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CONTROL OF TWO-STAGE CYCLIC ANAEROBIC PROCESSES BASED ON OPTIMIZATION OF CYCLE TIME

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Abstract : A process control index is proposed for the determination of an optimum cycle time, which is a crucial control parameter of Two-Stage Cyclic (TSC) process. Experiments were performed in 13 liter working volume hybrid anaerobic filter sets that were connected two reactors in series at several organic loading rates between 2.0-9.0 g soluble COD/L/d. The results indicated that the performance of the two-stage cyclic anaerobic filter system was closely related to the cycle time. The optimum cyclic time could be determined by using an operational index, [$S_O\tau_c$/(M], which includes influent wastewater COD (S_o), a cycle time (τ_c), a hydraulic retention time (θ), and biomass inventory (M). This optimum cycle time also corresponded to the time when the gas production in the second (follow) stage became less than 10 % of the total gas production. Tracer tests also were conducted for the clean bed and dirty bed conditions. The results showed that the hydraulic pattern of the upflow hybrid anaerobic filters under normal operating conditions demonstrated performance similar to that of a plug flow reactor with strong dispersion. Therefore, biogas mixing and biomass accumulation under normal operating conditions had a significant effect on reactor hydraulics.

Key Words : Anaerobic filter, anaerobic treatment, methane formation, biogas, tracer test, two-stage

INTRODUCTION AND OBJECTIVES

The two-stage cyclic anaerobic filter system is an advanced process that two equal-volume reactors are operated in series (Howerton and Young, 1987[1]; Siddique and Young, 1995[2]; Dahab and Kalagiri, 1996[3]). After a given period of operation, the flow regime is reversed so that the follow reactor becomes the lead reactor and *vice versa* (Figure 1). The basic concept behind the two-stage, cyclic (TSC) process is that biomass synthesis is high in the lead reactor because of the high waste strength and high loading. The second, or follow, reactor operates in a decay mode so that active biological solids remaining from the previous lead position continue to remove soluble organic constituents to lower concentrations than that would be provided by a single-stage unit. In order to maintain the advantages the frequency of flow reversal seems important; if the flow reversal occurs too frequently, the reactors will perform like a single stage reactor. On the other hand, if the flow reversal occurs rarely, the two-reactor system will perform as if it were a two-stage unidirectional flow system. However, no research has been conducted previously to provide guidance on the flow reversal time. Consequen-

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tly, studies were conducted using laboratory-scale, two-stage cyclic anaerobic filters that received brewery wastewater as a test substrate. Specific objectives were: 1) to investigate the optimum reversal time of the two-stage cyclic anaerobic filter process, 2) to analyze the effect immediately after the flow reversal on both lead and follow reactors, 3) to study the hydraulic patterns under clean and dirty bed conditions by tracer tests, and 4) to evaluate the performance of the ceramic tubular shaped media used in the laboratory-scale reactors.

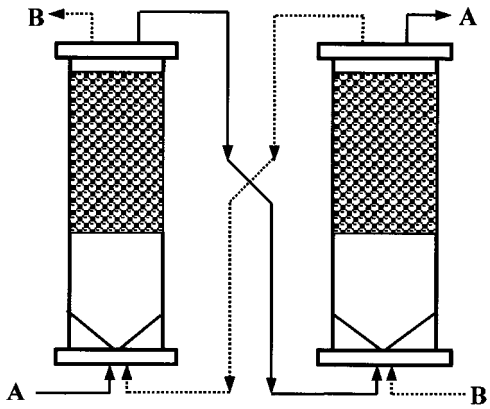


Figure 1. Schematic diagram of the two-stage cyclic process.

Tracer Study. The performance of reactors is affected by the degree of contact between incoming substrate and a viable bacterial population (Heertjes and Kuijvenhoven, 1982[4]; Samson *et al.*, 1985[5]; Chiang *et al.*, 1992[6]). Mixing characteristics depend on the reactor geometry, the type of medium, inlet and outlet design, the biogas production rate, the hydraulic aspects, flow velocities of the feed, and the effluent recycle and any other source of mixing. In anaerobic fixed film reactors, mechanical mixing is not generally possible. Tracer studies have been performed frequently to assess the hydraulic characteristics of reactor vessels and also to evaluate the effect of biofilm development on mixing in the upflow anaerobic filter. Results that were obtained from the tracer tests were used to produce a dispersion number as follows:

$$\sigma_\theta^2 = 2D/\mu L - 2(D/\mu L)^2(1-e^{-\mu L/D}) \quad (1)$$

in which,

$$\sigma_\theta^2 = \sigma^2/\bar{t}^2 \quad (2)$$

and

$$\sigma^2 = \frac{\sum t_i^2 C_i}{\sum C_i} - \bar{t}^2 \quad (3)$$

$$\bar{t}^2 = \left(\frac{\sum t_i C_i}{\sum C_i}\right)^2 \quad (4)$$

where, $(D/\mu L)$ = dispersion number,
σ^2 = variance
t_i = time at i, hr
$\bar{t}$ = mean residence time, hr
C_i = tracer concentration in effluent at time i, mg/L

The dispersion number is a measure of the deviation from the perfect plug flow pattern. Hence, dispersion number, $D/\mu L=0$, represents perfect plug flow and $D/\mu L = \infty$ is a completely mixed flow. According to Levenspiel (1957[7]), $D/\mu L = 0.2$ represents a large amount of dispersion, $D/\mu L = 0.025$ is an intermediate amount of dispersion, and $D/\mu L = 0.002$ is a small amount of dispersion. To take into account the dead space for the portion of the fluid, which was poorly mixed, the following expression was used:

$$C/C_o = V/V_m\, e^{-(V/V_m)t/\bar{t}} \quad (5)$$

where, C = measured concentration of tracer, mg/L
C_o = initial tracer concentration, mg/L
V = total volume of reactor, mL
V_m = volume of mixed zone, mL
t = time, hr

To take into account the portion of the fluid, which passed as short-circuiting, the following expression was used:

$$C/C_o = (v_1/v)^2 e^{-(v_1/v)t/\bar{t}} \quad (6)$$

where v_1 = the portion of fluid which was well mixed, mL/day
v = the total flow rate, mL/day

If both short-circuit and dead space were present, the following expression was used:

$$C/C_o = [(v_1 V/v)/t]\, e^{-t/\bar{t}} + (v_2/v)\delta_{t=0} \quad (7)$$

where, v_2 = bypass flow rate
$\delta_{t=0}$ = Dirac delta function

V_m, v_1 and v_2 in this equation were estimated using linear regression analysis.

Gas Production. The destruction of organic matter is the primary objective of anaerobic digestion, and the quantity of gas produced during digestion is directly proportional to the amount of organic matter destroyed, so methane production is probably the most sensitive process performance indicator. Therefore, the ratio of gas production between the lead reactor and the follow reactor can also be indicator of the performance of two-stage cyclic system. The influent (S_o) and effluent (S) concentration (usually measured as COD) provide the quantity of substrate utilized per unit volume, which in turns leads to the estimation of gas production. The gas production ratio can be formulated as follows:

$$\frac{dG}{dt} = P_g \frac{dS}{dt} \quad (8)$$

Process Load Factor (Substrate to Microorganism Ratio). The growth of biomass in the reactor was assumed to follow Monod kinetics with a first-order decay rate. The process load factor was calculated as follows:

$$\frac{dS}{dt} = \frac{k_{max} SM}{(K_s + S)} \quad (9)$$

Substituting Eq. 9 in Eq. 8 yields the following:

$$\frac{dG}{dt} = P_g \frac{k_{max} SM}{(K_s + S)} \quad (10)$$

Where :
G = gas production, L.
P_g = conversion constant, L gas/g COD removed.
M = biomass concentration, mg/L.
k_{max} = maximum substrate utilization rate, mg COD/mg VSS/day
K_s = half saturation coefficient, mg/L.
S = effluent substrate concentration, mg/L.
S_o = influent substrate concentration, mg/L.

Process Control Index. The optimum cyclic time could be determined by using an operational index, $[S_o\tau_c/\theta M]$, which includes influent wastewater, COD (S_o), cycle time (τ_c), hydraulic retention time (θ), and biomass inventory (M). The index that applied to this study was as follows:

$$\alpha = \frac{\tau_c S_o}{\theta M} \quad (11)$$

Eq. 11 can be expressed as

$$\alpha = \tau_c (F/M) \quad (12)$$

Assuming that the substrate degradation followed Monod kinetics (Eq. 9), the relationship between the substrate removal efficiency and the control index at steady state becomes:

$$\frac{S_o - S}{S_o} = \frac{k_{max} S}{K_s + S} \frac{VM}{QS_o} \quad (13)$$

Therefore, the process efficiency can be estimated by Eq. (14).

$$E = \frac{k_{max} S}{K_s + S} \frac{100}{F/M} \quad (14)$$

Substituting for F/M from Eq. 12 into Eq. 10 and 14 the process efficiency can be expressed in terms of dG/dt:

$$E = \frac{1}{P_g} \left(\frac{dG}{dt} \right) \frac{\theta}{S_o} 100 \quad (15)$$

Experimental Setup of the Two-Stage Cyclic Reactors

To conduct the two-stage cyclic operation, six laboratory-scale filters (three two-stage sets) were used. Each filter was 12 cm in diameter and 74 cm height, as shown in Figure 2. The reactors were packed with 25 mm and 10 mm ceramic tube shaped media. The two of three sets were operated in the cyclic mode. The other set was operated in unidirectional mode as a control set. The reactors were fed with wastewater from a brewery using peristaltic pumps and operated at 35°C in an incubator. Analytical measurements were made using standard or otherwise published procedures.

Wastewater Characteristics. The chemical characteristics of the brewery wastewater are listed in Table 1. Nitrogen and phosphorus were present in sufficient quantities to support biological growth; those were 40 and 23 mg/L, respectively. The wastewater was passed through a #140 standard sieve (sieve opening; 106㎛) to remove large particles, and was stored at 4°C in a refrigerator. The wastewater was fortified with beer to achieve the desired COD concentration depending on reactor loading requirements. Sodium bicarbonate was used for maintaining pH between 6.8 and 7.2.

Filter Media. The reactors were packed with two different sizes of ceramic tube-shaped media. Each media has the specific surface areas of 6.17 m^2/g and 12.51 m^2/g, respectively. The larger media was 25 mm in external diameter, and the smaller was 10 mm in external diameter. The smaller media was placed on the top of the larger media zone, which was supported by stainless steel wire mesh. The media occupied

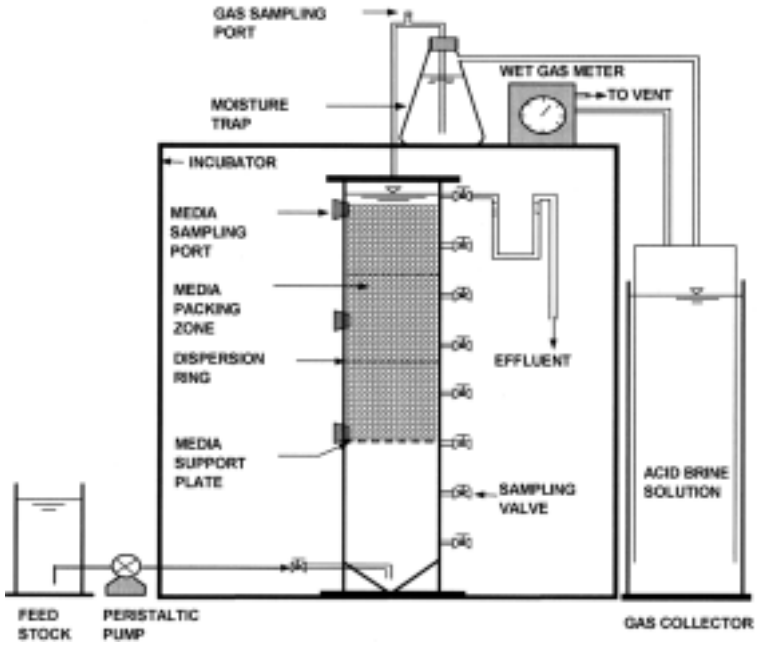


Figure 2. Schematic diagram of the laboratory- scale upflow anaerobic filter.

Table 1. Composition of brewery wastewater used in this study (units: mg/L)

Parameters	Values
pH	4.5 ~ 5.0 (unitless)
$TCOD_{cr}$	2,000 ~ 3,000
$SCOD_{cr}$	1,500 ~ 2,500
TSS	270 ~ 350
VSS	150 ~ 220
T-N	20 ~ 60
T-P	19.4 ~ 26.5
Cu	0.02 ~ 0.05
Fe	0.51 ~ 0.62
Si	4.75 ~ 6.10
Zn	0.76 ~ 0.12
Al	1.05 ~ 1.52
Mg	29 ~ 50
Ca	95 ~ 110
Na	250 ~ 350
K	110 ~ 130
Mn	N.D.
Cr	N.D.

56 percent of the empty bed volume of the reactor.

Tracer Test. The tracer test was carried out before inoculation (i.e. clean bed) and after 500 days of operation (i.e. dirty bed) to evaluate the fluid flow pattern in the reactors with respect to biomass attachment on the media. The flow pattern in the biogas-producing reactor was compared with that of the non-biogas producing reactor. The tracer was lithium chloride, which was injected 50 mL of 200 mg Li^{+}/L as a pulse into the reactors through the influent lines. Air injection was done by running a # 16 Masterflex pump into the bottom influent line with a single point application. Although this approach did not exactly simulate normal anaerobic filter gassing patterns, it was considered appropriate because most of biogas is produced in the lower parts of the filter. Comparisons between the clean bed with gassing and dirty-bed studies helped to show the effect of biomass accumulation on the hydraulics of the anaerobic filter system. Upon injection of the lithium tracer, 25-mL samples were collected over a period of about twice the detention time (about 24 hours). The sampling frequency was every thirty minutes within the period of the theoretical detention time and every hour thereafter. For the dirty-bed studies, samples were filtered with a fiberglass filter soon after collection to prepare the sample for lithium analysis. The reactor flow rate was also measured every hour during the study. The average flow rate was used to calculate the theoretical detention time.

Start up. The reactors were seeded with the sludge originating from the primary anaerobic digester at Cheongju, Korea, municipal sewage treatment plant located near the study area. The sludge was screened through a #25 sieve to remove large chunks. The final seed sludge contained 9,300 mg/L volatile suspended solids (VSS) and 6,400 mg/L fixed suspended solids (FSS). Prior to inoculation, each reactor was filled with water and then flushed with nitrogen to expel air. Seed sludge was added to the lower one-third of each reactor to avoid accumulation of the particulate solids in the voids of the media bed. All six reactors were operated in parallel during the initial acclimation period. After acclimation, two reactors were connected in series to set-up the two-stage system. Therefore, three separate systems were formed using six reactors. One was used as a control, which was operated in a unidirectional mode. The other two systems were operated in cyclic mode so that the lead reactor periodically became the follow reactor and *vice versa*. All tests were conducted at 35°C when operating at system organic loading rates varying from 2.0 to 9.0 g SCOD/L-day and at system HRTs ranging from 1.0 to 6 days.

Analytical Methods. Samples were withdrawn in succession from the top to the bottom of the filters using a syringe. In this manner, an undisturbed sample could be obtained at each sampling port. The pH of each sample was measured within five minutes to minimize pH changes caused by the loss of dissolved carbon dioxide. An Orion 720A pH meter (U.S.A) equipped with a glass electrode was used to make this determination. Chemical Oxygen Demand (COD) was determined by the closed tube dichromate reflux method described by

Standard Methods (APHA, 1995[8]).

Sample pretreatment for the SEM photographs of the biofilm on the surface of the media was accomplished with the following procedure; place sludge samples in sealed 50-mL serum bottles, which contained 2.5% glutaraldehyde in anaerobic 0.05 M Cacodylate buffer. Fixation was carried out overnight at 4°C. The fixed sludge samples were washed with buffer three times, and again fixed with 1% (OsO_4) for 1.5 hours, and again washed with buffer. The sludge was then dehydrated using a graduated series of ethanol in distilled water from 50 to 100% (v/v). The specimens were critical point dried in CO_2 and gold coated. The prepared specimens were examined with a SEM (Hitachi S-2500C, Japan) (Lazarova, 1992[9]).

Metals contained in the influent were determined by Inductively Coupled Plasma (ICP) (JY-38 Plus). The pH, biogas quantity, and effluent volumes, the temperature of the reactors were monitored daily. Influent COD, effluent COD, total alkalinity and the reactor sludge bed height were determined on alternate days. This study used pH 4.0 instead of pH 4.3 as proposed by Jenkins *et al.* (1983[10]) to utilize the direct titration method for an alkalinity endpoint as volatile acids. COD concentrations versus reactor height were evaluated by withdrawing samples from seven sampling ports on the last day of each flow reversal. Analysis of samples was performed in accordance with Standard Methods (APHA, 1995).

RESULTS

After about 60 days of operation, all reactors seemed to be acclimated judging from gas production. Table 2 summarizes the pseudo steady-state performance during unidirectional (control) and two-stage cyclic operation. A soluble COD removal efficiency of 98.2 % was obtained in the control system when it was operated at a system loading rate of 2.1 g SCOD/L-day. Before the first flow reversal for Systems I and II, soluble COD removal efficiencies were 97.0 % and 96.7 % at a loading rate of 2.1 g SCOD/L-day, respectively. At the higher organic loading rate, the differences of performance between the control system and the cyclic systems were distinct. This observation demonstrated that the two-stage cyclic process provided improved efficiency over a single stage system.

Mixing Pattern in the Hybrid Anaerobic Filters. Figure 3 and Table 3 show the results of the lithium tracer studies conducted on the clean bed with and without gas mixing. As indicated, all the reactors showed very large dispersions ($T/T_d << 1.0$ in Table 4). Tracer recovery ratios (R) were 96 (99% under clean bed conditions. Figure 4 and Table 3 show the responses of the tracer washout when operating at a COD loading rate of 6.5 g SCOD/L-day at that moment the test was conducted after 500 total days of operation. Mixing characteristics of hybrid anaerobic reactors are a function of growth of bacteria on the surface of media. Compared with the clean bed condition, the dirty-bed condition indicates that short-circuiting occurred due to the accumulated biomass (lower T/T_d).

The terms in Figures 3 and 4 and Table 3 are defined as follows: C_i = concentration of

Table 2. Pseudo steady-state performance for the control and two-stage, cyclic reactors

OLR (gSCOD/L/day)		Soluble COD removal efficiency (%)					
		Control		System I		System II	
		Lead	Follow	Lead	Follow	Lead	Follow
Unidirectional operation	2.1	88.1	98.2	84.5	97.0	85.3	96.7
Cyclic operation	2.7	86.3	98.1	85.2	98.4	84.5	98.5
	3.5	63.5	92.3	87.2	97.5	85.2	98.3
	6.2	64.0	85.3	66.4	94.3	65.3	92.8
	8.5	59.0	70.2	62.4	85.3	63.0	87.1

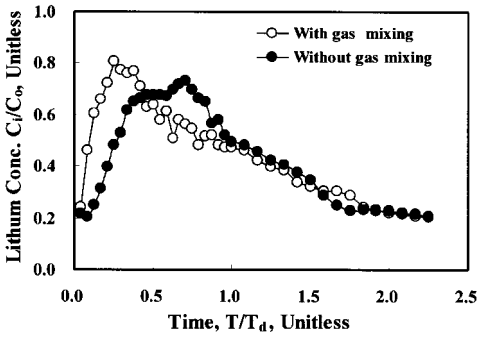


Figure 3. Lithium tracer washout characteristics of clean-bed.

lithium in reactor effluent at any time, C_o = average initial reactor lithium concentration (total slug input divided by clean-bed reactor liquid volume), T = mean residence time. T_d = theoretical reactor detention time (clean bed/flow rate). In an ideal plug flow condition, no lithium should appear in the effluent until $T = T_d$ and then it should appear at the same time, C_i should be equal to C_o instantly upon the slug feeding of lithium chloride ($C_i/C_o = 1$). Variations from these two extremes reflect intermediate degrees of short-circuiting in the reactors (Chiang, 1992). As seen in Table 4, both reactors in clean-bed condition had a very large dispersion with a dispersion number of 0.27 to 0.33. Even in clean-bed condition, short- circuiting was observed based on the T/T_d ratio of the reactors. Interestingly, gas-mixing produced more short-circuiting and less dispersion. It was probably due to the fact that gas bubbles contributed tracer washout through channels. In clean-bed tracer recovery ratio was 99.1 % for gas mixing and 95 % for non-gas mixing. The tracer recovery ratios of the dirty-bed studies ranged from 91 to 98 % except for the system II-lead reactor. This high recovery ratio indicates that lithium chloride was an appropriate tracer except System II resulted from an unknown reason. Figure 4 shows the responses of the six dirty-bed reactors under operation with a system organic loading rate of 6.5 g SCOD/L-day. These results illustrate the effect of attached biomass on mixing characteristic of anaerobic filters. Comparison between the flow pattern under clean bed conditions with gas mixing and the flow pattern under actual operating conditions showed a similar dispersion. The dispersion numbers (μL/D) ranged from 0.27 to 0.36. On the other hand, the ratio of mean retention time to theoretical retention time was remarkably different. The T/T_d ratio of the actual condition was much less than that of the gas mixing clean-bed condition. These facts suggest that, under normal operating conditions, biogas production and biomass accumulation significantly affected reactor hydraulics and caused short-circuiting.

Biomass Development and Distribution. Scanning electron micrographs (SEM) were taken in order to study biofilm formation on the

Table 3. Results of tracer studies

Reactors	T_d[a] (hr)	T[b] (hr)	Gas mixing (mL/L/min)	D/μL	T/T_d	R[c] (%)
Clean bed						
With gas mixing	11.7	9.15	1.5	0.27	0.78	99
Without gas mixing	11.8	9.91	0.0	0.33	0.83	96
Dirty bed (Day 500)						
Control-lead	16.7	8.91	2.0	0.27	0.53	97
Control-follow	15.1	6.66	0.3	0.33	0.46	94
System I-lead	16.6	9.44	2.3	0.27	0.67	98
System I-follow	15.1	8.02	0.3	0.33	0.55	97
System II-lead	16.8	8.90	2.0	0.27	0.67	66
System II-follow	14.8	7.64	0.4	0.36	0.52	90

a = Residence time based on clean bed packing liquid volume
b = Mean residence time.
c = Lithium recovery ratio.

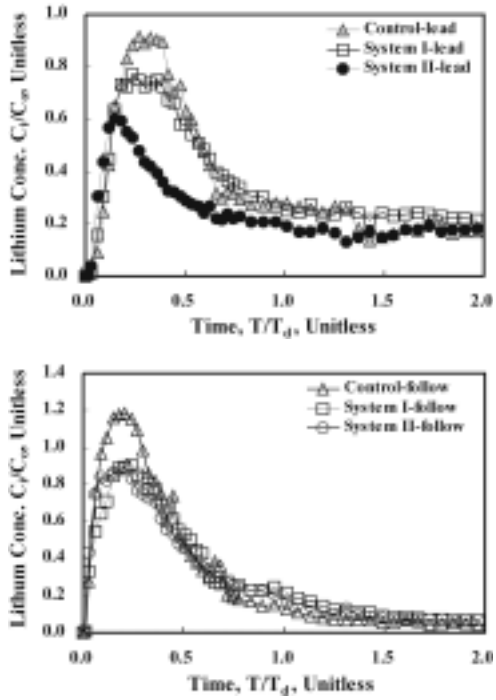


Figure 4. Effects of attached biomass on mixing characteristics in the lead and follow reactors.

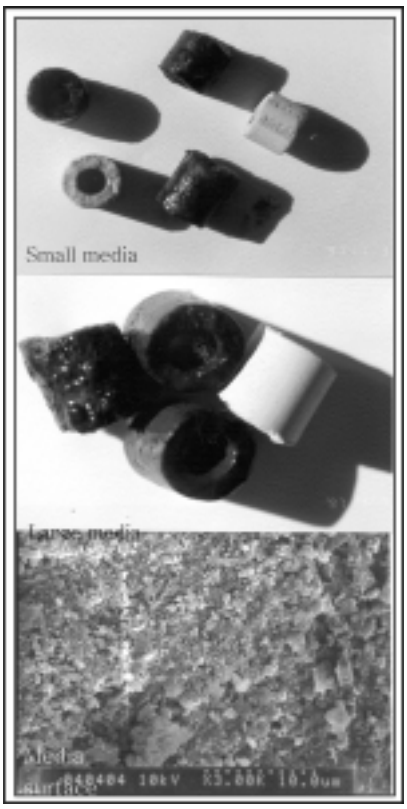


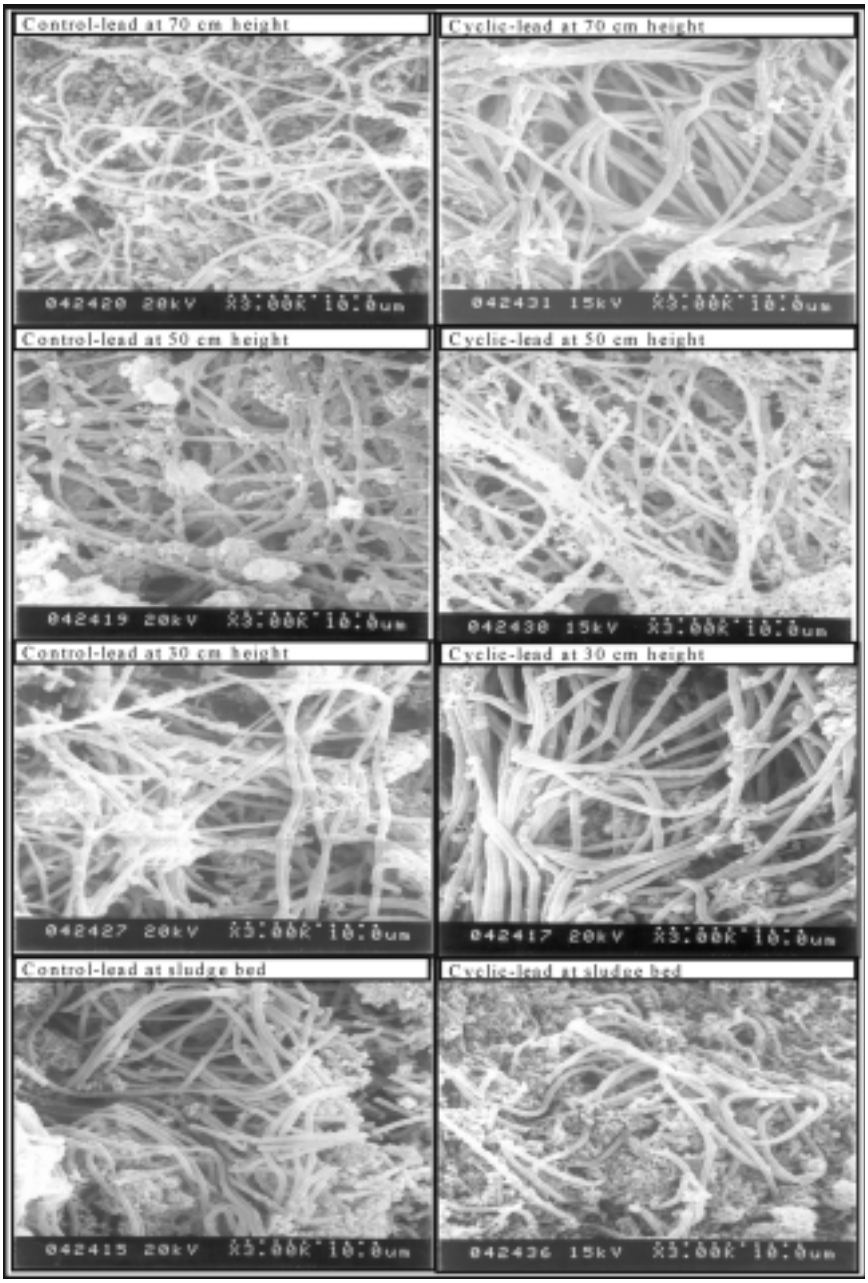
Figure 5. Photographs of media used in laboratory-scale anaerobic filters.

surface of media and to establish the biomass distribution along the reactor height. The biofilm samples were taken out through the media sampling ports of each reactor. Samples were collected from the sludge bed zone near the bottom to examine the suspended solids. Figure 5 shows the two different sized media used in this study. The large media was installed in the lower part of the column, and the small media was installed on top of the large media. The black color on the surface of the media indicates the biofilm, which was not detached by gentle shaking in liquid. As seen in this picture, the surface of the blank media was very rough. It seemed to be effective to attach the bacteria on the surface of the media during the start-up period. After a certain period of operation, sludge granules, the black or gray color, were found in the sludge bed.

Figures 6 and 7 compare bacterial forms in the biofilm in the control reactor system with those in the cyclic reactor system. Scanning electron microscopy revealed that the biofilm sampled at different reactor heights of the cyclic system was rather homogenous. The three major bacterial shapes were identified: filamentous shaped, rod-shaped, and coccus. Of these, the filamentous were predominant. The clumping tendencies and diameters (1.0 μm) of the coccus-shaped bacteria suggest that theses microorganisms were often enmeshed in flocs of filaments (possibly *Methanothrix*).

Rod-shaped bacteria had an average diameter of 0.3 to 0.5 μm and an average length of 1.0 to 4.0 μm. Judging by its size and shape, this bacterium could be related to *Methanobrevibacter, Methanobacterium,* and *Methanomicrobium.* Filamentous bacteria had a diameter of 0.3 to 0.4 μm and total lengths of 20 to 50 μm. They mainly consisted of networks of long multicelluar filamentous morphorogically identical to *Methanothrix spp.*, with diverse groups

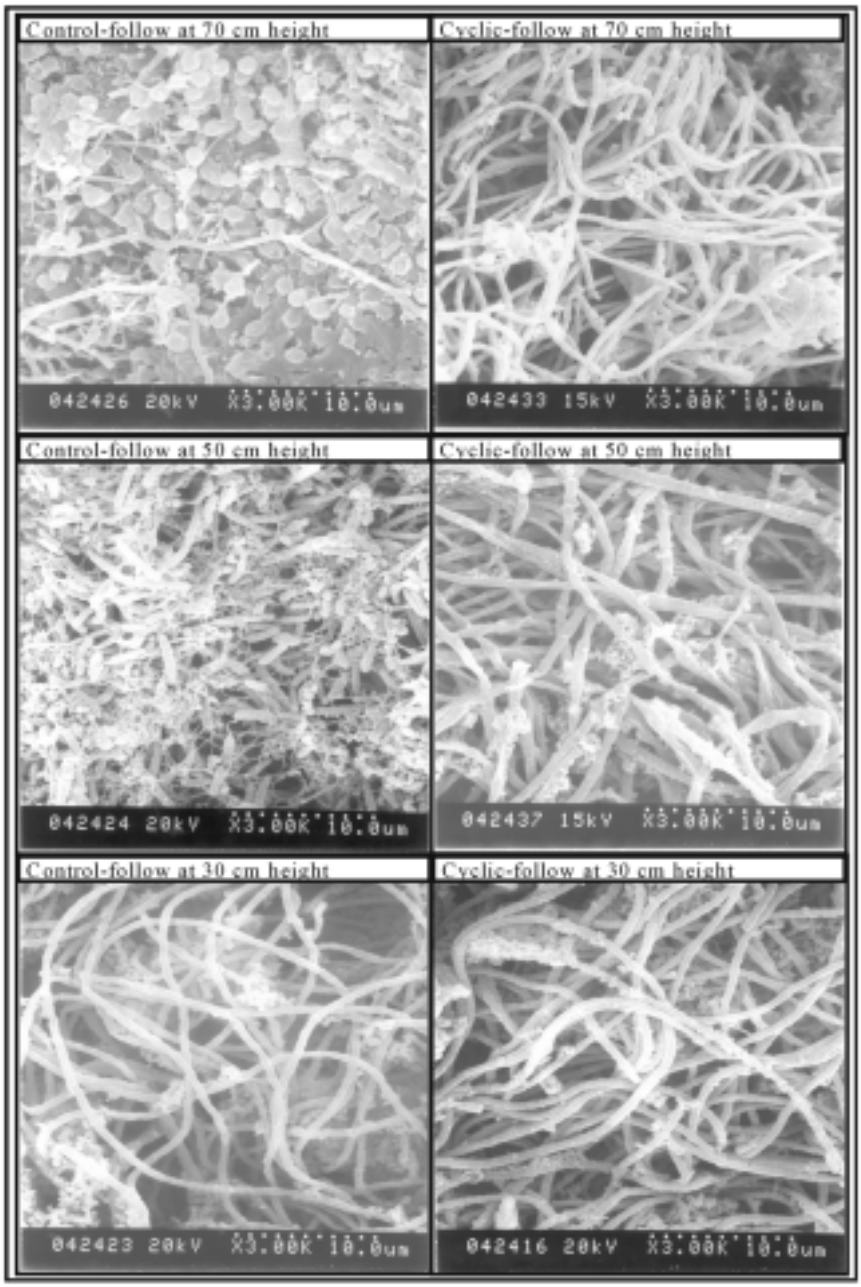
Figure 6. Scanning electron microphotographs of the biofilm sampled from the control-lead and cyclic-lead reactor at different reactor heights.



of rod and coccus shape bacteria entrapped therein.

Filamentous organisms resembling *Methanothrix soehngenii* were the most abundant although to a much lesser extent cocci and rod shaped bacteria were also present. Attached biofilm of the lead reactor of the control system consisted of bacteria that resembled in shape the *Methanothrix soehngenii* (acetate-fermenting methane bacterium), while the follow reactor con-

Figure 7. Scanning electron microphotographs of the biofilm sampled from the control-follow and cyclic-follow reactor at different reactor heights.



sisted of various bacterial shapes at above 50 cm height of the reactor. In the low zone, most of the bacteria were a filamentous shape while in the middle zone shorter species of bacteria were found. In the highest zone of the follow reactor of the control system bacteria resembling *Methanosacina appeared* were predominant.

In the follow reactor of the control system, the species of bacteria were distributed differently along the reactor height. Since different

environmental conditions exist at different heights in a reactor may favor the development of a particular group of microorganisms and thereby increase the rate of consumption of a particular component of a waste. The follow reactor of the control system showed an example.

Table 4 shows average biomass concentrations at three different positions along the reactor height. These samples were examined after operation for 420 days and 480 days. As expected, both the attached and the suspended biomass in the lead reactor were much more than that of following reactor. In the cyclic system, relatively more biomass existed in the follow reactor because it received a high influent COD concentration when the flow was reversed. The difference in attached biomass thickness depended on position, substrate concentration, hydraulic retention time, and operational mode. The attached biomass at the lowest zone of the reactor was the thickest. In the control system's lead reactor, the attached biomass was thicker than that in the cyclic operating system. In the control system, the lead reactor and follow reactor showed markedly different amounts of attached biomass, while no such difference was noted in the cyclic system. These results imply that the cyclic mode is effective in maintaining evenly distributed biofilm on the surface of the media.

The ratio of attached versus suspended biomass in the follow reactor was much higher than that in the lead reactor. The cyclic operation seemed to improve the distribution of biomass in both lead and follow reactors. Scanning electron microphotographs of samples taken from different position along the reactor height proved that the ceramic media provided a good surface for biofilm attachment.

Response to Flow Reversal. System II was operated for 34 days after the first reversal of the flow direction at a system loading rate of 3.53 g SCOD/L-day and HRT of 1.47 days. Figure 8 illustrates the behavior of the effluent soluble COD concentration immediately after the flow reversal. As shown in this figure, the follow reactor's effluent soluble COD was higher than that of the lead reactor for about five hours from the point of the flow reversal. Figure 9 demonstrates the fact that the soluble COD removal efficiency of the follow reactor was recovered up to the level of 99 % within 8hours after flow reversal.

System I was operated for 26 days after the flow reversal at a system loading rate of 3.89 g SCOD/L-day and HRT of 1.47 days. The soluble COD removal behavior was similar to the System II; the follow reactor's effluent soluble COD concentration was higher than that of the lead reactor during the transition period as shown in Figure 10. It took about eight hours until the follow reactor's soluble COD concentration became lower than that of the lead reactor. The higher effluent COD concentration of the follow reactor lasted until the former lead reactor's highly concentrated liquid was washed out after the flow reversal.

This problematic phenomenon was supposed

Table 4. Comparison of biomass at pseudo steady-state during two-stage unidirectional and cyclic operation. (Average value)

Parameter	Unidirectional operation		Cyclic operation	
	Lead reactor	Follow reactor	Lead reactor	Follow reactor
Total Biomass (gVSS/reactor)	96.21	27.65	61.85	38.89
Attached Biomass (gVSS/reactor)	37.28	19.04	26.8	25.26
Suspended Biomass (gVSS/reactor)	58.93	8.61	35.05	13.63
Height of sludge bed (cm)	18.9	7.7	19.8	13.9
Ratio of Attached to Suspended biomass	0.63	2.21	0.76	1.85
System Biomass (gVSS/reactor)	123.86		101.30	
Ratio of A/S^{a}_{lead} to A/S_{Follow}	0.29		0.41	

[a] A/S = Attached biomass/Suspended biomass

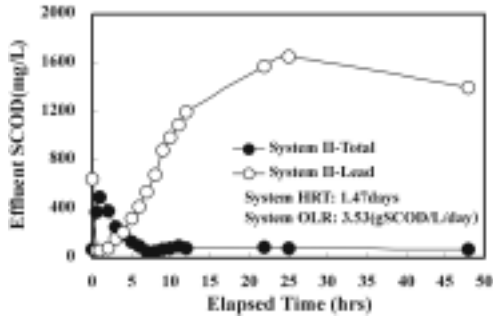


Figure 8. Behavior of the effluent COD concentration during transition period immediately after flow reversal at a system loading rate of 3.53 g SCOD/L-day and HRT of 1.47 days in system II.

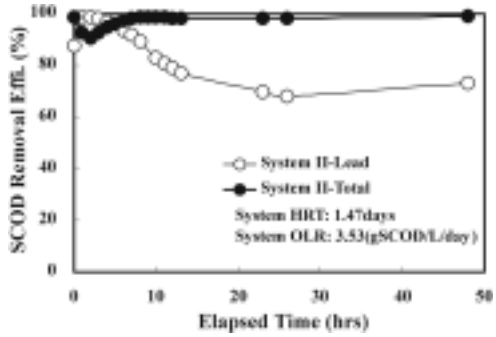


Figure 9. COD removal efficiency during transition period immediately after flow reversal at a system loading rate of 3.53 g SCOD/L-day and HRT of 1.47 days in System II.

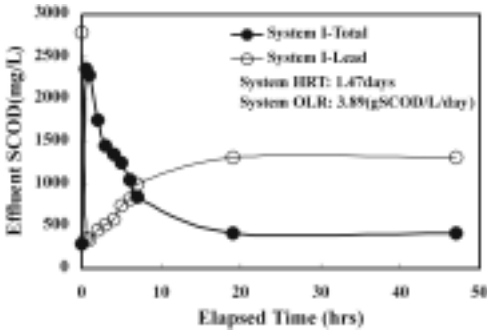


Figure 10. Behavior of the effluent COD concentration during transition period immediately after flow reversal at a system loading rate of 3.89 g SCOD/L-day and HRT of 1.47 days in System I.

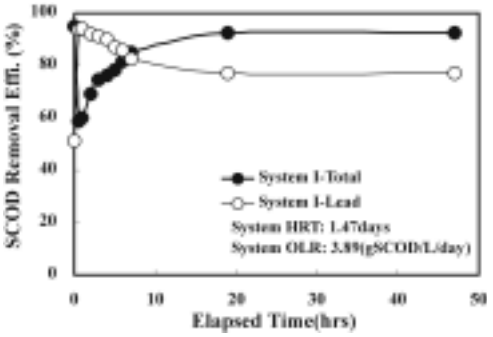


Figure 11. COD removal efficiency during transition period immediately after flow reversal at a system loading rate of 3.89 g SCOD/L-day and HRT of 1.47 days in System I.

to deteriorate the quality of effluent during the transition period in the two-stage cyclic mode operation. Even though recovering time of the effluent quality did not last over eight hours, the develop of a proper operational solution was required in order not to discharge of the lower quality of effluent. Therefore, this study suggests that the problem might be solved by recirculation of the follow reactor effluent into the lead reactor, or discharge of from the lead reactor instead of the following reactor for the transition period.

Process Control Index. Figure 12 shows the COD removal efficiencies for both the lead reactor and the total system during three cycles. The lead reactor's COD removal efficiency became stable after the gas production ratio of the follow reactor reached about 10 %. Thereafter, the follow reactor did not contribute much to total COD removal; providing longer cycle time, or operation to the low G_f/G_t ratios does not seem necessary.

If the cycle time is too short, the performance of the two-stage system will resemble that of a completely mixed reactor. On the other hand, if the cycle time is too long, the two-stage system will lose the advantages of the cyclic operation of the two-stage system. Therefore, the author assumed that the two-stage cyclic process had an optimum cycle time. Besides the cyclic time, due to the fact that the performance of the two-stage cyclic system is related to many other factors, such as hydraulic retention time, sludge retention time, organic loading rate, and biomass

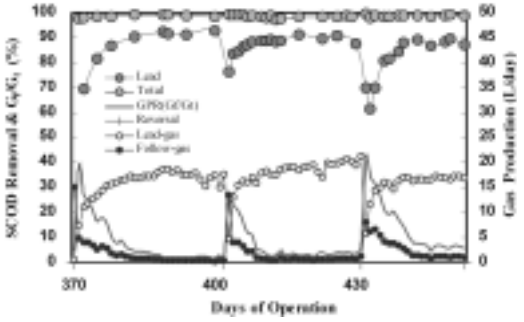


Figure 12. COD removal efficiencies for both lead reactor and total system during three cycles.

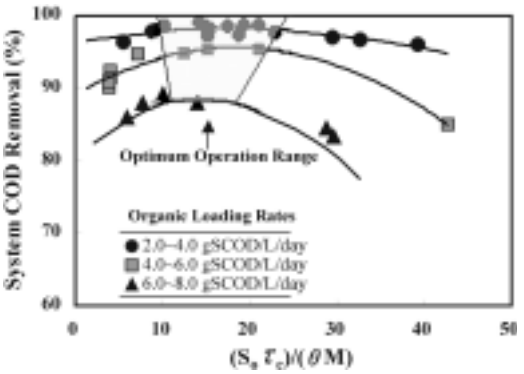


Figure 13. Optimum operating range of the two-stage cyclic operation at a number of organic loadings. Shaded section represents optimum ranges.

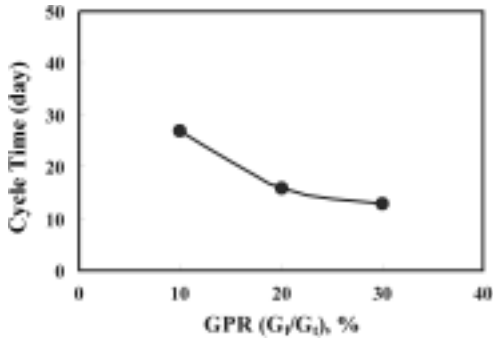


Figure 14. Relationship between cycle times and gas production ratios.

concentration, in order to control the system an integral process control index that includes above factors is required. Consequently, the range of the optimum cycle time for the two-stage cyclic operational system was shown when the system COD removal percent versus the operational index, $S_o\tau_c/\theta M$, was plotted in Figure 13. Where, S_o is COD, τ_c is the cyclic time, θ is the hydraulic retention time, and M is the biomass concentration. According to the relationship, the optimum range became narrow with increasing organic loading rates. Optimum ranges were 12-16, 14-24, and 14-22 for conditions receiving, 2.0-4.0, 4.0-6.0, and 6.0-8.0 g SCOD/L-day, respectively. Furthermore, it was also possible to use the gas production ratio of the lead and follow reactor as a cycle time determination parameter. Figure 14 shows an example of using gas production rates as a way of cycle time determination. At an organic loading rate of 3.0 COD/L-day, optimum cycle time was between 10 and 67 days based on the process control index, α. This range corresponds approximately to the point at which the gas production in the follow reactor is 10% of the total gas production. Therefore, the ratio of follow to total gas production, G_f/G_t, seems be used easily to estimate cycle time.

CONCLUSIONS

The following conclusions were drawn from this study.

- The performance of the two-stage cyclic anaerobic filter system was closely related to the cycle time. The optimum cyclic time could be determined by using an operational index [$S_o\tau_c/\theta M$], defined in this study. This optimum cycle time also corresponded to the time when the gas production in the second (follow) stage became less than 10 % of the total gas production.
- The hydraulic pattern of the upflow hybrid anaerobic filter under normal operating conditions was similar to that of a plug flow reactor with strong dispersion. Compared with the clean bed condition with gas mixing, the dirty beds displayed more short-circuiting. These results suggest that, under normal operating conditions, biogas mixing and biomass accumulation had a significant effect on reactor hydraulics, thereby causing short-

circuiting.

- The ceramic media provided a good surface for biofilm development: however, more research is needed to solve the clogging problem after the attachment of biofilm.

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