The operating cost of an anaerobic membrane bioreactor (AnMBR) treating sulphate-rich urban wastewater

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Abstract

The objective of this study was to evaluate the operating cost of an anaerobic membrane bioreactor (AnMBR) treating sulphate-rich urban wastewater (UWW) at ambient temperature (ranging from 17 to 33ºC). To this aim, energy consumption, methane production, and sludge handling and recycling to land were evaluated. The results revealed that optimising specific gas demand with respect to permeate volume (SGD_P) and sludge retention time (for given ambient temperature conditions) is essential to maximise energy savings (minimum energy demand: 0.07 $kWh \cdot m^{-3}$). Moreover, low/moderate sludge productions were obtained (minimum value: 0.16 kg $TSS \cdot kg^{-1}$ $COD_{REMOVED}$), which further enhanced the overall operating cost of the plant (minimum value: €0.011 per m^3 of treated water). The sulphate content in the influent UWW significantly affected the final production of methane and thereby the overall operating cost. Indeed, the evaluated AnMBR system presented energy surplus potential when treating low-sulphate UWW.

Keywords

Energy consumption; industrial-scale hollow-fibre membranes; operating cost; anaerobic membrane bioreactor (AnMBR); sulphate-rich urban wastewater.

1. Introduction

Nowadays, a key issue in global sustainable development is the dependency on fossil fuels for electricity production, which represents up to the 80% of the global energy consumption [1]. In this respect, electricity consumption is a key element in the overall environmental performance of a wastewater treatment plant (WWTP) [2]. Hence, it is particularly important to implement new energy-saving technologies that reduce the overall energy balance of the WWTP, such as anaerobic membrane bioreactors (AnMBRs). This technology focuses on the sustainability benefits of anaerobic processes compared to aerobic processes, such as: minimum sludge production due to low biomass yield of anaerobic organisms; low energy demand since no aeration is required; and methane production that can be used to fulfil process energy requirements [3].

Several issues have been recognised elsewhere as potential drawbacks which may affect the sustainability of AnMBR technology treating urban wastewater (UWW). One key issue is the competition between Methanogenic *Archaea* (MA) and Sulphate Reducing Bacteria (SRB) for the available substrate [4] when there is significant sulphate content in the influent, reducing therefore the available COD for methanisation [5]. For urban wastewater, which can easily present low COD/SO_4–S ratio, this competition can critically affect the amount and quality of the biogas produced. Specifically, 2 kg of COD are consumed by SRB in order to reduce 1 kg of influent SO_4-S (see, for instance,

[5]). According to the theoretical methane yield under standard temperature and pressure conditions (350 $L_{CH4}\cdot kg^{-1}COD$), SRB reduces the production of approx. 700 L of methane per kg of influent SO_4-S (considering reduction of all sulphate to sulphide). Therefore, higher biogas productions would be achieved when there is little sulphate content in the influent (typical sulphate concentration in UWW fluctuates around 7-17 mg SO_4-$S\cdot L^{-1}$ [6]). On the other hand, due to the low-growth rate of anaerobic microorganism, high sludge retention times (SRTs) are required when operating at low temperatures in order to achieve suitable organic matter removal rates, especially for low-strength wastewaters like urban ones (typical COD levels below 1 $g\cdot L^{-1}$ [6]). However, as regards filtration process, operating AnMBRs at high SRT may imply operating at high mixed liquor total solid (MLTS) levels. This is considered to be one of the main constraints on membrane operating because it can result in a high membrane fouling propensity and therefore high energy demand for membrane scouring by gas sparging [7].

The objective of this study was to evaluate the operating cost of an AnMBR system treating sulphate-rich urban wastewater (UWW) at ambient temperature (ranging from 17 to 33ºC). To this aim, power requirements, energy recovery from methane (biogas methane and/or methane dissolved in the effluent), and sludge handling and recycling to land were evaluated at different operating conditions. In order to obtain reliable results that can be extrapolated to full-scale plants, this study was carried out in an AnMBR using industrial-scale hollow-fibre membrane units. This system was operated using effluent from the pre-treatment of the Carraixet WWTP (Valencia, Spain).

2. Materials and methods

2.1. AnMBR plant description

A semi-industrial AnMBR plant was operated using the effluent of a full-scale WWTP pre-treatment. The average AnMBR influent characteristics are shown in Table 1. This influent UWW was characterised by a low COD (around 650 $mg \cdot L^{-1}$) and high sulphate concentration (around 105 mg SO_4-$S \cdot L^{-1}$).

The AnMBR plant consists of an anaerobic reactor with a total volume of 1.3 m^3 connected to two membrane tanks (MT1 and MT2) each one with a total volume of 0.8 m^3. Each membrane tank includes one ultrafiltration hollow-fibre membrane commercial system (PURON®, Koch Membrane Systems, 0.05 µm pore size, 30 m^2 total filtering area). The filtration process was studied from experimental data obtained from MT1 (operated recycling continuously the obtained permeate to the system), whilst the biological process was studied using experimental data obtained from MT2 (operated for the biological process without recycling the obtained permeate). Hence, different 20 ºC-standardised transmembrane fluxes (J_{20}) were tested in MT1, without affecting the hydraulic retention time (HRT) of the plant.

In addition to conventional membrane operating stages (filtration, relaxation and back-flushing), two additional stages were considered in the membrane operating mode: degasification and ventilation. Further details on this AnMBR can be found in Giménez et al. [5] and Robles et al. [8].

2.2. AnMBR operating conditions

The AnMBR plant was operated for around 920 days within a wide range of operating conditions for both filtration and biological process.

2.2.1 Filtration process

Five operating scenarios related to filtration process (FP1-FP5) were considered to evaluate the energy consumption of the AnMBR plant (see Table 2). As Table 2 shows, the main operating conditions in these five scenarios were as follows: transmembrane pressure (TMP) during filtration: from 0.09 to 0.35 bar; J_{20} from 9 to 20 LMH; MLTS entering the membrane tank: from 12.5 to 32.5 $g \cdot L^{-1}$; sludge recycling flow in anaerobic reactor and membrane tank (SRF_{MT} and SRF_{AnR} respectively): 2.7 and 1 $m^3 \cdot h^{-1}$ respectively; specific gas demand per square metre of membrane area (SGD_m): controlled at 0.17 and 0.23 $m^3 \cdot h^{-1} \cdot m^{-2}$; and biogas recycling flow to the anaerobic reactor (BRF_{AnR}): 1.5 $m^3 \cdot h^{-1}$.

2.2.2. Biological process

Variations in SRT and seasonal temperature were studied to account for the dynamics in methane and sludge productions over time. During the 920-day experimental period the plant was operated at ambient temperature ranging from 17 to 33 ºC and SRT varied from 30 to 70 days. Three different experimental scenarios related to biological process ($BP_{33°C,\ SRT\ 70days}$, $BP_{22°C,\ SRT\ 38days}$ and $BP_{17°C,\ SRT\ 30days}$) were considered to evaluate the energy consumption of the AnMBR plant (see Table 3): (1) a summer period of two months of operation resulting in high methane and low sludge productions ($BP_{33°C,\ SRT\ 70days}$) due to operating at high temperature (33 °C in average) and high SRT (70 days);

(2) one year of operation resulting in moderate methane and sludge productions ($BP_{22°C, SRT\ 38days}$) due to operating at variable temperature (22 °C in average) and moderate SRT (38 days); and (3) a winter period of two months of operation resulting in low methane and moderate sludge productions ($BP_{17°C, SRT\ 30days}$) due to operating at relatively low temperature (17.1 °C in average) and moderate SRT (30 days). These three scenarios represent boundary ($BP_{33°C, SRT\ 70days}$: best conditions; and $BP_{17°C, SRT\ 30days}$: worst conditions) and average ($BP_{22°C, SRT\ 38days}$) of the operating conditions evaluated in the plant.

In addition, several simulation scenarios were calculated in order to assess the AnMBR performance within the whole range of temperature (17-33 ºC) and SRT (30-70 days) evaluated in this study. Simulation results were obtained using the WWTP simulating software DESASS [9]. This simulation software features the mathematical model BNRM2 [10], which was previously validated using experimental data obtained in the AnMBR plant. Figure 1 shows the resulting effluent COD without including dissolved methane concentration (see Figure 1a); total methane production (see Figure 1b); and sludge production (Figure 1c) for the different temperature and SRT conditions simulated.

2.2.2.1. Influent sulphate concentration

The effect of the influent sulphate on the AnMBR operating cost was also evaluated. As mentioned before, the UWW fed to the AnMBR plant was characterised by relatively low COD and high sulphate concentrations (see Table 1). Therefore, an important fraction of the influent COD was consumed by SRB. To be precise, the sulphate content in the influent was approx. 105 mg $S\text{-}SO_4\cdot L^{-1}$, from which approx. 98% was reduced to hydrogen sulphide (around 103 mg $S\text{-}SO_4\cdot L^{-1}$). Therefore, about 206 $mg\cdot L^{-1}$ of influent COD were consumed by SRB.

The results obtained in this study were compared to the theoretical results obtained in an AnMBR system treating low-sulphate UWW (10 mg $S\text{-}SO_4\cdot L^{-1}$). To this aim, the methane production when treating low-sulphate UWW was calculated on the basis of the theoretical methane yield under standard temperature and pressure conditions: 350 $L_{CH4}\cdot kg^{-1}$COD. Table 4 shows the theoretical methane production (including both biogas methane and methane dissolved in the effluent) obtained for cases $BP_{33°C,\ SRT\ 70days}$, $BP_{22°C,\ SRT\ 38days}$ and $BP_{17°C,\ SRT\ 30days}$ when treating low-sulphate UWW (10 mg $S\text{-}SO_4\cdot L^{-1}$). The distribution between gas and liquid phase of the produced methane was established on the basis of the experimental distribution obtained in the AnMBR plant.

2.3. Analytical monitoring

The following parameters were analysed in mixed liquor and influent stream according to Standard Methods [11]: total solids (TS); total suspended solids (TSS); volatile suspended solids (VSS); sulphate (SO_4-S); nutrients (ammonium (NH_4-N) and orthophosphate (PO_4-P)); and chemical oxygen demand (COD). The methane fraction of the biogas was measured using a gas chromatograph equipped with a Flame Ionization Detector (GC-FID, Thermo Scientific) in accordance with Giménez et al. [5]. The dissolved methane fraction of the effluent was determined in accordance with Giménez et al. [12]. AMPTS® (Automatic Methane Potential Test System, Bioprocess Control) was employed for evaluating the biochemical methane potential (BMP) of the wasted sludge. Due to the low microbial activity of this sludge, BMP tests were inoculated using biomass coming from the anaerobic digester of the Carraixet WWTP. VSS and TSS levels in the wasted sludge were measured at the beginning and at the end of the BMP test, allowing the percentage of biodegradable volatile suspended solids (%BVSS) to be calculated. In this study, the sludge stabilisation criterion was set to 35% of BVSS.

2.4. Energy balance description

The energy balance of the AnMBR system consisted of: power requirements (W), and energy recovery from both biogas methane (E_{biogas}) and methane dissolved in the effluent ($E_{dissolved\ methane}$). The heat energy term (Q) was assumed negligible since the process was evaluated at ambient temperature conditions.

Therefore, the AnMBR energy consumption was evaluated in this study assuming the following terms: (1) energy consumption when non-capture of methane is considered; (2) net energy consumption including energy recovery from biogas methane; and (3) net energy consumption including energy recovery from both biogas methane and methane dissolved in the effluent.

The equipment considered in the W term consisted of the following: one anaerobic reactor feeding pump; one membrane tank sludge feeding pump; one anaerobic reactor sludge mixing pump; one permeate pump; one anaerobic reactor biogas recycling blower; one membrane tank biogas recycling blower; one rotofilter; and one dewatering system.

The energy requirements for each of the scenarios evaluated in this study were calculated using the simulation software DESASS, which includes a general tool that enables calculating the energy consumption of the different units comprising a WWTP.

2.4.1. Power requirements (W)

As proposed by Judd and Judd [13], the energy consumption related to pumps and blowers (adiabatic compression), was calculated by applying the corresponding theoretical equations (Equations 1, 2 and 3, respectively).

$$P_B\left(\frac{J}{S}\right)=\frac{(\mathrm{M}\cdot\mathrm{R}\cdot\mathrm{T_{gas}})}{(\alpha-1)\cdot\eta_{blower}}\left[\left(\frac{p_2}{p_1}\right)^{\frac{\alpha-1}{\alpha}}-1\right] \qquad \text{(Equation 1)}$$

where P_B is the blower power requirement (adiabatic compression), M (mol·s^{-1}) is the molar flow rate of biogas, R (J·mol^{-1}·K^{-1}) is the gas constant for biogas, P_1 (atm) is the absolute inlet pressure, P_2 (atm) is the absolute outlet pressure, T_{gas} (K) is the biogas temperature, α is the adiabatic index and η_{blower} is the blower efficiency.

P_1 and M were taken from the data obtained in the AnMBR plant; P_2 and T_{gas} were calculated by the simulation software; and a value of 0.8 was considered for η_{blower} as a theoretical typical value.

$$P_g\left(\frac{J}{S}\right)=q_{imp}\cdot\rho_{liquor}\cdot g\cdot\frac{\left\{\left[\left(\frac{(\mathrm{L}+\mathrm{Leq})\cdot\mathrm{f}\cdot\mathrm{V}^2}{D\cdot 2\cdot g}\right)\mathrm{asp.}+\left(\frac{(\mathrm{L}+\mathrm{Leq})\cdot\mathrm{f}\cdot\mathrm{V}^2}{D\cdot 2\cdot g}\right)\mathrm{imp.}\right]+\left[Z_1-Z_2\right]\right\}}{\eta_{pump}} \qquad \text{(Equation 2)}$$

where P_g is the power requirement by the general pump, considering both pump aspiration and pump impulsion section, calculated from the impulsion volumetric flow rate ($q_{imp.}$ in m^3·s^{-1}), liquor density (ρ_{liquor} in kg·m^{-3}), acceleration of gravity (g in m·s^{-1}), pipe length (L in m), pipe equivalent length of the punctual pressure drops (L_{eq} in m), liquor velocity (V in m·s^{-1}), friction factor (f, dimensionless), diameter (d in m), difference in height (Z_1-Z_2, in m) and pump efficiency (η_{pump}).

q_{imp} and ρ_{liquor} were taken from the data obtained in the AnMBR plant; L, L_{eq}, D and Z_1-Z_2 were taken from the dimensions of the AnMBR plant; V and f were calculated by the modelling software; and a value of 0.8 was considered for η_{pump} as a theoretical typical value.

$$P_{stage(filtration, degasification\ or\ back-flushing)}(\frac{J}{S}) = \frac{q_{stage} \cdot TMP_{stage}}{\eta_{pump}} \quad \text{(Equation 3)}$$

where P_{stage} is the permeate pump power requirement during filtration, degasification or back-flushing calculated from transmembrane pressure (TMP_{stage} in Pa), pump volumetric flow rate (q_{stage} in m^3·s^{-1}) and pump efficiency (η_{pump}).

TMP_{stage} and q_{stage} were taken from the data obtained in the AnMBR plant

To calculate the net power required by the permeate pump ($P_{permeate}$), the sum of the power consumed in the following four membrane operating stages was considered: filtration ($P_{filtration}$), back-flushing ($P_{back\text{-}flushing}$), degasification ($P_{degasification}$) and ventilation ($P_{ventilation}$). Equation 4 was used to calculate the power in filtration, back-flushing and degasification. Equation 3 was used to calculate the power in ventilation since the fluid does not pass through the membrane.

The energy consumption related to the rotofilter was obtained from a catalogue for full-scale implementation [14].

Concerning sludge handling, centrifuges with an average power consumption of 45 kWh·t^{-1} TSS [15] were selected in our study as sludge dewatering system.

2.4.2. Energy recovery from methane

Since microturbines can run on biogas, they were selected as combined heat and power (CHP) technology [16]. Microturbine-based CHP technology has an overall efficiency of around 65.5%, assuming power energy efficiency of about 27% (see Equation 4).

$$W_{biogas}(kW) = \frac{V_{biogas} \cdot \left(\%CH_4 \cdot CV_{CH4}\right) \cdot \%_{power\,effciency\,CHP}}{1000 \cdot 24 \cdot 3600} \qquad \text{(Equation 4)}$$

where W_{biogas} is the power generated by the Microturbine-based CHP system using biogas, V_{biogas} (L·d^{-1}) is the biogas volume, $\%CH_4$ is the methane percentage and CV_{CH4} (KJ·m^{-3}) is the methane calorific power.

It must be said that methane dissolved in the effluent was considered to be captured for obtaining power energy by using the Microturbine-based CHP system. Theoretical capture efficiency for the dissolved methane of 100% was considered in order to assess the maximum energy potential.

2.5. Operating cost assessment

The operating cost analysis was limited in this study to net energy demand, and sludge handling and recycling to land.

The net energy demand in scenarios FP1-FP5 was evaluated for cases $BP_{33°C, SRT 70days}$, $BP_{22°C, SRT 38days}$ and $BP_{17°C, SRT 30days}$ assuming, as previously mentioned, the following terms: (1) non-capture of methane; (2) energy recovery from biogas methane; and (3) energy recovery from both biogas methane and methane dissolved in the effluent. The energy term considered in this study was €0.138 per kWh (according to the current electricity rates and prices in Spain [17]).

Concerning sludge handling and recycling to land, centrifuges require the use of polyelectrolyte for proper sludge conditioning. The dose of polyelectrolyte considered in our study was 6 $kg \cdot t^{-1}$ TSS [18], and the assumed polyelectrolyte cost was €2.52 per kg Polyelectrolyte [19]. The produced sludge was considered to be used as a fertiliser in agricultural land. The assumed cost for sludge recycling to land was €4.81 per t TSS [19].

3. Results and discussion

3.1. Overall process performance

Figure 2 shows the 20 ºC-standardised membrane permeability (K_{20}) and the MLTS level in the anaerobic sludge fed to the membrane tanks during 920 days of operation. Both K_{20} and MLTS are referred to its daily average value. This experimental period is divided into two stages, represented in Table 2 by a horizontal dashed line. Energy consumption was firstly evaluated in a period of about 790 days, which was mostly operated at sub-critical filtration conditions (scenarios FP1 to FP3). Overall, during this stage K_{20} decreased due to increasing membrane fouling over time (see days 300 to 790 in Figure 2). Around day 790 the membranes were chemically cleaned. After this chemical cleaning, the energy consumption was evaluated in a period of about 140 days, which was operated at critical filtration conditions (scenarios FP4 and FP5). During this second stage higher J_{20} were applied (see days 790 to 920 in Figure 2), making the AnMBR performance comparable to full scale aerobic MBRs [13].

Regarding the biological process, methane production increased significantly when operating at both high temperature and high SRT ($BP_{33°C,\ SRT\ 70days}$). To be precise, the average experimental methane production was 41.1, 16.8 and 8.5 $L_{CH4}·m^{-3}$ for case $BP_{33°C,\ SRT\ 70days}$, $BP_{22°C,\ SRT\ 38days}$ and. $BP_{17°C,\ SRT\ 30days}$ (see Table 3), respectively. It can be considered that an increase in the ambient temperature and/or SRT leads to offset the low growth rate of MA [20]. In this respect, simulation results in Figure 1 show adequate effluent COD concentrations and increasing methane productions and decreasing sludge productions as temperature and/or SRT increases, and reducing sludge production as temperature and/or SRT increases, within the range of operating conditions evaluated in this study.

Concerning sludge production, low/moderate amounts of sludge were generated. As Table 3 shows, the sludge production resulted in 0.16, 0.43 and 0.55 $kgTSS·kg^{-1}$ $COD_{REMOVED}$ in average for cases $BP_{33°C,\ SRT\ 70days}$, $BP_{22°C,\ SRT\ 38days}$ and. $BP_{17°C,\ SRT\ 30days}$, respectively. The minimum sludge production corresponded to case $BP_{33°C,\ SRT\ 70days}$, due to operating at high temperature (33 ºC) and high SRT (70 days). On the other hand, the experimentally determined %BVSS resulted in values below 35% within the whole range of evaluated operating conditions, which indicated adequate sludge stabilities of the wasted sludge. For instance, %BVSS resulted in the highest value (31%) when operating under the most unfavourable conditions evaluated in this study (i.e. $BP_{17°C, SRT 30days}$). It is important to highlight that one key sustainable benefit of AnMBR technology is that the produced sludge is stabilised and no further digestion is required for its disposal on farmland. In addition, sludge production in anaerobic processes is expected to be lower than in aerobic processes.

3.2. Energy consumption and operating cost of the AnMBR system

3.2.1. Power requirements

Table 5 shows the power requirements of the AnMBR plant for each of the five scenarios shown in Table 2 (FP1-FP5). This table also illustrates the weighted average distribution for the energy consumption of each particular equipment, i.e. pumps, blowers and rotofilter. The dotted line between scenario FP3 and FP4 differentiates the scenarios evaluated before and after chemically cleaning the membranes. Comparing the different scenarios assessed, it is worth to say that scenarios studied prior to chemically cleaning the membranes present higher energy consumptions (0.44, 0.32 and 0.49 $kWh·m^{-3}$ for FP1, FP2 and FP3, respectively) than those studied afterwards (0.20 and 0.19 $kWh·m^{-3}$ for FP4 and FP5, respectively). This is mainly due to the higher J_{20} applied in the second operating stage whilst operating at similar SGD_m. Specifically, the specific gas demands per permeate volume (SGD_P) resulted in the range from 21 to 32 in scenarios FP1-FP3, decreasing to approx. 14 in scenarios FP4 and FP5.

Figure 3 shows the weighted average distribution for the power requirements in the first (scenarios FP1 to FP3 in Table 5) and second operating period (scenarios FP4 and FP5 in Table 5). This figure shows that the most important item contributing the power input was the membrane tank biogas recycling blower, representing about two-thirds (60-75%) of the total AnMBR power requirements. The next in importance was the membrane tank sludge feeding pump, which represented about 15-20% of the total AnMBR power requirements. Therefore, the main terms contributing the total AnMBR power requirements were related to filtration (representing about 85-90%). This highlights the need of optimising filtration in any operating range to improve the feasibility of AnMBR technology to treat UWW.

To keep long operating periods without applying membrane chemical cleaning (i.e. minimising irreversible fouling problems: first 790 days in Figure 2), low J_{20} and/or high SGD_m are required. On the other hand, increasing the chemical cleaning frequency allows operating at high J_{20} and/or low SGD_m (i.e. low SGD_P), which reduces considerably the net energy demand (days 790 to 920 in Figure 2). To be precise, scenario FP5 was operated with the lowest SGD_P (14.4), resulting in the lowest power input (0.19 $kWh \cdot m^{-3}$). Hence, it is of vital importance to reduce the energy consumption by minimising SGD_P, which indirectly increases the membrane chemical cleaning frequency. Nevertheless, increasing the frequency of membrane chemical cleaning means high chemical reagent consumption and may affect the membrane lifetime, resulting therefore in an increase in membrane replacement and maintenance costs. Therefore, further research is required to evaluate the most suitable AnMBR operating strategy from an economical and environmental point of view including not only energy consumption but also investment and maintenance costs.

3.2.2. Net energy consumption

Figure 4 shows the net energy consumption of the AnMBR for each of the five scenarios shown in Table 2 (FP1-FP5). This net energy consumption includes both power requirements and energy recovery from methane. As mentioned earlier, each scenario (FP1-FP5) was evaluated for three different methane productions ($BP_{33°C, SRT 70days}$, $BP_{22°C, SRT 38days}$ and $BP_{17°C, SRT 30days}$) and two different levels of energy recovery (biogas methane, and biogas methane and methane dissolved in the effluent).

Figure 4 shows considerable reductions in the AnMBR energy demand (in comparison with results shown in Table 5) whenever the generated methane is used as energy resource. For example, the energy consumption in scenario FP5 was 0.19 $kWh·m^{-3}$ when methane was not captured (see Table 5); whilst the net energy demand in scenario FP5 decreased to 0.17 $kWh·m^{-3}$ for case $BP_{17°C, SRT 30days}$ when capturing both the biogas methane and the methane dissolved in the effluent. In addition, operating at high ambient temperature and/or high SRT further enhances the energy balance of the system. For instance, the energy consumption in scenario FP5 could be reduced up to 0.07 and 0.14 $kWh·m^{-3}$ when recovering energy from both biogas methane and methane dissolved in the effluent for cases $BP_{33°C, SRT 70days}$ and $BP_{22°C, SRT 38days}$, respectively (see Figure 4b).

Therefore, operating at high ambient temperature and/or high SRT allows achieving significant energy savings whenever the methane generated is captured and used as energy resource.

3.2.3. Operating cost

Figure 5 shows the operating cost of the AnMBR system including energy recovery from methane (biogas methane and methane dissolved in the effluent) and sludge handling and recycling to land. As Figure 5 illustrates, the most favourable situation as regards operating cost corresponded to case $BP_{33°C,\ SRT\ 70days}$. By way of example, the operating cost in scenario FP5 when capturing both the biogas methane and the methane dissolved in the effluent was €0.011, €0.027 and €0.032 per m^3 of treated water for cases $BP_{33°C,\ SRT\ 70days}$, $BP_{22°C,\ SRT\ 38days}$ and $BP_{17°C,\ SRT\ 30days}$, respectively. In this respect, savings of up to 64% from winter to summer seasons could be achieved. This highlights the feasibility of AnMBR technology to treat UWW in warm climate regions, as well as the necessity of optimising SRT for a given ambient temperature to maximise methane production and minimise sludge production.

On the other hand, it is worth pointing out the reduction in the operating cost if energy is recovered from methane. To be precise, scenario FP5 for case $BP_{33°C,\ SRT\ 70days}$ resulted in an operating cost of €0.028, €0.017 and €0.011 per m^3 of treated water when considering non-energy recovery from methane, energy recovery from biogas methane, and energy recovery from biogas methane and methane dissolved in the effluent, respectively (see Figure 5).

Therefore, the energy recovery from methane enables reducing considerably the operating cost of AnMBRs treating sulphate-rich UWW at ambient temperature. This highlights the need of developing feasible technologies for capturing the methane dissolved in the effluent stream not only to reduce its environmental impact (e.g. due to methane release to the atmosphere from the effluent), but also to enhance the economic feasibility of AnMBR technology.

As previously commented, several simulation scenarios were calculated in order to assess the AnMBR performance within the whole range of temperature and SRT evaluated in this study. Figure 6 shows the simulation results regarding the theoretical influence of temperature and SRT on the AnMBR operating cost (when treating sulphate-rich UWW), including energy recovery from methane (biogas methane and methane dissolved in the effluent) and sludge handling and recycling to land. Specifically, this study shows the results obtained for three SGD_P levels (22.3, 33.4 and 14.4) corresponding to scenarios FP2, FP3 and FP4, respectively. As shown in Figure 6, from a biological process perspective, the operating cost is reduced when temperature and/or SRT increase; whilst, from a filtration process perspective, the operating cost is reduced when SGDp decreases.

3.3 Effect of influent sulphate content on AnMBR operating cost

As mentioned before, Table 4 shows the total volume of methane produced (including both biogas methane and methane dissolved in the effluent) for the cases referred as $BP_{33°C,\ SRT\ 70days}$, $BP_{22°C,\ SRT\ 38days}$ and $BP_{17°C,\ SRT\ 30days}$ when treating low-sulphate UWW (10 mg $S\text{-}SO_4\cdot L^{-1}$). Similar to treating high-sulphate UWW, methane production increases significantly when operating at high ambient temperature and/or high SRT ($BP_{33°C,\ SRT\ 70days}$). When treating low-sulphate UWW, since a little amount of COD is consumed by SRB, the amount of influent COD transformed into methane increases significantly compared to treating high-sulphate UWW (see Table 4 and Table 3).

Figure 7 illustrates the operating cost of the AnMBR system when treating low-sulphate UWW. As Figure 7 shows, a significant decrease in the AnMBR operating cost could be achieved when treating low-sulphate UWW in comparison with treating high-sulphate UWW. For instance, for scenario FP5 and case $BP_{33°C,\ SRT\ 70days}$, the operating cost could be reduced from €0.017 per m^3 (see Figure 5c) to €0.001 per m^3 (see Figure 7c) when recovering energy from biogas methane. This highlights the possibility of improving the feasibility of AnMBR technology when treating low/non sulphate-loaded wastewaters.

Mention must also be made of the potential of AnMBR to be net energy producer (surplus electricity that can be exploited in other parts of the WWTP) when treating low-sulphate UWW. Specifically, Figure 7c shows that when methane is captured from both biogas and effluent, scenario FP5 presents very low operating cost (€0.006 per m^3) for case $BP_{17°C,\ SRT,\ 30days}$; whilst this cost decreases up to €0.002 per m^3 for case $BP_{22°C,\ SRT\ 38days}$. Moreover, null operating cost (or even income if the surplus energy is exploited and/or sold to the market) could be achieved for case $BP_{33°C,\ SRT\ 70days}$: theoretical maximum benefit of up to €0.014 per m^3.

Therefore, in mild/warm climates (i.e. tropical or Mediterranean), AnMBR technology is likely to be a net energy producer when treating low/non sulphate-loaded wastewaters: a theoretical maximum energy production of up to 0.11 $kWh \cdot m^{-3}$ could be obtained by capturing the methane from both biogas and effluent.

3.4 Comparison with other existing technologies

According to recent literature [13], the full-scale aerobic MBR from Peoria (USA) has a membrane and total aeration energy demand of around 0.34 and 0.55 kWh·m^{-3}, which is low compared to the consumption of other full-scale municipal aerobic MBRs (e.g. Running Springs MBR WWTP, USA, consuming around 1.3-3 kWh·m^{-3}). On the other hand, the conventional activated sludge system in Schilde (Belgium) consumed 0.19 kWh·m^{-3} [21]. In our study, the theoretical minimum energy requirements treating sulphate-rich UWW resulted in 0.07 kWh·m^{-3}. Therefore, from an energy perspective, AnMBR operating at ambient temperature is a promising sustainable system compared to other existing urban wastewater treatment technologies. Nevertheless, it is important to consider that the energy demand from the AnMBR system evaluated in our study does not take into account the energy needed for nutrient removal, which it is considered in the wastewater treatment plants that has been mentioned as references.

According to Xing et al. [22], sludge production in activated sludge processes is generally in the range of 0.3-0.5 kg TSS·kg^{-1} $COD_{REMOVED}$. As expected, low/moderate amounts of sludge were obtained in our study (0.16, 0.43 and 0.55 kg TSS·kg^{-1} $COD_{REMOVED}$ for cases $BP_{33°C,\ SRT\ 70days}$, $BP_{22°C,\ SRT\ 38days}$ and. $BP_{17°C,\ SRT\ 30days}$, respectively). Moreover, the produced sludge was considered stabilised, which allows, as mentioned before, its direct disposal on farmland without requiring further digestion.

4. Conclusions

The results obtained reinforce the importance of optimising SGD_P and SRT (for given ambient temperature conditions) to minimise the energy requirements of AnMBRs treating sulphate-rich UWW (minimum value: 0.07 kWh·m^{-3}). Operating at high ambient temperature and/or high SRT allows achieving significant energy savings whenever the methane generated is used as energy resource. Moreover, low/moderate sludge productions were obtained (minimum value: 0.16 kg TSS·kg^{-1} $COD_{REMOVED}$), which further enhanced the AnMBR operating cost (minimum value: €0.01 per m^3). On the other hand, the sulphate content in the UWW significantly affected the final production of methane and thereby affected the overall energy consumption. Indeed, AnMBR technology is likely to be a net energy producer when treating low/non sulphate-loaded wastewaters in warm/hot climates: theoretical maximum energy productions of up to 0.11 kWh·m^{-3} could be achieved.

Acknowledgements

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References

[1] X.M. Guo, E. Trably, E. Latrille, H. Carrère, J.P. Steyer, Hydrogen production from agricultural waste by dark fermentation: A review, Int. J. Hydrog. Energy 35 (2010) 10660–10673.

[2] A. Gallego, A. Hospido, M.T. Moreira, G. Feijoo, Environmental performance of wastewater treatment plants for small populations, Resour. Conser. Recy. 52 (2008) 931–940.

[3] L. Raskin, Anaerobic Membrane Bioreactors for Sustainable Wastewater Treatment, WERF Report U4R08, June 2012.

[4] L.W. Hulshoff Pol, Treatment of sulphate-rich wastewaters: microbial and process technological aspects TMR Summer School Programme, The Biological Sulfur Cycle: Environmental Science and Technology, April, Wageningen, The Netherlands, 1998.

[5] J.B. Giménez, A. Robles, L. Carretero, F. Durán, M.V. Ruano, M.N. Gattib, J. Ribes, J. Ferrer, A. Seco, Experimental study of the anaerobic urban wastewater treatment in a submerged hollow-fiber membrane bioreactor at semi-industrial scale, Bioresource Technol. 102 (2011) 8799–8806.

[6] Miliarium Aureum, S.L, Portal de referencia para la Ingeniería Civil, Construcción y el Medio Ambiente, Available on: http://www.miliarium.com/prontuario/Tablas/Aguas/CaracterizacionAguasResidualesUrbanas.asp, (accessed 6 february 2014).

[7] A. Robles, M.V. Ruano, J. Ribes, J. Ferrer, Sub-critical long-term operation of industrial scale hollow-fibre membranes in a submerged anaerobic MBR (HF-AnMBR) system, Sep. Purif. Technol. 100 (2012) 88–96.

[8] A. Robles, M.V. Ruano, J. Ribes, J. Ferrer, Factors that affect the permeability of commercial hollow-fibre membranes in a submerged anaerobic MBR (HF-AnMBR) system, Water Res. 47 (2013) 1277–1288.

[9] J. Ferrer, A. Seco, J. Serralta, J. Ribes, J. Manga, E. Asensi, J.J. Morenilla, F. Llavador, DESASS: A software tool for designing, simulating and optimising WWTPs, Environ. Modell. Softw. 23 (2008) 19–26.

[10] R. Barat, J. Serralta, M.V. Ruano, E. Jiménez, J. Ribes, A. Seco, J. Ferrer, Biological Nutrient Removal Model Nº 2 (BNRM2): A general model for Wastewater Treatment Plants, Water Sci. Technol. 67 (2013) 1481–1489.

[11] *Standard methods for the Examination of Water and Wastewater,* 21th edn, American Public Health Association/American Water Works Association/Water Environmental Federation, Washington DC, USA, 2005.

[12] J.B. Giménez, N. Martí, J. Ferrer, A. Seco, Methane recovery efficiency in a submerged anaerobic membrane bioreactor (AnMBR) treating sulphate-rich urban wastewater: Evaluation of methane losses with the effluent, Bioresource Technol. 118 (2012) 67–72.

[13] S.J. Judd, C. Judd, Principles and Applications of Membrane Bioreactors in Water and Wastewater Treatment. Second Edition, Elsevier, London, UK, 2011, chapter 3 and 5.

[14] AGUA TÉCNICA, Equipos mecánicos para tratamiento de aguas, Tamiz rotativo para sólidos finos, Available on: http://www.aguatecnica.es/index.php?option=com_content&view=article&id=50&Itemid=59, (accessed 7 September 2012).

[15] X. Elias-Castell, Vías de tratamiento y valorización de fangos de depuradora. Ediciones Díaz de santos, Madrid, España, 2012.

[16] EPA, Environmental Protection Agency, Combined Heat and Power Partnership, Agency of the United States federal government, 2012.

[17] Spanish electricity tariff, Iberdrola S.A, 2013.

[18] J.A. Sainz-Lastre, Tecnologías para la sosteniblidad, Procesos y operaciones unitarias en depuración de aguas residuales, Colección EOI, 2005.

[19] Miliarium Aureum, S.L, Portal de referencia para la Ingeniería Civil, Construcción y el Medio Ambiente, Available on: http://www.miliarium.com/Proyectos/Depuradoras/introduccion/edar.asp, (accessed 5 may 2013).

[20] Lettinga. G, Rebac. S, Zeeman. G, Challenge of psychrophilic anaerobic wastewater treatment. Trends Trends Biotechnol. 19 (2001) 363–370.

[21] A. Fenu, J. Roels, T. Wambecq, K. De Gussem, C. Thoeye, G. De Gueldre, B. Vand De Steene, Energy audit of a full scale MBR system, Desalination 262 (2010) 121–128.

[22] C.-H. Xing, W.-Z. Wu, Y. Quian, E. Tardieu, Excess sludge production in membrane bioreactors: a theoretical investigation, J. Environ. Eng. ASCE, 129 (2003) 291–297.

Table and Figure captions

Table 1. Average characteristics of AnMBR influent.

Table 2. Main operating conditions in scenarios FP1-FP5. **TMP**: transmembrane pressure; $\mathbf{J_{20}}$: 20 ºC-standardised transmembrane flux; **MLTS**: mixed liquor total solids; $\mathbf{SRF_{MT}}$ and $\mathbf{SRF_{MT}}$**:** sludge recycling flow to membrane tank and anaerobic reactor, respectively; $\mathbf{SGD_m}$: specific gas demand per square metre of membrane area; and $\mathbf{BRF_{AnR}}$: biogas recycling flow to anaerobic reactor.

Table 3. Operating temperature (T) and sludge retention time (SRT), total methane production (V_{CH4}), biogas methane ($V_{CH4,BIOGAS}$), and methane dissolved in the effluent ($V_{CH4,EFFLUENT}$) per m^3 of treated water, and sludge production, for cases $BP_{33°C,\ SRT\ 70days}$, $BP_{22°C,\ SRT\ 38days}$ and $BP_{17°C,\ SRT\ 30days}$.

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Table 5. Power requirements in scenarios FP1-FP5.

Figure 1. AnMBR performance at different temperature and SRT conditions: **(a)** effluent COD (without including dissolved methane concentration); **(b)** total methane production (V_{CH4}) (biogas methane and methane dissolved in the effluent); and **(c)** sludge production measured in kg TSS·kg^{-1} COD removed.

Figure 2. Evolution of K_{20} and MLTS throughout 920 days of operation.

Figure 3. Weighted average distribution for the AnMBR power requirements in scenarios: **(a)** FP1 to FP3; and **(b)** FP4 and FP5.

Figure 4. Net energy consumption in scenarios FP1-FP5 for cases $BP_{33°C,\ SRT\ 70days}$ (■), $BP_{22°C,\ SRT\ 38}$ (■) and $BP_{17°C,\ SRT\ 30days}$ (■) including energy recovery from: **(a)** biogas methane; and **(b)** biogas methane and methane dissolved in the effluent.

Figure 5. Operating cost (net energy consumption and sludge handling and recycling to land) in scenarios FP1-FP5 for cases $BP_{33°C,\ SRT\ 70days}$ (■), $BP_{22°C,\ SRT\ 38days}$ (■) and $BP_{17°C,\ SRT\ 30days}$ (■): **(a)** non-capture of methane; **(b)** energy recovery from biogas methane; and **(c)** energy recovery biogas methane and methane dissolved in the effluent.

Figure 6. AnMBR operational cost (power requirements, energy recovery from total methane production, and sludge handling and recycling to land) at different temperature and SRT conditions for three SGD_P levels: (■) SGDp 33.4; (■) SGDp 22.3; and (■) SGDp 14.3.

Figure 7. Operating cost (net energy consumption and sludge handling and recycling to land) in scenarios FP1-FP5 for cases $BP_{33°C,\ SRT\ 70days}$ (■), $BP_{22°C,\ SRT\ 38days}$ (■) and $BP_{17°C,\ SRT\ 30days}$ (■) when treating low-sulphate UWW: **(a)** non-capture of methane; **(b)** energy recovery from biogas methane; and **(c)** energy recovery biogas methane and methane dissolved in the effluent.

Table 1. Average characteristics of AnMBR influent.

Parameter	Mean ± SD
Treatment flow rate ($m^3 \cdot day^{-1}$)	3.2 ± 0.7
TSS ($mg \cdot L^{-1}$)	313 ± 45
VSS ($mg \cdot L^{-1}$)	257 ± 46
COD ($mg \cdot L^{-1}$)	650 ± 147
SO_4-S ($mg \cdot L^{-1}$)	105 ± 13
NH_4-N ($mg \cdot L^{-1}$)	35 ± 3
PO_4-P ($mg \cdot L^{-1}$)	4 ± 1

Table 2. Main operating conditions in scenarios FP1-FP5. **TMP**: transmembrane pressure; $\mathbf{J_{20}}$: 20 ºC-standardised transmembrane flux; **MLTS**: mixed liquor total solids; $\mathbf{SRF_{MT}}$ and $\mathbf{SRF_{MT}}$**:** sludge recycling flow to membrane tank and anaerobic reactor, respectively; $\mathbf{SGD_{m}}$: specific gas demand per square metre of membrane area; and $\mathbf{BRF_{AnR}}$: biogas recycling flow to anaerobic reactor.

Scenario	**Period (days)**	**TMP (bar)**	$\mathbf{J_{20}}$ **(LMH)**	**MLTS** $(\mathbf{g \cdot L^{-1}})$	$\mathbf{SRF_{MT}}$ $(\mathbf{m^3 \cdot h^{-1}})$	$\mathbf{SRF_{AnR}}$ $(\mathbf{m^3 \cdot h^{-1}})$	$\mathbf{SGD_{m}}$ $(\mathbf{m^3 \cdot h^{-1} \cdot m^{-2}})$	$\mathbf{BRF_{AnR}}$ $(\mathbf{m^3 \cdot h^{-1}})$
FP1	137-170	0.35	10.0	32.5	2.7	1	0.23	1.5
FP2	361-404	0.13	13.3	12.5	2.7	1	0.23	1.5
FP3	556-600	0.26	9.0	22.5	2.7	1	0.23	1.5
FP4	807-850	0.09	15.0	14	2.7	1	0.17	1.5
FP5	853-896	0.20	20.0	13	2.7	1	0.23	1.5

Table 3. Operating temperature (T) and sludge retention time (SRT), total methane production (V_{CH4}), biogas methane ($V_{CH4,BIOGAS}$), and methane dissolved in the effluent ($V_{CH4,EFFLUENT}$) per m^3 of treated water, and sludge production, for cases $BP_{33°C,\ SRT\ 70days}$, $BP_{22°C,\ SRT\ 38days}$ and $BP_{17°C,\ SRT\ 30days}$.

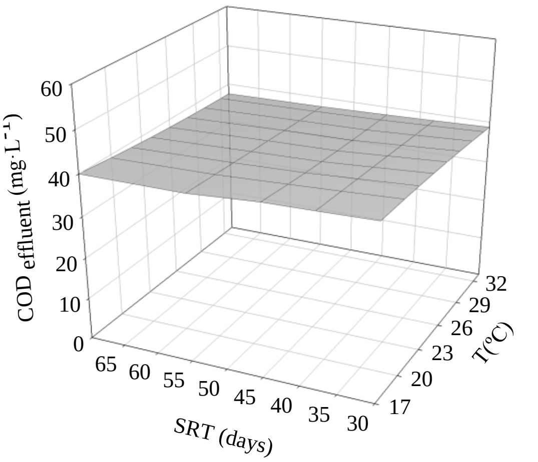
	T (°C)	**SRT (days)**	**V_{CH4} (BIOGAS+EFFLUENT) ($L·m^{-3}$)**	**$V_{CH4,BIOGAS}$ ($L·m^{-3}$)**	**$V_{CH4,EFFLUENT}$ ($L·m^{-3}$)**	**Sludge production (kg TSS·kg^{-1} COD removed)**
$BP_{33°C,\ SRT\ 70days}$	33	70	41.1	26.5	14.6	0.16
$BP_{22°C,\ SRT\ 38days}$	22	38	16.8	8.4	8.4	0.43
$BP_{17°C\ ,\ SRT\ 30days}$	17	30	8.5	1.4	7.1	0.55

Table 4. Theoretical methane production (V_{CH4}), biogas methane ($V_{CH4,BIOGAS}$), and methane dissolved in the effluent ($V_{CH4,EFFLUENT}$) per m^3 of treated water for cases $BP_{33°C,\ SRT\ 70days}$, $BP_{22°C,\ SRT\ 38}$ and $BP_{17°C,\ SRT\ 30days}$ when treating low-sulphate UWW.

	$V_{CH4,\ (BIOGAS+EFFLUENT)}$ (L·m^{-3})	$V_{CH4,BIOGAS}$ (L·m^{-3})	$V_{CH4,EFFLUENT}$ (L·m^{-3})
$BP_{33°C,\ SRT\ 70days}$	105.8	68.1	37.7
$BP_{22°C,\ SRT\ 38days}$	81.5	40.8	40.7
$BP_{17°C,\ SRT\ 30days}$	73.2	11.7	61.5

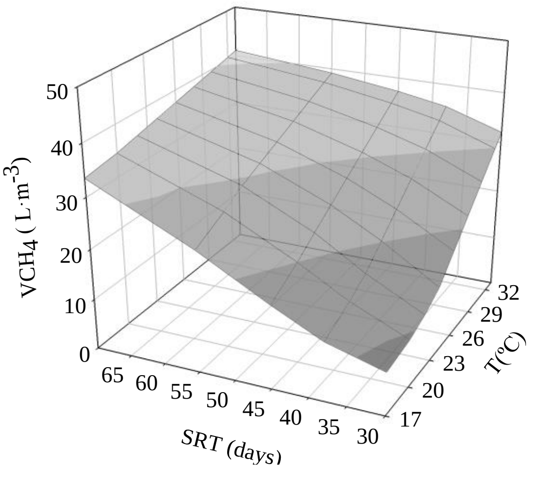
Table 5. Power requirements in scenarios FP1-FP5.

SCENARIO	TOTAL ENERGY CONSUMPTION ($kWh·m^{-3}$)	PERMEATE PUMP (%)	MEMBRANE TANK BIOGAS RECYCLING BLOWER (%)	MEMBRANE TANK SLUDGE FEEDING PUMP (%)	STIRRING POWER REACTOR (%)	ANAEROBIC REACTOR FEEDING PUMP (%)	ROTOFILTER (%)
FP1	0.44	2.34	73.15	14.54	8.20	0.52	1.25
FP2	0.32	1.26	73.18	14.69	8.43	0.72	1.73
FP3	0.49	1.61	73.94	14.58	8.27	0.47	1.13
FP4	0.20	1.38	61.73	21.02	11.89	1.17	2.81
FP5	0.19	3.06	67.46	16.19	9.18	1.21	2.90



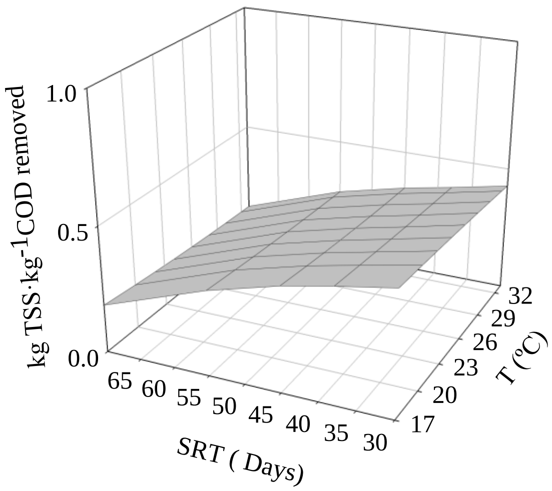
60
50
40
30
20
10
0
COD effluent (mg·L⁻¹)
65
60
55
50
45
40
35
30
SRT (days)
17
20
23
26
29
32
T(ºC)

(a)



50
40
30
20
10
0
VCH4 (L·m-3)
65
60
55
50
45
40
35
30
SRT (days)
17
20
23
26
29
32
T(ºC)

(b)



(c)

Figure 1. AnMBR performance at different temperature and SRT conditions: **(a)** effluent COD (without including dissolved methane concentration); **(b)** total methane production (V_{CH4}) (biogas methane and methane dissolved in the effluent); and **(c)** sludge production measured in kg TSS·kg^{-1} COD removed.

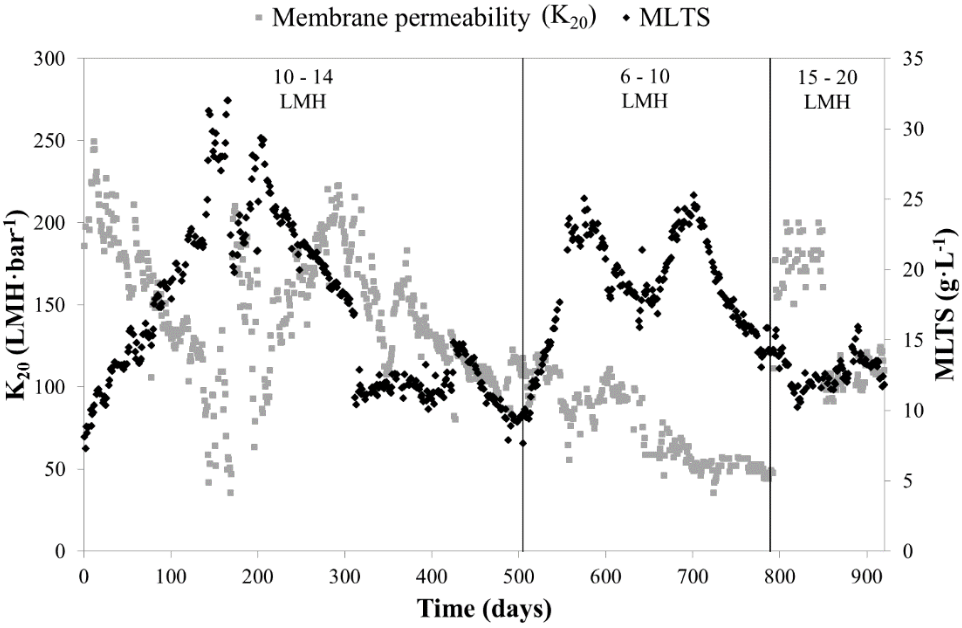
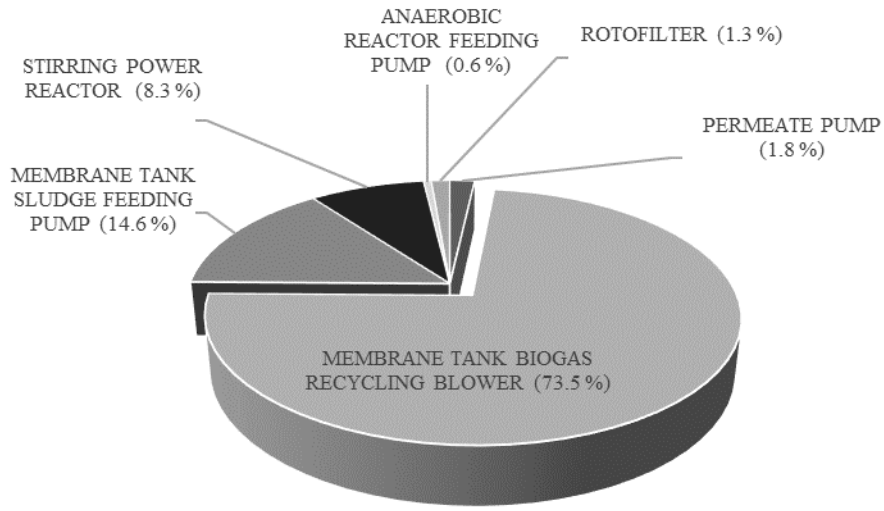


Figure 2. Evolution of K_{20} and MLTS throughout 920 days of operation.



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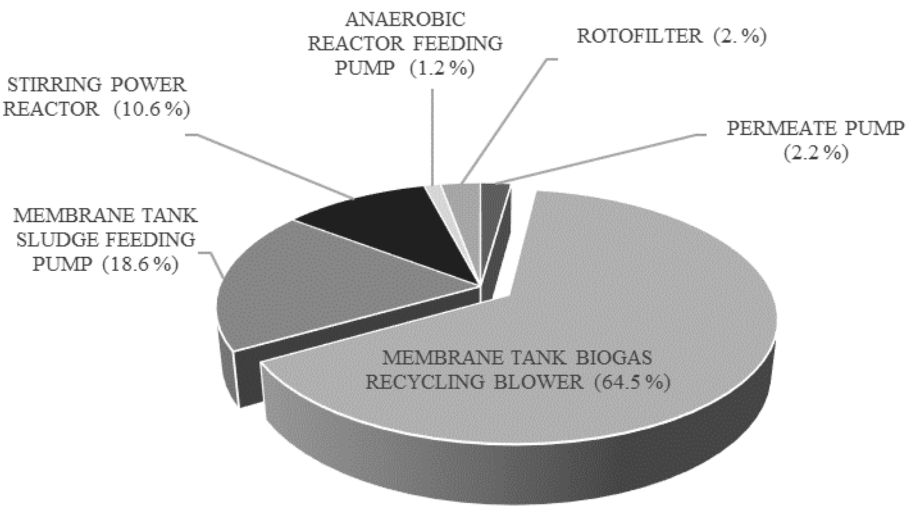
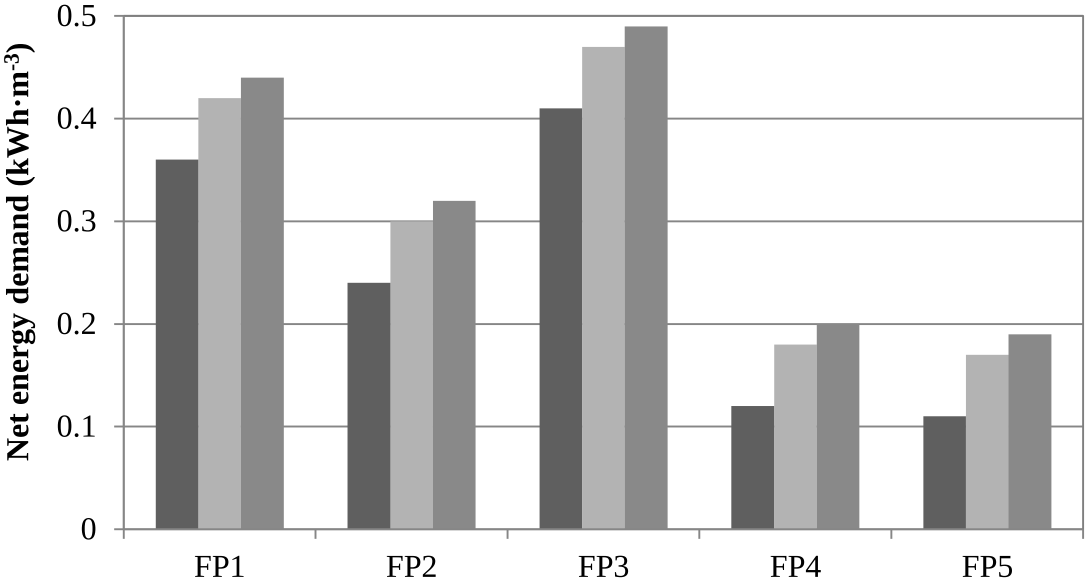
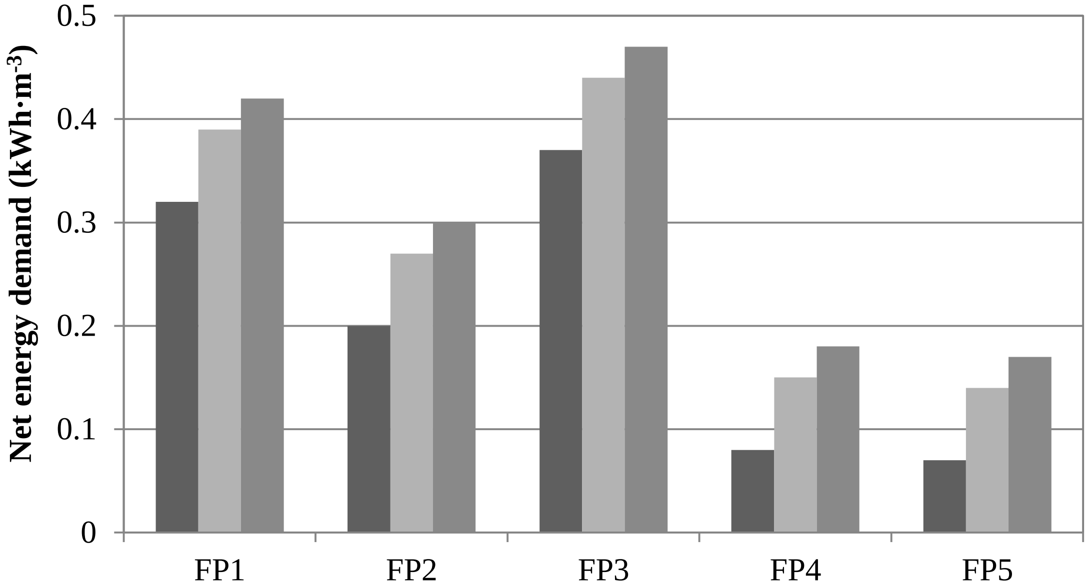


(b)

Figure 3. Weighted average distribution for the AnMBR power requirements in scenarios: **(a)** FP1 to FP3; and **(b)** FP4 and FP5.

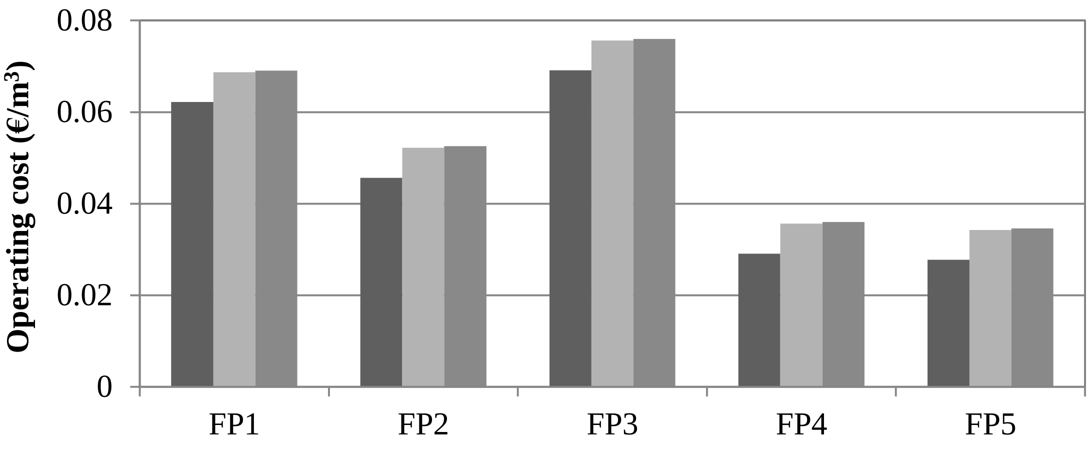


(a)



(b)

Figure 4. Net energy consumption in scenarios FP1-FP5 for cases $BP_{33°C,\ SRT\ 70days}$ (■), $BP_{22°C,\ SRT\ 38}$ (■) and $BP_{17°C,\ SRT\ 30days}$ (■) including energy recovery from: **(a)** biogas methane; and **(b)** biogas methane and methane dissolved in the effluent.



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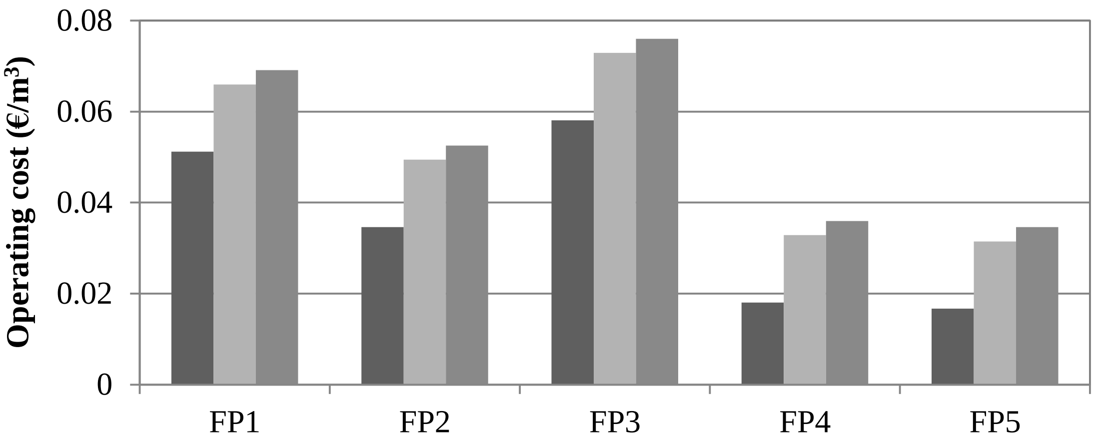

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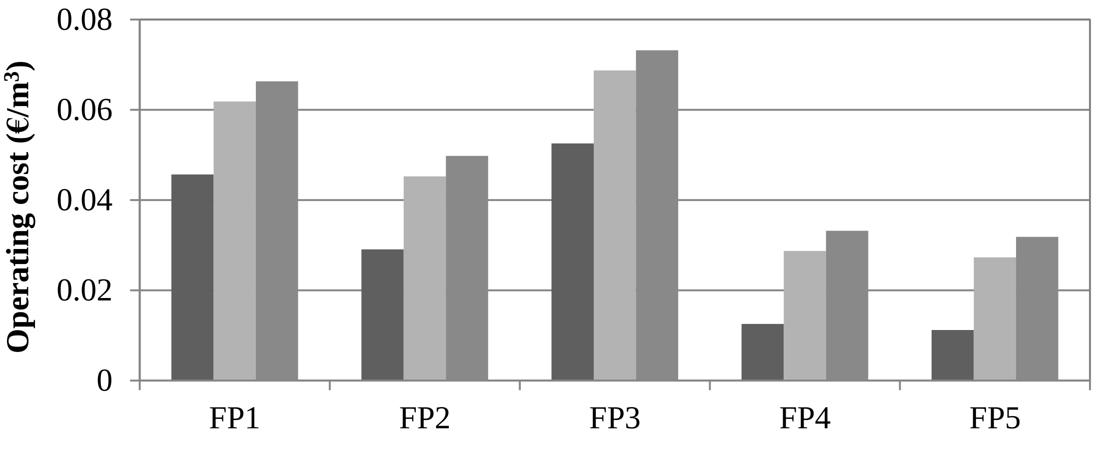
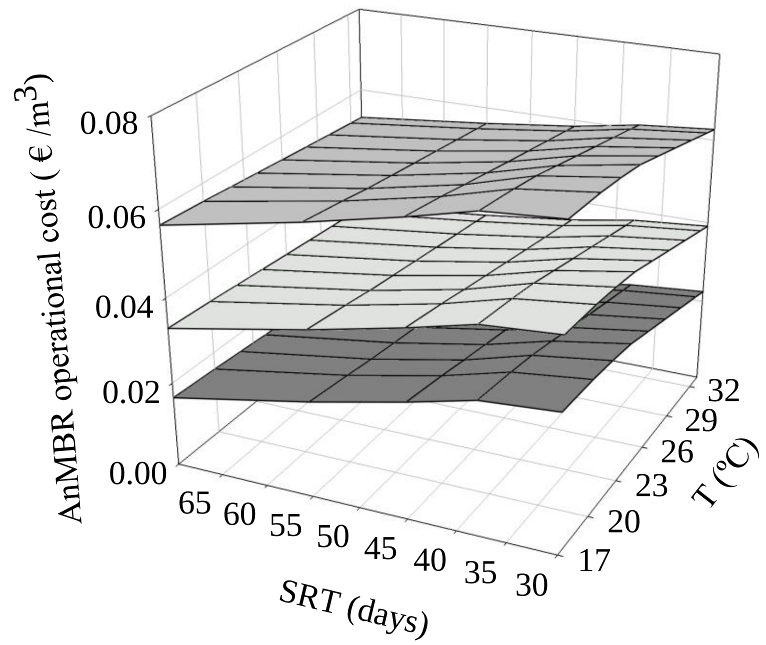


(c)

Figure 5. Operating cost (net energy consumption and sludge handling and recycling to land) in scenarios FP1-FP5 for cases $BP_{33°C,\ SRT\ 70days}$ (■), $BP_{22°C,\ SRT\ 38days}$ (■) and $BP_{17°C,\ SRT\ 30days}$ (■): **(a)** non-capture of methane; **(b)** energy recovery from biogas methane; and **(c)** energy recovery biogas methane and methane dissolved in the effluent.



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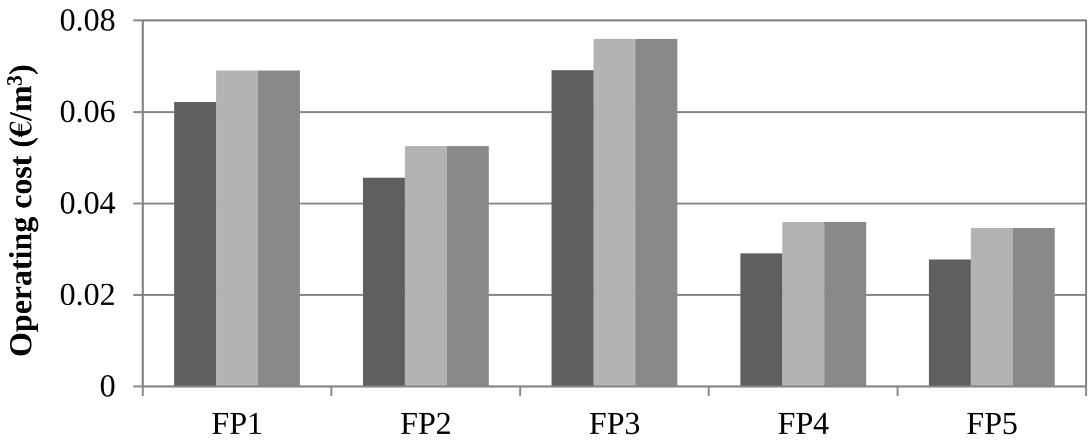
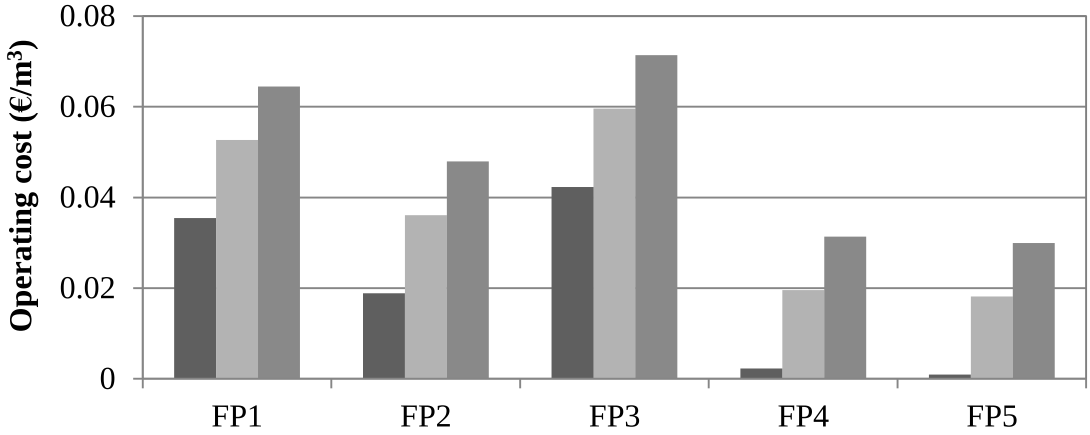


(a)



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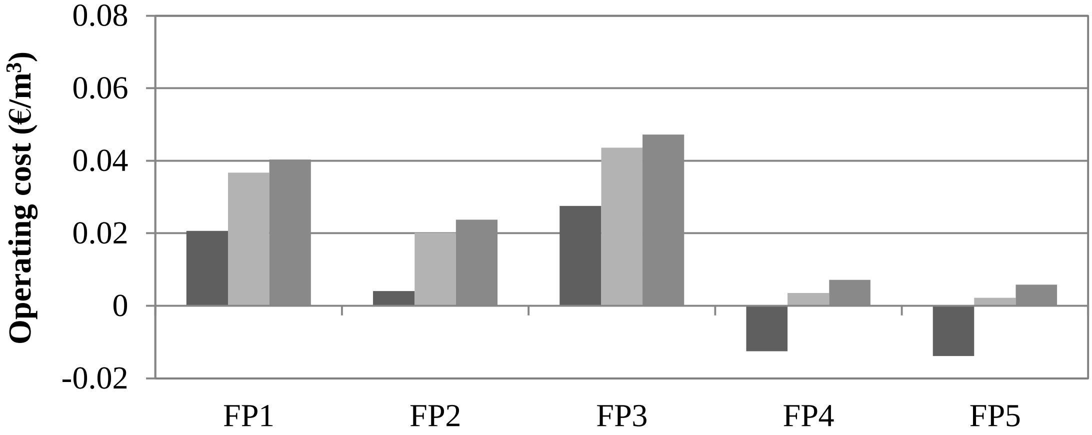

(c)

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