Economic analysis of desalination technologies in the context of carbon pricing, and opportunities for membrane distillation

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Abstract

The economics membrane distillation (MD) and common seawater desalination methods including multi effect distillation (MED), multistage flash (MSF) and reverse osmosis (RO) are compared. MD also has the opportunity to enhance RO recovery, demonstrated experimentally on RO concentrate from groundwater. MD concentrated RO brine to 361,000 mg/L total dissolved solids, an order of magnitude more saline than typical seawater, validating this potential. On a reference 30,000 m^3/day plant, MD has similar economics with other thermal desalination techniques, but RO is more cost effective. With the inclusion of a carbon tax of $23 per tonne carbon in Australia, RO remained the economically favourable process. However, when heat comes at a cost equivalent of 10% of the value of the steam needed for MD and MED, under a carbon tax regime, the cost of MD reduces to $0.66/$m^3$ which is cheaper than RO and MED. The favour to MD was due to lower material cost. On low thermally, high electrically efficient installations MD can desalinate water from low temperature (<50°C) heat sources at a cost of $0.57/$m^3$. Our assessment has found that generally, MD opportunities occur when heat is available at low cost, while extended recovery of RO brine is also viable.

Key words: membrane distillation, desalination, carbon tax, price, economics and waste heat.

1. Introduction

Desalination is a means of producing fresh water from saline or brackish water by removing dissolved salts to make it suitable for human use, agricultural and industrial or manufacturing purposes [1]. With water shortages emerging across the world, communities are turning to desalination as a solution to reliable water supply. Cost and energy reductions for desalination are therefore considered an important factor to minimise the environmental impact of desalinated fresh water supply especially in arid and semi arid regions where there is little alternative. Commercial technologies for desalination include membrane separation processes such as reverse osmosis (RO) and electrodialysis (ED), as well as thermal processes, specifically multi effect distillation (MED), multi stage flash (MSF) and vapour compression distillation (VCD). These technologies are the most widely used desalination processes with MSF and RO dominating the market for both brackish and sea water with a total share of about 78% [2].

The techno-economic performance of these processes favours RO due to the continual advances made to reduce energy consumption and lower cost of water produced [3, 4]. While most authors report RO as the less expensive process to recover fresh water these studies do not take into account imminent rises in energy prices. RO uniquely relies on electricity to operate, while the thermal processes can utilize a waste heat or solar thermal energy more conveniently [3-6].

The US Bureau of Reclamation Desalination Roadmap 2003 [7] indicated that in RO, energy consumption accounts for 44% of the produced water cost, and fixed charges account for 37%. Together, these account for over 81% of the total desalination cost [7, 8]. Similarly, it is known that for thermally driven desalination processes MSF and MED, the capital cost of the large metallic evaporators is very high, in the range of 40% to 50% of the total cost of water produced [1-9]. These systems thus conform to very different economics, and it is of interest to know where they fit economically under rising energy prices and the recent emergence of carbon pricing. Furthermore, alternative desalination processes that are not commercialized (or widely used) may be more economical from the perspective of capital and energy costs. They may also be easier to use and can potentially utilize a low grade heat source making them of considerable interest. One commercially emerging desalination technology that has different cost metrics and can harness waste heat sources is membrane distillation.

Membrane distillation (MD) is a thermally driven membrane process and may find an economically feasible niche amongst the commercialized desalination processes (MSF, MED and RO) which are considered to be technologically mature and therefore have very little space for major performance improvements [2]. The advantages of MD over commercialised desalination technologies are as follows [6]: (i) lower operating temperatures and vapour space required than MSF and MED (ii) lower operating pressure than RO (iii) more than 99.9% theoretical salt rejection (iv) the performance is not limited by high osmotic pressure or concentration polarization. Four MD configurations have been identified: Direct Contact Membrane Distillation (DCMD), Air Gap Membrane Distillation (AGMD), Vacuum Membrane Distillation (VMD) and Sweep Gas Membrane Distillation (SGMD) [5]. The DCMD configuration was selected in our experimental work because of its simplicity and high water flux.

To date the commercial uptake of MD has not been significant and further work is needed to uncover real opportunities. Furthermore, other emerging technologies which are still in the research and development phase such as forward osmosis and freeze/thawing [1] indicate that despite the variety of commercial desalination systems, there is still a driver for more diversity in desalination options. In order to foresee the ‘economic niche’ of these emerging technologies, a cost analysis is needed to understand how they will fit within the desalination industry. The emerging technology that is the focus in this economic assessment is MD.

A desalinated water cost model for MD, like the benchmark RO and MED systems, is sensitive to several economic and technical factors such as energy source, plant capacity, salinity, and design features [5]. Among those factors, energy source and plant capacity have a dominating influence in addition to feed seawater salinity for the RO process [5-11]. The energy requirement of desalination has an important effect on the overall process economics that is more prone to suffer from variation in the cost of fossil fuels [12].

An extensive study of membrane distillation by Obaidani et al [3], reports exergy analysis, sensitivity study and economical evaluation carried out to assess the feasibility of the direct contact membrane distillation (DCMD) process with heat recovery. They estimated a water cost \$1.17/m^3, which is comparable to the cost of water produced by conventional thermal processes, i.e. \$1.00/m^3 for MED and \$1.40/m^3 for MSF [10]. The study also reveals that there is a high possibility of significant savings when a low-grade thermal energy source is used. The study claims that the cost is competitive to the cost of water produced by RO, which is about \$0.5/m^3 [11].

The Memstill project presented in 2006 by Hanemaijer et al [13] claims to have the potential to reduce the cost of existing desalination technologies for seawater and brackish water, by replacing MSF and MED modules by an air gap MD module. The process proposes to reduce the desalination costs to \$0.26 /m^3 using low grade thermal steam or heat as the driving force. Similarly, in a recent (2012) study by M.R. Qtaishat and F. Banat [14], the costs in sourcing the heat from solar energy was explored.The economics were found to be dependent on the cost and efficiency of the solar panels indicating that waste heat for MD is currently a more economically viable concept.

Despite these costing reports in literature, it is uncertain what desalinated water by any technology will cost in a carbon constrained society. In 2012, Australia implemented a \$23 per tonne carbon cost. A variable price with commence in 2015 when the Australian Government converts this to an Emission Trading Scheme [15]. With policies coming into practice to tax carbon emissions, the economics of each desalination process are therefore undergoing change particularly with the concept of MD using waste heat. Therefore one of the purposes of this work is to explore how carbon taxing will influence the cost of desalination and how the waste heat concept can give opportunities for MD.

1.1 Membrane distillation progress and technological challenges

MD is a hybrid of membrane and thermal desalination. The MD process classically uses membranes that are hydrophobic and microporous. The driving force is a vapour pressure difference across the membrane. The vapour evolved from the feed solution passes through the pores of the membrane and is collected as the condensate. Liquid water is prevented from passing the membrane thus creating a desalination effect over a very small space. Due to the convenient containment of the liquid surface using the membrane, higher packing densities bring it in line with state of the art RO compactness. This is typically achieved via different MD configurations, which are DCMD, VMD, AGMD and SGMD, which have been well described in the literature [16-19].

The standard thermal energy required to operate an MD system is 628 kWh/m^3 [20]. This value equates to a performance ratio (PR), or gain output ratio (GOR) of 1, being the mass ratio of water produced to the amount of steam energy (i.e., latent heat) fed to the process. This can be compared to state-of-the-art MED requiring about 2 kWh/m^3 of electric energy and 60 kWh/m^3 thermal energy as shown in Table 1. In the last few years, MD has emerged with numerous commercially oriented devices and novel process integrations to try to match MED thermal efficiencies. The most notable organisations specialising in MD modules or high efficiency systems are: Fraunhofer ISE (AGMD), Memstill and Aquastill (AGMD), Scarab (AGMD), Memsys (vacuum enhanced multi effect AGMD) [20]. The thermal energy required through Memstill's trials, is as low as 56 to 100 kWh/m^3 of water produced (GOR up to 11.2). This is the lowest value reported from real testing (or highest GOR), but to achieve this, the water must be heated to 80–90 °C. [51].

In addition to high energy requirements, the other technological challenges of MD include module design, membrane fouling and scaling. These are well described in the literature [16, 17, 21-23]. Attractive advantages of MD are related to the possibility of overcoming the RO limit of around 70,000 mg/L (due to trans-membrane flux independent from feed concentration), process intensification and also its ability to operate at relatively low temperatures [7, 17]. This enables MD to be a compact operation for further recovery of RO brines at low pressure, and reduce discharge volumes in areas where this is a significant cost (e.g. inland groundwater desalination).

Despite the potential of MD, it has not been significantly implemented since it was patented in the late 1960's. Research intensity picked up in the 1980's [5] due to rising water, energy and environmental issues. We have previously explored polymer and ceramic membranes for desalination, and explored MD in dairy processing and industrial process integration [24-28]. While MD researchers have already focussed on relative technology costs, process optimisation, module design and fouling, this paper presents results on a niche operation of extended RO recovery, as well as the relative price of MD under a carbon tax and in a modified operational mode.

1.2 Objectives of the study

The objectives of this study are as follows:

1. To carry out testing on RO brines to explore potential to recover beyond the limit of RO systems (70,000 mg/L). This is done in the context of inland groundwater desalination to reduce discharge volumes;

2. To assess the potential impact of a carbon price on the cost of desalination technologies as well as the impact of utilising waste heat to drive MED and MD; and

3. To assess opportunities for low thermally/high electrically efficient MD setups that can effectively harness abundant lower grade heat sources.

2. Methodology

2.1 Desalination testing on groundwater RO concentrate

Experiments were conducted in DCMD mode to confirm the viability of MD to further concentrate beyond the limit of RO at 70,000 mg/L. The flow diagram of the experimental rig is shown in Figure 1. In this test, 20 L of feed solution was initially added to the feed reservoir. Following batch concentration, another 20 L of raw feed water was added after each run for five different experimental runs; making a total of 100 L raw feed solution processed using DCMD. The membranes used were flat sheet PTFE supported on polypropylene scrim backing as optimised in previous work [16]. The membranes had an active area of 0.0169 m^2, pore size of 0.45 μm and were supplied by Ningbo Chanqi, China. A cartridge filter with filtration size of 0.5 μm was used on the hot loop to collect precipitated matter prior to entering the MD module. The flow rate into the module hot and cold sides was 900 mL/min, the feed temperature was 60 ^oC and the cold temperature was maintained at 20 ^oC. Permeate build-up was measured by the accumulated mass of water in the permate tank. Electrical conductivity was measured using a conductivity meter in the permeate tank to ensure membrane intactness (conductivity < 100 μS/cm). The concentrated brine for the MD experiments was obtained from a RO plant operating in Edenhope, Victoria, Australia. Groundwater is fed to the RO plant with a total dissolved solids (TDS) of around 1,400 mg/L, and the brine TDS concentrations was approximately 3,300 mg/L. The RO concentrate was further concentrated by an in-house RO rig using a 2.5" DOW FILMTEC BW30 membrane operating at pressures between 1.5 and 1.8 MPa to achieve a TDS of approximately 11,000 mg/L. This became the feed water to the MD experiment. TDS was determined gravimetrically on 5 mL aliquots of sample dried at 105 ºC overnight. Cations in samples were determined by ICP–OES analysis, which was performed on a range of serial dilutions to reduce the concentration of each species to a suitable level. Cs matrix buffering @ ~ 5 g/L and a Cs internal standard was employed to accommodate the wide range of matrix variation in the dilutions. Chloride concentration was determined via titration with $AgNO_3$ after acidification with 1:1 nitric acid solution. The end point was detected electrochemically.

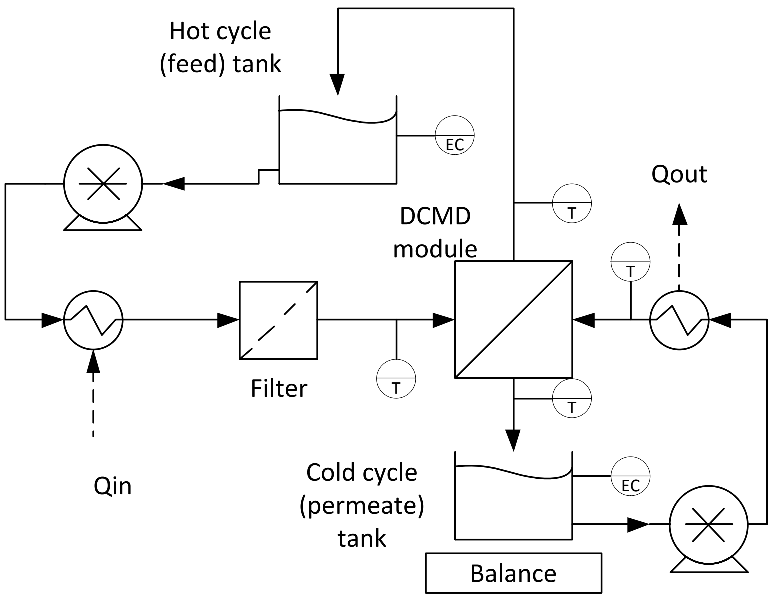

Figure 1: DCMD Experimental flow diagram

2.2 Cost modelling

2.2.1 Thermal and electrical energy usage and emissions of established desalination

Energy consumption in MD systems includes both the thermal energy necessary to heat the feed solution and the electrical energy required to run the circulation pumps, vacuum pumps or compressors. The thermal energy requirement is around 90% of the total energy but can come at a relatively low cost. Meanwhile electrical energy is more expensive than low grade heat. Desalination requires about 0.8 kWh/m^3 energy for seawater desalination based on a thermodynamic minimum [12, 14, 29, 30].

The energy values and carbon intensity used in our economic study are shown in Table 1. Here the total energy requirement for seawater desalination is at least an order of magnitude higher than the thermodynamic minimum, and mostly sourced from fossil fuel (electric power generation and by-product steam). The pollution associated with energy production from fossil fuel is as follows: NO_x, SO_2, volatile compounds, particulates, and CO_2. Other environmental impacts include cost from effluents disposal, including chemicals, brine and possible sludge. The CO_2 emission from fossil fuel is the pollution of greatest interest in this work and also considered to be the highest contributor to greenhouse gas emissions (largest volume emitted). The carbon footprint of desalination systems is a combination of emissions associated with power used in the desalination process and the embodied associated chemicals used in production, treatment and disposal of solid waste and manufacture and replacement of membrane components. In our cost model, a carbon tax of $23 per tonne carbon was used reflecting the initial fixed price of carbon introduced in Australia on 1 July 2012. This will increase gradually, then transition to a cap-and-trade emission trading system by 2015 [15].

Table 1: Data and assumptions used in the economic study

Input Parameter	MD	RO	MED	Reference
Plant availability (%)	90	90	90	[3, 31, 32]
Interest rate (%)	5	5	5	[3, 5, 33]
Amortization	0.08	0.08	0.08	[3, 5, 33]
Electrical cost ($/kWh)	0.09	0.09	0.09	[30, 33]
Steam cost ($/kg)	0.007	0.007	0.007	[3]
Labour cost ($/m^3)	0.03	0.02	0.03	[3, 5, 34, 35]
Brine disposal ($/m^3)	0.0015	0.04	0.0015	[3, 36, 37]
Maintenance cost (%)	2	2	2	[32, 34, 38]
Pre-treatment cost ($/m^3)	0.019	0.05	0.03	[3, 32, 39, 40]
Thermal energy requirement (kWh/m^3)	100	0	60	[2, 41-43]
Emission factor for natural gas (kg CO_2-e/kWh)	0.184	0.184	0.184	[44, 45]
Emission factor for Electricity (kg CO_2-e/kWh)	1.22	1.22	1.22	[44, 45]
Electrical energy requirement (kWh/m^3)	2	3.5	2	[2, 33, 46, 47]
Carbon Tax ($/tonne carbon)	23	23	23	[15]
Exponent ‘m’ (Scale index)	0.8	0.81	0.83	[30]

2.2.2 Economic model setup

The major cost elements for desalination plants are capital cost and annual operating costs.

Capital cost covers purchasing cost of equipment, auxiliary equipment, land and installation charges [5]. Annual operating cost represents the total yearly costs of owning and operating a desalination plant. These include amortization or fixed charges, operating and maintenance costs, energy costs and membrane replacement costs. This cost study is only for isolated plant cases and does not include distribution. The combined environmental impact of desalination includes on and off site pumping. However, this report only focuses on the impact of CO_2 emission from the desalination plant itself to allow weighing up of the truly different aspects. Other environmental costs would include effluent disposal, chemicals, brine and sludge.

The calculation setup is presented in the following sections. A sample calculation of water production cost using the MD system based on this setup, when a low grade heat source is available and carbon tax applied is presented in the Appendix.

2.2.3 Capital cost (CAPEX) of a plant

The CAPEX is estimated using a Capacity Factored Estimate. The cost of new plant is derived from the cost of a similar plant of known capacity, with similar production route, but not necessarily the same end product (the product should be relatively similar, however). It relies on the nonlinear relationship between capacity and cost as shown in Equation 1 [34, 39]:

$$\left[\frac{\text{capital cost plant 1}}{\text{capital cost plant 2}}\right] = \left[\frac{\text{plant capacity 1}}{\text{plant capacity 2}}\right]^{m} \qquad (1)$$

m = the scale index [exponent]. The *m* used in the capacity factor equation is the slope of the log curve that has been drawn to reflect the change in the cost of a plant as it is made larger or smaller [34]. The value varies depending on the type of plant as shown in Table 1. The methodology of using capacity factor is sometimes referred to as the "six tenth factor" method because of the reliance on an exponent of 0.6 if no other information is available [39]. However, for desalination plants the exponent *m* is usually closer to 0.8 [48]. The capital cost of various desalination plants as reported in the literature is as shown in Table 2.

Table 2: Capital cost of various desalination processes

Process	Plant Capacity m^3/day	Unit-Capital Cost $/($m^3$/day)	Reference
MD	24,000	1,131	[3]
MED	37,850	1,860	[5]
MSF	37,850	1,598	[5]
RO	37,850	1,313	[5]

2.2.4 The cost of capital

The annual capital cost reflects the cost associated with servicing the capital cost used to build the new desalination plant. This is estimated by multiplying the total capital cost of treatment and conveyance by appropriate capital recovery factor [30, 36, 45].

The capital recovery factor (CRF) is calculated using the Net Present Value (NPV) Method. The net present value (NPV) of the asset is defined for a given discount rate (r), and (n), a series of future payments over a defined period of time:

$$\mathrm{CRF} = \left[\frac{\mathrm{r}(1+\mathrm{r})^{\mathrm{n}}}{(1+\mathrm{r})^{\mathrm{n}} - 1}\right] \qquad (2)$$

In terms of cost per amount of water produced, the capital cost is then determined by:

$$\text{Normalised capital cost}\left[\frac{\$}{\text{m}^3}\right] = \frac{\text{CRF} \times \text{Capital cost}\left[\frac{\$}{\text{y}}\right]}{\text{Plant capacity}\left[\frac{\text{m}^3}{y}\right]} \quad (3)$$

2.2.5 Operating cost for the desalination process

Operating costs are those expenditures incurred after plant commissioning and during the actual operation. These include energy, brine disposal, membrane replacement, pre-treatment, labour, and maintenance cost determined as follows:

2.2.5.1 Electrical and thermal energy

Determining the electrical energy requirement requires the current industrial cost of electricity as shown in Table 1. However thermal energy requirement is less obvious since this energy is typically taken from low pressure steam lines in a thermal process as opposed to specifically burning fuel for desalination [46]. A value of \$0.007 per kg steam wash used in 2008 [3], which indexed to 2012 is \$0.0078 per kg. Using latent heat, this value converts to \$0.0124 per kWh. The energy (electricity or thermal) cost is determined by:

$$\text{Energy cost}\left[\frac{\$}{\text{m}^3}\right] = \text{Cost of energy}\left[\frac{\$}{\text{kWh}}\right] \times \text{Specific energy consumption}\left[\frac{\text{kWh}}{\text{m}^3}\right] \quad (4)$$

2.2.5.2 Emission cost

Carbon costs for both electrical and thermal (via natural gas) is given as follows:

$$\begin{aligned}\text{Carbon cost}\left[\frac{\$}{\text{m}^3}\right] &= \text{Energy requirement}\left[\frac{\text{kWh}}{\text{m}^3}\right] \times \text{ Emission factor}\left[\frac{\text{kg CO}_2 - \text{e}}{\text{kWh}}\right] \\ &\times \text{Carbon tax}\left[\frac{\$}{\text{tonne CO}_2 - \text{e}}\right] \times \frac{1}{1000}\left[\frac{\text{tonne}}{\text{kg}}\right] = \quad (5)\end{aligned}$$

Both electrical and thermal emissions are determined separately and added to produce the total emission cost.

2.2.5.3 Membrane replacement cost

In this analysis, the membrane cost for MD is estimated based on principle flux of 6 kg/m^2/h, a membrane cost of \$9 per m^2, and 20% replacement per year. This leads to the operating cost of \$0.034 per m^3 water treated:

$$\text{Membrane cost}\left[\frac{\$}{m^3}\right] = \text{Membrane price}\left[\frac{\$}{m^2}\right]\times \text{Replacement rate}\left[\frac{1}{y}\right]\times\frac{1000\left[\frac{l}{m^3}\right]}{\text{Membrane flux}\left[\frac{l}{m^2\,h}\right]\times 8760\left[\frac{h}{y}\right]} \qquad (6)$$

2.2.5.4 Brine disposal cost/Pre-treatment cost/ Maintenance cost /Labour cost

The specific cost for brine disposal, pre-treatment costs, maintenance costs and labour costs for various desalination technologies is given in Table 1.

3. Results and Discussion

3.1 Experimental viability of MD for RO brine reduction

Figure 2 shows the flux as a function of time (and increasing concentration) for three batch DCMD processes, while Table 3 shows the concentration of various species at the start and end of each batch concentration. In each run 20 L of feed was concentrated until a volume of < 2 L was achieved. This represents a recovery greater than 90%. Fluxes were in the range of 20 to 37 kg/m^2/h which were largely consistent and quite high for all salinities (up to 361,000 mg/L TDS) in these experiments. Importantly, this data shows that the flux is not significantly dependent on the feed water salinity in this concentration range. While flux declined during the run, since it returned to the original value as the previous run at lower concentration after replacing the membrane, we conclude this is due to changes in the new membrane's performance, and not salinity. This is strong evidence to support the concept that the flux of MD (and hence capital cost) is not as strongly linked to concentration as in the case of RO.

It is also important to note that the fluxes are higher than the value chosen in our economic model (Section 2.2.5.3) since practical MD installations typically have heat recovery systems or operate at lower temperatures, which tend to reduce flux. These experiments were performed, not to determine an appropriate flux but to validate to potential for RO brine concentration using MD.

The results in Table 3 also show that the majority of the TDS was sodium chloride, which accounted for 72% to 77% of the TDS for all samples with the exception of the initial feed of Feed 1. In this case the NaCl represented 99% of the TDS. We believe, based on a mass balance that this higher value is likely due to an underestimate of the TDS. Comparison of the concentration factors of sodium and chloride, measured by different techniques, reveals similar values for all three runs (17, 11 and 11 respectively). This implies that the elemental analysis is more reliable than TDS. We also see that sodium chloride concentration for the final concentrated solution (Feed 3) was 259,000 mg/L which is below the saturation concentration of approximately 373,000 mg/L at 60°C [49]. As stated earlier, the NaCl enrichment was very high. An approximate 17 fold increase was measured in the first batch concentration process, while 11 fold increases were seen in the second and third batch concentration processes. This aligns well with the greater than 90% water recovery estimated. The calcium concentration factor on the other hand was more limited. A 4.5-fold increase was observed in the first experiment reducing to a 1.6-fold increase for the third experiment. This is a strong indication that precipitation was occurring. No scale was observed on the membrane, but was observed in the 0.5 µm filter and was thus efficiently captured at the highest temperature point in the hot cycle to avoid membrane scaling. MD can therefore operate at salinities well beyond where RO fails, showing increasing potentials to enhance water recovery. Similar results have been demonstrated experimentally, using MD process to concentrate feed water salinity up to 76,000 mg/L TDS, which is twice the salinity of seawater [6, 21]. Our costing in this work is benchmarked using seawater desalination to allow comparison to RO, but clearly MD can exceed the limitations of RO due to its non-reliance on overcoming osmotic pressure. MED could also achieve such high salinities, but is limited due to scaling issues. An interesting feature of MD as a thermal process is the separation of the saline water heating zone (heat exchanger) from the evaporation zone (membrane). The separation has allowed the convenient placement of the filter between the zones to capture precipitating salts immediately after heating, but prior to evaporation. As the water enters the membrane module, it begins to cool due to evaporation effectively enhancing the solubility of common scale species as they concentrate. Also, the membrane surface itself is cooled by the cold permeate side leading to the temperature polarisation effect. Both effects assist in the avoidance of calcium scaling of the membrane.

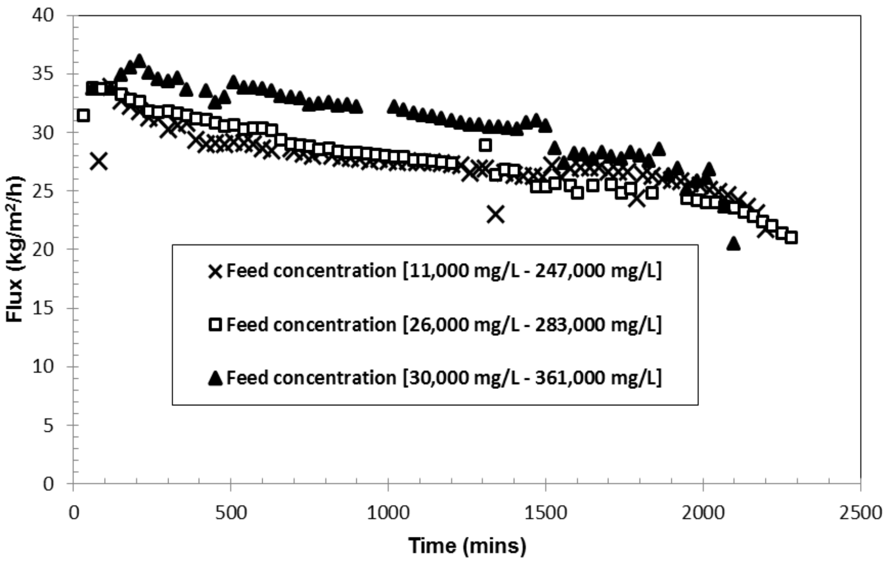

Figure 2: Flux over time during DCMD experiment fed with RO groundwater concentrate

Table 3: Concentration of TDS and major species in the concentrated groundwater samples

Sample solution	Feed solution (mg/L)	TDS (mg/L)	Sodium (mg/L)	Chloride (mg/L)	Sulphur (mg/L)	Calcium (mg/L)	Magnesium (mg/L)
Feed 1	Initial Feed concentration (mg/L)	11,000	3,820	7,080	260	250	520
	Final Feed concentration (mg/L)	247,000	63,700	120,000	4,340	1,130	8,110
Feed 2	Initial feed concentration (mg/L)	26,000	6,880	12,700	440	330	910
	Final feed concentration (mg/L)	283,000	76,000	143,000	5,040	750	10,300
Feed 3	Initial feed concentration (mg/L)	30,000	8,080	14,800	540	320	-
	Final feed concentration (mg/L)	361,000	91,200	168,000	5,410	500	13,100

3.2 Desalination technology costing

Figure 3 shows the treated water cost as a function of plant size. At a production capacity of 30,000 m^3/day, the cost of MD is \$1.72/m^3 while the MED cost is \$1.48/m^3 and RO cost is \$0.69/m^3, indicating that as a plant supplied with steam, MD is not economically favourable against RO or MED for seawater desalination. In this comparison, the higher cost of MD is due to the assumption that waste heat is not available on site and water is heated to high temperatures (>60 °C) by steam. Figure 3 also shows the economy of scale for all MD, RO, MED and MSF. MSF has a strong economy of scale compared to others, but at any scale was determined the most expensive method to desalinate seawater and the most energy intensive. MD costs are slightly higher than the cost of MED. Similar results were reported by Obaidani et al [3] in 2008, where the cost of MD was estimated as \$1.17/m^3 and MED is \$1.00/m^3. MED and MD consume greater amounts of energy (thermal and electrical combined) than RO. However, both MED and MD can utilize waste heat or solar thermal energy.

The cost comparison of both technologies presented in Figure 4 shows that a large cost saving is expected when a high temperature (~60 ^{o}C) waste heat source is available for MD desalination. In such a case, the cost of water produced by MD decreases to \$0.61/m^3 and MED \$0.81/m^3. MD is now cheaper and more competitive than the cost of water produced by

RO. We assumed that waste heat costs 10% of the price of steam used in this analysis. Therefore, MD appears cheaper for desalination if a low cost (i.e. waste) heat source is available.

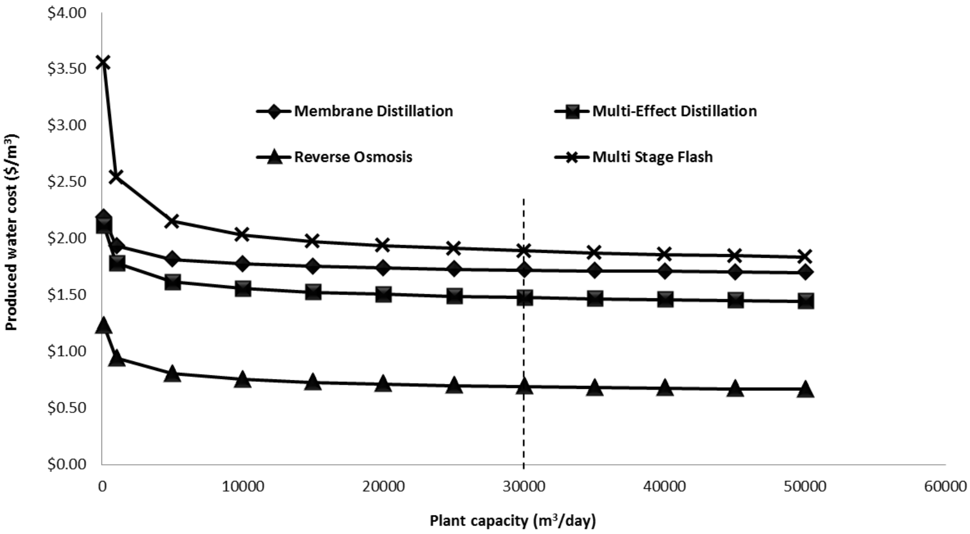

Figure 3: Water production cost of MD, RO, MED and MSF driven by steam and/ or electricity. Dotted line indicates reference plant capacity of 30,000 m^3/day used in this study.

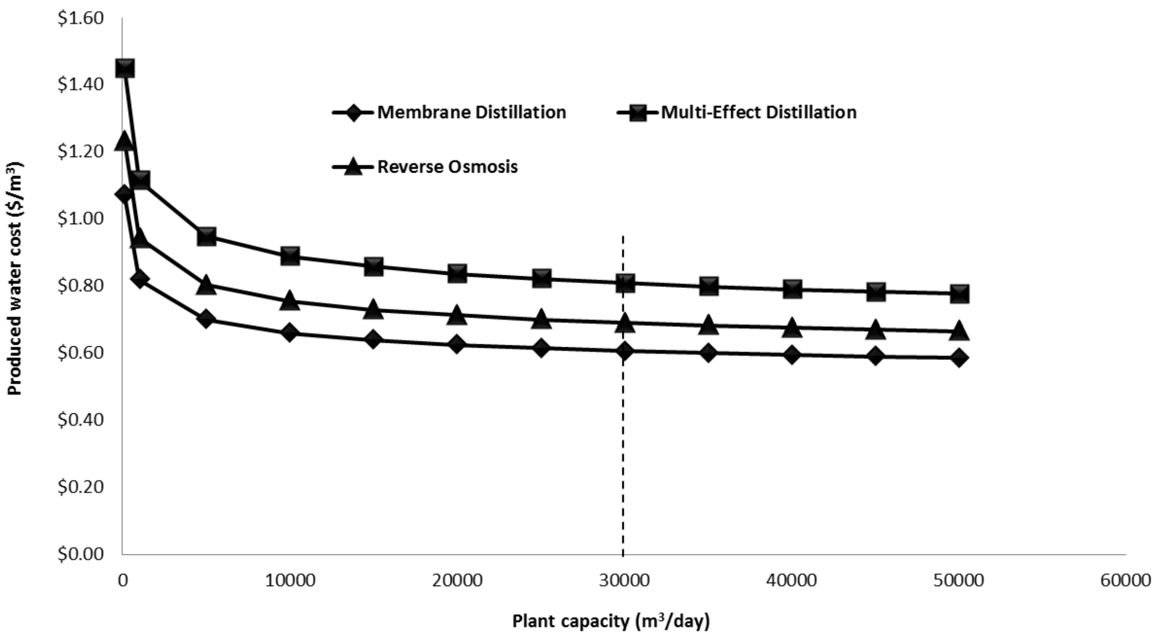

Figure 4: Cost of MD, RO and MED when high temperature waste heat (at 10% steam cost) is available for both MD and MED. Dotted line indicates reference plant capacity of 30,000 m^3/day used in this study.

3.3 Cost composition of MD and MED

Figures 5 and 6 show the cost break down for each contributor to the water price for MD and MED respectively. At any capacity, the cost is highly sensitive to thermal energy, but is more significant for MD and hence offers a better opportunity for the use of waste thermal heat as compared to MED (Figure 7). Although our choice of the nominal 100 kWh/m^3 thermal energy requirement is higher than the 60 kWh/m^3 value used for MED (Table 1), the advantage of MD over MED is its ability to be built from cheaper materials than the MED.

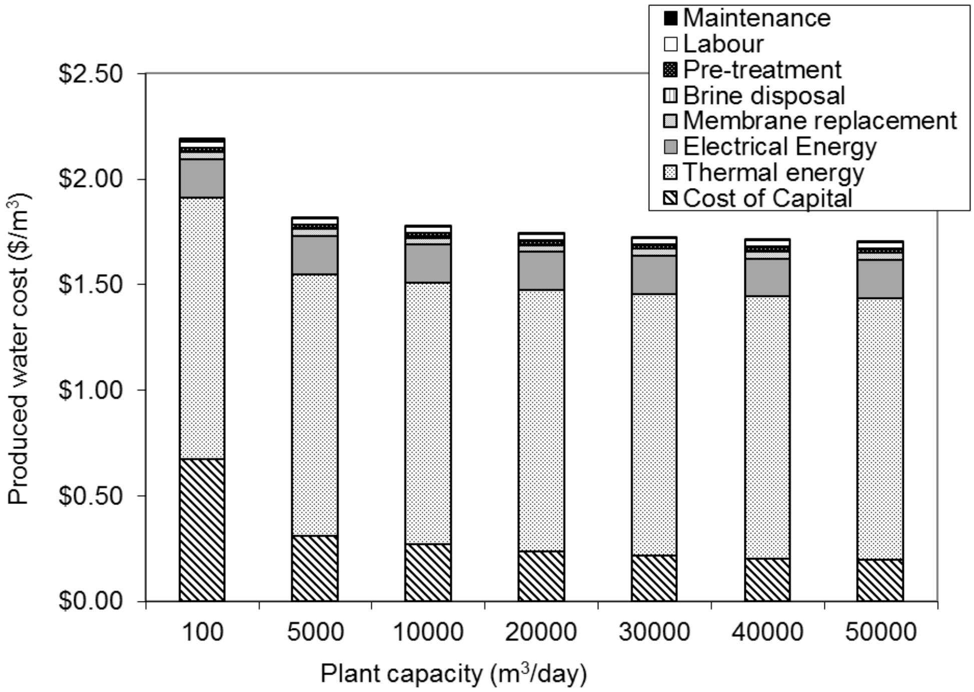

Figure 5: Breakdown of water production cost treated by MD

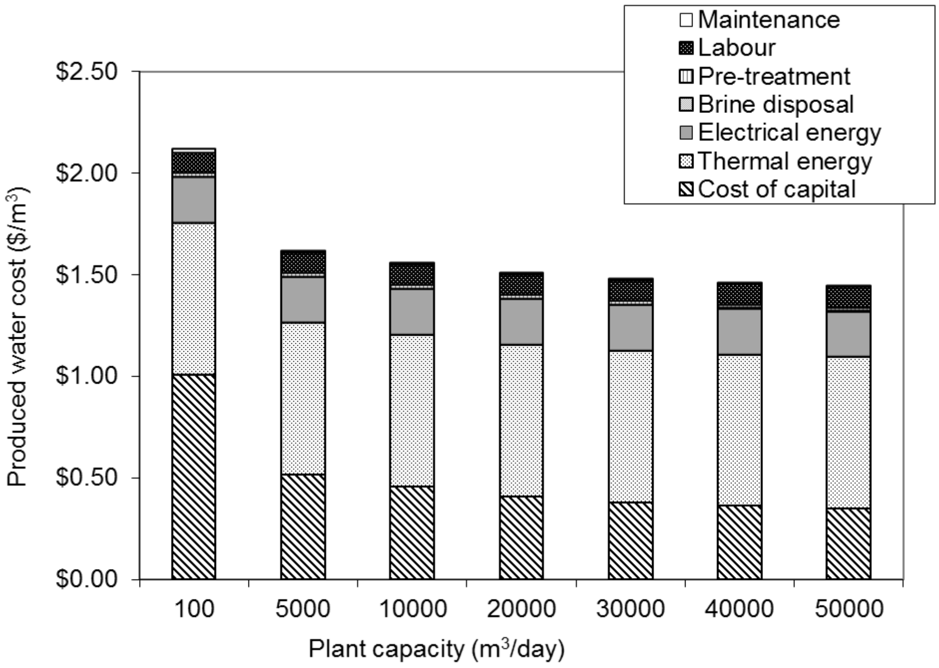

Figure 6: Breakdown of water production costs treated by MED

3.4 Cost impact of carbon tax on MD, RO and MED

The amount of carbon dioxide emitted to produce a cubic metre of potable water by desalination will depend on the source of energy used for both thermal and electrical energy, the amount of chemical used in the process and life of consumable items such as the membrane. Offsetting the carbon emissions associated with energy increases the total cost of desalination for MD, MED and RO as shown in Figure 7. With the inclusion of carbon of $23 per tonne carbon, and production capacity of 30,000 m^3/day, the overall production cost for MD is $2.20/$m^3$, MED is $1.77/$m^3$ and RO is $0.80/$m^3$. Again, RO is still the most cost effective desalination technology when a carbon tax is applied when no waste energy is used. In this context, RO would be most viable for desalination but reaches its TDS limit of around 70,000 mg/L where the thermal processes like MED and MD (demonstrated by data in Figure 2 and Table 3) can concentrate much beyond this. If the thermal energy had to be produced for these processes (i.e. as steam in Figure 3), the higher thermal energy requirement for MD over MED means it is more sensitive to the carbon tax. However this assessment does not take into account changes in emissions produced from the different construction materials of MD systems (utilising mostly polymeric materials) compared to MED (mostly metallic materials). A life cycle assessment (LCA) is therefore needed to assess this possibility for reduced carbon emissions of MD. Such assessments have already been conducted, and do indicate overall environmental benefits of the MD process [50, 51].

However, there is a cost saving for the thermal technologies when they are driven by a waste heat source while a carbon tax is applied. As shown in Figure 7, the cost of MD desalinated water is still more economical, with a total cost of $0.66/$m^3$, compared with $0.88/$m^3$ for MED and $0.80/$m^3$ for RO. Again, MD becomes the cost effective desalination technology when waste heat is applied due to the cheaper materials of construction and strongest sensitivity to thermal energy cost.

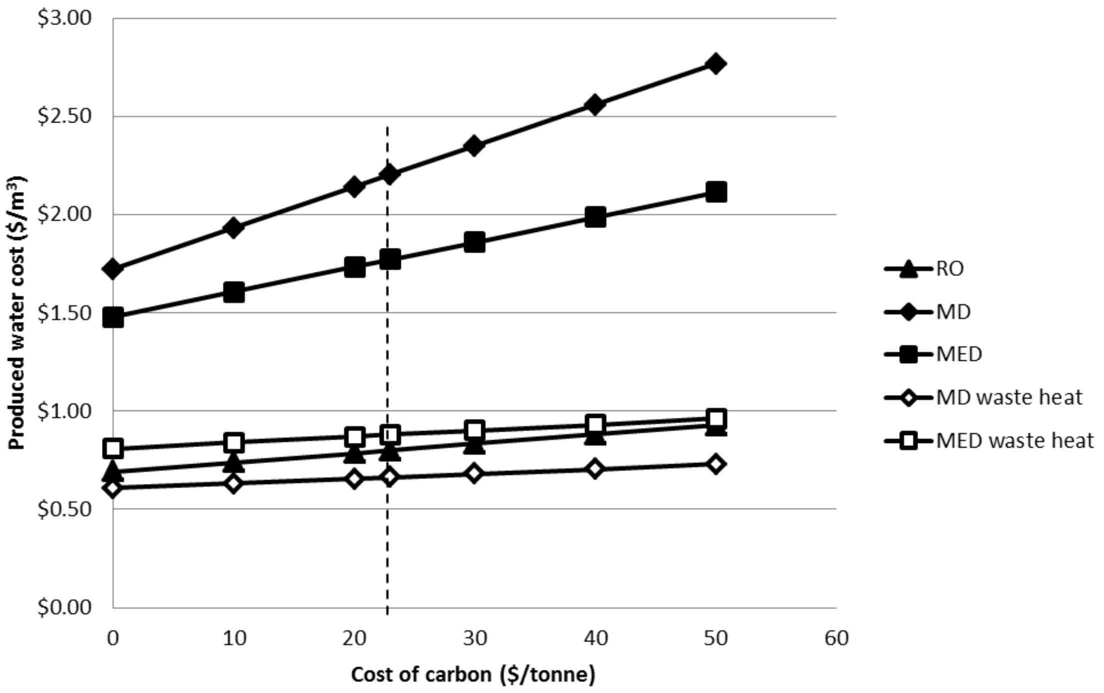

Figure 7: Impact of carbon tax on water price for RO, MED and MD with steam (solid symbols) and with waste heat (hollow systems). Dotted line indicates $23 per tonne carbon tax.

3.5 Thermal vs. electrical efficiency in MD: case for low grade waste heat

State of the art MD systems feature innovative arrangements which minimise the loss of the latent heat that is an essential part of the MD process. The internal heat recycling systems borrow from the thinking in conventional thermal processes [46]. These are ideally placed when the cost of the thermal energy is higher, for example higher temperature sources, or when supplied from solar panels (i.e. high grade). However thermal processes have optimal economics depending on the balance between the cost of the energy (operating cost) and the system complexity (capital cost) [45].

Based on this concept, when the cost of low grade heat is low due to lower temperature (<50 ^{o}C), a less thermally efficient system is therefore more economical. Further, heat recycling becomes less efficient when lower temperatures are supplied. Such cases are significant in industry, as less thermally efficient MED has better economics if it leads to cheaper capital [46]. For example in closed cycle power station condensers, anaerobic digesters, and industry heating services, exhaust heat is available at much lower temperatures and can be relatively abundant. As an example of this abundance, 500 MW of thermal energy from power station condensers can (assuming 100% to latent heat) evaporate 19 million litres of water per day even at low temperature (<50°C). This is a significant desalinated water volume from a low temperature waste heat that is commonly discharged to the sea or other large water body.

To apply this opportunity to MD, the MD plant works better without heat recycling. This is because lost flux due to the lower temperature can be restored by operating at higher membrane cross flow velocity. This however means heat recycling is not possible. A recent trial of MD on a sea water cooled power station's waste heat from our group [52] demonstrated that electrical efficiency can be 1.9 kWh/m^3 based on: DCMD with no heat recovery, a heat supply of around 35 ^{o}C, cooled by seawater, and flux of 4 kg/m^2/h (membrane replacement cost of \$0.051/m^3 by Equation 6). If this system required 1,200 kWh/m^3 of this 35 ^{o}C heat, the 500MW thermal load to the sea could desalinate 10 million litres of water per day. This quantity would take 0.8 MW of the power station's 500 MW electrical capacity to run the circulation pumps. Assuming the equivalent of 1% of the full price of steam to pay for this heat, the desalinated water cost becomes \$0.57/m^3 (carbon tax applied, scaled to 30,000 m^3/day reference plant). Further electrical energy reductions are possible through MD, for example in a MDHX module; electrical demand can be as low as 0.01 kWh/m^3 [28, 53]. Assuming this electrical demand is in practice 1 kWh/m^3, under a carbon tax and reference plant scale, we see cost drop to \$0.45/m^3. Therefore, MD seems economically viable in cases where low thermally efficient, but high electrically efficient systems are employed to make best use of abundant lower temperature heat. For this low grade waste heat case, MED has not been included in the costing as it is assumed MED will not effectively function at such low temperatures. A simple evaporator would be the technical equivalent, but subject to similar economics as MED due to the greater cost of the materials of construction.

This investigation acknowledges the efforts to offset greenhouse gas emissions by constructing renewable energy harnessing facilities (such as solar collectors and wind turbines). However despite these efforts, it is valid to argue that the power from these facilities would be better offsetting emissions of the existing high value electricity demand instead of a newly constructed desalination plant. Unless our electricity is sourced in a major way from renewables, the power grid still is supplied by 90% fossil fuels [46, 54], so although a positive outcome of building a desalination plant is the construction of a renewable energy power station, the power delivered to the desalination plant from the grid is still majority supplied by fossil fuels, and realistically will be the major proportion for decades to come. This makes the ability to desalinate water from existing energy sources that are currently discarded more attractive as a means to produce low carbon treated water today. This is the outcome we have found in our economic assessment that compares MD with the more established desalination technologies RO, MSF and MED. It is therefore concluded that MD is viable both economically and environmentally (low carbon context) to desalinate water when a low cost waste heat source is found. However this heat source would need to be considered as 'waste', i.e. it cannot be minimised by the source process' efficiency improvements. A water treatment operation could be considered a valuable use of currently discarded heat. Despite efforts to improve process efficiencies, there are viable sources of such heat, for example the abundant waste heat exhausted from a power station at 40 °C, which has no other value internally, but valuable as a source of energy for MD. With increased process integration, it is possible to explore options on utilising the existing heat paths in processes to become the heat source and sink to treat a separate process stream by MD[28]. Therefore the constraint to find 'waste' heat sources is lifted, but the compromise is increased process complexity.

4. Conclusion

- The cost of desalination schemes will increase by introduction of a price for carbon, but RO still remains lowest cost;
- Under a carbon tax, MD has best economics when the heat source has a low cost (e.g. waste heat). Specifically compared to MED, MD has lower cost materials. Compared to RO, MD has lower reliance in electricity;
- When fed with steam, MSF is the most costly desalination process, while MED and MD are similar and RO has the best economics;
- MD has the capacity to desalinate RO groundwater concentrate to hypersaline concentrations demonstrating its viability as a high recovery desalination technology;
- MD can also cost effectively harness abundant low grade heat sources or be integrated into existing processes.

Table 4 summarises the outcomes of the cost modelling.

Table 4: Summary costing of MD, RO and MED desalination for 30,000 m^3/day reference plant. Units in $\$/m^3$. Carbon tax of $23 per tonne carbon used when applied to costing.

Desalination case	**MD**	**RO**	**MED**
No carbon tax and driven with low pressure steam and electricity	1.72	0.69	1.48
Carbon tax applied and driven with low pressure steam and electricity	2.20	0.80	1.77
No carbon tax and driven with high temperature waste heat and electricity	0.61	0.69	0.81
Carbon tax applied and driven with high temperature waste heat and electricity	0.66	0.80	0.88
Carbon tax applied and driven with low temperature waste heat and electricity (low thermally, high electrically efficient mode)	0.57	0.80	N/A

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Appendix

A sample calculation of water production cost using MD system when a low grade heat source is available and a carbon tax

Plant capacity: 30,000 m^3/day

The annual production capacity:

$$30{,}000\,\frac{m^3}{day}\times 365\,\frac{days}{yr}\times 0.9 = 9855000\,\frac{m^3}{yr}$$

Capital cost: The capital cost is estimated using Equation 1, capacity based factor equation (see section 2.2.3)

$$= 1131\times 24{,}000\left[\frac{30{,}000}{24{,}000}\right]^{0.8} = \$32{,}449{,}040$$

Normalised capital cost: The cost of capital per m^3 water produced is estimated using Equations 2 and 3 in section 2.2.4, the Net present Value method

$$\frac{0.05\times(1+0.05)^{30}}{(1+0.05)^{30}-1}\times\frac{32{,}449{,}040}{9{,}855{,}000} = \frac{\$0.21}{m^3}$$

Cost of electricity: The cost of electricity is estimated using Equation 4:

$$0.09\times 2 = \frac{\$0.18}{m^3}$$

Thermal energy cost: Thermal energy cost is estimated using the same method for annual cost of electricity. However, when a low or high grade heat is available, we assumed the cost is equal to 10% of the total cost of thermal energy.

$$0.0124\times 100 \text{ x } 10\% = \frac{\$0.124}{m^3}$$

Brine disposal cost: The brine disposal cost from Table 1 is \$0.0015/m^3.

Membrane replacement cost: The membrane replacement cost is estimated using Equation 6

$$9\times 20\%\times\frac{1{,}000}{6\times 8760} = \frac{\$0.034}{m^3}$$

Pre-treatment cost: The pre-treatment cost from Table 1 is \$0.019/m^3.

Labour Cost: The labour cost taken from Table 1 is $0.03/m^3.

Maintenance cost: maintenance cost is estimated as 2% of the normalised capital cost.

$$0.21 \times 2\% = \frac{\$0.0043}{\mathrm{m}^3}$$

Emission cost

Emission cost takes into account both the thermal and electrical energy. However, when waste heat is available, the cost of emission using thermal energy is assumed to zero. Only electrical energy is therefore considered. This is estimated using Equation 5

$$2 \times 1.22 \times \frac{23}{1000} = \frac{\$0.056}{\mathrm{m}^3}$$

Total water treatment cost:

=0.21 + 0.12+ 0.18 + 0.0015 + 0.034 + 0.019 + 0.03+ 0.0043 + 0.056 =$0.66/m^3