Techno-economic assessment of fertiliser drawn forward osmosis process for greenwall plants from urban wastewater

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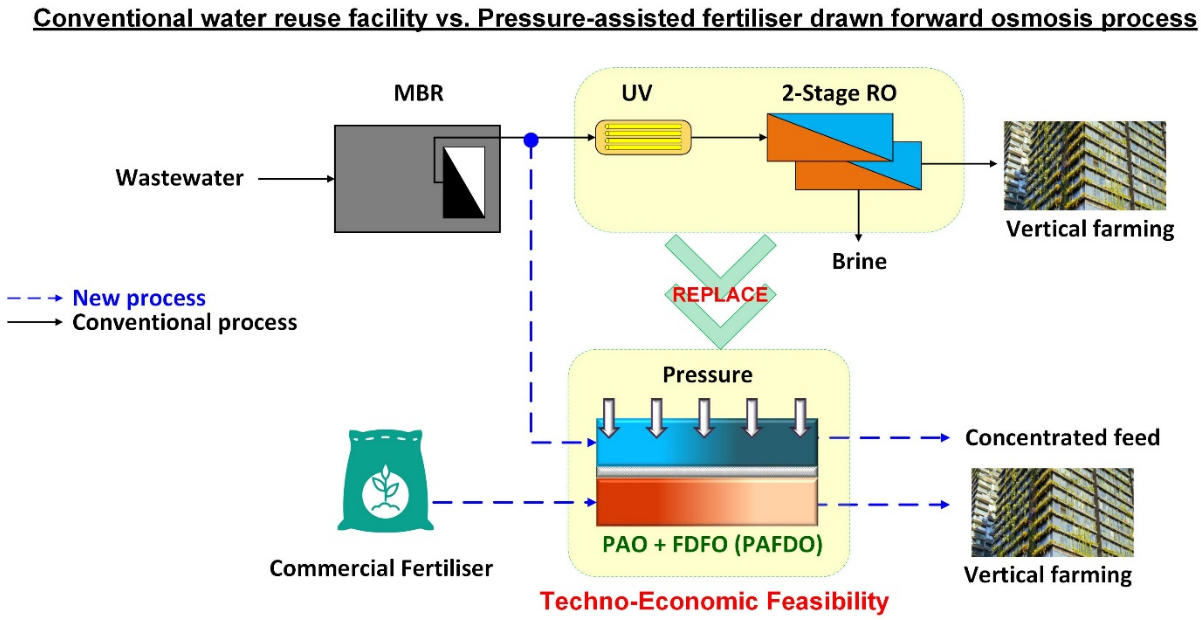
Highlights

- Fertiliser drawn forward osmosis (FDFO) for urban wastewater reuse was investigated
- Applying additional pressure in the FDFO was considered as an alternative
- Pressure applied at lower than a 60-fold fertiliser dilution was recommended
- Water flux of 10 $Lm^{-2}h^{-1}$ was required to make the FDFO economically feasible
- Pressure-assisted FDFO could be competitive with the existing water reuse facility

Graphical abstract



Abstract

Pressure-assisted osmosis (PAO) has been suggested to integrate with fertiliser driven forward osmosis (FDFO) to improve the overall efficiency of simultaneous wastewater reuse and fertiliser osmotic dilution. This study aims to demonstrate the techno-economic feasibility of pressure-assisted fertiliser driven forward osmosis (PAFDO) hybrid system compared to the existing ultraviolet and reverse osmosis (UV-RO) process. The results showed that coupling FDFO with PAO (i.e. PAFDO) could help fulfill the water quality required for greenwall fertigation. An economic analysis on capital and operational costs for the PAFDO showed that the PAO mode application at a lower FDFO dilution stage could significantly reduce the costs. However, when considering the different applied pressures in PAO (i.e. 2, 4, and 6 bar), the increase in the total water cost was not significant. This indicates that the dilution stage for applying PAO is more sensitive to the total water cost of the PAFDO than the applied pressure. A coupling of higher average water flux (>10 L/m^2h) and lower draw solution (DS) dilution factor (DF<60) is recommended. Therefore, this could make the PAFDO system economically viable compared to the benchmark for the UV-RO disinfection system.

Keywords

Forward osmosis; Fertigation; Pressure assisted osmosis; Wastewater reuse; Techno-economic assessment.

1 Introduction

The world population is projected to cross 9 billion by 2050 (Diaz et al. 2017). The rapid population growth coupled with climate change and urbanization have placed an increasing demand for limited potable water resources throughout the world. As the agricultural sector accounts for around 70 % of the world freshwater consumption (Wisser et al. 2008), food production may therefore soon be hindered by water availability. To guarantee food and water security, robust and sustainable methods to supply clean water are increasingly needed while mitigating the impact on the environment (Zhang et al. 2017, Zhang et al. 2018).

To date, reverse osmosis (RO) process has worldwide attention in both wastewater reclamation and desalination mainly due to the development of good performance membranes and its lower environmental impact compared to the thermal technologies (Al-Obaidi et al. 2017, Bunani et al. 2015). A recent study investigated the performances of two different types of RO membranes in removal of various dissolved species in secondary effluent stream and showed that the quality of the RO permeates is suitable for agricultural irrigation (Bunani et al. 2015). In addition, the use of reclaimed water produced from an integrating RO system consisting of RO-ultraviolet (UV) or UV-RO was demonstrated for irrigation or non-potable applications (Kargari and Mohammadi 2015, Ordóñez et al. 2011, Von Gottberg 2005). Although such RO integrated system can produce high-quality water for reuse, this still leads to high operational costs as the pressurised system requires more pumping costs and cleaning operations (Chekli et al. 2016).

One of the most promising technologies is fertiliser drawn forward osmosis (FDFO) process, which has recently gained global attention. In the concept of the FDFO process, when a highly concentrated fertiliser solution (i.e. draw solution, DS) and a low saline water (i.e. feed solution, FS) are separated by a selectively permeable membrane, this allows passage of fresh water from the FS to the DS by osmotic concentration differential. The diluted fertiliser, containing fertiliser nutrients, can thus be directly used for irrigation of crops. The concentration of the fertiliser after dilution in the FDFO process must be acceptable for direct application and this has however been found to be challenging (Phuntsho et al. 2013 a). The final diluted fertiliser produced from the FDFO process is limited by the feed stream concentration (i.e. osmotic pressure) based on the osmotic equilibrium between the feed and draw streams (Phuntsho et al. 2014). When feed water solution with a high content of salt is used for such application, the final product at osmotic pressure equilibrium could have much higher concentration of nutrients than allowable levels for irrigation. Reclaimed water with the relatively low salinity can be good candidates for enhancing water flux (i.e. dilution effect).

FDFO can be used as a stand-alone process or coupled with a post-treatment process such as RO and nanofiltration (NF) for draw solution recovery and water purification. In the latter case, the post-treatment process provides further purification of the product water. For example, Phuntsho et al. (Phuntsho et al. 2013 a) demonstrated that NF as post-treatment was found to be more effective in reducing the nutrient concentrations in the final product. Including this, different approaches were proposed in our previous investigation to mitigate the nutrient concentrations in the final product fertiliser such as mixing with fresh water at the final stage (Phuntsho et al. 2012), hybrid FO process for treating wastewater treatment (Phuntsho et al. 2012) and applying additional pressure on the feed side during the FO operation (Sahebi et al. 2015).

The pressurization of the FS of the FDFO process can offer a range of potential benefits over the limitations of the stand-alone FDFO system such as low water flux and high reverse salt flux. The pressure assisted FDFO system can thus take an advantage of the synergetic effect of the driving force to improve the permeate flux and thus further dilution of the DS. Chekli et al. (Chekli et al. 2017) recently reported that integrating FDFO with pressure assisted osmosis (PAO) could provide an insight into an opportunity for a cost-effective FDFO process and also assessed the applicability of the FDFO process to yield an irrigation solution for a hydroponic grow system, which is a widely applied technique for growing plants in the water/fertiliser solution. However, no study has directly evaluated the techno-economic feasibility of FDFO in the reuse of real wastewater effluent to a desirable quality for greenwall plants growth (i.e. vertical farming).

This work examined the techno-economic feasibility of the FDFO process for irrigation to greenwall plants. This includes short and long-term operations of the FDFO process with real urban wastewater of different qualities (i.e. primary and secondary effluents) as a feed solution candidate and commercial fertiliser as a draw solution to demonstrate its technical feasibility. The effect of a hydraulic pressure on the FS (i.e. pressure assisted osmosis, PAO) was also evaluated as an alternative way of reducing the final diluted fertiliser concentration. In addition, the economic performance of the pressure assisted FDFO (PAFDO) process was delineated to provide a better understanding of the applicability of this technology and its implications to make it economically feasible.

2 Materials and methods

2.1 Commercial fertiliser draw solution and real waste feed water

The commercial fertiliser diamond blue (denoted as DB) used in this study as DS was obtained from Campbells Fertilisers Australasia. The fertiliser DS was prepared to obtain 175 g/L as total dissolved solids (TDS, pre-filtered with 0.45 μm filters), corresponding to an osmotic pressure of 92.48 bar. The osmotic pressure of DB fertiliser was calculated using the thermodynamic modelling software OLI Stream Analyser (OLI Systems Inc., USA). Table 1 shows the characteristics of the DB. The bench-scale experiments were conducted using deionized water (DI water) as FS to elucidate the performance of the DB as DS in the FDFO process.

The wastewater streams used in this study as FS were collected from the Central Park Wastewater Treatment Plant (CPWTP) in Sydney, New South Wales, Australia. The CPWTP consists of a screen mesh, a membrane bioreactor (MBR) followed by an ultraviolet (UV) disinfection unit, RO system and chlorine contact, before finally being stored in the treated water storage tank for reuse. Three types of wastewater streams with different qualities were evaluated for their performances as a FS candidate in the FDFO process: raw wastewater, MBR supernatant and MBR effluent. The characteristics of the wastewater streams are presented in Table 2.

Table 1. Characteristics of the commercial fertiliser diamond blue.

Parameters	Diamond Blue Fertiliser
Electrical conductivity (EC, µS/cm)	150.4
Total dissolved solids (TDS, g/L)	175.0
pH	3.91
Total organic carbon (TOC, mg/L)	1,102.5
Osmotic pressure (bar)	92.48
Total Nitrogen (TN, mg/L)	72,000
NO_3^--N (mg/L)	22,700
NH_4^+-N (mg/L)	16,900
Total Phosphorus (TP, mg/L)	7,300
SO_4^{2-} (mg/L)	17,000
K^+ (mg/L)	26,440
Na^+ ((mg/L)	3,090
Mg^{2+} (mg/L)	3,860
Ca^{2+} (mg/L)	470

Table 2. Central Park Wastewater characteristics used in this study as FS.

Parameters	Raw wastewater*	MBR supernatant *	MBR effluent
EC (µS/cm)	1299.0	820.0	759.0
pH	7.90	7.38	7.50
TDS (mg/L)	646.0	357.0	336.0
Turbidity (NTU)	63.6	1.30	0.39
Osmotic pressure (bar)	0.307	0.226	0.194
NO_2^- (mg/L)	0.19	0.08	0.10
NO_3^--N (mg/L)	0.20	2.90	3.30
NH_4^+-N (mg/L)	65.6	1.30	1.90
TP (mg/L)	22.4	7.0	3.0
SO_4^{2-} (mg/L)	38.0	38.0	45.1
K^+ (mg/L)	25.4	20.45	18.24
Na^+ ((mg/L)	142.4	122.5	113.0
Mg^{2+} (mg/L)	9.25	8.0	5.56
Cl^- (mg/L)	72.0	67.0	21.54

* Pre-treated only by sedimentation to collect the supernatant.

2.2 Forward osmosis experimental procedure

A bench-scale crossflow FO experimental process is shown schematically in Fig. 1. Low-pressure variable speed gear pumps (Cole Palmer, USA) were installed to circulate the feed and draw streams. Each pump was connected to a membrane cell (2.6 cm width, 7.7 cm length, and 0.3 cm depth). A thin film composite (TFC) FO membrane supplied by Toray Industry Inc, which is made of a polyamide active layer deposited on a polysulfone support layer, was used for all FO experiments. The intrinsic properties of the TFC membranes are pure water permeability (A) of 2.47 $Lm^{-2}h^{-1}bar^{-1}$ and NaCl rejection (B) of 96%. The temperatures of the feed and draw solutions were maintained at 25±0.5 ºC using a heater/chiller system. The cross-flow velocity for the feed and draw streams was circulated at 10.68 cm/s in counter-current configuration. The water flux was determined by changing the weight of the draw tank collected on the digital weighing scale connected to a computer for the data recorder. During all FO experimental work, both FS and DS were recirculated back to their respective reservoirs. The initial volume of FS and DS solutions was 1 L for each short-term experiment that lasted for 5 hr. The long-term operations were conducted for 5 days with the initial volume of FS and DS of 5 L and 200 mL, respectively.

The reverse salt flux (RSF) was investigated either by measuring the electrical conductivity (EC) using a multi-meter (Hach, Germany) or by analyzing the major anions and cations using inductively coupled plasma mass spectrometry (ICP-MS) and microwave plasma-atomic emission spectroscopy (MP-AES). Nutrient concentrations were also evaluated with a spectrophotometer (Spectroquant NOVA 60; Merck, Germany). A total organic content analyser (TOC analyser, Analytikjena, Jena, Germany) was used to measure the TOC of the FS and DS. The turbidity of wastewater was measured with a 2100P Portable Turbidimeter (Hach, USA).

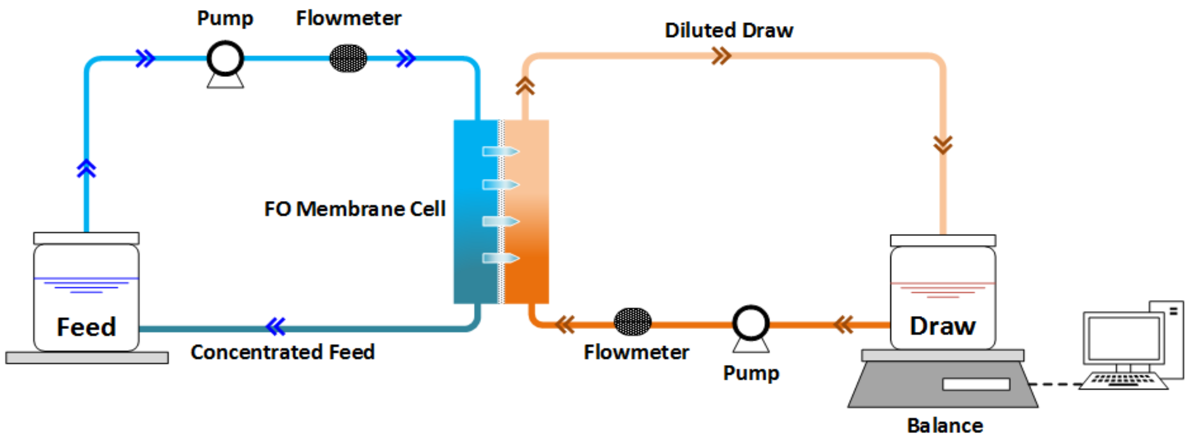


Fig. 1. A bench-scale FO experimental set-up and flow direction.

2.3 Determination of water flux, reverse salt flux and specific reverse salt flux

Water flux (J_w, $Lm^{-2}h^{-1}$, LMH) across the membrane was calculated automatically every 5 minute from the increase in DS weight recorded by a digital balance, on which the DS tank was placed, and connected to PC for real-time data collection.

$$J_w = \Delta W_D / (S_m * \Delta t) \quad (1)$$

where ΔW_D is the weight change of the draw, S_m the effective membrane surface (m^2) and Δt the time interval (hr).

The dilution factor *DF* (i.e. how many times the DS is diluted) of the DS is determined by the following equation:

$$DF = (V_{D,i} + \Delta V_D)/V_{D,i} \quad (2)$$

where $V_{D,i}$ (L) is the initial DS volume and ΔV_D (L) is the increase in draw volume over time.

RSF represents the amount of draw solutes that pass across the membrane to the feed side in a unit membrane area and in a unit operating time. RSF J_s ($gm^{-2}h^{-1}$) was calculated by monitoring the increase in the electrical conductivity of the FS using DI water as feed with a conductivity and pH meter (HACH, Germany) connected to the computer for data logging.

$$J_s = (F * \Delta(EC_F * V_F))/(S_m * \Delta t) \quad (3)$$

where F is a conductivity calibration factor for the conversion between conductivity and concentration and $\Delta(EC_F*V_F)$ the feed conductivity differential per feed volume change (µS*L/cm). It should be noted here that 'F' is valid for DI water as feed water. When the feed water contains various components, the detailed characteristics of the feed water need to be conducted. The amount of fertiliser lost per volume of water that permeates the membrane is represented by the specific reverse solute flux (SRSF) (g/L) calculated dividing J_s by J_w.

2.4 Economic analysis

The scope of this study was to conduct an economic feasibility study that compares an existing UV-RO disinfection system with a FDFO process in terms of the capital and operational expenses in Australian dollar (i.e. CAPEX and OPEX in AUD) by considering different FDFO process configurations. Several assumptions have been considered as follow:

- The CAPEX cost includes the cost of FO and RO modules in the FDFO and UV-RO systems respectively. Other components such as pipeline, pumps and valves are not included in this study due to its minor contribution to the total cost (Zhou et al. 2014). The

FO element cost was assumed to be the same as the RO element cost of $700 with a lifetime of 3 years for RO and 7 years for FO. It has to be noted here that the lifetime of the FO element was assumed longer than that of the RO element since fouling propensity and cleaning frequency of RO are more significant than FO due to the use of high pressure (Lee et al. 2010). A unit cost of $1,000 for one pressure vessel was assumed for both FO and RO processes.

- The annualized CAPEX cost ($/yr, $CAPEX_a$) was determined at an interest rate of 6% (i.e. i) and a plant availability of 0.95 for a 20-year plant lifetime (i.e. n) (Kim et al. 2018). The $CAPEX_a$ cost in $/yr is therefore calculated based on the following equation:

$$CAPEX_a = Total\ CAPEX\ cost \frac{i(1+i)^n}{(1+i)^n - 1}$$

- The annual OPEX cost ($/yr) comprises the energy consumption, membrane maintenance, and chemical consumption costs. The annual energy cost was estimated at an electricity cost of 0.29 $/kWh (Kim et al. 2017).
- Based on the real capacity of the Central Wastewater Treatment Plant, all configurations were set at 400 m^3/day of product water.
- In the case of PAO application, the applied pressure on the feed side was assumed to be 2 bar because higher operating pressure could result in additional costs and more severe fouling (Blandin et al. 2015, Kook et al. 2018). However, for a sensitivity analysis, the applied pressure was assumed to be varied from 2 to 6 bar.
- The total cost ($/yr) is the sum of $CAPEX_a$ and OPEX costs. The total product cost ($/$m^3$) can be therefore calculated from the following equation:

$$Total\ product\ cost\ (\$/m^3) = \frac{Total\ cost\ \left(\frac{\$}{yr}\right) \times Plant\ lifetime\ (yr)}{Plant\ capacity\ \left(\frac{m^3}{day}\right) \times 365 \times Plant\ availability}$$

2.5 Process description

2.5.1 Conventional recycled water plant

In Central Park WTP, the primary water (i.e. after screen process and biological processes) is first filtered through two membrane tanks and both contain one GE 500 membrane cassette with a surface area of 1099.8 m^2 followed by UV disinfection unit and RO. Two UV disinfection units provide disinfection of the filtered water. The setpoint of UV dosage rate in the system is 250 J/m^2. The RO system operates as a two stage-one pass process, which comprises six RO vessels containing four spiral wound membrane elements (BW30-400, Dow Filmtech Chemicals, USA) in each pressure vessel. The RO system is designed based on a design recovery of 80%, to produce 400 m^3/day of permeate. In the first stage, two sets of two RO vessels operate in parallel while two pressure vessels operate in series in the second stage. A booster pump between the two stages is required to compensate for osmotic pressure increase. Based on the current plant design, the RO system simulation and cost analysis to produce 400 m^3/day were conducted using WAVE simulation software (Water Application Value Engine, Dow Filmtech Chemicals, USA). A schematic diagram of the conventional WTP and flow directions is presented in Fig. 2. Calculations of the CAPEX and OPEX for UV-RO system are based on the percentage contribution to the total annual cost adapted from the literature (Holloway et al. 2016) and the results achieved from the WAVE analysis.

2.5.2 FDFO system

A full-scale FDFO system was simulated using lab-scale FDFO experimental results, equations developed by Deshmukh et al. (Deshmukh et al. 2015) and a mass balance relationship in the process. The feed water for the FDFO process was considered an MBR effluent with an osmotic pressure of 0.2 bar as shown in Table 2, determined by the results obtained from the lab-scale FDFO experiments. As mentioned above, the draw solution was commercial fertiliser with TDS of 175 g/L and an osmotic pressure of 92.48 bar (Table 1). The FO system was designed to produce 400 m^3/day of the product permeate. Membrane fouling was indirectly considered through physical cleaning and membrane replacement intervals. A schematic diagram of a hybrid PAO and FDFO process and flow directions is presented in Fig. 2.

The reliability of economic impact assessment is highly dependent on the selected background data due to a large number of input parameters including permeation flux, total membrane area required and membrane element and pressure vessel costs. (Kim et al. 2018). Among the various input parameters, the average permeation fluxes play a crucial role in economic feasibility of FO (Blandin et al. 2015, Kim et al. 2017). In addition, operating in PAO mode causes the advanced driving force thus resulting in a significant saving in total water product cost even with increased energy requirements (Blandin et al. 2015, Chekli et al. 2017, Sahebi et al. 2015). However, it is important to determine a certain dilution stage to apply PAO mode in the FDFO process because this can influence the economic viability of the PAFDO system. Therefore, a sensitivity analysis was carried out based on two different approaches: (i) PAO mode in different DS dilution stages in the FDFO process and (ii) PAO mode with three different applied pressures (2, 4, and 6 bar). The results were finally compared with the conventional wastewater reuse plant to evaluate its economic feasibility.

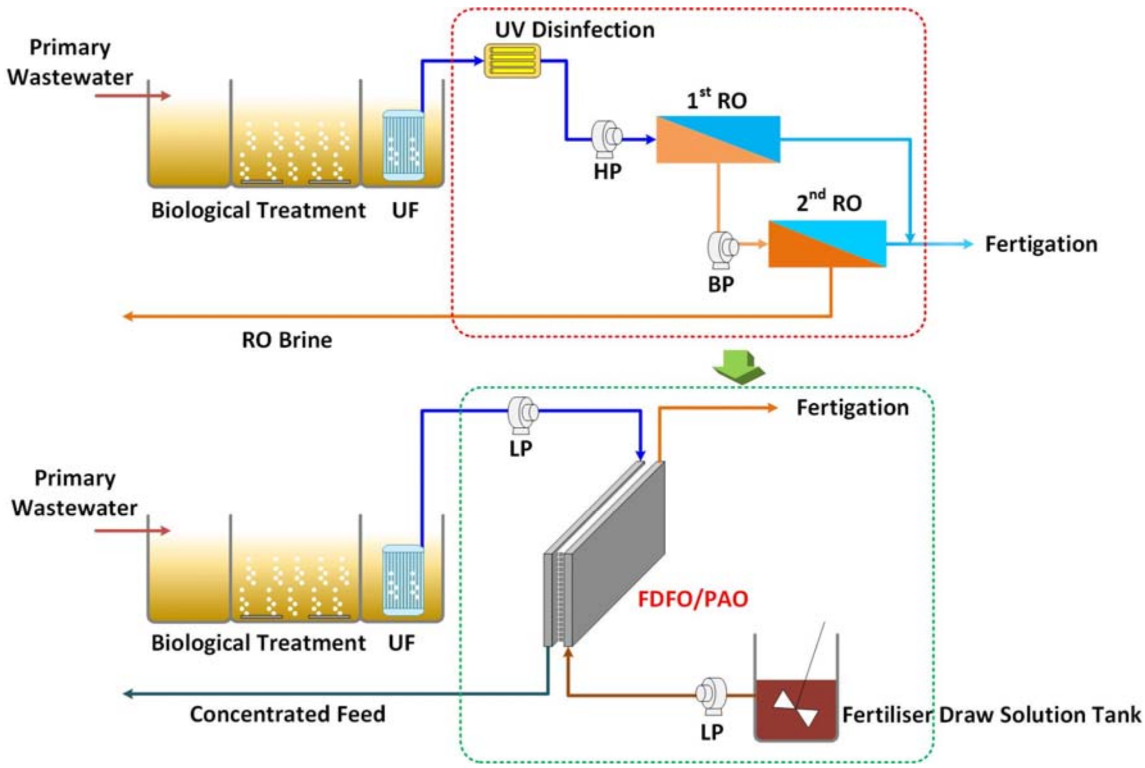


Fig. 2. Schematic diagram of MBR-UV-RO (Central Park Wastewater Treatment Plant) and MBR-PAFDO considered in this study. Plant capacity: 400 m^3/day. HP: High pressure, BP: booster pump, LP: low pressure pump.

3 Results and discussion

3.1 FDFO performance evaluation

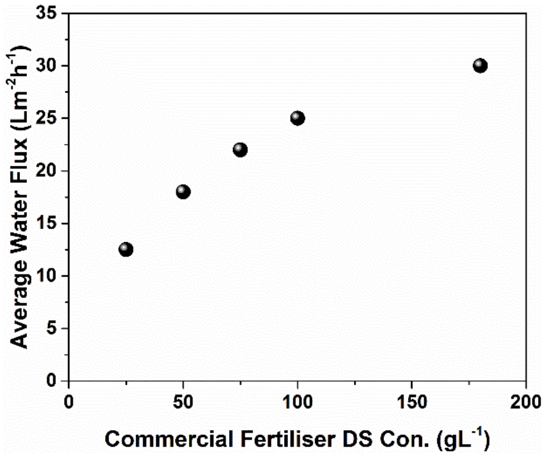
3.1.1 Short-term FDFO operation with DI water as FS

The short-term FO tests were conducted to demonstrate the ability of the DB fertiliser DS using DI as FS. The average water flux as a function of the concentration of the fertiliser DS is presented in Fig. 3 (a). Results showed that the water flux increased non-linearly with the increase in the DS concentrations, which is similar to our earlier study with ammonium sulphate fertiliser (Sahebi et al. 2015). This non-linear correlation between water flux and DS concentration (i.e. osmotic driving force) can be attributed to the severity of dilutive internal concentration polarisation (ICP) that significantly reduces the effective osmotic pressure difference across the FO membrane (Cath et al. 2006).

In addition, the RSF (J_s) for the commercial fertiliser was determined as explained in Section 2.3. The ratio of RSF to water flux (J_s/J_w, SRSF) was therefore found for the major nutrients in the DB fertiliser and shown in Fig. 3 (b). The result shows that PO_4^{3-}, Mg^{2+} and SO_4^{2-} had SRSF comparatively much lower than K^+, NO_3^- and NH_4^+, which can be explained by their larger hydrated radius and thus lower reverse diffusion toward the FS. In fact, monovalent ions are more subjected to reverse permeation than multivalent ions.

Moreover, larger-sized hydrated anions such as PO_4^{3-} and SO_4^{2-} diffuse less across the semi-permeable membrane because of electrostatic repulsion forces (less than 0.1 g/L). The difference in SRSF between potassium (K^+) and ammonia (NH_4^+) can be explained by their concentration in the initial fertiliser DS. Lower solute concentration in the DS can result in lower SRSF and vice versa. As shown in Table 1, the concentration of ammonia (i.e. 16.9 g/L) in the commercial DB fertiliser was lower than the one of potassium (i.e. 26.44 g/L). Consequently, the reverse permeation of NH_4^+ was lower than for K^+. The ratio J_s/J_w plays an important role in determining the draw solute loss during FO operation. This is directly related to the draw solute replenishment cost. The recent study conducted by Chekli et al. (Chekli et al. 2017) demonstrated that the enhanced water permeability reduces the RSF across the membrane. For instance, under PAO mode (2 bar applied pressure), the reverse diffusion of NH_4^+ was reduced by 80% and that of K^+ was reduced by more than 90%. The result clearly showed that the RSF can be reduced by integrating PAO in the FDFO process. Therefore, in this study, the DS replenishment cost was not considered for economic evaluation of the PAFDO process as it is a minor contributor to the total cost of the process.

(a)



(b)

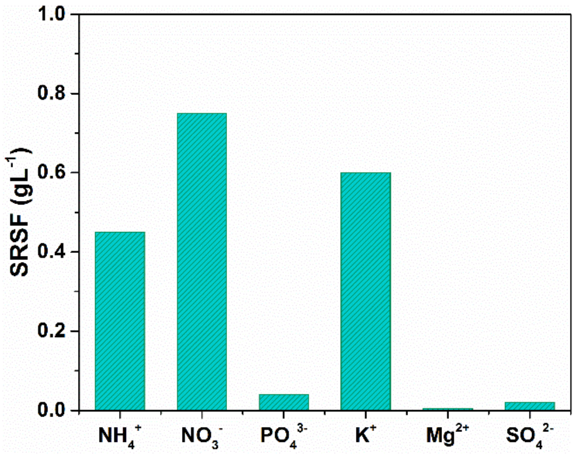


Fig. 3. (a) Experimental water flux of each concentration of commercial DB fertiliser (25, 50, 75, 100, and 175 g/L) and (b) specific reverse solute flux of the commercial DB fertiliser of 175 g/L. Feed and draw flow rate: 500 mL/min. The temperature of feed and draw sides: 25 °C. Feed: DI water.

3.1.2 Long-term FDFO operation with different wastewater streams as FS

The key parameter for FDFO technology is the draw dilution factor which can be calculated based on Equation (2) presented in Section 2.3. The principle of the osmotic equilibrium between the FS and the DS limits this pivotal factor during the FO process since the initial FS concentration governs the final DS concentration (i.e. final osmotic equilibrium), which shows direct implications for the end use of the final product. A series of long-term FDFO experiments were carried out to identify the optimal FS with the goal of diluting as much as possible the DS (i.e. maximum dilution factor).

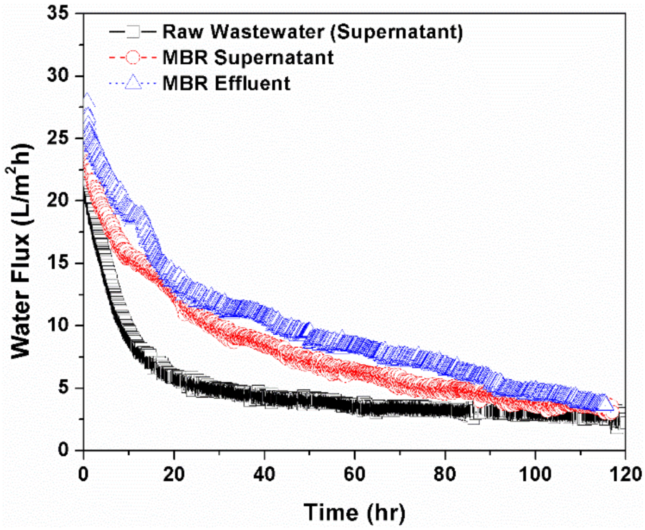
Fig. 4 shows the variations of the water flux and dilution factor of the fertiliser DS when different wastewater streams are used in the FDFO process. Fig. 4 (a) shows that the water flux with MBR supernatant and effluent appeared quite similar for operation time, demonstrating a consistent performance of the FDFO process under each long-term test. This can be seen that the flux decline with MBR supernatant and MBR effluent mainly caused by DS dilution effect rather than membrane fouling.

However, the water flux with raw wastewater is considerably lower than that with other streams (i.e. the sharper flux decline). Such flux decline with raw wastewater in Fig. 4 (a) was expected since the feed water used for the FDFO process had higher turbidity of 63.6 NTU (Table 2) and much lower dilution factor compared to the others. It is important to note that although MBR effluent and supernatant showed lower turbidity, organic compounds in the feed streams may cause severe fouling, thus resulting in flux decline. Meanwhile, there was a sharp flux decline in the first 10 h and this is mainly attributed to the effect of DS dilution while after 10 h operation, the flux decline with raw wastewater was more severe than others. This indicates the occurrence of fouling caused by high turbidity in the feed water at the initial stage of the operation.

In addition, Fig. 4 (b) presents the dilution factor with the operation, indicating the maximum dilution factor achieved during the same operation time. Corresponding to the water flux trend, when raw wastewater is used as FS, it shows the lowest DS dilution factor (DF 7.07) and followed by the MBR supernatant and MBR effluent (11.20 and 13.11, respectively). In general, a higher dilution factor is expected to have a higher flux decline; however, the performance using raw wastewater shows the lowest dilution factor but the highest flux decline among three feed solutions (Fig. 4). This indicates that the flux decline with raw wastewater was mainly due to fouling occurrence on the membrane surface rather than the dilution effect. The lowest diluted DS concentration (i.e. the highest dilution effect) can be achieved when the MBR effluent is used as FS and thus showing that the MBR effluent is the best FS candidate for FDFO application.

A recent study by Sahebe et al. (Sahebi et al. 2015) proved the application of PAO in the FDFO process to improve the dilution of the fertiliser DS. Results from this investigation indicated that the PAO application can provide further dilution of the final fertiliser DS due to increased water flux thereby achieving the fertigation standard. Consequently, PAO could potentially remove the need for additional treatment.

(a)



(b)

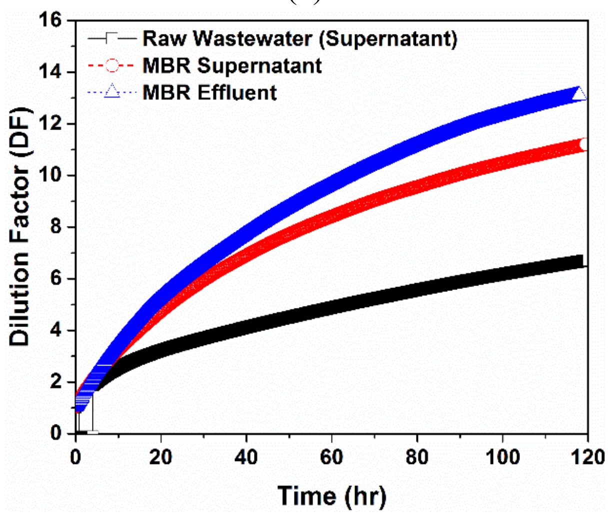


Fig. 4. The effect of wastewater quality (feed solution) on the water flux (a) and the dilution factor (b) as a function of the operation time. The initial FS and DS volumes were 5 L and 200 mL, respectively. Operation time: 5 days.

3.1.3 Suitability of the final FDFO nutrient solution for direct fertigation of greenwall plants

Based on the results of the short- and long-term experiments with different FSs (Section 3.1.1 and 3.1.2), the final long-term FDFO experiment was conducted using the MBR effluent as FS and DB fertiliser with an initial concentration of 175 g/L as DS. When the MBR effluent was used as the FS, it was observed that the fouling effect on the flux decline was not significant compared to the DS dilution effect. As mentioned earlier in Section 2.4, the final permeate water quality should be less than 1 g/L. Thus, it is obvious that the commercial fertiliser solution needs to be diluted 175 times to fulfill the requirement. The results in terms of water permeation and dilution factor during the operation of FDFO process are presented in Fig. 5. The operational parameters and the schematic diagram of the experimental procedure for this long-term FDFO operation can be found in Figure S1 in the supplementary information (SI).

The osmotic pressure difference between the DS and the FS became lower and lower which decreased the driving force across the FDFO process. It is worth noting that after 8-day operation, the water flux was almost zero meaning that the draw solution could not be further diluted because the osmotic equilibrium between the FS and the DS occurred (Phuntsho et al. 2014). At the end of the operation, the final DS concentration was reached to 3.83 g/L (total dilution of around 84.41). This results in the final fertiliser solution that contain insufficient nutrients for irrigation purpose (i.e. less than 1 g/L total dissolved solids).

In this study, PAO has been therefore considered as an integrated process to FDFO (referred to as PAFDO). Operating the PAO mode in FDFO can provide a trade-off between savings in total membrane area required (i.e. CAPEX cost) and the increased energy consumption (i.e. OPEX cost). Blandin et al. (Blandin et al. 2015) already reported the effect of hydraulic pressure on reducing the total membrane area. The results also showed that at the same recovery rate the required membrane area could be significantly reduced by increasing the applied pressure and thus savings in CAPEX cost. Recently, Kook et al. (Kook et al. 2018) investigated the optimum operating condition of PAO for PAO-RO hybrid system at a pilot scale level and its economic potential for wastewater purification and seawater dilution. From the practical aspect, the results showed that the PAO-RO hybrid system can be economically favourable if two FO membrane elements are connected in series in a housing. So, this is the first study to investigate an optimum point to apply PAO in the fertiliser driven FO application and its economic effect. The following section will discuss the optimum PAFDO process configuration for osmotic dilution of fertiliser draw solution using wastewater to reach economic viability.

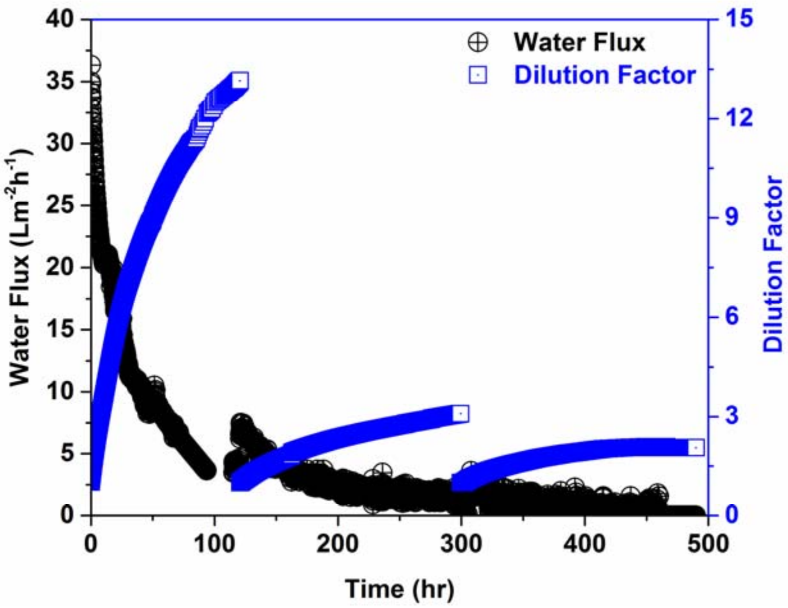


Fig. 5. Water flux and dilution factor with operating time. Experimental conditions: Commercial DB fertiliser as DS and MBR effluent as FS, initial FS and DS volumes were 15 L and 0.2 L respectively, and operation time: 500 hr.

3.2 Economic evaluation

3.2.1 MBR-UV-RO disinfection system

Fig. 6 (a) displays that the UV-RO accounts for 40% of the total energy consumption of the plant with 25% from RO process and 15% from the UV disinfection step. Fig. 6 (b) shows the RO and UV systems made up the largest proportion for all three costs. Compared to the MBR, the RO and UV systems had a 28% higher OPEX and a 10% higher CAPEX. Consequently, the contribution of the RO and UV units to the total cost of the plant was significant, around 60%.

For OPEX, the RO cost (\$0.321/m^3) was significant, followed by MBR and UV, \$0.277/m^3 and \$0.167/m^3 respectively. For CAPEX, the MBR cost (\$0.302/m^3) was highest and followed by RO and UV disinfection units (\$0.242/m^3 and \$0.130/m^3 respectively). As expected, the major factors responsible for such high OPEX cost of the plant are the energy consumption of the RO and UV units (Fig. 6 (a)). From the total water cost (\$1.439/m^3), the most economically feasible scenario would be ultimately made by reducing the operating cost of RO and UV units (\$0.86/m^3) under given plant operating conditions. Hence, one way of reducing the operating cost of the plant is to replace UV-RO unit with a low-energy technology, like FDFO. Several FDFO configurations have been therefore proposed and evaluated in the following section.

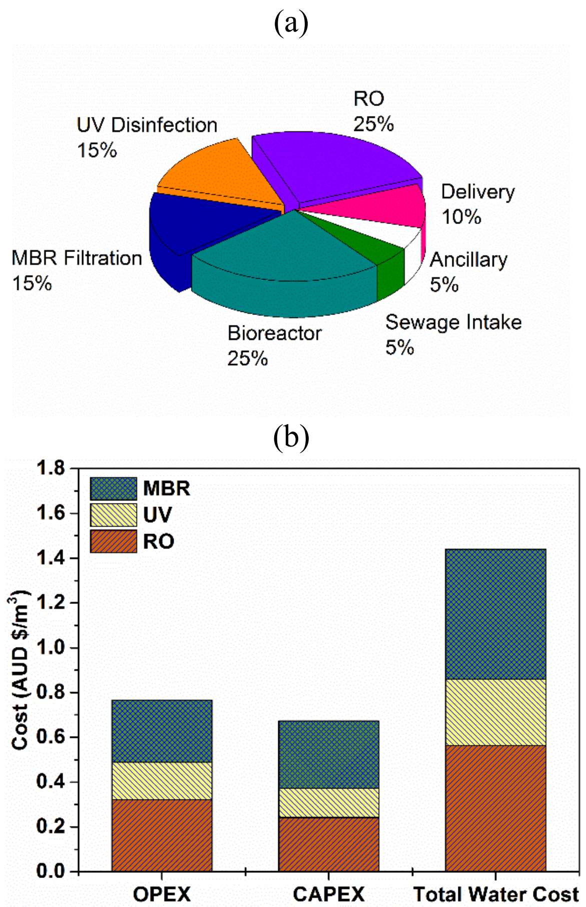


Fig. 6. (a) Energy consumption breakdown in Central Park wastewater treatment plant and (b) CAPEX and OPEX costs contributed to the total water cost of the plant.

3.2.2 FDFO system: sensitivity analysis

A sensitivity analysis is conducted on two approaches that can be used to make the PAFDO process economically favourable as mentioned above in Section 2.5.2; (i) PAO mode in different DS dilution stages in the FDFO process and (ii) PAO mode with three different applied pressures (2, 4, and 6 bar).

The benchmark for the UV-RO system is $0.86/m^3 and thus the cost of PAFDO should be lower than that of UV-RO. Fig. 7 shows the estimation of capital and operational costs for PAFDO as a function of FDFO dilution factor based on the results of the long-term experiments (as shown in Fig. 5). In the figure, "FDFO" refers to a FDFO stand-alone system while

"PAFDO (DF)" refers to a system in which PAO is integrated at a certain DS dilution factor of 5, 10, 20, 50 and 80. These values correspond to the FO water flux of 23.04, 14.89, 11.34, 5.57, and 4.37 LMH, respectively. It was also assumed that once the PAO mode is applied at a certain point in FDFO, the DS is continuously diluted to reach the final DS concentration of less than 1 g/L TDS, irrigation standard (Bauder et al. 2011).

Fig. 7 clearly shows that the FDFO stand-alone process had a 23% lower total water cost compared to the UV-RO process. When PAO was applied at different dilution factors in the FDFO process, the total water cost of the PAFDO was shown to be lower than that of the FDFO stand-alone, except when applying PAO at the dilution factor of 80. For example, when PAO is applied at a 5-fold DS dilution stage in the FDFO, the cost of PAFDO showed the lowest cost at \$0.293/m^3 among PAFDO configurations. However, with increasing the FDFO dilution factor from 5 to 80, the total cost of the PAFDO system was significantly increased to \$0.920/m^3, which is 6.5% higher than the UV-RO. This increase is mostly because the FDFO average water flux in the FDFO process became lower as increasing the dilution of the fertiliser DS and thus increasing the total membrane area required (Blandin et al. 2015, Phuntsho et al. 2014). This result indicates that in order to make the PAFDO process economically favorable compared to the UV-RO (\$0.86/m^3), the PAO mode should occur before fertiliser DS dilution reaches to 80.

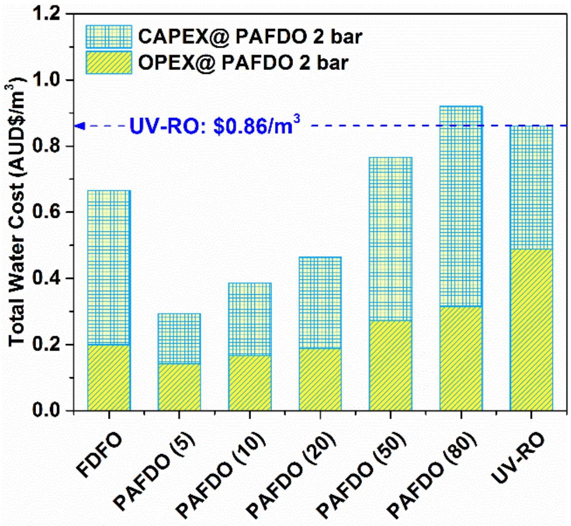


Fig. 7. Total water cost per m^3 of water produced for FDFO stand-alone and PAFDO with different FDFO recovery rates and for the existing UV-RO. PAFDO (DF) refers to PAO application at different DS fertiliser dilution factors with 2 bar applied.

From the results presented in Fig. 7, coupling FDFO with PAO is confirmed to be a promising strategy to overcome current limitations of FDFO, and could help reduce the total water cost even though additional energy is required for feed pressurization. However, in such a hybrid system, an optimum trade-off between the total membrane area (i.e. CAPEX and OPEX costs) and the additional energy (i.e. OPEX cost) is important for a practical application of the PAFDO process (Blandin et al. 2015, Sahebi et al. 2015).

In order to determine the significance of applying PAO mode of FDFO operation and the FDFO average water permeation in relation to the total product cost for the PAFDO system, total water cost calculation for FDFO stand-alone and PAFDO was made over the DS dilution factor from 0 to 80 as shown in Fig. 8. Detailed calculations on the CAPEX and OPEX can be found in Figure S2 in the SI.

Fig. 8 (a) shows a strong response of the average permeation fluxes to the total water cost as the driving force inevitably decreased due to the dilution of the fertiliser and the concentration of the wastewater stream. The FDFO cost is also seen to highly influence the total water cost of the PAFDO. This indicates that requiring the higher DS dilution leads to a significant cost increase of the FDFO process and thus it is essential to take into consideration a certain dilution factor for applying PAO mode of the FDFO process (i.e. PAFDO). For instance, when the DS was diluted 20-fold in the FDFO process, the water cost increases to 61% (from \$0.107/m^3 to \$0.273/m^3). This is because an FDFO average water flux decreased from 34 L/m^2h to 11.3 L/m^2h (67% decrease). With further dilution of DS fertiliser (i.e. a 40-fold dilution), the cost of the FDFO further increased to \$0.503/m^3 considering the average water flux of 5.9 L/m^2h.

Compared to the FDFO, although the PAO cost increases with increasing the dilution stages in the FDFO process, its contribution to the PAFDO cost is not significant. For example, when the FDFO dilution increased from 20 to 40, the PAO cost increased from \$0.213/m^3 to \$0.255/m^3 (i.e. 16.5% increase). This corresponded to the trend of the PAO average water flux reduction, suggesting that the flux at a 20-fold DS dilution was 22.4 L/m^2h while that at a 40-fold DS dilution was 17.0 L/m^2h (i.e. 24% decrease) shown in Fig. 8 (a). This again confirms the importance of average permeation flux to the total water cost of the PAFDO process and thus resulting in its economic sustainability.

These results indicate that when the DS dilution factor increased, the total cost significantly increased due to the much lower average permeation flux. This indicates that the total water cost of the PAFDO is highly influenced by dilution factor in the FDFO process. Therefore, a suitable stage to apply the PAO mode in FDFO process should be below a 60-fold DS dilution with considering a threshold flux of above 10 L/m^2h.

To clarify the effect of the PAO application on the total cost of PAFDO, a sensitivity analysis has been conducted based on three different hydraulic pressures of 2, 4, and 6 bar. This range was considered because of concern for the possibility of membrane deformation when applying a pressure higher than 6 bar (Blandin et al. 2013). The results of the sensitivity analysis are further presented in Fig. 8 (b).

The increase in the total water cost of the PAFDO due to additional hydraulic pressure is higher at lower DS dilution factor than at higher DS dilution factor. Specifically, as expected, when applying the PAO at a 10-fold DS dilution stage in the FDFO, the water cost of the PAFDO due to the applied pressure of 2 bar is \$0.364/m^3 while that of 6 bar is \$0.395/m^3 (7.9% increase). However, it is notable that for higher DS dilution factors of 50 and 80 the lowest total cost of the PAFDO was observed with the applied pressure of 4 bar. This can be explained by the results presented in Fig. 8 (a). It has been shown that the contribution of the FDFO cost to the PAFDO cost is more significant than that of the PAO. This is attributed to the possibility that higher DS dilution in the FDFO process could require more FO membrane areas thus increasing CAPEX and OPEX costs of the FDFO process.

Concerning this, as shown in Fig. 8 (b), when applying the PAO at a 20-fold DS dilution stage in the FDFO, the PAFDO cost with 2 bar applied pressure became similar to that with 4 bar applied pressure (\$0.442/m^3 and \$0.441/m^3 respectively). These results, therefore, demonstrate that the trade-off between savings in CAPEX costs (i.e. reduced total membrane area) when using pressure and the increased OPEX costs required for feed pressurization could occur at a certain DS dilution stage or factor. Thus, this finding can help define optimal operating conditions for an integrated FDFO and PAO system thus economically feasible.

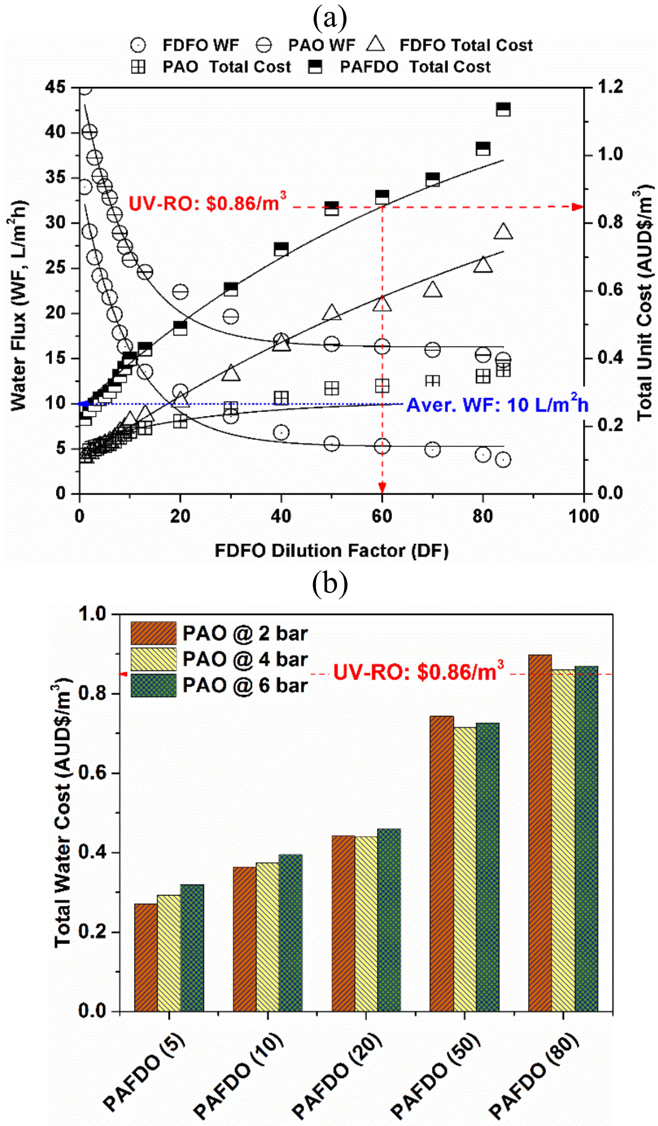


Fig. 8. Sensitivity analysis on (a) FDFO and PAO average water fluxes and total unit cost and (b) the applied pressure in PAO mode of FDFO operation related to the total water cost per m^3 of water produced. PAFDO (5) refers to PAO application at the FDFO dilution factor of 5. Plant water production capacity: 400 m^3/day.

4 Conclusions

Techno-economic analyses were conducted for a fertiliser driven forward osmosis process for commercial fertiliser dilution by biologically treated urban wastewater to produce irrigation water. Experimental investigations under the conditions used in this study showed that urban wastewater in particular the MBR effluent is the best feed stream for the FDFO application. In addition, the commercial diamond blue fertiliser, whose nutrient composition is well balanced for plants, turned out to generate an osmotic pressure similar to one of the inorganic salts well-known for their good performance as DS. In that way, this study demonstrated that the FDFO process is technically feasible with the potential to simultaneously reuse some amounts of wastewater and produces water for greenwall irrigation. Economic evaluation results showed that the integration of the PAO process can make PAFDO process economically favorable due to the enhancement of the average FDFO water flux and thus reduction in CAPEX. However, the sensitivity analysis proved that the average FDFO water flux and PAO application point in the FDFO process play a crucial role in economic feasibility of the PAFDO system, indicating that a coupling of higher average FDFO water flux ($\geq$ 10 L/m^{2}h) and PAO application at lower DS dilution factor (DF < 60) in the FDFO process is recommended. Finally, PAO operation with less than 4 bar applied is recommended since the energy penalty (i.e. increased OPEX cost) caused by the PAO application of FDFO operation is compensated by improved water permeation (i.e. reduced CAPEX cost). Although further work is required to validate the application of PAO in FDFO process in terms of fouling behaviour, FO element arrangement configurations, and a full-scale system design to control water flux and dilution factor, this study offers a better understanding for the process engineers to design and operate the collaborative process for the dilution of the fertiliser DS and the strategic management to lower the wastewater reuse cost for greenwall fertigation.

Acknowledgments

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