Fouling assessment of tertiary palm oil mill effluent (POME) membrane treatment for water reclamation
Mohd Syahmi Hafizi Ghani, Teow Yeit Haan, Ang Wei Lun, Abdul Wahab Mohammad, Rahmat Ngteni and Khairul Muis Mohamed Yusof

ABSTRACT
In order to minimize the adverse impacts of palm oil mill effluent (POME) towards the environment and to cope with the stress associated with water scarcity, membrane technology has been employed to reclaim water from POME. This study investigated the performance and fouling propensity of membranes in treating tertiary POME with the aim to recycle and reuse the reclaimed water as boiler feed water. Three types of membranes (NF270, BW30, and XLE) were used and their performances were evaluated based on the removal of chemical oxygen demand (COD), color, turbidity, total dissolved solids, phosphorus, and conductivity. All parameters were significantly reduced through XLE and BW30 membrane filtration processes in which the permeate was complied with the boiler feed water standard, except NF270 membrane where the COD value exceeded the allowable limit. High permeation drag of NF270 and rougher surface of XLE membranes resulted in the accumulation of foulant on the membrane surfaces which eventually reduced the permeate flux, whereas BW30 membrane was encountered for lower fouling propensity due to its low permeation rate. Hence, BW30 was deemed as the best candidate for water reclamation due to its low fouling propensity and because the production of permeate complied with boiler feed water standard.

INTRODUCTION
Malaysia is one of the world’s largest palm oil producers, accounting for 39% of world palm oil production and 44% of world export (MPOC 2014). From 2008 to 2015, the production of crude palm oil has increased from 17.73 to 19.96 million tonnes (MPOB 2016). Along with the increase of production capacity, a large amount of wastewater was also being generated. It is estimated that to produce 1 tonne of crude palm oil, 5–7.5 tonnes of water is required and more than 50% of this water will be disposed as palm oil mill effluent (POME) (Ahmad et al. 2003). POME is brownish in color with a distinct offensive odor. It contains a large amount of fatty acids, proteins, carbohydrates and other plant materials, high biochemical oxygen demand (21,500–28,500 mg/L) and chemical oxygen demand (COD) (45,500–65,000 mg/L) levels (Shian et al. 2009). Discharging of POME into the river will contaminate the waterway, destroy the river ecosystem and impair the community that relies on the river water (Khalid & Mustafa 2015). Thus, POME has to be treated properly before it can be discharged into the environment.
In Malaysia, more than 85% of palm oil mills have adopted the ponding system to treat the POME due to its lower cost compared to other technologies (Khalid & Mustafa 1992; Teng et al. 2015). However, the associated issues such as long retention time and difficulty in ensuring process efficiency have driven the search for alternative treatment technology (Chin et al. 1996; Yacob et al. 2005). One of the potential candidates is membrane technology. Membrane technology has been proven as a reliable process in treating a wide range of water such as wastewater, groundwater and surface water. The increasing acceptance of membrane technology is associated with stringent legislation for wastewater discharge (Peinemann & Nunes 2010; Ang et al. 2015). It offers many benefits such as wide applicability, invariable quality of produced water, efficient, economical, easy automation and does not require highly skilled operators (Cheryan & Rajagopal 1998; Xia et al. 2004).

Membrane-treated water with good quality has the potential to be recycled and reused for palm oil mill consumption such as feed water for cooling and boiler towers or for manufacturing practices such as washing the floors, external part of trucks and rinsing outside areas (Asano et al. 2007; Azmi & Yunus 2014). In this context, the practice of reclaiming water from POME could be economically beneficial to the palm oil industry by reducing the demand of tap water (Andrade et al. 2015). In addition, this practice would greatly help to reduce the quantity of wastewater discharge (Mavrov 2000). Such benefits have been widely covered by researchers in other industries. For instance, Dolar et al. (2011) reported that reverse osmosis (RO)/nanofiltration (NF) membrane treatment can effectively remove the fluoride and phosphate from fertilizer wastewater. The permeate water quality is good enough to be reused or safely discharged into the river. Vourch et al. (2008) claimed that the quality of RO purified water from dairy wastewater can be reused for heating, cleaning and cooling applications.

However, despite the expansion and successful application of membrane technology in the water industry, membrane fouling, which degrades the performance of membrane filtration process, remains a critical problem. Membrane fouling can be defined as the deposition of particles inside or on top of the membrane surface. It may decrease the permeate flux and the permeate quality, increase the overall energy consumption and frequency of chemical cleaning, incur additional expenses due to shorter membrane lifespan and additional labor for maintenance (Al-Amoudi & Lovitt 2007). Hence, in order to harvest the benefits of reclaimed water from POME, thorough investigation on practicality and performance of membrane filtration process for long-term operation has to be conducted.

In our preliminary short term study, it has been found that the water reclaimed from diluted aerobic digested POME, after undergoing the membrane treatment process, fulfilled the criteria for boiler feed water. Therefore, in this study, long-term performance of the membrane filtration process towards the reclamation of diluted aerobic digested POME will be investigated with particular attention being paid to the membrane performance and membrane fouling propensity. The finding from this study is important in optimizing the membrane treatment process for POME reclamation and to provide an insight into the impact of membrane fouling on the permeate quality.

MATERIALS AND METHODOLOGY

Membranes

The NF and RO membranes employed were the products of Dow FilmTech (USA). The characteristics of the membranes used are listed in Table 1. XLE and BW30 membranes belong to brackish water reverse osmosis (BWRO)

<table>
<thead>
<tr>
<th>Membrane</th>
<th>MWCO (Da)</th>
<th>Zeta potential at pH 9 (mV)</th>
<th>Root mean square (RMS) roughness (mm)</th>
<th>Water permeability coefficient (L/m² h)</th>
<th>Contact angle (°)</th>
</tr>
</thead>
<tbody>
<tr>
<td>NF270</td>
<td>200–400</td>
<td>–41.3</td>
<td>9.0 ± 4.2</td>
<td>21.18</td>
<td>36.6 ± 13.17</td>
</tr>
<tr>
<td>XLE</td>
<td>~100</td>
<td>–27.8</td>
<td>142.8 ± 9.6</td>
<td>14.45</td>
<td>62.8 ± 17.26</td>
</tr>
<tr>
<td>BW30</td>
<td>~100</td>
<td>–10.1</td>
<td>68.3 ± 12.5</td>
<td>6.54</td>
<td>56.2 ± 20.57</td>
</tr>
</tbody>
</table>

*Zeta potential values and RMS roughness were taken from Tang et al. (2009b).
membranes whereas the NF270 membrane is categorized as NF membrane.

**Preparation of feed solution**

The feed solution used in this study was collected from the first aerobic digester pool after the closed anaerobic digester system at East Mill Sime Darby Palm Oil Plantation located at Carey Island, Selangor, Malaysia. The collected aerobic digested POME was preserved in a cold room, at temperatures below 4 °C but above the freezing point, immediately after sampling to prevent the POME from undergoing microbial biodegradation.

During the membrane filtration study, the collected aerobic digested POME was diluted to around 150 mg/L COD value to imitate the quality of POME after treating by biofilm. The diluted aerobic digested POME is also known as tertiary POME since it has undergone several treatment stages. The typical characteristics of the diluted aerobic digested POME are summarized in Table 2.

**Cross-flow membrane filtration system**

A laboratory bench scale cross-flow membrane filtration system, as shown in Figure 1, was used for this study. The commercial flat sheet membranes were cut into a rectangular shape with an effective filtration area of 0.0042 m² (excluding the area covered by the O-ring). The membrane was then laid on top of the CF 042 membrane holder (Sterlitech, USA) and tightened by a rubber O-ring. Before membrane filtration was started, the newly cut membrane was soaked in ultra-pure water and left for 1 day to remove the residual solvent/chemical from the membrane.

In order to alleviate the impact of compaction, pre-filtration with ultra-pure water was first conducted at a constant pressure of 6 bars for 1 hour until steady-state flux was achieved. Diluted aerobic digested POME was then charged into a 10 L feed tank. Retentate was recycled into the feed tank in which the feed solution temperature was maintained at 27 °C using a re-circulating water chiller (SPH-20, Malaysia). The applied pressure of the membrane filtration system was generated using the high pressure pump (Blue Clean, BC 610, Italy) and controlled at 3 bars for all experiments. Two pressure gauges were used to indicate the operating pressure of the feed and retentate streams.

The permeate flux \( J \) was determined by direct measurement of permeate volume over time:

\[
J = \frac{V}{At}
\]

where \( J \) is the permeate flux (L/m² h), \( V \) is the permeate volume (L), \( A \) is the membrane effective surface area (m²), and \( t \) is the permeation time (h).

The membrane rejection \( R \) was calculated using the following equation:

\[
R = \frac{C_i - C_f}{C_i} \times 100\
\]

where \( R \) denotes the membrane rejection (%) and \( C_i \) and \( C_f \) indicate the concentration of feed solution and permeate, respectively.

The membrane fouling study was conducted for 6 hours with all operating conditions being controlled and

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**Table 2 | Typical characteristics of diluted aerobic digested POME**

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>COD (mg/L)</td>
<td>152</td>
</tr>
<tr>
<td>TDS (mg/L)</td>
<td>394.68</td>
</tr>
<tr>
<td>Color (PtCo)</td>
<td>626</td>
</tr>
<tr>
<td>Phosphorus (mg/L)</td>
<td>3.63</td>
</tr>
<tr>
<td>Turbidity (NTU)</td>
<td>18.1</td>
</tr>
<tr>
<td>Conductivity (μs)</td>
<td>778</td>
</tr>
<tr>
<td>pH</td>
<td>7.68</td>
</tr>
</tbody>
</table>
maintained at the aforementioned values. Fresh diluted aerobic digested POME was added every 2 hours to maintain the concentration of feed solution. In order to evaluate the fouling propensity of membranes, relative flux reduction (RFR), which indicates the fouling tendency of the membrane, was calculated as follows:

\[
RFR(\%) = 1 - \frac{J_p}{J_{w1}} \times 100\% 
\]  

where \( RFR \) is the relative flux reduction (%), \( J_p \) is the instantaneous permeate flux (L/m² h), and \( J_{w1} \) is the initial permeate flux (L/m² h).

**Analytical methods**

The performance of each membrane in treating the diluted aerobic digested POME was evaluated by assessing the permeate water quality based on several parameters such as COD, total dissolved solids (TDS), phosphorus (P), color, turbidity, conductivity, and pH. COD was measured by preheating the samples at 150 °C in a Hach digital reactor RBC 200 (Hach Company, Colorado, USA) for 2 hours and analyzed using DR3900 Benchtop spectrophotometer with RFID* Technology (Hach Company, Colorado, USA) at a high range (3–1,500 mg/L). Phosphorus (P) and color were measured based on the PhosVer 3 method and platinum-cobalt standard method using a DR3900 Benchtop spectrophotometer with RFID* Technology (Hach Company, Colorado, USA). Turbidity was measured using a 2100 N Laboratory Turbidimeter (Hanna, USA). Conductivity, TDS and pH of the permeate were measured using an HI 2550 Benchtop Meter (Hanna, USA).

**Membrane characterization**

In order to characterize the morphology of neat and fouled membranes (surface and cross-sectional views), a field emission scanning electron microscope (FESEM) was employed (Merlin Compact, Zeiss, Germany). For FESEM observation, the membrane samples were cut into an appropriate size and mounted on the sample holders. The samples were examined under the electron microscope at potentials of 3.00 kV at various magnifications.

**Membrane fouling mechanism**

Four blocking models have been proposed to describe the membrane fouling phenomena. These models comprise complete blocking, intermediate blocking, standard blocking and a cake filtration mechanism.

**Standard blocking model**

Standard blocking occurs when small size solutes are adsorbed into the membrane pores (Bowen et al. 1995). The solutes that trapped in the membrane pore will eventually contribute to the decrease of membrane pore size. The volume per unit membrane filtration area for standard blocking model is depicted by the equation below:

\[
V = \left( \frac{1}{J_0 t} + \frac{K_s}{2} \right)^{-1} 
\]

where \( J_0 \) is the initial permeate flux (L/m² h), \( t \) is time (s) and \( K_s \) (m⁻¹) is the standard blocking constant.

**Complete blocking**

In the case of complete blocking, the particles are larger than the membrane pore. In this case, the particles are deposited on the membrane surface and block or seal the pores without superposition of the particle. The filtration resistance will increase as the number of blocked pores increases. The correlation between volume per unit area membrane \( V \) and time \( t \) for complete blocking is given by Equation (5):

\[
V = \frac{J_0}{K_b} (1 - \exp (-K_b t)) 
\]

where \( J_0 \) is the initial permeate flux (L/m² h), \( t \) is time (s) and \( K_b \) (s⁻¹) is complete blocking constant.

**Intermediate blocking**

Intermediate blocking is somewhat similar to complete blocking. A particle can deposit on top of the particles that had already deposited on the membrane surface to
form multi-layers on the membrane surface. The volume per unit area membrane \( (V) \) versus time \( (t) \) is given by Equation (6):

\[
V = \frac{1}{K_i} \ln (1 + K_i J_0 t)
\]

where \( J_0 \) is the initial permeate flux \( (L/m^2 \cdot h) \), \( t \) is time \( (s) \), and \( K_i \) \( (m^{-1}) \) is intermediate blocking constant.

**Cake filtration model**

Cake formation occurs when the solutes are larger than the membrane pore size. The solutes settle down on the membrane surface that are already covered with solutes. Over time, a layer of cake consisting of deposited solutes will be formed. The relation between time \( (t) \) and volume per unit membrane filtration area \( (V) \) for the cake filtration model is given by Equation (7):

\[
V = \frac{1}{K_c J_0} \left( \sqrt{1 + 2K_c J_0^2 t} - 1 \right)
\]

where \( J_0 \) is the initial permeate flux \( (L/m^2 \cdot h) \), \( t \) is time \( (s) \) and \( K_c \) \( (s/m^2) \) is the cake filtration constant.

### RESULTS AND DISCUSSION

**Performance evaluation of cross-flow membrane filtration system**

The characteristics of the feed solution and permeate after 6 hours of filtration are summarized in Table 3, while Figure 2 shows the percentage removal of each parameter after the membrane filtration process. In order to confirm the usability of the treated water, the quality of permeate was compared to the boiler feed water standard for low to moderate pressure boilers set by the US Environmental Protection Agency (USEPA).

BW30 and XLE membranes managed to reject almost 98 and 94% of COD and TDS in the feed solution, respectively, while NF270 membrane showed slightly lower removal percentages of COD and TDS, at 95 and 93%, respectively. This is because the NF270 membrane has a looser structure (larger MWCO), as depicted in Table 1, compared to XLE and BW30 membranes. Therefore, the NF270 membrane is more vulnerable towards rejected particles with a smaller size. Similar findings have been reported by Coskun et al. (2010) where NF membrane (NF270) was found to have lower COD rejection compared to RO membranes (XLE and BW30) in treating olive mill wastewater. On the other hand, more than 98% of the suspended particles and impurities were found to be successfully removed from the diluted aerobic digested POME. Permeate produced after each membrane filtration process was recorded with turbidity less than 0.5 NTU, which fell within the allowable range for boiler feed water. Due to the great removal of suspended particles and organic substances which contributed to the turbidity of water sample, the color of the diluted aerobic digested POME changed from brownish to colorless after undergoing membrane filtration, as illustrated in Figure 3, whereas the initial phosphorus concentration in feed solution was much lower than the allowable limit for boiler feed water. However, further removal of phosphorus by the membrane filtration process in this study is expected to prohibit the scaling problem in the boiler which is mainly attributed to the deposition of phosphate during boiler operation (Chemtreat.com 2016).

In order to maintain the operation efficiency of a boiler tower, it must be fed with good quality boiler feed water. Based on the results presented in Table 3, it can be concluded that all parameters of the permeate obtained after filtering with BW30 and XLE membranes met the boiler feed water standard set by USEPA. Owing to the superior permeate quality obtained, the treated water can be recycled and reused as boiler feed water. However, the COD

**Table 3 | Characteristics of the feed solution and permeate after 6 hours of filtration**

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Feed</th>
<th>NF270</th>
<th>XLE</th>
<th>BW30</th>
<th>Boiler Feed</th>
</tr>
</thead>
<tbody>
<tr>
<td>COD (mg/L)</td>
<td>152</td>
<td>8</td>
<td>5</td>
<td>5</td>
<td>5</td>
</tr>
<tr>
<td>TDS (mg/L)</td>
<td>394.67</td>
<td>24.00</td>
<td>21.10</td>
<td>21.50</td>
<td>100</td>
</tr>
<tr>
<td>Color (PtCo)</td>
<td>626</td>
<td>5</td>
<td>5</td>
<td>4</td>
<td>–</td>
</tr>
<tr>
<td>Phosphorus (mg/L)</td>
<td>3.63</td>
<td>0.73</td>
<td>0.72</td>
<td>0.61</td>
<td>30–50</td>
</tr>
<tr>
<td>Turbidity (NTU)</td>
<td>18.100</td>
<td>0.233</td>
<td>0.141</td>
<td>0.209</td>
<td>0–3</td>
</tr>
<tr>
<td>Conductivity (μs)</td>
<td>778.00</td>
<td>49.30</td>
<td>41.07</td>
<td>42.89</td>
<td>–</td>
</tr>
<tr>
<td>pH</td>
<td>7.68</td>
<td>7.98</td>
<td>7.84</td>
<td>7.62</td>
<td>7–10</td>
</tr>
</tbody>
</table>
concentration of the permeate obtained after filtering with the NF270 membrane exceeded the standard limit by 3 mg/L. The excessive COD in the boiler feed water will cause scale formation. Scale formation results in a decrease in heat transfer and lower efficiency. Scale deposits can also cause overheating and failure of boiler metal (The National Board of Boiler and Pressure Vessel Inspectors 2015). Nevertheless, this water can be used for other purposes such as for factory cleaning or machine cleaning.

Membrane fouling for long term filtration process

Figure 4 shows the profiles of membrane permeate flux for NF270, XLE and BW30 membranes during 6 hours of filtration. The NF270 membrane had the highest initial permeate flux at 50.32 L/m² h followed by XLE membrane
(36.69 L/m² h), and BW30 membrane had the lowest initial permeate flux at 15.06 L/m² h. This observation was in correspondence with the membrane water permeability reported in Table 1. The NF membrane has been known as ‘loose’ RO membrane as its structure is less dense than RO membrane. Therefore, water can easily pass through the membrane and lead to high permeate flux. On the other hand, XLE is the abbreviation of extra low energy RO membrane, which indicates that it could be operated at lower operating pressure but is able to produce equal permeate flux as compared to the standard brackish water RO membrane, BW30. In other words, under similar operating pressure, XLE can produce higher flux over BW30 (Baker 2004).

During the first 200 minutes of filtration, both NF270 and XLE membranes exhibited an obvious gradual flux decline compared to BW30. After 6 hours of membrane filtration period, NF270, XLE and BW30 membranes recorded RFR values of 16, 14 and 8%, respectively. Fouling propensity of a membrane is generally affected by several factors including membrane hydrophilicity (wettability), membrane surface charge, and membrane surface roughness (Teow et al. 2013). Theoretically, a membrane with higher hydrophilicity, higher surface charge, lower surface roughness is less likely to foul. As depicted in Table 1, the NF270 membrane had the highest membrane hydrophilicity (reflected by the lowest contact angle value), highest membrane surface charge (reflected by the lowest zeta potential value), and lowest surface roughness (reflected by the lowest root mean square (RMS) value). Thus, the NF270 membrane was expected to have the least fouling tendency. Surprisingly, a contradictory result was obtained in Figure 4 where NF270 membrane experienced the most severe fouling among the other membranes. This could be attributed to the high permeation drag of NF270 and XLE membranes which are about 3.6 times and 2.6 times higher than BW30 membrane, dominantly leading to fast cover up of membrane surface by the organic matter (foulant) in the POME membrane, dominantly leading to fast cover up of membrane surface and will severely restrict the permeation of water through the membrane. Consequently, a sharp decline in flux was obtained.

After 200 minutes of filtration, membrane permeate flux had reached a constant level due to the compaction and thickening of the fouling cake layer (Nghiem & Hawkes 2009). Figure 5 shows the cross-sectional FESEM images of neat and fouled NF270, XLE and BW30 membranes, whereas Figure 6 depicts the top surface FESEM images of neat and fouled NF270, XLE and BW30 membranes. As presented in Figure 5(a), the cross-sectional structures of all NF270, XLE, and BW30 membranes are similar to each other. Thus, the membrane cross-sectional morphology is expected to not alter the fouling mechanism in this study any obvious impact towards membrane fouling. Figure 5(b) shows that the foulant layer on NF270 membrane was much thicker (1,206 nm) compared to XLE membrane (122.8 nm). However, the foulant layer on BW30 membrane was almost negligible. On the other hand, the top surface view of the membranes depicted in Figure 6 indicate that the top surface of NF270 and XLE membranes were fully covered by a foulant layer, whereas the valleys on BW30 membranes surface were partially filled with foulant. The observation obtained from FESEM images was affirmatively supporting the membrane permeate flux profiles depicted in Figure 4.

Although both BW30 and XLE membranes are categorized as BWRO membrane, there was a marked difference in their membrane fouling severity. The factor contributing to this observation could be the membrane surface roughness. As shown in Table 2, XLE membrane had a rougher surface compared to BW30 membrane in which the RMS roughness for XLE and BW30 membranes was 142.8 ± 9.6 and 68.3 ± 12.5 nm, respectively. Generally, foulant is more easily adsorbed onto the rougher membrane surface or into the membrane valley (Vrijenhoek et al. 2001; Boussu et al. 2006; Idris et al. 2007; Cao et al. 2010). Consequently, foulant will have a higher tendency to deposit on the rougher XLE membrane surface and will significantly reduce the membrane permeate flux (Vrijenhoek et al. 2001; Tang et al. 2009a; Hurwitz et al. 2010). Though membrane surface roughness played an important role in affecting the membrane fouling propensity, NF270 membrane, which has the smoothest surface, experienced the most severe fouling.
This could be attributed to the relatively high permeate drag of NF270 membrane where the frictional drag forced the foulant to attach onto the membrane surface.

As an outcome from the membrane fouling study, BW30 membrane, which has a low fouling propensity as compared to NF270 and XLE membranes, could be a good option for long period membrane filtration in acquiring treated water with boiler feed standard without degrading the membrane performance.

**Membrane fouling mechanism**

Figure 7 shows the fitting of four blocking models (complete blocking, intermediate blocking, standard blocking and cake
filtration) to the experimental data in explaining the membrane fouling phenomenon of each membrane filtration process in this study, while Table 4 summarizes the degree of model fitness (R² value) for different fouling mechanism prediction. The best fitted model was determined by the highest value of R².

For BW30 and XLE membranes, standard pore blocking and complete pore blocking models provide the best fit with R² values above 0.99. However, the pore blocking mechanism in this context might be different from the pore blocking fouling mechanism which occurs on ultrafiltration (UF) or microfiltration (MF) membranes because NF and RO
membranes do not have a discrete pore size as UF and MF membranes do. The pore blocking of BW30 and XLE membranes in this study was interpreted as the flux shut off by the foulant deposited on the membrane surface, whereby the filtrate can only pass through the unblocked area (Ang et al. 2018). Indeed, the top surface FESEM images of fouled BW30 and XLE membranes in Figure 6(b) show that the foulant only partially covered the membrane surface which had agreed well with this postulation. On the other hand, none of the blocking models were able to fit to the experimental permeate flux data collected from NF270 membrane filtration as the R² value obtained from each model was extremely low. By referring to the cross-sectional FESEM image of fouled NF270 membrane in Figure 5(b), the foulant layer on top of the surface of the NF270 membrane is thick. This could possibly be due to more than one fouling mechanism being involved. Consequently, none of the individual pore blocking models can be used to describe the fouling mechanism occurring in the NF270 membrane filtration process.

Table 4 | The summary of R² value for different fouling mechanism prediction

<table>
<thead>
<tr>
<th>Membrane</th>
<th>Standard blocking</th>
<th>Complete blocking</th>
<th>Intermediate blocking</th>
<th>Cake formation</th>
</tr>
</thead>
<tbody>
<tr>
<td>NF270</td>
<td>&lt;0.1</td>
<td>&lt;0.1</td>
<td>&lt;0.1</td>
<td>&lt;0.1</td>
</tr>
<tr>
<td>XLE</td>
<td>0.999943</td>
<td>0.999932</td>
<td>&lt;0.1</td>
<td>&lt;0.1</td>
</tr>
<tr>
<td>BW30</td>
<td>0.996025</td>
<td>0.995968</td>
<td>0.795613</td>
<td>&lt;0.1</td>
</tr>
</tbody>
</table>

Figure 7 | Volume vs time data for standard model, intermediate model, complete model and cake formation model: (a) NF 270 membrane; (b) XLE; (c) BW 30.
CONCLUSIONS

This research investigated the performance of NF and RO membranes in claiming water from POME and reusing it as boiler feed water. It was found that the membrane technology significantly reduced the concentration of all parameters: color, turbidity, TDS, phosphorus, COD and conductivity. Permeate produced from RO, BW30 and XLE membranes showed great potential to be reused as the boiler feed water as all parameters fitted to the boiler feed standard set by USEPA, whereas for the membrane fouling study, the high initial permeate flux NF270 membrane encountered more severe flux decline (fouling) compared to RO membranes (XLE and BW30) in accordance with the high permeation drag. However, permeate flux data collected from the NF270 membrane filtration did not fit with any blocking model as it might involve several fouling mechanisms. The BW30 membrane, a BWRO membrane which has lower membrane surface roughness hence lower fouling propensity compared to the XLE membrane, could be a good option for long period membrane filtration in acquiring treated water with boiler feed standard without degrading the membrane performance. Standard blocking and complete blocking models which fitted well with the permeation data collected from BW30 membrane filtration indicate that ongoing fouling is progressively developed and proper cleaning has to be carried out at this stage to sustain the membrane performance and to extend the membrane lifespan. Overall, although the BW30 membrane had the lowest permeate flux, it was deemed as the best option for tertiary POME treatment for water reclamation due to less fouling propensity, more consistent performance and good grade of permeate produced.

ACKNOWLEDGEMENTS

The authors wish to gratefully acknowledge the financial support for this work by Geran Gerakan Penyelidikan Muda (GGPM-2016-030) from Universiti Kebangsaan Malaysia and Sime Darby Research Grant (KK-2014-006).

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First received 18 November 2016; accepted in revised form 1 March 2017. Available online 1 June 2017