Recovery of reusable water from sewage using aerated flat-sheet membranes

V. I. Diamantis, I. Antoniou, E. Athanasoulia, P. Melidis and A. Aivasidis

ABSTRACT

Continuous developments and advances in membrane technology allow recovering to large extent reusable water from untraditional water sources, such as municipal effluents. In this paper, operational results and preliminary cost-analysis of a microfiltration (MF) process used for raw sewage and secondary effluent polishing are given. The research was conducted with a 22 sandwich-type flat-sheet membrane module (0.45 μm and 4.5 m²) employing aeration for fouling control. During raw sewage filtration the majority of the organics were retained (effluent COD < 90 mg/L and SS ~ 0 mg/L). However, the fluxes were significantly lower [~7 L/(m² h)] compared to those achieved during secondary effluent filtration [~29 L/(m² h)] (at TMP 0.14 bar). In the second case, aeration was found to be suitable mean for fouling control. The total costs (capital and operational) for water recovery from raw sewage and secondary effluent were estimated to be in the order of 2 and 0.75 €/m³ respectively.

Key words | activated sludge free process, decentralized wastewater treatment, direct membrane filtration, municipal wastewater, primary membrane filtration, raw sewage filtration, secondary effluent, tertiary treatment, water recovery

INTRODUCTION

The worldwide freshwater shortage has made that currently municipal discharges are considered as valuable resources. The latter already receive attention in arid and semi-arid environments and representative examples are given by Asano (1998). Reuse schemes usually implement aerobic biological processes (i.e. activated sludge), as the core technology for water purification, followed by microfiltration (MF)/ultrafiltration (UF) and reverse osmosis (RO) (Dewettinck et al. 2001; Van Houtte & Verbauwhede 2008). The latter treatment train is capable to produce high quality water but is complex and costly (Verstraete et al. 2009).

At conventional aerobic sewage treatment plant the municipal wastewater costs about 0.6 €/m³ to be treated (Van Haandel & van der Lubbe 2007). The costs reported above, incorporate both the annualized capital expenses (CAPEX) over the expected lifetime of the treatment plant and the operational expenses (OPEX) per m³ of water treated. Thus, primary clarification, activated sludge treatment and sludge disposal are included. The total costs associated with RO (including MF/UF pre-treatment) of secondary effluent have been reported to be 0.46 €/m³ (Van Houtte & Verbauwhede 2008). This value was based on actual capital and operational expenses of the Torrel Reuse Plant in Belgium. Because of the additional expenditures for conventional activated sludge (0.6 €/m³) and the reuse treatment, the total process costs ~1.0 €/m³ of water treated. On the other hand, it is possible to optimize the overall process by implementing the membrane bioreactor (MBR) technology. The total costs (CAPEX and OPEX) for an MBR treatment system are slightly higher than those of activated sludge, but in the same range, i.e. ~0.6 €/m³ (Cote et al. 2004). Adding an extra cost of 0.2–0.3 €/m³ for treating the effluent from the MBR with RO gives a total of 0.8–0.9 €/m³ of wastewater treated.
Membranes are able to separate a wide range of suspended, colloidal and soluble constituents from water. Although considerable knowledge, from both long-term-pilot and full scale plants, exists on membrane filtration of secondary effluent (tertiary applications) and activated sludge (MBR systems), applications directly on sewage (Primary Membrane Filtration, PMF) are currently limited to laboratory and pilot testing (e.g. Bendick et al. 2005; Ravazzini et al. 2005; Gan & Allen 1999). These studies have been performed with microfiltration, ultrafiltration and nanofiltration systems, with the former being the most frequently tested. The main focus was to evaluate the permeate quality (removal of pathogens, suspended solids and organic matter) and process performance regarding steady-state permeate fluxes at different pore size and operational conditions (transmembrane pressure-TMP, crossflow velocity, temperature, etc).

Primary membrane filtration (PMF) provides the possibility to recover reusable water in one step (suitable for gardening or secondary applications) and to up-concentrate the organics present in domestic effluents for a waste-to-energy strategy (Verstraete et al. 2009). It is especially suitable for decentralised and satellite applications (Gikas & Tchobanoglous 2009a). However, the limitation of membrane filtration is fouling (Judd & Till 2000; Schafer et al. 2006) and different mitigation strategies are summarized in Table 1.

In this paper a flat-sheet microfiltration module with aeration as a means for fouling control, was operated on raw sewage and secondary effluent. The aim of the study was to evaluate, from the engineering point of view (technically and economically) process performance (permeate quality and fluxes obtained) and the major costs involved (capital and operational), for a small-scale installation (20 population equivalent).

**MATERIALS AND METHODS**

**Wastewater**

Original wastewater received by the municipal treatment plant of Xanthi (Greece) was used for the study. The water was transported to the laboratory and stored in a cold (4°C) completely mixed tank. The secondary effluent was obtained by a pilot-scale activated sludge process operated with the same sewage (Kapagiannidis & Aivasidis 2008).

**Membrane plant design**

Commercial microfiltration equipment was used for the study (ATB Umwelttechnologien GmbH) consisting of membranes and housing, permeate pump (0.12 kW), air-compressor (0.12 kW), backwash unit, tubing, pressure sensor and control panel. A total of 22 sandwich-type flat-sheet membranes with a pore size of 0.4 μm and a total surface area 4.5 m² were used. The operation capacity of the permeate pump was 7.2 m³/d and the operational TMP lower than 0.20 bar (according to the supplier).

**Short-term permeation experiments**

The membrane tank (120 L) was fed with a batch of raw sewage or secondary effluent and the permeate pump was set in operation (Figure 1). During the experimental period, the system was operating for up to 5–6 hours per day and the flux was monitored until steady-state values were achieved (three stable subsequent flux measurements). The operational program of the permeate pump was in all cases set at 9 min discharge and 1 min relaxation. The permeate stream was continuously returned to the membrane tank. Three different aeration programs were tested: i) continuous aeration, ii) zero aeration and iii) intermittent aeration (9 min without aeration followed by 1 min aeration). Thus, in the last case the membranes were aerated during the relaxation phase. The effect of TMP on process...
performance was also examined by gradually increasing transmembrane pressure from 0.12 to 0.14 and 0.16 bar.

Clean water flux (CWF)

The CWF was determined using tap water while the membranes were continuously aerated. The TMP was gradually increased from 0.12 to 0.14 and 0.16 bar and the permeate flux was monitored continuously until steady-state conditions were achieved.

Membrane cleaning

It was performed with chlorine (and surfactant) or citric acid. Initially 250 ml of commercial chlorine solution (5%) were added into the membrane tank (120 L) and left overnight, while the membranes were continuously aerated. Subsequently, the water was rejected and freshwater was added with 0.5 kg citric acid (pH = 2.8) for one more night. In all cases the clean water flux was recovered after membrane cleaning.

Analytical methods

During the course of the experiment, samples were obtained from the membrane tank and the permeate stream. The parameters examined were: suspended solids (SS), chemical oxygen demand (COD), nitrogen (NH₄⁺-N, NO₃⁻-N) and phosphorus (PO₄³⁻-P, TP). The analyses were performed according to Standard Methods (APHA 1998).

RESULTS AND DISCUSSION

Membrane filtration on secondary treated sewage

The quality of the raw wastewater, the secondary effluent and the membrane permeate are given in Table 2. The removal of COD during membrane filtration of secondary effluent was maintained on average at 55 (±25) %. The retention of suspended solids was complete. During membrane filtration, the NH₄⁺-N concentration decreased from 5 to 2 mg/L and the respective NO₃⁻-N concentration showed an increase from 5 to 7 mg/L. An explanation for this behavior could be nitrification, sustained by the action of nitrifying bacteria, either attached on the membrane surface or suspended inside the membrane tank (Li et al. 2005). A slight decrease of TP was also observed and this was the result of particulate-P retention (Ravazzini et al. 2005).

In Figures 2 and 3 the normalized permeate flux (corrected at 20°C) during secondary effluent processing is given. Increasing the trans-membrane pressure from 0.12 to 0.16 bar a linear increase of the clean water flux was observed (see Figure 2). However, this was not the case for the secondary effluent. Due to the formation of the fouling layer, the steady-state permeate flux values were significantly lower especially when the TMP was higher than 0.12 bar. In general, aerations of the membranes was beneficial to maintain high permeate flux. The intermittent aeration program was efficient to achieve steady-state fluxes in the order to 29 L/(m² h) (see Figure 3).

Considering a design flux value equal to 29 L/(m² h) and the total surface area of the module (4.5 m²), the pilot plant was capable to treat approximately 3 m³ of water per day.
This amount of water corresponds to ~20 population equivalents (PE) at 150 L/(PE d).

Membrane filtration directly on raw sewage

In Table 3 data concerning the efficiency of the microfiltration process directly on raw sewage are presented. In this case the system was capable to remove the majority of the organics giving a final effluent COD between 45–83 mg/L at influent concentrations ranging from 330–500 mg/L. The retention of suspended solids was complete similarly to the filtration of secondary sewage. MF membranes are by definition capable to remove the majority of the suspended, colloidal particles and pathogens (Judd & Till 2000; Bendick et al. 2004; Bendick et al. 2005; Sayed et al. 2006; Schafer et al. 2006). The high COD removal observed in our study during raw sewage filtration, can be attributed to the formation of the cake-fouling layer (additional water filtration) which resulted in the retention of compounds having smaller size than the membrane pores. Similarly, varying membrane pore size from 0.05 to 1.4 μm did not resulted in significant changes in permeate COD and SS during treatment of primary effluent and this was closely directed by the formation of a cake layer (Bendick et al. 2005).

In Figure 4 the normalized permeate flux (corrected at 20°C) is given under different operational conditions. The respective data for secondary effluent filtration are also presented for comparison. During the continuous aeration program, the permeate flux decreased continuously and steady-state conditions were not achieved within the first 120 min of operation. Consequently, air-slugs were not efficient to maintain optimum hydraulic performance. In order to avoid biomass growth, the test was continued without aeration. In this case, the permeate flux showed a continuous decrease, which was attributed to the formation of the cake layer. The flux values achieved after 2.5 h continuous operation was ~5 L/(m²h). Intermittent aeration resulted in a slight increase of process performance and steady-state values were approached (see Figure 4). Under these conditions, the permeate flux with raw sewage was 4 times lower, compared to the values obtained with secondary effluent [7 and 29 L/(m²h) respectively].

Table 3 | Permeate quality during membrane filtration of raw sewage (Numbers in parenthesis for standard deviation, n = 5)

<table>
<thead>
<tr>
<th>Parameter (units)</th>
<th>Raw sewage</th>
<th>Permeate</th>
</tr>
</thead>
<tbody>
<tr>
<td>SS influent (mg/L)</td>
<td>207 (23)</td>
<td>0 (0)</td>
</tr>
<tr>
<td>COD influent (mg/L)</td>
<td>413 (97)</td>
<td>64 (27)</td>
</tr>
<tr>
<td>SS removal (%)</td>
<td>100 (0)</td>
<td></td>
</tr>
<tr>
<td>COD removal (%)</td>
<td>85 (2)</td>
<td></td>
</tr>
</tbody>
</table>

Figure 2 | Steady-state permeate flux values as a function of TMP using clean water and secondary effluent under different aeration programs.

Figure 3 | Effects of aeration frequency on permeate flux during continuous microfiltration of secondary effluent.

Figure 4 | Normalized permeate flux during membrane filtration of raw sewage and secondary effluent (TMP at both experiments 0.14 bar).
The results presented in Figure 4 indicate that although aeration was able to remove the fouling layer during secondary effluent filtration, this was not the case for raw sewage. It was therefore speculated that the adsorption/deposition of fine colloids and macromolecules (proteins, lipids, polysaccharides) play a significant role. Modise et al. (2005) determined that 72% of the total resistance to flow during primary effluent filtration was due to cake build-up while 27% accounted for internal MF membrane fouling. The formation of cake layer was also reported as the predominant mechanisms for ultrafiltration (UF) of raw sewage and primary effluent (Ravazzini et al. 2005).

Assuming a design flux value equal to 7 L/(m² h) the plant was capable to treat approximately 0.7 m³/d of raw sewage (≈200 m³/yr). This amount of water corresponds to ≈5 population equivalent (PE) at 150 L/(PE d). Thus, a total of 4 membrane modules are required for a 3 m³/d capacity process (20 PE, see previous section).

**Preliminary economic considerations**

The capital (CAPEX) and operational expenses (OPEX) for the membrane filtration plant are summarized in Table 4.

<table>
<thead>
<tr>
<th>Capital expenses (CAPEX, €)</th>
<th>Raw sewage</th>
<th>Second. effl.†</th>
<th>MBR‡</th>
</tr>
</thead>
<tbody>
<tr>
<td>Membrane module (all included)</td>
<td>12,000</td>
<td>5,000</td>
<td>7,500</td>
</tr>
<tr>
<td>Annualized CAPEX§</td>
<td>1,200 €/yr</td>
<td>500 €/yr</td>
<td>750 €/yr</td>
</tr>
<tr>
<td>CAPEX/(PE yr)</td>
<td>60 €/(PE yr)</td>
<td>25 €/(PE yr)</td>
<td>37.5 €/(PE yr)</td>
</tr>
<tr>
<td>CAPEX/m³</td>
<td>1.33 €/m³</td>
<td>0.56 €/m³</td>
<td>0.83 €/m³</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Operational expenses (OPEX annualized, €/yr)</th>
<th>Raw sewage</th>
<th>Second. effl.†</th>
<th>MBR‡</th>
</tr>
</thead>
<tbody>
<tr>
<td>Energy for permeate discharge (0.071 €/kWh)</td>
<td>224</td>
<td>56</td>
<td>112</td>
</tr>
<tr>
<td>Energy for aeration (0.071 €/kWh)</td>
<td>32</td>
<td>8</td>
<td>128</td>
</tr>
<tr>
<td>Chlorine and surfactants (0.50 €/L)</td>
<td>24</td>
<td>6</td>
<td>12</td>
</tr>
<tr>
<td>Citric acid (1.0 Euro/kg)</td>
<td>96</td>
<td>24</td>
<td>48</td>
</tr>
<tr>
<td>Membrane replacement</td>
<td>320</td>
<td>80</td>
<td>160</td>
</tr>
<tr>
<td>Labor</td>
<td>Not included</td>
<td>Not included</td>
<td>Not included</td>
</tr>
<tr>
<td>Total OPEX/m³</td>
<td>696 €/yr</td>
<td>174 €/yr</td>
<td>460 €/yr</td>
</tr>
<tr>
<td>Total CAPEX and OPEX/m³</td>
<td>0.77 €/m³</td>
<td>0.19 €/m³</td>
<td>0.51 €/m³</td>
</tr>
<tr>
<td>Total</td>
<td>2.10 €/m³</td>
<td>0.75 €/m³</td>
<td>1.34 €/m³</td>
</tr>
</tbody>
</table>

*Four (4) membrane modules required.
†One (1) module required.
‡Two (2) modules required.
§Life span: 15 years; interest rate: 6%.

The manufacturer provides all the necessary equipment (membranes and housing, permeate pump, air-compressor, control panel, pressure sensor, automations, valves, tubings etc) at a cost 5,000 €/module. The capital expenses can be further decreased to 3,000 €/module when four (4) modules are purchased. Considering the life time of the plant (15 years) and the interest rate (6%), the annualized capital expenses are equal to $R = 0.103 \times 5,000 = 500$ and $1,200 €/yr$ respectively.

The operational expenses derive from the permeate pump (0.12 kW) and the air-compressor (0.12 kW) operation, the chemicals used for membrane cleaning, membrane replacement and the required labor. Continuous operation of the permeate pump for 22 h/d (25 d/month, 12 month/yr), gives total electricity consumption per year equal to $(22 \text{ h/d}) \times (25 \text{ d/m}) \times (12 \text{ m/yr}) \times (0.12 \text{ kW}) = 792$ kWh. Accordingly, the operation of the air-compressor for 3 h/d results in $(3 \text{ h/d}) \times (25 \text{ d/m}) \times (12 \text{ m/yr}) \times (0.12 \text{ kW}) = 108$ kWh per year. Four membrane modules (required for raw sewage filtration) consume four times higher electrical energy.

Cleaning of the membranes requires chlorine, surfactant and citric acid. Cleaning with chlorine (0.25 L of...
commercial 5% NaOCl solution per cleaning) was accomplished once per week during secondary effluent filtration and the total annual consumption was determined at 12 L/yr per module. Cleaning with citric acid (0.50 kg per cleaning) was performed under the same frequency resulting in an annual consumption of 24 kg/yr. According to Table 4, the chemical cleaning costs were estimated to be in the order of 30 €/module/yr, which corresponds to 0.03 €/m³ of secondary effluent, and −0.13 €/m³ of raw sewage. In the second case, similar costs (~0.15 €/m³) were reported for chemical cleaning by Bourgeois et al. (2005).

According to the supplier, the membrane replacement is necessary once per 5 years and the costs involved are 500 €/module. The membranes will be replaced twice over the projected 15 year lifespan of the plant. Thus, an annual cost of approximately 80 € was used for the calculations for each membrane module.

The data presented in Table 4 do not include the required labor. The supervision and operation of the microfiltration process includes monitoring of the permeate flux, macroscopic observation of the permeate and concentrate quality, adjustment of the operational parameters (aeration frequency, etc) and plant cleaning (once per week).

From the results presented in Table 4 it is evident that the recovery of reusable water from raw sewage, in one step (through PMF) is not economical, under the conditions of this study.

However, a small-scale conventional aerobic treatment plant designed for 3 m³/d, costs ~5,000 € (0.55 €/m³) including the tanks, the air-compressor (0.24 kW) and the installation. The process is completely aerobic and the OPEX (mainly due to aeration) are in the order of 1728 kWh/yr or 122 €/yr (0.13 €/m³). Because of the additional expenditures for the conventional aerobic treatment (CAPEX + OPEX = 0.68 €/m³) and the reuse (MF) treatment (~0.75 €/m³, see Table 4), the overall process to recover water is costly (1.45 €/m³) and moreover complex.

In Table 4, the CAPEX and OPEX for an MBR treating 3 m³/d sewage are included for comparison. The calculations were performed according to the following assumptions: a design permeate flux equal to 15 L/(m² h) (given by the supplier), two membrane modules with capital cost 7,500 €, a continuous aerated process (24 h/d operation of two air-compressors, 0.12 kW each), similar cleaning frequency with the secondary effluent process (weekly basis). Surprisingly, the total costs (CAPEX and OPEX) for water processing using the membrane bioreactor (MBR) were in the same range (1.34 €/m³) with the conventional activated sludge/MF process. Moreover, the process is much more simple and compact.

However, for an objective evaluation of the reuse technology the cost of freshwater, which is in most cases used for gardening, irrigation and secondary applications, should be also taken into consideration. The latter may vary from 0.5 up to 3 €/m³ in different EU countries (www.oecd.org), while water importation can raise the costs up to 5–7 €/m³, like in case of the Greek islands (Gikas & Tchobanoglous 2009b). Thus, the higher the price of freshwater and the lower its availability, makes water recovery from municipal wastewater via microfiltration an interesting option.

CONCLUSIONS

Primary membrane filtration (PMF) of municipal wastewater was possible using a flat-sheet membrane module with aeration as a means for fouling control. According to short-term permeation experiments, the fluxes obtained during PMF of sewage were significantly lower (~7 L/(m² h)) compared to the filtration of secondary effluent (~29 L/(m² h)) and as a consequence the total costs (capital and operational) for water recovery were estimated to be in the order of 2 and 0.75 €/m³ respectively. PMF was capable to achieve low permeate COD and SS qualifying disposal standards for non-sensitive environments. This is very important since the water can be treated without the need for aerobic biological processes. However, novel membrane technologies are needed to make this a viable process.

ACKNOWLEDGEMENTS

We would like to thank ATB Umweltechnologien GmbH and Dipl. Ing. Andreas Parisis (ENYA Engineering, Greece) for supplying the membrane module, A. Kapagiannidis and E. Gazani for the assistance with the laboratory work, and Prof. Verstraete for the fruitful discussions.
REFERENCES


Gikas, P. & Tchobanoglous, G. 2009a The role of satellite and decentralized strategies in water resources management. *J. Environ. Manage.* 90, 144–152.


