The innovative Osmotic Membrane Bioreactor (OMBR) for reuse of wastewater


ABSTRACT

An innovative osmotic membrane bioreactor (OMBR) is currently under development for the reclamation of wastewater, which combines activated sludge treatment and forward osmosis (FO) membrane separation with a RO post-treatment. The research focus is FO membrane fouling and performance using different activated sludge investigated both at laboratory scale (membrane area of 112 cm$^2$) and at on-site bench scale (flat sheet membrane area of 0.1 m$^2$). FO performance on laboratory-scale (i) increased with temperature due to a decrease in viscosity and (ii) was independent of the type of activated sludge. Draw solution leakage increased with temperature and varied for different activated sludge. FO performance on bench-scale (i) increased with osmotic driving force, (ii) depended on the membrane orientation due to internal concentration polarization and (iii) was invariant to feed flow decrease and air injection at the feed and draw side. Draw solution leakage could not be evaluated on bench-scale due to experimental limitation. Membrane fouling was not found on laboratory scale and bench-scale, however, partially reversible fouling was found on laboratory scale for FO membranes facing the draw solution. Economic assessment indicated a minimum flux of 15 L.m$^{-2}$.h$^{-1}$ at 0.5 M NaCl for OMBR-RO to be cost effective, depending on the FO membrane price.

Key words | forward osmosis, membrane fouling, osmotic membrane bioreactor, salt flux, water flux

INTRODUCTION

Owing to more stringent regulations, extensive treatment of wastewater is becoming increasingly important. An energy-effective innovative osmotic membrane bioreactor (OMBR) is currently under development (Cornelissen et al. 2008; Qin et al. 2008; Achilli et al. 2009). An OMBR is an innovative membrane bioreactor for the reclamation of wastewater, which combines activated sludge treatment and forward osmosis (FO) membrane separation with a post-treatment, e.g. reverse osmosis (RO) (Figure 1). The driving force in FO is a concentration difference over the membrane surface using a draw solution (usually concentrated salt or sugar solutions) at the membrane product side, which draws pure water from the feed water side towards the product side. This is in contrast to RO processes, in which feed water is pressed through the membrane towards the product side at high pressures, resulting in an energy intensive process.

OMBR is a multiple barrier water reuse system which combines the advantages of conventional membrane bioreactors, and is expected to overcome the drawbacks of conventional membrane bioreactors, i.e. membrane fouling and a high energy demand (Cath et al. 2006). For the OMBR to be both technically and economically viable, the performance of FO membranes has to be sufficiently high, while membrane fouling and draw solution leakage (expressed as reverse salt flux) has to be low. Furthermore, transmission of pollutants from the feed towards the draw solution should be minimised, since this can compromise the product water quality. The effect of different types of activated sludge, concentration of the NaCl draw solution and membrane orientation on the performance of the FO membranes was investigated. Furthermore, membrane fouling and draw solution leakage was investigated using activated sludge.
The focus of this research is FO membrane fouling and performance investigated both at laboratory scale (using a FO membrane area of 112 cm$^2$) and at on-site bench scale (using a FO membrane area of 0.1 m$^2$) (Figure 2).

MATERIALS AND METHODS

Experimental set-up

FO laboratory scale experiments were carried out in an U-tube set-up and are described elsewhere in more detail (Cornelissen et al. 2008). FO bench-scale experiments were carried out in a newly developed set-up designed for continuous experiments (Figure 2). A see-through Perspex membrane cell for a flat membrane sheet with membrane area of 0.1 m$^2$ was constructed and vertically mounted into the bench-scale set-up. The flow channels at both feed side and draw solution side were originally the same (7 mm channel height), however, a spacer was inserted at the draw solution side for mechanical support. The cross section area of the flow channels at the feed side and draw solution side (with a spacer) of the flat sheet membrane were $7 \times 10^{-4}$ m$^2$ and $3.75 \times 10^{-4}$ m$^2$, respectively (Qin et al. 2009).

Fresh activated sludge was pumped through a 0.4 mm metal mesh screen and stored in a continuously stirred feed vessel (1.0 m$^3$) to prevent sedimentation. The pre-screened activated sludge was fed to the feed side of the membrane cell in a single pass mode. The concentration of the draw solution was kept constant during bench-scale testing by an automatic control of the conductivity in the draw solution loop interlocked with dosage of a concentrated salt solution (1.5 M or 3 M sodium chloride). The volume decrease of the concentrated salt solution vessel was recorded by a mass balance. The volume increase in the draw solution loop as a result of FO permeation was recorded by a mass balance registering the overflow of an open vessel within the draw solution loop. An air compressor was provided at the feed and draw side of the membrane cell to increase the turbulence in the flow channels for minimization of external concentration polarization (ECP). A data-logger was installed at the control panel of the bench-scale set-up to continuously monitor and record (i) permeate flow from the overflow vessel (ii) feed water conductivity and temperature, (iii) draw solution conductivity and temperature before and after dosing concentrated draw solution into the draw solution loop and (iv) the mass of the two balances.

Membrane material

A forward osmosis (FO-type) membrane (‘Explorer’ type of Hydration Technologies, Albany, OR) was obtained for the laboratory (U-tube) and bench-scale experiments. This membrane consisted of a thin active layer supported by an open support layer. The membranes were used on two sides, with the active layer on the feed solution side and with the active
layer on the draw solution side to study the influence of orientation on the membrane performance.

**Draw and feed solutions**

*Draw solutions.* Different sodium chloride solutions prepared in deionised (DI) water (laboratory scale experiments) or in locally available tap water (bench-scale experiments) were used as draw solutions. DI water was supplied by a Milli-Q plus device (Millipore Academic A10), and locally available tap water (conductivity of approximately 400\(\mu\)S/cm) was prepared from anaerobic groundwater, treated by aeration and rapid sand filtration and distributed without disinfectant residual. Osmotic pressure values of the draw solutions, theoretically calculated from the Van’t Hoff equation (Moore 1972), ranged of 25–75 bar.

*Feed solutions.* Activated sludge from different locations were investigated in laboratory scale experiments i) Ulu Pandan MBR Demonstration Plant (UP) (PUB, Singapore) (Qin et al. 2008) ii) WWTP Horstermeer (HM) (Waternet, the Netherlands) and iii) MBR pilot plant at Hilversum (HS) (Waternet, the Netherlands). Fresh activated sludge (period October 2007–February 2008) from the anoxic zone at WWTP Horstermeer (Waternet, the Netherlands) was obtained for bench-scale experiments. The average compositions of the different activated sludge types are provided in Table 1. The conductivity of the sludge was measured using a conductivity probe and varied during the research period (average values are included in Table 1).

**Forward osmosis experiments**

*Laboratory experiments.* Forward osmosis laboratory scale experiments were carried out to determine (i) the temperature dependency of the water and salt flux and (ii) the influence of activated sludge type on the FO performance. Determination of the water and salt flux from these laboratory scale experiments are described elsewhere (Cornelissen et al. 2008). The draw solution was 0.5 M sodium chloride and the active layer of the membrane was facing the feed solution during the laboratory scale experiments (unless stated otherwise). Additional surface characterization of virgin and fouled FO membrane surfaces was carried out with an X-ray photo-spectroscopy device (Quantera XPS from Physical Electronics). Samples were dried in ultrahigh vacuum (<6.7 \(\times\) 10\(^{-9}\) torr) and no degradation was observed as a result of the X-rays (Al K\(\alpha\), monochromatic at 1486.6 eV).

*Bench-scale experiments.* Forward osmosis bench-scale experiments were carried out to determine the influence of (i) orientation, (ii) draw solution concentration and (iii) the air and feed flow velocity on the performance of the FO process. Membrane fouling was investigated both on laboratory scale (Cornelissen et al. 2008) and on bench-scale. Feed water and draw solution flow during the bench-scale experiments was 800 L/h (feed flow velocity of 0.3 m/s) and 80–120 L/h (draw solution velocity of 0.06–0.08 m/s) respectively. Pressurized air of 150 NL/h was introduced at the feed water and draw side of the membrane cell during one of the bench-scale experiments. The draw solution was 0.5 M sodium chloride and the active layer of the membrane was faced to the draw solution during the laboratory scale experiments (unless stated otherwise). Although the set-up was originally designed to operate continuously, in this research tests were conducted with a maximum duration of 48 hours as a result of practical restrictions.

Water flux and salt flux were determined by mass balances in the draw solution loop (Figure 3) given by:

\[
J_w = \frac{V_p - V_s}{A_m \Delta t} \times 1000
\]

\[
J_s = \frac{c_p Q_s - c_p Q_p}{A_m} \times 1000
\]

Where, \(J_w\) is the water flux (L.m\(^{-2}\)h\(^{-1}\)), \(J_s\) is the salt flux (g.m\(^{-2}\)h\(^{-1}\)), \(V_p\) is the permeate volume (m\(^3\)), \(V_s\) is the

<table>
<thead>
<tr>
<th>WWTP parameter</th>
<th>Abbrev.</th>
<th>HS value</th>
<th>HM value</th>
<th>UP value</th>
<th>Unit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mixed Liquor Suspended Solids</td>
<td>MLSS</td>
<td>5</td>
<td>5</td>
<td>5</td>
<td>g/L</td>
</tr>
<tr>
<td>Conductivity</td>
<td>-</td>
<td>630</td>
<td>780</td>
<td>730</td>
<td>(\mu)S/cm</td>
</tr>
<tr>
<td>Chem. oxygen demand</td>
<td>COD</td>
<td>550</td>
<td>1300</td>
<td>230</td>
<td>mg/L</td>
</tr>
<tr>
<td>Biol. oxygen demand</td>
<td>BOD</td>
<td>250</td>
<td>380</td>
<td>140</td>
<td>mg/L</td>
</tr>
</tbody>
</table>
concentrated salt solution volume (m$^3$), $A_{\text{mbr}}$ is the membrane area (m$^2$), $\Delta t$ is the elapsed time, $c_s$ is the concentrated salt solution concentration (kg.m$^{-3}$), $Q_s$ is the concentrated salt solution flow (m$^3$.h$^{-1}$), $c_p$ is the permeate concentration (kg.m$^{-3}$) and $Q_p$ is the permeate flow (m$^3$.h$^{-1}$). Density variations due to the use of concentrated salt solutions were neglected (for 0.5 M and 1.5 M NaCl the increase in density in compared to water is respectively 3% and 9%).

**FO PROCESS PERFORMANCE ON LABORATORY SCALE**

**Temperature dependency of water and salt flux**

Both water and salt flux increased linearly with temperature between 10°C and 30°C (Figures 4 and 5), with slopes ranging between 0.09 and 0.22. The water flux increase was due to a decrease in viscosity of the feed water which increased the diffusion rate of water through the membrane resulting in a higher permeability (McCutcheon and Elimelech 2006). A higher slope (0.22) between water flux and temperature was observed for the FO membrane facing the draw solution. Temperature will also affect internal concentration polarization (ICP) which is lower for the FO membrane facing the draw solution resulting in higher water fluxes (Gray et al. 2006; McCutcheon and Elimelech 2006; McCutcheon et al. 2006). The influence of temperature on the water flux is more pronounced for the FO membrane facing the draw solution.

The salt flux increase was due to an increase of the diffusion rate of salt through the membrane resulting in a higher salt permeability (Ghiu et al. 2003). A higher slope (0.17) between salt flux and temperature was observed for the FO membrane facing the draw solution. This was again ascribed to the effect of ICP which is lower for the FO membrane facing the draw solution resulting in higher salt fluxes. The influence of temperature on the salt flux is more pronounced for the FO membrane facing the draw solution. Flux and salt flux values presented below were corrected for temperature (20°C) using the presented slopes (Figures 4 and 5).

**Influence of activated sludge type on the FO performance**

Activated sludge from different locations was investigated (UP, MH and MS). The performance of the forward process was similar for these different types of activated sludge. Water flux values were around 6–7 L.m$^{-2}$.h$^{-2}$ for the different activated sludge types with different COD concentrations and similar conductivity values (Figure 6), which was in the same order of magnitude of values found in literature (Cornelissen et al. 2008; Achilli et al. 2009). The osmotic pressures of the different activated sludge types were estimated at approximately 0.3 bar (Cornelissen et al. 2008), which was negligible compared to the osmotic pressure of the draw solution.

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**Figure 3** | Depiction of the water and salt flows in the bench-scale set-up during the FO experiments.

**Figure 4** | Water flux in function of temperature on laboratory scale (feed: DI water; draw solution: 0.5M NaCl; membrane orientation: active layer facing draw solution (●) and feed solution (○)).

**Figure 5** | Salt flux in function of temperature on laboratory scale (feed: DI water; draw solution: 0.5M NaCl; membrane orientation: active layer facing draw solution (●) and feed solution (○)).
solution. Because of the short experimental duration (7–8 hours) negative effects of potential membrane fouling phenomena could not be assessed (see Membrane Fouling Studies).

Salt flux values varied among the different activated sludge types between 2 and 7 g.m⁻²h⁻² (see Figure 7), in accordance with salt flux values found in literature for activated sludge (Cornelissen et al. 2008; Achilli et al. 2009). The highest salt flux was found for activated sludge with the lowest conductivity, which could not be explained. A thorough analysis of the ion balance of the different activated sludge types is necessary to explain the observed phenomena.

FO PROCESS PERFORMANCE ON BENCH-SCALE

Influence of draw solution concentration of FO performance

An increase in FO water flux was found for on-site bench scale experiments using HM-type activated sludge as a feed when the sodium chloride draw solution concentration increased from 0.5 M to 1.5 M, due to an increase in the osmotic driving force over the membrane (Figure 8). The shape of the curve is non-linear as a result of the effects of ICP, which reduces the effective driving force of the draw solution. Similar results have been obtained for activated and digested sludge in literature with water flux values in the same order of magnitude (Cath et al. 2006; Cornelissen et al. 2008; Qin et al. 2008; Achilli et al. 2009; Tang et al. 2010).

Salt fluxes could not be evaluated during the bench-scale experiments as a result of the measurement method. Salt flux values were determined from a mass balance in the draw side loop, by subtraction of the salt load towards the draw side loop and the salt load out of the draw side loop via the overflow vessel (see equation 2), resulting in strongly fluctuating and non realistic values. Salt flux values could not be obtained from the feed side because of a single pass mode. For further investigation, a recirculation approach of the feed side flow was suggested.

Influence of membrane orientation of FO performance

The FO water flux was higher when the active membrane layer faced the draw solution due to less ICP (Figures 8 and 9). For the FO membrane facing the draw solution ICP is caused by concentration of feed water components into the porous substructure of the membrane, while for the FO membrane facing the feed solution ICP is caused by dilution of the draw solution in the porous substructure of the membrane (McCUTCHEON and ELIMELECH 2006). The latter phenomenon is more severe due to the asymmetry of the porous substructure of the investigated FO membrane at a draw solution concentration of 0.5 M NaCl. FO membranes
with an improved support layer structure are expected to show a lower influence of the membrane orientation.

Similar results have been obtained when locally available drinking water was used for the bench-scale FO experiments (Figure 9 FS(DW) and DS(DW)). Higher water flux values were obtained with the FO membrane facing the draw solution. Higher water flux values were found for drinking water compared to HM-type activated sludge probably as a result of a lower conductivity leading to (i) slightly higher driving forces and (ii) less ICP.

The FO flux values during bench-scale experiments were higher compared to the FO flux values during laboratory scale experiments for both drinking water and activated sludge as feed water (Figure 9). This was ascribed to more favourable hydrodynamic conditions at the membrane surface during bench-scale experiments resulting into lower ECP phenomena.

**Influence of air and water flow on FO performance**

ECP can potentially decrease the driving force in FO processes in either way (i) dilution of the draw solution with the active membrane layer facing the draw solution or (ii) concentration of feed components with the active membrane layer facing the feed. ECP in FO is determined by the flow conditions outside membrane surface at the feed side and draw side. The influence of (i) the feed flow rate and (ii) air flow at the feed side and draw side was investigated in on-site bench-scale experiments for a FO membrane facing the draw solution. Water flux remained constant at approximately 5 L.m\(^{-2}\).h\(^{-1}\) despite reducing the feed flow (from 800 L/h to 400 L/h) and applying air flow (15 NL/h) on both sides of the membrane (Figure 10). ECP did not play a prominent role in the on site bench-scale experiments using HM-type activated sludge.

**MEMBRANE FOULING STUDIES**

Membrane fouling experiments were performed on laboratory scale by measuring the DI water flux before and after the water flux of HS-type activated sludge. This experiment was performed for the active membrane layer facing the feed and draw solution (Figures 11 and 12 respectively). For the FO membrane facing the feed side all water flux values were in the same range (5–6 L.m\(^{-2}\).h\(^{-1}\)) indicating the absence of membrane fouling during FO within 7–8 hours (Figure 11). For the FO membrane facing the draw side, the water flux of activated sludge (9 L.m\(^{-2}\).h\(^{-1}\)) was substantially lower than the DI water flux values before (11 L.m\(^{-2}\).h\(^{-1}\)) and after (approximately 10 L.m\(^{-2}\).h\(^{-1}\)) indicating the occurrence of partially reversible fouling (Figure 12). For both membrane orientations the water flux value of the DI experiment after the experiment with activated sludge was similar to the water flux value of the DI experiment before the activated sludge experiment. From this, it was concluded that no membrane
fouling occurred during FO of activated sludge facing the feed and draw side within 7–8 hours, possibly as a result of operating below critical flux conditions.

Additional XPS membrane surface characterization was carried out to investigate potential membrane surface fouling of the FO membranes. Carbon, nitrogen, oxygen and silica peaks were identified for virgin and fouled FO membranes after laboratory scale FO of HS-type activated sludge for 7–8 hours facing the feed side (Table 2). The FO membrane consisted of higher carbon peaks compared to the theoretical chemical structure of cellulose triacetate (also indicated by a higher C/O ratio). No difference in peak areas was observed between the virgin and fouled FO membrane, which confirmed an absence of membrane fouling. Contrary to these observations, respectively organic and silica fouling of FO membranes was observed by Mi and Elimelech (2008) and Lay et al. (2010). This fouling, however, was proven to be almost fully reversible under different laboratory conditions (Mi and Elimelech 2010). Similar results on the reversibility of membrane fouling of FO membranes were recently obtained by the same authors for gypsum scaling (Baoxia and Elimelech 2010).

On-site bench-scale experiments on HM-type activated sludge did not result in significant FO membrane fouling problems over a period of 2 weeks indicated by relatively stable water flux values during 2 weeks (Figure 13). Differences in water flux values were observed as a result of membrane orientation, resulting in higher flux values for the FO membrane facing the draw solution. An increase of the flux in time was subscribed to (i) fluctuations in the feed water composition, (ii) temperature effects and (iii) adjustments of the membrane in time (e.g. rinsing out preservation liquid).

### ECONOMIC CONSIDERATIONS

Capital cost estimations for the osmotic membrane bioreactor followed by reverse osmosis (OMBR-RO) compared to a conventional membrane bioreactor followed by reverse osmosis (MBR-RO) were carried out according to cost functions which are generally applied in the Netherlands (Wessels et al. 1998). These cost functions estimate the capital cost of the major equipment, such as civil, mechanical and electrical engineering costs excluding land costs, with an accuracy of 30%. Additional capital costs are assumed to be 34% of the major equipment cost, and includes auxiliary equipment, land costs, construction, management overheads, contingency costs, building interest, etc. Ray 2001; Zhou and Tol 2004; Diaz-Caneja et al. 2005. Amortized annual capital costs are calculated from Owen et al. (1995):

\[
C = \frac{i \cdot P}{1 - (1 + i)^{-n}}
\]

where C is the amortized capital cost, P the capital cost in the original year, i the interest rate and n amortization years. An interest rate of 7% and amortization times of 30, 15 and 10 years for respectively civil, mechanical and electrical engineering costs were assumed (Nederlof et al. 2000; Diaz-Caneja et al. 2005). The amortization time of the membranes

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**Table 2** | XPS results of virgin and fouled FO membrane samples with HS-type activated sludge for 7–8 hours

<table>
<thead>
<tr>
<th></th>
<th>C</th>
<th>N</th>
<th>O</th>
<th>Si</th>
<th>C/O</th>
</tr>
</thead>
<tbody>
<tr>
<td>Theoretical</td>
<td>60</td>
<td>-</td>
<td>40</td>
<td>-</td>
<td>1.5</td>
</tr>
<tr>
<td>Virgin</td>
<td>68.3</td>
<td>0.5</td>
<td>30.0</td>
<td>0.9</td>
<td>2.3</td>
</tr>
<tr>
<td>Fouled</td>
<td>66.5</td>
<td>0.5</td>
<td>32.5</td>
<td>0.4</td>
<td>2.0</td>
</tr>
</tbody>
</table>

*Cellulose triacetate: [C₆H₁₀O₂]₀OC(O)C₆H₃]*
depends on membrane replacement and was assumed to be 5 years. To compare OMBR-RO and MBR-RO, capital costs were calculated for a 3.5 Mm$^3$/year (400 m$^3$/h) waste water treatment plant, and a 1.0 ratio between maximum and average capacity of these plants was assumed.

In general, the operational costs include costs for energy consumption, labor, chemical cleaning and dosing chemical consumption (anti-scalant, sodium hypochloride, hydrogen peroxide), consumables (e.g. ultra filtration (UF), forward osmosis and reverse osmosis membranes) and spare parts (Ray 2001; Zhou and Tol 2004; Adham and Kumar 2005). In this work the following operational costs related to membrane cleaning will be considered; (i) chemical consumption cost, (ii) energy cost for heating up cleaning solutions, (iii) operational labor cost for chemical cleaning and (iv) membrane replacement cost. The following assumptions were made for the membrane cleaning costs estimates:

- energy cost of 0.065 €/kWh (Owen et al. 1995; Chellam et al. 1998);
- NF/RO membrane cost of 20 €/m$^2$ (Nederlof et al. 2000; Ray 2001);
- UF membrane cost of 60 €/m$^2$ (Nederlof et al. 2000);
- FO membrane cost of 30 €/m$^2$ or 60 €/m$^2$ (assumption)
- formulated cleaning product cost of 2.5 €/kg (Adham and Kumar 2005);
- commodity cleaning product cost of 0.2 €/kg (Adham and Kumar 2005);
- man-hour cost of 70 € for plant operators (Owen et al. 1995; Ray 2001; Adham and Kumar 2005);
- Disposal costs of chemical cleaning and concentrate are not included.

For the economical assessment, a MBR system was designed with an UF system operating at a flux of 20 L.m$^{-2}$.h$^{-1}$ at a trans membrane pressure of 0.1 bar. The subsequent RO system is operating at a flux of 15 L.m$^{-2}$.h$^{-1}$ at a trans membrane pressure of 14.5 bar with a recovery of 75% including an energy recovery device. The OMBR system was designed to operate at different water fluxes (10, 15 and 20 L.m$^{-2}$.h$^{-1}$ at 0.5 M), and the coupled RO system is operating at a flux of 30 L.m$^{-2}$.h$^{-1}$ at a trans membrane pressure of 31.7 bar with a recovery of 65% including an energy recovery device.

From the economic assessment, a 5–25% cost saving was calculated for reuse of wastewater with the OMBR with RO post-treatment compared to conventional MBR-RO. Cost savings predominantly depend on the FO membrane cost and on the FO membrane performance (Table 3). A minimum water flux values of 15 L.m$^{-2}$.h$^{-1}$ at 0.5 M is required to obtain a cost saving for OMBR-RO. This is not possible for existing FO membranes, and therefore the development of FO membranes with improved permeation and separation properties is essential for the OMBR-RO concept to be economically feasible.

**CONCLUSIONS**

- The OMBR seems a promising new development which facilitates the reuse of both industrial and municipal wastewater with a potential double barrier approach for pathogens, organic micro pollutants and particulate matter.
- FO performance (i.e. water flux) depended on the temperature as a result of viscosity effects of the feed solution, while the draw solution leakage depended on the temperature as a result of salt diffusion effects of the draw solution;
- FO performance was independent of the type of activated sludge with different compositions in COD and conductivity, while draw solution leakage varied for different types of activated sludge;
- FO performance on bench-scale increased in function of the osmotic driving force and depended on the membrane orientation;
- Higher flux values were obtained for FO membranes facing the draw solution side because of lower internal concentration polarization effects;
- External concentration polarization did not play a prominent role during bench-scale FO testing;
- No membrane fouling was observed during laboratory and on-site bench scale experiments, while partially reversible

**Table 3 | Relative cost benefit in cost of OMBR-RO with alternative (MBR-RO)**

<table>
<thead>
<tr>
<th>FO membrane price (€/m$^2$)</th>
<th>60</th>
<th>30</th>
</tr>
</thead>
<tbody>
<tr>
<td>FO-flux (L.m$^{-2}$.h$^{-1}$)</td>
<td>10</td>
<td>15</td>
</tr>
<tr>
<td>Energy cost @ 0.065 €/kWh</td>
<td>–20%</td>
<td>5%</td>
</tr>
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membrane fouling only occurred for the FO membrane facing the draw solution side;
- A minimum water flux value of 15 L.m\(^{-2}\).h\(^{-1}\) at 0.5 M was required to obtain a cost saving for OMBR-RO compared to MBR-RO, which will only be possible by improving existing FO membranes.

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