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Impact of membrane configuration on fouling in anaerobic membrane bioreactors

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ABSTRACT

The filtration performance of flocculated and granulated configured anaerobic membrane bioreactors (MBR) treating domestic wastewater has been evaluated and compared to conventional aerobic MBR. Immersed hollow fibre (HF) and external tubular membrane geometries were additionally compared with the latter operated in both pumped and gas-lift mode. After 200 d of operation, both granular and flocculated anaerobic MBR (AnMBR) suspensions were characterised by an increased population of colloidal particles while the aerobic MBR retained a unimodal particle size distribution with a d_{50} of 20 μm. Consequently, the flocculated AnMBR supernatant was characterised by a soluble microbial product (SMP) concentration ca. 500% higher than the aerobic MBR, such that the lowest critical fluxes for both HF and tubular membranes were recorded for the AnMBR. In comparison, the granulated AnMBR sludge was characterised by a low mixed liquor suspended solids concentration and an SMP concentration below 50% that of the flocculated anaerobic MBR. Consequently, similar fluxes to those of the aerobic MBR were achieved with the granulated anaerobic sludge using immersed HF membranes. Operating external tubular membranes in gas-lift appeared less effective for the granular AnMBR than the Aerobic MBR. However, critical fluxes >40 L m^{-2} h^{-1} were achieved using pumped mode. Results suggest granular AnMBR systems to be most suited to domestic wastewater treatment using either immersed HF membranes or external tubular membranes in pumped crossflow mode.

1. Introduction

Anaerobic treatment of domestic wastewater is constrained by the low organic strength of the wastewater, the quality of the available carbon and the high K_s value associated with the anaerobic community which impair bacterial growth and effective treatment. Anaerobic membrane bioreactor (AnMBR) technology can enhance effluent quality from increased rejection of solids and colloidal organic matter, and also achieve higher biomass concentrations by minimising washout. AnMBR thus increases the viability of anaerobic treatment, offering treatment at reduced energy demand and sludge yield over that of aerobic processes, for which the aeration energy demand is significant. However, AnMBR studies undertaken to date have generally found that membrane fouling is significant due to the high concentration of colloidal material, inclusive of protein and polysaccharide, the concentrations of which are further exacerbated at extended solids retention times (SRTs) [1].

Previous AnMBR fouling studies have primarily investigated the effect of pumped crossflow velocity on filtration performance in external tubular membranes operated under pumped [2–5] or gas-lift [6–10] condition. While gas sparging in immersed membrane systems is known to present the most cost effective means of fouling control in aerobic wastewater treatment applications, permeability data for anaerobic immersed flat-sheet and hollow fibre (HF) geometries is more limited [11–14]. Comparisons of liquid pumping and gas sparging in external tubular membranes [15] and gas sparging in immersed flat-sheet and HF membranes have been undertaken in AnMBR [11]; however, a direct comparison of external and immersed configured AnMBRs has yet to be made.

Direct comparison of current AnMBR studies also remains challenging as two reactor configurations now predominate: conventional completely mixed flocculated reactors [8–10] and upflow anaerobic sludge blanket (UASB) reactors [10,12,14]. In the latter case, membrane filtration is sited at the head of the UASB or in a subsequent stage, thus the membrane is challenged with the sludge supernatant. While UASB based AnMBR are less studied, lower irreversible fouling than completely mixed reactors has been demonstrated in a comparative study on the treatment of blackwater [10]. The aim of this present study is therefore to provide a direct comparison of external and immersed membrane operation in both flocculated and UASB configured AnMBR treating domestic wastewater. Previous AnMBR studies primarily utilised synthetic

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Table 1
Module characteristics for the tubular and hollow fibre membranes tested.

Parameter	Unit	Experimental reactor		MBR pilot plant		
Pilot plant				AeMBR	AnMBR	G-AnMBR
Geometry		Tubular	HF	HF	HF	HF
Filtration area	m^2	0.022	0.93	12.5	12.5	0.93
Material	–	PVDF	PVDF	PVDF	PVDF	PVDF
Pore size	μm	0.03	0.04	0.08	0.08	0.04
Module length	m	1	0.7	1	1	0.7
Fibre OD/LD	mm	8	1.9/0.8	1.3	1.3	1.9/0.8
CSA	m^2	–	0.0074	0.0177	0.0177	0.0074
Packing density	m^2/m^3	–	300	710	710	300

ID, internal diameter; OD, outer diameter; LD, lumen diameter; HF, hollow fibre.

wastewaters comprising of soluble carbon sources [9,11,14]. While such studies present significant insight, real wastewaters present more complex colloidal and particulate matrices. Furthermore, to date most published studies have used an external thermal source to stabilise reactor operation to mesophilic or thermophilic conditions [2–4]. Energetic modelling has demonstrated that methane generated from AnMBR treating domestic wastewater is insufficient to support heating the influent wastewater due to the low organic strength and high liquid flow rate [16]. Consequently, to provide environmentally relevant conditions, the flocculated AnMBR, granular AnMBR (G-AnMBR) and a reference aerobic MBR (AeMBR) were operated on real domestic wastewater and without external temperature control.

2. Materials and methods

2.1. Pilot plants

Three MBR comprising suspended growth (flocculated) aerobic and anaerobic MBR and a granular anaerobic MBR were operated in parallel for 250 d fed with wastewater having mean chemical oxygen demand (COD), biochemical oxygen demand (BOD_5), suspended solids (SS) and ammonia concentrations of 338 mg COD L^{-1} (range 197–553 mg COD L^{-1}), 167 mg BOD L^{-1} (range 155–285 mg BOD L^{-1}), 84 mg SS L^{-1} (range 51–186 mg SS L^{-1}) and 35 mg NH_4-N L^{-1} (range 15–48 mg NH_4-N L^{-1}) respectively. The AeMBR and AnMBR pilot plants comprised two tanks of 1.5 m^3 total volume divided between the biological (1 m^3) and membrane (0.5 m^3) compartments. Wastewater was introduced in the biological tank through a floating valve which controlled the level of sludge at a height of 1.5 m, creating a total working volume of 1.2 m^3. The AeMBR was continuously aerated at a flow rate of 50–100 L min^{-1} through cylindrical fine bubble diffusers at the biological tank base. The AnMBR tank was sealed with a PVC lid. Biomass was cycled between both compartments through external pumps, and an additional pump was operated in cycles of 15 min on/15 min off to mix the reactor contents by recycling biomass through venturi nozzles located at the base of this chamber. Both flocculated MBRs were fitted with polyvinyldifluoride (PVDF) HF membranes of 12.5 m^2 surface area, 0.08 μm nominal pore size, 1 m length, 0.0177 m^2 cross sectional area and a packing density of 710 m^2 m^{-3} (Table 1). Permeate was continuously extracted using peristaltic pumps (620 Du, Watson Marlow, Falmouth) to provide an instantaneous flux of 6 L m^{-2} h^{-1}. Membrane fouling was controlled by continuous gas sparging at gas velocities ranging from 0.02 m s^{-1} to 0.078 m s^{-1} in both aerobic and anaerobic systems, the latter employing nitrogen enriched air (>99% N_2) generated from an industrial gas membrane nitrogen generator unit (Atlas Copco, Herts, UK) fed with compressed air. During the 250-d trial, 12 L of sludge per day were withdrawn from the suspended aerobic and anaerobic MBRs in order to set SRT based on bioreactor volume at 100 d. This value could be taken as representative for the mean residence time in the anaerobic system since no sludge accumulation was observed. For the AeMBR, which was started with a low MLSS concentration of 1 g MLSS L^{-1}, the dynamic SRT reached a value of 80–85 d for the last 100 d of operation taking into account biomass accumulation within the reactor for the estimation of the sludge age.

The granular anaerobic MBR (G-AnMBR) consisted of a 85 L perspex cylindrical vessel (1.75 m depth × 0.25 m diameter) seeded with 40 L of granular sludge from an UASB sited at a sugar refinery (British Sugar, Suffolk, UK), with a 38 L perspex cylinder housing the membrane (Fig. 1). The granular UASB contactor was operated in the expanded mode (EGSB) using an external recirculation pump (620s, Watson Marlow, Falmouth, UK) to maintain a superficial upflow velocity (V_{up}) of less than 1 m h^{-1}. Effluent from the granule contactor was recirculated through the membrane tank during permeation. At the fixed upflow velocity, the granule bed expanded to a depth of 0.60 m, or approx. 30% of the column depth; the effluent entering into the membrane tank was characterised by a relatively low solids fraction compared to the flocculated reactors. A PVDF hollow-fibre module with a surface area of 0.93 m^2 and nominal pore size 0.04 μm, 1 m length, cross sectional area of 0.0074 m^2 and a packing density of 300 m^2 m^{-3} was used in this study at a set instantaneous flux (J) of 6 L m^{-2} h^{-1} to provide a biotank hydraulic retention time (HRT) of 16 h. In the G-AnMBR membrane fouling was controlled by gas sparging nitrogen at gas velocities ranging from 0 to 0.057 m s^{-1}. During the 250-d trial, no granular biomass was withdrawn from the biotank; samples were only collected from the membrane tank for analysis.

2.2. Short-term fouling experiments

Short term fouling experiments for the AeMBR and AnMBR were conducted in an external 38 L cylindrical tank (0.20 m diameter × 1.2 m height) using a 30 L slurry from either reactor (Fig. 2). For immersed trials, an identical PVDF HF membrane to that used in the granular AnMBR was tested (Table 1). A 0.022 m^2 tubular PVDF membrane with a nominal pore size of 0.03 μm was used for external (sidestream) experiments. The external tubular membrane was operated in both gas lift and pumped mode. During immersed and gas lift operation, nitrogen-enriched air for anaerobic experiments or natural air for aerobic experiments were employed for membrane scouring and biomass mixing. For gas lift experiments, the base of the membrane inlet was connected through a 'T' junction to provide a port for gas injection. The same range of specific gas demand, 0.2–1.2 m^3 m^{-2} h^{-1}, was used for both membrane geometries. For pumped crossflow operation, a centrifugal pump (EHEIM Gmbh and Co., Deizisau, Germany) was employed to generate a cross flow velocity (CFV) controlled by a throttling valve sited upstream of the membrane module to values between 0.4 and 2 m s^{-1}. For all studies both the retentate mixed liquor and permeate product streams from the membrane module were returned to the tank to maintain a constant mixed