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Influence of feed concentration and transmembrane pressure on membrane fouling and effect of hydraulic flushing on the performance of ultrafiltration

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HIGHLIGHTS

- A diaphragm pump was dual-used in this experimental setup.
- Membrane fouling varies with feed SDS concentration and transmembrane pressure.
- Suitable flushing time is important for periodic hydraulic flushing.
- Compared with backwashing and forward flushing, the combined flushing is more effective to alleviate membrane fouling.

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ABSTRACT

Micellar-enhanced ultrafiltration (MEUF) is a promising technology developed for treating the wastewater containing metal ions or organic pollutants. One of the greatest problems in MEUF is membrane fouling which is mainly caused by concentration polarization, gel layer or cake formation caused by the deposition of surfactant micelles on the membrane surface and surfactant adsorption in the membrane interior. In this study, surfactant sodium dodecyl sulfate (SDS), which was used in membrane separation as colloidal particles, caused the flux decline. The transmembrane pressure (TMP) and feed concentration of SDS had significant influences on the flux. This paper presented that the lower TMP had a smaller effect on membrane fouling, and when SDS concentration was around the critical micelle concentration (CMC), lower permeate flux and higher additional membrane fouling resistance were obtained. The effects of three kinds of hydraulic flushing methods on membrane permeate flux were investigated, including periodic forward flushing, periodic backwashing and forward flushing followed by backwashing. It was found that when the periodic combined flushing interval was 10 min, forward flushing and backwashing phase times were 150 s and 90 s, respectively, and that combined flushing was more conductive to permeate flux recovery in this study.

1. Introduction

Micellar-enhanced ultrafiltration (MEUF) is a new technology developed for treating the wastewater containing metal ions and organic matters. In MEUF, surfactants are added to wastewater at levels equal to or higher than their critical micelle concentrations (CMCs) and surfactant monomers will aggregate to form micelles, then the micelles are able to solubilize organic solutes or bind ions on the surface of the opposite charged micelle via electrostatic interactions [1,2] which aims to promote the removal of metal ions or organic matters. MEUF has its own advantages such as simple operation, high removal efficiency, economical and practical, and small space requirement [3]. During the ultrafiltration, however, the flux decreases dramatically due to the existence of membrane fouling which results in increased operating costs, decreased membrane lifetime and reduced removal efficiency. There are many factors that lead to membrane fouling, such as the adsorption of solids onto the membrane, gel formation, pore blocking, and concentration polarization [4–6]. Particulate matter which is larger than the pores in commercial MF and UF membranes, forms a cake at the membrane surface; dissolved matter which can penetrate pores forms a surface cake, penetrates and clogs pores and reduces the pore diameter due to adsorption within pores [7]. The formation mechanism

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Nomenclature

List of symbols

J_w	the pure distilled water flux of clean membrane ($L/m^2 \cdot h$)
J_f	the pure water flux of membrane after flushing ($L/m^2 \cdot h$)
ΔP	the transmembrane operating pressure (Pa)
μ_m	the viscosity of distilled water (10^{-3} Pa·s)
R_m	the hydraulic resistance of the membrane (m^{-1})
$R_{f,}$	the additional fouling resistance (m^{-1})
J	the synthetic wastewater permeate flux ($L/m^2 \cdot h$)
A	the area of membrane (m^2)
V	the permeate volume (L)
t	the filtration time (h)
μ	the dynamic viscosity of the solution (Pa s)

of the gel layer during MEUF can be summarized as follows: adsorption of the surfactant monomer and aggregates on the membrane surface and within the pores; blockage of the membrane pores and partial constriction of membrane pores by surfactant adsorption [8,9]. JÖnsson and JÖnsson [10] investigated the impact of hydrophilic and hydrophobic membrane materials on the membrane flux reduction and fouling. They found that flux decline caused by low-molecular weight hydrophobic solutes is generally because of adsorption. Broeckmann et al. [11] reported that bulk particles can be separated into two fractions by including small enough to enter the membrane pores and the deposit on the membrane surface. One remained on the outside of the membrane. The other entered the membrane pores and contributes to pore blocking. Membrane fouling can be categorized into reversible fouling and irreversible fouling. Reversible fouling which is due to accumulation of the particles and build-up of a cake on the membrane can be mitigated by physical cleaning such as backwash and rinsing. Irreversible fouling is caused by internal accumulation of particles during the penetration of small particles through the membrane [12]. Actually, reversible fouling can transform into irreversible fouling if the formation of fouling layer with the solute during continuous filtration process was not removed timely. Irreversible fouling is normally caused by the strong attachment of particles, which leads to progressive deterioration of membrane performance and cannot be removed by physical cleaning methods [13]. Various methods have been proposed to alleviate membrane fouling [14–18] such as pretreatment of feed water (coagulation), physical methods (forward flushing, backwashing, ultrasonic), chemical methods (chemical reagents: NaOH, HCl, EDTA and NaClO) and modification of the membrane surface. However, these methods cannot completely eliminate fouling. Therefore, to maintain an efficient process, we generally preferred physical methods to control membrane fouling, because it does not cause pollution again and the operation is simple. Niina Laitinen et al. [19] analyzed in their study the effects of different backflushing parameters (backflushing frequency, length of the backflush, and the backflushing pressure) on flux and retentions. They concluded that the highest permeate flux was obtained when the backflush of 1 s was made every 2 min with a 4 bar backflushing pressure.

Sodium dodecyl sulfate (SDS), an anionic surfactant, is often chosen for the effective removal of zinc, copper, nickel, cadmium and methylene blue [20–23]. Meanwhile, the influence of SDS on colloidal interactions in the cake layer on the ultrafiltration membrane surface was investigated. The relationship between fouling potential and SDS depends on the concentration of SDS [24]. Some researchers have pointed out that the low SDS concentration did not benefit the rejection of SDS because of the small number of micelles, and with the increasing feed SDS concentration, the SDS rejection increased dramatically, for the reason that the SDS concentration at the vicinity of the membrane surface reached the CMC of SDS, leading to the formation of more micelles. However, when the feed SDS concentration was 10 times the CMC of SDS (80 mM), the SDS rejection decreased, this was attributed to the change of micelle shape from spherical to cylindrical or plate like and then the micelles could easily pass through the membrane pores causing considerable drop in the rejection of SDS [24–26]. In this study, the anionic surfactant sodium dodecyl sulfate (SDS) was added into the aqueous solution to form micelles when the SDS concentration reaches the critical micelle concentration and can be retained by an ultrafiltration membrane with pore sizes smaller than the micelle size. During MEUF, micelles bind ions on the surface of the opposite-charged micelles via electrostatic interaction. Therefore, metal ions associated with micelles are removed effectively. In fact, UF membrane cannot reject free ions at any transmembrane pressure [3]. So our study only considers SDS as membrane pollutant.

In previous studies, we reported the effects of different conditions (pH, TMP, temperature, feed concentration) on removal efficiency, fouling mechanisms and flux decline [27–29].

The purpose of this study is an attempt to investigate the effects of feed concentration and TMP on membrane fouling which was mainly reflected by permeate flux, and the influence of rinsing time on permeate flux, including the effects of forward flushing and backwashing on permeate flux.

2. Experimental

2.1. Chemicals

SDS ($C_{12}H_{25}NaSO_4$) was obtained from Tianjin Kermel Chemical Reagents Development Center, China. Phosphoric acid (H_3PO_4), potassium hydrogen phthalate ($KHC_8H_4O_4$), Sodium bicarbonate ($NaHCO_3$) and Hydrochloric acid (HCl) used in these experiments were obtained from Guoyao Chemical reagent Plant, China and analytically pure. In all the experiments distilled water was used as solvents.

2.2. Membrane

The spiral-wound membrane module used in this study was supplied by Dalian Yidong Membrane Engineering Equipment Co., Ltd., Dalian, China. The membrane material was polyethersulfone (PES) which was hydrophobic in nature. The permeate flux of the membrane was measured under standard test conditions and was found to be 44 $L/m^2 \cdot h$ at 0.05 MPa. The properties of the membrane are given in Table 1.

2.3. Performance of the experiments

All the experiments were conducted at temperature 25 ± 5 °C. According to the experiment design, the synthetic wastewater was made by adding a pre-determined amount of SDS into distilled water. Before the experiments, the solutions were stirred fully using a magnetic stirrer for about 10 min to provide highly efficient mixing and settled for 45 min to ensure the formation of micelles of constant size. The schematic diagram of the experimental setup is shown in Fig. 1.

In MEUF experiments, 40 L of feed solution was filtered through the membrane by a diaphragm pump under different pressures (0.05 MPa, 0.10 MPa, 0.15 MPa, 0.20 MPa) and the retentate was recirculated to

Table 1
Characteristics of the used spiral-wound ultrafiltration membrane module.

Type	JU1812-41
Membrane material	Polyethersulfone
MWCO (Da)	10 k
Contour size φ × L (m)	0.046 × 0.305
Effective membrane area (m^2)	0.4
Operating pressure (MPa)	<0.3
Operating temperature (°C)	5–50
pH	1–13

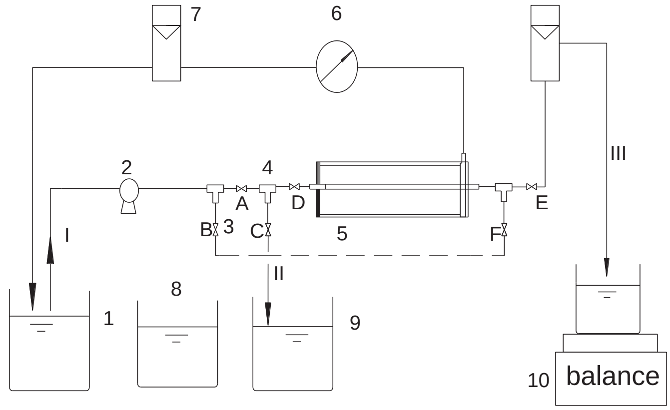

Fig. 1. Schematic diagram of the experimental setup: (1) feed tank, (2) diaphragm pump, (3) valve, (4) tee joint, (5) membrane, (6) manometer, (7) rotameter, (8) reservoir, (9) waste tank and (10) balance.

the feed tank. The TMP was monitored with a manometer. In the ultrafiltration experiment process, valves B, C, and F were closed and the retentate was recycled to the feed tank. In the experiment of forward flushing, the valves B, C, E, and F were closed and the retentate was not recirculated to the feed tank. During backwashing experiments, valves A and E were closed. During all the flushing experiments, the tube I was placed into the reservoir (tap water) and the diaphragm pump was used to pump the tap water from the reservoir to the flush membrane, then the diaphragm pump dual-used was used to deliver the feed solution into membrane when tube I was put into feed tank. In these experiments, the feed solution and permeate were sampled at frequent intervals, based on the permeate weight. Rotameters just only reflected the instantaneous flux, so we recorded the static permeate volume times to ensure the permeate flux accuracy.

After each run, the membrane was thoroughly washed to recover its permeability. First of all, the membrane was washed with ultrapure water without pressure to remove the surface deposits, and then was washed with distilled water for 30 min under 0.05 MPa, finally with distilled water for 10 min without pressure. After then, distilled water was filtered to determine the permeate flux in order to check the permeability of the membrane.

2.4. Methods

The synthetic wastewater only contained one kind of organic matter and the concentration of surfactant SDS was measured by total organic carbon (TOC), using a Shimadzu TOC-VCPH analyzer (Japan) [30]. Before measurement, the samples were filtered through a 0.2 μm filter membrane. The CMC of SDS was measured with a conductivity meter (mode DDS-11A, made in Shanghai, China) and it was found to be 8.0 mM [3].

2.5. Calculations

When distilled water is filtered, the distilled water flux (J_w) follows the equation:

$$J_w = \frac{\Delta P}{\mu_w R_m} \quad , \quad R_m = \frac{\Delta P}{J_w \mu_w} \tag{1}$$

where J_w is distilled water flux (L/m^2·h), ΔP is the transmembrane operating pressure (Pa); μ_m is the viscosity of distilled water (10^{-3} Pa·s at 25 °C), and R_m is the hydraulic resistance of the membrane (m^{-1}).

In the resistance-in-series model [31], the permeate flux in an ultrafiltration process follows the equation:

$$J = \frac{1}{A}\frac{dV}{dt} = \frac{\Delta P}{\mu(R_m + R_f)} \tag{2}$$

where J is the synthetic wastewater permeate flux (L/m^2·h), A is the area of the membrane (m^2),V is the permeate volume (L), t is the filtration time (h), μ is the dynamic viscosity of the solution (Pa s), and R_f is additional fouling resistance (m^{-1}).

Removal efficiency (R) of SDS was defined as:

$$R = \left(1 - \frac{C_p}{C_f}\right) \times 100\% \tag{3}$$

where C_p and C_f (mg/L) represent the concentration of SDS in the permeate solution and the feed solution, respectively.

The flux recovery (FR) was defined as:

$$FR(\%) = \left(\frac{J_f}{J_w}\right) \times 100 \tag{4}$$

where J_f and J_w (L/m^2·h) denote the pure water flux of the membrane after flushing and the pure water flux of the clean membrane, respectively.

3. Results and discussion

3.1. Membrane intrinsic resistance and pure water flux

To measure membrane intrinsic resistance, the pure water flux under different pressures before each fouling experiment was tested. Fig. 2 shows the pure water flux before each fouling experiment. These measurements provided a benchmark for gaging the changes in permeate flux in terms of varying TMP when the clean membrane was fouled. Apparently the pure water flux increased linearly with TMP. The measured membrane intrinsic resistance was 4.12×10^{12} m^{-1}, which indicated that the cut-off capacities of the membranes before each group were in substantial agreement and the influence of the membrane itself on fouling could be neglected.

3.2. Effects of operating pressure and initial feed SDS concentration on permeate flux and additional membrane fouling resistance

3.2.1. Effects of operating pressure on permeate flux and additional membrane fouling resistance

The initial feed SDS concentration was fixed at 10 mM, at room temperature (25 ± 5 °C) and the transmembrane pressure is ranged from

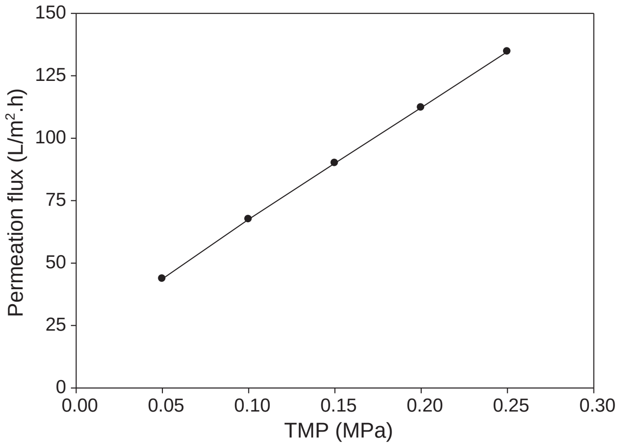

Fig. 2. Relationship between TMP and pure water flux.

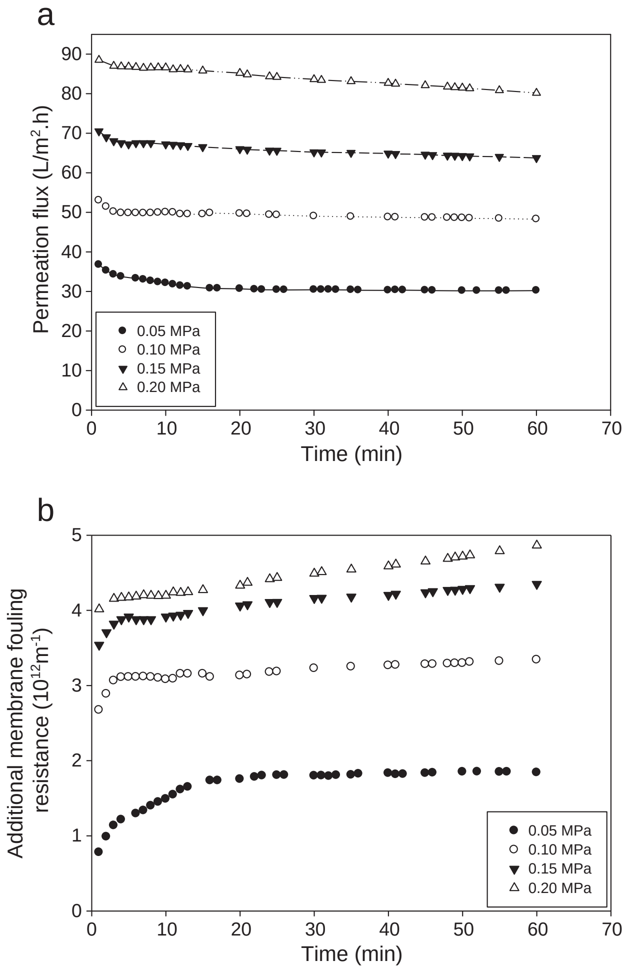

Fig. 3. Effects of the TMP on permeate flux (a) and additional membrane fouling resistance (b). Initial feed concentration of SDS, 10 mM; temperature, 25 °C.

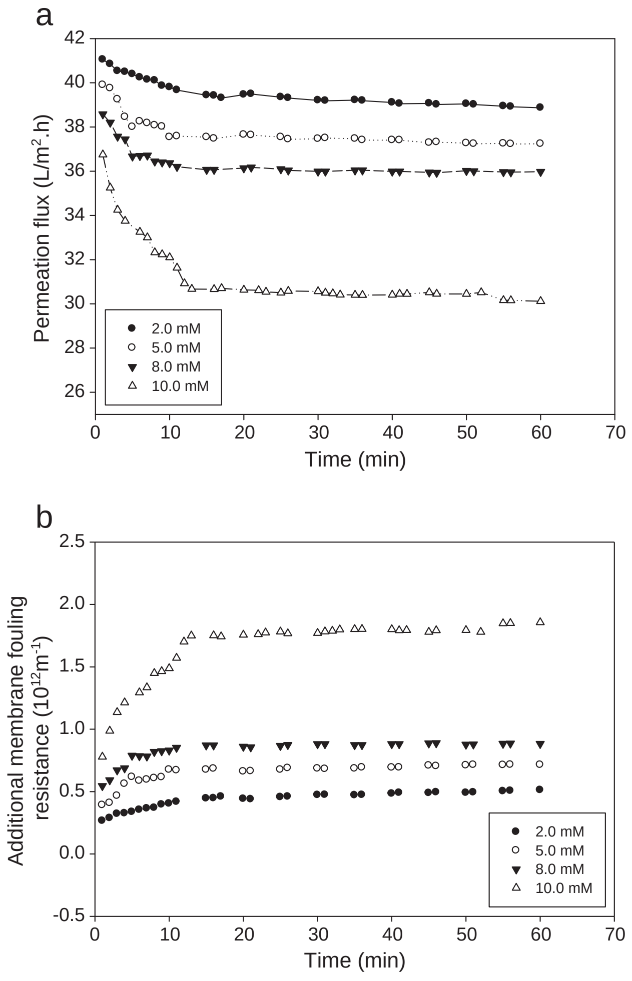

Fig. 4. Effects of initial feed SDS concentration on permeate flux (a) and additional membrane fouling resistance (b). Operating pressure, 0.05 MPa; temperature, 25 °C.

0.05 to 0.20 MPa. We found that the permeate flux increased with the increasing of TMP and flux decline occurred during the initial stages of filtration for all the pressures and then came to a steady state value (Fig. 3(a)), ranging from 30.20 $L/m^2 \cdot h$ at 0.05 MPa to 80.15 $L/m^2 \cdot h$ at 0.20 MPa after an hour of ultrafiltration, which was consistent with those studies [25,2]. The reasons are that the TMP between retentate and permeate played the role of an effective driving force and the increased pressure could overcome the resistance, hence compelling more solution to filter through the membrane and resulting in a higher permeate flux [28]. Additional membrane fouling resistance curves for different TMPs are shown in Fig. 3(b). It can be seen that resistance increased with the increase of TMP. With time increasing, the increased pressure resulted in a thicker gel layer. According to the above results, 0.05 MPa was selected for the membrane fouling experiment.

3.2.2. *Effects of initial feed SDS concentration on permeate flux and additional membrane fouling resistance*

During membrane filtration, the flux can be used as an indicator of membrane fouling. It is easy to estimate the degree of the membrane fouling from the changes of permeate flux. The pure water flux was measured with distilled water before and after experiments to evaluate the membrane fouling. Under constant pressure operation, flux declines as the result of membrane fouling.

The initial feed concentration of SDS was varied between 2 mM and 10 mM, at 0.05 MPa, at room temperature. As shown in Fig. 4(a), it was clear that after 1 h membrane filtration, the membrane permeate flux sequential order was 2 mM > 5 mM > 8 mM > 10 mM. Obviously, with the increment of initial feed SDS concentration, membrane permeate flux kept decreasing (Fig. 4(a)). It can be explained as follows: at 2 mM, which means the concentration below CMC, the feed solution existed as the monomer and the surfactant monomer was smaller than that of the membrane pore, consequently, the permeate flux was higher than other concentrations; at 5 mM, as the experiment proceeded, the retentate was recycled to the feed tank, contributed to the feed concentration of SDS increasing, then the concentration nearby the membrane surface increased gradually until reaching the critical micelle concentration, as a result, parts of SDS monomers began to form micelles, at the same time, there were still some monomers can get through membrane pores; at CMC (8 mM), nearby the membrane surface the SDS began to form micelles which was larger than membrane pore size, then the great mass of micelles can be rejected by membrane and these micelles cannot block the pores and then get much higher permeate flux; when the feed concentration was around CMC (10 mM), the permeate flux was lower than other concentrations. This could be explained by the fact that the feed solution which was just more than the critical micelle concentration (CMC), could form a large number of micelles, not only more and more SDS micelles deposited on the membrane surface then assembled into cake formation, but also the existence of surfactant monomer (these monomers appear to be formed smaller aggregations, called pre-micelles) and some of monomers could be rejected by the gel layer whereas the rest of them would likely to get through the pores or be adsorbed by pores. Therefore the permeate flux was lower than the others [32]. When the initial SDS feed concentration at 10 mM, in the first 10 min, the permeate flux decreased quickly. This behavior may be attributed to the concentration polarization [26], namely, SDS

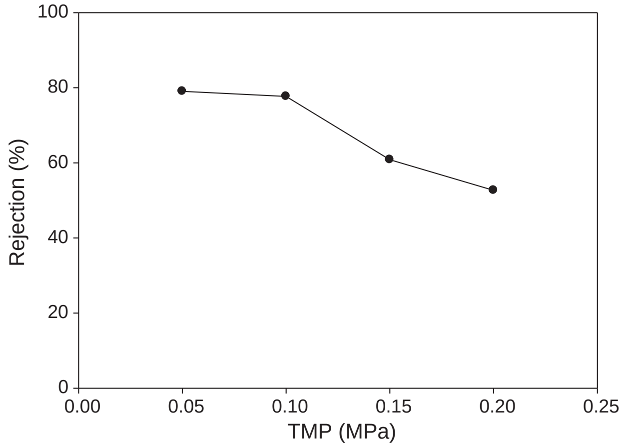

Fig. 5. Effect of the transmembrane pressure (TMP) on SDS surfactant rejection. Feed concentration of SDS, 10 mM; temperature, 25 °C.

micelles deposited quickly on the membrane surface and blocked in the membrane pores in a short time. Furthermore, the retentate stream was recycled to the feed tank, and the feed concentration of SDS increased gradually, so that membrane fouling assembled into cake formation along with the time. As a result of the permeate flux decreased more and more slowly and tended to be constant. However, below the CMC, the permeate flux decreased slowly than the concentration of 10 mM. The reason is that most of the SDS molecules existed as free monomers with the smaller size than pores, and then can easily get through the membrane pores.

The effects of feed SDS concentration on the membrane fouling resistance were investigated. As shown in Fig. 4(b), results for various SDS concentrations revealed that the sequential order of the additional membrane fouling resistance was 10 mM > 8 mM > 5 mM > 2 mM. According to Eqs. (1) and (2), additional resistance (R_f) increased from 0.78×10^{12} m^{-1} to 1.86×10^{12} m^{-1} at 10 mM. We observed that the additional membrane fouling resistance increased with time quickly at first, then increased slowly and gradually tended to be stable. The reason that contributed to this phenomenon was described below: with the increase of feed SDS concentration, the aggregation number of micelles increased and more micelles deposited on the membrane surface and blocked the membrane pores, increasing R_f by membrane fouling and concentration polarization.

3.2.3. Effect of the operating pressure on the SDS rejection

The effect of operating pressure on the rejection of SDS was investigated, as presented in Fig. 5. From the figure, it can be seen that SDS rejection decreased with an increase of operating pressure, ranging from 79% at 0.05 MPa to 53% at 0.20 MPa. It may be due to the following reasons: at higher pressure, surfactant micelles might be compacted and the increase of the effective driving force caused the increment of the convective transport of solutes filtered through the ultrafiltration membrane to the permeate solutions [25].

3.3. Membrane cleaning

Membrane fouling in membrane applications is inevitable. Therefore, necessary membrane cleaning is required. The usual cleaning process is hydraulic flushing. This study aims to analyze the effect of water rinse on membrane fouling. Tap water was used for forward flushing and backwashing. Forward flushing and backwashing are the two main ways of membrane cleaning. Fig. 6 shows the diagram of membrane filtration, forward flushing and backwashing.

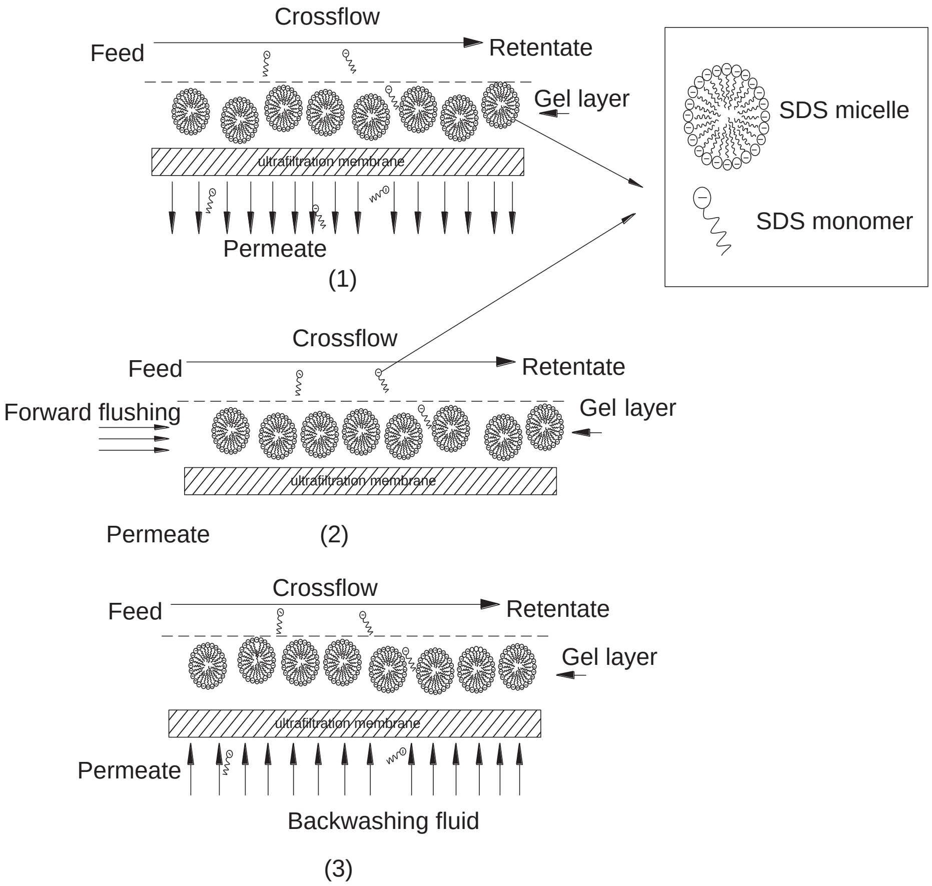

Fig. 6. Schematic diagram of removal of SDS from aqueous solution by micellar-enhanced ultrafiltration: (1) membrane filtration; (2) forward flushing; (3) backwashing.

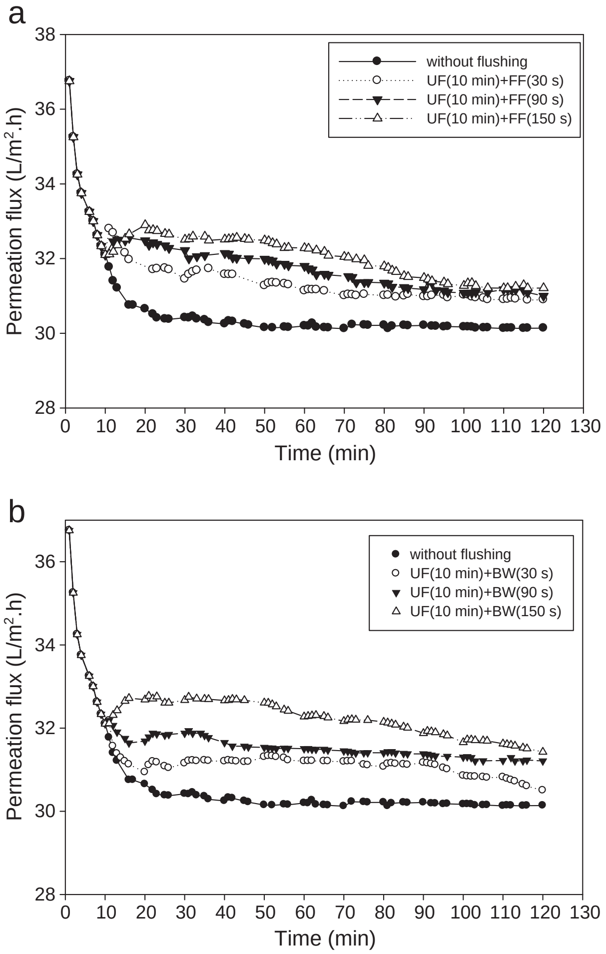

Fig. 7. Effects of forward flushing and backwashing on SDS permeate flux: (a) forward flushing, (b) backwashing.

Before each experiment, a clean membrane was compacted at 0.05 MPa for 40 min using distilled water, and then according to Eq. (1) calculated the pure water flux. When cleaning cycles were finished, the membrane permeability was measured with distilled water under the same conditions as the pure water flux provides a benchmark for gauging cleaning efficiency. SDS was added into 40 L distilled water and the SDS concentration is 10 mM (28.8 g/L). Apparently, the value of SDS permeate flux (J) reduced quickly in the first 10 min (Fig. 4(a)), and the additional membrane fouling resistance increased with the increase of time (Fig. 4(b)). So it means that after 10 min of filtration, it is imperative to wash the membrane. The following experiments were carried out for 2 h at room temperature, the feed SDS concentration of 10 mM, TMP of 0.05 MPa and periodic water flushing, namely, membrane filtration 10 min then rinse membrane a certain time, and then re-start filtering for a total of 12 cycles. To evaluate the effect of forward flushing and backwashing, filtration time was fixed at 10 min, forward flushing time (FF) and backwashing time (BW) were varied as 30, 90 and 150 s. Then the transmembrane pressure was fixed at 0.05 MPa, and permeate flux was calculated by weight of permeation solution.

3.3.1. Effects of forward flushing and backwashing on permeate flux

The cleaning interval between two successive filtration cycles is a very important parameter. As can be seen in Fig. 4(a), permeate flux decreased quickly at first, and then tended to be stable after 10 min filtration. Therefore, to save the cost of operation, the filtration time should be fixed at 10 min and cleaning cycle was 12.

Forward flushing tests were performed with the permeate valve closed in order to avoid permeation and obtain a high crossflow velocity without pressure. The results were shown in Fig. 7(a). This graph shows that as flushing time increases, the flushing effect of the membrane is getting better and better. Meanwhile, with the experiment progressing, membrane permeate flux reduced gradually until the leveled off. Without forward flushing, the flux was about 30.14 L/m^2·h after 2 h; with forward flushing, the flux increased about 30.90 L/m^2·h at FF (30 s), 30.95 L/m^2·h at FF (90 s), and 31.21 L/m^2·h at FF (150 s). This behavior can be explained by the fact that forward flushing can eliminate the deposition of foulants on the membrane surface due to concentration polarization. If forward flushing time is too short, it could not be able to flush foulants on the membrane surface completely. From the point of view of energy and water savings, forward flushing for 150 s is more appropriate. The three kinds of membrane flushing time effects tend to be close. This is because of the reason that during the forward flushing process, the permeate valve closed, then there is no convective transport, thus the fouling on the membrane surface was swept loosely and some SDS monomers were not washed away which can go through pores or adsorbed by the membrane internally. However, internal fouling in the pores cannot be swept away by forward flushing [33], which resulted in the phenomenon that forward flushing has no capacity to clear them. The process of backwashing is that push washing water (tap water) back through the membrane to remove the internal fouling present in the membrane pores. The permeation is forced in the reverse direction through the membrane. So the foulants on the membrane surface can be lifted off and the foulants lifted would be resuspended by tangential flow [12]. Fig. 7(b) shows the comparison of permeate flux in function of time when operated in the presence and absence of backwashing. When the backwashing was performed, the observed permeate flux increased significantly compared to without backwashing. The highest flux and lowest flux values were achieved at BW (150 s) and BW (30 s) respectively. The flux is 31.43 L/m^2·h at BW (150 s), 31.10 L/m^2·h at BW (90 s) and 30.50 L/m^2·h at BW (30 s) respectively. This phenomenon may be caused by the following reasons: pollutants inside the membrane were lifted by backwashing, backwashing could loosen and detach the particle from the membrane surface and the cross flow wiped debris away from the membrane surface [34]; with backwashing time increased, backwashing can preferably control the increase rate of membrane fouling. Nevertheless, backwash time should not be as long as possible.

Fig. 7 shows that the backwashing is more effective than forward flushing. Backwashing is capable of decreasing foulants and increasing the permeate flux, but if the backwashing time is too long, it could damage the membrane and shorten its service life, and requires higher energy than forward flushing [35].

3.3.2. Effects of combined flushing on permeate flux

To compare the effects of the single hydraulic cleaning mode on permeate flux, two combined washing methods (forward flushing followed by backwashing, backwashing followed by forward flushing) were investigated. The flux of backwashing followed by forward flushing is 28.13 L/m^2·h. Forward flushing followed by backwashing is 30.14 L/m^2·h after 12 cycles under the condition of flushing 30 s of time. So the following experiments were performed under the condition of forward flushing followed by backwashing.

The effects of combined flushing (forward flushing followed by backwashing) on membrane permeate flux was shown in Fig. 8 (a)–(c), in which the highest value of permeate flux could be maintained at FF (150 s) + BW (90 s) and was 33.05 L/m^2·h. Moreover, the lowest value could be found at FF (30 s) + BW (30 s), which was 31.04 L/m^2·h (as shown in Fig. 8(a)). These results indicate that hydraulic cleaning time may provide beneficial flux recovery. As indicated in Fig. 8(a), with the increases of backwashing time, the recovery efficiency of membrane flux increases gradually. That probably means forward flushing could be beneficial to flush

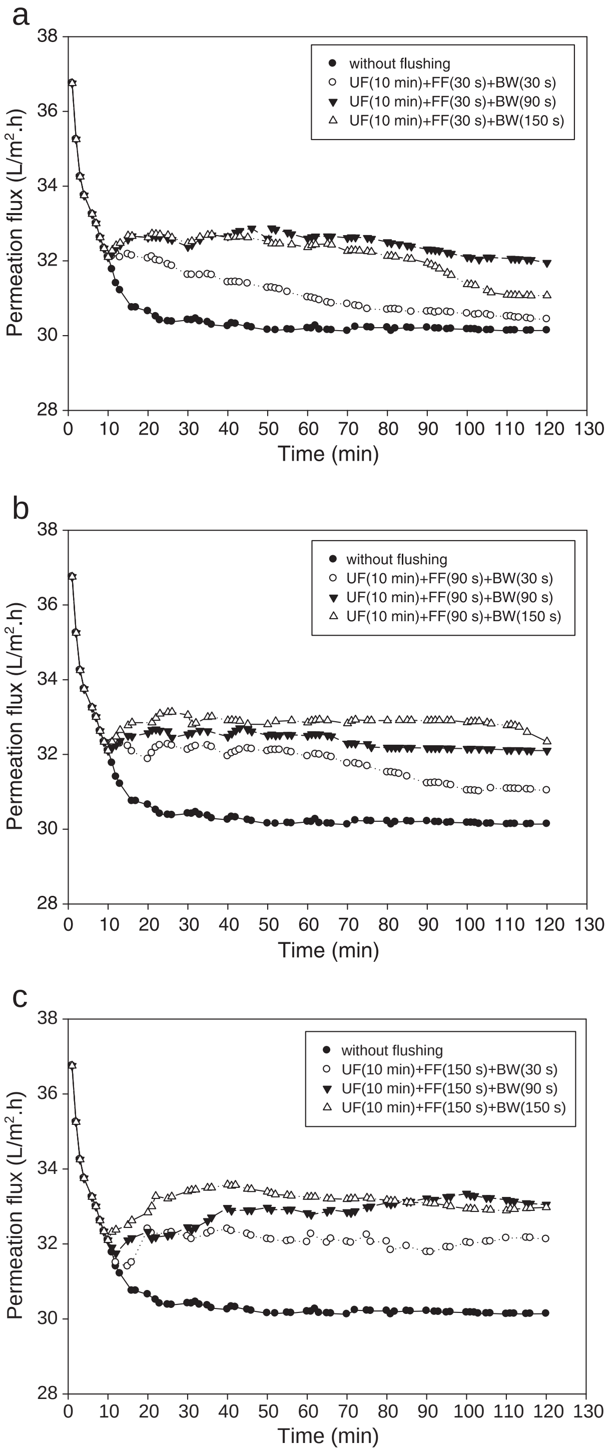

Fig. 8. Effects of combined flushing on SDS permeate flux: forward flushing followed by backwashing.

foulants on the membrane surface while backwashing could break up the foulants on the membrane pores and be helpful to rinse those foulants which are not being washed away by forward flushing. However, the longer backwashing time does not mean higher cleaning efficiency. The permeate flux of FF (150 s) + BW (150 s) is higher than FF (150 s) + BW (90 s) at first but slightly lower after 80 min. The impact of forward flushing works on particle and gel layers at the membrane surface and backwashing loosens contaminates which adsorbed the membrane internal, then the combination of forward flushing and backwashing could bring higher permeate flux than single flushing. The flux recovery values of the PES membrane for SDS solutions at various flushing modes are given in Table 2.

Table 2
The flux recovery values of PES membrane for SDS solutions at various flushing modes (Feed concentration of SDS: 10 mM, at room temperature, 0.05 MPa).

Flushing mode	J_w ($\mathrm{L/m^2\cdot h}$)	J_f ($\mathrm{L/m^2\cdot h}$)	FR (%)
Without flushing	44.00	30.15	68.52
FF (30 s)	43.35	36.00	83.04
FF (90 s)	43.65	36.75	84.19
FF (150 s)	43.50	37.20	85.52
BW (30 s)	43.50	34.65	79.66
BW (90 s)	43.35	36.90	85.12
BW (150 s)	43.65	37.50	85.91
FF (30 s) + BW (30 s)	43.95	34.95	79.52
FF (30 s) + BW (90 s)	43.50	38.25	87.93
FF (30 s) + BW (150 s)	43.73	37.20	85.07
FF (90 s) + BW (30 s)	43.80	37.20	84.93
FF (90 s) + BW (90 s)	43.65	38.70	88.66
FF (90 s) + BW (150 s)	43.43	39.60	91.18
FF (150 s) + BW (30 s)	43.50	39.00	89.66
FF (150 s) + BW (90 s)	43.80	40.73	92.99
FF (150 s) + BW (150 s)	43.73	40.50	92.61

4. Conclusions

The performance of MEUF with hydraulic flushing for treatment of SDS and some parameters related to membrane fouling in MEUF such as feed concentration and TMP were investigated. With the increase of TMP, membrane flux increased from 30.20 $\mathrm{L/m^2\cdot h}$ at 0.05 MPa to 80.15 $\mathrm{L/m^2\cdot h}$ at 0.20 MPa. However, SDS rejection decreased with an increase of TMP, ranging from 79% at 0.05 MPa to 53% at 0.20 MPa. In this study, the feed concentration of SDS has an influence on the additional membrane fouling resistance, and 10 mM SDS causes the biggest fouling resistance and lowest permeate flux. The effects of forward flushing, backwashing and combined flushing (forward flushing followed by backwashing) on membrane permeate flux were discussed. Backwashing is an available rinsing method by dislodging the particles which formed a cake layer blocking the membrane pores, but it may also damage the membrane. Forward flushing could easily remove membrane surface foulants by means of crossflow rinse. It was proved in this study that the combination of flushing (forward flushing followed backwashing) could alleviate fouling caused by the deposition of particles on the membrane surface and adsorbed in the membrane interior and be more effective than the other two hydraulic flushing modes.

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