Fouling mitigation in anaerobic membrane bioreactors using fluidized resin beads

B. Düppenbecker, J. Behnisch, M. Engelhart and P. Cornel

ABSTRACT

This study focuses on the use of fluidized resin beads to mitigate fouling during ultrafiltration (UF) of the effluent of an anaerobic bioreactor. Two different module configurations were tested: A fluidized bed of resin beads was generated in a tubular UF membrane, and a hollow fiber (HF) UF membrane was submerged into a fluidized bed, respectively. During filtration of anaerobically treated synthetic wastewater using the tubular module, fluidized resin beads with a diameter of 0.5–0.71 mm did not show any beneficial effect. In contrast, the presence of fluidized resin beads (diameter of 0.5–0.71 and 1.00–1.25 mm) in the HF module reduced the fouling rate significantly. Furthermore, particle diameter and the bed voidage affected the cleaning efficiency of a pre-fouled membrane in the HF module. Interestingly, short-term filtration tests (<2 h) of a dextran solution showed that fluidized resin beads are able to minimize concentration polarization of a macromolecule, even in the tubular module. Therefore, it is supposed that fouling of the anaerobically treated synthetic wastewater was mainly attributed to the deposition of colloidal and particulate matter.

Key words | anaerobic membrane bioreactor, energy demand, fluidized bed, fouling, resin beads

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INTRODUCTION

A promising approach to mitigate fouling in anaerobic membrane bioreactors (AMBR) might be the use of fluidized granular activated carbon (GAC) as a turbulence promoter and mechanical cleaning agent, respectively (Kim *et al.* 2011; Yoo *et al.* 2012; Gao *et al.* 2014; Shin *et al.* 2014). However, fluidized GAC can damage the membrane surface and release fine particles that will block membrane pores irreversibly (Wu *et al.* 2014; Shin *et al.* 2016). Therefore, it appears appropriate to replace GAC by particles with comparable density, but with a smoother surface and a higher resistance to mechanical degradation. A previous study showed that resin beads represent a promising option (Imasaka *et al.* 1989) to mitigate fouling in AMBRs.

Fouling mitigation by fluidized beds in AMBRs can be implemented by (1) submerging a hollow fiber (HF) or flat-sheet membrane into a fluidized bed or by (2) creating a fluidized bed inside a tubular membrane. The first case was widely investigated using GAC as fluidization medium, for instance by Kim *et al.* (2011), Yoo *et al.* (2012), Gao *et al.* (2014) and Shin *et al.* (2014). In contrast, only a few studies focused on the second case; Seib *et al.* (2016) used fluidized GAC and Düppenbecker *et al.* (2016) fluidized glass beads, respectively. However, so far no study has focused on the comparison of both configurations.

Although the effect of fluidized GAC on fouling has extensively been investigated recently (Aslam *et al.* 2014; Wang *et al.* 2016a; Wang *et al.* 2016b; Wang *et al.* 2017; Wu *et al.* 2016; Wu *et al.* 2017), the impact of fluidized bed parameters (bed voidage, upflow velocity, particle diameter and density, particle shape, etc.) on fouling mitigation is not yet understood in detail. Van der Waal *et al.* (1977), Rios *et al.* (1987) and Noordman *et al.* (2002) investigated the use of fluidized glass and steel beads as turbulence promoters in tubular membranes during filtration of macromolecular solutions. Rios *et al.* (1987) reported an optimum bed voidage of about 68% (highest flux and rejection at constant pressure filtration), which can be explained by the fact that the wall-to-liquid mass transfer in fluidized beds depends on the bed voidage and reaches a maximum at about 70% (Schmidt *et al.* 1999). Models exist to predict the wall-to-liquid mass transfer (e.g. Schmidt *et al.* 1999); however, it must be considered that these models are valid only in case diffusion is the driving force, i.e. for

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molecular compounds and colloids <0.1 μm (Belfort *et al.* 1994; Davies 1992). Mikulášek & Filandrová (1995) reported an optimum bed voidage of 80% during filtration of alumina suspensions (particle diameter about 0.5–1.0 μm) using fluidized glass beads with a diameter of 1.5 mm. Furthermore, de Boer *et al.* (1980) observed an optimum bed voidage of about 70% during filtration of cheese whey using fluidized glass beads with diameters between 0.7 and 3 mm. More recently, Wu *et al.* (2017) showed that the scouring action of fluidized GAC is a function of the mass concentration (i.e. the bed voidage). De Boer *et al.* (1980) and Wu *et al.* (2016), using fluidized glass beads and GAC, respectively, reported an influence of the particle diameter as well.

Based on these results, the present study focuses on the suitability of fluidized resin to mitigate fouling in AMBRs. The influence of bed voidage and particle diameter on the cleaning efficiency of fluidized resin beads was investigated. A further objective of the present study is the comparison of the cleaning efficiency in the following configurations: (1) HF membrane placed into a fluidized bed and (2) generation of a fluidized bed inside a tubular membrane. Furthermore, to focus on the effect of fluidized resin beads on concentration polarization in case diffusion is the driving force, filtration tests with a dextran (as a surrogate for macromolecular wastewater compounds) solution were carried out as well.

MATERIAL AND METHODS

Experimental set-up

Experiments were carried out with a polymeric tubular membrane (ID = 8 mm, L = 1,000 mm, mean pore size = 0.03 μm, PVDF, Pentair X-Flow, The Netherlands) and a bundle of five polymeric HF membranes (OD = 2.6 mm, L = 300 mm, mean pore size = 0.03 μm, PVDF, Koch Membrane Systems, USA). The resulting total membrane surface areas were 0.025 m² (tubular module) and 0.012 m² (HF module). The tubular membrane was fitted into a housing made of acrylic glass. The HF membrane bundle was submerged into an acrylic glass tube (ID = 34 mm, L = 565 mm). The membranes were cleaned after each experiment, but not replaced. To monitor the effectivity of the cleaning procedure the clean water permeability was determined before each experiment. Figure 1 shows a schematic diagram of the laboratory setup. A fluidized bed of resin beads was generated in the tubular membrane and the HF membrane was submerged into a fluidized bed, respectively. Tubing pumps (Ecoline, Ismatec) were used for feed recirculation and permeate drawing. Permeate mass flow was measured gravimetrically using a balance (Entris, Sartorius). The volume flow of the feed line was measured with a flowmeter (Gemü 855, Gemü (HF module) and RGC 2207, Yokogawa (tubular module)). Pressure transmitters (33 X, Keller) were installed at the feed and the permeate line to monitor the transmembrane pressure (TMP).

The bed voidage of the fluidized bed was varied by changing the resin beads' mass and the crossflow velocity. Hence, the depth of the fluidized bed was kept constant. In the case of the tubular membrane module, the height of the fluidized bed was controlled by an inspection glass at the top of the membrane module. The resin beads were not recirculated. Since the fluidized beds showed a sharp boundary, further efforts to avoid discharge of the resin beads from the membrane modules were not required.

Strongly acidic gel-type cation exchange resin beads (Resinex K-8, Jacobi Carbons, Sweden) were used as fluidizing

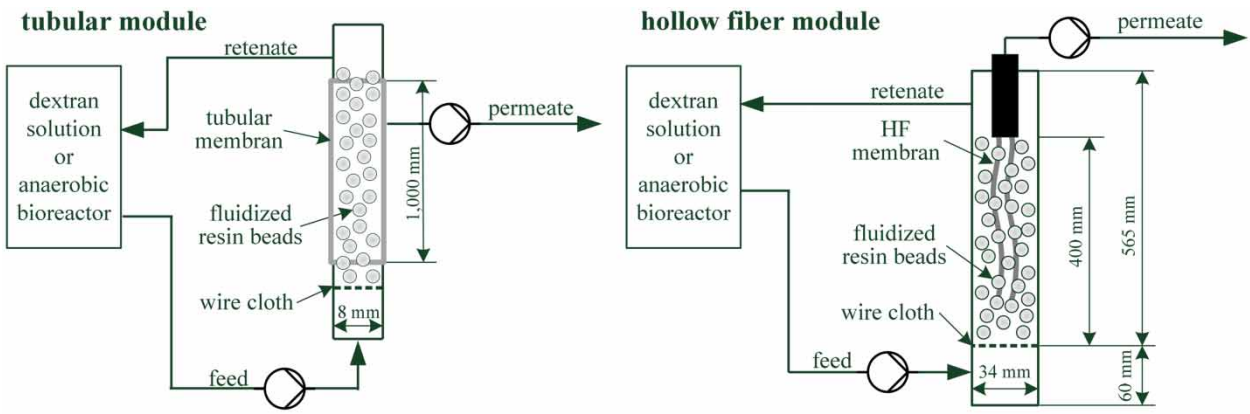

Figure 1 | Schematic diagram of the laboratory setups.

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medium. The resin beads were used as delivered in the Na^+-form (dextran filtration tests) or in Ca^+-form (filtration tests with anaerobically treated synthetic wastewater). The resin bead diameters were 0.5–0.71 and 1.00–1.25 mm. The density of the wet resin beads (ρ_P) was determined to be about 1,260 kg/m^3. The resin beads showed no ability for the sorption of dextran and organic matter of the anaerobic bioreactor effluent (data not shown). All experiments were carried out at 20 °C.

Filtration of bioreactor effluent

The above-mentioned membrane modules were coupled to an anaerobic fluidized bed (FB) reactor. The reactor was described in detail by Düppenbecker & Cornel (2016) and was fed with a synthetic medium containing peptone (200 mg/l), yeast extract (200 mg/l), K_2HPO_4 (17 mg/l), NaCl (113 mg/l), $CaCl_2{\cdot}2H_2O$ (110 mg/l), $MgSO_4{\cdot}7H_2O$ (100 mg/l), sodium acetate (67 mg/l), glutamic acid (50 mg/l), tartaric acid (10 mg/l), and tap water. The mean total chemical oxygen demand (COD_{tot}) of the medium during the filtration tests was 536 mg/l and consisted only of dissolved COD (COD_{diss}). The operating conditions are shown in Table 1. The organic loading rate (OLR) was higher during the filtration test with the tubular membrane and varied due to fluctuations of the feed volume flow. After each experiment, the membrane was cleaned with a NaOCl solution (195 ppm) and citric acid (1% monohydrate by mass).

COD_{diss} was determined after filtration with a 0.45 μm syringe filter (PES, VWR). Samples were homogenized with a dispersion tool (S25N-10G, IKA) before determining COD_{tot}. The particulate COD (COD_{part}) results from the difference between COD_{tot} and COD_{diss}. COD was determined by cuvette tests (Hach). Before analyzing samples of the feed (equals the effluent of the FB), retentate, and permeate, hydrogen sulfide and dissolved methane were stripped according to Düppenbecker & Cornel (2016). Membrane rejection (%) was calculated according to Equation (1). Here, c_{Feed} is the feed COD concentration and $c_{Permeate}$ the permeate COD concentration.

$$\text{Rejection} = \frac{(c_{Feed} - c_{Permate})}{c_{Feed} \cdot 100} \quad (1)$$

Dextran filtration tests

To investigate solely the effect of fluidized resin beads on concentration polarization of a macromolecular compound, filtration tests with a dextran solution were carried out. Dextran is a neutral homopolymer of glucose (Heinze *et al.* 2006) and widely used as model solute to characterize UF membranes (Tkacik & Michales 1991; Bakhshayeshi *et al.* 2012). A 0.1% (by mass) solution of dextran (500.000 Da, Dextran 500, Serva Electrophoresis) and $NaNO_3$ (Emsure, Merck) was prepared using DI water. The prepared solutions were 0.45 μm filtered (Whatman ME 25, GE) before use. Operating conditions of the dextran filtration tests are shown in Table 2. After each experiment, the membrane was cleaned with a NaOCl solution (195 ppm).

Membrane rejection was calculated according to Equation (1) using the total organic carbon (TOC) concentrations of the feed and permeate. TOC of dextran solutions was determined using a vario TOC Cube (Elementar). Feed samples were 0.45 μm filtered (PES syringe filter, VWR) before analysis.

Prediction of pumping power required for fluidization

The pumping power required for fluidization P (W) of the resin beads was determined according to Equation (2)

Table 1 | Operating conditions during filtration of bioreactor effluent

Membrane module (–)	Flux (l/(m²·h))	Diameter resin beads (mm)	Mass resin beads (g)	Bed voidage (%)	Crossflow velocity (m/h)	OLR[a] (kg COD/(m³ d))
Tubular	13.1	0.50–0.71	10	75	32	10.8
Tubular	13.2	–	0	100	31	12.4
HF	12.0	–	0	100	37	8.5
HF	12.3	1.00–1.25	100	58	37	8.3
HF	12.0	1.00–1.25	130	45	17	8.2
HF	12.0	–	0	100	17	7.5
HF	11.9	0.50–0.71	90	63	18	9.0

[a]Organic loading rate of bioreactor.

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Table 2 | Operating conditions of dextran filtration tests

Membrane module (–)	Flux (l/(m²·h))	Diameter resin beads (mm)	Mass resin beads (g)	Bed voidage (%)	Crossflow velocity (m/h)
Tubular	14.6	0.50–0.71	10	72	30
Tubular	12.6	0.50–0.71	5	86	45
Tubular	12.0	–	0	100	45
HF	10.1	–	0	100	41
HF	9.9	1.00–1.25	125	50	19
HF	9.6	1.00–1.25	90	63	41

(Zamani *et al.* 2015). Here, Q is the volume flow (m³/s) of the recirculation required for fluidization and Δp (Pa) the pressure loss across the fluidized bed.

$$P = Q\Delta p \tag{2}$$

The pressure loss of the fluidized bed can be calculated according to Equation (3) (Epstein 2003). Here, H (m) is the depth of the fluidized bed, ϵ (–) the bed voidage (the ratio of the void volume to the total volume of the bed), ρ_P the density of the fluidized particles, ρ_L (kg/m^3) the density of the fluid and g (m/s^2) the acceleration of gravity. Due to the fact that the product of bed height and particle concentration is constant for a fluidized bed (Epstein 2003), the pressure drop is determined by the initial depth of the packed bed.

$$\Delta p = H(1-\varepsilon)(\rho_P - \rho_L)g \tag{3}$$

The volume flow Q in Equation (2) can be calculated predicting the superficial liquid velocity u (m/s), which is required to achieve the desired expansion of the fluidized bed, according to Equation (4) (Epstein 2003). By assuming that the resin beads are smooth spheres, the free-settling velocity u_0 was calculated according to Turton & Clark (1987). Furthermore, a mean particle diameter of 1.1 mm and 0.6 mm (for particles with diameters between 1.00–1.25 mm and 0.50–0.71 mm, respectively), and a density of 1,000 kg/m³ for water (ρ_L) were assumed. The parameters n (–) and k (–) in Equation (4) were calculated according to Khan & Richardson (1989).

$$u/u_0 = k\varepsilon^n \tag{4}$$

RESULTS AND DISCUSSION

Filtration of bioreactor effluent

The results of the filtration test using the tubular module are shown in Figure 2. The presence of fluidized resin beads did not show any beneficial effect (Figure 2(a)): after 6 operating hours, the TMP increased from about 1 kPa to 25.4 kPa (with resin beads) and 22.2 kPa (without resin beads). Moreover, the addition of fluidized resin beads during empty tube filtration after 6 h (bed voidage =75%) did not show any beneficial effect, as the slope of the TMP between operating hour 6 and 10 (Figure 2(a)) shows. Hence, fluidized resin beads did not mitigate fouling under the given operating parameters. The distinct drop of the TMP after 6 h was caused by cake layer removal during the addition of the resin beads (Figure 2(a)). Since permeate drawing was interrupted, the cake layer could easily break away from the membrane surface. Note that,

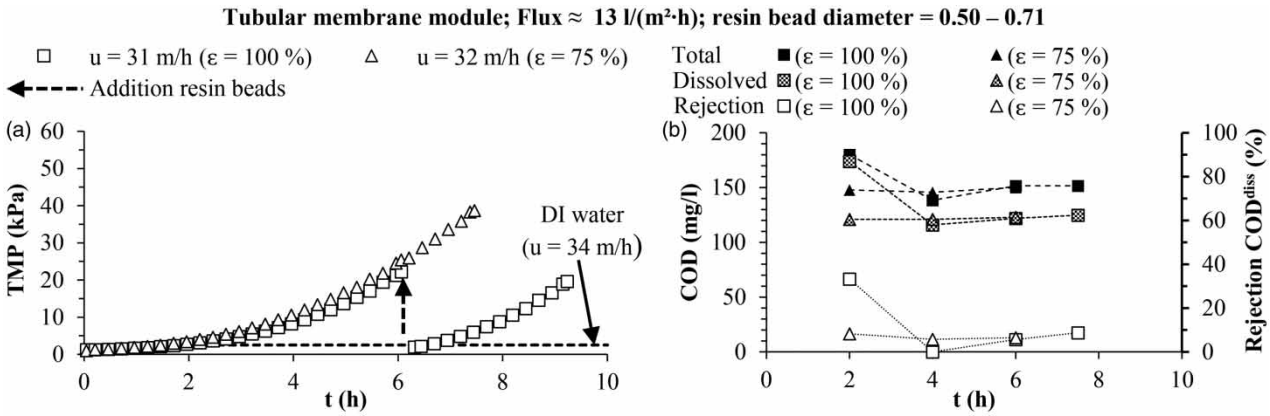

Figure 2 | TMP during filtration of bioreactor effluent with tubular membrane in the presence and absence ($\varepsilon = 100\%$) of fluidized resin beads (a). After 6 hours, the empty-tube filtration test was stopped and resin beads were added (marked with arrow); a bed voidage of $\varepsilon = 75\%$ was adjusted. U is the crossflow velocity in the membrane module. COD_{tot} and COD_{diss} concentrations of the feed (inlet membrane module) and COD_{diss} rejection (b).

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due to plugging in the presence of gas bubbles, it was not possible to create a stable fluidized bed with a lower bed voidage or with larger particles. However, the applied bed voidage of 75% is in the range of the optimum reported by van der Waal *et al.* (1977), de Boer *et al.* (1980), Rios *et al.* (1987) and Mikulášek & Filandrová (1995). The COD_{diss} rejection of the tubular membrane was below 10%, except that at the beginning of the filtration test without fluidized resin beads a rejection of about 30% was observed (Figure 2(b)).

Seib *et al.* (2016) examined the effect of fluidized GAC (0.6–1.7 mm) while filtrating treated synthetic wastewater with tubular UF membranes. The fluidized GAC did not show any positive effect in a ceramic tubular membrane with an inner diameter of 16 mm (Seib *et al.* 2016). In contrast, fouling in a polymeric tubular UF membrane with an inner diameter of 12.5 mm was reduced by the fluidized GAC (Seib *et al.* 2016). The results of Seib *et al.* (2016) might indicate that fluidized GAC is more effective in terms of fouling mitigation than fluidized resin beads. However, the diameter of the GAC particles (0.6–1.7 mm) used by Seib *et al.* (2016) was larger than that of the resin beads (0.5–0.71 mm) used in the present study. Consequently, the crossflow velocities in the study of Seib *et al.* (2016) were higher (between 64 and 90 m/h) than in the present study.

In contrast, in the case of the HF membrane, fluidized resin beads mitigated fouling distinctly. As Figure 3(a), 3(c) and 3(e) show, TMP increased during empty tube filtration at crossflow velocities of 17 m/h (bed voidage = 45%) and 37 m/h (bed voidage = 58%) up to about 60 kPa within 11–12 h. When fluidized resin beads were employed, there was only a slight increase in TMP within 8 h under all tested operating conditions (Figure 3(a), 3(c) and 3(e)). In the absence of fluidized resin beads, the fouling rates (calculated based on the change in TMP within the first 8 operating hours) were 29.7 Pa/min (u = 37 m/h) and 35.9 Pa/min (u = 17 m/h). In the same time interval, the fouling rates were 0.5 Pa/min (u = 37 m/h) and 0.6 Pa/min (u = 17 m/h) in the presence of fluidized resin beads with a diameter between 1.00 and 1.25 mm. A slightly higher fouling rate of 2.0 Pa/min (u = 18 m/h) was observed for the smaller resin beads. Hence, fluidized resin beads, even with diameters between 0.5 and 0.71 mm, are able to prevent fouling in the HF membrane module. The results indicate that it seems to be more effective to submerge a membrane in a fluidized bed than creating a fluidized bed in a tubular membrane.

Due to fluctuating influent COD concentrations (Figure 3(b), 3(d) and 3(f)), it might be possible that the fouling ability of the feed varied as well. To exclude any effect of the feed characteristics and irreversible membrane fouling (which could not be removed by chemical cleaning), the fluidization was stopped after 8 h and restarted as soon as the TMP reached a value of about 10 kPa. Hence, the fouling rate of the feed could be quantified and in each experiment, the membrane was fouled to a comparable level before restarting the fluidization. As Figure 3(a) shows, using particles with a diameter of 1.00–1.25 mm at a bed voidage of 58% showed the strongest cleaning efficiency. Even though the highest fouling rate (after stopping fluidization) was observed during this experiment, which might be attributed to the increased COD concentration of the feed, after restarting fluidization, the TMP decreased nearly to the initial level. Kim *et al.* (2011) reported total recovery of the membrane permeability after restart of GAC fluidization. In the present study, a slight increase in TMP after 20 h indicates a weaker cleaning efficiency of the resin beads. The stronger effect of fluidized GAC might be attributed to higher density, different shape, and higher material hardness of GAC. At lower crossflow velocity (u = 17 m/h) and bed voidage ($\varepsilon = 45\%$), respectively, after restart of fluidization only a slight decrease in TMP was observed until the TMP started to increase further (Figure 3(c)). Using smaller particles at a bed voidage of 63% showed a comparably poor cleaning efficiency (Figure 3(e)). Hence, the cleaning efficiency seems to increase with increasing bed voidage (at a constant particle diameter) and particle diameter (at a constant bed voidage). The first observation agrees well with the results of Mikulášek & Filandrová (1995), Rios *et al.* (1987) and de Boer *et al.* (1980). However, it should be mentioned that at a bed voidage of about 70–80%, a transition point was observed and a further increase of the bed voidage resulted in lower fouling mitigation (de Boer *et al.* 1980; Rios *et al.* 1987; Mikulášek & Filandrová 1995). The influence of the particle diameter was previously reported by de Boer *et al.* (1980) and Wu *et al.* (2016) using fluidized glass beads and GAC, respectively.

Dextran filtration tests

In the absence of fluidized resin beads, TMP increased (u = 45 m/h) up to about 70 kPa within one hour when filtrating with the tubular membrane (see Figure 4(a)). Interestingly, fluidized resin beads mitigated fouling distinctly. In the presence of fluidized resin beads, the fouling rates (calculated based on the change in TMP within the first operating hour) were 15.8 Pa/min (u = 30 m/h) and 16.2 Pa/min (u = 45 m/h). In contrast, in the same time interval, the fouling rate was 1,193.9 Pa/min (u = 45 m/h) for empty tube filtration. Furthermore, TOC rejection in the presence of fluidized resin beads was

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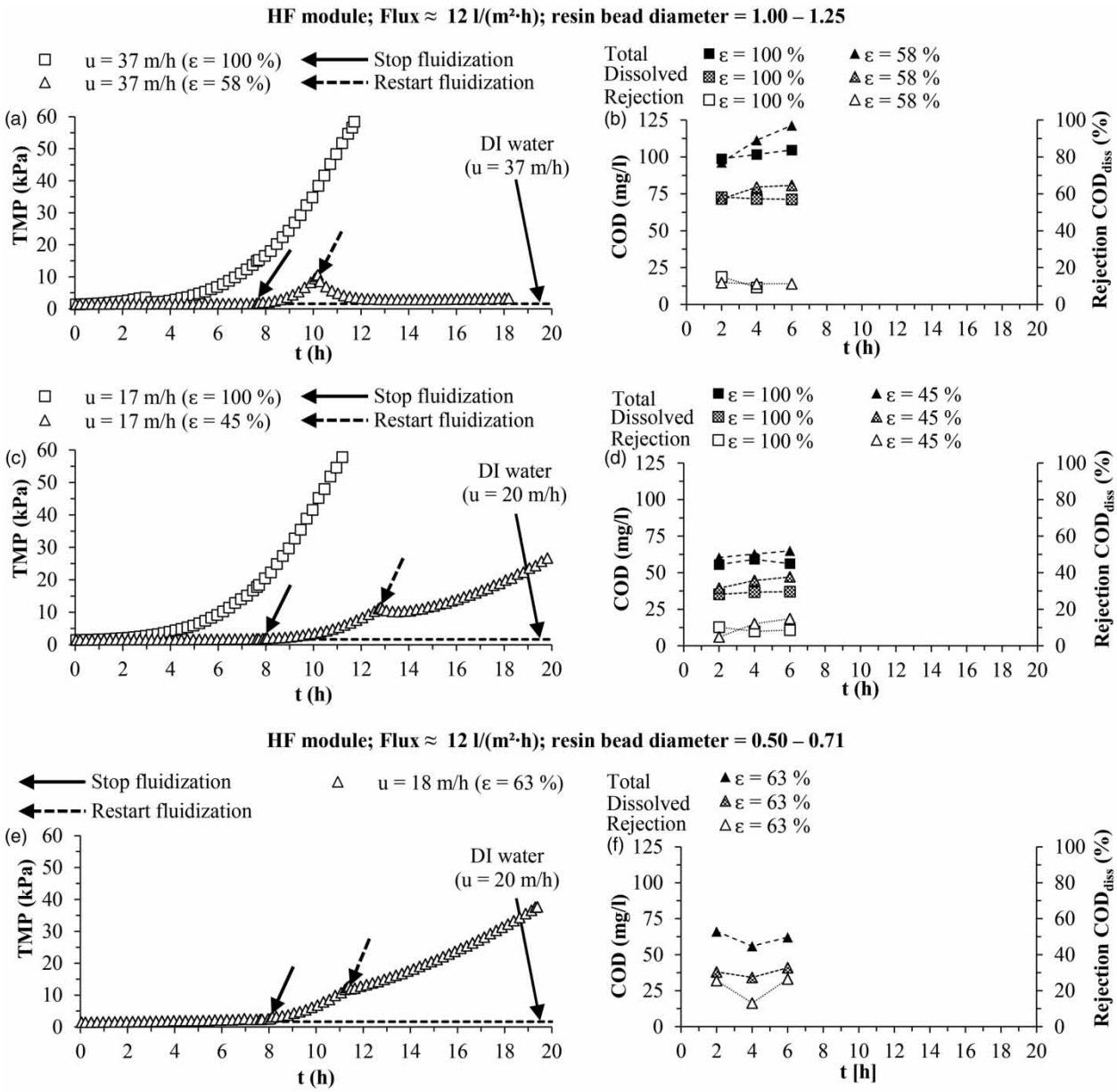

Figure 3 | TMP during filtration of bioreactor effluent with HF module in the presence and absence ($\varepsilon = 100\%$) of fluidized resin beads (a), (c), (e). U is the crossflow velocity in the membrane module. COD_{tot} and COD_{diss} concentrations of the feed (inlet membrane module) and COD_{diss} rejection (b), (d), (f).

about 60% as Figure 4(b) shows. In contrast, during empty tube filtration the rejection decreased to about 20%. The higher rejection for the tubular membrane at the beginning can be attributed to washout of clean water from the shell-side volume of the membrane module. The low fouling rate and the distinctly higher rejection in the presence of fluidized resin beads leads to the conclusion that fluidized resin beads could effectively minimize concentration polarization. This is a remarkable fact, considering the above discussed result that fluidized resin beads did not show any positive effect during filtration of the bioreactor effluent using the tubular module (see Figure 2(a)). It is assumed that fouling during filtration of the bioreactor effluent is mainly attributed to deposition of colloids and particles on the membrane surface. Apparently, the fluidized resin beads were not able to detach the resulting cake layer by scouring under given operating conditions in the tubular module. A possible reason for this phenomenon might be non-uniform lateral particle concentrations. A CFD (computational fluid dynamics) study on the

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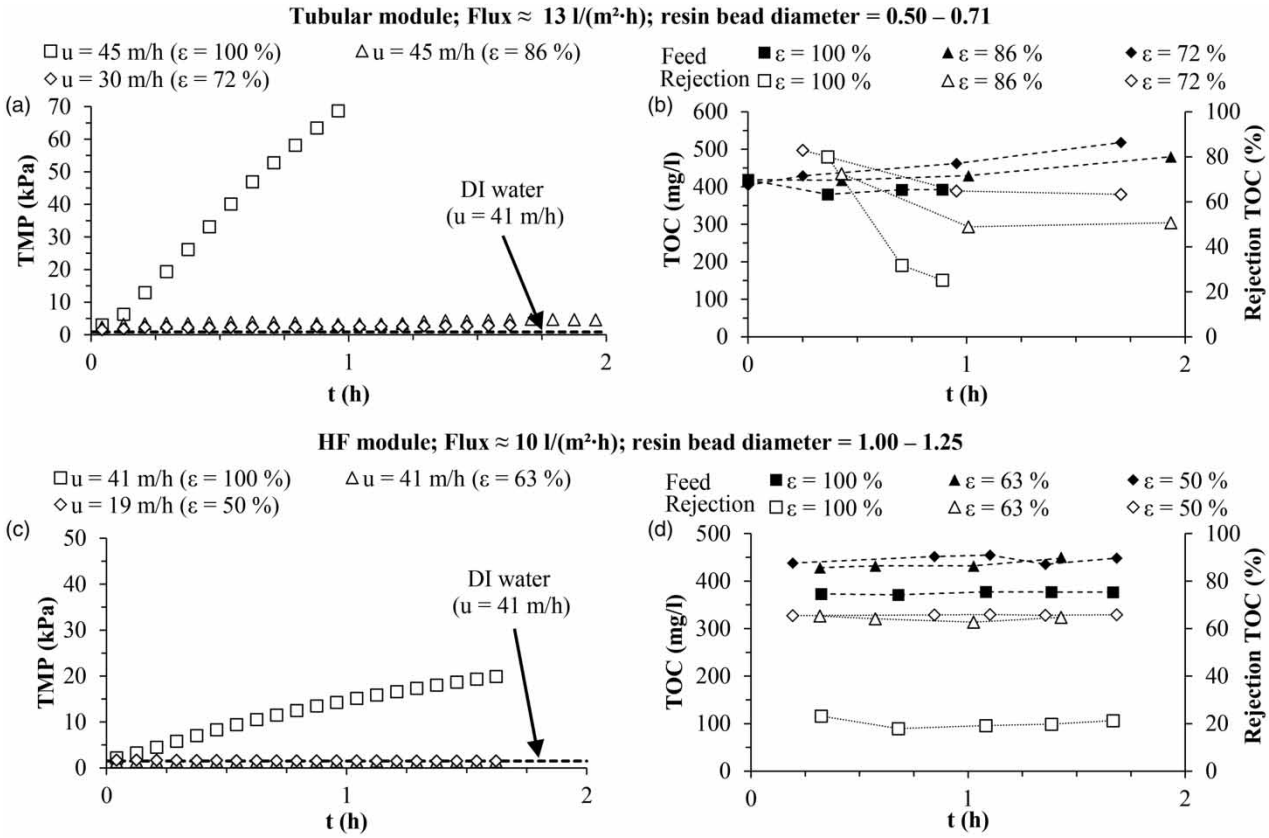

Figure 4 | TMP increase during filtration (u = crossflow velocity) of dextran solution with tubular (a) and HF (c) membrane in the presence and absence ($\varepsilon = 100\%$) of fluidized resin beads. Total organic carbon (TOC) concentration of the feed (inlet membrane module) and TOC rejection (b) and (d).

hydrodynamics of fluidized GAC indicated insignificant particle concentrations near the column wall (Cahyadi *et al.* 2017). Hence, the scouring effect of the fluidized resin beads might be limited near the surface of the tubular membrane. However, Cahyadi *et al.* (2017) considered only a two-dimensional system and assumed idealized uniform inlet flow.

The strong effect of the resin beads on mitigation of concentration polarization was confirmed during the experiments with the HF membrane module. As Figure 4(c) and 4(d) show, in the presence of fluidized resin beads no increase in TMP (Figure 4(c)) and a distinctly higher TOC rejection (Figure 4(d)) were observed.

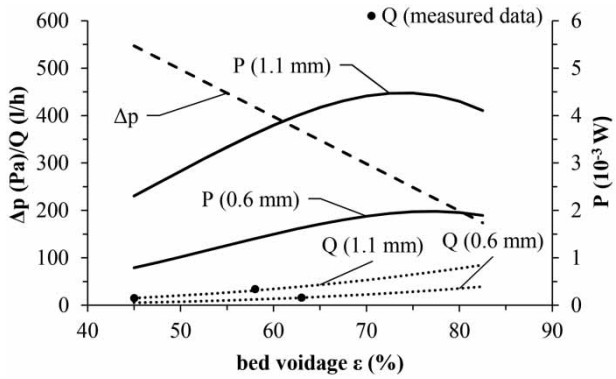

Figure 5 | Predicted (theoretical) volume flow, pressure loss and required pumping power (HF membrane module) at filtration of bioreactor effluent. A mean particle diameter of 1.1 and 0.6 mm was assumed. Additionally, measured data for volume flow are shown (dots).

Energy considerations

The predicted (theoretical) volume flow, pressure loss (of the fluidized bed), and required pumping for the HF module are shown in Figure 5. Furthermore, experimental data of the volume flow are shown. The depth of the fluidized bed was kept constant at 0.39 m (required to cover the membrane while filtrating bioreactor effluent). Consequently, the initial packed bed depth and the pressure drop, respectively, decreased with increasing bed voidage. In contrast, the volume flow required for fluidization increased with increasing bed voidage. The required

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pumping power (the product of both) increased initially with increasing bed voidage and reached a maximum at approximately 75% (1.1 mm bead diameter) and 77.5% (0.6 mm bead diameter). A further increase in bed voidage would result in a decreasing pumping power demand. Consequently, the required pumping power for fluidization can be reduced by adjusting bed voidage below or above the maximum between 75 and 77.5%. However, a minimum fouling rate was reported at a bed voidage between about 70 and 80% by de Boer *et al.* (1980), Rios *et al.* (1987), and Mikulášek & Filandrová (1995) using glass and steel beads, respectively. Furthermore, Wang *et al.* (2017) showed that the effect of GAC on fouling mitigation of latex particles (5 μm) correlates positively with the power input at given particle size, indicating an optimum mass transfer at bed voidage between 70 and 75%. Hence, at given bead diameter the fouling rate tends to correlate positively with bed voidage and power input, respectively.

Table 3 compiles the required pumping energy during filtration of the bioreactor effluent using the HF membrane module. At a bead diameter of 1.00–1.25 mm and a bed voidage of 58%, both the required pumping energy (0.027 kWh/m³) and the cleaning efficiency of the pre-fouled membrane is highest (see Figure 3(a)). By decreasing the bed voidage to 45% the required pumping energy is reduced to 0.016 kWh/m³, but the cleaning efficiency decreased as well (see Figure 3(c)). At a bead diameter of 0.5–0.71 mm and a bed voidage of 63%, the required pumping energy is lowest (0.011 kW/h). However, at this operating state, the cleaning efficiency was lowest as well (Figure 3(e)). Hence, in the present study, the cleaning efficiency of the fluidized resin beads tends to correlate positively with the power input.

Overall, the resulting electrical energy demand for the filtration process is between 0.020 and 0.046 kWh/m³, which is distinctly lower than reported values between 0.25 and 7.3 kWh/m³ for conventional AMBR processes (Liao *et al.* 2006). However, note that the predicted energy demand of the present study based on laboratory-scale experiments and comparison of these data with data from pilot and large-scale plants is only possible to a limited extent. Yoo *et al.* (2012) achieved a value of 0.036 kWh/m^3, which lies within the range of the present study, for a laboratory-scale AMBR using fluidized GAC for fouling mitigation.

CONCLUSIONS

Fouling during filtration of anaerobically treated synthetic wastewater was distinctly reduced by placing an HF

Table 3 | Required energy for fouling control (HF membrane)

Bead diameter	(mm)	1.00–1.25	1.00–1.25	0.50–0.71
Bed voidage	(%)	58	45	63
Fluidization resin beads				
Volume flow recirculation[a]	(l/h)	34.0	15.0	16.0
Pressure loss[b]	(kPa)	0.42	0.55	0.37
Required pumping power	(W)	0.004	0.002	0.002
Required pumping energy[c]	(kWh/m³)	0.027	0.016	0.011
Permeate drawing				
Permeate flow	(l/h)	0.14	0.14	0.14
Pressure loss[d]	(kPa)	1.6	1.6	1.6
Required pumping power	(W)	0.00006	0.00006	0.00006
Required pumping energy[c]	(kWh/m³)	0.00044	0.00044	0.00044
Total				
Required pumping energy	(kWh/m³)	0.028	0.016	0.012
Required electrical energy[e]	(kWh/m³)	0.046	0.027	0.020

[a]Based on measured data.
[b]Predicted according to Equation (3).
[c]Related to permeate volume flow at flux of 12 l/(m² h).
[d]Assuming complete fouling prevention.
[e]Assuming pump efficiency of 60% (Martin *et al.* 2011).

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module in a bed of fluidized resins beads. The cleaning efficacy of a pre-fouled membrane seems to depend on the bed voidage and the particle diameter. At a bed voidage of 58%, the effect was distinctly stronger than at a bed voidage of 45%. Resin beads with a diameter of 1.00–1.25 mm were more effective than particles with a diameter of 0.5–0.71 mm at nearly the same bed voidage. In contrast, when generating a fluidized bed within a tubular membrane module, no beneficial effect was observed. Presumably, the fluidized resin beads could not detach deposited particular matter from the tubular membrane's surface. Interestingly, fluidized resin beads mitigated concentration polarization of dextran in both modules.

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