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Direct contact membrane distillation for anaerobic effluent treatment

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ABSTRACT

This study was conducted using direct contact membrane distillation configuration for treating effluent from a thermophilic anaerobic membrane bioreactor. The effluent was taken as feed to better understand the effect of treating anaerobic wastewater (with and without biomass) on the flux and fouling interaction on membrane distillation in batch mode. While treating the effluent without biomass, permeate fluxes varied from 1.41 to 9.22 L/m^2 h at a temperature of (40–70 °C) and cross flow velocities of 0.005, 0.008, 0.01, 0.012 and 0.014 m/s. The COD and ammonia rejection at the stage was greater than 90%. In conditions where biomass was introduced in the feed at 6 and 12 g/L MLVSS, it was observed that permeate flux decreased from 2 to 0.6 L/m^2 h after 72 h operation at 12 g/L MLVSS wastewater, thus maximum observable fouling was also at 12 g/L. Fouling investigation concluded that the major foulant on the MD membrane surface was removable fouling, accounting for 71–77.5% of the total fouling at a loading rate of 6 and 12 g/L MLVSS, respectively. After cleaning, the membrane could recover 96% of the initial flux.

1. Introduction

High rate anaerobic treatment processes, such as upflow anaerobic sludge blanket (UASB), upflow anaerobic filter process and anaerobic fluidized-bed reactor have proven to be highly effective in treating medium-to-high strength industrial wastewater. Generally, these industrial wastewater streams are discharged with high temperature. At such high temperatures, granule and biofilm formation are less prominent due to the decline in extra-cellular polymers (EPS) formation [1–3]. Thus, sludge settling becomes difficult, leading to processes instability through conventional anaerobic routes. This sludge settlement issue can be overcome by coupling anaerobic processes with membranes termed as anaerobic membrane bioreactor (AnMBR). With the aid of membranes, biomass washout can be minimized leading to better effluent quality. However, membranes used for AnMBR are generally microfiltration (MF) and ultrafiltration (UF), hence small molecular weight organic compounds and some trace organic contaminants (e.g., steroid hormones, pharmaceuticals, personal care products, pesticides, and disinfection by-products) [4] are able to pass through these membranes. As a consequence, retention time of these compounds becomes the same as hydraulic retention time (HRT) and hence microbes are not able to degrade such compounds.

Membrane distillation (MD) is a non-isothermal membrane processes, where the driving force is the temperature difference induced by vapor pressure gradient between feed and permeate side. MD simultaneously combines both heat and mass transfer through a hydrophobic porous membrane which allows only the vapor to pass through. Direct contact membrane distillation (DCMD) is a membrane configuration where the warm feed and the cold permeate are in direct contact with hydrophobic porous membrane, thus creating a vapor pressure gradient and low vapor pressure liquid (water) preferentially passing through to the cold permeate side. This configuration is best suited for applications in which the major flux is water, such as desalination or to concentrate aqueous solutions. However, heat loss is one of the major drawbacks of this configuration. Due to extensive study made available by various researchers [5,6] on the configuration and ease for process control, DCMD was selected for this research.

With the successful application of thermophilic anaerobic membrane bioreactor (TAnMBR) for industrial wastewater treatment [7], it is now possible to add MD as a thermally driven filtration unit to achieve higher effluent quality. Membrane distillation (MD) is a novel separation process that mimics distillation process while using less energy, hence making it possible to retain and treat compounds that anaerobic systems generally failed to treat effectively, e.g., effluent of

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TAnMBR was taken to assess the effectiveness of MD for further treatment. The TAnMBR with its ability to retain biomass was highly effective in removing all suspended solids (SS), nearly 95.2% COD, 81.6% TS and 73.6% TDS, but still had an effluent with an average of 951.5 mg/L COD, 2918.5 mg/L TS and 3008 mg/L TDS. Thus, even though the current application of membrane distillation is primarily focused on desalination, one can harness MD potential for only allowing compounds with low vapor pressure (volatiles) to selectively pass through its hydrophobic surface. This would result in very high rejection in COD, TS and TDS. Till date only limited number of studies have been published with respect to membrane distillation bioreactor (MDBR) both for aerobic and anaerobic treatment (present study) as compared to MBR [8] and AnMBR[1,7,9]. Thus in this research, effluent from a thermophilic anaerobic membrane bioreactor (TAnMBR) treating high strength wastewater [10] was used as feed for the direct contact membrane distillation (DCMD) process to assess the challenges, flux and fouling interaction while treating thermophilic anaerobic effluent.

2. Material and methods

2.1. Direct contact membrane distillation design configuration

The membrane used in this study was a hydrophobic polytetrafluoroethylene (PTFE) flat sheet membranes (Sumitomo Electric Industries, Ltd. POREFLON®) with a nominal pore size of 0.1 μm (Model number HP-010-30). The membrane characteristics were as follows: contact angle 112°, thickness 30 μm, liquid entry pressure (LEP) 180 kPa and an operational temperature tolerance of (−100)–(+260) °C. The membrane module was fabricated from chlorinated polyvinyl chloride (cPVC-temperature tolerance up to 90 °C) material, with an active membrane area of 0.0185 m^2 (0.170 m × 0.109 m). Two 3 L stainless steel tanks were insulated and used as feed and permeate tanks. The permeate tank was placed on an electronic balance (T-scale QHW-6-R) for continuous flux measurement. Feed and permeate streams were circulated using a peristaltic pump (Masterflex L/S) in co-current conditions to achieve DCMD configuration. Spacers were introduced at both feed and permeate sides to protect the membrane from operational variation. An immersion heater (Cole Parmer, EW-03046-54) was directly submerged into the feed tank to control the temperature of the feed solution. The permeate stream from the membrane module was circulated to double wall spiral heated exchanger to control the permeate temperature at 10 °C using a cooler (TTK Science, CTL 911). The schematic of the experimental setup is presented in Fig. 1. The overall system is automated with PID controllers.

2.2. Feed characteristics

The wastewater was obtained from the effluent of a two-stage thermophilic anaerobic membrane bioreactor (TAnMBR) and diluted as per need using deionized water. The TAnMBR used tapioca starch-based synthetic wastewater as a feed with COD:N:P ratio of 100:5:1 and operated at 55 °C. The feed water characteristics used in the current study (collected from the effluent of the TAnMBR operating at OLR 8 kg COD/m^3 d^1 with a HRT of 58.37 h, SRT ∞ and a methane yield of 17.3 L/d) were as follows: pH 7.5 ± 0.1, COD 1240 ± 360 mg/L, TS 2918 ± 236 mg/L, TDS 3008 ± 172 mg/L, TSS 0, NH_3-N 512 ± 70 mg/L and VFA 512 ± 112 mg/L [10].

2.3. Operational and analytical methods

Permeate flux (J) in the study was measured by the change in weight at the permeate side. The flux was calculated using the following equation:

$$J = \frac{(W_2 - W_1)}{A \times \mathrm{HRT}} \tag{1}$$

where W_1 (kg) is the weight of the permeate solution before the batch experiment, W_2 is the weight of the permeate solution after the batch experiment, A (m^2) is active surface area of the membrane, and HRT (h) is a batch operation time. The experiment flux obtained from Eq. (1) was used to calculate experimental MD coefficient (B_m) or resistance (R_m) from the following equation:

$$J = B_m(p_{mf} - p_{mp}) = (p_{mf} - p_{mp})/R_m \tag{2}$$

The vapor pressure difference in DCMD could be improved by the increasing feed temperature or decreasing permeate temperature. The vapor pressure (Pa) was calculated with the temperature (K) using the following equation:

$$p = \exp\left(23.1964 - \frac{3816.44}{T - 46.13}\right) \tag{3}$$

In order to assess B_m and R_m, boundary layer resistance had to be taken into account. By evaluation of membrane surface temperature at feed and permeate sides, Eqs. (4), (5) [11] and (6) [12] were used.

$$T_{mf} = \frac{k_m/\delta(T_{bp} + (h_f/h_p)T_{bf}) + h_f T_{bf} - JH_v}{\frac{k_m}{\delta} + h_f(1 + (k_m/\delta h_p))} \tag{4}$$

$$T_{mp} = \frac{k_m/\delta(T_{bf} + (h_p/h_f)T_{bp}) + h_f T_{bp} + JH_v}{(k_m/\delta) + h_p(1 + (k_m/\delta h_f))} \tag{5}$$

$$H_v = 1.7535(T) + 2024.3\ \mathrm{kJ/kg} \tag{6}$$

where T, J, k, h, H_v and δ are temperature, permeate flux, thermal conductivity, heat transfer coefficient, heat of vaporization and membrane thickness, respectively. The subscripts m, f and p stand for membrane, feed and permeate, respectively. Heat transfer coefficient highly depends on flow condition in flow channel, indicating either laminar, transition or turbulent conditions. Thus Reynolds number was calculated by using the following equation [13]:

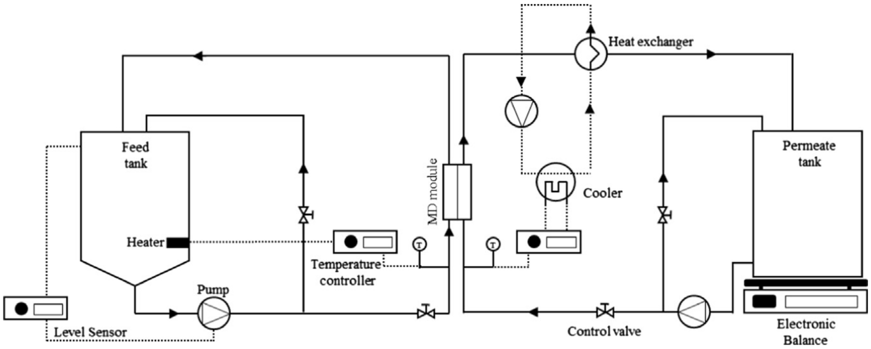

Fig. 1. Experimental setup of the DCMD module.