MEMBRANE AS SOLID/LIQUID SEPARATOR AND AIR DIFFUSER IN A BIOREACTOR

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Abstract: A bench scale pilot plant was operated to evaluate the potential of using microfiltration hollow fiber membrane modules as an air diffuser and solid/liquid separator in an alternate cycle within a bioreactor treating domestic wastewater. Two modules capable of air backwashing were immersed in a bioreactor. Compressed air backwashing and filtration by suction were effected alternatively. The experimental results reveal that application of the air backwashing technique to submerged membrane modules is capable of not only declogging the membranes but also aerating the mixed liquor. Thus better filtration flux rate and aeration without a separate aeration device were attained simultaneously. Further it was also noted that the introduction of an anoxic zone enhanced the removal of nitrogen. In addition, operation at low HRT with high sludge concentration and the absence of a sedimentation tank promise a considerable saving on plant area.

Key Words: membrane bioreactors, air diffusers, microfiltration, biological processes, gas transfer, declogging, water reuse

INTRODUCTION

Standards for effluent discharge are becoming more and more stringent in order to satisfy the constraints of the receiving bodies. Use of treated wastewater for secondary purposes in densely populated urban centres is also increasing due to the scarcity of available potable water as well as the capacity limitations of the water and wastewater conveyance systems. Thus in both cases achieving a high level of treatment is imperative.

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The use of biological treatment can be traced back to the late nineteenth century and by the 1930s it was a standard method of wastewater treatment (Rittmann, 1987). Basically the activated sludge process consists of an aeration tank for biological oxidation of organic matters and nitrogen compounds, followed by a sedimentation tank for solid/liquid separation. However, the quality of the effluent depends very much on the hydrodynamic conditions in the sedimentation tank and the settling characteristics of the sludge. Consequently, large volume sedimentation tanks offering residence times of several hours are required to effect adequate solid-liquid separation (Fane and Fell, 1978). At the same time, it is necessary to avoid conditions in the aeration stage, which lead to poor settleability or bulking of sludge. On the other hand, if the sludge concentration of 3,000 – 6,000 mg MLSS/L maintained in the aeration tank can increased to well over 10,000 mg MLSS/L, the loading rate to the plant also can be increased, consequently plant size become more than four fold smaller. Further operating at a lower sludge concentration means the lower sludge age. As a result high sludge production rate and insufficient removal of nutrients are experienced. In such situation, the maximum retention of suspended solids and colloids appears to be a necessary step. Therefore, to achieve enhanced sludge retention, a separation technique different from secondary clarification needs to be applied.

For years now, membrane technology has been used to replace secondary clarifiers to overcome the limits of the conventional activated sludge. The application of membrane coupled biological wastewater treatment was reported nearly 30 years ago by Smith *et al.* (1968). By using the membrane system, the settling characteristics of the solids are no longer of importance, hence there is greater flexibility in the operation of the aeration stage. This can lead to more efficient pollutant (organic/nutrients) removal as sludge residence time (SRT) is independent of hydraulic retention time (HRT) (Smith *et al.*, 1969; Suwa *et al.*, 1989), and importantly, development of waste-specific, rather than generic, microbial populations becomes possible. With membrane-coupled bioreactors (MBR) good treatment efficiencies have been reported even at a HRT as low as 2 hours (Chaize and Huyard, 1991). In terms of treatment efficiency of MBR, more than 90% of BOD/COD (Smith *et al.*, 1969; Trouve *et al.*, 1994; Muller *et al.*, 1995; Lübbeke *et al.*, 1995) and more than 90% of nitrogen (Cheimchaisri *et al.*, 1992) removal have been reported. In general, the treated water quality is very much suitable for secondary reuse purposes (Aya, 1994; Yokomizo, 1994).

However, this process is not without problems; concentration polarisation and membrane fouling are the major hurdles for the application of this process. Among the several methods proposed to overcome these problems, crossflow filtration was the first and the most important. During crossflow filtration, the filter cake is continuously swept away as the fluid flows along the membrane surface. The full-scale application of this configuration is now very commonly used in wastewater re-use systems in Japanese buildings (Aya, 1994). However, the method consumes a great amount of energy to maintain the crossflow velocity and filtration pressure. Moreover, it uses 10 to 20 times as much circulation feed as filtrate by volume to maintain high filtration flux, and the use of the circulation pump with this discharge rate generates excessive shear stress, which in turn damage the biocatalyst, thus the microbial activities are hindered (Shimzu *et al.*, 1992).

A turning point was achieved when Yamamoto *et al.* (1989) proposed the submersion of the membrane in an aeration tank and filtration of the treated water by suction. Energy consumption was reduced significantly, as suction pressure in a submerged MBR is generally lower than that in a crossflow MBR, and an essential part of the crossflow filtration, the recirculation pump, is absent here. The mechanism used to create the crossflow stream across the membrane surface was air diffusion, which is usually the part of the activated sludge process. The shear stress in the mixed liquor of the submerged MBR is therefore comparable to that experienced in a conventional activated sludge system, and as a result, microbial activities are not hindered by coupling membrane technology with the activated sludge process.

A submerged MBR system with 0.1 μm pore size microfiltration, was reported to achieve more than 88% organic oxidation (Chiamchaisri *et al.*, 1992; Yamamoto *et al.*, 1989) and more than 90% nitrogen removal (Chiamchaisri and Yamamoto, 1993; Chiamchaisri *et al.*, 1992). In all these studies, the improved flux was obtained by the intermittent suction operation (mostly on a five-minute cycle), which eliminated the compaction of the cake layer. Because of this, half of the time membrane modules are not in use. If the air-backwashing technique is used during this idling time, it is able to not only eliminate the cake layer formation on the surface but is also able, to some extent, to remove the internal fouling without impairing any useful filtering time of the module. Moreover, if this backwash air is sufficient to aerate the mixed liquor, the expenditure on the conventional aeration system can be saved. On the other hand, to achieve the reported nitrogen removal, intermittent aeration in a 90 min. cycle was used for nitrification and denitrification in a single compartment. Because of this intermittent aeration there was no turbulence in the aeration tank in certain periods, and as a result, therefore, there was a tendency for the sludge cake to accumulate on the membrane surface. To overcome these problems, a rotating paddle and intermittent jet-aerating device were incorporated to create a localised turbulence. However, this led again to increasing capital, operation and maintenance costs, a continuation of recent trends. The incorporation of a separate anoxic zone might provide similar nitrogen removal while eliminating some of these problems.

The objectives of the research reported here were to examine the potential for using hollow fiber microfiltration membrane modules for air diffusion and filtration in an alternate cycle within a bioreactor treating domestic wastewater. The extent of total nitrogen removal by the introduction of the anoxic zone was also investigated. For these purposes, a bench scale plant fed with domestic wastewater was operated. Carbon, nitrogen, total phosphate and coliform removal were studied to evaluate the reactor performance at different HRTs.

MATERIALS AND METHOD

Experimental Set-up

The schematic diagram of the bench scale membrane bioreactor is shown in Fig. 1. The system basically consists of an anoxic tank and an aeration tank. The working volume of the aeration tank was 80 L, and that of the anoxic tank was 40 L for the first three runs and 120 L for the last run. Three pumps for substrate feeding, recycling of mixed liquor from the aeration tank to the anoxic tank, and suction of the filtrate were attached to the unit. Two hollow fiber microfiltration membrane modules with a nominal pore size of 0.2 μm were immersed into the aeration tank. The surface area of each module was $1m^2$ based on lumen area. These membrane modules were connected to the suction pump and compressed air supplied through solenoid valves. These solenoid valves were then connected to a timer arrangement to regulate the intermittent opening and closing operation in such a way that while one module was used for filtration the other module was supplied with compressed air for backwashing. This was achieved by closing valve

V_1 & V_4 (Fig. 1), while the valves V_2 & V_3 were left open and vice versa. The compressed air supply was connected to the modules through a regulator, air filter and air flow meter. The anoxic tank was provided with a turbine mixer to keep the contents in suspension. The aeration of mixed liquor throughout the study was solely achieved by the backwash air only.

FIG. 1. Experimental Setup

Feed and Seed

The seed sludge used for inoculating MBR was collected from an activated sludge plant treating municipal wastewater. The MLSS concentration of the sludge was increased to the desired level (about 15,000 mg/L) by feeding raw septage. The domestic wastewater collected on the AIT campus was used as substrate feed. Since this wastewater had a very low strength, it was supplemented with septage to reflect the typical medium strength domestic wastewater.

Operating Conditions

The experimental investigation in this study consists of two phases. In the first phase, the effect of backwash air pressure on flux recovery was considered. In the second phase the bioreactor performance in terms of treatment efficiency was studied at different HRTs. For both phases of study, the cycle time for membrane filtration and air backwashing was fixed at 15 min. (each module underwent 15 min. filtration and 15 min. backwashing alternatively). This cycle time was identified by Maythanukhraw (1995) as optimal for a similar kind of setup. For the first phase of the study, the MLSS concentration in the aeration tank was maintained at about 17,000 mg/L and there was no substrate feed. The anoxic tank was temporarily disconnected during this phase of study. The trials were carried out at different backwash air pressures, varying from 50 kPa to 200 kPa at intervals of 50 kPa. At each backwash pressure, filtrate flux was measured frequently for 5 h and the filtrate was returned to the aeration tank continuously to maintain the MLSS concentration.

In the second phase of the study, complete set-up was operated with the substrate feed. Trials were carried out at 15, 10, 6 and 3 hours of HRT in the aeration tank, while the SRT was maintained at 50 days. The desired HRTs were maintained by adjusting the suction pressure (transmembrane pressure) and the calculation of HRT was based on the cumulative volume collected. For the MBR, since there is no biomass lost in the effluent and the biomass concentration in both the reactor and wasted sludge stream is the same, the SRT was calculated as suggested by Li *et al.* (1984).

$$\mathrm{SRT} = \frac{V}{W} \tag{1}$$

in which V= reactor volume; W = volume of the sludge wasted. Thus the SRT of the MBR was controlled hydraulically by deliberately wasting 1.6 L of sludge every day from the aeration tank. In the case of the anoxic tank, the HRT was maintained such that the recycled mixed liquor resided in the anoxic tank for 90 min. and no sludge was wasted. The DO concentration in the aeration tank was always more than 2.0 mg/L and in the anoxic tank below 0.1 mg/L. The operating temperature was in the range of 28 to 32 ^{o}C.

Measurement of Initial Membrane Resistance

Each membrane module was immersed separately in a rectangular tank with a working volume of 80 L. The tank was filled with ultrafiltered tap water. The water was withdrawn through the membrane by a suction pump and the suction pressure was measured by a vacuum gauge. Flux through the membrane was recorded at different suction pressures. During the experiment, the withdrawn water was returned to the tank to keep the water level in the tank constant.

The relationship between the flux and transmembrane pressure is given in the following equation :

$$J = \frac{\delta P}{\mu R_m} \qquad (2)$$

in which J = flux ($L/m^2.h$); δP = transmembrane pressure (kPa); μ = viscosity ($kN.s /m^2$); R_m = apparent membrane resistance. However, the apparent membrane resistance is supplemented by initial membrane resistance, and the resistance due to the solid deposition and internal fouling as in the equation below.

$$R_m = Rm_o + R_C + R_{IF} \qquad (3)$$

in which Rm_o = initial membrane resistance; R_c = membrane resistance due to the deposition of solids (cake layer); R_{IF} = membrane resistance due to internal fouling. When using ultrafiltered water, there is no solid deposition or internal fouling, so the term R_C and R_{IF} in equation (3) is redundant. The modified equation to find the initial membrane resistance when clean water is used is:

$$\delta P = \mu.Rm_o.J \qquad (4)$$

Based on the equation (4), Rm_o was determined by plotting the variation of flux with transmembrane pressure.

Membrane Cleaning

Membrane modules were cleaned after the second and third runs, and twice during the fourth run. The cleaning was carried out in two steps. First, the membrane was removed from the reactor and washed with water followed by air diffusion for 5-6 hours. Immediately after this, membrane resistance was measured to check the cleaning efficiency of this method. After this the membrane was chemically cleaned, as follows. First, the cleaning solution was prepared by mixing 720 mL of 50% sodium hydroxide solution and 280 mL of solution provided by the membrane manufacturer in 50 L dechlorinated ultrafiltered tap water at 35^{o}C. Membrane modules were submerged in the solution and the solution was filtered at 30 kPa for half an hour. After this, the modules were soaked in the solution for another two hours. The modules were then washed with clean water and the membrane resistance was measured.

Analytical Method

Reactor performance was monitored by measuring COD, BOD, TKN, NO_3-N, NO_2-N and TP in the influent and in the effluent, according to the standard procedures (APHA, 1989). Total and faecal coliform content in the influent and effluent were detected using MPN technique (multiple tube permentation technique) to determine the pathogenic organism removal in this system.

RESULTS AND DISCUSSION

This section comprises three parts. In the first part, the performance of air backwashing at different pressures is presented. Next, the membrane performance in the bioreactor in the long term experiment is reported and discussed. Finally, treatment efficiencies are addressed.

Effect of Backwash Air Pressure

A series of trials were conducted to determine the effect of air pressure on backwashing efficiencies of membrane modules during filtration. For this purpose, filtration flux through membrane modules was observed at 15:15 operation mode (15 minutes filtration and 15 minutes air backwashing), with different backwash air pressure for five hours. Variation of flux with time at each backwash air pressure investigated is shown in Fig. 2. In addition, the decline in flux with time during continuous filtration without air backwashing also presented in Fig. 2 to show the effect of air backwashing in comparison to continuous filtration.

FIG. 2 Variation of Flux at Different Backwash Air Pressure

From these experimental results, it can be observed that, at the beginning, all backwash air pressures, except 150 kPa give similar recovery. For 150 kPa backwash air pressure, Module II gives a slightly higher initial flux. However, in the later stage, the flux of this module also shows a similar trend to other cases. In addition, in a long run, flux recovery is better with higher backwash air pressure. This is evidenced by the flux variation in the fifth hour of the experiment for 150 and 200 kPa backwash air pressures. During this period, flux with 200 kPa backwash air pressure tends to improve further while the flux with 150 kPa backwash air pressure seems to stabilize at a lower level. A similar trend could be seen at 50 and 100 kPa backwash air pressures. In addition, the slow flux decline rate for cyclic operations with air backwashing can clearly be seen in comparison with the continuous suction.

From these observations, it can be concluded that cyclic operation with air backwashing improves the flux recovery compared to continuous filtration. Further, an increase in backwash air pressure improves flux recovery in a longer run. To study the long term effects of flux recovery, trials were carried out with air backwashing (150 kPa) in a cyclic operation and with continuous filtration without air backwashing. With air backwashing, a 90% flux improvement compared to continuous operation can be achieved after 26 hours of operation; this percentage improvement continues to increase with longer periods of operation.

Gas Transfer Efficiency

In order to investigate the dual purpose of the backwash air namely, membrane pore declogging and air diffusion, experimental runs were conducted to measure K_La values. Figure 3, presents the K_La values of the conventional stone air diffusers and membrane modules, at different air flow rates, in pure water. Here, it was noted that in both case the K_La value increases with air flow rate, but the gas transfer rate efficiency of the membrane modules are superior to the stone air diffusers for all air flow rates. The larger effective gassed area and formation of fine air bubbles could be the cause for better performance of the membrane modules in term of gas transfer.

FIG 3. Comparison of Gas Transfer Coefficient (K_La) in Pure Water at 20^o C for Stone and Membrane Diffusers

Operating Conditions of the Membrane During Long Term Run

Transmembrane pressure

Fig. 4 shows the variation of transmembrane pressure with time during different HRT. For the first three runs, the transmembrane pressure increased at the beginning and stabilised at constant levels at the end of each run. In the case of the fourth run, the transmembrane pressure increased to 96 kPa within a day or two, and remained constant at that level for the rest of the operation. Chemical cleaning of the membrane twice and an increase in backwash air pressure by 50 kPa after each chemical cleaning during the fourth run gave a similar pattern of transmembrane pressure increase. The rapid increase in the transmembrane pressure during the run could be attributed to the colloidal fouling of membrane pores due to the deposition of macromolecules and colloids. It is well known that the colloidal fouling is a function of filtrate volume i.e. more the volume filtered the tendency for the colloidal fouling is high. This is what happened during the fourth run. In this run the HRT was reduced to half compared to the previous run, thus the filtered volume doubled. Therefore more colloids and macromolecules carried to the membrane pores, caused the membrane fouling which could not be removed by air backwashing and consequent increase in transmembrane pressure. The membrane fouling further discussed in latter section considering the membrane resistance.

FIG. 4. Transmembrane Pressure Variation

From Fig. 4 it can be further noted that during long HRT (Run I) the time taken to reach the stabilised transmembrane pressure is long compared to the shorter HRT runs. This can be explained by the fact that at long HRT, flux was low, so the chances of fouling materials being carried towards membrane modules are lower.

Permeate flux

The average permeate flux obtained throughout the study is presented in Fig. 5. For the first three runs, flux was obtained to maintain the desired HRT. However, at the beginning of the fourth run, flux was obtained to maintain the desired HRT, but subsequently continued to decrease; cumulative flux after a week's operation led to the situation of 6 h HRT. With the aim of recovering flux, membrane modules were chemically cleaned twice, but in each occasion the HRT reached 6 h after seven days of operation. So the run was continued and the flux started to get more or less a steady state when the HRT has reached to 8 h. This so happened after 14 days of last chemical cleaning of membranes.

FIG. 5. Permeate Flux of Membrane Modules

It should be noted that the flux could be maintained to get a HRT of 6 h at a moderate transmembrane pressure of 42 kPa but it was not possible during fourth run with the highest possible transmembrane pressure (96 kPa). From this, it can be concluded that the increase in transmembrane pressure has led to cake layer compaction and internal fouling, most of which could not be removed by air backwashing. This seems to indicate that the specific resistance caused by cake layer and internal fouling is a strong function of the applied suction pressure and filtrate flux, and the increase in resistance to filtration more than offsets the increased driving force. A similar phenomenon of lower flux at higher transmembrane pressure (80 kPa) was observed by Benítez et al (1995). Based on the above discussions on flux and transmembrane pressure at different HRTs, it can be concluded that the optimum HRT lies between 3 and 6 h.

Membrane resistance

The influence of internal fouling on membrane resistance can be explained by the measured values after each cleaning. Fig. 6 shows the resistance of membrane before being used in any experiments (*Initial*), cleaned with tap water and air diffusion after the second run (*First (1a)*) subsequent chemical cleaning (*First (1b)*), after the third run (*Second*), after 11 days of the fourth run (*Third*), and after 21 days of the fourth run

(*Fourth*). Once again the membrane resistance model described in equation (3) can be considered here. Considering the membrane cleaning, it could be said that the tap water cleaning with air diffusion created sufficient turbulence to remove any cake layer formation on the membrane surface. Therefore the term R_C in equation (3) becomes redundant at this stage. After this cleaning, the membrane resistance showed an increase of 178% of the initial membrane resistance. After subsequent chemical cleaning, by contrast, the resistance merely increased by about 10% over the initial membrane resistance. These observations indicate that there was a certain amount of colloidal fouling, which could not be removed by air backwashing, and that this could be the cause for these reported filtrate flux reductions.

FIG. 6. Membrane Resistance (Initial and after Cleanings)

Bioreactor Performance

Color and turbidity

Variation of color and turbidity in effluent is shown in Fig. 7. In all experiments the effluent turbidity varied between 0.15 and 0.3, and the color between 20 and 30 Hazen color unit. This measured turbidity is below even the standard set for drinking water (0.5-1 NTU) by USEPA (Sawyer *et al*, 1994). Thus the membrane with pore size of 0.2 μm effectively prevents escape of most colloidal matters with effluent. However, membrane may not be particularly efficient in preventing the escape of micromolecules probably originated from septage and this could be the cause of the slight color observed in the effluent.

FIG. 7. Color and Turbidity of Effluent

Removal of organic matter

Variation of influent and effluent COD concentration and the percentage removal efficiency are shown in Fig. 8. The influent COD varied between 540 to 625 mg/L. The effluent COD concentration always maintained a level below 25 mg/L, with a lowest value of 7 mg/L throughout the study. This corresponds to more than 95% organic matter being removed. Steady organic matter removal efficiency through out the study indicates that the change in operating conditions does not have any effect on the organic matter removal.

FIG. 8. Feed and Permeate COD and Removal Efficiency

Further analysis reveals that influent BOD varies from 295 to 380 mg/L and effluent varies in a small range of 1.3 to 3.5 mg/L. In addition, the percentage biodegradable matter (BOD/COD ratio) in the influent (52-64%) and very low percentage of biodegradable matter in the effluent (7-17%) shows that almost all the biodegradable matter is already removed in the bioreactor. The long SRT allows the biodegradable substances with high and low molecular weight to be taken up, broken down and gassified by micro-organisms or converted into polymers as constituents of bacterial cells, thereby raising the quality of treated effluent.

Total nitrogen (TN)

Fig. 9 shows the TN in influent and effluent, and its removal efficiency. From these results it can be seen that the removal efficiency is more than 80% with a few exceptions at the beginning of Run I and IV. Moreover, removal efficiency improved during each run. This lower removal efficiency at the beginning may be due to the change in operating conditions. However, it can be concluded that the removal of TN would be more than 85% in a long run with stable operating condition (i.e. constant HRT). Further, it should be noted that, although during the change of operating condition there was a reduced removal efficiency, in a long run the HRT at which the process takes place has no effect on total nitrogen removal. Except for the third run, effluent TN concentration meets the EC (European Community) effluent standard (10 mg/L or an 80% reduction) for NO_3^--N (Morris and Bird, 1994). The non-compliance during the third run is mainly because of the high concentration of TN in the feed (due to the low C:N ratio in the septage added to adjust the feed wastewater strength), which is normally not the case for domestic wastewater.

FIG. 9. Feed and Effluent Total Nitrogen and Removal Efficiency

Nitrogen mass balance

A schematic diagram of the nitrogen balance in a unit of the biological treatment plant is given in Fig. 10. There are two pathways for the removal of nitrogen: assimilation into the biomass, and nitrification and denitrification. The amount of nitrogen lost in assimilation depends on the amount of biodegradable mater removed. Based on these facts nitrogen mass balance can be written as follows:

$$TN_i = TN_e + \text{N assimilated} + \text{N lost due to denitrification} \qquad (5)$$

The equation (5) can be re-written to determine nitrogen lost due to denitrification.

$$\text{N lost in denitrification} = (TN_i - TN_e) - \alpha(BOD_{5i} - BOD_{5e}) \qquad (6)$$

in which α = conversion factor (ratio of nitrogen to BOD_5); TN_i = total nitrogen influent; TN_e = total nitrogen in effluent; BOD_{5i} = influent biological oxygen demand at five days; BOD_{5e} = influent biological oxygen demand at five days. The term $\alpha(BOD_{5i}$ - $BOD_{5e})$ represents the removal of nitrogen by assimilation. It is generally accepted that during the aerobic process of organic matter removal, each 100 mg/L of BOD_5 need 5 mg/L of nitrogen and 1 mg/L of phosphorous (Klopping *et al.*, 1995). Based on this, BOD : N = 100 : 5, α can be determined to be 0.05. But this ratio cannot be used directly to calculate the nitrogen assimilated into the biomass in a MBR process due to its long SRT operation. With a longer SRT operation the biomass yield becomes lower thus nitrogen assimilated per kg of BOD_5 removed become lesser. However, it is still conservative to use this ratio to show there is denitrification took place.

The mass balance of nitrogen for the MBR system presented in Fig. 11 shows the total nitrogen into the system (in the form of TKN, NO_3-N and NO_2-N) and the loss of nitrogen from the system (in the form of TKN, NO_3-N, NO_2-N, assimilation and denitrification). The influent nitrogen mainly in the form of TKN and it should be noted that the amount of the nitrogen in the form of NO_2-N is significant neither in influent nor in effluent. Further, the mass balance indicates that denitrification has taken place though not completely (note the presence on NO_3-N in the effluent) but to a very significant level, and that it is more than likely the main cause of the more than 80% TN reduction in the effluent. Prevalence of favourable pH and temperature during all runs could have enhanced this high level of denitrification. The denitrification is further evidenced by no drastic pH drop in the system, despite nitrification having taken place. The denitrification process should have compensated the pH drop by producing alkalinity. In addition, the presence of air bubbles were observed on the anoxic tank surface indicates the escape of the nitrous gases.

FIG. 10. Nitrogen Pathway in Biological Wastewater Treatment

FIG. 11. Nitrogen Mass Balance

Phosphate removal

Total phosphate removal varied from 52 - 92%. A similar mass balance calculation indicates that the major mechanism for total phosphate removal was assimilation. The low amount of sludge wasted and the low F/M ratio maintained in the system could be used to explain this.

Pathogenic micro-organism removal

To determine the disinfecting ability in this process, total and faecal coliform content of influent and effluent during each run were checked. The analysis indicates that while there were more than six log number coliforms in the influent, no colonies were detected in the effluent. Such a result cannot be achieved in the conventional activated sludge process. In addition, the removal efficiency is independent of operating conditions such as HRT, transmembrane pressure etc. Since no chemicals are involved in the disinfecting, this process does not have the disadvantage of residual disinfectants, toxic or carcinogenic by-product formation.

Reusability of the treated water

Table 1 shows the summary of influent and effluent water quality on Membrane Bioreactor as well as the guidelines for the reuse of treated wastewater for different purposes. From this comparison it can be noted that the MBR effluent meets the guidelines in every aspects. Therefore the application of MBR system has great potential where reuse of treated wastewater is in practice.

TABLE 1. Comparison of Treated Water Quality from MBR with Reuse Guidelines

SUMMARY OF FINDINGS AND CONCLUSIONS

The study shows that filtration through membrane in a cyclic operation with air backwashing plays an important role in the improvement of permeate flux stability by removing external deposits on the membrane surface, preventing the compaction of cake layer and reducing the internal pore clogging of the membranes. After 26 hours of operation in cyclic mode of 15:15 (15 minutes filtration and 15 minutes air backwashing) with 150 kPa backwash air pressure shows a 90% improvement in flux compared to continuous suction. Study with various backwash air pressure reveals that an increase in backwash pressure will lead to a greater improvement of flux in a long run. In addition, the use of the air backwashing technique can eliminate the need for conventional air diffusers because backwash air itself sufficient enough to aerate the mixed liquor in the reactor. Backwash air at 250 kPa is alone sufficient to aerate the mixed liquor (MLSS = 13,000 mg/L) to a DO level of 3.5 mg/L when system is loaded with 0.19 kg BOD/kg MLVSS.d. Further the measurement of gas transfer efficiencies indicate that the membrane modules are better air diffusers than the stone air diffusers.

Therefore membrane filtration with air backwashing could return cost savings by many means: (1) improved flux rate obtained by the air backwashing application; (2) elimination of the capital cost involved for the conventional air diffusers; and (3) with the increased volumetric loading rate plant size become smaller thus the lower investment on construction cost plus less cost for the space required. Further scale-up data should be similar since the modules used in these experiments are essentially of commercial scale modules. In fact, it is anticipated that the air transfer efficiency could be further enhanced by more liquid depth could be used in scale-up instead of 50 cm used in the bench scale experiments. However longer term experiments with in-plant studies needed to develop suitable data to optimise operation and make an economic analysis.

During this study, the flux could be maintained to obtain a HRT of 6 h at a moderate transmembrane pressure of 42 kPa. However, this was not possible after seven days of operation during the last run with the transmembrane pressure of 96 kPa, where the original intention was to maintain 3 h HRT. It can be concluded that the increase in transmembrane pressure led to cake layer compaction and internal fouling, most of which could not be removed by air backwashing. This indicates that the specific resistance caused by cake layer and pore plugging is a strong function of the applied suction pressure and filtrate flux, and the increase in resistance to filtration more than offsets the increased driving force.

Considering the reactor performance, COD removal in all experimental runs was observed to be more than 95% with the maximum effluent COD value of 27 mg/L. Similarly, BOD removal was also more than 98%, with a maximum effluent BOD of 4 mg/L. The effluent turbidity was extremely good, with a maximum value of 0.3 NTU. This is below the drinking water standard set by USEPA. Effluent quality in terms of SS was also very good, since no solids were lost in the effluent. With more than 2 mg/L of DO concentration in the system throughout the study, the TKN removal was more than 95%. In addition, total nitrogen removal efficiencies of more than 80% were observed. The mass balance for nitrogen indicates the occurrence of denitrification.

From the process efficiency point of view, the membrane bioreactor produced better quality effluent than a conventional activated sludge process, so the treated water has a great potentials. This study establishes that using hollow fiber membrane capable of air backwashing for solid/liquid separation will lead to aeration of mixed liquor and declogging of membrane modules simultaneously, so conventional aerators can be eliminated. By having an efficient anoxic zone, total nitrogen removal can also be achieved. With this revolution, the full scale application of MBR system will become an attractive option. However, further studies on fouling characteristics of membrane with floc characterisation will pave a way to selecting the most suitable membrane pore size, and thus optimising the process further.

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APPENDIX II. NOTATION

The following symbols are used in this paper:

BOD_{5e} = five day biochemical oxygen demand of effluent;

BOD_{5i} = five day biochemical oxygen demand of influent;

J = flux;

K_La = Gas Transfer Coefficient

R_C = resistance caused by cake layer;

R_{IF} = resistance caused by internal fouling;

R_m = apparent membrane resistance;

R_{mo} = initial membrane resistance;

TN_e = total nitrogen in effluent;

TN_i = total nitrogen in influent;

V = volume of reactor;

W = volume of the sludge wasted;

α = conversion factor (ratio of nitrogen to BOD_5);

μ = viscosity;

δP = transmembrane pressure;

TABLE 1. Comparison of Treated Water Quality from MBR with Reuse Guidelines

Parameters (1)	Concentration		Criteria/Guidelines *		
	Influent (2)	Effluent (3)	Toilet Flush Water (4)	Landscape Irrigation (5)	Environmental Water (6)
Total coliform/(Count/mL)	$> 10^7$	ND	≤ 10	ND	ND
Fecal coliform/(Count/mL)	$> 10^5$	ND	-	-	-
Chlorine residual combined/(mg/L)	-	ND	TA	≤ 0.4	-
Appearance	NP	NU	NU	NU	NU
Turbidity /(NTU)	> 1000	< 0.3	-	-	≤ 10
Biological Oxygen Demand (BOD)/(mg/L)	295-375	< 4	-	-	≤ 10
Odor	NP	NU	NU	NU	NU
PH	7.6-8.5	7.3-8.4	5.8-8.6	5.8-8.6	5.8-8.6
Chemical Oxygen Demand (COD)/(mg/L)	530-625	< 25	-	-	-
Total Kjedahl Nitrogen (TKN)/(mg/L)	26-165	< 3	-	-	-
Total Nitrogen (TN)/(mg/L)	26-165	< 6	-	-	-
Total Phosphate (TP)/(mg/L)	2.2-9.0	0.2-4	-	-	-
Color/(Hazen color unit)	> 5000	<30	-	-	-

NP - Not pleasant NU - Not Unpleasant ND - Not Detected TA - Trace Amount

* adopted from Japan Sewage Work Association, 1993.

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