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Membrane Concentrate Thickening by Hollow-fiber Microfilter in Drinking Water Treatment Processes

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Abstract: A novel system to thicken the concentrated colloidal solution from membrane water treatment processes was developed. A hollow-fiber microfilter(hydrophilic polyethylene, nominal pore size 0.1 μm, total surface area 0.42 m^2) was installed in an acrylic housing that has an aeration port 5 cm below the membrane, and a clarifier in the bottom. The concentrate was uniformly supplied from the top of the housing. Vacuum filtration caused downward flow of concentrate and, as a result, thickening interface. The addition of poly-aluminum chloride (PAC) resulted in rapid increase of trans-membrane pressure (TMP) and in no improvement of the filtered water turbidity and thickening process. Two types of concentrate and concentrate turbidity had little effect on the increase of TMP and concentrate thickening. It was observed that, for the same height of membrane housing, membrane surface area to housing volume (A/V) ratio had significant effect on the increase of TMP. When the housing volume was increased ten times, the increasing rate of TMP was three times faster as compared to the original housing. A hydraulic model successfully simulated the formation and sedimentation of thickening interface.

1. Introduction

Although water treatment processes using either microfilter or ultrafilter guarantee high quality drinking water [1,2], the problem of treatment and disposal of concentrate has not yet been resolved [3]. Normally, membrane water treatment processes do not use coagulant to reduce cost and sludge production. Five to ten percent of raw water is discharged as concentrate, containing up to several hundreds mg/l of suspended solids. The suspended colloidal solids in concentrate, therefore, cannot be thickened in a conventional gravity thickener without flocculation.

Hollow-fiber membrane filtration of high colloidal solution such as activated sludge in aeration tank has been difficult because of inter-fiber clogging on the membrane surface by suspended particles [4,5]. In this study, a novel system that can prevent inter-fiber clogging and cake-layer formation was developed for concentrate thickening in the membrane water treatment processes.

2. Materials and Methods

Fig. 1 shows the schematic diagram of experimental apparatus. A hollow-fiber membrane made of hydrophilic polyethylene and having nominal pore size of 0.1 μm and total surface area of 0.42 m^2 is used in this experiment. Table 1 shows the specification of membrane. This hollow-fiber membrane was installed in an acrylic housing that has an aeration port 5 cm below the membrane. The filtration section was separated from the clarifier section to prevent sediment from being stirred by air scrubbing. The feed was uniformly supplied

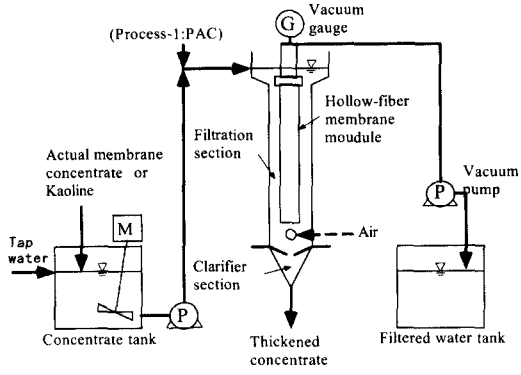

Fig. 1. Schematic diagram of experimental apparatus.

Table 1. The Specification of Membrane

Type	Hollow-fiber Microfilter
Material	Hydrophilic Polyethylene
Membrane Surface Area	0.42 m^2
Nominal Pore Size	0.1 μm

from the top of the membrane housing and filtration was driven by a vacuum pump. The membrane was physically washed by air-scrubbing in every 30 minutes, and thickened concentrate was partially removed from the clarifier to lower the surface of thickened concentrates.

2.1. Experiment I: Effect of Addition of PAC

Effect of addition of poly-aluminum chloride (PAC) as a coagulant was investigated. There were two processes, namely Process-1 and Process-2. PAC was added at a concentration of 10 mg/l in Process-1, whereas PAC was not added in Process-2. Membrane filtration was operated at a flux of 0.5 m/d using synthetic (kaoline) concentrate with 100 NTU. Table 2 shows the operating conditions.

2.2. Experiment II: Effects of Type of Concentrate, Concentrate Turbidity and A/V Ratio

Effects of the types of concentrate, concentrate turbidity and membrane-surface area to housing volume (A/V) ratio were investigated. Two types of concentrate, actual membrane concentrate (Process-1) and synthetic concentrate with kaoline (Process-2) were used in the experiments. The concentrate turbidity was raised to 200 NTU in Run-2 from 100 NTU in Run-1. Housing volume was increased by a factor of 10 in Run-3 though the height remained the same as before. Table 3 shows the operating conditions.

Table 2. Operating Condition of Experiment I

	Process-1	Process-2
PAC	10 [mg/l]	-
Type of Concentrate	Kaoline	
Concentrate Turbidity	100 [NTU]	
Flux	0.5 [$m^3/(m^2day)$]	
Operation cycle	Filtration: 30[min] Air-scrubbing & Addition of PAC: 1[min] Sedimentation: 3[min] Draining: 1[sec] (80[ml/cycle])	

Table 3. Operating Condition of Experiment II

		Run-1	Run-2	Run-3
Type of Concentrate	Process-1	Actual membrane concentrate		
	Process-2	Kaoline		-
Concentrate Turbidity		100 [NTU]	200 [NTU]	100 [NTU]
A/V ratio		600 [m^2/m^3]		60 [m^2/m^3]
Flux		0.5 [$m^3/(m^2day)$]		
Operation cycle		Filtration: 30[min] 4 Cycles Air-scrubbing: 3[min] 4 Cycles Draining: 1[sec] (80[ml/cycle])		

3. Results and Discussions

3.1. Experiment I: Effect of addition of PAC

Fig. 2 shows time course of turbidity and the trans-membrane pressure(TMP) in experiment I. Both Process-1 (added with PAC) and Process-2

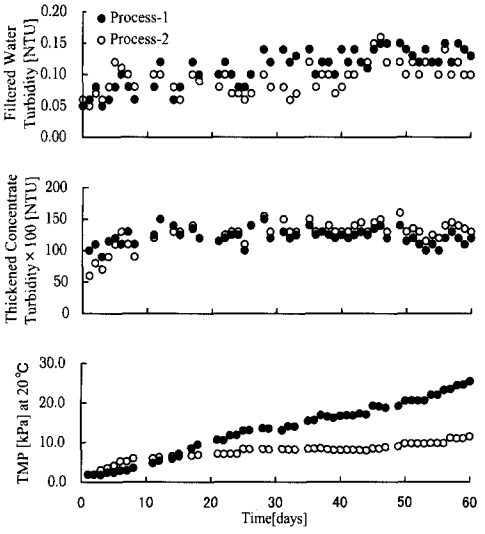

Fig. 2. Time course of turbidity and TMP in experiment I.

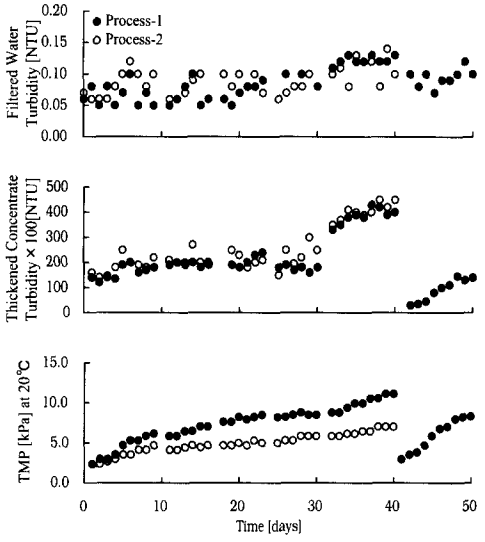

Fig. 3. Time course of turbidity and TMP in experiment II.

(without PAC), were operated for 60 days.

Filtrate from both processes had turbidity of less than 0.15 NTU, and concentrate was thickened 120 times in the clarifier. TMP in Process-2 increased to 11.5 kPa after 60 days of operation, whereas TMP in Process-1 increased to 25.6 kPa.

After the experiment, the membrane filter was removed for analysis. The cakelayer was not observed on the membrane surface in Process-2, whereas thick cake layer was observed in Process-1 due to floc attachment on the membrane surface. Addition of PAC caused no improvement in thickening. It was found that flocs are more likely to adhere to the membrane surface than colloidal particles.

3.2. Experiment II: Effects of Type of Concentrate, Concentrate Turbidity and A/V Ratio

Fig. 3 shows time course of turbidity and TMP in experiment II. The membrane thickener was operated for nearly 50 days including three Runs.

The turbidity of filtrate was consistently less than 0.15 NTU regardless of the type of concentrate and concentrate turbidity. Both in Run-1 and Run-2, the thickened concentrate was about 200 times higher than the influent concentrate. TMP increase rate in Runl and Run-2 were 0.21 kPa/d and 0.26 kPa/d for Process-1, and 0.12 kPa/d and 0.12 kPa/d for Process-2, respectively.

The effect of membrane surface area to housing volume (A/V) ratio was investigated. The housing was replaced with another one having volume ten times higher than that of Process-1 on the 40th day (Run-3). When new membranes were installed and filtration was commenced, TMP increased rapidly at 0.60 kPa/d. A lower A/V ratio (from 600 m^2/m^3 to 60 m^2/m^3) caused an increase in TMP (from 0.21 kPa/d to 0.60 kPa/d) because downward flow velocity had decreased and cake-layer formation was not prevented. The water in the housing moved horizontally when A/V ratio was 60 m^2/m^3 as opposed to the downward flow in the filtration with the high A/V ratio (600 m^2/m^3).

3.3. Simulation of Thickening Interface Settling

A hydraulic model has been simulated for the formation and sedimentation of thickening interface. Fig. 4 illustrates the hollow-fiber membrane in the housing.

According to the *Bernoulli theory* and fraction of

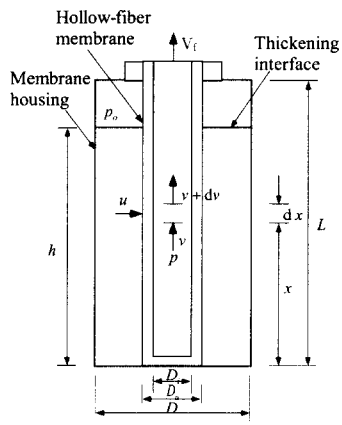

Fig. 4. Hollow-fiber membrane in housing.

Hagen-Poliseuille fluid flow, the following equation is given:

$$p+\frac{v^2}{2g}=p+dp+\frac{(v+dv)^2}{2g}+\left(\frac{32\mu v}{\rho g D_i^2}\right)dx \quad (1)$$

x = distance from bottom of hollow-fiber membrane (m)

p = pressure at x (m)

v = velocity of fluid flow at x (m/sec)

g = gravity (=9.80 m/sec^2)

μ = coefficeint of viscosity (=10^{-3} kg/(msec))

ρ = water density (=10^3 kg/m^3)

D_i = diameter of hollow-fiber membrane inside (=0.27×10^{-3} m)

By omitting minute term in Eq.(1) and simplifying, Eq.(2) is obtained.

$$\frac{dp}{dx}=-\frac{v}{g}\frac{dv}{dx}-\frac{32\mu v}{\rho g D_i^2} \quad (2)$$

By the law of mass conservation, the flow rate inside the hollow-fiber membrane is the same as the flow rate into the hollow-fiber membrane from the outside surface.

$$\frac{\pi}{4}D_i^2 dv=\pi D_o u dx \quad (3)$$

D_o = outer diameter of hollow-fiber membrane (= 0.41×10^{-3} m)

u = membrane permeability (m/sec), therefore,

$$\frac{dv}{dx}=\frac{4D_o u}{D_i^2} \quad (4)$$

The membrane permeability is proportional to the pressure difference across the membrane.

$$u=\kappa(P_o-P) \quad (5)$$

κ =coefficient of permeability (κ_o=3.0×10^{-5} sec^{-1})

P_o=ambient pressure (m)

By substituting Eq.(5) into Eq.(4) and differentiate for x on both sides, Eq.(6) is derived.

$$\frac{dp}{dx}=-\frac{D_i^2}{4\kappa D_o}\frac{d^2v}{dx^2} \quad (6)$$

Substitution of Eq.(6) into Eq.(2) and simplification gives the following equation.

$$\frac{dp}{dv}=-\frac{v}{g}\left\{1+\frac{8\mu}{\rho D_o\kappa(p_o-p)}\right\} \quad (7)$$

The first term on the right side of Eq.(7) could be ignored because it is much smaller than second term.

$$\frac{dp}{dx}=-\frac{32\mu v}{\rho g D_i^2} \quad (8)$$

Then, under the boundary conditions [v=0 when x=0, v=v_f when x=L] the following equation is obtained.

$$v=v_f\frac{\sinh(\sqrt{\alpha}\cdot x)}{\sinh(\sqrt{\alpha}\cdot L)} \quad (9)$$

$$\alpha=\frac{128\mu\kappa_o D_o}{\rho g D_i^4}$$

L = length of hollow-fiber (=0.33 m)

v_f = velocity of fluid flow at L (m/sec)

The velocity of thickening interface sedimentation

is proportional to fluid flow into hollow-fiber membrane at t.

$$-\frac{dh}{dt} = \beta \cdot v_f \frac{\sinh(\sqrt{\alpha} \cdot h)}{\sinh(\sqrt{\alpha} \cdot L)} \quad (10)$$

$$\beta = \frac{nD_i^2}{D^2 - nD_o^2}$$

D= inner diameter of housing (= 0.44×10^{-1} m)

h = height from hollow-fiber bottom to thickening interface (m)

n = number of hollow-fiber (=960).

By integrating Eq.(10) and substituting the initial condition [$h=h_o$ when $t=0$], the height of thickening interface at time t could be expressed as in Eq. (11)

$$h = \frac{2}{\sqrt{\alpha}} \tanh^{-1}\left[\tanh\left(\frac{\sqrt{\alpha} h_o}{2}\right) \exp\left\{\frac{\sqrt{\alpha} \cdot \beta \cdot v_f}{\sinh(\sqrt{\alpha} \cdot L)} t\right\}\right] \quad (11)$$

As the filtration proceeds, the water moved downward due to withdrawal of water by a vacuum filtration pump. Accordingly, suspended solids concentration has gradually increased because solid particles were rejected by the membrane.

The simulated heights of thickening boundary were drawn in Fig. 5. The measured height were found to be in close agreement to the calculated ones. Concentrate turbidity of filtration and clarifier sections was also measured during the filtration process. The suspended solids concentration around the membrane filter was approximately the same as the influent, whereas it was about 200 times higher underneath in the clarifier. The above simulation based on hydraulic model proved that the settling of thickening boundary was driven by water displacement due to membrane filtration.

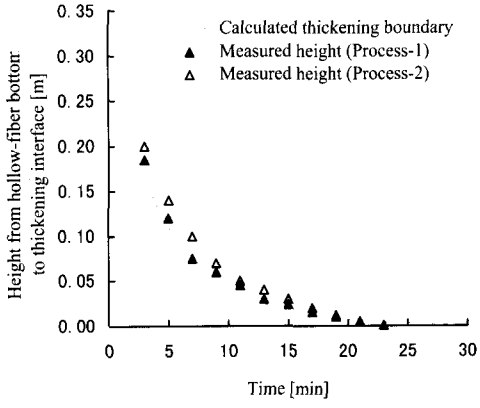

Fig. 5. Height of thickening interface with time.

4. Conclusions

In order to develop a thickening process for the concentrated colloidal solution in membrane water treatment processes, a novel system based on a new concept of solids thickening has been developed. This system showed that cake-layer formation and inter-fiber clogging on the membrane surface were successfully averted by downward flow that was created by vacuum filtration. The downward flow was herein caused by the displacement of water that was filtered out through the membrane. As a result, thickened colloidal solution was collected at the bottom of membrane housing, and concentrate turbidity around the membrane remained nearly the same as the influent solution. Simulation of downward flow in the housing demonstrated that settling of thickened interface was caused by downward flow due to water withdrawal through the membrane filter.

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