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# Factors that affect the permeability of commercial hollow-fibre membranes in a submerged anaerobic MBR (HF-SAnMBR) system

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**Abstract**

A demonstration plant with two commercial HF ultrafiltration membrane modules (PURON®, Koch Membrane Systems, PUR-PSH31) was operated with urban wastewater. The effect of the main operating variables on membrane performance at sub-critical and supra-critical filtration conditions was tested. The physical operating variables that affected membrane performance most were gas sparging intensity and back-flush (BF) frequency. Indeed, low gas sparging intensities (around 0.23 $Nm^3\ h^{-1}\ m^{-2}$) and low BF frequencies (30 second back-flush for every 10 basic filtration-relaxation cycles) were enough to enable membranes to be operated sub-critically even when levels of mixed liquor total solids were high (up to 25 $g\ L^{-1}$). On the other hand, significant gas sparging intensities and BF frequencies were required in order to maintain long-term operating at supra-critical filtration conditions. After operating for more than two years at sub-critical conditions (transmembrane flux between 9 and 13.3 LMH at gas sparging intensities of around 0.23 $Nm^3\ h^{-1}\ m^{-2}$ and MLTS levels from around 10 to 30 $g\ L^{-1}$) no significant irreversible/irrecoverable fouling problems were detected (membrane permeability remained above 100 LMH $bar^{-1}$ and total filtration resistance remained below $10^{13}\ m^{-1}$), therefore no chemical cleaning was conducted. Membrane performance was similar to the aerobic HF membranes operated in full-scale MBR plants.

**Keywords**
Back-flush frequency; biogas sparging; commercial hollow-fibre membranes; submerged anaerobic membrane bioreactor; membrane permeability.

# 1. Introduction

## *1.1. Anaerobic treatment of urban wastewater using MBR technology*

In recent years there has been increased interest in assessing the feasibility of the anaerobic treatment of urban wastewater at ambient temperatures. This interest focuses on the sustainable advantages of anaerobic rather than aerobic processes, i.e. anaerobic processes generate little sludge due to the low anaerobic biomass yield; consume little energy because no aeration is needed; and generate biogas that can be used as an energy resource. The total greenhouse gas emissions of this technology are, therefore, low because low energy consumption indirectly means low gas emissions. The main challenge of anaerobic biotechnology is to develop treatment systems that prevent biomass loss and enable high sludge retention times (SRTs) in order to offset the low growth rates of anaerobic biomass at ambient temperatures (Lin *et al.*, 2010). In this respect, submerged anaerobic membrane bioreactors (SAnMBRs) are a promising technology for urban wastewater treatment. However, operating membrane bioreactors with high SRTs may lead to high mixed liquor total solids (MLTS) at a specific reacting volume. This is one of the main constraints of using membranes (Judd and Judd, 2011) since it can result in high membrane fouling propensities.

## *1.2. Membrane fouling in SAnMBRs*

The key challenge in SAnMBR technology is how to optimise membrane operating in order to minimise any kind of membrane fouling, especially the irrecoverable/permanent component that cannot be eliminated by chemical cleaning. The extent of irrecoverable/permanent fouling is what ultimately determines the membrane lifespan (Judd, 2008; Drews, 2010a; Patsios and Karabelas, 2011). Several strategies to control fouling (see, for example, Liao *et al*., 2006) aim to optimise filtration whilst minimising investment and operating costs. In this respect, the SAnMBR design strategy must be carefully selected. Depending on the design strategy, different design criteria can be adopted. One such criterion is based on operating membranes in sub-critical filtration conditions that are limited by the so-called critical flux ($J_C$) (Bachin *et al*., 1995; Field *et al*., 1995). Operating membranes sub-critically gives membranes long lifespans, which reduces replacement and maintenance costs (by minimising physical cleaning costs, i.e. membrane scouring or back-flush). In this respect, MLTS has been widely identified as one of the factors that affect $J_C$ most. Thus, an investment compromise between operating reactor volume and filtration area should be selected in order to keep MLTS at sub-critical levels for a given transmembrane flux (J). Another design criterion is based on operating membranes at critical or supra-critical filtration conditions. This reduces initial investment costs because it requires lower operating volumes and/or lower membrane surfaces than when operating membranes at sub-critical filtration conditions, however, replacement, maintenance and operating costs are probably higher.

Regardless of the design criterion adopted, it is necessary to determine which filtration conditions (Drews *et al*., 2010b) are most suitable in order to optimise the membrane module design and configuration. An exhaustive analysis in the different potential operating conditions is, therefore, necessary in order to optimise both membrane lifespan (i.e. membrane replacement cost) and operating and maintenance costs (i.e. the cleaning mechanism). In this respect, it is necessary to assess the impact of the main operating variables upon membrane performance, i.e. frequency and duration of the physical cleaning stages (back-flush and relaxation); gas sparging intensity; cross-flow sludge velocity over the membrane surface (for cross-flow membrane configurations); up-flow sludge velocity in the membrane tank (submerged membrane configurations) which determines the sludge concentration factor when the membranes are located in external tanks; and maximum operating transmembrane pressure (TMP).

### *1.3. Full-scale implementation of SAnMBRs*

Membrane technology has been used increasingly to treat wastewater over the last decade (Lesjean and Huisjes, 2007) even in large urban WWTPs. The treatment capacity of urban MBR WWTP has significantly increased (to maximum design flow rates of more than 150000 $m^3$ $day^{-1}$) in just a few years (Huisjes et al., 2009). As regards membrane configuration, flat sheet (FS) membranes are used mostly in small plants (< 5000 $m^3$ $d^{-1}$), whilst hollow fibre (HF) membranes are used for the entire flow range and prevail in large plants (> 10000 $m^3$ $d^{-1}$) and account for about 75% of all total MBR installed capacity (Cote et al., 2012).

Nevertheless, it is important to highlight that all these urban MBR WWTPs are aerobic wastewater treatments. Although MBR technology has not yet been applied to full-scale anaerobic urban wastewater treatment, the scientific community is showing increasing interest in the feasibility of its full-scale implementation because of the above-mentioned advantages. Indeed, several studies which assess the feasibility of using SAnMBR technology to treat urban wastewater at the laboratory scale have been published (Jeison and van Lier, 2007; Huang et al., 2008; Lew et al., 2009). However, the impact of the main operating conditions upon membrane fouling cannot be determined exactly at the lab scale because they depend heavily on the membrane size. In HF membranes in particular, HF length is a key performance parameter. In this respect, there is still a lack of thorough knowledge about fouling mechanisms, mainly as regards hydraulic performance and membrane permeability (Guglielmi *et al*., 2007; Di Bella *et al*., 2010; Mannina *et al*., 2011). In addition, it is expected that membrane fouling will be affected to a considerable degree by the different characteristics of aerobic and anaerobic mixed liquors, such as particle size distribution, extracellular polymeric substances (EPS), soluble microbiological products (SMP), biomass concentration, inorganic and organic compounds (Lin *et al*., 2009), or pH values affecting both biofouling (Sweity *et al*., 2011) and formation of chemical precipitates.

Therefore, since membrane performance cannot be scaled up directly from laboratory to plant dimensions, especially in the case of HF-based technology (Liao *et al*., 2006), further studies of HF-SAnMBR technology on an industrial scale are needed in order to facilitate its design and implementation in full-scale wastewater treatment plants (WWTPs).

To gain more insight into the optimisation of the physical separation process in a SAnMBR system at the industrial scale, this paper shows the impact of the main operating variables upon the performance of industrial HF membranes. Gas sparging intensity, up-flow sludge velocity in the membrane tank, duration and frequency of the different physical cleaning stages (relaxation and back-flush), and length of filtration stage were evaluated in an SAnMBR system featuring commercial HF membrane modules. The effect of these variables at two different membrane operating conditions (sub-critical and critical/supra-critical filtration conditions) was assessed. The plant was operated using Carraixet WWTP pre-treatment effluent (Valencia, Spain). On the basis of the results obtained this study aims to provide guidelines for the sub-critical and critical operation of commercial HF membranes in an SAnMBR system.

## 2. Materials and methods

### *2.1. Demonstration plant description*

Figure 1 shows the flow diagram of the HF-SAnMBR demonstration plant used in this study. It consists mainly of an anaerobic reactor with a total volume of 1.3 $m^3$ (0.4 $m^3$ head space) connected to two membrane tanks each with a total volume of 0.8 $m^3$ (0.2 $m^3$ head space). Each membrane tank has one industrial HF ultrafiltration membrane unit (PURON®, Koch Membrane Systems (PUR-PSH31) with 0.05 µm pores). Each module has 9 HF bundles, 1.8 m long, giving a total membrane surface of 30 $m^2$. Normal membrane operating entails a specific schedule involving a combination of different individual stages taken from a basic filtration-relaxation (F-R) cycle. In addition to traditional membrane operating stages (filtration, relaxation and back-flush), another two stages of membrane operation were considered: degasification and ventilation.

For further details of this SAnMBR demonstration plant see Giménez *et al*. (2011) and Robles *et al.* (2012a).

### *2.2. Operating conditions*

The demonstration plant was fed with effluent from pre-treatment of a full-scale WWTP (screening, degritter, and grease removal), which main component is domestic type. It is important to emphasise the great variation in the characteristics of the anaerobic reactor influent (e.g. 186 ± 61 mg $L^{-1}$ of TSS and 388 ± 95 mg $L^{-1}$ of total COD), which is reflected by the high standard deviation of each parameter. The plant was operated using an SRT of 70 days on operating days 1 to 445, and an SRT of 40 days on operating days 446 to 600. Hydraulic retention times (HRTs) ranged from 5 to 24 hours. The temperature varied from around 33 to 15 ºC. The pH of the mixed sludge ranged from 6.5 to 7.1, and carbonate alkalinity remained at values of around 600 $mgCaCO_3\ L^{-1}$.

We studied how membranes operate in both short-term trials and in the long term. In the latter instance, the membrane underwent 300-second basic F-R cycles (250 s filtration and 50 s relaxation) with 30 seconds of back-flush every 10 cycles, 40 seconds of ventilation every 10 cycles, and 30 seconds of degasification every 50 cycles. In addition, six different $J_{20}$ and temperature conditions were tested: 13.3, 10, 12, 13.3, 11 and 9 LMH, at controlled temperatures of 33, 33, 25, and 20 ºC, spring and summer ambient temperatures (from approx. 20 to 30 ºC), and autumn and winter ambient temperatures (from approx. 30 to 14 ºC), respectively. Hence, the overall operating period was divided into six experimental periods (periods i, ii, iii, iv, v and vi) taking into account both $J_{20}$ and temperature. The average specific gas demand per membrane area ($SGD_m$) was 0.23 $Nm^3\ h^{-1}\ m^{-2}$. A maximum TMP safety value of 0.4 bars was set. The flow of sludge through the membrane tank was set to 2700 L $h^{-1}$, giving an up-flow sludge velocity of 2.7 mm $s^{-1}$.

In order to evaluate the critical filtration conditions throughout the long-term membrane performance, different short-term trials (flux-step type, see Robles et al., 2012a) were carried out. For instance, on day 125 and day 590 (operating with MLTS of 23 g $L^{-1}$ and $SGD_m$ of 0.23 $Nm^3\ h^{-1}\ m^{-2}$), the critical flux resulted in 14 and 10.5 LMH, respectively. Therefore, the critical flux remained generally at values over 10.5 – 14 LMH during the operating period since $SGD_m$ was maintained at 0.23 $Nm^3\ h^{-1}\ m^{-2}$ and MLTS remained mostly below 23 g $L^{-1}$. Hence, the long-term operating shown in this study was mainly carried out at sub-critical filtration conditions ($J_{20}$ was varied from 9 to 13.3 LMH).

In addition, several short-term trials were conducted at sub-critical and supra-critical filtration conditions with varying gas sparging intensities, up-flow sludge velocities in the membrane tank, durations and frequencies of the different physical cleaning stages (relaxation and back-flush), and lengths of filtration. Normally, the membrane was operated with 300-second basic F-R cycles (250 s filtration and 50 s relaxation), 30 seconds of back-flush every 10 cycles, 40 seconds of ventilation every 10 cycles, and 30 seconds of degasification every 50 cycles, whilst the operating $J_{20}$ was 10 LMH, the average $SGD_m$ was 0.23 $Nm^3\ h^{-1}\ m^{-2}$, and the up-flow sludge velocity was 2.7 mm $s^{-1}$. Table 1 summarises the values of the operating variables studied in each short-term trial. In each trial, the sub-critical and supra-critical conditions were determined by the different levels of MLTS. The $J_{20}$ in these short-term trials was set to 10 LMH, whilst the other operating variables of the membrane operating mode were established in the same way as the general, long-term operating conditions mentioned above.

## *2.3. Analytical methods*

### *2.3.1. Analytical monitoring*

In addition to the on-line process monitoring, the performance of the biological process was assessed by taking 24-hour composite samples of the influent and effluent streams, and taking grab samples of anaerobic sludge once a day. The following parameters were analysed daily: total solids (TS); total suspended solids (TSS); volatile suspended solids (VSS); carbonate alkalinity; and nutrients (ammonium ($NH_4$-N) and orthophosphate ($PO_4$-P)). The total and soluble chemical oxygen demand ($COD_T$ and $COD_S$, respectively) were determined once a week.

Solids, COD, and nutrients were determined according to Standard Methods (2005). Carbonate alkalinity was determined by titration according to the method proposed by WRC (1992).

*2.3.2. Membrane performance indices*

The 20 ºC-normalised membrane permeability ($K_{20}$) was calculated using a simple filtration model (Equation 1) that takes into account the TMP and J data monitored online. This simple filtration model includes temperature correction (Equation 2) to account for the dependence of permeate viscosity on temperature (Rosenberger *et al.*, 2006), and therefore the 20 ºC-normalised transmembrane flux ($J_{20}$) was calculated by applying Equation 3. Relative membrane permeability ($K_0$) was used to assess the effect of the different operating factors on membrane performance. This relative permeability was defined as shown in Equation 4. Total membrane resistance ($R_T$) was theoretically represented by the following partial resistances (Equation 5): membrane resistance ($R_M$); cake layer resistance ($R_C$); irreversible layer resistance ($R_I$); and irrecoverable layer resistance ($R_{IC}$).

$$K_{20} = \frac{J_T f_T}{TMP} \qquad \text{(Eq. 1)}$$

$$f_T = \exp(-0.0239\,(T-20)) \qquad \text{(Eq. 2)}$$

$$J_{20} = J_T \cdot \exp(-0.0239\,(T-20)) \qquad \text{(Eq. 3)}$$

$$K_0(t) = \frac{K_{20}(t)}{K_{20}(t=0)} \qquad \text{(Eq. 4)}$$

$$R_T = R_M + R_C + R_I + R_{IC} \qquad \text{(Eq. 5)}$$

Moreover, $J_C$ was determined by applying a modified flux-step method based on the method proposed by van der Marel *et al.* (2009). $J_C$ was calculated according to its weak concept: the flux below which TMP and J are not directly related. When applying this method, the duration of both filtration and relaxation stages was set to 15 min. Flux-stepping was arbitrarily set to 1.22 LMH of $J_{20}$ (equivalent to a permeate flow rate of 50 L $h^{-1}$). The relaxation stages were conducted using the same $SGD_m$ as in the filtration stages. For further details about the applied flux-step method, see Robles *et al*. (2012a).

## 3. Results and discussion

### *3.1. Long-term performance*

Figure 2 depicts the average daily $K_{20}$ (Figure 2a) and the average daily $R_T$ (Figure 2b) obtained during the operating period, and the average daily MLTS in the anaerobic sludge entering the membrane tank. It must be said that the MLTS level in the membrane tank increased by up to 5 g $L^{-1}$, depending on the ratio between the net permeate flow rate and the sludge flow rate entering the membrane tank. The results shown in Figure 2 can be divided in two different long-term operating periods according to the irreversible/ irrecoverable fouling component observed: (1) days 1 to 300; and (2) days 300 to 600. It is important to note that since no chemical cleaning was conducted throughout the operating period, it was not possible to determine the single contribution to $R_T$ of both $R_I$ and $R_{IC}$.

Up to operating day 300, no significant irreversible/irrecoverable fouling was observed, since $K_{20}$ and $R_T$ recovered to values very close to the values obtained at the beginning of the long-term operation as MLTS decreased. This behaviour indicated that throughout this operating period, $R_T$ was mainly related to $R_C$ ($R_I$ was negligible), whilst a relatively constant contribution of about $5 \cdot 10^{11}$ $m^{-1}$ (at 650 LMH $bar^{-1}$ of $K_{20}$ treating clean water in similar operating conditions) was attributed to $R_M$. This behaviour means that the MLTS level is a key factor as regards membrane permeability in this HF-SAnMBR system (Robles *et al*., 2012b). In this respect, Figure 2a illustrates how every variation in MLTS was inversely reflected by $K_{20}$. Nevertheless, it is important to note that even at high MLTS levels (up to 25 g $L^{-1}$) and relatively low $SGD_m$ values (around 0.23 $Nm^3$ $h^{-1}$ $m^{-2}$), $K_{20}$ and $R_T$ remained at sustainable values, above 100 LMH $bar^{-1}$ and below $3 \cdot 10^{12}$ $m^{-1}$, respectively. At MLTS of more than 25 g $L^{-1}$, $K_{20}$ showed fell sharply because the 20 ºC-normalised $J_C$ was exceeded: 10 and 13 LMH when operating at 0.23 $Nm^3$ $h^{-1}$ $m^{-2}$ of $SGD_m$ and MLTS levels of 28 and 23 g $L^{-1}$, respectively (close to the operating membrane fluxes). Thus, at MLTS levels higher than 25 g $L^{-1}$, an $SGD_m$ of 0.23 $Nm^3$ $h^{-1}$ $m^{-2}$ was not enough to maintain sub-critical filtration conditions.

After operating day 300, a slight downward trend in $K_{20}$ and a slight upward trend in $R_T$ were observed even when operating at a more or less stable MLTS level (see period day 300 to 400). This behaviour revealed a progressive accumulation of irreversible/irrecoverable fouling over the membrane surface after one year of operation and $R_I$/$R_{IC}$ was detected along this period. Nevertheless, it must be emphasised that the membranes did not require any chemical cleaning after more than two years of operation, even with high MLTS and temperature shocks affecting the biomass population and the derived compounds. These results revealed that reversible fouling was successfully removed from the membrane surface and that irreversible fouling was low, mainly due to applying physical cleaning mechanisms (relaxation, back-flush and shear intensity gas sparging) and operating membranes under sub-critical filtration conditions. These results suggested that operating membranes under sub-critical filtration conditions during long-term operation minimises the likelihood of membranes being irreversibly fouled. However, it is well known that operating membranes under sub-critical rather than critical levels implies a higher total filtration area at a given $J_{20}$. Nevertheless, this larger filtration area will probably increase the membrane lifespan whilst decreasing maintaining necessities. Hence, a reduction in replacement, maintenance and operating costs can be achieved.

## ***3.2. Short-term trials: main factors affecting membrane performance***

### *3.2.1. Effect of gas sparging intensity*

Different sub-critical short-term trials were carried out at 0.17, 0.23, 0.33 and 0.40 $Nm^3\ h^{-1}\ m^{-2}$ of $SGD_m$ and MLTS of 20 g $L^{-1}$. An almost stable $K_0$ close to 1 (i.e. $K_{20}(t)$ remained very close to $K_{20}(t=0)$) was achieved in all trials. Thus, low fouling rates were observed (lower than 10 mbar $min^{-1}$). Membrane permeability recovered to the initial value of the short-term trial, indicating that no irreversible fouling component was detected. These results reveal that a $SGD_m$ of 0.17 $Nm^3\ h^{-1}\ m^{-2}$ (equal to 5 cm $s^{-1}$ i.e. the minimum value supplied by the blower) was enough to completely remove the reversible fouling from the membrane surface.

*3.2.1.1. Major role of gas sparging intensity when operating supra-critically*

Figure 3 shows the resulting $K_0$ at different $SGD_m$ when the membranes were operated with high MLTS, and thereby at supra-critical filtration conditions. The $SGD_m$ was set to 0.23 $Nm^3$ $h^{-1}$ $m^{-2}$ when the MLTS level fed to the membrane tank was 28 and 31.5 g $L^{-1}$, whilst the $SGD_m$ was set to 0.17 $Nm^3$ $h^{-1}$ $m^{-2}$ when operating at an MLTS of 30 g $L^{-1}$. As it can be observed in Figure 3, even operating at similar MLTS levels, the two short-term trials carried out at MLTS of 28 and 30 g $L^{-1}$ and different $SGD_m$ resulted in quite different behaviours. A sharp decrease in $K_0$ was detected in the short-term trial conducted at the lowest $SGD_m$ (0.17 $Nm^3$ $h^{-1}$ $m^{-2}$). In these operating conditions, a considerable increase of the reversible fouling rate was observed throughout the trial (up to 80 mbar $min^{-1}$). In this case, the $SGD_m$ applied was not enough to fulfil the membrane scouring necessities, and the filtration process was stopped because the maximum TMP (safety value set to 0.4 bars) was reached. On the other hand, in the trial carried out at 0.23 $Nm^3$ $h^{-1}$ $m^{-2}$ (equal to 7 cm $s^{-1}$) and an MLTS of 28 g $L^{-1}$, $K_0$ did not reach unsustainable values (the reversible fouling rate remained at values lower than 25 mbar $min^{-1}$.). However, $K_0$ decreased continuously throughout the trial as the reversible fouling was accumulated over the membrane. This accumulation could lead to a high irreversible/irrecoverable fouling propensity.

It is important to highlight that the $SGD_m$ applied in these short-term trials (0.23 $Nm^3$ $h^{-1}$ $m^{-2}$) was quite low compared to the typical operating range the supplier proposed for aerobic processes (from 0.3 to 0.7 $Nm^3$ $h^{-1}$ $m^{-2}$).

As Figure 3 shows, an increase in MLTS from 28 to 31.5 g $L^{-1}$ at the same $SGD_m$

($0.23\ Nm^3\ h^{-1}\ m^{-2}$) resulted in a significant decrease of $K_0$. In the trial with an MLTS of 31.5 g $L^{-1}$, an increase in the membrane reversible fouling rate of up to 80 mbar $min^{-1}$ was observed. The maximum TMP value was reached after 140 minutes and the trial was stopped. Hence, as a result of operating under critical filtration conditions, small variations in the MLTS concentration affected membrane performance considerably. Nevertheless, previous studies (flux-step type, see Robles *et al.*, 2012c) showed that it is theoretically possible to operate sub-critically at 10 LMH of $J_{20}$ and MLTS of 28 g $L^{-1}$ when $SGD_m$ is about $0.25\ Nm^3\ h^{-1}\ m^{-2}$. Therefore, it is assumed that gas sparging intensities around 0.3 – 0.5 $Nm^3\ h^{-1}\ m^{-2}$ may keep $K_0$ at proper values when operating at MLTS levels around 30 – 31.5 g $L^{-1}$.

The results shown in Figure 3 suggest that when membranes are operated supra-critically at low specific $SGD_m$ the duration and/or frequency of the physical cleaning stages (relaxation and back-flush) must be increased considerably. On the other hand, in order to operate the membranes sub-critically at 13.3 LMH of $J_{20}$ and MLTS levels of 23 and 28 g $L^{-1}$, the theoretical $SGD_m$ required is approx. 0.23 and 0.45 $Nm^3\ h^{-1}\ m^{-2}$ (both calculated using the flux-step method), respectively. In contrast, $SGD_m$ values lower than 0.1 $Nm^3\ h^{-1}\ m^{-2}$ are theoretically needed when operating sub-critically at 13.3 LMH of $J_{20}$ and MLTS levels of 11.5 g $L^{-1}$, which are quite low when compared with aerobic MBR technology operating in similar conditions.

*3.2.1.2. Gas sparging intensity as a key operating parameter for optimising SAnMBRs at the industrial scale*

The results obtained confirm the need to optimise the gas sparging intensity in all membrane operating conditions. The gas sparging intensity poses a major challenge since it must be minimised in order to maximise energy savings. It is important to emphasise that aeration energy can account for up to 50 - 75% of all the energy consumed by aerobic MBR technology (Verrecht *et al*., 2010). Not only can considerable energy savings be achieved but also appropriate long-term operating because the onset of irreversible/irrecoverable fouling problems can be minimised.

Hence, controlling gas sparging to ensure appropriate membrane scouring is mandatory in order to optimise the economic feasibility of operating HF membranes in full-scale SAnMBR systems. In this respect, several recently-published studies assess different monitoring strategies designed to save energy in aerobic MBR technology (see e.g. Huyskens *et al*., 2011; Ferrero *et al*., 2012). Nevertheless, the applicability of these control strategies for saving energy in SAnMBR technology on an industrial scale has yet to be evaluated.

### *3.2.2. Effect of up-flow sludge velocity in the membrane tank*

Figure 4 illustrates how the up-flow sludge velocity in the membrane tank affects membrane performance. This operating variable is related to the sludge concentration factor resulting from the ratio between the sludge flow entering the membrane tank and the net permeate flow. For instance, Equation 6, 7 and 8 show the expected MLTS level in the membrane tank as a function of the MLTS level in the sludge fed to the membrane tank when the up-flow sludge velocity is set to 1.0, 2.2 and 2.7 mm $s^{-1}$, respectively. This expected MLTS was calculated on the basis of a mass balance according to the above-mentioned ratio between the sludge flow entering the membrane tank and the net permeate flow (i.e. according to the applied up-flow sludge velocity in the membrane tank). The permeate flow rate was set to a constant value of 300 L $h^{-1}$ ($J_{20}$ of 10 LMH).

As indicated by Equations 6 to 8, MLTS could theoretically rise to 43, 16 and 12% when the up-flow sludge velocity is set to 1.0, 2.2 and 2.7 mm $s^{-1}$, respectively. Hence, the MLTS in the membrane tank could reach prohibitive values when the concentration in the sludge entering the membrane tank has considerably high values. For instance, when the MLTS entering the membrane tank is 25 g $L^{-1}$ and the up-flow sludge velocity is 2.7 mm $s^{-1}$ (the maximum studied value), the MLTS recycled to the anaerobic reactor is expected to be around 28 g $L^{-1}$.

$$Theoretical\ MLTS_{Outlet} = 1.43 \cdot MLTS_{Inlet} \qquad \text{(Equation 6)}$$

$$Theoretical\ MLTS_{Outlet} = 1.16 \cdot MLTS_{Inlet} \qquad \text{(Equation 7)}$$

$$Theoretical\ MLTS_{Outlet} = 1.12 \cdot MLTS_{Inlet} \qquad \text{(Equation 8)}$$

Figure 4 shows the short-term trials carried out with MLTS of 18 g $L^{-1}$ and an up-flow sludge velocity of 1 mm $s^{-1}$ (i.e. sub-critical conditions), and MLTS of 28 g $L^{-1}$ and an up-flow sludge velocity of 1, 2.2 and 2.7 mm $s^{-1}$ (i.e. critical/supra-critical conditions). Figure 4 shows that $K_0$ remained at values close to 1 when operating membranes sub-critically. In this respect, the reversible fouling rate remained at values lower than 10 mbar $min^{-1}$. Up-flow sludge velocities of less than 1 mm $s^{-1}$ when operating with MLTS of 18 g $L^{-1}$ resulted in critical filtration conditions (data not shown) as a result of the corresponding increase in MLTS in the membrane tank. On the other hand, this figure illustrates that the up-flow sludge velocity had a significant effect on $K_0$ when the membranes were operated at high MLTS levels (around 28 g $L^{-1}$, i.e. critical/supra-critical filtration conditions). For instance, at an up-flow sludge velocity of 1 mm $s^{-1}$, the maximum TMP value was reached at minute 50, so the filtration process was promptly stopped. In this case, a maximum reversible fouling rate of about 90 mbar $min^{-1}$ was observed. On the other hand, when the up-flow sludge velocity was set to 2.7 and 2.2 mm

$s^{-1}$ a maximum reversible fouling rate of around 10 and 20 mbar $min^{-1}$ was achieved, respectively. In both cases, $K_0$ recovered to values lower than 10 mbar $min^{-1}$ after back-flushing. Hence, it was possible to keep the filtration process operating at appropriate TMP values.

These results show that in order to keep the filtration process working properly, the operating up-flow sludge velocity must be selected carefully depending on the operating conditions. When the membranes are operated at high MLTS levels, the up-flow sludge velocity in the membrane tank has to be high enough not only to keep MLTS at suitable levels, but also to minimise the energy consumption needed to keep $J_{20}$ at sub-critical levels (e.g. required $SGD_m$). Nonetheless, up-flow sludge velocity must be minimised in order to maximise energy savings since pumping energy accounts for up to 15 – 20% of all the energy consumed by aerobic MBR technology (Verrecht *et al*., 2010). Hence, it is advisable for the up-flow sludge velocity to be regulated in order to optimise the economic feasibility of HF membranes in full-scale SAnMBR systems.

Another aspect that must be taken into account is whether or not the up-flow sludge is well distributed over the filtration area. A sludge flow distributed evenly across the membrane tank helps remedy any death zones and minimises the likelihood of clogging. Consequently, a minimum up-flow sludge velocity is required to ensure that the sludge is adequately distributed over the filtration area. The configuration of the membrane tank is important in this respect.

*3.2.3. Effect of back-flush frequency*

Several short-term trials were carried out in order to assess the effect of the duration and frequency of the different physical cleaning stages (relaxation and back-flush) on membrane performance. Figure 5 shows the effect of back-flush frequency on membrane permeability with MLTS of 24, 28 and 31.5 g $L^{-1}$. Two different back-flush frequencies were tested: 30 seconds of back-flush every 10 F-R basic cycles (1:10) and 30 seconds of back-flush every 30 F-R basic cycles (1:30) (Figure 5a and Figure 5b, respectively).

Figure 5 shows that at MLTS levels of less than 24 g $L^{-1}$, the $K_0$ performance was independent of the back-flush frequency, at the selected operating conditions. In these short-term trials, the reversible fouling rate remained less than 5 mbar $min^{-1}$, and no residual fouling component was observed. Hence, a complete recovery of $K_0$ was achieved after each relaxation stage. At MLTS levels above 24 g $L^{-1}$, a significant decrease in $K_0$ was detected, making it necessary to increase the back-flush frequency from 1:30 to 1:10 in order to keep the filtration process working below the TMP safety value mentioned earlier. In this respect, when the back-flush frequency was set to 1:30, the maximum reversible fouling rate reached was around 25 and 80 mbar $min^{-1}$, at MLTS levels of 28 and 31.5 g $L^{-1}$, respectively. On the other hand, when the back-flush frequency was increased to 1:10, the maximum reversible fouling rate at an MLTS level of 28 g $L^{-1}$ decreased to values around 20 mbar $min^{-1}$. However, this higher back-flush frequency had no noticeable effect on membrane performance when the MLTS level was 31.5 g $L^{-1}$, i.e. $K_0$ quickly returned to its previous values after back-flushing.

Hence, these results showed that values of MLTS above 30 g $L^{-1}$ are not advisable since increasing the back-flush frequency from 1:30 to 1:10 did not improve the membrane performance. However, at MLTS levels lower than 28 g $L^{-1}$, it was possible to improve membrane performance considerably without significantly increasing the back-flush frequency.

It is a well-known fact that back-flush frequency affects the economic feasibility of the process not only because of the pumping cost but also due to the resulting decrease in the net $J_{20}$. Hence, it is essential to control the back-flush frequency to ensure that the membrane is physically cleaned correctly and thereby maximise the $J_{20}$ at minimum operating costs.

*3.2.4. Effect of relaxation stage duration*

Different sub-critical short-term trials were carried out to assess how the duration of the relaxation stage affects membrane performance. Relaxation stages of 50 and 30 seconds were tested when operating at MLTS of 25 g $L^{-1}$ and sub-critical filtration conditions, and at MLTS of 28 g $L^{-1}$ and supra-critical filtration conditions. An almost complete recovery of $K_0$ was achieved when the MLTS level was 25 g $L^{-1}$. Hence, it was observed that membrane performance was not critically affected by relaxation stages of between 30 and 50 seconds at the selected operating conditions when operating at MLTS levels below 25 g $L^{-1}$. In this case, reversible fouling rates lower than 5 mbar $min^{-1}$ were achieved. On the other hand decreasing the relaxation stage duration from 50 to 30 seconds at MLTS levels of 28 g $L^{-1}$ slightly affected membrane performance, i.e. a slight increase in the reversible fouling component that accumulated on the membrane surface was observed (see Figure 6). In this case, the reversible fouling rate reached values of around 10 mbar $min^{-1}$. However, the TMP recovered to values lower than 0.1 bars after back-flushing.

These results showed that combining relaxation stages with an appropriate back-flush frequency keep TMP stable at quite low values. However, the prolonged accumulation of reversible fouling components upon over the membrane surface could lead to an increased likelihood of irreversible fouling. For that reason, reducing the relaxation stage from 50 to 30 seconds when membranes are operated at MLTS levels higher than 25 g $L^{-1}$ is not recommendable, if the other operating conditions are kept constant (250-second filtration stage, 30 seconds of back-flush every 10 F-R cycles, 10 LMH of $J_{20}$, and $SGD_m$ at 0.23 $Nm^3\ h^{-1}\ m^{-2}$).

Relaxation stage duration affects the economic feasibility of the process because of the resulting decrease in the net $J_{20}$. Hence, it is advisable to control the length of the relaxation stage in order to ensure that the membrane is correctly cleaned physically and thereby minimise the decrease in the net transmembrane flux whilst minimising the operating cost per unit of treated water (e.g. reducing the specific gas demand per volume of permeate).

### *3.2.5. Effect of filtration stage duration*

Figure 7 shows an example of the short-term trials carried out in order to assess the effect of filtration stage duration on membrane performance. Duration was set to 250, 350 and 450 seconds. In this case, the MLTS level in the sludge fed to the membrane tank was 23 and 31.5 g $L^{-1}$ and the back-flush frequency was set to 30 seconds of back-flushing every 10 F-R basic cycles (Figure 7a), and 30 seconds of back-flushing every 30 F-R basic cycles (Figure 7b).

Figure 7 shows that with MLTS of 31.5 g $L^{-1}$, increasing the back-flush frequency from 1:30 to 1:10 did not improve membrane performance. It was not possible to test filtration lasting more than 250 seconds because the maximum TMP value was reached in both 250-second trials.

In the short-term trials carried out with MLTS of 23 g $L^{-1}$, an increase in the filtration stage duration from 250 to 450 seconds resulted in the incomplete removal of the reversible fouling component from the membrane surface. Despite no high reversible fouling rates having been reached, a slight and continuous decrease of $K_0$ over time was observed when the filtration stage duration was set to 450 seconds: an effect that was slightly accentuated when the back-flush frequency was reduced from 1:10 to 1:30.

Increasing the duration of the filtration stage causes an increase in net $J_{20}$. However, as observed in this trial, it is essential to strike a balance between maximising net $J_{20}$ and minimising maintenance and operating costs.

*3.2.5. Overall effect of MLTS and sustainable operating MLTS level*

On the basis of the results shown in this study, we established a critical value for long-term membrane operating of around 20 – 25 g $L^{-1}$. Since several operating variables considerably affect the appearance of reversible fouling at short-term, this maximum operating MLTS was established for the following scenario: 300-second basic F-R cycles (250 s filtration and 30-50 s relaxation) with 30 seconds of back-flush every 10 cycles; $J_{20}$ of about 10 LMH; $SGD_m$ of 0.23 $Nm^3\ h^{-1}\ m^{-2}$; and up-flow sludge velocity of 2.7 mm $s^{-1}$. For this specific scenario, increasing the filtration stage duration over 250 s will lead to a progressively reduction in $K_0$ over time, being this effect greater when the back-flush frequency is decreased to 1 back-flush every 30 F-R cycles (see Figure 7). On the other hand, back-flushing can be decreased from a frequency of 1 back-flush every 10 F-R cycles to 1 back-flush every 30 F-R cycles when membranes are operated at MLTS levels lower that 24 g $L^{-1}$, whilst the same increase considerably affects $K_0$ when the MLTS is around 28 g $L^{-1}$ (see Figure 5). As regards gas sparging intensity, an $SGD_m$ of around 0.17 – 0.23 $Nm^3\ h^{-1}\ m^{-2}$ allows maintaining proper long-term operation when the MLTS entering the membrane tank is around 20 g $L^{-1}$, whilst this value is not enough to properly operate at MLTS levels around 28 g $L^{-1}$ (see Figure 3). Finally, reducing the up-flow sludge velocity from 2.7 to 2.2 mm $s^{-1}$ may lead to a considerable decrease in $K_0$, due to a significant increase in the MLTS level when its concentration entering the membrane tank is around 28 g $L^{-1}$ (see Figure 4).

Functioning with similar operating modes may allow reducing the offset of irreversible/irrecoverable fouling at long-term operation since the accumulation of reversible fouling component over the membrane surface can be minimised.

### *3.3. Overall membrane operation compared to full-scale aerobic MBR plant.*

On the basis of the long-term results obtained in this work, MLTS levels above 25 g $L^{-1}$ are not recommended for commercial HF membranes because $J_C$ drops to less than 10 LMH, making the filtration process unnecessarily expensive. On the basis of the short-term results, two opposite design strategies could be applied depending on the operating regime adopted. If the design strategy is based on sub-critical operating, the installed filtration area must be increased – which increases the initial investment. On the other hand, if the design strategy selected is based on supra-critical operating, then high back-flush frequencies and/or unsustainable $SGD_m$ are required – which increases operating and maintenance/replacement costs. This may result in low process efficiency per unit of treated water (i.e. a decrease in net $J_{20}$) or high energy consumption, respectively.

The long-term membrane performance shown in this study, demonstrates that working at sub-critical filtration conditions is an adequate operating strategy for SAnMBR technology because no considerable irreversible/irrecoverable fouling component was detected after operating for almost two years. Indeed, membranes did not required chemical cleaning. Nevertheless, an exhaustive economic analysis is needed to accurately demonstrate the feasibility of working at sub-critical or critical/supra-critical levels in an specific scenario. However, in order to shed more light upon the economic feasibility of SAnMBR technology for treating urban wastewater, the long-term operating strategy proposed in our study is compared in tables 2 and 3 with some available data related to full-scale aerobic MBR operations.

*3.3.1. Average operating values for transmembrane flux, membrane permeability and specific gas demand*

Table 2 shows a summary of data for full-scale aerobic plants treating both urban and industrial wastewater with submerged MBR (extracted from Judd and Judd, 2011) and the average values obtained throughout the long-term operation of our study.

Using FS membranes in urban wastewater treatment enables higher transmembrane fluxes (19.4 LMH) and membrane permeability (261 LMH $bar^{-1}$) in comparison with the results obtained in our work: transmembrane fluxes of around 11 LMH, resulting in membrane permeability of 135 LMH $bar^{-1}$ in average. However, higher $SGD_m$ and higher specific gas demand with respect to permeate volume ($SGD_P$) are commonly required in FS technology (see Table 2). It is well known that HF technology allows some degree of lateral movement which enables greater cake layer detachment at lower gas sparging intensities than in FS technology. On the other hand, when using FS membranes to treat industrial wastewater, commonly operating at high MLTS levels, the transmembrane fluxes and membrane permeability are similar to those obtained in our study. However, considerably higher $SGD_m$ and $SGD_P$ are reported when using FS membranes.

As regards HF technology, the results from full-scale aerobic operation are similar to the results obtained in our study. Both aerobic and anaerobic operation result in reasonably adequate transmembrane fluxes and membrane permeability by applying low air/gas demands. In this respect, even though the resulting $J_{20}$ was lower in the case of anaerobic HF membranes (around 11 LMH vs. approx. 17 LMH), higher $K_{20}$ levels were obtained (around 135 LMH $bar^{-1}$ vs. approx. 75 LMH $bar^{-1}$) whilst applying similar $SGD_m$ (around 0.25 $Nm^3\ h^{-1}\ m^{-2}$) in both anaerobic and aerobic HF. Moreover, $SGD_P$ remained in a similar range (from approx. 15.5 to approx. 20.5) in both cases. Some studies suggested that the cake layer formed with aerobic and anaerobic sludge might have different removability (see e.g. Meng *et al.*, 2009). Nevertheless, based on comparable results for the aerobic and anaerobic operation of HF membranes, it can be assumed in this study that differences between anaerobic and aerobic sludge properties (i.e. particle size distribution, EPS, SMP and biomass concentration, etc.) did not critically determine the removability of the cake layer from the membrane surface. In this respect, HF technology is a promising, competitive technology for the anaerobic treatment of urban wastewater.

### *3.3.2. Physical and chemical cleaning requirements*

A summary of the physical cleaning protocols for full-scale aerobic MBRs treating urban wastewater (extracted from Judd and Judd, 2011) and the average values applied throughout the long-term operation of our study are shown in table 3. Full-scale results from aerobic MBR technology reveal a relaxation downtime of around 10% of the operating time in both FS and HF configurations. This value is significantly lower than the resulting relaxation downtime obtained in our study (around 16.7% of the operating time). However, the relaxation stage duration applied in our work can be considered as a quite conservative value – selected in order to avoid possible problems when operating at large MLTS concentrations – since the results from the short-term trials showed that it is possible to reduce this parameter to a value of 30 seconds with a minimum impact on membrane performance when operating at MLTS levels below 25 g $L^{-1}$. This decrease in the duration of the relaxation stage results in a downtime of around 10% of operating time, which is similar to the average downtime shown in table 3 for full-scale aerobic MBRs. On the other hand, a back-flush downtime of around 6 – 9% of operating time was reported by Judd and Judd (2011) in the aerobic treatment of urban wastewater in full-scale MBRs. In this respect, only an additional downtime of around 1% of the operating period was obtained in our study (carried out with a back-flush frequency of 0.5 min every 10 F-R cycles). This gives a total average downtime for physical cleaning of 17.7% of operating time throughout the long-term operation of HF membranes shown in our study (instead of an average downtime of around 16 – 19% when using HF technology to treat urban wastewater aerobically). Moreover, it is important to emphasise that the membranes in our study did not require chemical cleaning after operating for more than two years – despite operating at high MLTS levels and with temperature shocks that affected mixed sludge properties – which is a considerably longer than the periods usually employed in aerobic MBR technology.

Hence, the results of our study predict that HF membranes will result in a sustainable approach for SAnMBR technology compared to the full-scale results reported for aerobic MBR technology.

## 4. Conclusions

The membrane performance demonstrated that HF-SAnMBR may be a promising technology for urban wastewater treatment since low maintenance and operating costs related to membrane separation process can be achieved. According to the results, gas sparging intensity and back-flush frequency are the physical variables that affect membrane performance most. In our study, low gas sparging intensities (around 0.23 $Nm^3\ h^{-1}\ m^{-2}$) and low BF frequencies (30 seconds of BF every 10 basic F-R cycle) were enough to operate membranes sub-critically even at high levels of MLTS (up to 25 $g\ L^{-1}$). On the other hand, operating at critical filtration conditions involves significant physical cleaning (gas sparging intensity and BF frequency) to ensure that membranes operate correctly. The results of our study show that establishing a suitable physical cleaning schedule (relaxation, back-flush and gas sparging intensity) enhances the removal of the reversible fouling component accumulated on the membrane surface, and thus minimises the irreversible fouling propensity. After more than two years of sub-critical operation (transmembrane flux between 9 and 13.3 LMH at gas sparging intensities of around 0.23 $Nm^3\ h^{-1}\ m^{-2}$ and MLTS levels in the mixed liquor entering the membrane tank of around 10 to 30 $g\ L^{-1}$) no significant irreversible/irrecoverable fouling problems were detected (membrane permeability remained above 100 LMH $bar^{-1}$ and total filtration resistance remained below $10^{13}\ m^{-1}$), thus no chemical cleaning was conducted. Membrane performance was similar to the aerobic HF membranes operated in full-scale MBR plants. On the basis of the different experiments carried out, different control strategies will be developed with a view to optimising membrane performance in both sub-critical and critical/supra-critical operating. Nevertheless, an exhaustive economic analysis is needed to make the best choice between the two different operating regimes in a specific scenario: working at sub-critical levels or critical/supra-critical levels.

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## Figure and table captions

**Table 1.** Short-term trials operating conditions (Nomenclature: **MLTS**: mixed liquor total solid; **$SGD_m$**: specific gas demand per membrane area; **BF**: back-flush; **F-R**: filtration-relaxation)

**Table 2.** Summary of full-scale plant data for urban wastewater treatment: Average data for submerged MBRs (adapted from Judd and Judd (2011)). Nomenclature: **FS**: Flat-sheet; **HF**: hollow-fibre; **WW**: wastewater; **J**: transmembrane flux; **K**: membrane permeability; **$S(A/G)D_m$**: specific air/gas demand per membrane area; **$S(A/G)D_P$**: specific air/gas demand per permeate volume.

**Table 3.** Summary of full-scale urban physical cleaning protocols. Average data on submerged MBRs (adapted from Judd and Judd (2011)). Nomenclature: **FS**: Flat-sheet; **HF**: hollow-fibre; **WW**: wastewater; **F**: Filtration stage duration; **R**: Relaxation stage duration; **BF**: Back-flush stage duration.

**Figure 1.** Flow diagram of the pilot plant. Nomenclature: **RF**: rotofilter; **ET**: equalization tank; **AnR**: anaerobic reactor; **MT**: membrane tanks; **DV**: degasification vessel; **CIP**: clean-in-place; **P**: pump; and **B**: blower.

**Figure 2.** Long-term operation: evolution of **(a)** $K_{20}$ and MLTS; and **(b)** $R_T$ and MLTS. Experimental periods: **(i)** $J_{20}$ of 13.3 LMH and temperature of 33 ºC; **(ii)** $J_{20}$ = 10 LMH and T= 33 ºC; **(iii)** $J_{20}$ = 12 LMH and T= 25 ºC; **(iv)** $J_{20}$ = 13.3 LMH and T= 20 ºC; and **(v)** $J_{20}$ = 11 LMH and ambient temperature (spring and summer, from about 20 to 30 ºC); and **(vi)** $J_{20}$ = 9 LMH and ambient temperature (autumn and winter, from about 30 to 15 ºC).

**Figure 3.** Short-term trial 1: Effect of gas sparging intensity on membrane permeability at MLTS level of 28, 30 and 31.5 g $L^{-1}$. Nomenclature: **MLTS**: mixed liquor total solids; **$K_0$**: unit-normalised membrane permeability; **BF**: back-flush.

**Figure 4.** Short-term trial 2: Effect of up-flow sludge velocity on membrane permeability at MLTS levels of 18 and 28 g $L^{-1}$, and up-flow sludge velocity of 1.0, 2.2, and 2.7 mm $s^{-1}$. Nomenclature: **MLTS**: mixed liquor total solids; **TS**: total solids; **$K_0$**: unit-normalised membrane permeability; **BF**: back-flush.

**Figure 5**. Short-term trial 3: Effect of back-flush frequency on membrane permeability at MLTS of 24, 28 and 31.5 g $L^{-1}$ and **(a)** 30 seconds of back-flush every 10 F-R cycles; and **(b)** 30 seconds of back-flush every 30 F-R cycles. Nomenclature: **$K_0$**: unit-normalised membrane permeability; **BF**: back-flush.

**Figure 6.** Short-term trial 4: Effect of relaxation stage duration on membrane permeability at MLTS level

of 28 g $L^{-1}$. Nomenclature: $\mathbf{K_0}$: unit-normalised membrane permeability; **BF**: back-flush.

**Figure 7.** Short-term trial 5: Effect of filtration stage duration on membrane permeability at **(a)** MLTS levels of 23 and 31.5 g $L^{-1}$ and back-flush frequency of 1 back-flush every 10 F-R cycles; and **(b)** MLTS levels of 23 and 31.5 g $L^{-1}$ and back-flush frequency of 1 back-flush every 30 F-R cycles. Nomenclature: $\mathbf{K_0}$: unit-normalised membrane permeability; **BF**: back-flush.

Table 1
Click here to download Table: Table 1.doc

**Table 1.** Short-term trials operating conditions (Nomenclature: **MLTS**: mixed liquor total solid; **$SGD_m$**: specific gas demand per membrane area; **BF**: back-flush; **F-R**: filtration-relaxation)

| Trial | Variable studied | Sub-critical conditions | | Supra-critical/Critical conditions | |
|---|---|---|---|---|---|
| | | Value | MLTS ($g\ L^{-1}$) | Value | MLTS ($g\ L^{-1}$) |
| 1 | $SGD_m$ ($Nm^3\ h^{-1}\ m^{-2}$) | 0.17, 0.23, 0.3, 0.4 | 20 | 0.17 | 30 |
| | | | | 0.23 | 28, 31.5 |
| 2 | Up-flow sludge velocity ($mm\ s^{-1}$) | 1.3 | 18 | 1.0, 2.2, 2.7 | 28 |
| 3 | BF frequency (BF:F-R) | 1:10, 1:30 | 24 | 1:10, 1:30 | 28, 31.5 |
| 4 | Relaxation stage duration (seconds) | 30, 50 | 25 | 30, 50 | 28 |
| 5 | Filtration stage duration (seconds) | 250, 350, 450 (1BF:10F-R) | 23 | 250 (1BF:10F-R) | 31.5 |
| | | 250, 350, 450 (1BF:30F-R) | | 250 (1BF:30F-R) | |

**Table 2.** Summary of full-scale plant data for urban wastewater treatment: Average data for submerged MBRs (adapted from Judd and Judd (2011)). Nomenclature: **FS**: Flat-sheet; **HF**: hollow-fibre; **WW**: wastewater; **J**: transmembrane flux; **K**: membrane permeability; **$S(A/G)D_m$**: specific air/gas demand per membrane area; **$S(A/G)D_P$**: specific air/gas demand per permeate volume.

| **Technology** | **Treatment** | **J (LMH)** | **K (LMH $bar^{-1}$)** | **$S(A/G)D_m$ ($Nm^3$ $h^{-1}$ $m^{-2}$)** | **$S(A/G)D_P$** |
|---|---|---|---|---|---|
| FS | Aerobic; Urban WW | 19.4 | 261 | 0.57 | 27.5 |
| FS | Aerobic; Industrial WW | 13.4 | -- | 0.80 | 91.9 |
| HF | Aerobic; Urban WW | 19.5 | 104 | 0.30 | 15.4 |
| HF | Aerobic; Industrial WW | 15.4 | 47 | 0.23 | 16.5 |
| This study (HF) | Anaerobic; Urban WW | 11.1 | 135 | 0.23 | 20.7 |

**Table 3.** Summary of full-scale urban physical cleaning protocols. Average data on submerged MBRs (adapted from Judd and Judd (2011)). Nomenclature: **FS**: Flat-sheet; **HF**: hollow-fibre; **WW**: wastewater; **F**: Filtration stage duration; **R**: Relaxation stage duration; **BF**: Back-flush stage duration.

| **Technology** | **Treatment** | **F (min)** | **R (min)** | **BF (min)** |
|---|---|---|---|---|
| FS | Aerobic; Urban WW | 22.0 | 2.2 | --- |
| HF | Aerobic; Urban WW | 10.0 | 1.0 | 0.43 |
| This study (HF) | Anaerobic; Urban WW | 4.2 | 0.8 | 0.50 |

Figure 1

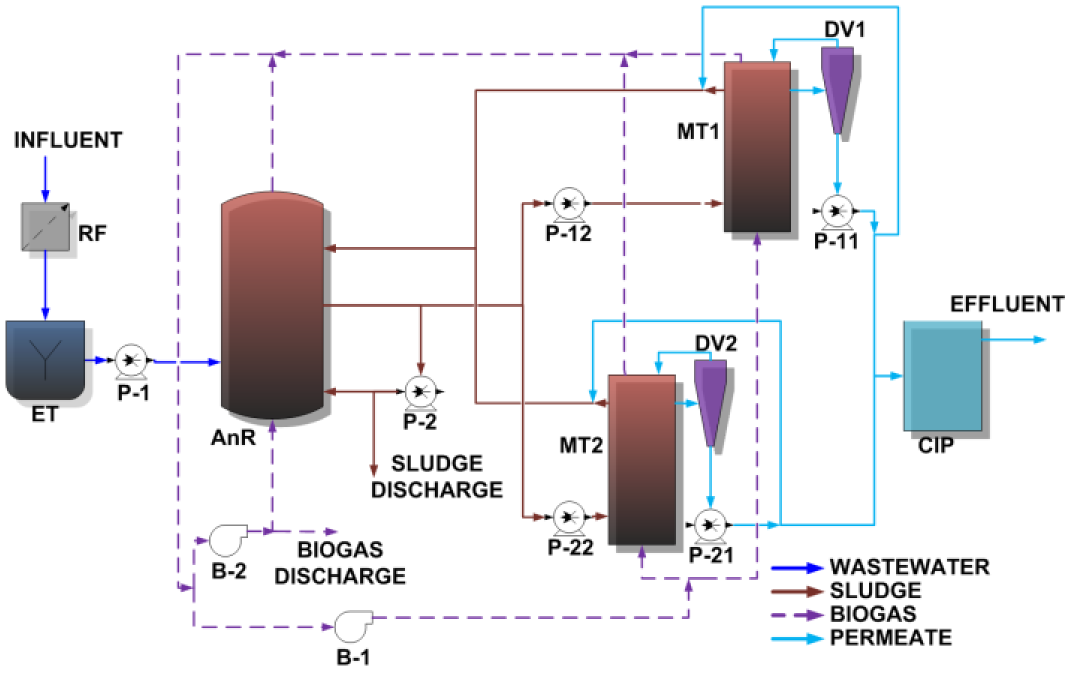
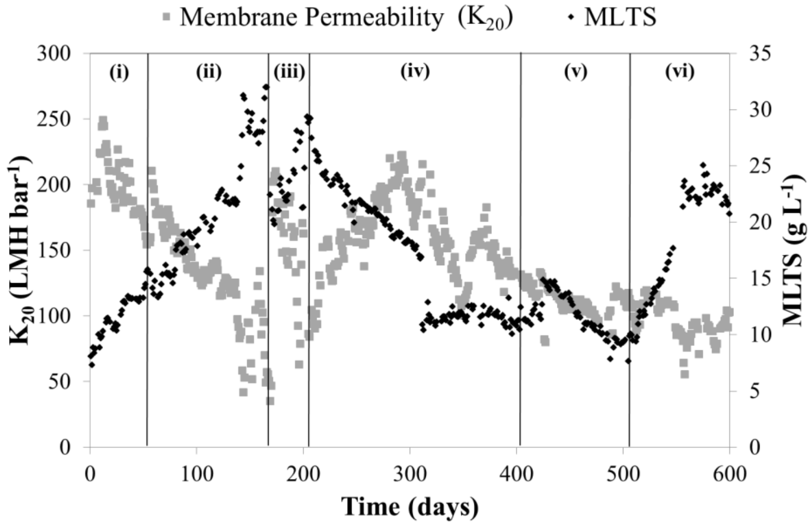
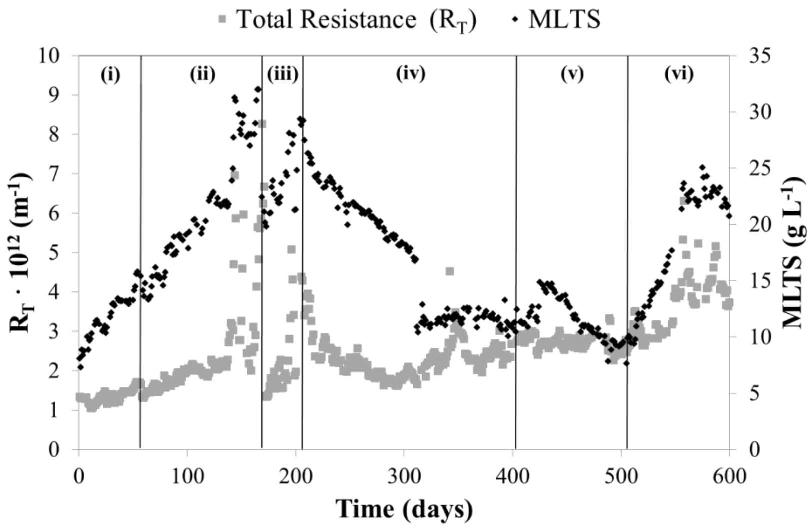



**Figure 1.** Flow diagram of the pilot plant. Nomenclature: **RF**: rotofilter; **ET**: equalization tank; **AnR**: anaerobic reactor; **MT**: membrane tanks; **DV**: degasification vessel; **CIP**: clean-in-place; **P**: pump; and **B**: blower.

Figure 2




(a)




(b)

**Figure 2.** Long-term operation: evolution of **(a)** $K_{20}$ and MLTS; and **(b)** $R_T$ and MLTS. Experimental periods: **(i)** $J_{20}$ of 13.3 LMH and temperature of 33 ºC; **(ii)** $J_{20}$ = 10 LMH and T= 33 ºC; **(iii)** $J_{20}$ = 12 LMH and T= 25 ºC; **(iv)** $J_{20}$ = 13.3 LMH and T= 20 ºC; and **(v)** $J_{20}$ = 11 LMH and ambient temperature (spring and summer, from about 20 to 30 ºC); and **(vi)** $J_{20}$ = 9 LMH and ambient temperature (autumn and winter, from about 30 to 15 ºC).

Figure 3

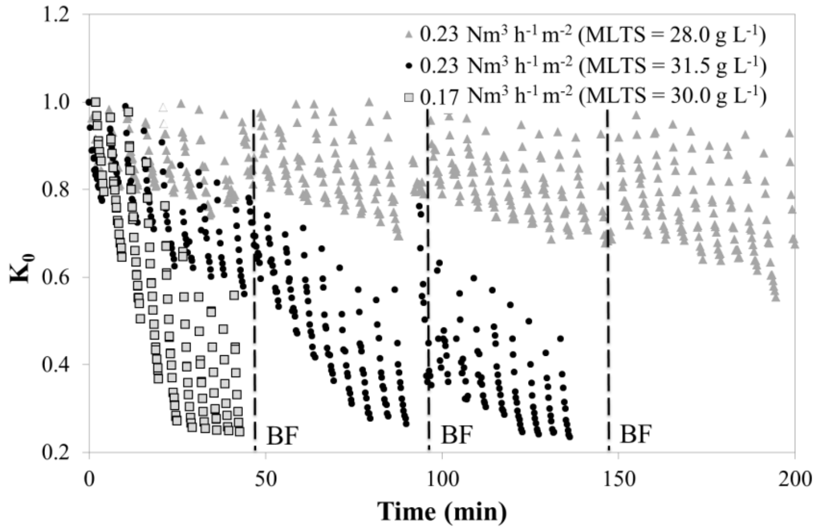



**Figure 3.** Short-term trial 1: Effect of gas sparging intensity on membrane permeability at MLTS level of 28, 30 and 31.5 g $L^{-1}$. Nomenclature: **MLTS**: mixed liquor total solids; $\mathbf{K_0}$: unit-normalised membrane permeability; **BF**: back-flush.

**Figure 4**

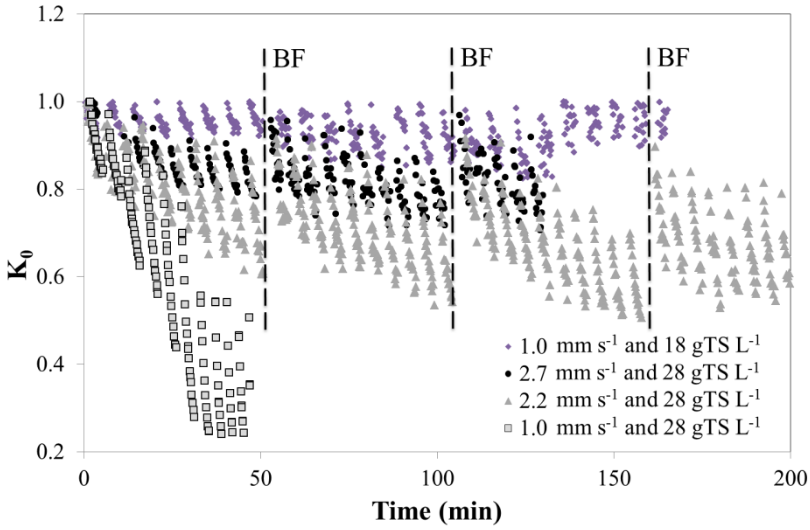
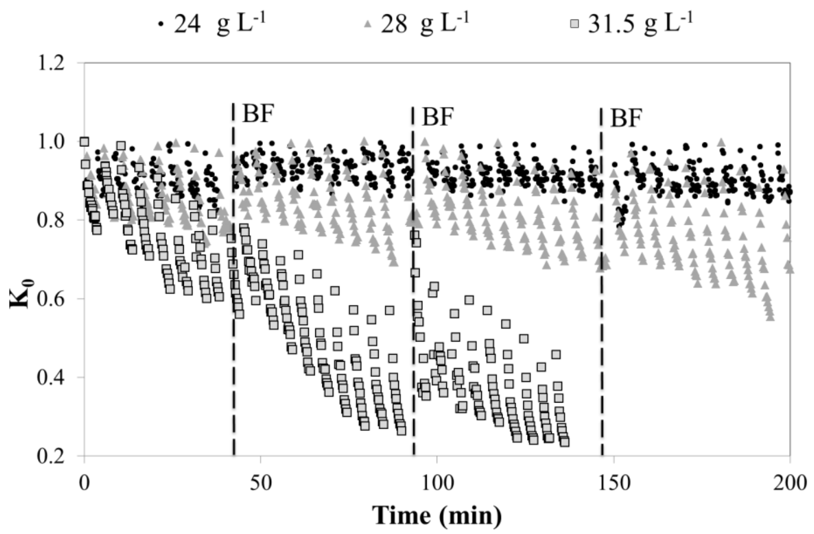
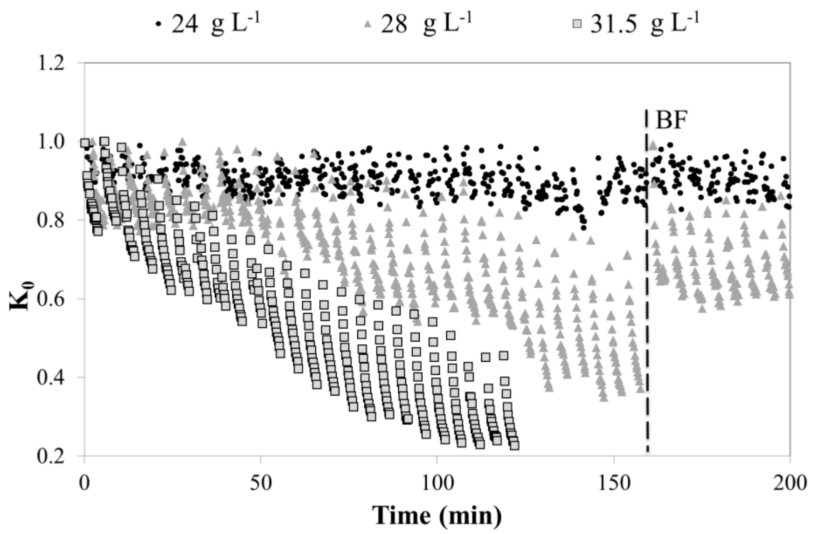



**Figure 4.** Short-term trial 2: Effect of up-flow sludge velocity on membrane permeability at MLTS levels of 18 and 28 g $L^{-1}$, and up-flow sludge velocity of 1.0, 2.2, and 2.7 mm $s^{-1}$. Nomenclature: **MLTS**: mixed liquor total solids; **TS**: total solids; **$K_0$**: unit-normalised membrane permeability; **BF**: back-flush.

Figure 5




**(a)**




**(b)**

**Figure 5**. Short-term trial 3: Effect of back-flush frequency on membrane permeability at MLTS of 24, 28 and 31.5 g $L^{-1}$ and **(a)** 30 seconds of back-flush every 10 F-R cycles; and **(b)** 30 seconds of back-flush every 30 F-R cycles. Nomenclature: $\mathbf{K_0}$: unit-normalised membrane permeability; **BF**: back-flush.

**Figure 6**

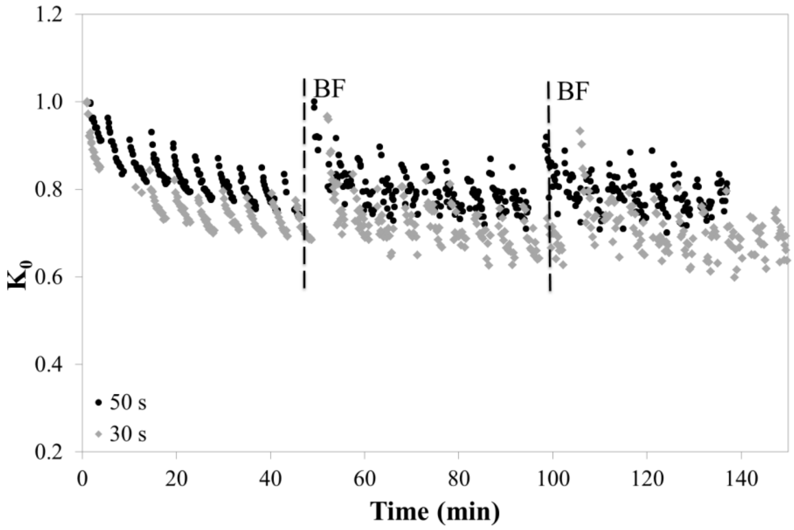



**Figure 6.** Short-term trial 4: Effect of relaxation stage duration on membrane permeability at MLTS level of 28 g $L^{-1}$. Nomenclature: $\mathbf{K_0}$: unit-normalised membrane permeability; **BF**: back-flush.

**Figure 7**

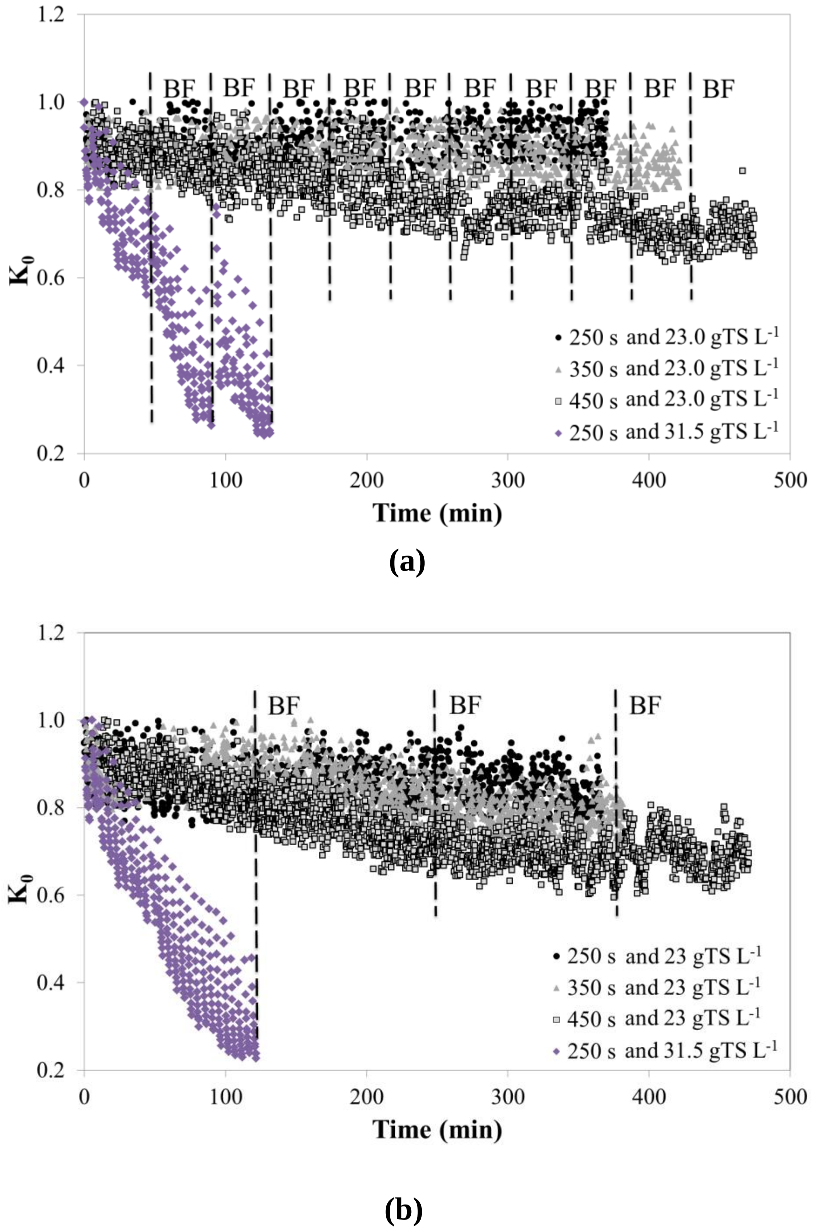



**Figure 7.** Short-term trial 5: Effect of filtration stage duration on membrane permeability at **(a)** MLTS levels of 23 and 31.5 g $L^{-1}$ and back-flush frequency of 1 back-flush every 10 F-R cycles; and **(b)** MLTS levels of 23 and 31.5 g $L^{-1}$ and back-flush frequency of 1 back-flush every 30 F-R cycles. Nomenclature: $\mathbf{K_0}$: unit-normalised membrane permeability; **BF**: back-flush.