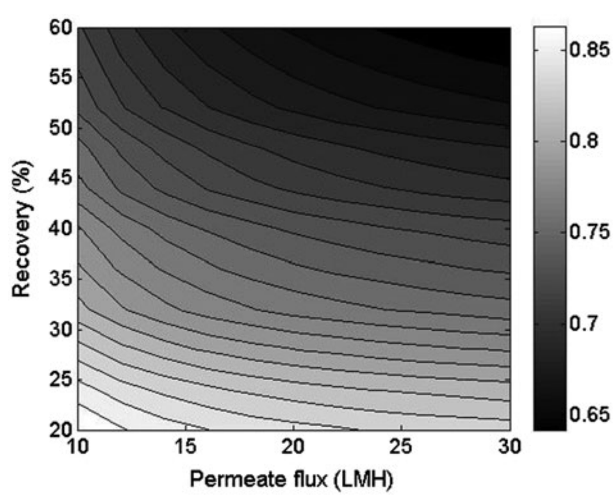

Fig. 11. Contours of water cost ($/m^3) of RO-MD hybrid plant at different permeate flux and recovery.

results are shown in Fig. 11, which showed water cost of RO-MD hybrid system under different permeate flux and recovery of MD system. In this calculation, the permeate flux, feed pressure and recovery of RO process were set to 12 LMH, 55.0 bar, and 40%, respectively. Moreover, the recovery of MD process changed from 20 to 60% and the permeate flux changed from 10 to 30 LMH. The steam cost and membrane cost were set to 1 \$/ton and 50 \$/m^2, respectively. The simulation results showed that the water cost of RO-MD hybrid system ranges from 0.65 to 0.85 \$/m^3 with the changes in the permeate flux and recovery of MD system. The water cost decreases with increasing recovery and permeate flux of MD system due to the reduction in the capacity of intake and pretreatment, thermal energy, and the required membrane area.

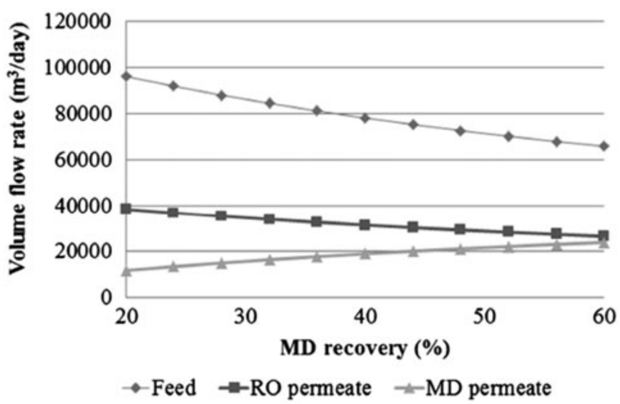

Fig. 12. Dependence of feed flow rate, RO permeate, and MD permeate on MD recovery in RO-MD hybrid system.

The changes in the feed flow rate, RO permeate, and MD permeate are shown as a function of MD recovery in Fig. 12. The feed flow rate and RO permeate decrease as the MD recovery increases. To decrease the water cost for the RO-MD hybrid system lower than that of RO system, the permeate flux and recovery of MD should exceed 10 LMH and 35%, respectively, under these simulation conditions. The calculation results show that the RO-MD hybrid system allows lower water cost than MD plant in the similar operating conditions. This is attributed to the fact that the recovery of hybrid plant is higher than that of MD plant. In fact, the recovery of hybrid plant increases from 52 to 76% with increasing MD recovery ranging from 20 to 60%.

Figs. 13 and 14 show contours of annual capital cost and annual operating cost of RO-MD hybrid system at different permeate flux and recovery of MD system, respectively. The trend of the annual capital cost and operating cost contours is same as the contours for water cost. The recovery is inversely proportional to capital and operating cost of intake and pre-treatment process, vacuum pressure pump, and feed pump. Moreover, the permeate flux is inversely proportional to membrane cost and membrane replacement cost. At the same recovery (40%) and permeate flux (12 LMH) conditions to RO and MD, the capital cost of RO-MD hybrid plant decreased by approximately 15% because the recovery of the hybrid

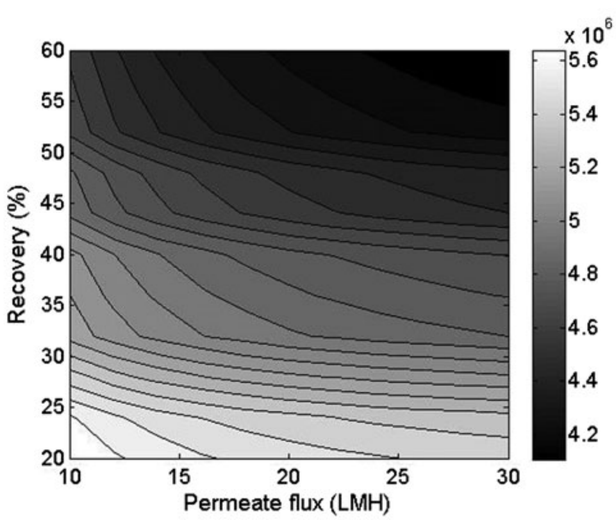

Fig. 13. Contours of annual capital cost (\$/year) of RO-MD hybrid system at different permeate flux and recovery.

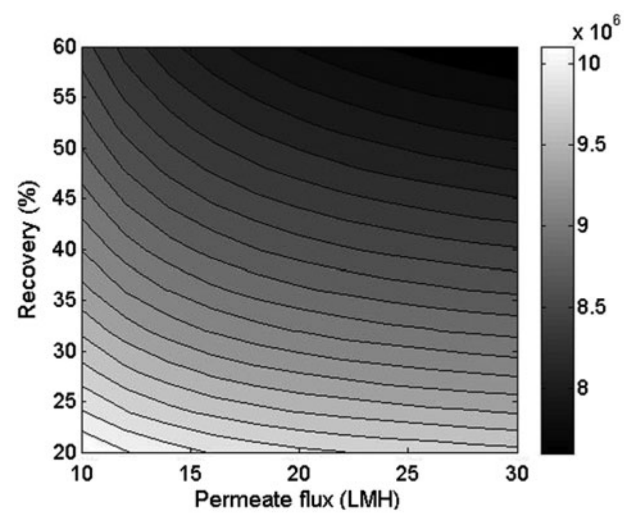

Fig. 14. Contours of annual operating cost (\$/year) of RO-MD hybrid system at different permeate flux and recovery.

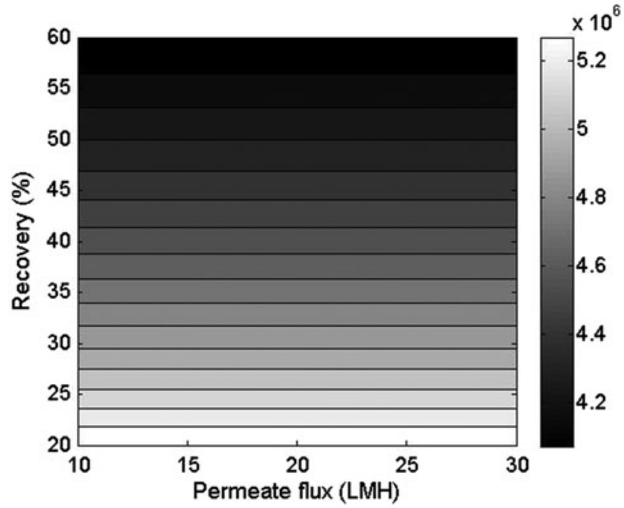

Fig. 15. Contours of annual power cost ($/year) of RO-MD hybrid system at different permeate flux and recovery.

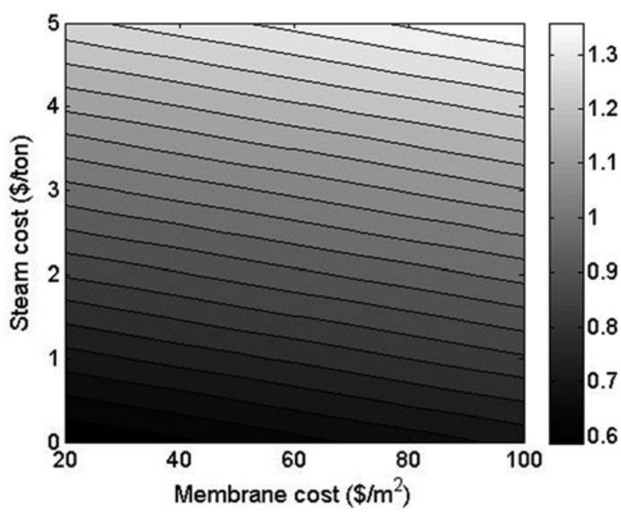

Fig. 16. Contours of water cost ($/m^3) of RO-MD hybrid system at different membrane cost and steam cost.

plant was 64% under the given conditions. But the operating cost increased to approximately 20% since the power cost of MD system is higher than that of RO system.

In Fig. 15, the power cost of RO-MD hybrid system is shown at different permeate flux and recovery. Compared with the results in Fig. 6, which is for the stand-alone MD system, the dependence of power cost on recovery is less significant in RO-MD hybrid system. According to the simulation results, the power

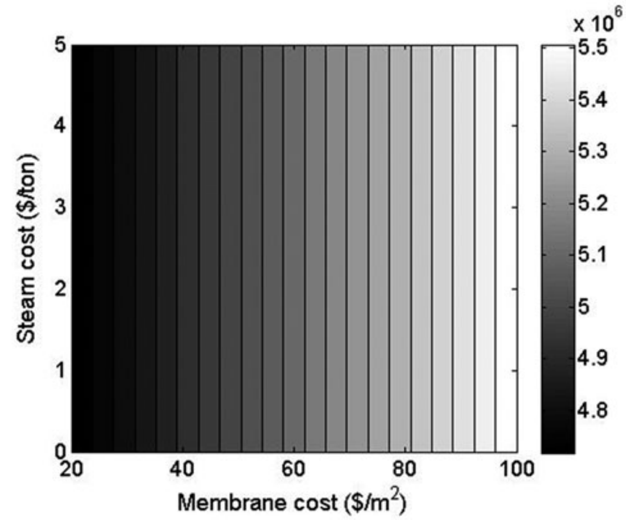

Fig. 17. Contours of annual capital cost ($/year) of RO-MD hybrid system at different membrane cost and steam cost.

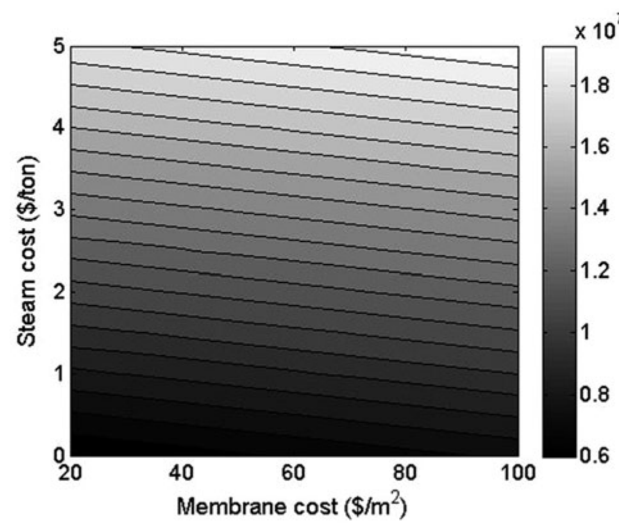

Fig. 18. Contours of annual operating cost ($/year) of RO-MD hybrid system at different membrane cost and steam cost.

cost in RO-MD hybrid system is reduced by 30% with an increase in recovery from 20 to 60%.

Fig. 16 shows contours of water cost of RO-MD hybrid system at different membrane and steam cost. As expected, the water cost of RO-MD hybrid system was sensitive to the membrane and steam cost. The water cost of RO-MD hybrid system was estimated between 0.6 and 1.4 $/m^3, with the changes in the steam cost and membrane cost. In addition, the water cost of RO-MD hybrid system is lower than that

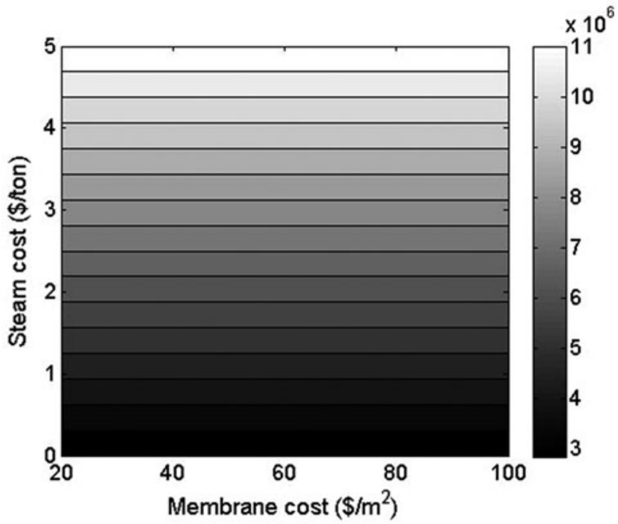

Fig. 19. Contours of annual power cost ($/year) of RO-MD hybrid system at different membrane cost and steam cost.

of stand-alone MD system in the same simulation conditions. Moreover, the steam cost had less significant effect on the hybrid system than stand-alone MD system. To make the water cost of RO-MD hybrid system cheaper than that of the RO system, the membrane and steam cost should not exceed 1 $/ton 90 $/m^2, respectively.

The annual capital cost and annual operating cost of RO-MD hybrid system are analyzed at different membrane and steam cost. It is evident from Fig. 17 that the annual capital cost only depends on the membrane cost. On the other hand, the annual operating cost affected both membrane cost and steam cost as shown in Fig. 18. Under the same conditions, the operating cost is much higher than capital cost due to the consumption of large amount of thermal energy.

Fig. 19 shows the power cost of SWRO-MD hybrid system at different membrane and steam cost. The power cost of hybrid system increases significantly with increasing steam cost. The power cost of hybrid system increases from 3,000,000 to 11,000,000 $/year with an increase in steam cost from 0 to 5 $/ton. However, the power cost of RO-MD hybrid system is still much lower than that of stand-alone MD system.

Table 3 shows the summary of the economic evaluation of 50,000 m^3/d RO, MD, and RO-MD hybrid plant under similar conditions. The flux and recovery of RO and MD were set to 12 LMH and 40% equally. The membrane cost of RO and MD was set to 50 $/m^2 and the steam cost was set to 1 $/ton. In this case, the water cost of RO-MD hybrid system is equal to the water cost of RO system. On the other hand, the water cost of MD is higher than that of RO system. Nevertheless, both MD and RO-MD hybrid system may have cost-competiveness when the steam cost is reduced. If renewable energy or waste heat from industrial process can be used, the steam cost may be lowered. This will lead to reduction in water cost for MD and RO-MD hybrid system.

Table 3
Comparison of calculation results of RO, MD, and RO-MD hybrid plant

	RO	MD	RO-MD
Energy consumption			
Electric energy	5,714 kW	3,248 kW	3,572 kW
Thermal energy	–	331,600 kW	113,000 kW
Capital cost			
Direct capital cost	44,580,800 $	33,603,000 $	38,000,000 $
Indirect capital cost	13,374,200 $	10,082,000 $	11,300,000 $
Capital cost	57,955,000 $	43,685,000 $	49,300,000 $
Annual capital cost	5,903,000 $/y	4,450,000 $/y	5,020,000 $/y
Operating cost			
Power cost	3,604,557 $/y	6,825,000 $/y	4,430,000 $/y
Annual operating cot	7,633,400 $/y	12,230,000 $/y	8,600,000 $/y
Water cost			
Water cost	0.75 $/m^3	0.91 $/m^3	0.75 $/m^3

4. Conclusions

This study undertook evaluations of the economic feasibility of RO, MD, and RO-MD hybrid systems to propose guidelines for the MD having price competitiveness. The following conclusions can be drawn from this work:

(1) The theoretical model was successfully applied to estimate the economics of three different desalination systems including RO system, MD system, and RO-MD system. The model allowed the calculation of water cost as a function of permeate flux, recovery, membrane cost, and steam cost for these systems.
(2) MD and RO-MD hybrid systems can have economic feasibility compared with RO system when the recovery and flux of MD is higher than RO and the steam cost is inexpensive.
(3) Under similar conditions, the water costs of RO-MD hybrid and RO system were calculated to be almost same. However, the water cost of MD system is higher than that of RO-MD hybrid and RO systems. If the steam cost becomes cheaper, the MD system can also have economic feasibility. In other word, the most important fact affecting the economics of MD system and RO-MD hybrid system is the cost of thermal energy source.

Acknowledgment

This research was supported by a grant(code 15IFIP-B065893-03) from Industrial Facilities & Infrastructure Research Program funded by Ministry of Land, Infrastructure and Transport of Korean government.

References

[1] M. Parfit, Water, The Power, Promise, and Turmoil of North America's Fresh Water, National Geographic Special Edition, Washington, DC, 1993.
[2] R.F. Service, Desalination freshens up, Science 313 (2006) 1088–1090.
[3] A. Cipollina, G. Micale, L. Rizzuti, Seawater Desalination: Conventional and Renewable Energy Processes (Green Energy and Technology), Springer Science & Business Media, New York, NY, 2009.
[4] C. Fritzmann, J. Löwenberg, T. Wintgens, T. Melin, State-of-the-art of reverse osmosis, Desalination 216 (2007) 1–76.
[5] R. Rautenbach, T. Linn, D.M.K. Al-Gobaisi, Present and future pretreatment concepts—Strategies for reliable and low-maintenance reverse osmosis seawater desalination, Desalination 110 (1997) 97–106.
[6] P. Sukitpaneenit, T.-S. Chung, High performance thin-film composite forward osmosis hollow fiber membranes with macrovoid-free and highly porous structure for sustainable water production, Environ. Sci. Technol. 46 (2012) 7358–7365.
[7] L.F. Greenlee, D.F. Lawler, B.D. Freeman, B. Marrot, P. Moulin, Reverse osmosis desalination: Water sources, technology, and today's challenges, Water Res. 43 (2009) 2317–2348.
[8] J.R. McCutcheon, R.L. McGinnis, M. Elimelech, Desalination by ammonia–carbon dioxide forward osmosis: Influence of draw and feed solution concentrations on process performance, J. Membr. Sci. 278 (2006) 114–123.
[9] K.W. Lawson, D.R. Lloyd, Membrane distillation, J. Membr. Sci. 124(1) (1997) 1–25.
[10] A.M. Alklaibi, N. Lior, Membrane-distillation desalination: Status and potential, Desalination 171(2) (2005) 111–131.
[11] L. Mariah, C.A. Buckley, C.J. Brouckaert, E. Curcio, E. Drioli, D. Jaganyi, D. Ramjugernath, Membrane distillation of concentrated brines—Role of water activities in the evaluation of driving force, J. Membr. Sci. 280 (1–2) (2006) 937–947.
[12] C.R. Martinetti, A.E. Childress, T.Y. Cath, High recovery of concentrated RO brines using forward osmosis and membrane distillation, J. Membr. Sci. 331(1–2) (2009) 31–39.
[13] P. Wang, T.-S. Chung, Recent advances in membrane distillation processes: Membrane development, configuration design and application exploring, J. Membr. Sci. 474(0) (2015) 39–56.
[14] M.A. Darwish, M. Abdel-Jawad, G.S. Aly, Technical and economical comparison between large capacity multi stage flash and reverse osmosis desalting plants, Desalination 72 (1989) 367–379.
[15] Y. Dreyzin, Ashkelon seawater desalination project—Off-taker's self costs, supplied water costs and benefits, Desalination 190 (2006) 104–116.
[16] C. Sommariva, Desalination Management and Economics, Mott Mac Donald, Faversham House Group, Surrey, 2004.
[17] A. Malek, M.N.A. Hawlader, J.C. Ho, Design and economics of RO seawater desalination, Desalination 105 (1996) 245–261.
[18] M.G. Marcovecchio, P.A. Aguirre, N.J. Scenna, Global optimal design of reverse osmosis networks for seawater desalination: Modeling and algorithm, Desalination 184 (1–3) (2005) 259–271.
[19] F. Marechal, E. Aoustin, P. Bréant, Multi-objective optimization of RO desalination plants, Desalination 222 (2008) 96–118.
[20] F. Maskan, D.E. Wiley, L.P.M. Johnston, D.J. Clements, Optimal design of reverse osmosis module networks, AIChE J. 46(6) (2004) 946–954.
[21] L. Yan-Yue, H. Yang-Dong, X.-L. Zhang, W. Lian-Ying, Q.-Z. Liu, Optimum design of reverse osmosis system under different feed concentration and product specification, J. Membr. Sci. 287 (2007) 219–229.
[22] S. Al-Obaidania, E. Curcio, F. Macedonio, G. Di Profio, H. Al-Hinai, E. Drioli, Potential of membrane distillation in seawater desalination: Thermal efficiency, sensitivity study and cost estimation, J. Membr. Sci. 323 (2008) 85–98.