Forward osmosis desalination from laboratory to market
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ABSTRACT
A two-step forward osmosis (FO) desalination process combining both FO and reverse osmosis (RO) systems has been developed by the Centre for Osmosis Research and Applications at the University of Surrey and commercialised by Modern Water plc. In the FO + RO process seawater was used as feed water (FW) and a concentrated aqueous solution was used as a draw solution (DS). By taking advantage of natural osmosis, pure water is transferred from the FW to the DS and then recovered from the DS by the RO process utilising low resistance membranes, and hence lower specific energy consumption (SEC). This paper presents results of FO experiments conducted on flat sheet membrane using a bench-scale rig. The osmotic agent investigated in this study was magnesium sulphate, which is non-toxic, and highly soluble in water. Furthermore experiments were carried out on the RO pilot in order to regenerate the DS for reuse in the FO process and produce clean water. This paper also presents some pilot plant results and data from commercial plants in Oman and Gibraltar. The data demonstrates the efficiency of the FO + RO compared with the conventional RO process in terms of SEC and membrane fouling performance.

Key words | draw solution, forward osmosis desalination, low fouling, reverse osmosis, specific energy consumption

NOMENCLATURE

<table>
<thead>
<tr>
<th>Abbreviation</th>
<th>Definition</th>
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<tr>
<td>FO</td>
<td>forward osmosis</td>
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<tr>
<td>DS</td>
<td>draw solution</td>
</tr>
<tr>
<td>MW</td>
<td>modern water</td>
</tr>
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<td>FW</td>
<td>feed water</td>
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<td>CORA</td>
<td>Centre for Osmosis Research and Applications</td>
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<td>UN</td>
<td>United Nations</td>
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<td>ED</td>
<td>electro dialysis</td>
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<td>RO</td>
<td>reverse osmosis</td>
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<td>MSF</td>
<td>multi-stage flash distillation</td>
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<tr>
<td>NF</td>
<td>nano-filtration</td>
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<tr>
<td>WHO</td>
<td>World Health Organization</td>
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<tr>
<td>SEC</td>
<td>specific energy consumption</td>
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<tr>
<td>TDS</td>
<td>total dissolved solids</td>
</tr>
<tr>
<td>M</td>
<td>mole</td>
</tr>
<tr>
<td>ICP</td>
<td>internal concentration polarisation</td>
</tr>
<tr>
<td>ECP</td>
<td>external concentration polarisation</td>
</tr>
<tr>
<td>EPC</td>
<td>engineering, procurement and construction</td>
</tr>
<tr>
<td>CDR</td>
<td>coefficient of desalination reality</td>
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</tbody>
</table>

INTRODUCTION
Shortage of freshwater is one of the greatest challenges, particularly in developing countries, where the population figures rise exponentially. Most Middle Eastern countries, for example, have less than 500 m$^3$ per capita water requirement of renewable natural water sources, while the minimum limit set by the UN is 1,700 m$^3$ (United Nations World Water Development Report 2005). Keeping up with the requirements for proper sanitation and water treatment
in these countries is found to be increasingly complicated. Freshwater can be produced from seawater, brackish water, or wastewater, using different desalination technologies. As the cost of desalination is determined by water salinity, feed water (FW), containing the minimum amount of impurities, is most favourable for use in the desalination processes leading to lower costs.

Seawater and brackish water are desalinated by various methods such as pressure-driven membrane separation processes, thermal distillation and electro dialysis (ED) while all these methods involve high operating and capital costs. The high operating cost of membrane-based methods such as reverse osmosis (RO) are due to essential pretreatment, scaling, bio-fouling and high-energy consumption while scaling and low thermal efficiency are the main constraints of thermal distillation techniques such as multi-stage flash distillation (MSF). The RO process, commercialised in 1960, became more competitive with the historical thermal desalination techniques in the 1980s, and since the mid-1990s the worldwide installed capacity of RO plants has been exceeding that of thermal plants (Global Water Intelligence 2010). A review by the National Research Council of the National Academies (2004) strongly recommended the support of further research and development in the application of novel membrane-based technologies for desalination to reduce energy and capital costs and brine disposal. The proposed process would overcome many of the practical problems associated with these conventional processes and hence reduce the operating costs. The review states that the most optimistic limit of achievement is a 50–80% capital and operating cost reduction, coupled with a similar increase in energy efficiency using the application of new ‘break-through’ technologies over the next 10 years. By the year 2020, the review states that desalination and water purification technologies will contribute significantly to ensuring a safe, sustainable, affordable, and adequate water supply.

The review states that to obtain further reductions in energy consumption, a different desalination approach is recommended.

Forward osmosis (FO) process is one of the recent developments and the most promising desalination technique which has the potential to provide a reliable and cost-effective method for producing freshwater with low energy consumption. In the FO process, pure water flows out of seawater or any impure water that has lower osmotic pressure, across a selective permeable membrane to dilute the draw solution (DS) with a higher osmotic pressure. Then, the diluted DS goes to the regeneration unit in order to separate and recycle DS for reusing and extract freshwater as the product (Figure 1). FO desalination process has faced two main challenges: (a) selecting a suitable DS as the main source of the driving force with a sustainable regeneration method; and (b) the structure and material of the appropriate membrane (Wang et al. 2012).

In the first stage of this study, a laboratory-scale FO setup was employed to investigate the effect of varying operation conditions such as temperature, concentration and cross-flow rate of FW and DS on the water flux and salt rejection using commercially available RO ultra-low pressure membrane. For the second stage, the performance of a FO coupled with RO pilot unit was determined in terms of specific energy consumption (SEC). The RO pilot unit was used to regenerate the DS for reuse in the FO process and produce clean water. Furthermore, this paper presents the background development of the hybrid FO + RO process that has been further developed and taken from the laboratory to the market for seawater desalination in Gibraltar and Oman (Sharif 2007; Nicoll 2015).

MATERIALS AND METHODS

The FO + RO pilot plant, which has been installed in CORA at the University of Surrey, was designed for hybrid FO and RO processes (Sharif 2007). In this study, the FO experiments were carried out using a dual channel flatbed cell in a recycle manner creating a counter current flow of feed and DS on both sides of membrane. In FO laboratory-scale, firstly deionised water, secondly a solution containing 10,000 ppm and finally 40,000 ppm NaCl to resemble brackish water and seawater were applied as FW, respectively. Furthermore, magnesium sulphate (MgSO₄), with concentration varying from 2.076 to 1.744 M, was used as the DS. The OLI Stream Analyzer 2.0 (2005) was applied to calculate the osmotic pressures of MgSO₄ based on its thermodynamic modelling on published experimental data to calculate the properties of solutions under different conditions. The flat
A sheet membrane TFC-ULP produced by Koch Membrane Systems was considered for the FO laboratory-scale unit. The FW and DS flowed from the beaker to the chamber of the cell and back to the beaker independently by FW and DS centrifugal pumps (Totton pump, AD 4/90). The flow rates were monitored with a flow metre (Rotameter, Key Instrument, FR 4500) while the volumetric flow rate of feed and DS were held at 1, 2 and 3 L/min. Temperature of both feed and DS were varied at 15, 25 and 35 °C using a heater (Hydro, Theo 200W) and an immersion cooler (SBS Labscience, TUR-1) with a temperature controller (Thermometer, Fisher, FB 68605) and were kept homogeneous by two independent magnetic stirrers (Thermolyne, Cimarec 3). Two analytical balances (Kern, CB12K1N) were used to monitor the weight changes of the DS and FW and calculate the water flux across the membrane quantitatively. The conductivities of FW and DS were measured with conductivity metre (Windaus Labortechnik, Winlab Data Line) before and after each run to determine the solute rejection by the membrane.

In the RO pilot test unit, the FW containing 6,300 ppm of NaCl and 125,000 ppm of MgSO₄ was tested to simulate the expected obtained DS from the FO unit. The testing involved investigation of the effects of varying the feed pressure and flow rate on the performance of the RO system based on permeate flow rate and SEC. The RO experiments were conducted using high-pressure hollow fine fibre (Toyobo) and spiral wound (Dow Filmtec) membranes in a pilot unit. The FW (diluted DS) was circulated into the membrane module by a reciprocating high-pressure pump (Grundfos) and the hydraulic feed pressure was varied up to 60 bar. The pump discharge pressure was controlled by adjusting the back-pressure regulator. The applied pressure and flow rate of feed, concentrate and permeate lines were monitored by pressure gauges (Wika, EN 837–1) and flow metres (E&H Instruments), respectively. All tests were done at 25 °C and feed flow rate was maintained constant at ~22 L/min through all experiments by setting the feed pump on a constant speed of 1,000 rpm. The FO + RO set-up is illustrated in Figure 2.
RESULTS AND DISCUSSION

The performance of the FO process

The performance of the FO process was investigated to achieve optimum operating conditions including DS and FW concentrations, operating temperature and cross-flow rate. Magnesium sulphate (MgSO₄) (MW₆₇₄: 120.4) was considered to be used as a DS in the FO process. A suitable DS has high osmotic pressure, high solubility in water, low reverse diffusion and has a relatively high molecular weight for easy separation in the regeneration system. Furthermore, the DS must be non-corrosive and should not be harmful to the human body. MgSO₄ has large molecular weight and high charged ions such as SO₄²⁻ which are more highly rejected than monovalent ions by a negatively charged NF membrane (Ng et al. 2006). A larger and highly charged DS can be separated more easily and cost-effectively, when utilising membranes with high porosity and high mean pore diameter, to yield potable water. Although the osmotic pressure of MgSO₄ leads to higher feed concentration requirement, acceptable water flux and maximum solute rejection for the hybrid FO + RO process resulting in good quality product water was the main target in this study.

Effect of feed and DS concentration

The effect of varying concentrations of the feed and DS on the water flux is illustrated in Figure 3. Three concentrations of FW were obtained to simulate deionised, brackish and seawater as feed solution while DS concentration was varied from 0.35 to 2.05 M. The cross-flow velocity and temperature of both feed and DS were maintained constant at 3 L/min and 25 °C, respectively.

It is observed that as the concentration or osmotic pressure of the DS increases, the water flux for all FW also rises. However, the increase in water flux is significantly declined when the FW concentration is increased simultaneously. The feed solution containing 40,000 ppm NaCl resembling seawater produced much lower water flux than the remaining two feed solutions which suggests that the combined concentration polarisation effects are much more severe for seawater. Osmotic pressure is a function of the concentration of salts contained in aqueous solution and can be enhanced by increasing the concentration of
solute. Higher salt concentration of FW, i.e. seawater results in lower water flux through the membrane; therefore, higher concentrations of DS is required to achieve high effective osmotic driving force. Furthermore, the results showed that increasing the DS concentration leads to a decrease in the percentage utilisation of the available bulk osmotic driving force due to rising in the severity of the internal concentration polarisation (ICP). Therefore, a trade-off between the water flux and DS concentration should be considered in designing the FO process.

Effect of feed and DS flow rates

In the next step, the effect of cross-flow velocity on water flux in the FO process was investigated by changing the FW and DS flow rates from 1 L/min to 3 L/min, respectively. All the conditions were kept the same including the operating temperature of FW and DS which were maintained at 25 °C. The result of 40,000 ppm NaCl, as FW was discussed for the FO seawater desalination. Figure 4 illustrates that the water flux rises non-linearly as the cross-flow velocity of feed or DS increases while varying the concentration of MgSO₄ from 1.7 to 2.1 M.

By varying the flow rate of the DS the permeate water flux was relatively higher than varying the FW flow rate. It was concluded that dilutive ECP on the DS side has a significant impact on the osmotic driving force compared to the minor effects of concentrative ECP on the feed side.

Effect of operating temperature

The water flux data were collected at 15, 25 and 30 °C in order to determine the temperature effect on the FO performance. Figure 5 shows the non-linear trend raising the water flow rate when changing the DS concentration from 1.7 to 2.1 M.

The gathered data showed that by increasing the operating temperature in both membrane sides especially in the FW, the water flux is enhanced accordingly. By increasing the temperature, the viscosity of water decreases which raises the diffusion rate of water through the membrane and water permeability coefficient. Raising the temperature results in increasing both diffusion and mass transfer coefficients of the FW. Figure 6 shows that although water flux in the FO process trend rose at high temperatures, the salt rejection decreased due to increasing mass transfer coefficient of the NaCl feed solution and reducing the impact of external concentration polarisation.

The temperature effect is also similar on the ICP where the increased mass transfer coefficient reduces solute resistivity and increases ICP exponentially. Furthermore, the
effect of dilutive ICP was found to have a significant impact on the driving force due mostly to the fact that the phenomenon was acting on the concentrated DS. The results have indicated that as the temperature of the DS is increased, the flux through the membrane rises and this can be partly attributed to the reduction in the effect of the dilutive ICP. The water flux across the membrane is sensitive to change in the temperature of FW and DS, more so for the DS due to decreasing the effect of dilutive ICP. For instance, the flux produced is slightly greater for a DS of 35 °C than for a FW of 35 °C. As the temperature of the DS was raised, the water flux across the membrane increased which is partly due to the reduction in the effect of dilutive ICP. The drawback is that the increased temperature also resulted in lower salt rejection and salt selectivity due to a higher diffusion rate of salt through the membrane.

The performance data of the FO process showed that the overall performance in terms of water flux and salt
rejection was satisfactory; however the development of a more specialised membrane will lead to improved performance of the process. The sustainable regeneration stage for the selected DS is described in the next section.

Performance of DS regeneration based on RO unit

The performance of the RO pilot unit as the regeneration system for DS in FO process was determined in terms of evaluating the quality of produced clean water and SEC in this investigation. Two types of membrane, including hollow fine fibre and spiral wound membranes, were used in the RO test and involved investigation of the effect of varying the feed pressure and flow rate of FW on the RO performance. Osmotic pressure values for the saline solutions were calculated using OLI software depending on salt concentrations. Salt concentrations for feed, concentrate and permeate solutions have been calculated by laboratory chemical analysis.

RO experiments at constant feed flow rate

Table 1 provides the results of the set of experiments which were carried out at a constant feed flow rate with different feed pressure values.

As expected with the ordinary RO processes, water flux and salt rejection were increased with rise in FW pressure.

Table 1 | RO experiments results at 25°C and constant feed flow rate under different feed pressures

<table>
<thead>
<tr>
<th>Feed pressure, bar</th>
<th>40</th>
<th>47</th>
<th>54</th>
<th>60</th>
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<tbody>
<tr>
<td>Hollow fine fibre membrane</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Water flux, l/m² h</td>
<td>0.184</td>
<td>0.323</td>
<td>0.431</td>
<td>0.521</td>
</tr>
<tr>
<td>Permeate TDS, ppm</td>
<td>667</td>
<td>346</td>
<td>247</td>
<td>211</td>
</tr>
<tr>
<td>SEC, kW h/m³</td>
<td>8.9</td>
<td>5.9</td>
<td>5.1</td>
<td>4.6</td>
</tr>
<tr>
<td>Spiral wound SW-4040 membrane</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Flux, l/m² h</td>
<td>2.736</td>
<td>6.367</td>
<td>11.005</td>
<td>13.506</td>
</tr>
<tr>
<td>Permeate TDS, ppm</td>
<td>988</td>
<td>417</td>
<td>290</td>
<td>224</td>
</tr>
<tr>
<td>SEC, kW h/m³</td>
<td>34.6</td>
<td>17.3</td>
<td>11.4</td>
<td>10.2</td>
</tr>
</tbody>
</table>

Furthermore, the performance of RO process in terms of the SEC was increased as the FW pressure augmented. The hollow fibre membrane shows lower SEC due to its high productivity and recovery rate compared with the spiral wound membrane.

RO experiments at constant feed pressure

Table 2 illustrates the numerous experiments undertaken with different feed flow rates at a constant feed pressure of 60 bars to find the optimum operating conditions in the RO stage of the hybrid FO + RO process.
The water flux and membrane salt rejection (within the permeate TDS) increased with a rise in feed flow rate. Moreover, the SEC increased as the FW flow rate rises. The hollow fibre membrane showed lower SEC at constant feed pressure with variable feed flow rates due to its high productivity and recovery rate compared with the spiral wound membrane.

Results obtained from the RO test indicated that the hybrid FO + RO scheme has the potential to be a more efficient process for seawater desalination in terms of water flux and energy consumption (Sharif 2007, 2008a, b, 2010; Sharif & Al-Mayahi 2007, 2011). The work was then taken from the laboratory to the field through the Modern Water (MW) plc development programme (Thompson & Nicoll 2011). A summary of the FO process development from laboratory to a commercial facility is provided in the next section.

**FO desalination from laboratory to market**

MW is one of the few companies which have developed a FO desalination process from laboratory scale to a commercial scale. MW’s development programme began at the laboratory's pilot test facility (Figure 7) in the CORA centre at the University of Surrey.

The performance of various membranes, physical configuration and analytical techniques were investigated by CORA team members to provide the data required for scaling up. Specifically designed FO membranes were supplied to MW by a third party. It must be noted that the details of membranes are commercially sensitive and hence are not presented here. The component of the osmotic agent was based upon a low-cost, non-toxic, commodity chemical and in particular suitable for use in drinking water facilities. The regeneration system was also membrane based and its energy consumption was minimised by careful selection of the osmotic agent and the system operating conditions.

Table 2 | RO experiment results at 25°C under 60 bar feed pressure with different feed flow rates

<table>
<thead>
<tr>
<th>Feed flow rate, l/min</th>
<th>18.1</th>
<th>20.6</th>
<th>22.5</th>
<th>26.4</th>
<th>29.7</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hollow fine fibre membrane</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Flux, l/m²h</td>
<td>0.460</td>
<td>0.521</td>
<td>0.550</td>
<td>0.625</td>
<td>0.684</td>
</tr>
<tr>
<td>Permeate TDS, ppm</td>
<td>227</td>
<td>211</td>
<td>178</td>
<td>147</td>
<td>142</td>
</tr>
<tr>
<td>SEC, kW h/m³</td>
<td>4.6</td>
<td>4.6</td>
<td>4.8</td>
<td>4.9</td>
<td>5.0</td>
</tr>
<tr>
<td>Spiral wound SW30-4040</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Flux, l/m²h</td>
<td>12.690</td>
<td>13.506</td>
<td>13.712</td>
<td>14.948</td>
<td>15.707</td>
</tr>
<tr>
<td>Permeate TDS, ppm</td>
<td>234</td>
<td>224</td>
<td>214</td>
<td>180</td>
<td>168</td>
</tr>
<tr>
<td>SEC, kW h/m³</td>
<td>9.7</td>
<td>10.2</td>
<td>11.0</td>
<td>11.8</td>
<td>11.9</td>
</tr>
</tbody>
</table>

Figure 7 | FO + RO pilot plant test rig.
The results were very promising and supported the need for scaling-up to large commercial applications. The first industrial-scale installation of the FO Desalination Technology was commissioned by MW in Gibraltar in September 2008 (Thompson & Nicoll 2011).

The test facility used the RO process to regenerate the DS; however there was also provision of a nano-filtration (NF) based system. The plant has been successfully delivering 18 m$^3$/day of water to the local drinking water system since 1 May 2009. The feed was delivered through a shared pretreatment facility with silt density index (SDI) figures of between 3 and 4. The product water has a TDS of less than 200 mg/L and boron levels of less than 0.6 mg/L and the plant is currently successfully operating with MW’s third generation of full-scale FO membranes (Thompson & Nicoll 2011).

Based on successful results from the trial site, this was followed by a much larger plant (100 m$^3$/day) located at the Public Authority for Electricity and Water’s site (PAEW) at Al Khaluf in Oman (Figure 8) in July 2009 (Thompson & Nicoll 2011).

The containerised water plant was designed to tie in with the existing pretreatment and post-treatment equipment. The FO system has been operating successfully with a seawater recovery of 35% and the product water has a TDS of less than 200 mg/L and boron content of between 0.6 and 0.8 mg/L. The benefits of the FO system may best be illustrated by comparison of the FO with the performance of the existing seawater reverse osmosis desalination (SWRO) facility on the site. Figure 9 shows the normalised output from the adjacent SWRO facility. Despite repeated cleaning of the membranes, there is a 30% decline in the output from new membranes in just 5 months (Thompson & Nicoll 2011).

In contrast, the normalised output from the FO system, over the same period has remained relatively unchanged (Figure 10). This is despite the fact that no chemical cleaning has been performed, and operation has been at a higher seawater recovery.

The plant has operated without any chemical cleaning, changing of membranes and decline in productivity since November 2009. This demonstrates significant advantages including lower energy consumption as well as particularly low fouling. Table 3 illustrates the performance of the FO + RO facility in comparison with the conventional SWRO facility at the Al Khaluf site in Oman.

**Figure 8** | The PAEW Water site at Al Khaluf (including MW’s containerised FO + RO plant).
The MW plc was granted an Engineering, Procurement and Construction contract for a 200 m$^3$/day FO desalination plant at Al Jardah again in Oman (Thompson & Nicoll 2011). In an independent study by the Global Water Intelligence (2010), the manipulated osmosis (FO + RO) process was ranked first and awarded the highest coefficient of desalination reality among all new desalination technologies.

The potential of FO desalination for low energy consumption could be summarised in a number of ways:

- FO units foul at a lower rate than conventional RO systems therefore they require less energy to maintain the same output than RO units. In addition, fouling in FO is reversible, i.e. can be removed by back flushing, while most fouling in RO is irreversible, which requires chemical cleaning and in severe conditions, membrane replacement.

- However, the majority of energy in the FO unit is used for regenerating the draw agent. Optimisation of the DS concentration and composition, configuration and type of membrane and applying an energy recovery system can minimise steps of RO or NF in the regeneration stage for saving energy up to 30% (Thompson & Nicoll 2011).

- Removal of boron in the RO desalination process to meet the varying international standards may require additional treatment steps such as ion exchange or multi-stage RO units. These increase not just capital costs but operating costs due to higher energy consumption. The regeneration steps.

### Table 3

<table>
<thead>
<tr>
<th>Permeate extraction from FW</th>
<th>Unit</th>
<th>SWRO</th>
<th>FO + RO</th>
</tr>
</thead>
<tbody>
<tr>
<td>FW recovery</td>
<td>%</td>
<td>25</td>
<td>35</td>
</tr>
<tr>
<td>Product water flow</td>
<td>m$^3$/d</td>
<td>71.4</td>
<td>100</td>
</tr>
<tr>
<td>FW pump pressure</td>
<td>bar</td>
<td>65</td>
<td>4</td>
</tr>
<tr>
<td>FW pump absorbed power</td>
<td>kW</td>
<td>25.3</td>
<td>1.6</td>
</tr>
<tr>
<td>DS regeneration DS recovery</td>
<td>%</td>
<td>47</td>
<td></td>
</tr>
<tr>
<td>Diluted DS feed pump pressure</td>
<td>bar</td>
<td>65</td>
<td></td>
</tr>
<tr>
<td>Diluted DS feed pump absorbed power</td>
<td>kW</td>
<td>18.8</td>
<td></td>
</tr>
<tr>
<td>Overall plant SEC</td>
<td>kWh/m$^3$</td>
<td>8.5</td>
<td>4.9</td>
</tr>
</tbody>
</table>

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**Figure 9** | Normalised output from adjacent SWRO Facility 2010 (Thompson & Nicoll 2011).

**Figure 10** | Normalised output from the FO plant 2010 (Thompson & Nicoll 2011).
step of the FO unit has the potential to decrease boron content because it is fed with a free foulant DS that improves salt rejection, eliminating the need to use extra treatment steps leading to lower energy consumption (Thompson & Nicoll 2011).

CONCLUSION

This paper presents the background development work of a patented FO desalination technology from laboratory to a commercial facility that addresses some of the practical problems associated with conventional desalination systems such as fouling and energy consumption. The process was taken from the laboratories of CORA at the University of Surrey and was commercialised by MW plc over the last 6 years, to be the world’s first developer of FO desalination process on a commercial scale. Results and data were offered to cover laboratory work in the UK, a trial facility installed in Europe on the Mediterranean Sea (Gibraltar) and a commercial plant operating in Oman on the Arabian Sea. Operational results taken from a commercial scale FO plant operating alongside a SWRO plant, located in Oman, utilising a common pretreatment system were outlined. The FO plant demonstrated significant advantages in performance, both in energy consumption and particularly very low fouling. The plant has operated over a 36-month period without any chemical cleaning, whereas the conventional plant had numerous chemical cleanings, a change of membranes and showed a marked decline in productivity over the same period. The FO plant also demonstrated the inherent capability for higher boron rejection than the conventional membrane plant. However, there are still areas in the ongoing programme with a number of membrane suppliers and research organisations for improving the performance of FO desalination with regards to the suitable membrane, the proprietary osmotic agent and regeneration stage.

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REFERENCES


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