Pellet Mass

The characteristics of the pellets formed in the pelletiser under different operating conditions are shown in Table 6 and Figure 5. More pellet mass (PM) was detected in the low NTU Cases A1 and A2 than that in the high NTU Scenarios B1 and B2 ; this is due to the different startups which in turn resulted in different initial pellet mass, i.e., about 188 to 234 and 53 to 75 g for the low and high NTU Cases A and B, respectively. In practice where the process will be run continuously for a long period, the pellet mass for the case of the same influent turbidity, should however eventually become similar.

Table 6a. Characteristics of pellets and Pellet Mass Loading (PML) and Pellet Retention Time (PRT) at the 84th hour of running for Case A (summer, 30-60 NTU)

Cases	Pellet Mass (g)	PML				PRT	Pellet*		
		1	2	3	4	(d)	d (mm)	v_s (m/h)	? ? g/cm^3?
A1) at 9.6 m/h upflow velocity Alum+0.3 mg/l nonionic polymer									
3	220	0.14	0.13	0.2	0.17	6.31	0.21	9.63	1.1
5	226	0.13	0.13	0.18	0.17	11.29	0.21	9.63	1.1
7	234	0.14	0.14	0.19	0.18	7.26	0.22	9.63	1.09
10	237	0.14	0.14	0.17	0.16	11.9	0.22	9.66	1.09
at 15 m/h upflow velocity									
3	238	0.19	0.19	0.19	0.17	4.74	0.2	15	1.17
5	250	0.18	0.18	0.15	0.13	4.36	0.21	15.16	1.16
7	234	0.2	0.2	0.16	0.14	6.22	0.21	15.06	1.16
10	269	0.21	0.21	0.15	0.14	5.56	0.22	15.12	1.14
A2) at 9.6 m/h upflow velocity Polymer (mg/l)									
0.1	233	0.13	0.11	0.19	0.15	7.13	0.18	9.54	1.14
0.2	245	0.12	0.11	0.19	0.15	11.87	0.19	9.6	1.12
0.3	272	0.1	0.1	0.17	0.14	9.02	0.19	9.63	1.12
at 15 m/h upflow velocity									
0.1	226	0.19	0.17	0.18	0.15	5.88	0.18	15	1.21
0.2	212	0.19	0.18	0.17	0.13	5.25	0.18	15	1.21
0.3	265	0.18	0.16	0.15	0.13	4.78	0.19	15	1.19

* taken from the top layer (or 1300 mm height) of the pellet blanket

Note : PML (1) = $\frac{\text{g SS applied to the system/day}}{\text{g pellet mass in reactor}}$, or day^{-1}

(2) = $\frac{\text{g SS removed by the system/day}}{\text{g pellet mass in reactor}}$, or day^{-1}

(3) = ave. NTU fed to the system/ g pellet mass in reactor

(4) = ave. NTU removed by the system/ g pellet mass in reactor

PRT = pellet retention time = $\frac{\text{PM in reactor (g)}}{\text{PM discharged (g/d)**}}$, or day

** including those sampled for lab analysis

d = pellet diameter, mm

v_s = settling velocity, m/h

ρ = pellet density, g/cc?

Table 6b. Characteristics of Pellets and Pellet Mass Loading and Pellet Retention Time at the 72th hour of running for Case B (rainy season, 100-200 NTU)

Cases	Pellet Mass (g)	PML				PRT	Pellet		
		1	2	3	4	(d)	d (mm)	v_s (m/h)	??????? ? g/cm^3?
B1) at 6 m/h upflow velocity Alum+0.3 mg/l nonionic polymer									
18.7	106	0.4	0.4	0.94	0.9	3	0.22	6.21	1.06
21.1	120	0.45	0.44	1.09	1.06	3.28	0.22	6.21	1.06
23.4	103	0.45	0.45	1.02	0.98	3.74	0.23	6.24	1.05
25.8	155	0.45	0.45	1.15	1.13	2.1	0.23	6.24	1.05
at 9.6 m/h upflow velocity									
18.7	135	0.67	0.66	1.21	1.18	1.34	0.21	9.66	1.11
21.1	133	0.69	0.68	1.16	1.13	1.6	0.22	9.69	1.09
23.4	160	0.68	0.67	1.16	1.13	1.55	0.22	9.75	1.09
25.8	157	0.76	0.75	1.18	1.16	1.49	0.23	9.75	1.08
B2) at 6 m/h upflow velocity Polymer(mg/l)									
0.1	96	0.6	0.56	1.87	1.76	2.93	0.2	6.18	1.07
0.2	95	0.6	0.56	2.02	1.94	3.01	0.2	6.24	1.07
0.3	87	0.52	0.46	1.65	1.57	3.13	0.21	6.24	1.06
At 9.6 m/h upflow velocity									
0.1	125	0.7	0.59	1.44	1.29	2.04	0.19	9.6	1.12
0.2	132	0.72	0.62	1.46	1.4	2.04	0.2	9.63	1.11
0.3	102	0.67	0.58	1.4	1.32	2.28	0.21	9.66	1.1

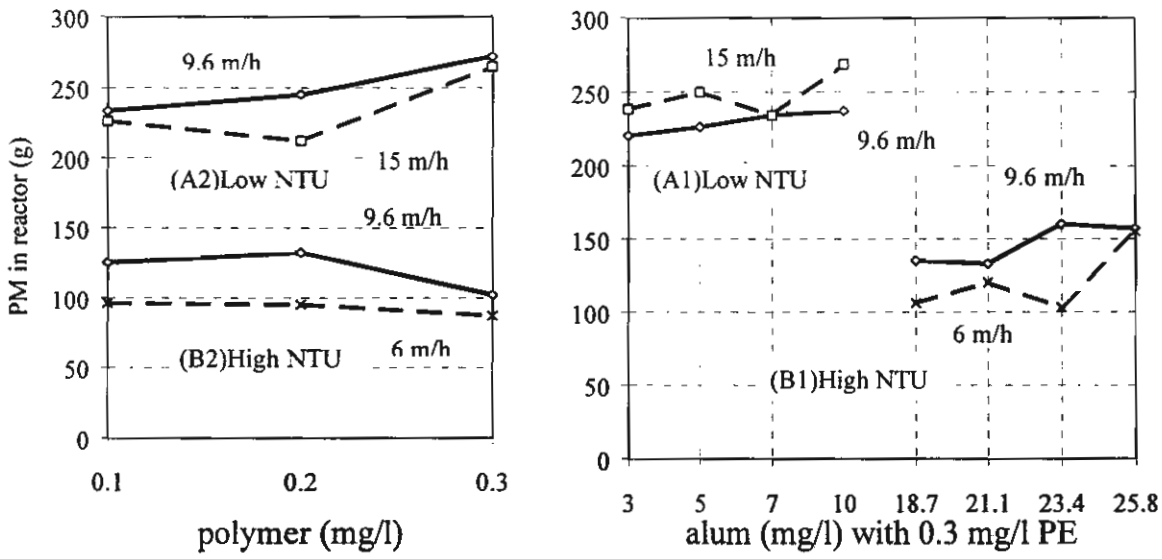

Figure 5 Pellet mass in the reactor at the steady state for different operating conditions.

The Pellet Mass Loading (PML) expressed in different units (see Table 6) was suggested as the criterion for optimum operating of the process. In the first two scenarios, the suspended solids applied or removed in g/d were compared with the pellet mass (PM, g) in the system. For the

application of the first unit, if the incoming SS expressed as g/d is known, the PM required to exist in the reactor can then be computed and targeted for. However, if only the suspended solids to be removed by the system are to be taken into account, the second expression can be used instead. For the last 2 units, the application is similar; the only difference is to use the NTU value in place of the SS number. For Case B or in the rainy season, the PM loading using alum and polymer as the coagulant and coagulant aid, respectively, was shown to be in the range of 0.4 to 0.76 day^{-1} whereas that for Case A (30-60 NTU) was 0.13 to 0.21 day^{-1}. The corresponding values expressed as NTU/g PM were 0.90 to 1.21 and 0.16 to 0.20, respectively. It is noted that the PM loading expressed by NTU under the 100-200 NTU scenario of Case B1 was about 2 times that with the SS expression. The numbers for these two criteria at the low turbidity of 30-60 NTU were nevertheless quite similar. This finding may be attributed to the higher NTU to SS ratio of the raw water at the low SS condition. A similar conclusion may be drawn for the scenario using the polymer as the sole coagulant, i.e., with 0.1 to 0.19 and 0.46 to 0.72 day^{-1}, and 0.13 to 0.19 and 1.3 to 2.0 NTU/g PM for Cases A2 and B2, respectively. Again, it should be borne in mind that with the application of the polymer of the said doses as the coagulant, the effluent turbidity was not very good and could merely go down to about 8 to 10 NTU. A conclusion on the optimum PML could not be fully drawn from this preliminary study. However, if a lower initial PM was targeted for lower operating costs, it can be concluded that as high as 0.76 day^{-1} of PML was possible for a good effluent quality of lower than 5 NTU. The PML as design criteria should be, on the other hand, developed from a long-term study with a larger pilot plant and with real water of different situations.

Another criterion suggested to be used as an indicator of the performance of the system was the Pellet Retention Time or PRT (in days) which can be calculated as shown in Table 6. When the alum was used as the coagulant, the PRT was found to be in the range of 1.3 to 12 days, depending on the initial PM after the start-up process. The corresponding numbers for the cases using the polymer as a coagulant were 3 to 12 days. It is noted that this was possibly the first work done on the pelletiser for turbidity removal with actual low-NTU real river water and it is also the first time these two criteria, i.e., PML and PRT, were suggested as process performance indices. Precise values for the said paramaters cannot readily be suggested now and more work on the fine tuning is recommended.

It is apparent from Table 6, which shows the characteristics of the pellets taken from the top level of the pellet blanket, that the diameters of the pellets formed in the cases using alum as a coagulant were only about 0.20 to 0.22 mm, probably due to the low turbidity in the feed water when compared with work performed by others. However, with the selection mechanisms due to various upflow velocities, the pellet density (calculated from Stokes' law) increased from 1.05 -1.06 to 1.08 - 1.11 and 1.14 - 1.17 for the 6, 9.6 and 15 m/h scenarios, respectively. For the polymer-as-coagulant conditions, the related numbers were 0.18 to 0.21 mm in diameter, and 1.06 - 1.07, 1.10 - 1.14 and 1.19 - 1.21 for those 3 upflow velocities, respectively. The pellet diameters reported here are much lower than those found by Tambo and Watanabe [17] (0.2 to 4 mm), Tambo and Matsui [2] (0.9 to 2 mm), and Tambo and Matsui [3] (0.7 to 2 mm), possibly because of a) the low NTU used in this study, b) real water used in this study, with different characteristics from the synthetic water used by others, and c) different heights from which pellets were taken for analysis (i.e., only from the top or a critical layer in this study versus all levels in other investigations). The density of the pellets formed in this experiment was in the range reported by others, i.e., 1.06 to 1.12 g/cm^3 versus approximately 1.002 to 1.10 [17] and 1.08 to 1.25 [2,3].

CONCLUSIONS

A pelletiser process was successfully and satisfactorily used for turbidity removal from a low NTU feed water taken from a local river, the Chao Phraya. The key was the start-up process which was done with 3000 mg/l kaolin and a dose of alum and a polymer. Even with low turbidity in the influent, the pellet mass in the reactor increased with time until a steady state after about 50 to 70 hours. With a low initial pellet mass (PM), 'highly' turbid water in the rainy season of 100-200 NTU and an upflow velocity of 9.6 m/h, the process proved capable of producing an effluent with quality better than the 5 NTU, requirement of the Bangkok Metro Waterworks Authority. The corresponding Al/T and P/T ratios were 0.004 to 0.022 and 0.0030 to 0.0054 g/g SS, respectively. However, when only a nonionic polymer was used as the coagulant, with the P/T ratio of 0.0006 to 0.0053 g/g SS, the product water was not of high quality and the NTU ranged from 7 to 18. Thus a higher chemical dose needs to be used if a better effluent quality is required. With the lower alum dose used in this study than in regular water plants, the product water had a low concentration of aluminium, i.e., 75 to 186 ?g/l, which was lower than the 200 ?g/l standard established by the WHO. Two new parameters, i.e., PML (pellet mass loading) and PRT (pellet retention time or pellet age, θ_c) were suggested as the criteria to determine the process performance. A PML as high as 0.76 day^{-1} and a PRT as low as 1.3 day were shown to be effective for turbidity removal for this low-NTU feed river water. More work is needed to determine the optimum value of these suggested parameters.

ACKNOWLEDGEMENT

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Decolourization of Azo-reactive Dye by Polyphosphate and Glycogen Accumulating Organisms in an Anaerobic-Aerobic SBR Process

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ABSTRACT

Reactive azo dyes can be effectively decolourized by anaerobic system. Most current conventional treatment technology for textile wastewater is, however, aerobic process. Simple modification of this aerobic system to anaerobic + aerobic configuration may be an answer for simultaneous organics and color removal. An anaerobic-aerobic sequencing batch reactor with a sludge age of 8 days and anaerobic+aerobic+settling times of 18+5+1 hr, respectively, was used to decolourize an azo-reactive dye wastewater. The nutrient broth and sodium acetate (NB+SA) solution at 500+0, 350+150, 250+250 and 0+500 mg/l as COD was fed to the system to promote the polyphosphate accumulating organisms (PAOs), while only glucose (500 mg/l COD) was used as a glycogen accumulating organisms (GAOs) promoting substrate. The decolourization capability of the process was about 73 to 77 and 59 to 64 percent in terms of ADMI for the two cases in which the PAOs and GAOs proliferated, respectively. For practical purpose, the tank configuration with anaerobic+aerobic phases was the only requirement for good decolourization, and a high P removal efficiency of the system was not necessarily the ultimate goal. The experiments using nutrient broth and sodium acetate as the substrates were a little better in decolourization than those using glucose. A longer anaerobic time (18 vs. 6 hrs) was also slightly superior for the colour reduction.

Key words: textile wastewater, azo dye, reactive dye, color removal, anaerobic-aerobic process, sequencing batch reactor, polyphosphate accumulating organisms, glycogen accumulating organisms

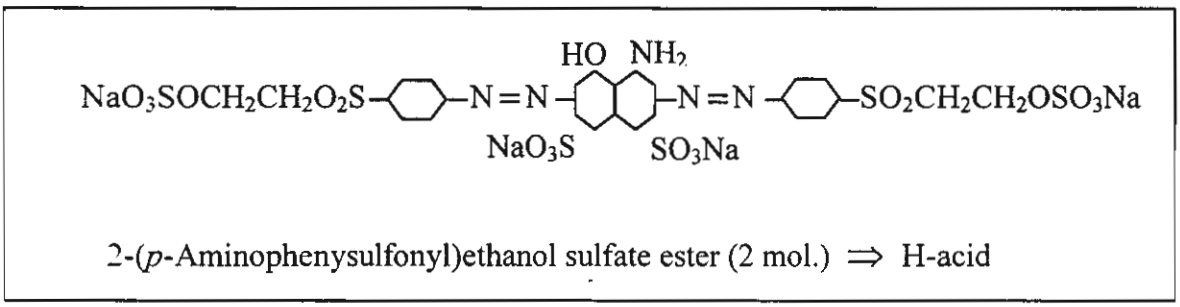
INTRODUCTION

Brown and Laboureur (1983) as well as Brown and Hamburger (1987) demonstrated the possibility of decolourization of azo-reactive dyes by an anaerobic process from which aromatic amines were produced as intermediates. Meyer (1981) stated that, in the process, the azo bond in reactive dye could act as electron acceptor, resulting in the cleavage of the bond and subsequent decolourization. However, these carcinogenic intermediates could be further degraded in any aerobic system (Brown and Hamburger, 1987; Zaoyan *et al.*, 1992). Meanwhile, most current treatment technology for textile wastewater is aerobic process, i.e., activated sludge system. A simple modification to this system can result in a process with anaerobic and aerobic configuration, which can then remove both organic and colour simultaneously. This process, if properly operated, can also be an enhanced biological phosphorus removal (EBPR) system in which polyphosphate accumulating organisms (PAO) proliferate. Pansuwan and Panswad (1997) and Pansuwan *et al.* (1999) concluded from a study on dye decolourization by an A/O or Phoredox process that external carbon sources such as sucrose or acetic acid could be used for the co-metabolism, resulting in a better colour reduction efficiency. This is in agreement with other investigators (Atlas, 1993; Chinwetkitvanich *et al.*, 1999). However, if glucose is used as a co-substrate glycogen accumulating organisms (GAOs) can outgrow the PAOs in the system and the phosphorus

removal efficiency possibly deteriorates, especially when phosphate is limiting (Cech and Hartman, 1990). The decolourization of the system with the anaerobic-aerobic tank configuration, with PAOs or GAOs as relatively dominating microorganisms (besides ordinary heterotrophic organisms, OHOs), was then evaluated in this investigation. The effect of the anaerobic hydraulic retention time (HRT) was also preliminarily studied.

MATERIALS AND METHODS

A 12 litre pilot anaerobic-aerobic (A-A) sequencing batch reactor (SBR) with the ratio of feed volume (V_f) to remaining volume (V_o) and the overall sludge retention time (SRT, θ_c) of 2 : 1 and 8 days, respectively, was used at room temperature of 28 to 31°C to treat a synthetic wastewater containing 10 mg/l Remazol Black B reactive dye, which has the following molecular structure:



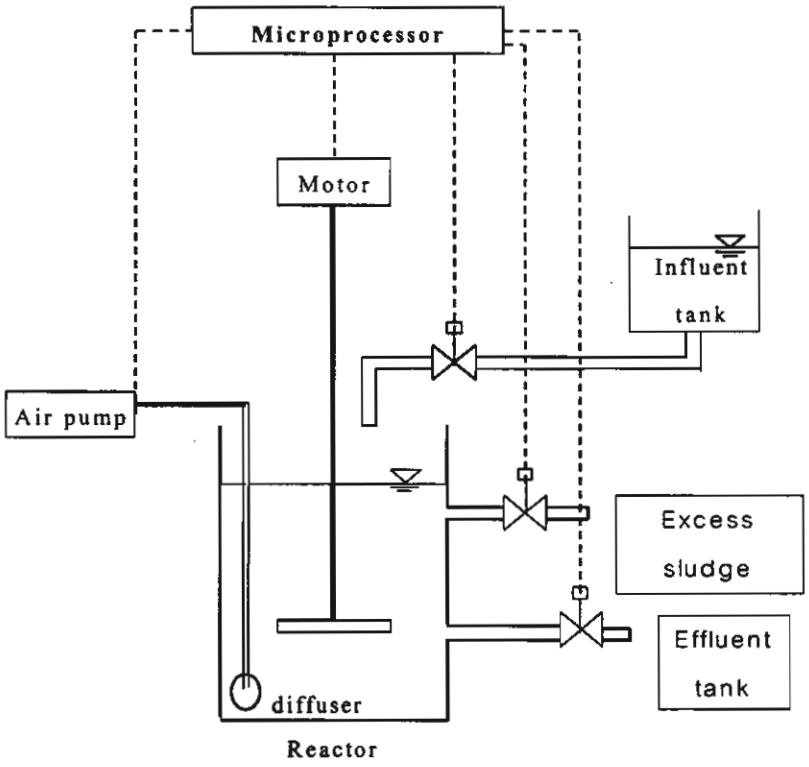


Figure 1 Experimental setup for this study.

Figure 1 illustrates the experimental setup, while Table 1 shows the operating SBR cycle. The anaerobic and aerobic θ_c was 6.3 and 1.7 days, respectively. The system was first inoculated with activated sludge taken from a local sewage treatment plant (STP) to have initial mixed liquor suspended solids (MLSS) of 2500 mg/l in all experiments. Table 2 shows the composition of the synthetic wastewater, while Table 3 tabulates various experimental scenarios. In the first four experiments with COD : N : P ratio of 500 : 50 : 15 and different concentrations of nutrient broth and sodium acetate (NB+SA), the latter of which was an EBPR promoting substrate, the PAOs in addition to OHOs were set to be the predominant species, whereas in the experiments using solely glucose at 500 mg/l COD (G1), the PAO growth was to be suppressed and GAOs could then outgrow PAOs. In the experiment G2, the nutrients were limitingly added (COD : N : P = 500 : 25 : 5) to ascertain that PAOs could definitely not proliferate in the reactor. The impact of a shorter anaerobic HRT was also investigated in experiment G3. The system was run batchwise in an anaerobic-aerobic mode for approximately 7 months to ensure the EBPR characteristics before the dye was added. In the glucose-only scenarioes (G1 to G3), however, this EBPR step was not done. Samples were taken at different points and some were filtered through GF/C papers before chemical analysis, following Standard Methods (APHA, 1995). On the other hand, certain samples were filtered through 0.45 μm membrane filter papers for the soluble phosphate and 'true' colour determination. The colour concentration was expressed as space units, SU (Gregor, 1992) and ADMI (Allen, 1973; APHA, 1995). The polyhydroxyalkanoate (PHA) was determined according to Lee (1995) by a Shimadzu 7AG gas chromatograph.

Table 1 Operating SBR cycle.

Operation	Duration
Mixed fill stage	< 5 minutes
Anaerobic stage	18 hours (6 hours)
Aerobic stage	5 hours
Excess sludge withdrawal*	< 5 minutes
Settling	55 minutes
Effluent drain	5 minutes
Total cycle time	24 hours (12 hours)

* at the end of the aerobic stage ( ) = for experiment G3

Table 2 Composition of the synthetic wastewater.

Constituents	Concentration (mg/l)
Nutrient Broth*	500 ; 350 ; 250 (as COD)
$CH_3COONa \bullet 3H_2O$*	150 ; 250 ; 500 (as COD)
Glucose*	500 (as COD)
$(NH_2)_2CO$	50 ; 25 (as N)
KH_2PO_4	15 ; 5 (as P)
$NaHCO_3$	500 (as $CaCO_3$)
$FeCl_3 \bullet 6H_2O$	2.5 (as Fe)
$MgSO_4 \cdot 7H_2O$	3.75 ; 1.25 (as Mg)
$CaCl_2$	7.5 ; 2.5 (as Ca)
Remazol Black B Dye	10

* total COD in feed = 500 mg/l

Table 3 Various experimental scenarios.

Substrates (500 mg/l as COD)	N (mg/l)	P (mg/l)	Anaerobic + aerobic + settling time (hrs)	Cycle time (hrs)	Remarks
NB+SA = 500+0	*				NB only
NB+SA = 350+150					-
NB+SA = 250+250	50	15	18+5+1	24	-
NB+SA = 0+500					SA only
Glucose = 500 (G1)					G only
Glucose = 500 (G2)	**25**	**5**			G only
Glucose = 500 (G3)	50	15	**6+5+1**	**12**	G only

NB = nutrient broth SA = sodium acetate G = glucose

* Nitrogen was 70 mg/l, which was more than the required 50 mg/l, because of high N content in the NB.

RESULTS AND DISCUSSION

Table 4 summarizes the 5-day average results of the study at the steady state. The pH was raised from 7.4 - 7.5 in the influent to 7.7 – 8.5 in the anaerobic stage of the NB+SA cases, except for the 500+0 (or only NB) scenario. This is due to the uptake of volatile fatty acids (VFA) available from the sodium acetate by the microorganisms, leaving alkaline sodium in the bulk liquid. On the other hand, in other experiments with glucose, the pH dropped a little to 7.2 - 7.4 because of fermentation in the anaerobic phase, and then raised again to 8.0 - 8.5 in the aerobic stage. Both the anaerobic and aerobic ORP of -220 to -300 and 33 to 139 mV, respectively, as well as the aerobic DO of 5.3 to 6.1 mg/l, illustrated that the process was run under the anaerobic-aerobic condition as desired.

At the steady state, which normally took about 30 to 50 days, the MLSS and MLVSS (mixed liquor volatile suspended solids) for the 7 experiments (4 NB+SA cases as well as G1, G2 and G3) were 1589, 1562, 1644, 877, 1300, 1231 and 2257 mg/l, and 1364, 1326, 1432, 698, 1110, 1104 and 2005 mg/l, respectively, with the corresponding f value (or MLVSS/MLSS) of 0.80 to 0.90. There was a tendency that the mass developed from a more suitable substrate (or NB+SA) was more than that from glucose. In run G3, however, the bacterial mass was greater due to more substrate feeding into the system. In the case of 0+500 NB+SA, a lower mass and f value was apparent. It is highly probable that only one carbon source, from the sodium acetate, was not suitable for the process. A lack of other essential micronutrients available in NB is another possible explanation for this occurrence. The glycogen, which is also a vital part of the PAOs, is not effectively produced from this carbon source (Mino *et al.*, 1998), while more energy is also required to uptake the volatile fatty acids (VFA) under this high pH (8.5) operating condition (Smolders *et al.*, 1995). Less mass was therefore developed. It has also been reported that acetic acid at the high concentration of 400 mg/l could be toxic to the PAOs (Randall and Rodney, 1994).

Table 4 Average results at the steady state[n].

Parameters	Sampling points	NB+SA				Glucose		
		500+0	350+150	250+250	0+500	GI	G2	G3
pH	Influent	7.5	7.5	7.4	7.5	7.4	7.5	7.4
	Anaerobic	7.4	7.8	7.7	8.5	7.4	7.3	7.2
	Aerobic	8.0	8.4	8.4	8.5	8.4	8.4	8.2
	Effluent	8.1	8.4	8.4	8.5	8.4	8.4	8.2
ORP (mV)	Anaerobic	-241	-221	-257	-296	-306	-286	-232
	Aerobic	130	134	51	33	118	88	139
DO (mg/l)	Aerobic	5.29	5.75	5.88	6.04	6.14	6.00	6.13
Alkalinity (mg/l as $CaCO_3$)	Influent	503	604	610	762	444	446	446
	t = 0*	4.79	606	600	824	435	447	425
	Anaerobic	659	731	730	950	594	496	538
	Δ	+180	+125	+130	+126	+159	+49	+113
	Aerobic	428	607	582	946	423	451	383
	Δ	-231	-124	-148	-4	-171	-41	-155
	Effluent	431	609	580	947	416	448	382
MLVSS (mg l)	Aerobic	1364	1326	1432	698	1110	1104	2005
MLSS (mg l)	Aerobic	1589	1562	1644	877	1300	1231	2257
f(MLVSS MLSS)	Aerobic	0.86	0.85	0.87	0.80	0.85	0.90	0.89
SV_{30} (ml/l)	Aerobic	100	108	110	110	96	100	96
SVI (ml/g)	Aerobic	63	69	67	125	74	81	43
SS (mg/l)	Effluent	18	15	15	57	21	27	12
COD (mg l)	Influent	542	552	499	492	510	512	510
COD_{rem} (%)	Overall	91.7	93.3	96	95.1	96.5	95.9	97.3
TKN (mg l)	Influent	72.3***	51.8***	49.6	49.8	48.4	24.8	48
TKN_{rem} (%)	Overall	97.6	96.3	96.6	42.6	92.6	90.7	95.6
P (mg/l)	Influent	14.9	14.9	15	14.8	15.5	5.3	15.5
P_{rem} (%)**	Overall	99.3	100	78.7	45.9	67.7	98.1	38.7

Remark: NB + SA = nutrient broth + sodium acetate
n = 5 (5 days average experimental results at the steady state)
* calculation [(remaining alkalinity x 41+influent alkalinity x 81)/121]
** [(TP-SP)/TP]x100
*** higher than the required 50 mg/l, due to the available high N content in NB
Δ = increment or reduction

Polyphosphate Accumulating Organisms

Since the combination of nutrient broth and sodium acetate is a PAO promoting substrate, PAOs can proliferate well in the system. The phosphate removal efficiency for the first four experiments with nutrient broth and sodium acetate as the substrates was 99.3, 100, 78.7 and 45.9 percent, respectively (see Table 4), while the profile of soluble P can be seen in Figure 2. A good anaerobic P release and subsequent aerobic P uptake were obvious for the 500+0 and

350+150 NB+SA scenarios. The soluble P concentration went up from 10 to some 40 mg/l at the end of anaerobic stage and was reduced to almost zero after the aerobic step. The EBPR performance was worse for the 250+250 case, and practically no substantial P release or uptake by the PAOs was seen in the 0+500 condition.

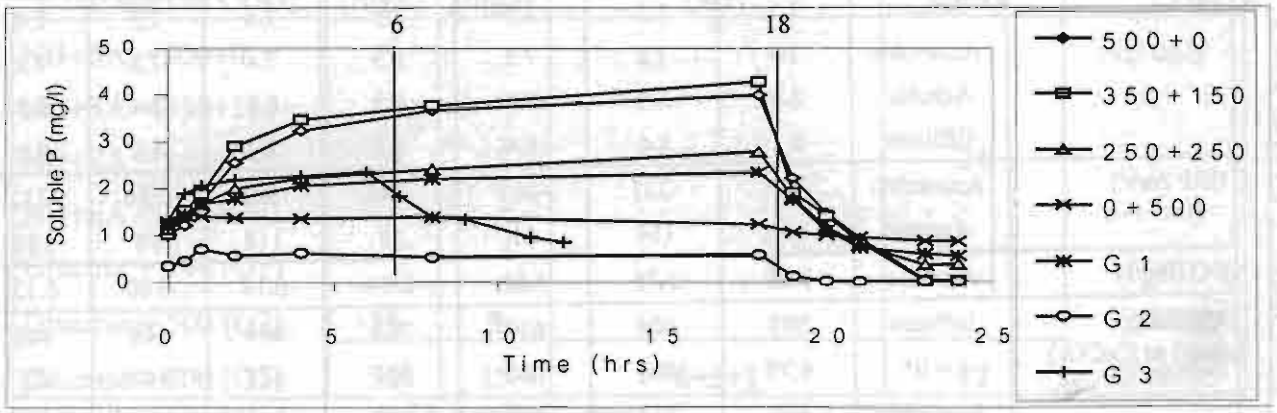
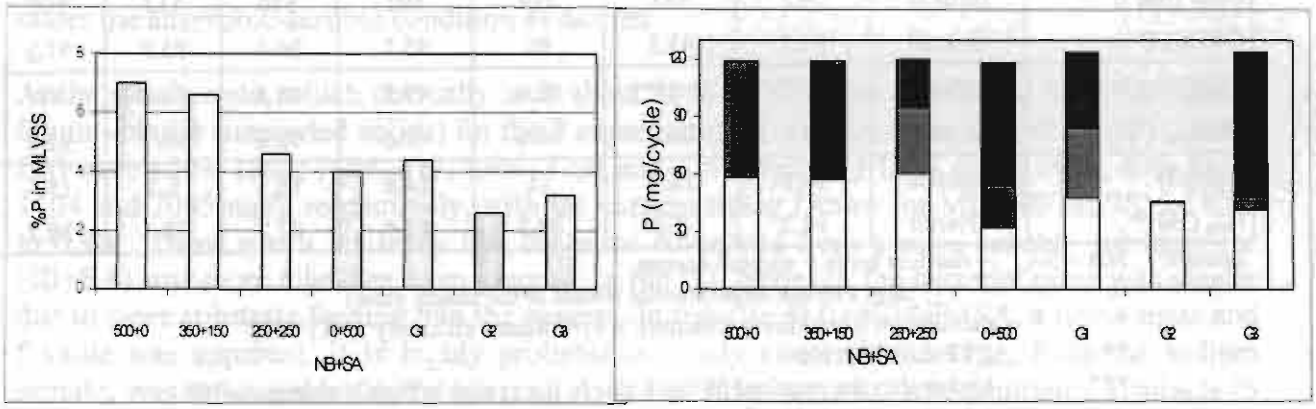


Figure 2 Soluble phosphorus profile.

At present, there is no satisfactory method to identify and quantify PAOs in enriched sludge of an EBPR process. However, the P content in MLVSS can be used as an approximate indicator of the concentration of PAOs in reactor, see Figure 3. For example, the system with NB+SA of 500+0 had P content as high as 7 percent in MLVSS, while that of 0+500 condition was only 4 percent, see Figure 3a. This implies that the first system had more PAOs in the enriched sludge than the latter one. This is confirmed by the quantity of phosphorus uptaken per cycle by microorganisms (or mg P in excess MLVSS), which was higher in the first scenario than in the latter one (118 VS 33 mg P/cycle), see Figure 3b. This explained why the effluent phosphorus of the first case was very low (0.1 mg/l) while that of the latter condition was as high as 8 mg/l P.



3a) P content in MLVSS. (%)

3b) Phosphorus balance.

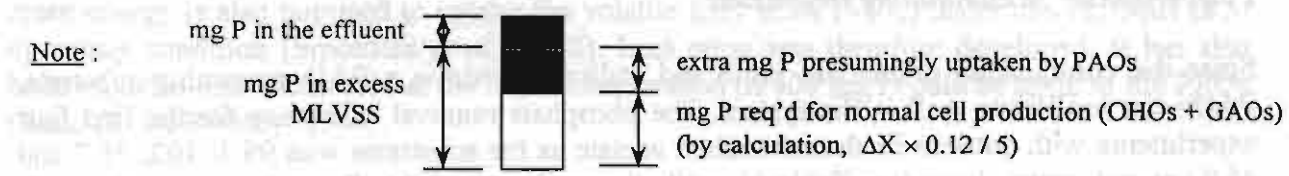
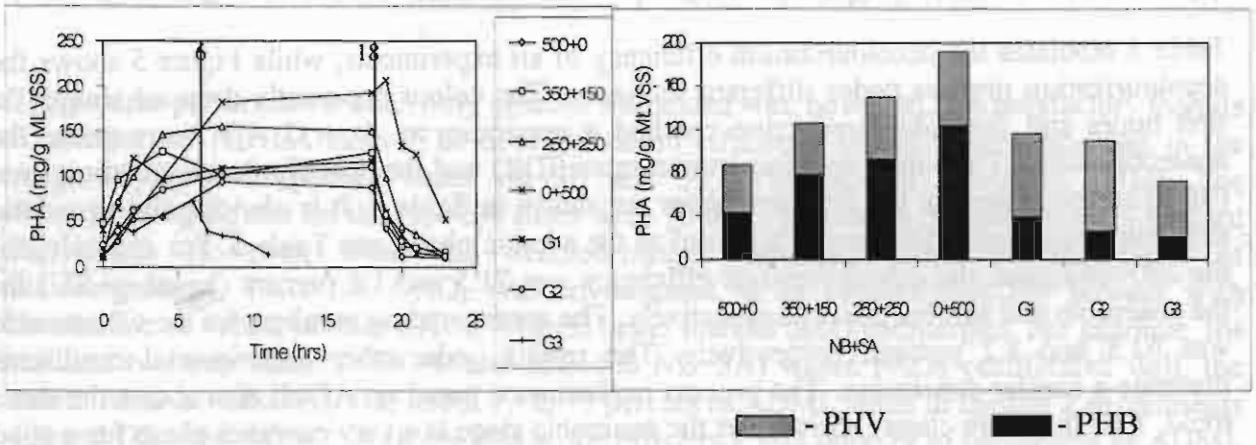


Figure 3 Phosphorus balance in the system.

In order to have a good mass balance, the phosphorus entering and leaving the process must be equal. The phosphorus leaving the system can, in turn, be separated into 2 streams, i.e., the effluent and the wasted sludge (ΔX or excess MLVSS). The P content in the wasted sludge can be further divided into 2 portions, i.e., uptaken for cell synthesis by OHOs+GAOs and PAOs. The difference between the total P content in MLVSS and the phosphorus presumingly utilized by OHOs+GAOs (roughly equal to 0.024 times mg VSS wasted per cycle) can be hence taken as an approximate indicator of the PAO portion in the mixed liquor, see the gray area in Figure 3b.

For the first 4 experiments with NB+SA, it is therefore evident from Figure 3b that more PAOs were present in the sludge when more nutrient broth (500+0 and 350+150 scenarios) was used for the preparation of synthetic wastewater, i.e., the PAO portion decreased when less nutrient broth (250+250 condition) was used, and became minimum when no nutrient broth was used at all (0+500 scenario). The relative portion of PAOs in the sludge will be used as the base for comparison of decolourization efficiency in subsequent paragraphs.

PHA as well as polyhydroxybutyrate (PHB) and polyhydroxyvalerate (PHV) can be a potential carbon source for manufacturing biodegradable plastics. As a result, although it is not related to the objective of this study, it should be noted that more PHB, PHV and PHA were produced when sodium acetate was used as substrate than when nutrient broth was, see Figure 4. In addition, the anaerobic PHA content in MLVSS increased from 85 to 190 mg/g when the NB+SA concentration changed from 500+0 to 0+500 mg/l.



4a) PHA profile. 4b) PHA at the end of the anaerobic stage.

Figure 4 PHA profile and PHA content in the enriched sludge.

Glucose Accumulating Organisms

Glucose was added into the system with the intention to suppress the growth of PAOs (Cech and Hartman, 1990). However, in run G1, the anaerobic time of 18 hours was long enough to provide good fermentation and some glucose was transformed into VFA, which was subsequently used as a promoting substrate for the PAOs. The phosphorus profile (Figure 2) and the sludge P content (Figure 3) show that the G1 system therefore acted, to a certain extent, as an EBPR process, while G2 and G3 did not. Under the G2 condition, phosphorus was added at a limiting concentration and PAOs could not effectively proliferate. In the G3

system, however, the anaerobic time of 6 hours was too short for a good fermentation, and, as a result, PAOs should also not be able to substantially grow. The phosphorus removal was 67.7, 98.1 and 38.7 percent for the last three scenarios, respectively. For the case of G2, almost complete P removal was achieved simply through a normal OHO+GAO cell synthesis (not by the PAOs), whereas the G3 anaerobic time was not sufficiently long for a good VFA conversion and a poor P removal resulted. It is known that the GAOs can also store and subsequently use up PHA in the anaerobic and aerobic phases, respectively (Cech and Hartman, 1990; Mino et.al., 1998). The profile of anaerobic PHA production and its subsequent aerobic utilization is shown in Figure 4a, while Figure 4b illustrates the overall amount of PHA at the end of the anaerobic stage. Considering Figures 2, 3 and 4, it is apparent then that besides OHOs, some of the organisms in the 250+250, 0+500 NB+SA and G1 cases and most of those in the G2 and G3 conditions were GAOs.

It is also noteworthy that PHV was synthesized relatively more than its PHB counterpart for the cases of glucose addition (see Figure 4b). This was because some glucose was converted to propionyl-CoA, and more alteration to PHV was achieved as a result (Sudiana *et al.*, 1997).

In summary, for the last 3 experiments with glucose as sole substrate, a mixture of PAOs, GAOs and OHOs was evident in the G1 case, while only GAOs and OHOs were present in the G2 and G3 scenarios. This information will also be used in the subsequent comparison of decolourization capacity of the process.

Colour Removal

Table 5 tabulates the decolourization efficiency of all experiments, while Figure 5 shows the decolourization profiles under different scenarios. The colour apparently dropped sharply in first hours and the colour reduction reached a maximum at about 2 – 3 hours under the anaerobic stage. The initial decolourization rates (IDR) and the specific decolourization rates (SDR) of the system in the first few hours are shown in Table 6. It is obvious that very little decolourization was additionally achieved at the aerobic phase, see Table 5. For example, for the 250+250 case, the colour removal efficiency was 71.5 and 1.6 percent (based on SU) for the anaerobic and aerobic phases, respectively. The corresponding numbers for the G2 scenario was 62.8 and 8.2 percent, respectively. The results under other experimental conditions illustrate a similar conclusion. The process performance based on ADMI also shows the same trend. This therefore clearly shows that the anaerobic stage is a very essential phase for a good decolourization.

Table 5 Colour removal efficiency, %.

Colour unit	Sampling Point	Nutrient Broth+Sodium Acetate				Glucose		
		500+0	350+150	250+250	0+500	G1	G2	G3
SU	Anaerobic	70.8	70.4	71.5	68.2	62.8	62.8	59.4
	Aerobic	8.0	3.6	1.6	3.5	10.6	8.2	10.2
	Overall	73.3	71.7	71.4	70.6	67.8	66.3	63.1
ADMI	Anaerobic	72.4	71.2	71.7	67.5	62.6	60.3	52.8
	Aerobic	12.8	7.0	5.0	3.5	8.8	9.6	13.4
	Overall	76.7	73.4	73.0	69.0	66.2	64.4	59.0

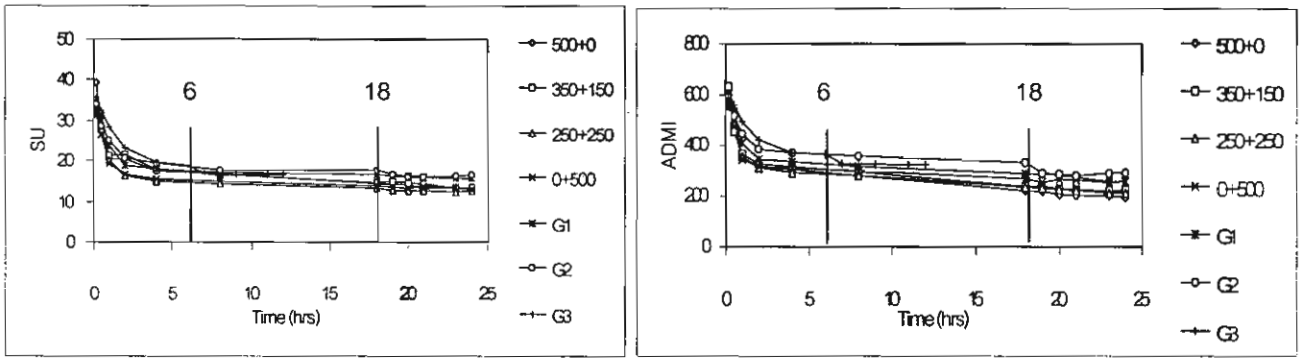


Figure 5 Colour profile.

Table 6 Initial Decolourization Rates and Specific Decolourization Rates.

Colour unit	Nutrient Broth+Sodium Acetate				Glucose		
	500+0	350+150	250+250	0+500	G1	G2	G3
Space Unit							
IDR (SU/h)	22.2	15.3	14.3	14.5	8.1	6.8	6.8
SDR (SU/g MLVSS-h)	1.6	1.1	1.0	2.0	0.7	0.6	0.3
ADMI							
IDR (ADMI/h)	315	311	247	268	124	111	99
SDR (ADMI/g MLVSS-h)	22.0	22.3	16.4	36.6	10.6	9.6	4.7

The process performance when only glucose was added was, however, less satisfactory than its counterpart of NB+SA cases. The decolourization efficiency of the G scenarios was in the range of 63 to 68 percent based on SU and 59 to 66 percent based on ADMI. The corresponding figures for the NB+SA cases were 70 to 73 percent as SU and 69 to 77 percent as ADMI. However, when the decolourization capability based on the types of microorganisms, PAO VS GAO, was investigated, it is evident that the process with domination of PAOs, besides OHOs, worked better than its GAO counterpart. For example, for the first two experiments (500+0 and 350+150 NB+SA) where PAOs proliferated well, the colour removal efficiency was about 73 to 77 percent as ADMI, while in the last 2 experiments G2 and G3, where GAOs outgrew PAOs, the efficiency was only 59 to 64 percent as ADMI, see Table 7. A drop of 13 to 14 percent was obvious in this condition. In the process where a mixture of PAOs and GAOs was apparent, the decolourization efficiency ranged from 66 to 73 percent as ADMI, depending on the portion of PAOs in the enriched sludge. As a result, in practice, the minimum requirement is that the tank configuration must be anaerobic + aerobic, and, if possible, the process should be operated as an EBPR system, by providing sufficient amount of phosphate for growth of PAOs, which can be easily achieved under the joint treatment between domestic and textile industrial wastewaters. On the other hand, when only dye house wastewater is to be treated and no external P source is available the process should be run as a simple anaerobic+aerobic system, with the GAOs and OHOs as target species. However, a somewhat lower efficiency is to be anticipated with this type of operation.

Additionally, the process with a longer anaerobic HRT of 18 hrs (G1) could achieve a somewhat better colour reduction efficiency (66 % in ADMI) than its 6-hr counterpart (G3, 59 %), possibly because of the fermentation effect and subsequent higher portion of PAOs in the

system. Moreover, in this study, it was discovered that the dominant wavelength (λ_{max}) observed during the colour measurement shifted from 595 nm in the raw wastewater to 555 nm in the treatment effluent (data not shown). This observation illustrates the change in the colour hue and it also shows that the azo bond was cleaved, resulting in the decolourization (Meyer, 1981; Luangdilok and Panswad, 1999).

Table 7 Summary of analysis.

Substrate	P removal (%)	Stored PHA * (mg/g MLVSS)	Microorganisms	Colour removal (%)	
				SU	ADMI
NB+SA					
500+0	99.3	87.2	PAOs	73.3	76.7
350+150	100	125.9	PAOs	71.7	73.4
250+250	78.7	149.8	PAOs + some GAOs	71.4	73.0
0+500	45.9	191.8	PAOs + some GAOs	70.6	69.0
Glucose					
G1	67.7	116.6	PAOs + some GAOs	67.8	66.2
G2	98.1	109.9	GAOs	66.3	64.4
G3	38.7	72.1	GAOs	63.1	59.0

* at the end of anaerobic step

CONCLUSIONS

The synthetic wastewater composed of only sodium acetate was a less promising option for good colour reduction in anaerobic-aerobic SBR process. The anaerobic-aerobic system with either GAO or PAO proliferation could remove the colour of the Remazol Black B reactive azo dye at an efficiency of about 60 to 77 percent. The decolourization capability with PAOs as relative dominating species, besides OHOs, was however, some 13 to 14 percent better than that with GAOs. For a good decolourization, an anaerobic compartment is essential and must be added ahead of the conventional aeration tank of an actived sludge system. Under this scenario, the colour reduction was mainly achieved under the first 2 hrs of the anaerobic step. A longer anaerobic HRT (18 vs. 6) could offer a little better decolourization performance for this 10 mg/l dye condition when glucose was used as the carbon source, probably due to a better fermentation in the anaerobic stage, which could transform some glucose to short chained volatile fatty acids, which in turn could be readily used by PAOs.

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ANAEROBIC DECOLORIZATION OF REACTIVE DYEBATH EFFLUENTS BY A TWO-STAGE UASB SYSTEM WITH TAPIOCA AS A CO-SUBSTRATE

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Short Title : Decolorization of Reactive Dye by 2 Stage UASB

ABSTRACT

This study was to investigate the anaerobic treatability of the reactive dyebath effluents in three different colors, namely, black, red and blue, by using a pilot scale two-phase UASB system which consisted of an acidification tank and a UASB reactor with a hydraulic retention time of about 12 hours for each phase. In the first three experiments, the feed wastewater was prepared from real wastewaters to have a constant coloration of approximately 150 SU (space units) with three different concentrations of tapioca starch as a supplemental carbon source or a co-substrate. The fourth was an experiment on synthetic blue wastewater and the fifth was to study the effect of an initial coloration of 0, 50 or 100 SU with a constant tapioca concentration of 500 mg/l. In the black-dye experiment, using the tapioca concentrations of 500, 1000 and 1500 mg/l, the decolorization efficiencies were not much different, i.e., 67, 71 and 69%, respectively. Subsequently, the concentrations of 0, 200 and 500 mg/l tapioca were used for the red-dye and the blue-dye experiments. The decolorization efficiencies of the red-dye experiment were 36, 57 and 56% and those of the blue-dye experiment were 48, 52 and 56%, respectively. In the synthetic wastewater, the decolorization efficiencies were 36, 54 and 57% for 0, 200 and 500 mg/l tapioca, respectively, and in the last experiment, also with the synthetic wastewater, the efficiencies of decolorization of 63% and 56% were found for the initial colorations of 50 and 100 SU, respectively. The supplement of tapioca starch as a co-substrate apparently gave a better color removal performance but the excessively high concentration of tapioca did not enhance the process capability in terms of color removal efficiency. Moreover, experimental results disclosed that sulfate reducing bacteria could increase the organic carbon comsumption of the system but they did not play an important role in color reduction. The acid forming bacteria could have some role in the decolorization process, and the strictly anaerobic methane producing bacteria were not the main and only microorganisms responsible for the color reduction.

Key words: anaerobic decolorization/ color removal/ reactive dye/ UASB/ sulfate reducing bacteria/ methane producing bacteria/ dye wastewater/ textile wastewater/ co-metabolism.

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INTRODUCTION

Reactive dyes are becoming more popular in the textile industry. They are colored molecules capable of forming a covalent bond between the dye molecule and the fibers, which are usually cellulosic fibers. However, reactive dyes hydrolyze easily, resulting in a high portion of unfixed (or hydrolyzed) reactive dyes which has to be washed off during the dyeing process. As much as 50% of the initial dye load is present in the dyebath effluent (Shore, 1995).

Color is the first contaminant to be recognized in textile wastewaters and has to be removed before discharging into the receiving water body. Many different methods have been used to purify this effluent, but none of them can be described as perfect (Idaka *et al.*, 1978). In the case of biological treatment. the effluent from the conventional treatment systems is low in BOD but still high in color. The traditional aerobic wastewater treatment systems do not substantially decrease the coloration of these wastewaters (Carliell *et al.*, 1995; 1996), while a number of research reports demonstrated the effectiveness of anaerobic decolorization (Gingell and Walker 1971; Bhattacharya *et al.*, 1989; Carliell *et al.*, 1994; Banat *et al.*, 1996). Under anaerobic conditions, many types of bacteria can reduce the dyes, cleaving the azo bond and forming two aromatic amines (Weber et al., 1987). Biodegradability studies on various dyes showed that anaerobic processes can also be used to decolorize dye wastewaters and improve the wastewater biodegradability for subsequent aerobic treatment (Brown and Hamburger, 1987; Boe *et al.*, 1993; Seshadri *et al.*, 1994; Carliell *et al.*, 1995).

In order to investigate the anaerobic decolorization capability by a two-stage UASB system, three different colors of reactive dyebath effluents were chosen and studied. This particular process was chosen to be investigated because the potential of decolorization of the acidification process, besides the conventional UASB, could be also preliminarily determined. The effect of different tapioca starch concentrations (as a co-substrate) on the color removal efficiency was simultaneously investigated. The co-substrate was added in order to improve the treatability of the more refractory matter, which in this case comprised dyestuffs (Atlas, 1993).

MATERIALS AND METHODS

Various experimental conditions, e.g., with and without additives as well as a co-substrate, and different initial colors and concentrations, were investigated in this research.

The two-phase UASB system

Three sets of the two-stage UASB process, consisting of 3 litre acidification tanks followed by 3 litre methane production or UASB units, with 12 hours hydraulic retention time (HRT) each, were used in this study (Figure 1). The acidification tanks were 0.14 m. ϕ shallow cylinder-typed reactors in which 20 mm. PVC overflow pipes were installed to have a down and up flow pattern as well as an effective liquid height of 0.2 m. The main parts of the UASB unit were the upflow 55 mm. ϕ x1 m. height (H) acrylic cylinders, completed with the 0.15 m. H internal gas-solids separator (GSS), making the overall height of the UASB 1.15 m.

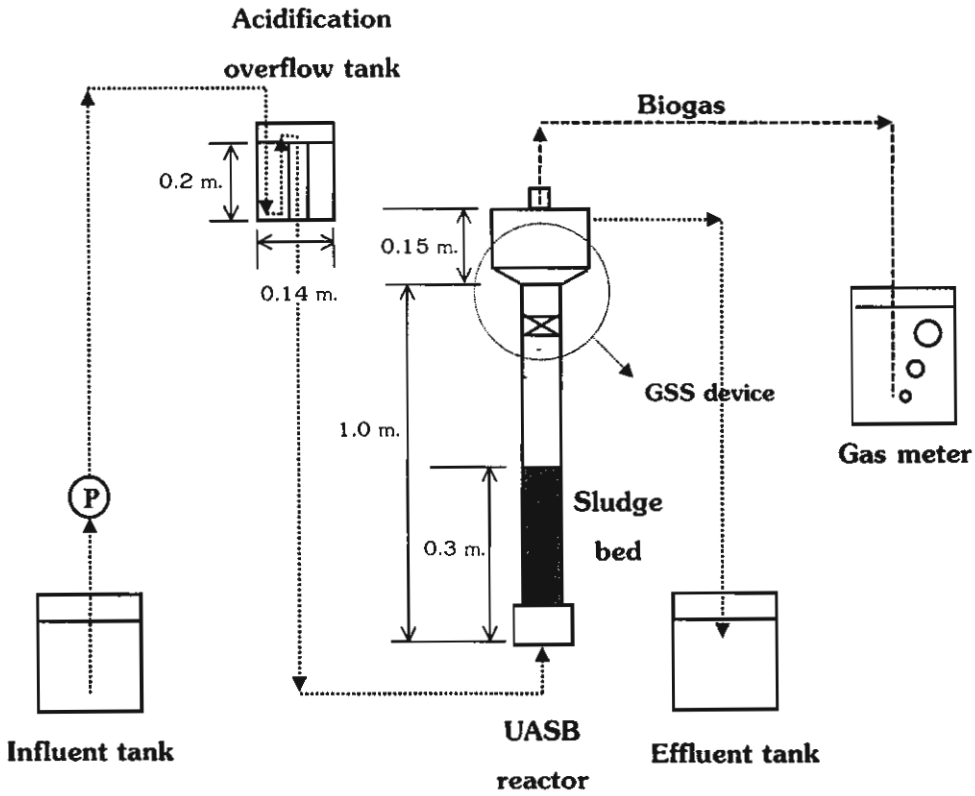

Figure 1. Schematic diagram of the two-phase UASB system

The bacterial seed for the main reactors came from 3-4 mm. ϕ brown pellets taken from existing UASB pilot models which have been operated for more than 3 years on synthetic sewage and another 2 years on tapioca and pineapple canning wastewaters. The pellet MLSS concentration ranged from 40 to 50 g/l. Equal volumes of the seed were added to the three UASB reactors to achieve a final pellet height of about 0.3 m. The acidification

tanks were, however, unseeded and the fermentation process was supposed to take place naturally through tapioca hydrolysis due to the suitable and relatively constant room temperatures of 29 to 32°C. Both reactors were also fully covered with black plastic sheets to block sunlight and subsequent algal growth.

Table 1. Characteristics of three dyebath effluents collected from a local dyehouse

Parameter	Black (BK)	Red (R)	Blue (BL)
pH	9.8	9.4	9.95
Color (SU)*	3,300	1,000	2,550
COD (mg/l)	2,250	2,700	1,650
Chloride (mg/l)	245	60,000	250

* Space units, see Gregor (1992).

Only reactive dyes which are very soluble and problematic were used in this study, and both real and synthetic wastewaters were investigated. The first one consisted of the dyebath effluents, i.e., black (BK), red (R) and blue (BL), from a local dyehouse. Because they were taken before the rinsing process, they were very high in COD and color (Table 1) as well as some additives such as acetic acid, sodium carbonate, sodium hydroxide, sodium chloride, etc. These wastewaters were diluted to a predetermined color level of 150 SU (space units) before being stored in the influent tank. It is noted that due to the very slow alteration in characteristics of these wastewaters (kept at room temperature, 29-32°C), a sufficient amount of the dyebath effluents was collected from a local dyehouse and subsequently used for the entire investigation period which lasted some 2 to 5 months for each experiment. The latter or synthetic one was prepared by dissolving the dye powder in tap water according to the dyeing specifications but without the supplementation of dye additives. The composition of these various wastewaters can be seen in Table 2.

It is known that the co-substrate is an alternate growth substrate which when supplied to a bio-reactor can enhance the degradation of some wastes or pollutants that cannot alone support the microbial growth (Atlas, 1993). It has also been reported that 1000 mg/l glucose could increase the color reduction efficiency of an anaerobic biomass inoculated in a serum bottle (Carliell *et al.*, 1995). Because tapioca is a cheap carbon source and produces the same intermediate metabolites, i.e., acetic acid, as glucose does, it was selected as the co-substrate at different concentrations in this study. The tapioca starch was dissolved in hot water prior to the addition of urea and KH_2PO_4 to achieve the COD : N : P ratio of not less

than 100 : 1.1 : 0.2 as recommended by McCarty (1964). Sodium bicarbonate at the rate of 0.5 mg $NaHCO_3$ per mg COD was also added as buffer to the system. In the first 3 experiments, which were run on real wastewaters, 0 to 1500 mg/l tapioca was used to investigate the effect of the co-substrate concentration, while the last 2 experiments, which were operated on the synthetic wastewater of blue color, were to investigate the effect of the dye additives and the initial color intensity (Table 2). During the start-up periods, the processes were operated at 500 or 1000 mg/l tapioca and a low color intensity of 25 SU, which was gradually increased until the pre-set value of 50, 100 or 150 SU, and then the supplemental tapioca was adjusted to the pre-determined concentrations. After the steady state was reached, the system was run further for at least 2 weeks, during which samples were drawn and analyzed daily.

Table 2. Wastewater composition and experimental details.

Expt.	Waste water	Color	Composition	Color intensity* (SU)	Tapioca conc" (mg/l)	Remarks
1	real	black	NA**	150	500, 1000, 1500	effect of co-substrate
2	real	red	Ciba Red FNR 86.7% Ciba Orange FR 11.2% Ciba Blue FR 2.1%	150	0, 200, 500	with and without co-substrate, red color
3	real	blue	Ciba Navy r-G 73.9% Ciba Red FB 15.2% Ciba Yellow F-SR 10.9%	150	0, 200, 500	with and without co-substrate, blue color
4	synthetic ***	blue	Ciba Navy r-G 73.9% Ciba Red FB 15.2% Ciba Yellow F-SR 10.9%	150	0, 200, 500	no additives
5	synthetic ***	blue	Ciba Navy r-G 73.9% Ciba Red FB 15.2% Ciba Yellow F-SR 10.9%	0, 50, 100	500	different color intensities, no additives

* 150 SU can be regarded as a low color intensity for dye-bath effluents. However, after dilution with wastewaters from other sources in the textile factory, this value can be in the middle range of real wastewaters from dyehouses.

**NA = not available

*** synthetic = no additives (e.g., sodium sulfate and acetic acid)

Note: the flowrate was kept constant at 6 l/d by 503S Watson Marlow peristaltic pumps.

The COD and alkalinity were analyzed according to Standard Methods (APHA, 1992), while the pH and ORP were measured by a F-13 Horiba pH meter and PHM80 Radiometer ORP meter (with a Ag/AgCl reference electrode in saturated KCl solution and a Pt electrode), respectively. The VFA-alkalinity was analysed and exposed as $CaCO_3$ by a titration method in the pH range of 5.75 to 4.3 (Moosbrugger, 1993). For the red dye case, in which the chloride content was very high, certain correction using excess $HgSO_4$ was done for the COD analysis, also according to the Standard Methods (APHA, 1992). The total organic carbon (TOC) was measured by a 5000 Shimadzu TOC analyser. The samples were filtered through GF/C paper prior to the true color (absorbance) measurement in visible 400-700 nm. wavelengths by a UV-1201 Shimadzu spectrophotometer. The aeras beneath the extinction curves were calculated and put in relation to one another. In other words, the summation of product of the absorbance and the wavelength throughout the visible range was computed and expressed in a single value as space units or SU (Gregor, 1992).

RESULTS AND DISCUSSION

Tables 3 to 7 tabulate the results of Experiments 1 to 5, respectively. They do not show a clear relationship between ORP values and the decolorization efficiencies. The results only indicate that this system was operated under strictly anaerobic conditions, both in the acidification tanks and the UASB reactors. Although the ORP values, ranging from -300 to -388 mV, were not in the same low range (-375 to -475 mV) as reported by Carliell et al. (1995) (probably because of different types of ORP probes and operating conditions,i.e., serum bottles by Carliell et al. and pilot reactor in this study), they are very close and that is why the decolorization did still occur in this study.

Table 3. Experimental results under high tapioca concentration for 'real' black dyebath effluent.*

Parameters	500 mg/l tapioca			1000 mg/l tapioca			1500 mg/l tapioca		
	I	A	E	I	A	E	I	A	E
pH	8.45	6.88	7.24	8.46	6.65	7.47	8.49	6.45	7.41
Temp. (°C)	31.2	30.5	30.8	31.2	30.4	31.2	31.2	30.5	31.1
ORP (mV)		-353	-358		-349	-360		-341	-359
Alkalinity (mg/l as $CaCO_3$)	458	621	657	616	832	907	791	998	1150
VFA-alkalinity (mg/l as $CaCO_3$)	129	254	215	173	401	282	223	563	363
COD_T (mg/l)	646	413	168	1172	729	147	1675	1140	172
COD_T removal (%)		36	74		38	87		32	90
Color (SU)^F	159	86	52	156	83	46	157	92	49
Color removal^F (%)		45	67		47	71		41	69

Remarks : * *ave. results of at least 2 weeks' work after a steady state condition*

I = influent, A = overflow from the acidification tank, E = effluent

COD_T = total COD of unfiltered samples

F = filtered through GF/C paper before analysis.

Table 4. Experimental results under no and low tapioca concentration for 'real' red dyebath effluent.*

Parameters	0 mg/l tapioca			200 mg/l tapioca			500 mg/l tapioca		
	I	A	E	I	A	E	I	A	E
pH	7.62	8.25	8.01	7.78	7.41	7.34	7.89	6.99	7.24
Temp. (°C)	29.7	29.7	29.9	30.0	29.8	30.0	30.0	29.8	29.9
ORP (mV)		-351	-388		-379	-382		-380	-384
Alkalinity (mg/l as $CaCO_3$)	509	598	630	570	693	773	674	867	942
VFA-alkalinity (mg/l as $CaCO_3$)	198	198	205	218	267	276	249	375	367
COD_T (mg/l)	408	329	298	619	458	368	940	719	519
COD_T removal (%)		19	27		26	40		23	45
Color (SU)^F	152	144	92	151	136	64	154	132	66
Color removal^F (%)		5	39		10	58		14	57

Remarks : * *ave. results of at least 2 weeks' work after a steady state condition*

I = influent, A = overflow from the acidification tank, E = effluent

COD_T = total COD of unfiltered samples

F = filtered through GF/C paper before analysis.

Table 5. Experimental results under no and low tapioca concentration for 'real' blue dyebath effluent.*

Parameters	0 mg/l tapioca			200 mg/l tapioca			500 mg/l tapioca		
	I	A	E	I	A	E	I	A	E
pH	7.45	7.67	7.48	7.57	7.14	7.32	7.79	6.83	7.14
Temp. (°C)	29.8	29.7	30.0	29.9	29.7	30.0	29.8	29.7	29.9
ORP (mV)		-328	-357		-367	-377		-370	-379
Alkalinity (mg/l as $CaCO_3$)	347	371	392	408	503	539	507	673	757
VFA-alkalinity (mg/l as $CaCO_3$)	117	124	133	134	185	177	162	300	252
COD_T (mg/l)	96	85	84	300	172	82	602	347	96
COD removal (%)		11	13		43	73		42	84
Sulfide (mg/l)	N/D	5.5	7.0	N/D	30.8	38.8	N/D	45.8	53.1
TOC (mg/l)	40.2	33.3	32.2	x	54.7	29.0	x	103.0	16.5
Color $(SU)^F$	152	142	79	154	132	73	155	121	69
Color removal (%)		6	48		14	52		22	56

Remarks : ** ave. results of at least 2 weeks' work after a steady state condition*

I = influent, A = overflow from the acidification tank, E = effluent

COD_T = total COD of unfiltered samples

F = filtered through GF/C paper before analysis

N/D = non detectable.

x = due to the fact that samples had to be filtered through GF/C paper in order to eliminate the suspended solids before analysis, the correct influent TOC values were not achieved when starch was added to the feed solution.

Table 6. Experimental results with synthetic blue dye wastewater (no dye additives)*.

Parameters	0 mg/l tapioca			200 mg/l tapioca			500 mg/l tapioca		
	I	A	E	I	A	E	I	A	E
pH	6.84	6.98	6.84	7.25	6.33	6.70	7.79	6.04	6.54
Temp. (°C)	29.5	29.6	29.9	29.5	29.5	30.0	29.5	29.5	29.9
ORP (mV)		-157	-301		-340	-346		-332	-347
Alkalinity (mg/l as $CaCO_3$)	62	68	81	136	173	197	246	279	352
COD_T (mg/l)	47	57	56	250	175	80	549	405	115
COD_T removal (%)		-21**	-20**		30	68		26	79
Sulfide (mg/l)	N/D	0.96	1.3	N/D	7.4	5.3	N/D	7.7	6.8
TOC (mg/l)^F	19.5	21.8	20.5	x	59.1	28.7	x	126.5	33.0
IC (mg/l)^F	15.6	17.1	20.5	34.8	37.8	54.9	65.0	54.0	99.6
TC (mg/l)^F	35.1	38.8	41.0	x	97.0	83.6	x	180.6	132.6
Color (SU)^F	152	152	97	152	134	70	153	124	66
Color removal^F (%)		0	36		12	54		18	57

Remarks : * *ave. results of at least 2 weeks' work after a steady state condition*

I = influent, A = overflow from the acidification tank, E = effluent

COD_T = total COD of unfiltrated sample

F = filtered through GF/C paper before analysis

***due to sulphide interference.*

N/D = non detectable.

x = due to the fact that samples had to be filtered through GF/C paper in order to eliminate the suspended solids before analysis, the correct influent TOC values were not achieved when starch was added to the feed solution.

Table 7. Experimental results for different initial synthetic color intensities with 500 mg/l tapioca*.

Parameters	0 SU color			50 SU color			100 SU color		
	I	A	E	I	A	E	I	A	E
pH	7.84	6.23	6.75	7.81	6.29	6.77	7.81	6.27	6.78
Temp. (°C)	30.2	29.9	30.1	30.3	30.0	30.3	30.1	29.8	30.2
ORP (mV)		-340	-342		-334	-347		-332	-345
Alkalinity (mg/l as $CaCO_3$)	265	294	373	267	294	363	263	286	357
COD_T (mg/l)	509	334	44	521	333	68	536	340	78
COD_T removal (%)		34	91		36	87		37	86
TOC (mg/l)^F	x	119.1	17.4	x	115.8	21.1	x	115.4	28.0
IC (mg/l)^F	71.0	56.1	106.4	70.6	54.4	103.1	69.2	56.5	100.3
TC (mg/l)^F	x	175.3	123.8	x	170.3	124.2	x	171.9	128.4
Biogas (l/d)			0.47			0.42			0.38
Color (SU)^F	-	-	-	56	34	21	106	90	47
Color removal^F (%)		-	-		39	63		15	56

Remarks : ** ave. results of at least 2 weeks' work after a steady state condition*

I = influent, A = overflow from the acidification tank, E = effluent

COD_T = total COD of unfiltrated sample

F = filtered through GF/C paper before analysis

x = due to the fact that samples had to be filtered through GF/C paper in order to eliminate the suspended solids before analysis, the correct influent TOC values were not achieved when starch was added to the feed solution.

Effects of co-substrate concentrations

It is apparent from Tables 3 to 5 that there were pH drops during the acidification stage, demonstrating that the fermentation process yielding volatile fatty acid (VFA) obviously occurred in these simple reactors and the higher VFA-alkalinity production was achieved at the higher tapioca concentrations, e.g., the VFA-alkalinity in the black dye case increased from 129 to 254 and from 223 to 563 mg/l as $CaCO_3$ for 500 and 1500 mg/l tapioca, respectively. It is also obvious that an appreciable level of decolorization, e.g., 41-47 percent for the black dye condition at 500 to 1500 mg/l tapioca, took place in the acidification units. This implies that the bacteria prevalent in the acidification units also had the capability to reduce the color, at least for this black dye scenario in which an ample

amount of readily biodegradable carbon, both from the additive (acetic acid) and tapioca, was available. On the other hand, the overall color reduction of the black dye was somewhat constant, i.e., 67, 71 and 69 per cent for the case of 500, 1000 and 1500 mg/l tapioca, respectively. One can therefore preliminarily conclude that the tapioca used here was excessive for this 150 SU black dye concentration.

In the subsequent experiments on the real wastewaters with red and blue dyes, only 200 or 500 mg/l tapioca was added into the systems. The process without the tapioca addition (0 mg/l) as the "control" unit was also investigated in order to see the effect of the co-substrate on the decolorization capability of the process. Tables 4 and 5 showed that the decolorization processes, though somewhat reduced (10-22%), still took place during the acidification step for the case of 200 and 500 mg/l tapioca concentration. The discrepancy between this and the black dye case is probably due to the different characteristics of those dyes and the level of added tapioca. The acid forming bacteria (AFB) could therefore play some role in the decolorization mechanisms in this process. The overall color reduction, in terms of SU, of 58 and 57% for the red dye and 52 and 56% for the blue dye at 200 and 500 mg/l tapioca, respectively, was comparable to those of higher tapioca dosages in the first experiment at 1000 and 1500 mg/l. Some 200 to 500 mg/l tapioca is therefore suggested to be the minimum requirement for good decolorization for this 150 SU condition.

Meanwhile, a small degree of decolorization (5-6%) was even possible in the acidification step of the real wastewaters in the 'control' or no co-substrate scenario, due to the availability of the readily degradable carbon source (acetic acid) which was an additive in the dyeing process (Tables 4 and 5). On the other hand, the overall efficiency of this "control" condition was 39 and 48 per cent for the red and blue dyes, respectively, which was regarded as quite high because there was no external carbon source available. However, the addition of a co-substrate (of at least 200 mg/l tapioca) could enhance the color removal performance of the acidification step, i.e., the decolorization efficiency in this acidification tank increased from 5 to 10 and 14% and from 6 to 14 and 22% for the 0 to 200 and 500 mg/l tapioca concentration for the red and blue dyes, respectively. The decolorization efficiency of the acidification step hence evidently increased with the tapioca or co-substrate concentration.

Figure 2 demonstrates the color reduction efficiency for the three **real** dye wastewaters at a constant 500 mg/l tapioca starch concentration. The removal efficiency of 14 percent was

seen in the acidification step of the red color case, which was less than 22 and 45 percent for the blue and black conditions, respectively. This was owing to less color intensity in the red dyebath effluent (Table 1), resulting in less dilution required and more dye materials in the feed water to the test units. This higher amount of red dye was presumably the reason behind the poorer performance in the acidification reactor. However, when this higher color intensity was loaded into the UASB reactor, a higher reduction efficiency was achieved, i.e., the UASB decolorization was 50, 43 and 40 percent for the red, blue and black dye scenarios, respectively, making the overall performance somewhat similar, i.e., 57, 56 and 67 percent, respectively. This happening indicated that the decolorization occurred more in the UASB reactor than in the acidification unit, naturally because much more bacterial mass was present in the UASB reactors than in the acidification units and this UASB mass was possibly excessive or more color could still be removed by these reactors.

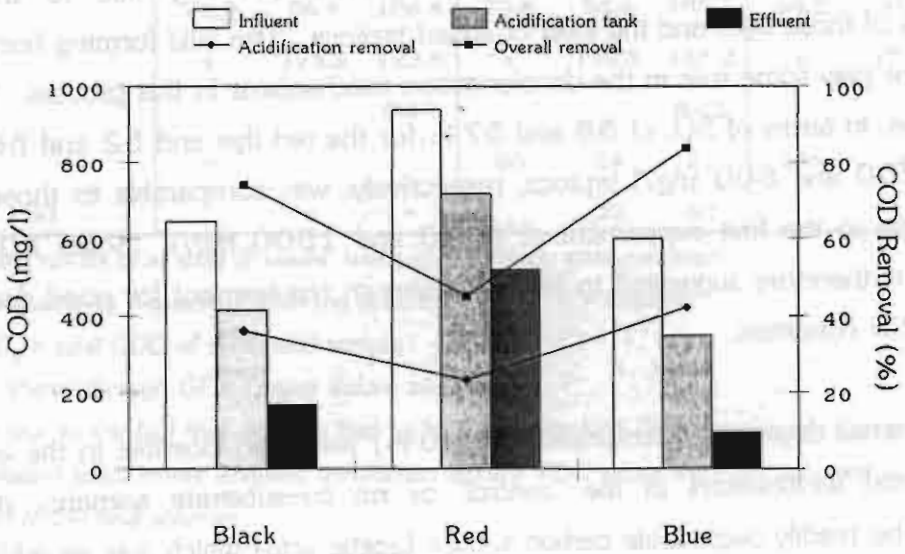

Figure 2. Decolorization for three dyebath effluents with 500 mg/l tapioca starch.

Effects of dye additives

The main dye additives were sodium sulfate and acetic acid. The effect of the sodium sulfate will be discussed in the next paragraphs. Tables 5 and 6 tabulate the experimental results under the same conditions (150 SU and 0, 200 and 500 mg/l tapioca) of the real and synthetic blue dyes, respectively. It can be seen from Table 6 that the decolorization did not happen in the acidification step of the control (0 mg/l tapioca) test for the synthetic dye case (no C source), while some 6 percent was achieved for the real wastewater condition in which a certain amount of acetic acid was readily available (Table 5). That is, the bacteria in

the acidification reactor needed one form or another of a co-substrate, be it the acetate added during the dyeing process or the external carbon source such as tapioca, to be able to decompose the dye molecules. If none was present, the decolorization in this tank was not possible. However, when the wastewater was sent further to the UASB units, a substantial color reduction (36 percent) was achieved for the "control" synthetic wastewater while 48 percent was apparent for the real wastewater condition. A very high amount of cell mass available in the UASB pellets can definitely be attributed to this decolorization whose mechanism was presumably the azo bond cleavage which resulted in different and smaller intermediate molecules which did not posses the color-showing property. However, when the co-substrate or tapioca was added in the synthetic wastewater case (where no other carbon source was present), a better color removal was evident for both the acidification and methane production steps. It can therefore be concluded that the bacteria in both steps needed a co-substrate at a certain concentration to enhance their decolorization capability. However, for the real wastewater, in which a carbon source in the form of acetic acid was already available and due to the high bacterial mass in the UASB reactor, a co-substrate could only slightly improve the color removal performance of the process for this 150 SU blue dye case (Figure 3). In such circumstances, only a single-stage UASB may be sufficient to perform the required job of decolorization.

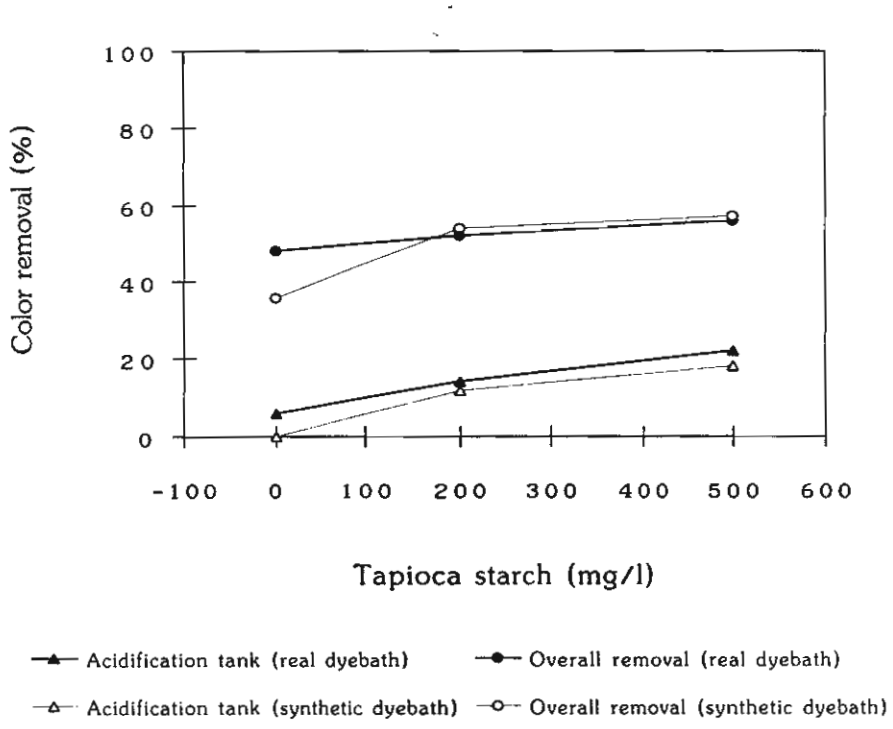

Figure 3. Decolorization efficiency of the real and synthetic wastewaters for the blue color scenario.

Sulfate Reducing Bacteria (SRB)

It is interesting to note that the sulfide production was apparent in both acidification and UASB reators (Tables 5 and 6). More production was clearer for the real than for the synthetic dye wastewaters, demonstrating that the sulfate reducing bacteria (SRB) co-existed with other bacteria in this process and could compete with those bacteria in the organic carbon consumption. This finding was more obvious when a co-substrate was added to the system, e.g., 6.8 and 53.1 mg/l $S^{=}$ were detected in the effluent for the synthetic and real dye wastewaters at 500 mg/l tapioca, respectively. When the sulfide production at the acidification and the UASB steps was compared for the 200 mg/l tapioca and real blue dye condition, it can be seen that there was not much difference, e.g., 30.8 and 38.8 mg/l $S^{=}$ was produced after the acidification and methane production steps, respectively. This implied that the SRB have already proliferated in the acidification reactor and the available sulfate was insufficient for more SRB growth in the subsequent UASB reactor.

It should be also noted from Tables 3 to 7 that the alkalinity concentrations increased in both the acidification and the methane production phases. However, the alkalinity production was reported to be hardly found in the acidogenesis step by the Emden-Meyerhof Pathway (Fenchel and Finlay, 1995) or in the methanogenesis stage (Gottschalk, 1988 and Fenchel and Finlay, 1995). The sulfate reducing bacteria which can release bicarbonate during their metabolisms (Widdle, 1988) can be the possible explanation for this happening.

However, there was no correlation between the quantity of the sulfide produced and the color removal efficiency. This meant that the SRBs were present but not the main decolorizing microorganisms in this two-stage process.

Since there was no carbon source other than the dye itself in the case of the synthetic dyebath without the co-substrate addition (Table 6), no significant TOC reduction was seen at the acidification and UASB reactors, i.e., the TOC was 19.5, 21.8 and 20.5 mg/l for the synthetic wastewater, the overflow from the acidification tank and the effluent of the UASB reactor, respectively. (Note: the COD was not here used as the interpretation parameter because of the sulfide interference.) But when a carbon source (acetic acid) was available as the dye additive for the case of the real wastewater without a co-substrate (Table 5), the TOC decreased from 40.2 to 33.3 and 32.2 mg/l, respectively. This means that

the SRB could compete with other microorganisms in the organic carbon consumption but with less efficiency of decolorization.

Methane Producing Bacteria (MPB)

Table 8. Decolorization and methane production for the experiments with real and synthetic dyebaths.

Parameters	Diluted real dyebath effluent		Synthetic dyebath effluent	
	Tapioca concentration		Tapioca concentration	
	200 mg/l	500 mg/l	200 mg/l	500 mg/l
- Biogas (l/d)	0	0.30	0.04	0.42
- Fraction of methane (%)	-	86	-	98
- Methane yield (l/g COD removed)*	0	0.18	-	0.23
- $Color^F$ removal (%)	52	56	54	57
- COD_T removal (%)	73	84	68	79

Remarks :

** Methane yield (l/g COD removed) = fraction of methane x biogas (l/d)/COD removed (g/d)*

Carliell et al. (1996) and Razo-Flores et al. (1997) reported that the decolorization took place under the methane production step in anaerobic processes. These bacteria used the azo bonds in chromophores of the dye as electron acceptor and, in the process, the azo bond was broken, resulting in the decolorization. As a result, the level of the methane formation should reflect the degree of the color removal. Table 8 shows the decolorization and the methane production of the system treating both real and synthetic wastewaters at 200 and 500 mg/l tapioca starch. Practically no biogas was produced under the 200 mg/l tapioca scenario despite the good COD removal. On the other hand, some gas was formed under the 500 mg/l tapioca case while the decolorization efficiency did not differ that much, i.e., 52 and 54 percent decolorization at 200 mg/l, 56 and 57 per cent at 500 mg/l for the real and synthetic wastewaters, respectively. It is therefore concluded that the strictly anaerobic MPB were not the main and only microorganisms responsible for the color reduction in this process. This statement contradicts those of Carliell et al. (1996) and Razo-Flores et al. (1997), but agrees with Eiamsamor (1999) and Panswad et. al., (2000) who worked on an anaerobic/aerobic SBR process in which facultative anaerobes

were predominant and reported approximately 50-60 percent decolorization in SU units. Moreover, it was reported earlier in this paper that the decolorization could also take place in the open acidification tank in which the facultative rather than strict anaerobes were present.

It is also noteworthy that the percentage of methane in the produced biogas was very high (86-98 percent) when compared with the theoretical value of 50 percent and the values found in practice or reported cases of approximately 60-70 percent (Metcalf and Eddy, 1991; Mosey, 1982). This very high percentage was shown to be possible under a low organic loading (Tantoolvest *et al.*, 1999).

Effect of initial color intensity

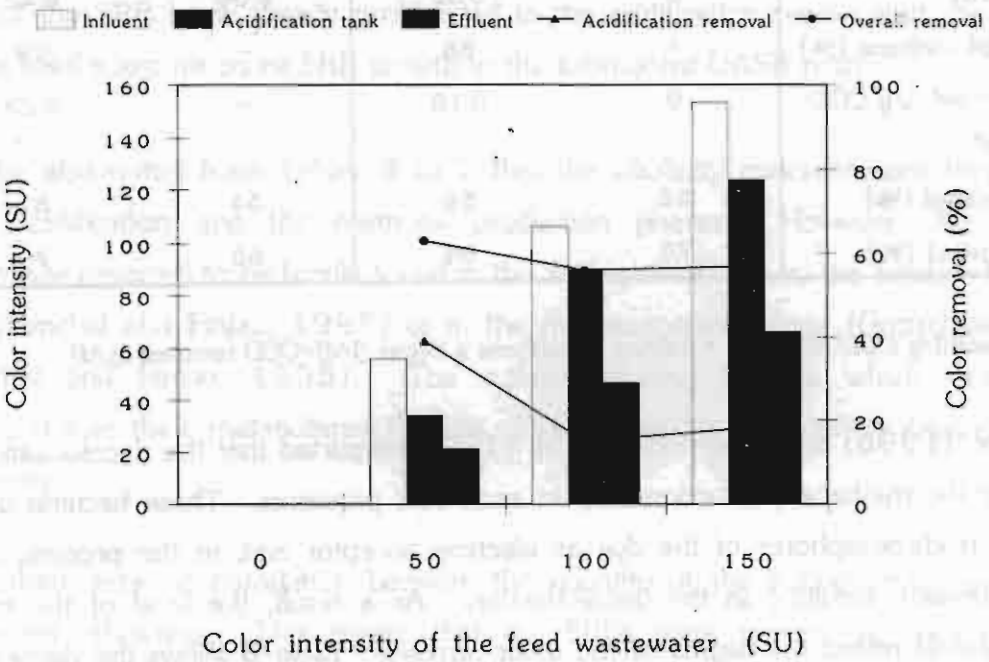

Figure 4. Decolorization efficiency at different blue color intensities and 500 mg/l tapioca.

Experimental results on the study of the effect of the color intensity at 500 mg/l tapioca are shown in Table 7. The color removal efficiency by the acidification process was 39 and 15 percent while the overall efficiency was 63 and 56 percent for 50 and 100 SU, respectively, i.e., a higher color reduction efficiency was possible for a lower color input. In addition, in comparing with the result of the 500 mg/l tapioca condition from Table 6, the decolorization performances of the 100 and 150 SU cases were quite similar for both the

acidification (12-18 percent) and the UASB (54-57 percent) reactors, whereas more color was removed in the first tank (39 percent) for the 50 SU condition (Figure 4). It is probable that the excessive dye concentration of 100 SU and over might be detrimental to the acid forming bacteria. On the other hand, since a large bacterial mass was present in the UASB pellets, the overall decolorization capability in terms of percentage was not much affected.

CONCLUSIONS

Three different reactive dyes, namely, black, red and blue, were used with a two-stage UASB system in a decolorization investigation with and without the addition of an external carbon source (tapioca starch) as a co-substrate. It is apparent that an optimum external carbon supplement could enhance the decolorization capacity of the process. On the other hand, the possibility of decolorization in the case of no external carbon source still existed, mainly because of the availability of the readily degradable acetic acid which is an additive normally used in the dyeing process. Certain sulfate reducing bacteria were present in the system and could wastefully use up the carbon sources without an efficient color reduction. Contrary to reports by others, this study illustrates that the strictly anaerobic methane forming bacteria were not the only microorganisms responsible for the decolorization. The acid forming bacteria may also have some role in the process.

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submitted paper

EFFECTS OF TEMPERATURE ON BIOLOGICAL PHOSPHORUS REMOVAL PERFORMANCE

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Abstract - The purpose of this research was to investigate the effects of temperature on biological phosphorus removal processes. Two 16.8 l automatic temperature controlled anaerobic/aerobic sequencing batch reactors were operated at 12 day sludge age and 12 hr cycle times to treat a synthetic wastewater of COD, TKN and TP of 300, 15 and 15 mg/l, respectively. The system was run for some 30 to 45 days before steady state was reached. Apparently, the high temperatures directly affected phosphorus removal efficiency because at 5, 15, 25, 35 and 40°C, P removal efficiencies at steady state dropped from 100, 100, and 100% to 72 and 61 percent, respectively. The P content in the mixed liquor volatile suspended solids (MLVSS) decreased from 10.8 and 10.4 at 5 and 15°C to 5.5. 3.1 and 3.5 percent for 25, 35 and 40°C respectively. Also, more PHA was anaerobically synthesized and stored in the cells at lower temperatures, i.e., 165, 140, 111, 58 and 58 mg/g VSS, respectively. The PHAs were utilized during the aerobic period and at the end were 70, 56, 64, 33 and 25 mg/g VSS, respectively. Very little anaerobic P release was detected at 35°C, while practically no P release occurred at 40°C. This clearly showed that the phosphorus accumulating organisms (PAOs) proliferated better at lower temperatures (5 to 25°C). However, temperature had less impact on the ordinary heterotrophs or on the filtered COD removal efficiency, which was 99, 99, 97, 99 and 93 percent. respectively, but temperature of 40°C was relatively unfavorable for COD reduction. The removal of TKN increased with temperature until 35°C, after which the efficiency also drastically dropped.

Key words - Activated sludge;biological phosphorus removal; temperature effects; anaerobic/aerobic; SBR; PHA

INTRODUCTION

There have been two schools of thought on the effects of temperature on the performance of enhanced biological phosphorus removal (EBPR) processes. The first group reported that the processes worked better at low temperatures (Sell *et al.*, 1981; Ekama and Marais, 1984; Barnard *et al.*, 1985; Kang *et al.*, 1985; Krichten *et al.*, 1985; Van Groenestijn and Deinema, 1985 (from Espanto and Jones, 1994); and Florentz *et al.*, 1987), while the second group stated otherwise (Jones *et al.*, 1987; Yoeman *et al.*, 1988; Waltrip, 1991; Mamais and Jenkins, 1992; McClintock *et al.*, 1993; Converti *et al.*, 1995; Jones *et al.*, 1996; Choi *et al.*, 1998; Helmer and Kunst, 1997). In some of these experiments, the system was a closed one, resulting in less potential to recycle and entrain the electron acceptors, DO and NO_x, into the anaerobic mixed liquor at the low temperature scenario, and consequently obtained better phosphorus removal at lower temperature (Ekama and Marais, 1984). In others, however, the system was run at an unacclimated condition to temperature alternation and, as a result,

experienced temperature shock (Mamais and Jenkins, 1992). However, McClintock *et al.*, 1993, and Mamais and Jenkins, 1992, also concluded that EBPR functions will "wash-out" of activated sludge at a higher sludge age than COD removal functions at any given temperature, just as nitrifiers can be "washed-out".

Literature review indicates that there is a paucity of information regarding the effects of temperatures above 30°C on EBPR. It was the intention of this study to evaluate EBPR efficiency under steady state conditions at different temperatures, ranging from 5 to 40°C, using an anaerobic-aerobic (AA) sequencing batch reactor (SBR) system. The synthesis and utilization of polyhydroxyalkanoate (PHA) was also investigated.

MATERIALS AND METHODS

Two 16.8 litre automatic temperature controlled AA - SBR systems (Figure 1) were first inoculated with 11 litres of an EBPR sludge, which was taken from a 30 litre batch AA culturing unit operated at 25-30°C. This unit was previously inoculated with pure cultures of phosphorus accumulating organisms (PAOs), i.e., *Pseudomonas fluorescens* and *Acinetobacter calcoaceticus* (Timm *et al.*, 1990 and Ohtake *et al.*, 1985), in addition to activated sludge from a local extended-aeration sewage treatment plant, and run separately for 7 months until the PAOs were in abundance and the system exhibited good EBPR characteristics.

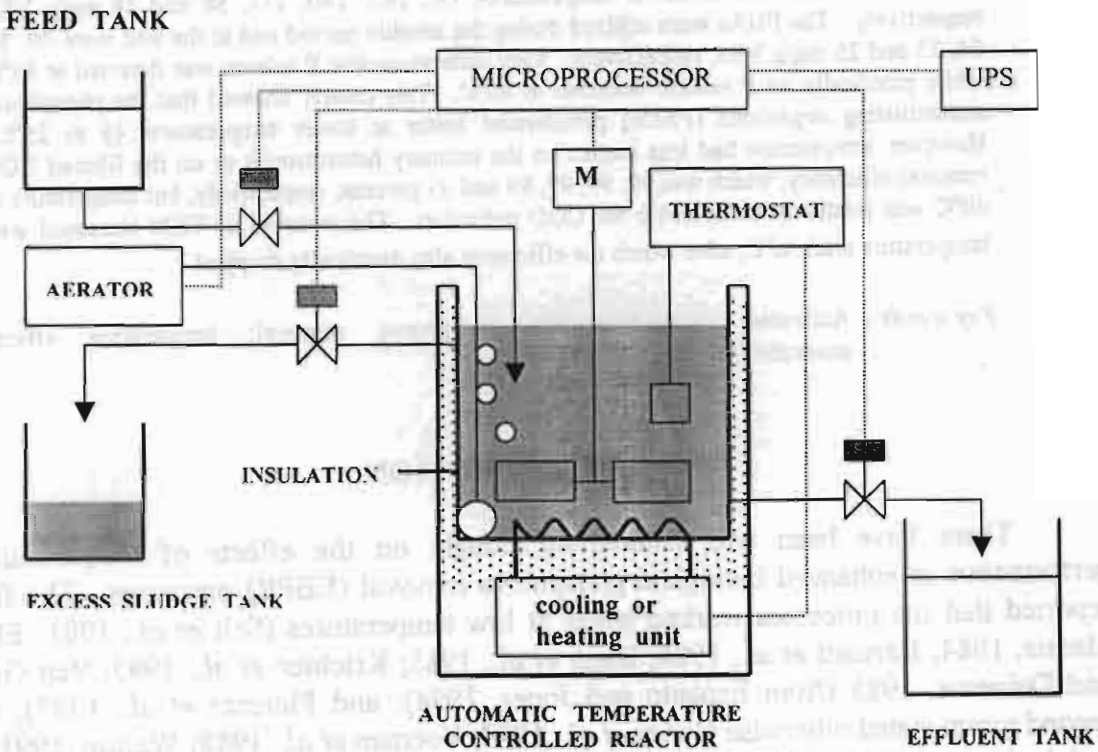


Figure 1. Automatic temperature controlled anaerobic-aerobic sequencing batch reactor.

A synthetic wastewater with COD, TKN and TP concentrations of 300, 15 and 15 mg/l (100:5:5), respectively, was used to promote the proliferation of PAOs while minimizing the

growth of nitrifiers (Table 1). Thus the effect of nitrate interference on EBPR efficiency was minimized. The 300 mg/l COD consisted of two readily biodegradable carbon sources, namely, sodium acetate, SA (190 mg/l COD), and nutrient broth, NB (110 mg/l COD). Sodium bicarbonate at 400 mg/l $CaCO_3$ was also added as a buffer to the system. The radio of total phosphorus to calcium, magnesium and potassium was 1:0.4:0.33:0.36 in terms of moles, which was approximately equal to the ratio of 1:0.5:0.25:0.23, suggested by Randall *et al.* (1992). The ratio of the feed volume (V_f) to the remaining volume in the tank (V_o) was 2:1, and the operating 12 hr cycle is described in Table 2. The subsequent anaerobic and aerobic solids retention times (SRTs) were 5.5 and 6.5 days, respectively, for a total of 12.0 days. The SRT was maintained by withdrawing 0.7 litres of mixed liquor per cycle, including the samples drawn for analysis, near the end of the aerobic step. To make sure that the dissolved oxygen (DO) was rapidly increased during the aerobic phase, two air pumps was used during the first 45 minutes. After that one air pump was turned off.

Table 1. Synthetic wastewater composition

Composition	Concentration
Nutrient broth	110 mg COD/l (including TKN of 15 mgN/l)
CH_3COONa	190 mg COD/l
KH_2PO_4	15 mg P/l ; 4.3 mg K/l
$NaHCO_3$	400 mg $CaCO_3$/l
$MgSO_4.7H_2O$	3.75 mg Mg/l
$FeCL_3$	0.9 mg Fe/l
$CaCl_2$	7.5 mg Ca/l

Table 2. Working cycle of system

Period	Time
Fill*	5 minutes
(Anoxic**) Anaerobic	4 hrs. 50 minutes
Aerobic	6 hrs.
Excess sludge ***	5 minutes
Settling	1 hr.
Decant	10 minutes
Total	12 hrs.

* mix-fill, done along with the anaerobic period

** anoxic condition before anaerobic phase, because of existing nitrate in V_o

***done near the end of the aerobic period.

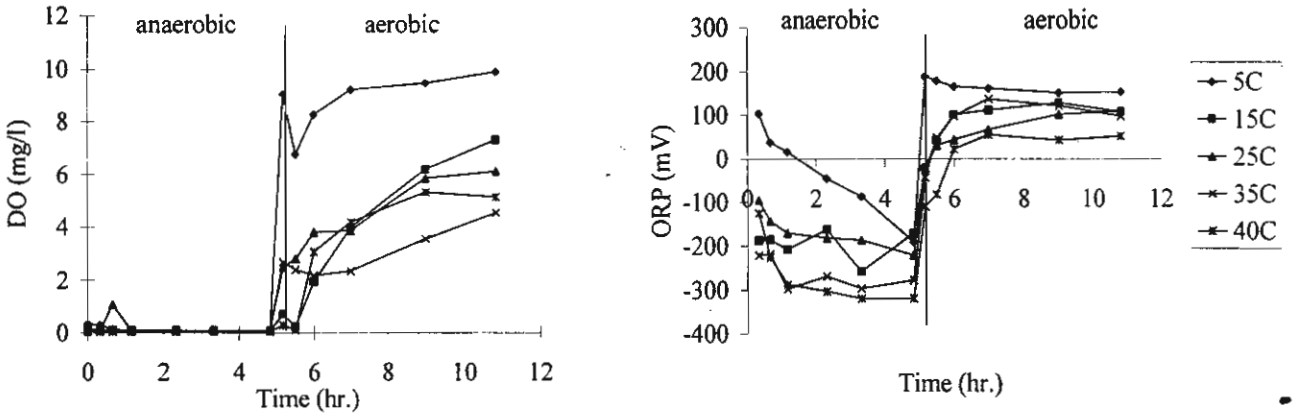
The EBPR enriched sludge was allowed to adjust from room temperature (28 – 30°C) to the desired temperature at the rate of one degree per day. The mixed liquor, after being dump fed with the wastewater, reached the specified temperature within 15 minutes. After that, the system was operated at the controlled temperature until steady state, which normally took another 1 to 1.5 months. Daily samples were then taken and analyzed for 12 consecutive days. Average results for at least 5 days during this steady state period were reported. Certain samples were filtered using GF/C paper while those for soluble phosphorus were 0.45 μ filtered before analysis, which were performed according to Standard Methods (APHA, AWWA and WPCF, 1992). The nitrite and nitrate tests were performed using a SQ 118 Merck Photometer. For the PHA determinations the samples were extracted by sulfuric acid as well as chloroform (Lee *et al.*, 1995), and subsequently analyzed by a Shimadzu 7G gas chromatograph. The volatile fatty acids (VFA) were analyzed for C1 to C6 (Ripley *et al.*, 1986), also by the Shimadzu 7G gas chromatograph.

RESULTS AND DISCUSSION

The development of steady state conditions at the temperatures of 5, 15, 25, 35 and 40°C, took 32, 42, 26, 42 and 46 days, respectively. Different MLVSS concentrations were

obtained for each temperature. The average steady state concentrations were 990, 958, 1264, 1979 and 692 mg/L, respectively, for the temperatures from low to high. Typical profiles of dissolved oxygen (DO) and oxidation reduction potential (ORP) are shown in Figure 2. The anaerobic DO was negligible while the aerobic DO was rapidly raised to concentrations of 9.9, 7.2, 6.1, 4.4 and 5.1 mg/l for the 5 temperatures from low to high, respectively. The final anaerobic ORP ranged from –179 to –315 mV.

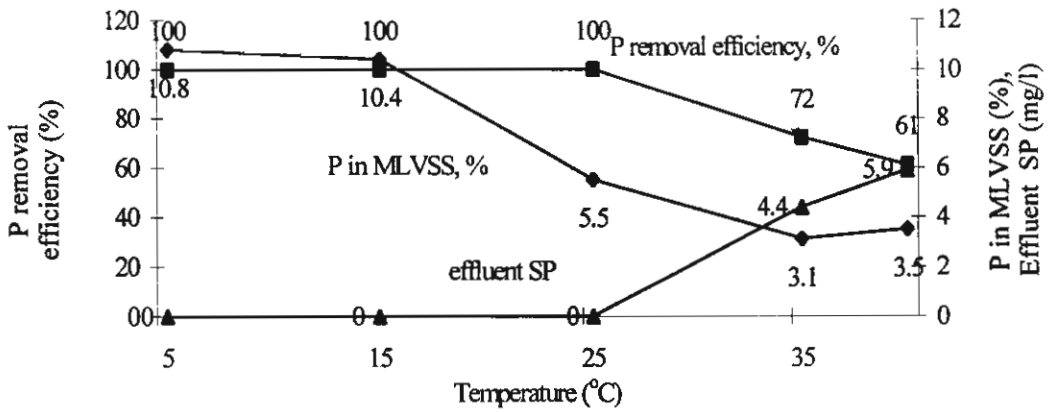
The 5 day average phosphorus (P) removal results are shown in Figure 3. It is evident that better P removal was achieved at low temperatures because the effluent soluble phosphorus (SP) was 0, 0, 0, 4.4 and 5.9 mg/l, while the P content of the VSS averaged 10.8, 10.4, 5.5, 3.1 and 3.5 percent for the said temperatures, respectively. Complete phosphorus removal at 5°C was also been reported possibly by others, via extension of the aerobic time (Baeten *et al.*, 1999) and through increase in sludge age (Brdjanovic *et al.*, 1998). The soluble phosphorus profiles during the anaerobic and aerobic stages also demonstrate better phosphorus removal at 5 to 25°C than at 35 and 40°C (Figure 4).



a) DO profile

b) ORP profile

Figure 2. DO and ORP profiles at the end of the anaerobic and aerobic steps under steady state conditions.



Note: Phosphorus removal (%) = (TP-SP) x 100/TP

Figure 3. Phosphorus removal, effluent soluble phosphorus concentrations and VSS phosphorus content at steady state.

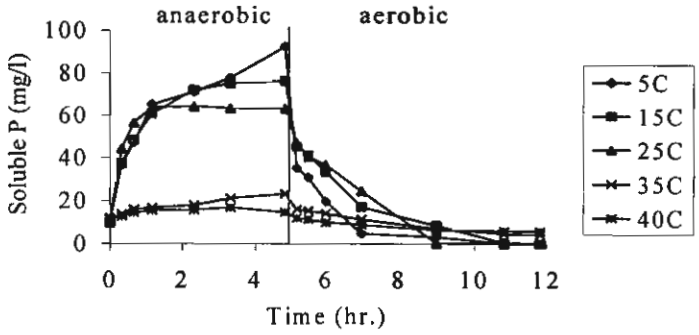
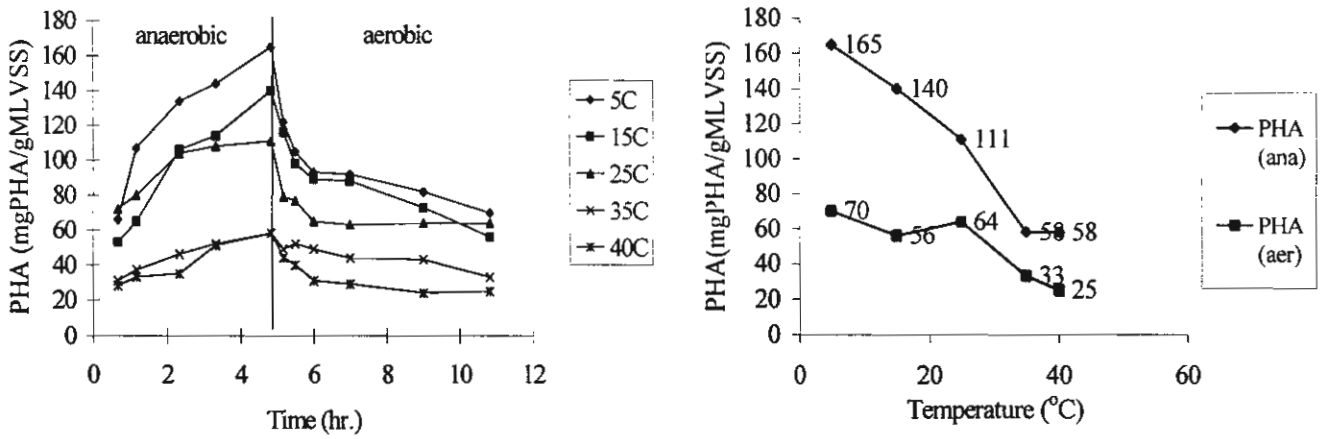


Figure 4. Profiles of soluble phosphorus at different temperatures.

More soluble phosphorus release (63.2 and 92.4 mg/l) and luxury P uptake (78.2 to 107.4 mg/l) were obvious during the anaerobic and aerobic steps at 5 to 25°C, while only about 23.8 mg/l was released at 35°C, and practically no soluble phosphorus release took place at 40°C. It is well known that aerobic phosphorus uptake correlates strongly with anaerobic phosphorus release, as recently noted by Helmer and Kunst (1997), and Sell *et al.* (1981) and Krichten *et al.*(1985) have reported that PAOs are psychrophilic. However, Helmer and Kunst (1997) also reported that phosphorus uptake did not depend on phosphorus release at the low temperature of 10°C, which does not agree with this study. The patterns of anaerobic P release and aerobic luxury P uptake coincided very much with the data on anaerobic PHA storage and aerobic PHA utilization (Figure 5a). More PHA was anaerobically produced and stored in the PAO cells at the lower temperatures of 5 to 25°C, which coincided with the already mentioned more soluble phosphorus release. Baetens *et al.* (1999) reported that anaerobic PHB (polyhydroxybutyrate) storage increased as temperature decreased, which is the same as found in this study (Figure 5b). They also obtained the greatest amount of PHB storage in the cells at 5°C. Additionally, Satoh *et al.*, (1992) reported that more PHA was aerobically utilized at low temperatures resulting in higher soluble P uptake, as observed during this study (Figure 5b).



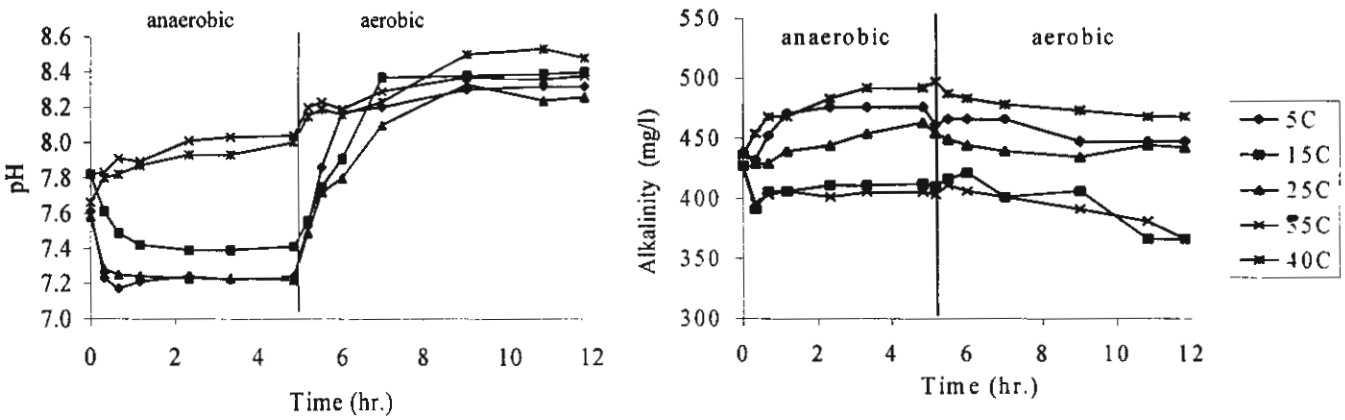
a) PHA profile

b) Anaerobic and aerobic PHA at steady state

Figure 5. PHA profiles and PHA VSS content at the end of the anaerobic and aerobic steps under steady state conditions.

It was noted that with the ratio of the readily biodegradable COD to phosphorus in the influent of about 20:1, organic carbon was in excess and complete phosphorus removal was possible in the process providing that the EBPR mechanism was developed (Hong *et al.*, 1984; Randall *et al.*, 1992). The stored PHA in this case was, therefore, not totally utilized (Figure 5b), and the system actually could have removed more phosphorus at the lower temperatures if more than 15 mg/l TP had been fed to the process. This is also why 100 percent phosphorus removal was achievable at 25°C even though the phosphorus content in the VSS was only 5.5, in contrast to contents of 10.8 and 10.4 percent at 5 and 15°C, respectively.

Figure 6a and 6b illustrate the typical profiles of pH and alkalinity in the system. The anaerobic alkalinity changed variably, but tended to increase a little. It varied from 420 – 457 mg $CaCO_3$/l in the influent (before dilution in the SBR units), with averages of 472, 411, 464, 405 and 491 mg $CaCO_3$/l for the five temperatures from low to high, respectively. Although relatively little nitrogen was added, the alkalinity subsequently decreased in the aerobic phase because of nitrification to 460, 368, 444, 379 and 471 mg $CaCO_3$/l, respectively. The reduction in alkalinity for the two extreme temperatures, i.e., 5 and 40°C was relatively small because the nitrifiers were inhibited (Figures 8b & 9B). This is consistent with the results of Choi *et al.*, (1998). The final anaerobic pHs for the low temperatures of 5, 15 and 25°C were lower than those at 35 and 40°C, i.e., 7.2 – 7.4 versus 8.0. The aerobic pHs were, however, similar for all temperatures, i.e., 8.25 to 8.47.



a) pH profile

b) Alkalinity profile

Figure 6. Profiles of pH and alkalinity under steady state conditions.

A good relationship between soluble phosphorus release, PHA production and VFA removal is evident in the data (Figures 4, 5, and 7). At lower temperatures, more phosphorus was anaerobically released (Figure 4), resulting in more energy available and more VFA uptake (Figure 7), as well as more PHA synthesized and stored (Figure 5). This happened when the pHs were optimal (7.2 to 7.4) for the EBPR process (Liu *et al.*, 1996). On the other hand, when the temperature was raised to 35 and 40°C, the anaerobic pH increased to 8.0 and poor EBPR was achieved. Smolders *et al.*, (1994) reported that anaerobic P release was pH dependent because of anaerobic substrate metabolism, and pH in the range of 5.5 – 8.2 had an effect on the kinetics and thermodynamics of P release, but not on acetate uptake. They also stated that the pH effect on the uptake of acetate could be related to the increasing electric

potential across the cell membrane as pH increased, and at high pH more P release is required to obtain the same level of energy to uptake an equal amount of VFA. However, in this study VFA uptake was not good at temperatures of 35 and 40°C (Figure 7), but the P releases also were low at the high temperature scenarios (Figure 4). The finding in this study then does not fully agree with the statement by Smolders *et al.* (1994), probably because the EBPR functions were more strongly inhibited by the temperature effects. A more focused study is needed to reach a final conclusion.

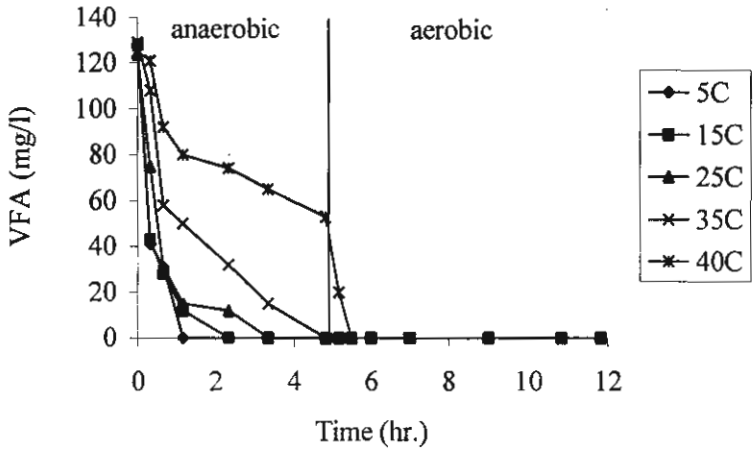
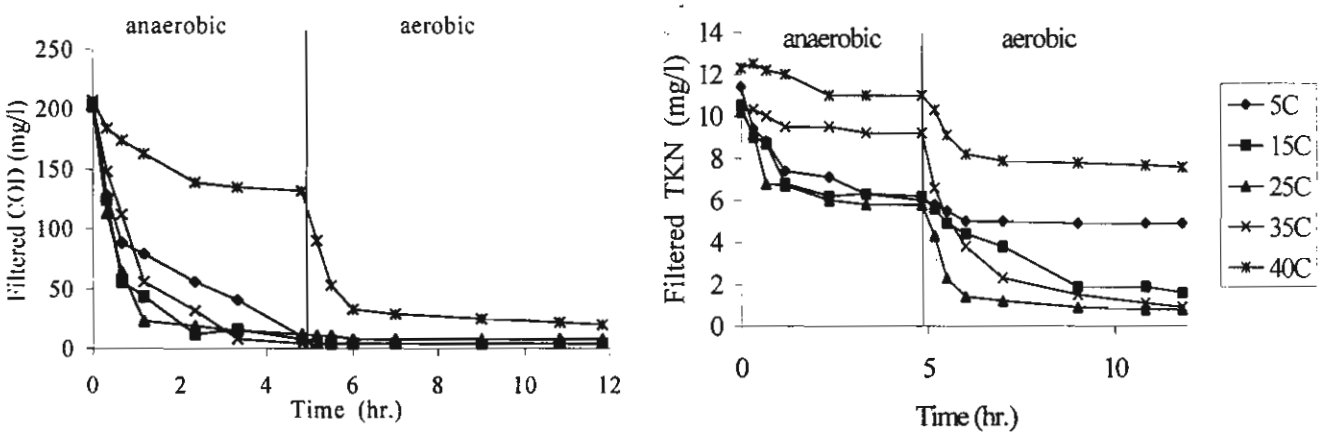


Figure 7. Profiles of filtered VFA under steady state conditions.



a) Filtered COD

b) Filtered TKN

Figure 8. Profiles of filtered COD and TKN under steady state conditions.

Figures 8a and 8b show the profiles of filtered COD and TKN removal in the system for the five temperatures, and they agree quite well with the VFA data. That is, complete filtered COD removal was achieved during the anaerobic phase for all temperatures except 40°C. A following aerobic phase was definitely required to decrease the filtered COD to about 20 mg/l at this temperature. Because of the higher organic carbon loading to the aerobic phase plus temperature inhibition, less nitrification was possible at this temperature. Also, the

nitrifiers did not perform very well at 5°C, as previously reported by Choi *et al.*, (1998), and good ammonium oxidation was not achieved in the aerobic step at this temperature. Very little nitrite and nitrate were present in both stages (Figure 9), and the impact of anoxic condition was negligible, as designed. Some nitrite accumulation did occur at 15 and 35°C, and effluent nitrates were highest at the three middle temperatures.

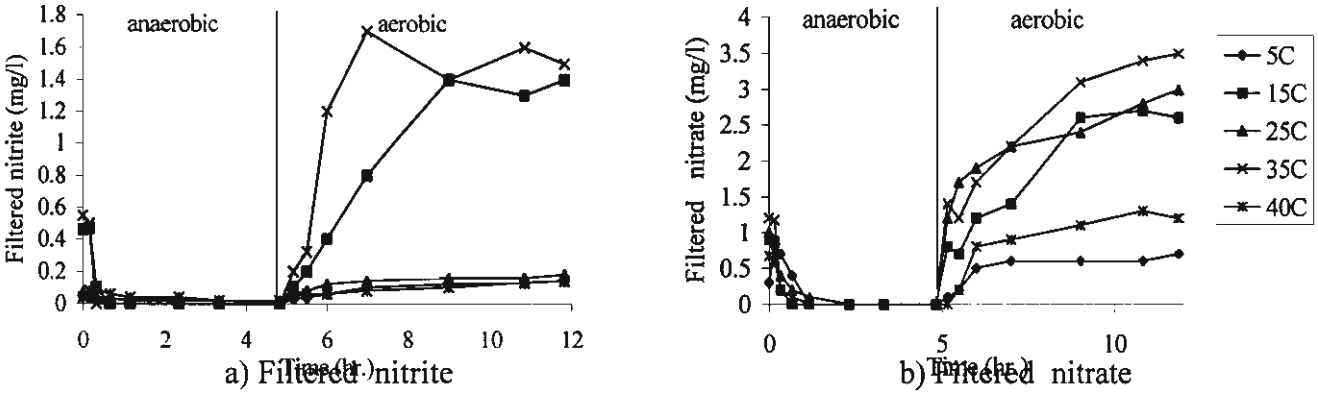


Figure 9. Profiles of filtered nitrite and nitrate under steady state conditions.

The overall filtered COD, TKN and TN removal efficiencies are shown in Figure 10. The TKN reduction in the anaerobic step probably was through the uptake and internal storage as polypeptides and amino acids as reported by Satoh *et al.*, (1998). Because of a sufficient sludge age, the filtered COD removal was very high (97 to 99 percent), except for the 40°C temperature which was a little less than normal and dropped to 93 percent. The filtered TKN removal, i.e., the nitrification capability of the process increased from 67 to 94 percent as temperature increased from 5 to 35°C, but then decreased to 57 percent at 40°C because of temperature inhibition of the nitrifiers (Wild *et al.*, 1971).

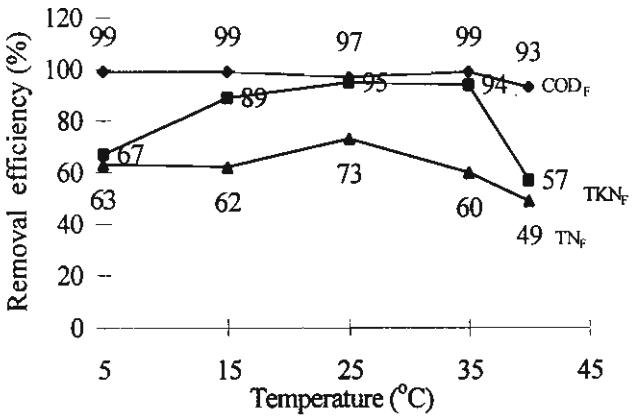


Figure 10. COD_F , TKN_F and TN_F removal efficiencies at different temperatures

From the above results and discussions, it is apparent that, through selective processes, the bacterial composition of activated sludge systems operated for EBPR changes with temperature. This agrees with the conclusion of Brdjanovic *et al.* (1998).

CONCLUSIONS

The results of this investigation strongly indicate that PAOs are physcrophilic bacteria, and phosphorus removal performance improves as temperature decreases down to 5°C. The phosphorus accumulating organisms (PAOs) proliferated and performed well at temperatures as low as 5°C, and EBPR efficiency at steady state was clearly better at the lower temperatures investigated (5 to 25°C) than at the high ones (35 to 40°C). Under the optimum ratio of biodegradable COD to P of 20:1 and with minimum interference from nitrates, the P removal efficiency dropped from 100 percent at 5, 15 and 25°C to 72 and 61 percent at 35 and 40°C, respectively. Also, higher soluble phosphorus release and greater PHA storage occurred during the anaerobic period at the lower temperatures, and higher luxury phosphorus uptake and PHA utilization occurred during the following aerobic stage.

It was concluded that EBPR processes were severely inhibited at 35°C, and were non-existent at 40°C for the operating conditions of these experiments. It also was observed that nitrification was strongly inhibited at both 5 and 40°C.

Only a slight difference in COD removal (97 – 99%) was seen over the 5 to 35°C range, but when the temperature reached 40°C, the COD reduction efficiency decreased to 93 percent.

ACKNOWLEDGEMENT

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2. ผลงานตีพิมพ์ในวารสารวิชาการในประเทศไทย

printed paper

ชฎารัตน์ อนันต์ *
ธงชัย พรรณสวัสดิ์**

กระบวนการกำจัดฟอสฟอรัสทางชีวภาพ

การบำบัดน้ำเสียในปัจจุบันฟอสฟอรัสเป็นปัญหาหนึ่งที่ต้องได้รับการพิจารณาอย่างรอบคอบกว่าเดิมที่ผ่านมา เนื่องจากฟอสฟอรัสเป็นธาตุอาหารชนิดหนึ่งของพืชน้ำ ดังนั้นหากน้ำเสียที่ได้รับการบำบัดแล้วแต่ยังไม่กำจัดฟอสฟอรัสออกไป เมื่อถูกทิ้งลงสู่แหล่งรับน้ำทั่วๆไป เช่นแม่น้ำ ลำคลอง หนองบึงต่างๆ ก็จะก่อให้เกิดสภาพยูโทรฟิเคชัน ซึ่งเป็นสภาพที่เกิดการเจริญเติบโตของสาหร่ายเซลล์เดียวมากผิดปกติ ทำให้แหล่งน้ำมีสีเขียวหรือแดง นอกจากนี้สาหร่ายเหล่านี้จะใช้ออกซิเจนในแหล่งน้ำในยามที่ไม่มีแสงแดดทำให้เกิดการขาดแคลนออกซิเจน เมื่อสาหร่ายตายไปก็เกิดการหมักทำให้น้ำเน่าเสียและมีกลิ่นเหม็นตามมาได้อีก ดังนั้นจึงจำเป็นอย่างยิ่งที่ต้องกำจัดฟอสฟอรัสออกจากน้ำเสียก่อนที่จะทิ้งสู่แหล่งรับน้ำ

การกำจัดฟอสฟอรัสกระทำได้ 2 วิธีด้วยกันคือ ก) วิธีทางเคมี ซึ่งใช้สารเคมีในการตกตะกอน ผลึก การสร้างตะกอนหรือการดูดซับฟอสฟอรัสก่อนที่จะปล่อยทิ้ง และ ข) วิธีทางชีววิทยาซึ่งเป็นการใช้จุลินทรีย์ในการบำบัดโดยควบคุมสภาพทางสิ่งแวดล้อมให้เหมาะสมต่อแบกทีเรียที่จะใช้กำจัดฟอสฟอรัส

กระบวนการกำจัดฟอสฟอรัสทางชีวภาพ

การกำจัดฟอสฟอรัสทางชีวภาพนั้นใช้วิธีการเลี้ยงเชื้อ โดยมีขั้นตอนเริ่มจากสภาพที่ไม่ใช้ออกซิเจนและตามด้วยสภาพที่เป็นแอโรบิก และสุดท้ายปล่อยให้จุลชีพจมตัวลงในถังตกจะกอน จึงทำให้สามารถกำจัดฟอสฟอรัสออกจากระบบด้วยการระบายสลัดจ์ส่วนที่มีฟอสฟอรัสเป็นส่วนประกอบนี้ออกจากระบบ

ในสภาวะที่เป็นแอนแอโรบิกคือไม่มีออกซิเจนทั้งในรูปออกซิเจนละลายน้ำ (DO) และในรูปไนเทรตนั้น แบกทีเรียที่สะสมฟอสฟอรัสได้เป็นพิเศษหรือที่เรียกว่า โพลี-พีแบกทีเรีย (poly - p bacteria) จะใช้กรดไขมันระเหยง่าย (volatile fatty acids , VFAs) เช่นอะซิเตท เป็นสารอาหารและปล่อยโพลีฟอสเฟตออกมานอกเซลล์ โดยโพลี-พีแบกทีเรียจะเก็บสำรองอาหารส่วนหนึ่งไว้ในรูปของพีเอชเอ (poly-(-hydroxyalkanoate, PHA) ซึ่งเท่ากับ PHB + PHV และเมื่อเข้าสู่สภาวะแอโรบิก พีเอชเอที่ถูกเก็บไว้นี้จะถูกแบกทีเรียใช้ไปและปล่อยพลังงานออกมา ซึ่งแบคทีเรียจะใช้พลังงานนี้ในการดึงโพลีฟอสเฟตมาเก็บไว้ในเซลล์ ซึ่งโพลีฟอสเฟต ที่โพลี-พีแบกทีเรียจัดเก็บได้มากเป็นพิเศษนี้จะถูกกำจัดโดยการทิ้งสลัดจ์ส่วนเกินออกจากระบบนั่นเอง

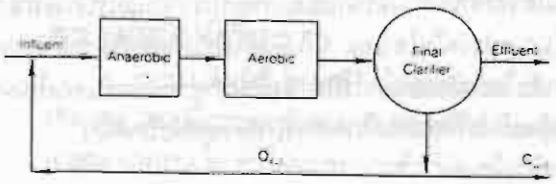
กระบวนการที่สามารถกำจัดฟอสฟอรัสทางชีวภาพได้นั้นมีด้วยกันหลายกระบวนการ คือมีทั้งกระบวนการที่สามารถกำจัดได้เฉพาะฟอสฟอรัสอย่างเดียว และกระบวนการที่สามารถกำจัดได้ทั้งฟอสฟอรัสและไนโตรเจน

1. กระบวนการที่กำจัดฟอสฟอรัสได้อย่างเดียวสามารถแบ่งได้เป็น 2 กระบวนการด้วยกันคือ

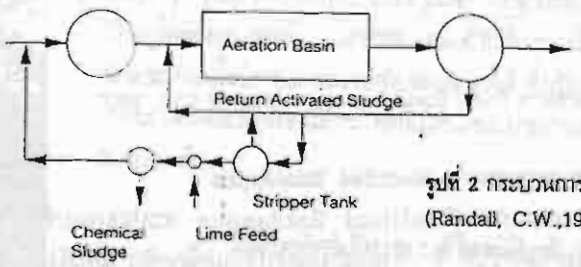
1.1 กระบวนการแอนแอโรบิก/ออกซิก (A/O process)

กระบวนการแอนแอโรบิก/ออกซิกประกอบด้วยถังแอนแอโรบิกตามด้วยถังแอโรบิก ดังรูปที่ 1 ซึ่งการปล่อยฟอสฟอรัสจะเกิดขึ้นในถังแอนแอโรบิก และเมื่อน้ำเสียผ่านไปยังถังออกซิกก็จะเกิดการจับใช้ฟอสฟอรัส จากนั้นน้ำเสียที่มีสลัดจ์ที่สะสมฟอสฟอรัสเป็นพิเศษนี้ก็จะไหลเข้าสู่ถังตกตะกอนขั้นที่สอง จุลชีพเกิดการจมตัวและถูกเวียนกลับเข้าระบบอีกครั้ง โดยที่สลัดจ์ส่วนหนึ่งจะถูกกำจัดออกจากระบบโดยการระบายทิ้งเพื่อควบคุมอายุสลัดจ์ (solid retention time , SRT) ซึ่งก็ทำให้ฟอสฟอรัสถูกกำจัดออกไปด้วยพร้อมๆกัน

สำหรับกระบวนการแอนแอโรบิก/ออกซิกนี้จะออกแบบโดยใช้อายุสลัดจ์สั้นๆและอัตราภาระอินทรีย์สูงๆ เพื่อให้อัตราการผลิตสลัดจ์เกิดได้มากและทิ้งออกไปมากด้วยเช่นกัน



รูปที่ 1 กระบวนการแอนแอโรบิก-ออกซิก
(WEF Manual and Practice, 1992)



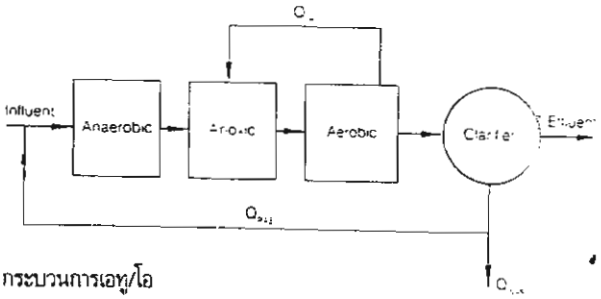
รูปที่ 2 กระบวนการ Phostrip
(Randall, C.W.,1992)

1.2 กระบวนการ Phostrip

กระบวนการ Phostrip เป็นกระบวนการกำจัดฟอสฟอรัสที่ใช้วิธีกำจัดทางชีววิทยาและทางเคมีร่วมกัน แสดงดังรูปที่ 2 ระบบบำบัดหลักของกระบวนการนี้ก็คือ กระบวนการแอกทิเวเต็ด สลัดจ์ ซึ่งประกอบด้วยถังแอนแอโรบิกกับถังตกตะกอนขั้นที่สอง ส่วนการกำจัดฟอสฟอรัสนั้นจะ

กำจัดออกจากน้ำเสียที่ออกจากจุดเวียนสลัดจ์กลับของถังตกตะกอนชั้นที่สอง โดยจะนำน้ำสลัดจ์นี้เข้าสู่ถัง Stripper ภายในถัง Stripper จะมีสภาพเป็นแอนแอโรบิก ดังนั้นก็จะเกิดการปล่อยฟอสฟอรัสออกจากเซลล์น้ำเสียที่มีฟอสฟอรัสความเข้มข้นสูง เป็นส่วนประกอบนี้จะถูกนำไปบำบัดโดยวิธีทางเคมี คือใช้ปูนขาวทำให้ตกตะกอน ซึ่งสลัดจ์ที่ได้นี้จะถูกนำไปกำจัดต่อไป ส่วนน้ำเสียที่อยู่ด้านบนก็จะถูกเวียนกลับเข้าสู่ระบบ

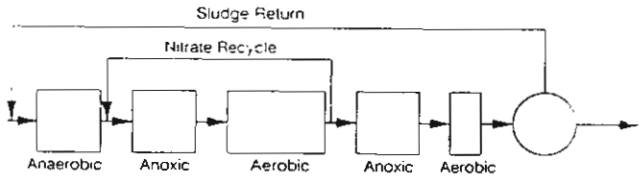
2. กระบวนการที่สามารถกำจัดได้ทั้งไนโตรเจนและฟอสฟอรัสมี 4 กระบวนการด้วยกันคือ



รูปที่ 3 กระบวนการเอทู/โอ
(WEF Manual and Practice, 1992)

2.1 กระบวนการเอทู/โอ (A2/O process)

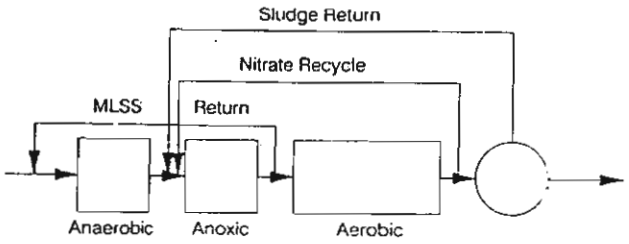
กระบวนการเอทู/โอ ประกอบด้วยถังแอนแอโรบิก/แอน็อกซิก/ออกซิก ดังรูปที่ 3 กระบวนการนี้จะเพิ่มถังแอน็อกซิกขึ้นระหว่างถังแอนแอโรบิกกับถังออกซิก ทั้งนี้เพื่อให้เกิดปฏิกิริยาดีไนตริ ฟิเคชัน ซึ่งเป็นสภาพที่น้ำเสียขาดแคลนออกซิเจนอิสระ มีเพียงออกซิเจนในรูปของไนเทรตหรือไนไทรต์เท่านั้นที่ไนตริฟายอิงแบกทีเรียจากถังแอโรบิกจะสามารถนำไปใช้ได้ในการกำจัดไนโตรเจน นี่ก็เป็นทางหนึ่งที่ลดปริมาณไนเทรตที่จะถูกเวียนกลับไปยังถังแอนแอโรบิก ซึ่งถ้าหากความเข้มข้นของไนเทรตมีมากเกินไปก็จะมีผลให้ประสิทธิภาพการกำจัดฟอสฟอรัสลดลง



รูปที่ 4 กระบวนการ Five - Stage Bardenpho (Randall C.W., 1992)

2.2 กระบวนการ Modified Bardenpho

กระบวนการ Modified Bardenpho จะประกอบด้วย 5 ขั้นตอนด้วยกันดังรูปที่ 4 ซึ่งถังแอนแอโรบิก/แอน็อกซิก และถังออกซิกนั้นจะใช้เพื่อการกำจัดฟอสฟอรัส ไนโตรเจน และ คาร์บอน สำหรับถังแอน็อกซิกถังที่สองนั้นมีเพื่อให้เกิดปฏิกิริยาดีไนตริฟิเคชันเพื่อกำจัดไนเทรตที่มาจากถังแอโรบิกถังแรก ส่วนถังแอโรบิกถังสุดท้ายจะใช้ในการถ่ายเทก๊าซไนโตรเจนที่ยังเหลืออยู่ออกจากสารละลาย และป้องกันไม่ให้สภาพแอนแอโรบิกไปเกิดในถังตกตะกอนชั้นที่สอง



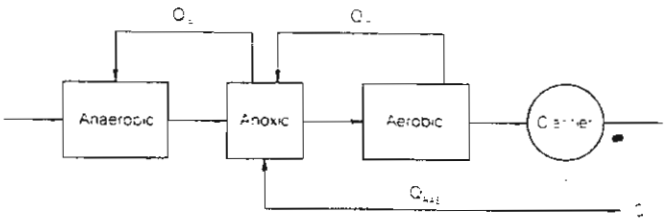
รูปที่ 5 กระบวนการยูซีที (Randall C.W.,1992)

2.3 กระบวนการยูซีที (University of Cape Town process)

กระบวนการยูซีทีรวม 3 ถังหลักที่ใช้ในการกำจัดฟอสฟอรัสและไนโตรเจนคือส่วนแอนแอโรบิก/แอน็อกซิก/ออกซิก ดังรูปที่ 5 การเวียนสลัดจ์ของกระบวนการนี้จะเวียนเข้าสู่ถังแอน็อกซิกแทนที่จะเข้าถังแอน แอโรบิกจากนั้นจึงเวียนจากถังแอน็อกซิกไปยังถังแอนแอโรบิก ทั้งนี้เพื่อลดภาระไนเทรตที่จะไปรบกวนการกำจัดฟอสฟอรัสในส่วนแอนแอโรบิก

สำหรับถังแอน็อกซิกนั้นจะเป็นส่วนที่รับการเกิดสภาพสลัดจ์ลอยตัวเนื่องจากปฏิกิริยาดีไนตริฟิเคชัน ดังนั้นการแยกส่วนนี้ออกจากส่วนที่จะเวียนตะกอนกลับสู่ถังแอนแอโรบิกก็จะทำให้ไนเทรตส่วนเกินที่จะถูกเวียนมายังส่วนนี้ไม่ไปรบกวนกระบวนการ

รูปที่ 6 กระบวนการวีไอพี
(WEF Manual and Practice, 1992)



2.4 กระบวนการวีไอพี (The Virginia Initiative Plant process)

กระบวนการวีไอพีแสดงดังรูปที่ 6 จะเห็นได้ว่ากระบวนการนี้มีลักษณะคล้ายกับกระบวนการเอทู/โอ และกระบวนการยูซีที ยกเว้นวิธีการเวียนสลัดจ์ของระบบ โดยการเวียนสลัดจ์จากถังแอโรบิกจะเวียนเข้าสู่ทางเข้าถังแอน็อกซิก และน้ำสลัดจ์จากถังแอน็อกซิกก็จะเวียนเข้าสู่ทางเข้าถังแอนแอโรบิกต่อไป

เอกสารอ้างอิง

Randall, C.W.,Burnard,J.L. and Stensel,H.D. Design and Retrofit of Wastewater Treatment Plants for Biological Nutrient Removal. , Technomic Publishing Company , Inc. Pennsylvania

WEF Manual and Practice . Integrated Biological Processes for Nutrient Removal Design of Municipal Wastewater Treatment Plants . 2:1992

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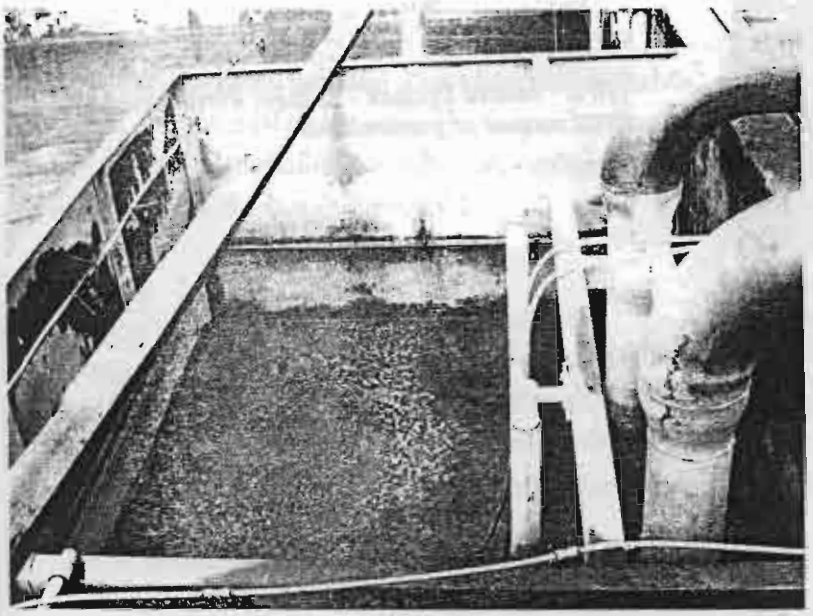
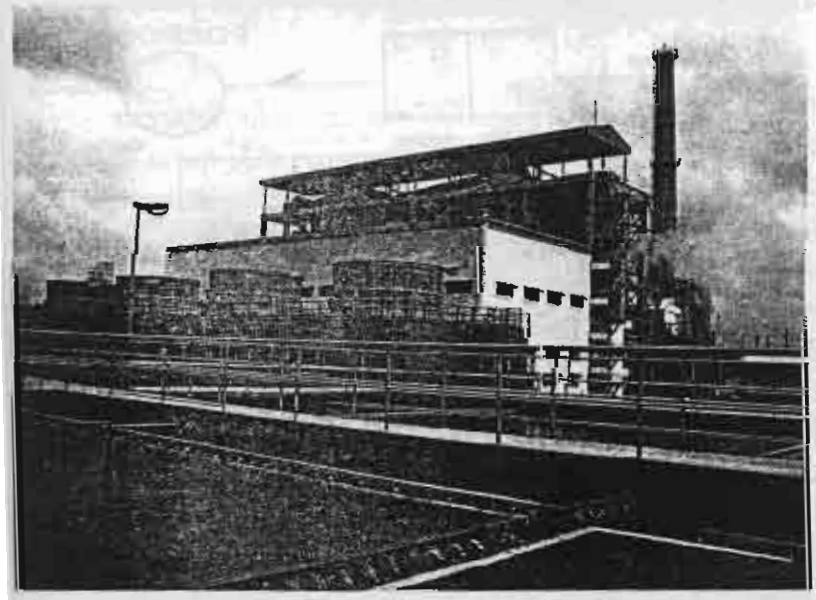
printed paper

WATER Pollution

ปัญญรัตน์ ยลพฤกษา *
องชัย พรรณสวัสดิ์ * *

การกำจัดไนโตรเจน ด้วยกระบวนการแอกทิเวเต็ดสลัดจ์ แบบแอน็อกซิก – ออกซิก





ในปัจจุบันกระบวนการแอกทิเวเต็ดสลัดจ์ซึ่งเป็นกระบวนการหนึ่งของระบบที่จุลินทรีย์เติบโตแบบแขวนลอย (suspended growth process) ได้ถูกนำมาประยุกต์ใช้ในการกำจัดไนโตรเจน โดยออกแบบถังปฏิกรณ์ในขั้นต่างๆ ให้มีสภาพแวดล้อมเหมาะต่อกระบวนไนตริฟิเคชัน และดีไนตริฟิเคชันซึ่งสามารถแบ่งออกได้เป็น 3 แบบด้วยกัน (WEF , 1992) คือ

1. ระบบสลัดจ์เดี่ยว(single - sludge system)
2. ระบบสองสลัดจ์ (dual - sludge system)
3. ระบบสามสลัดจ์ (triple - sludge system)

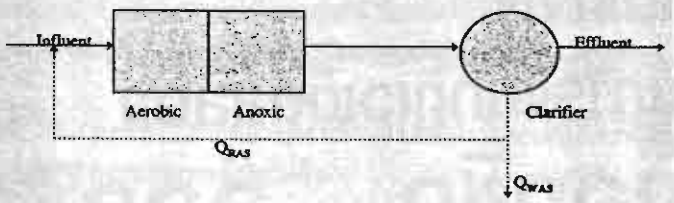
ระบบสลัดจ์เดี่ยว

ระบบสลัดจ์เดี่ยวอาจเรียกได้อีกอย่างหนึ่งว่า **ระบบเชื้อผสม'** ระบบนี้ประกอบด้วยแอกทิเวเต็ดสลัดจ์เพียงชุดเดียว ลักษณะของระบบโดยทั่วไปจะประกอบด้วยถังแอน็อกซิก (ไม่มีออกซิเจนอิสระ แต่มีออกซิเจนในรูปอื่นๆ เช่น เป็นองค์ประกอบอยู่ในไนเทรต เป็นต้น) ถังแอโรบิกหรือออกซิก (มีออกซิเจนอิสระ) และถังตกตะกอน โดยถังแอน็อกซิกและถังออกซิก อาจมีมากกว่า 1 ถัง แต่ถังตกตะกอนจะมีเพียงใบเดียวเท่านั้น โดยวางอยู่ในตำแหน่งสุดท้ายของระบบ ส่วนตำแหน่งของถังแอน็อกซิกและถังแอโรบิกสามารถวางสลับตำแหน่งได้ ซึ่งเป็นสาเหตุให้ระบบสลัดจ์เดี่ยวมีกลไกในการกำจัดไนโตรเจนแบ่งย่อยออกอีกเป็น 2 แบบ คือ

ก) ระบบสลัดจ์เดี่ยว**แบบดีไนตริฟิเคชันเกิดทีหลัง** (post denitrification)

เมื่อน้ำเสียเข้าสู่ระบบ **ถังแรกที่รองรับน้ำเสียคือถัง**แอโรบิกหรือถังออกซิก (ดูรูปที่1) **ในถังนี้**หากมีเวลากักพักนานพอจะเกิดกระบวนการ**กำจัดสาร**อินทรีย์คาร์บอนในภาวะที่มีออกซิเจนอิสระและเกิด**กระบวน**การไนตริฟิเคชันที่แปลงรูปของสารอินทรีย์ไนโตรเจน**ไปสู่รูปของ**ไนเทรต ปริมาณออกซิเจนที่เติมต้องเพียงพอต่อการ**กำจัด**สารอินทรีย์คาร์บอน

*นิสิตมหาบัณฑิต ภาควิชาวิศวกรรมสิ่งแวดล้อม จุฬาลงกรณ์มหาวิทยาลัย
** ศาสตราจารย์ ภาควิชาวิศวกรรมสิ่งแวดล้อม จุฬาลงกรณ์มหาวิทยาลัย

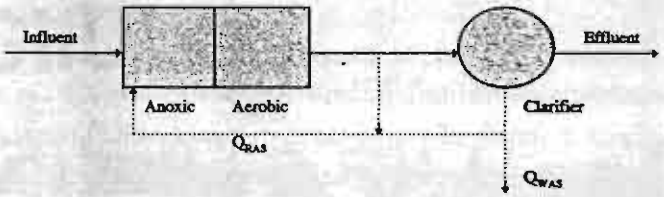


รูปที่ 1 กระบวนการ Wuhermann สำหรับการกำจัดไนโตรเจน (WEF manual of practice,1992)

และการเกิดไนตริฟิเคชัน น้ำที่ออกจากถังนี้จะมีสารอินทรีย์คาร์บอน (หรือบีโอดี) ต่ำมากและมีไนเทรตสูง จากนั้นจะไหลเข้าถังที่สองหรือถังแอน็อกซิกที่ซึ่งจะเกิดปฏิกิริยาดีไนตริฟิเคชันขึ้น โดยจุลินทรีย์จะดึงเอาออกซิเจนจากไนเทรตมาใช้และเกิดการลดรูปหรือรีดักชันของไนเทรตขึ้น แต่มักพบว่าปริมาณสารอินทรีย์คาร์บอนในถังแอน็อกซิกนี้จะไม่เพียงพอต่อการเกิดกระบวนการดีไนตริฟิเคชันอย่างสมบูรณ์ เนื่องจากสารอินทรีย์คาร์บอนส่วนใหญ่ถูกกำจัดออกไปในถังแอโรบิกก่อนหน้านี้แล้ว ทำให้จุลินทรีย์เฮเทอโรทรอฟ (ซึ่งต้องใช้อินทรีย์คาร์บอนเป็นแหล่งอาหารและพลังงาน) มีสารอาหารคาร์บอนไม่พอจึงทำให้อัตราการเกิดดีไนตริฟิเคชันมีค่าต่ำมาก ตัวอย่างของระบบได้แก่ Wuhermann process (ซึ่งแสดงรายละเอียดใน รูปที่ 1) ระบบนี้เป็นระบบที่ Wuhermann ได้เสนอขึ้นในปี 1964 ซึ่งมีประสิทธิภาพต่ำและอัตราการกำจัดไนโตรเจนต่ำ

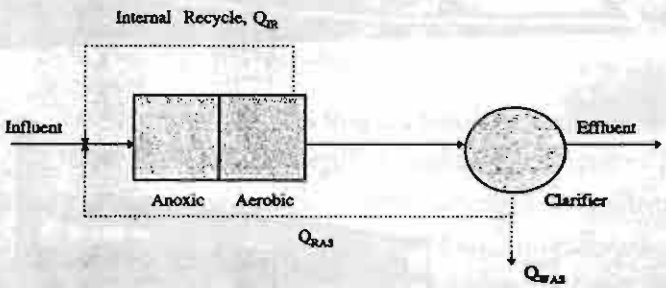
ข) ระบบสลัดจ์เดี่ยวแบบดีไนตริฟิเคชันเกิดก่อน (pre - denitrification)

ระบบสลัดจ์เดี่ยวแบบดีไนตริฟิเคชันเกิดก่อนจะสลับกับแบบ ก. คือมีถังแอน็อกซิกเป็นถังปฏิกรณ์ใบแรกและถังแอโรบิกเป็นถังปฏิกรณ์ใบหลัง(ดูรูปที่ 2) โดยที่จะเกิดกระบวนการดีไนตริฟิเคชัน เปลี่ยนรูปของไนเทรต ซึ่งได้จากการเวียนเอมแอลเอสเอสกลับจากถังแอโรบิกมาสู่ถังแอน็อกซิกให้เป็นก๊าซไนโตรเจน ทั้งนี้โดยใช้สารอินทรีย์ที่อยู่ในน้ำเสียที่เข้าสู่ถังแอน็อกซิกเป็นแหล่งของสารอินทรีย์คาร์บอน หลังจากผ่านถังแอน็อกซิกแล้ว สารอินทรีย์คาร์บอน(ที่เหลือจากกระบวนการดีไนตริฟิเคชัน)และสารอินทรีย์ไนโตรเจน (จากน้ำเสียดิบและที่ถูกเจือจางลงแล้ว ด้วยการเวียนเอมแอลเอสเอสกลับ)จะเข้าสู่ถังแอโรบิกซึ่งเป็นถังที่สอง ในถังนี้จะมีการเติมอากาศเพื่อให้เกิดกระบวนการกำจัดสารอินทรีย์คาร์บอนแบบใช้ออกซิเจนอิสระ ซึ่งเป็นภาวะที่กระบวนการไนตริฟิเคชันสามารถเกิดขึ้นได้ดี ในถังใบหลังนี้จะเกิดการแปลงรูปของสารอินทรีย์ไนโตรเจนให้อยู่ในรูปของไนเทรต เห็นได้ว่าในระบบนี้กระบวนการดีไนตริฟิเคชัน(ถังใบแรก)จะมีสารอินทรีย์คาร์บอนเพียงพอตลอดเวลาทำให้การกำจัดไนโตรเจนออกจากระบบสามารถกระทำได้อย่างมีประสิทธิภาพ ตัวอย่างของระบบนี้ได้แก่ กระบวนการ Ludzack - Ettinger (ดูรูปที่ 2) ซึ่งประสิทธิภาพการกำจัดไนโตรเจนของระบบนี้ค่อนข้างสูง อัตราการกำจัดไนโตรเจนเกิดขึ้นได้อย่างรวดเร็ว



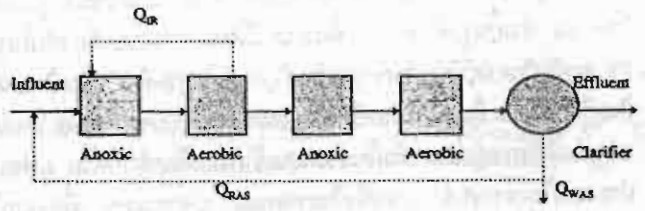
รูปที่ 2 กระบวน Ludzack - Ettinger สำหรับกำจัดไนโตรเจน (WEF manual of practice,1992)

ต่อมา Barnard ได้พัฒนากระบวนการของ Ludzack - Ettinger มาเป็นกระบวนการใหม่ที่เรียกว่า Modified Ludzack - Ettinger process ระบบนี้จะมีการเวียนเอมแอลเอสเอส (^{Q}IR) ภายในถังแอโรบิกตรงไปสู่ถังแอน็อกซิก ซึ่งทำให้มั่นใจมากยิ่งขึ้นว่าจะมีไนเทรตหลุดไปกับน้ำออกน้อยลง ลักษณะของระบบแสดงดังรูปที่ 3



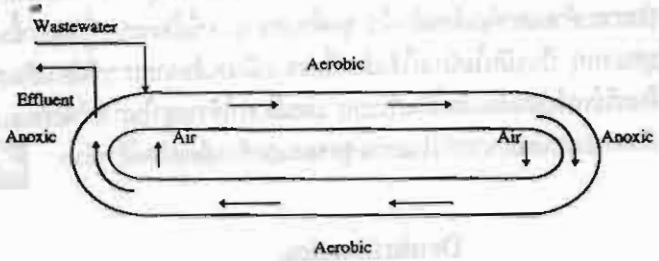
รูปที่ 3 กระบวนการ Modified Ludzack - Ettinger สำหรับกำจัดไนโตรเจน (WEF manual of practice,1992)

อีกกระบวนการหนึ่งในระบบนี้มีชื่อเรียกว่า Bardenpho process กระบวนการนี้ Barnard ได้พัฒนาขึ้นในปี1973(Randall,1992) ลักษณะของระบบประกอบด้วยกระบวนการแบบ Modified Ludzack- Ettinger และ Wuhermann โดยที่ระบบจะมีถังแอน็อกซิกและถังแอโรบิก จำนวน 2 ชุด ดังแสดงในรูปที่ 4 โดยชุดแรกจะมีการเวียนเอมแอลเอสเอสกลับ เช่นเดียวกับระบบ Modified Ludzack - Ettinger ถังแอน็อกซิกในชุดที่สองจะมีหน้าที่ในการลดปริมาณไนเทรตที่ยังคงมีเหลืออยู่ในน้ำเสียซึ่งออกมาจากถังแอโรบิกของชุดแรก และถังแอโรบิกในชุดที่สองยังทำหน้าที่ในการเป่าไล่ก๊าซไนโตรเจนให้แยกตัวออกจากน้ำเพื่อไม่ให้เกิดปัญหาการลอยตัวของสลัดจ์ในถังตกตะกอน



รูปที่ 4 กระบวนการ Bardenpho สำหรับกำจัดไนโตรเจน (WEF manual of practice,1992)

อนึ่ง ระบบคูวนเวียน (oxidation ditch) ก็สามารถใช้สำหรับกำจัดไนโตรเจนได้ กลไกการกำจัดไนโตรเจนโดยใช้ระบบคูวนเวียนนี้สามารถอธิบายได้จากทฤษฎีพื้นฐานเช่นเดียวกับระบบอื่น ๆ โดยการเกิดกระบวนการไนตริฟิเคชันและดีไนตริฟิเคชันจะเป็นไปตามรูปที่ 5

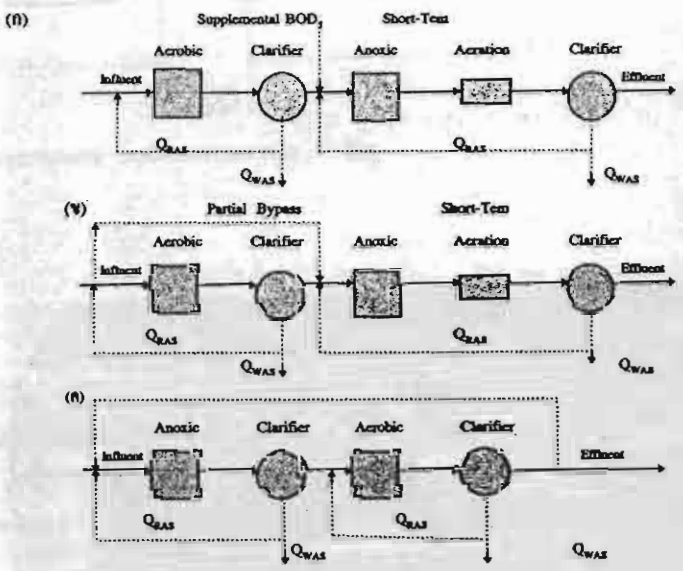


รูปที่ 5 กระบวนการคูวนเวียน สำหรับกำจัดไนโตรเจน (Sedlak,1991)

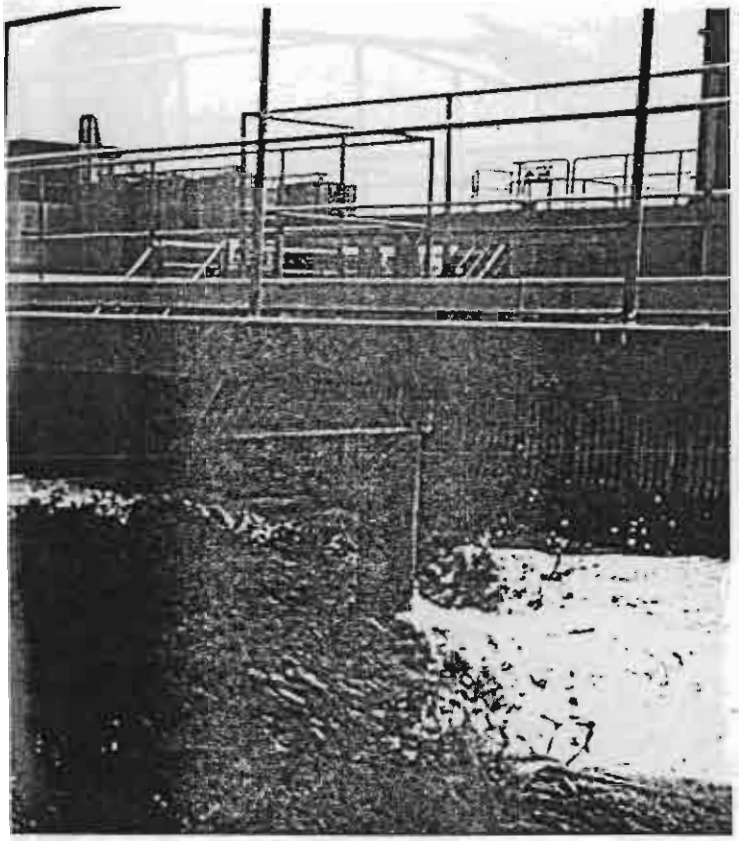
ระบบสองสลัดจ์

ระบบนี้ประกอบไปด้วยระบบแอกทิเวเต็ดสลัดจ์ 2 ชุด แต่ละชุดจะแยกหน้าที่ในการทำให้เกิดกระบวนการไนตริฟิเคชันและดีไนตริฟิเคชัน โดยแต่ละชุดจะต่อกันอย่างอนุกรมและมีถังตกตะกอนเป็นของตนเอง ในถังแอโรบิกนั้นจุลินทรีย์ที่อยู่ในถังจะได้แก่พวกเฮเทอโรโทรฟิก และออโทโทรฟิก (ใช้คาร์บอนไดออกไซด์เป็นแหล่งคาร์บอนและได้พลังงานจากปฏิกิริยาเคมี)ซึ่งมีบทบาทในการเกิดปฏิกิริยาไนตริฟิเคชัน จุลินทรีย์ทั้งสองชนิดสามารถอยู่ร่วมในถังเดียวกันได้ เนื่องจากจุลินทรีย์พวกแฟคัลเททีฟใช้สารอินทรีย์คาร์บอนในน้ำเสียเป็นแหล่งคาร์บอน ส่วนพวกออโทโทรฟิกใช้คาร์บอนจากคาร์บอนไดออกไซด์ในน้ำเสีย ฉะนั้นจึงไม่แย่งอาหารกัน ส่วนในถังแอน็อกซิกนั้นจุลินทรีย์ที่อาศัยอยู่จะเป็นพวกแฟคัลเททีฟที่ใช้กำจัดสารอินทรีย์คาร์บอนเหมือนในถังแอโรบิก ต่างกันตรงที่ในถังนี้ไนเทรตจะเป็นตัวรับอิเลคตรอนแทนออกซิเจนเท่านั้น

ตัวอย่างรูปแบบของระบบสองสลัดจ์แสดงไว้ดังรูปที่ 6 จากรูปที่ 6 (ก) ถังแอโรบิกจะทำหน้าที่กำจัดสารอินทรีย์คาร์บอน และเกิดกระบวนการไนตริฟิเคชัน จากนั้นน้ำที่ออกจากถังตกตะกอนของระบบแอโรบิก ซึ่งมี



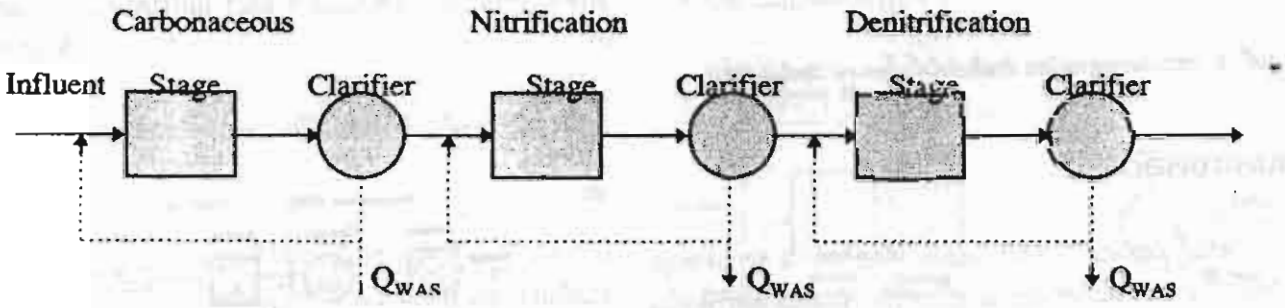
รูปที่ 6 กระบวนการสองสลัดจ์ สำหรับกำจัดไนโตรเจน (WEF manual of practice,1992)



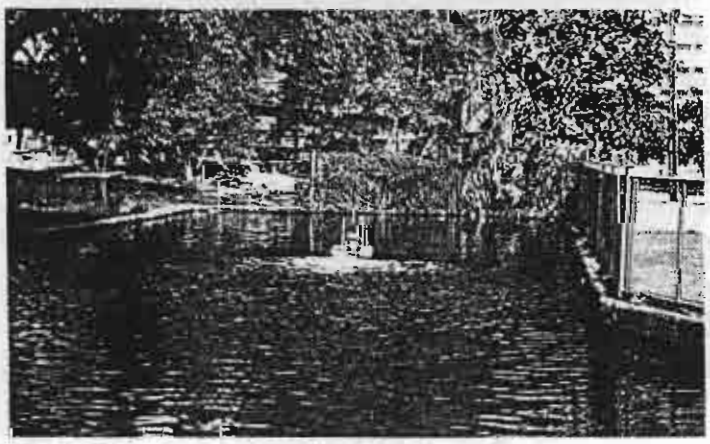
ไนเทรตอยู่ด้วยนั้น จะได้รับการเติมสารอินทรีย์คาร์บอนก่อนจะเข้าสู่ถังแอน็อกซิกต่อไป ส่วนในรูปที่ 6 (ข) นั้นระบบนี้ใช้หลักการแบบเดียวกับในรูปที่ 6 (ก) จะต่างกันตรงที่มีน้ำเสียเข้าบางส่วนที่ถูกแบ่งไปสู่ชุดที่ 2 เพื่อเป็นแหล่งสารอินทรีย์คาร์บอน ถึงแม้สารอินทรีย์คาร์บอนภายในระบบจะเพียงพอ โดยไม่ต้องหาจากแหล่งภายนอกมาเพิ่มในช่วงก่อนเข้าถังแอน็อกซิกก็ตาม แต่จะมีสารประกอบไนโตรเจนจำนวนหนึ่งที่จะหลุดออกไปจากระบบ เนื่องจากในถังแอน็อกซิกไม่สามารถออกซิไดส์แอมโมเนียได้ และในกรณีของรูปที่ 6 (ค) จะมีการ หมุนเวียนน้ำจากถังตกตะกอนชุดที่ 2 สู่ถังแอน็อกซิก เพื่อเป็นการนำไนเทรตที่ยังคงมีเหลืออยู่ในน้ำมาทำปฏิกิริยาในถังแอน็อกซิกอีกครั้ง ทำให้น้ำทิ้งมีไนโตรเจนลดลงและยังเป็นการช่วยลดปริมาณความต้องการออกซิเจนในถังแอโรบิกด้วย เนื่องจากมีออกซิเจนจากไนเทรตมาเป็นตัวเสริมในการกำจัดสารอินทรีย์คาร์บอนในถังแอน็อกซิกได้มากขึ้น

ระบบสามสลัดจ์

ระบบนี้เป็นการแยกกระบวนการกำจัดสารอินทรีย์คาร์บอน ไนตริฟิเคชัน และดีไนตริฟิเคชันออกจากกัน ดังแสดงในรูปที่ 7 โดยระบบจะประกอบด้วยแอกทิเวเต็ดสลัดจ์ 3 ชุดด้วยกัน ระบบนี้เป็นระบบที่พัฒนาขึ้นในยุคแรกๆ ปัจจุบันไม่นิยมใช้แล้วเนื่องจากมีผลเสียจากการที่ต้องเติมสารอินทรีย์คาร์บอนเพิ่มเติมให้แก่ระบบ และเสียค่าใช้จ่ายสูงในการที่จะต้องแยกถังตกตะกอนออกจากกันในแต่ละชุดของแอกทิเวเต็ดสลัดจ์อีกด้วย



รูปที่ 7 กระบวนการสามสลัดจ์ สำหรับกำจัดไนโตรเจน (WEF manual of practice,1992)



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3. การเสนอผลงานในที่ประชุมวิชาการนานาชาติ

presented paper

MASS BALANCE OF SOLIDS IN THE UPFLOW PELLETIZATION PROCESS TREATING 30-240 NTU RAW WATER

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ABSTRACT

From the study of an upflow pelletization process treating raw water from the Chao Phraya river was iniated at 2 conditions, i.e., a) during rainy season from September to November 1996, the turbidity of 100 to 240 NTU, upflow velocity of 6 and 9.6 m/hr, using alum as a coagulant at 18.73, 21.13, 23.44, and 25.79 mg/l coupled with nonionic polymer of 0.3 mg/l as a coagulant aid, and nonionic polymer of 0.1, 0.2 and 0.3 mg/l as an alternative coagulant, and b) during dry season from January to February 1997, the turbidity of 30 to 60 NTU, upflow velocity of 9.6 and 15 m/hr, using alum as a coagulant at 3, 5, 7 and 10 mg/l coupled with nonionic polymer of 0.3 mg/l as coagulant aid, and nonionic polymer of 0.1, 0.2 and 0.3 mg/l as an alternative coagulant, It was found that the solids or pellet mass in the reactor was continuously increase from the start-up until reaching steady state condition. The mass of the input solids to the reactor was more than its output counterpart until the process reached the steady state where the solids mass input was approximately equal to that in the output. The accumulation of pellet in the reactor was found to decrease till the solid mass difference in the process became zero. Meanwhile, the amount of alum required was not related to the increase in pellet mass but seemed to relate to mass of feed solids or raw water turbidity. In addition, it was apparent that the accumulation of pellet in the reactor with high turbidity influent was more than its counterpart in the low turbidity case.

KEYWORDS

Pelletization, Turbidity removal, Pellet mass

INTRODUCTION

One of the essential process in most water supply treatment system is coagulation of which the colloids are destabilized in rapids mixing unit by the addition of coagulant (commonly alum and iron compounds. The destabilized particle was then contacted and developed floc in slow

mixing unit and was removed by sedimentation. After that, process water is filtrated, disinfected and kept in clear water storage for later distribution. Each of the water treatment processes generally consume a cosiderably long perriod of time. As a result, several new techniques including upflow pelletizationn process have been developed to reduce times in treating water which exceeds turbidity standard. Pelletization consists of two main processes: colloid destabilization in rapid mixing unit and forming pellets from destabilized colloids in upflow pelletization reactor. The shape of pellet is sphere which is more dense than floc of conventional coagulation and flocculation and endurable to water turbulence very well. Smaller pellets will be in the upper part while larger pellets will be in the lower part of reactor. These large pellets will be discreted into small pellets and flow up by shear force of turbines and continuous turbulence along the length of reactor. Then, they will be caught by another pellet to form larger pellet and will sink down again. It will follow this cycle repeatedly.

The objective of this research is to study the mass balance of solids in upflow pelletization system treated raw water turbidity in the range of 30-240 NTU. The results from this study will be applied to large scale water treatment plant.

EXPERIMENTAL RESULTS AND DISCUSSION

This reseach was initiated by installing an upflow pelletization reactor (see Figure 1) at Bangkaen water treatment plant. Raw water was taken from the Chao Phraya river during September 1996 to Febuary 1997, experiment was seperated into 2 conditions, i.e., a) during rainy season period from September to November 1996, the turbidity of 100 to 240 NTU, upflow velocity of 6 and 9.6 m/hr, using alum as coagulant at 18.73, 21.13, 23.44, and 25.79 mg/l coupled with nonionic polymer of 0.3 mg/l as coagulant aid, and nonionic polymer of 0.1, 0.2, and 0.3 mg/l as an alternative coagulant ; and b) during dry season peroid from January to February 1997, the turbidity of 30 to 60 NTU, upflow velocity of 9.6 and 15 m/hr, using alum as coagulant at 3, 5, 7, and 10 mg/l coupled with nonionic polymer of 0.3 mg/l as coagulant aid, and nonionic polymer of 0.1, 0.2, and 0.3 mg/l as an alternative coagulant. The operation interval in each run was 84 hours and water samples were drawn very 6 hours. The results are shown in Tables 1 and 2 and Figures 2 to 5.

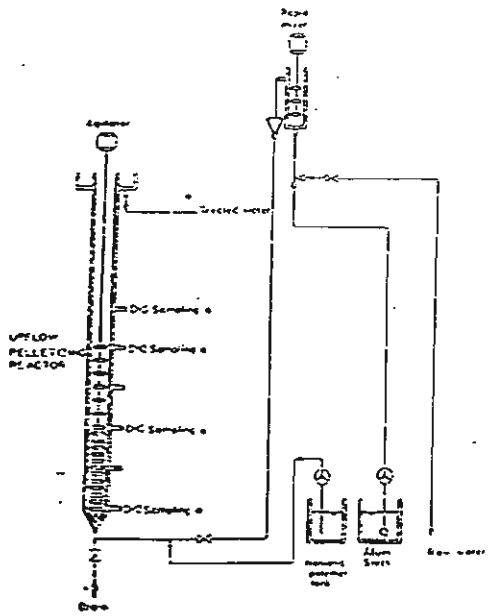

Figure 1 : Upflow pelletization reactor (● sampling point)

Table 1 : Raw water and effluent water turbidity and pellet mass in reactor (during rainy season).

	V_{up} 9.6 m/hr			V_{up} 6 m/hr		
Alum and/or PE (mg/l)	Raw Water (NTU)	Eff. Water (NTU)	Pellet Mass (g/cm^3-hr)	Raw Water (NTU)	Eff. Water (NTU)	Pellet Mass (g/cm^3-hr)
A0+P.1	179.7	18.22	5.68	179.7	10.38	6.97
A0+P.2	192.9	8.33	6.02	192.9	8.16	6.91
A0+P.3	143.2	9	4.67	143.2	7.11	6.29
A18.73+P.3	163.2	4.88	6.11	99.0	3.86	7.50
A21.13+P.3	153.9	4.45	6.02	131.6	4.12	8.60
A23.44+P.3	185.2	4.23	7.28	105.6	3.97	7.38
A25.79+P.3	184.9	3.71	7.13	178.1	3.45	11.09

Note ; A0+P.1 = alum 0 mg/l and nonionic polymer (PE) 0.1 mg/l

Table 2 : Raw water and effluent water turbidity and pellet mass in reactor (during dry season).

	V_{up} 15 m/hr			V_{up} 9.6 m/hr		
Alum and/or PE (mg/l)	Raw Water (NTU)	Eff. Water (NTU)	Pellet Mass (g/cm^3-hr)	Raw Water (NTU)	Eff. Water (NTU)	Pellet Mass (g/cm^3-hr)
A0+P.1	41.5	7.92	6.62	44.7	10.52	10.62
A0+P.2	36.8	8.35	6.20	45.9	8.03	11.16
A0+P.3	41.1	7.55	7.76	45.9	7.63	12.39
A3+P.3	45.9	6.63	6.96	44.7	6.99	10.05
A5+P.3	36.8	3.45	7.31	41.5	2.53	10.31
A7+P.3	36.8	3.04	6.86	44.7	2.70	10.64
A10+P.3	41.1	2.35	7.86	41.5	2.35	10.81

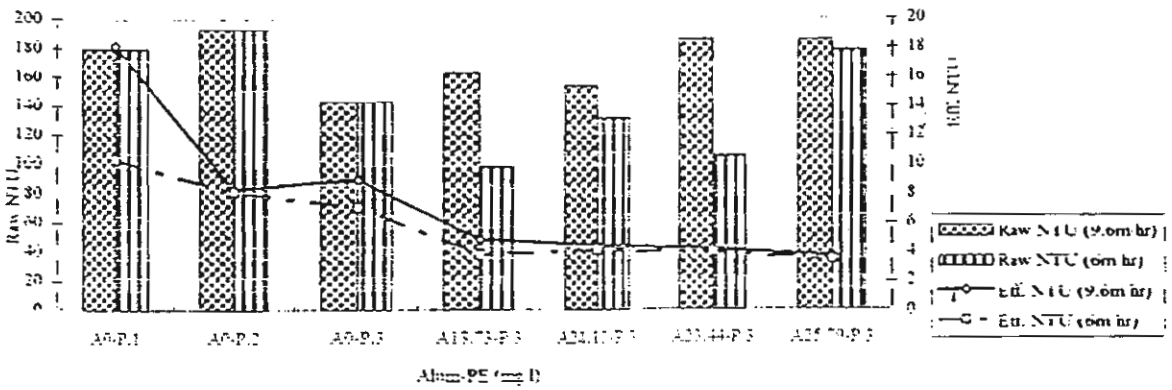

Figure 2 : Raw water and effluent water turbidity during rainy season with upflow velocity of 6 and 9.6 m/hr

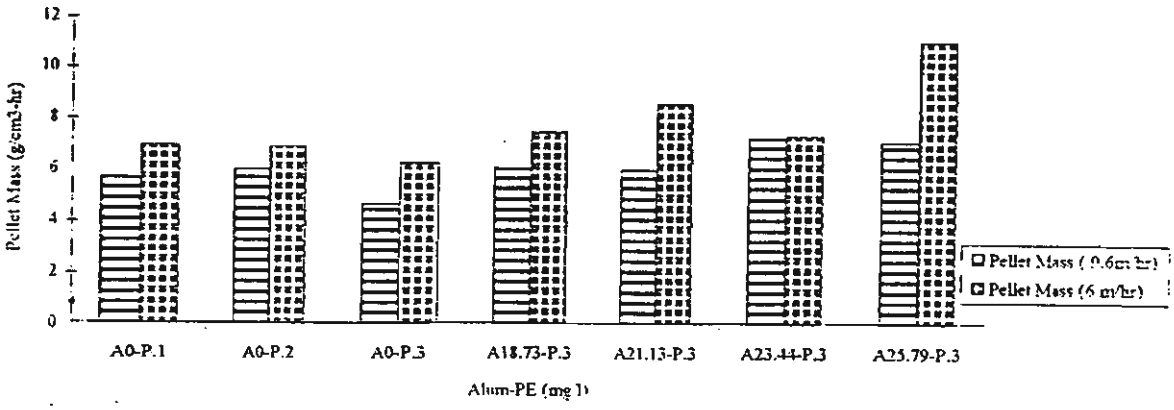

Figure 3 : Pellet mass in reactor during rainy season with upflow velocity of 6 and 9.6 m/hr

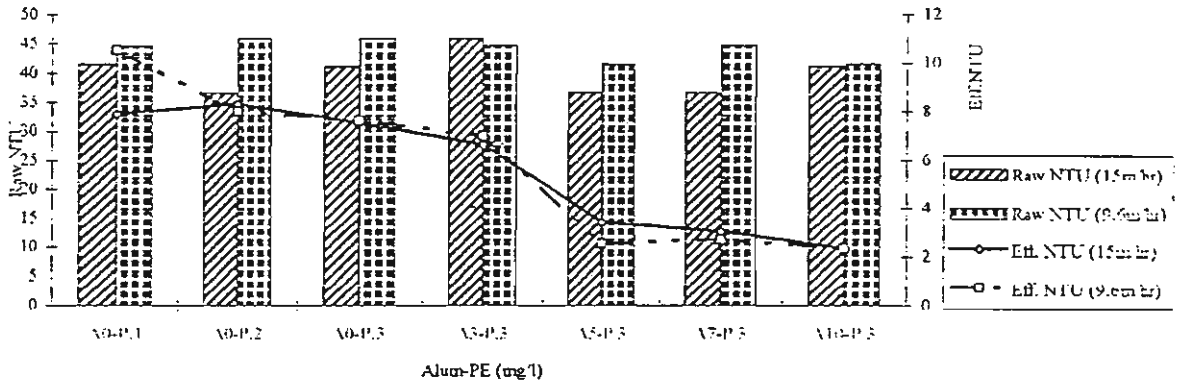

Figure 4 : Raw water and effluent water turbidity during dry season with upflow velocity of 9.6 and 15 m/hr

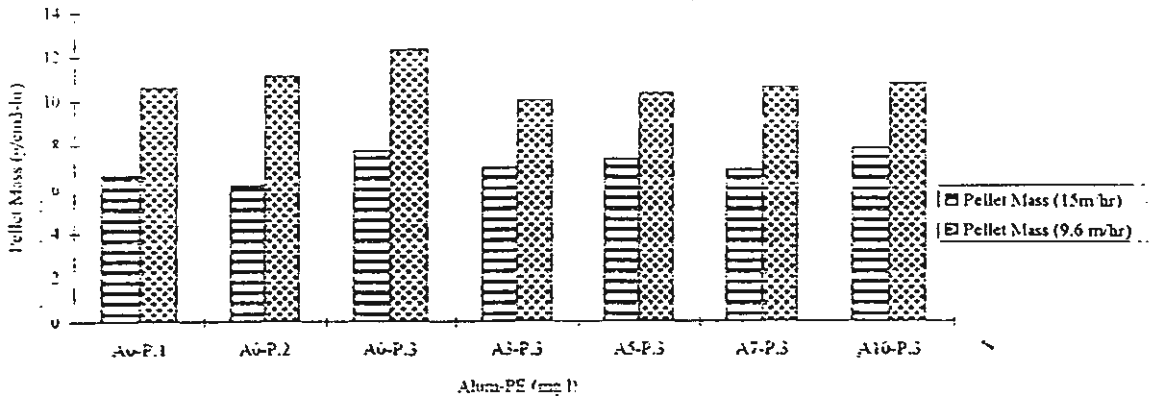

Figure 5 : Pellet mass in reactor during dry season with upflow velocity of 9.6 and 15 m/hr

After 84 hours of continuous operation for each run, the solids mass at steady state of condition (a) with the upflow velocity of 6 m/hr were 7.50, 8.60, 7.38 and 11.09 g/m^3-hr whereas were 6.11, 6.02, 7.28 and 7.13 g/m^3-hr for the upflow velocity 9.6 m/hr for 18.73, 21.13, 23.44, and 25.79 mg/l of alum, respectively. In the case of without alum, the solids mass at the steady state with upflow velocity of 6 m/hr was 6.97, 6.91 and 6.29 g/m^3-hr for 0.1, 0.2 and 0.3 mg/l of nonionic polymer, respectively, while the corresponding figure was 5.68, 6.02 and 4.67 g/m^3-hr in the 9.6 m/hr upflow velocity condition. In condition (b) with the upflow velocity of 9.6 m/hr the figure was 10.05, 10.31, 10.64 and 10.81 g/m^3-hr for the four doses of alum, respectively whereas the corresponding figure was 6.96, 7.31, 6.86 and 7.86 g/m^3-hr for the upflow velocity 15 m/hr case. In the case of without alum, the solids mass at the steady state with upflow velocity of 9.6 m/hr was 10.62, 11.16 and 12.39 g/m^3-hr, respectively, whereas the corresponding figure was 6.62, 6.20 and 7.76 g/m^3-hr in the 15 m/hr upflow velocity condition.

CONCLUSION

The solids or pellet mass in the reactor was found to continuously increase from the start-up until the steady state peroid input solid mass of the reactor was more than its output counterpart until the process reached the steady state where the solids mass input was approximately equal to that in the output. The accumulation of pellet in the reactor was shown to decrease till the solid mass difference in the process became zero. Meanwhile, the alum requirement was not related to the increase of pellet mass but seemed to relate to mass of feed solids due to raw water turbidity. In case of high turbidity raw water, it was apparent that the accumulation of pellet in the reactor was more than its counterpart in the low-turbidity raw water scenario.

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presented paper

POTENTIAL OF DECREASING SLUDGE DRYING TIME BY PREWITHDRAWAL OF SUPERNATANT FROM SLUDGE DRYING BEDS

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ABSTRACT

The sludge dewatering was investigated on 570 mm ∅ × 900 mm H pilot-scale drying beds, with 3 layers of filtering media, namely, 10-15 mm ∅ gravel of 100 mm height, 3-5 mm ∅ coarse sand of 100 mm height, and 0.36 mm ES with 1.48 UC sand of 350 mm height for the lower, intermediate, and upper layers, respectively. The applied volume of sludge was 76.5 litres which was equivalent to 300 mm of the sludge initial height on the beds. Tested sludge samples were collected from different types of industries and pre-prepared into four different concentrations by thickening or dilution. Two experimental sets were initiated in this study, i.e. without treatment(control units) and with the withdrawal of supernatant after sludge settling on the sludge drying beds(test units)

It was observed that the tested sludge with good settling property yielded a considerable amount of supernatant which could be siphoned off. As a result, it was dryer than that without this property. However, the sludge in the control unit (without pre-withdrawal of supernatant) dewatered readily in the next few days and its final solid concentration was closed to that of the test unit.

KEYWORDS

dewatering
drying beds
pre-withdrawal
sludge drying time
supernatant

INTRODUCTION

Drying beds are physical operation units used to reduce the moisture content of the sludge. Water is released from sludge by drainage and evaporation. It normally takes a fews days until the sand becomes clogged with fine particles or until all the free water has been drained away. Once the supernatant has formed, it tries to remove this part of water that accummulated on the surface of sludge layers comparative with others that allowed supernatant to acccumulate on the surface. It is possible to reduce time for drying sludge or reduce land for dewatering sludge .

THEORY

Sludge Drying Beds

Sand drying beds is one of the most economical methods of dewatering sludge if adequate land area is available and climatic conditions are favorable (Teerawat luangurai, 1990). Moisture could be removed from sludge by 2 means (Eckenfelder, 1966 quoted in Khomsilp kaewtungkhaun, 1983)
1). Gravity drainage of free water
2). Evaporation of moisture depending on solar radiation, wind and relative humidity of air in the area of drying beds (USEPA., 1987)

Mechanisms of Sludge Dewatering on Sand Beds

Dewatering on sand beds occurs by two mechanisms: filtration of water through the subsurface underdrain, and evaporation of water due to radiation and convection. (Adam et al , 1981 quoted in Teerawat luangurai, 1990). The rate of air drying is usually slower and is related to temperature, relative humidity, and wind velocity .(Teerawat, 1990). Figure 1 shows drying phases of sludge.

During the constant rate period, the sludge surface is wetted and the rate of evaporatin is relatively independent of the nature of sludge. When the critical moisture content is reached water no longer migrates to the surface of the sludge as rapidly as it evaporates, hence, the falling rate period occurs. The rate of drying during this period is related to the thickness of the sludge, its physical and chemical properties, and ambient conditions. Subsurface drying continues until an equilibrium moisture content is obtained. (Teerawat luangurai, 1990)

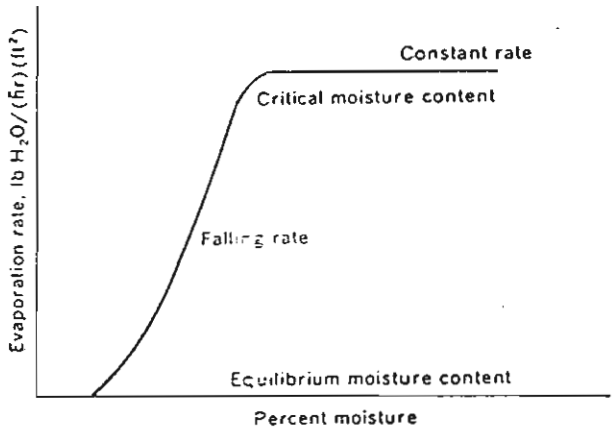

Fig 1. Drying phases for a sludge

RESULT AND DISCUSSION

1) Effect of pre-withdrawal on sludge drying time.
In this study, eigth dewatering runs with sand drying beds were conducted at different initial sludge concentrations. Figure 2 to 9 show the percents of solids in sludge during the experimental period. It can be seen that the drying rate for the test units (with pre- withdrawal of supernatant) was less than those of control units (without pre-withdrawal of supernatant) only a short time on the initial few days. Control unit could drained off water including existing supernatant readily within few days and the solids percentage in sludge were close to the test units.

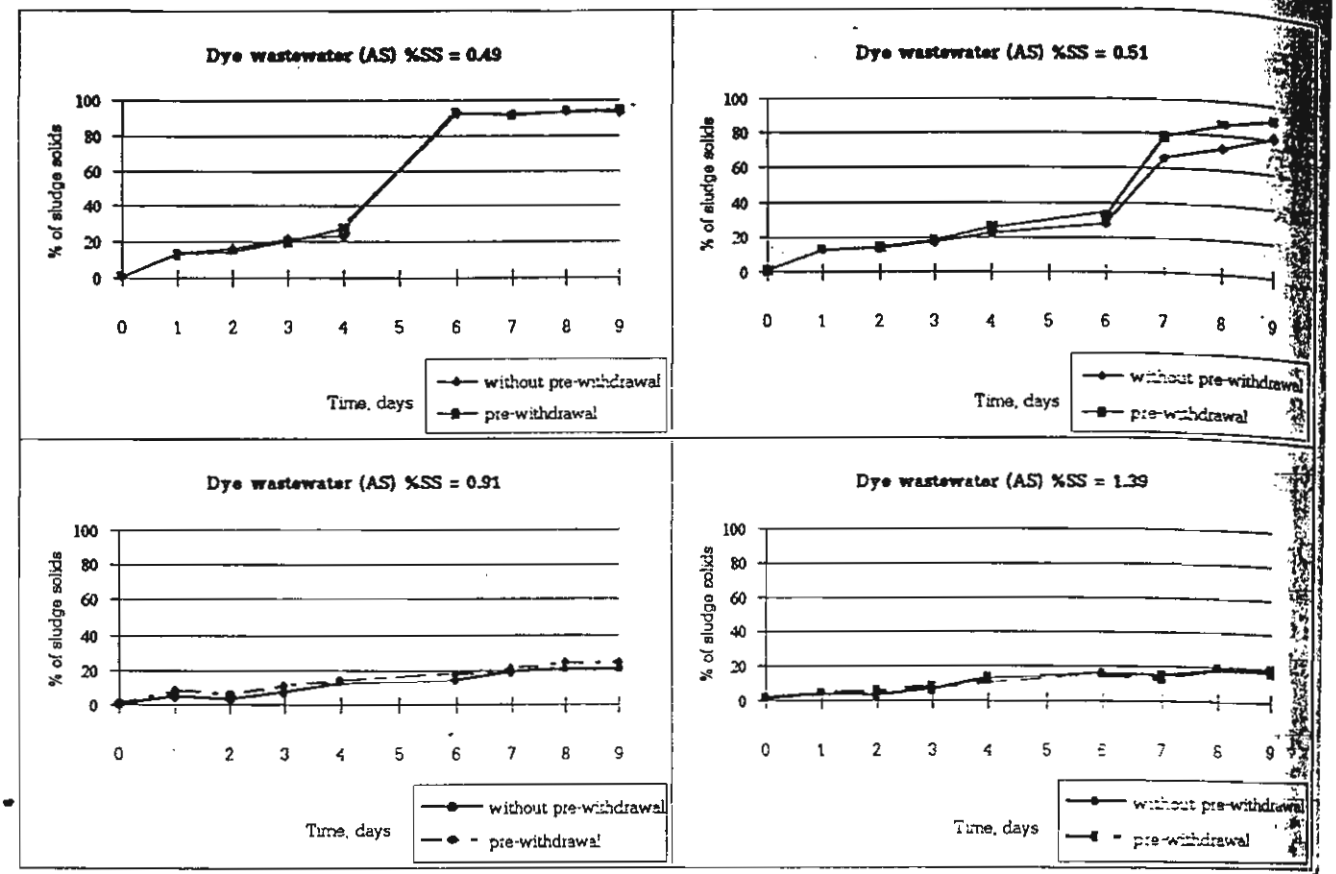

Fig 2 % of sludge solids vs. times (Dye wastewater/AS)

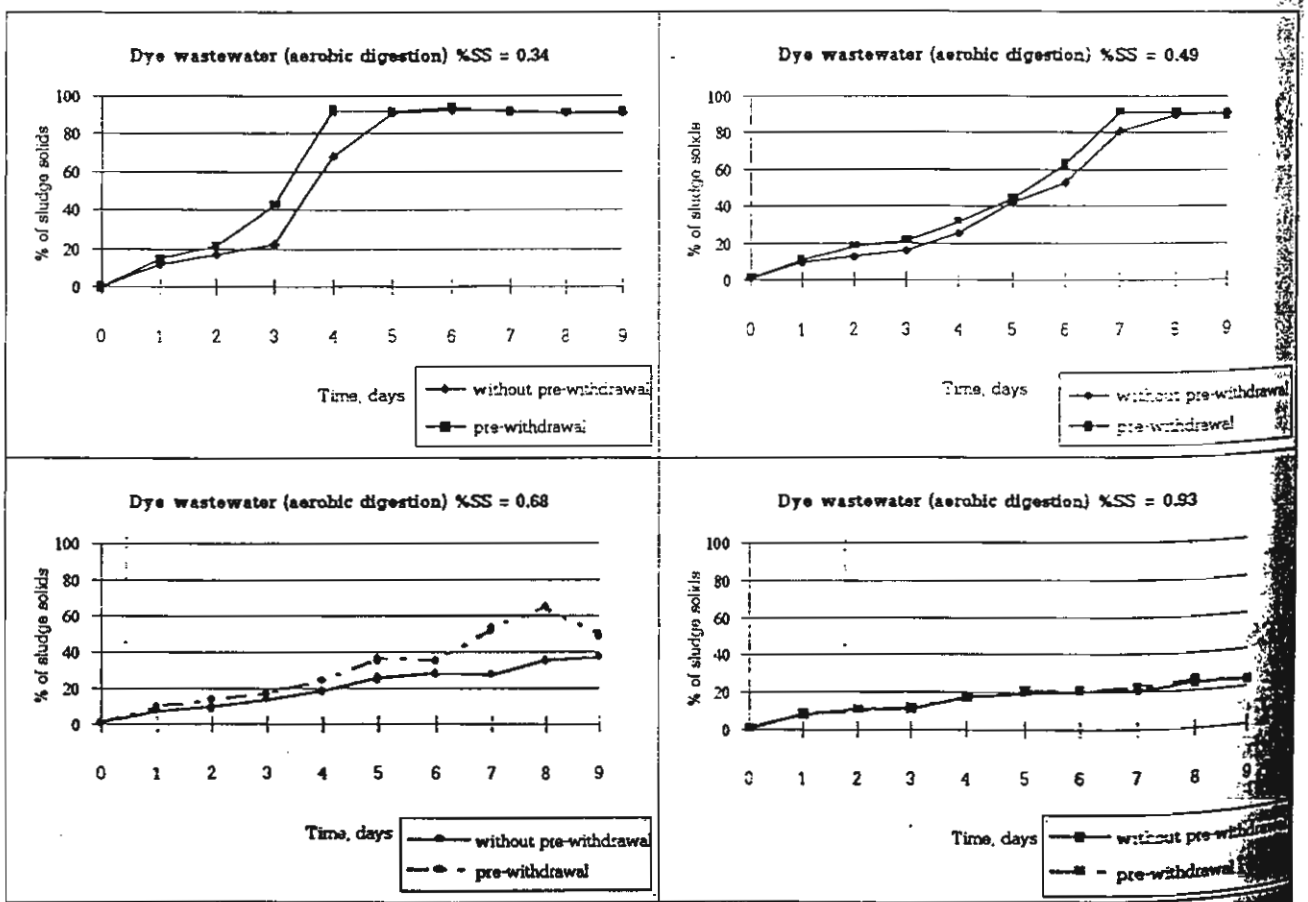

Fig 3 % of sludge solids vs. times (Dye wastewater/digestion)

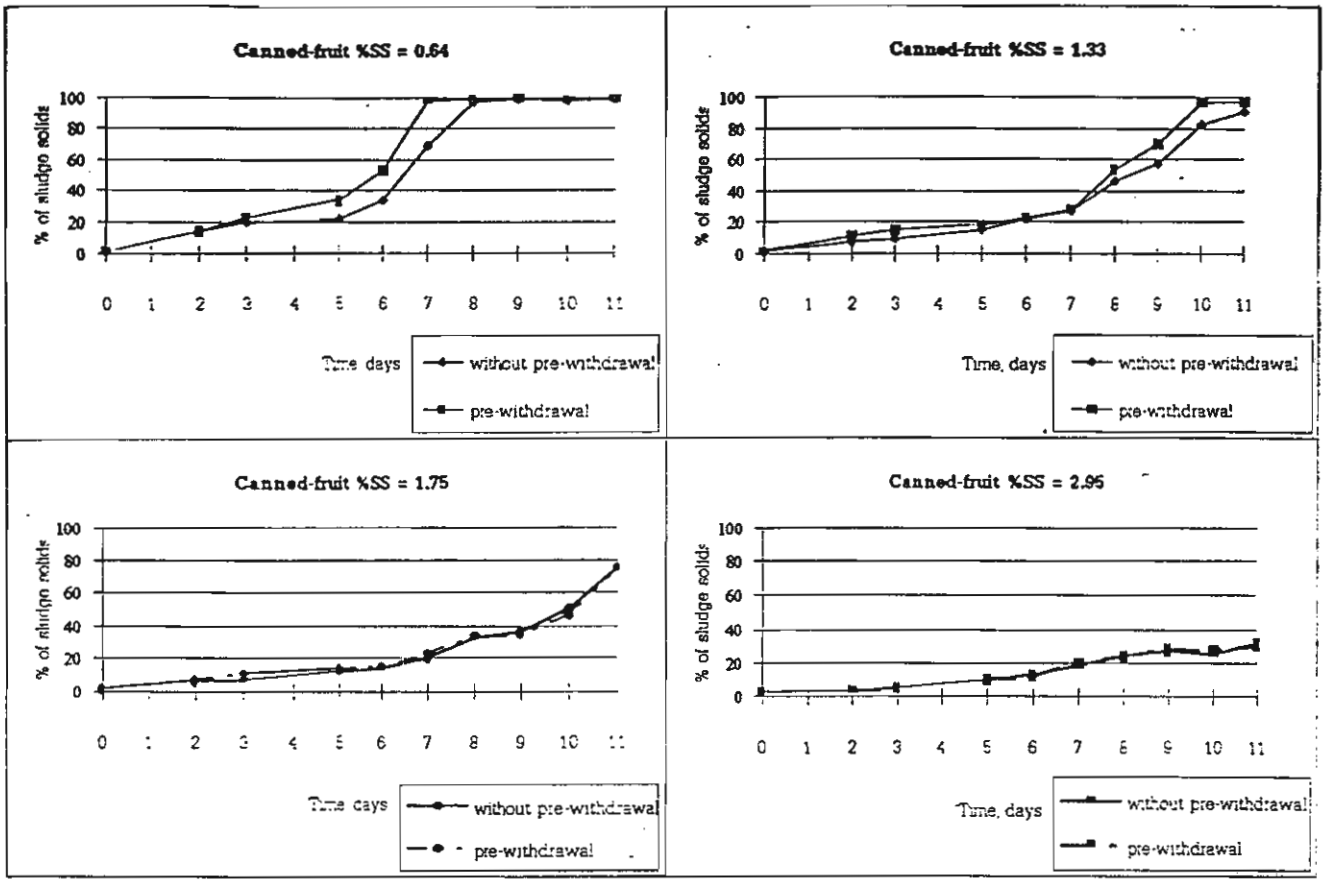

Fig 4 % of sludge solids vs. times (canned-fruit)

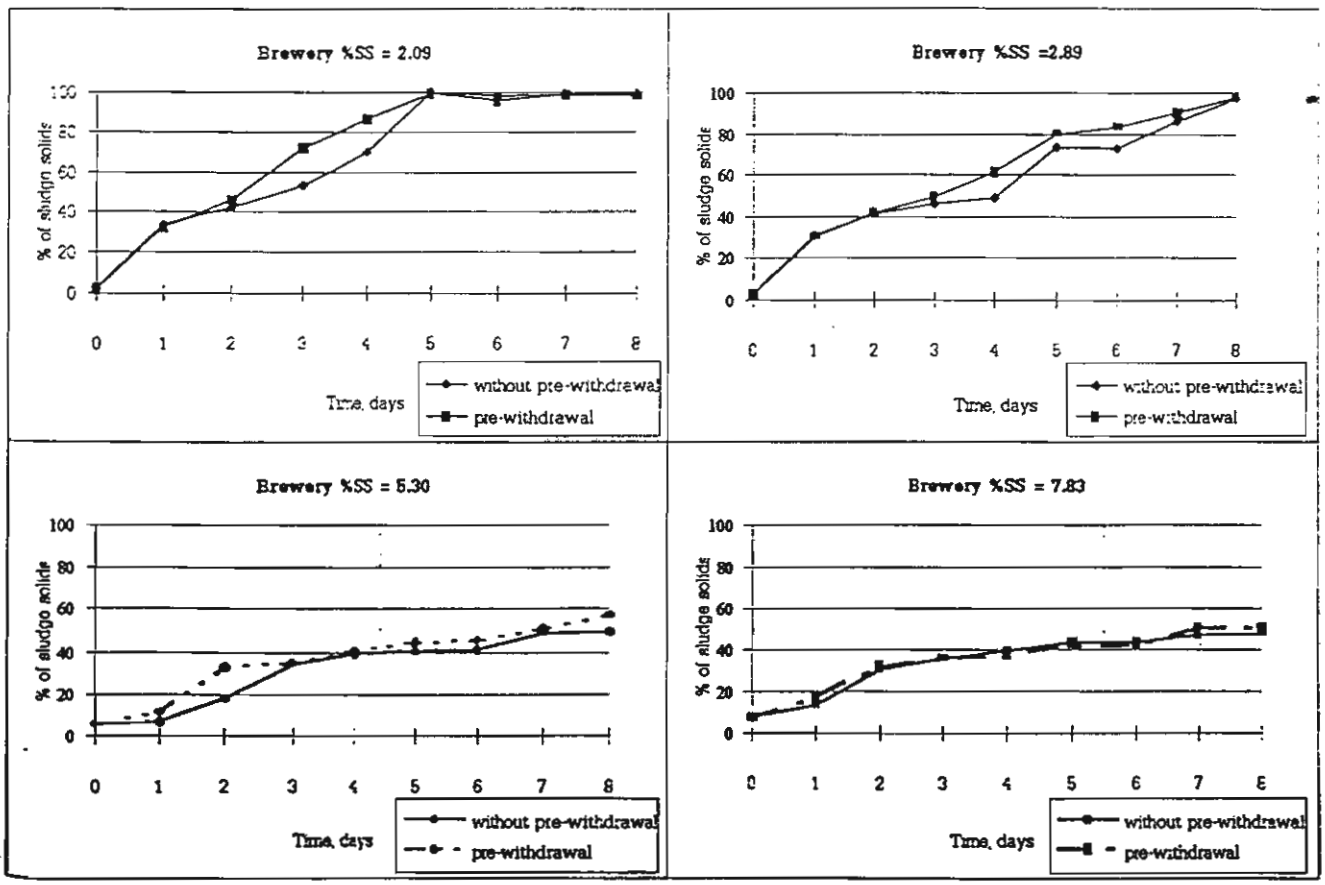

Fig 5 % of sludge solids vs. times (Brewery)

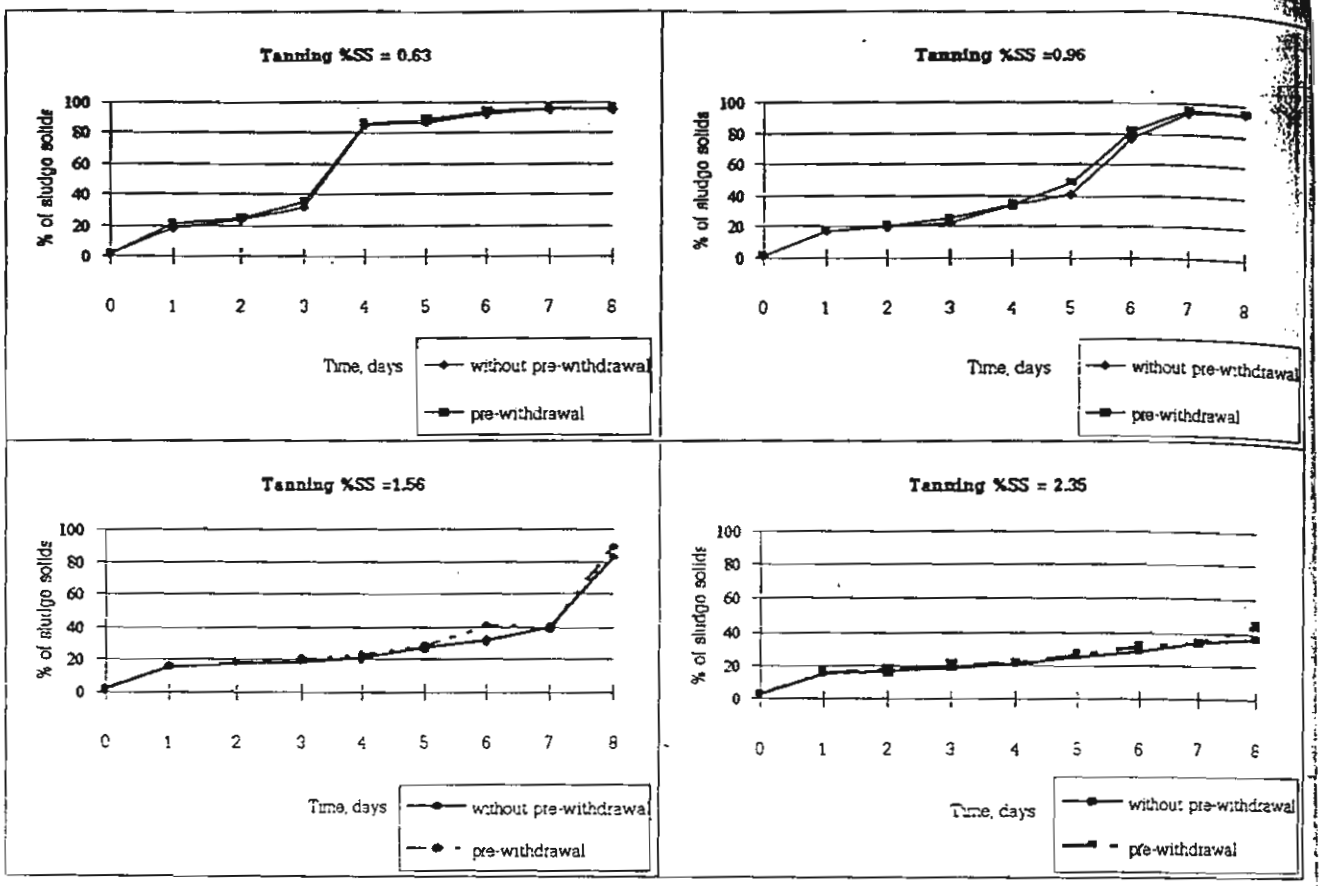

Fig 6 % of sludge solids vs. times (Tanning)

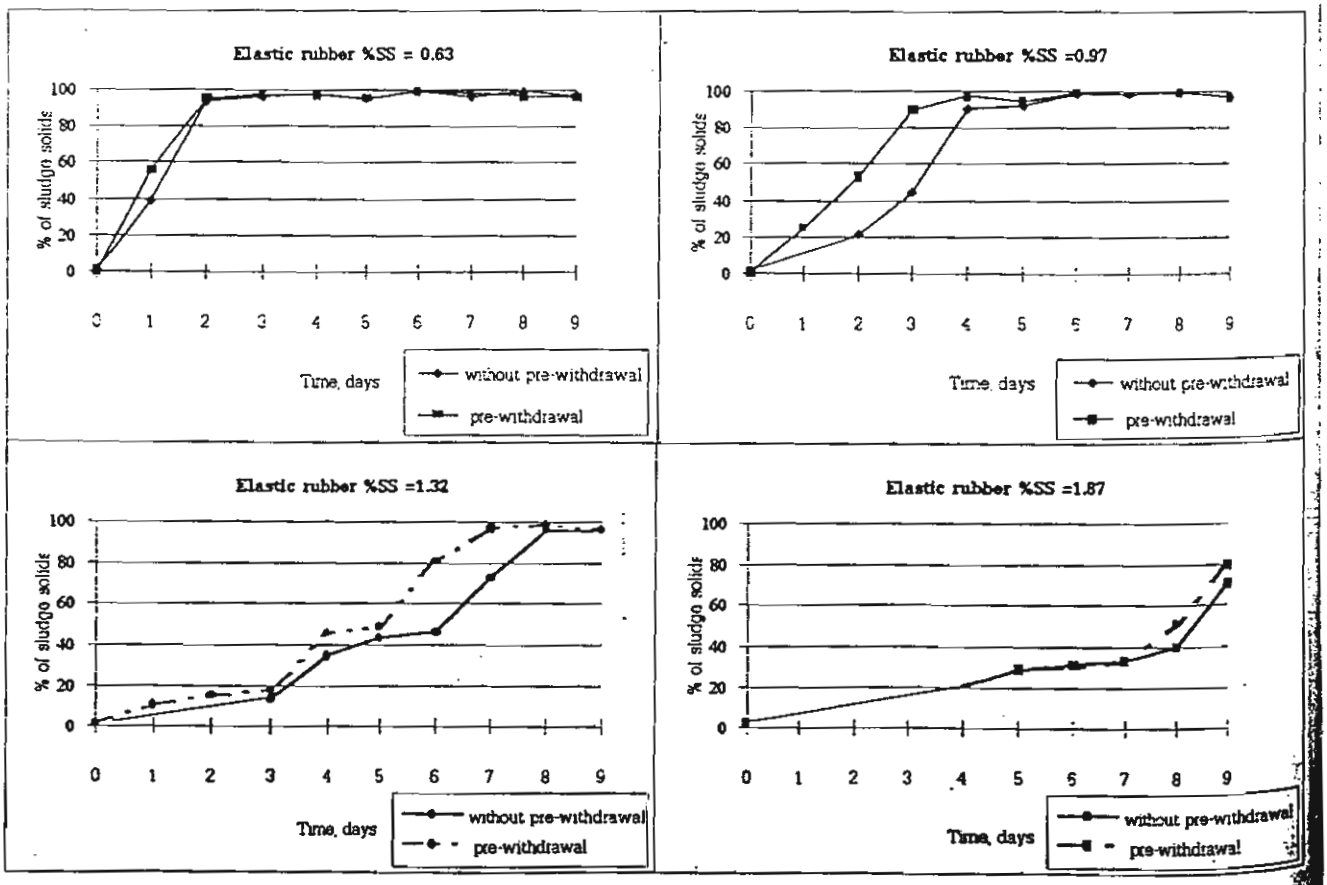

Fig 7 % of sludge solids vs. times (Elastic rubber)

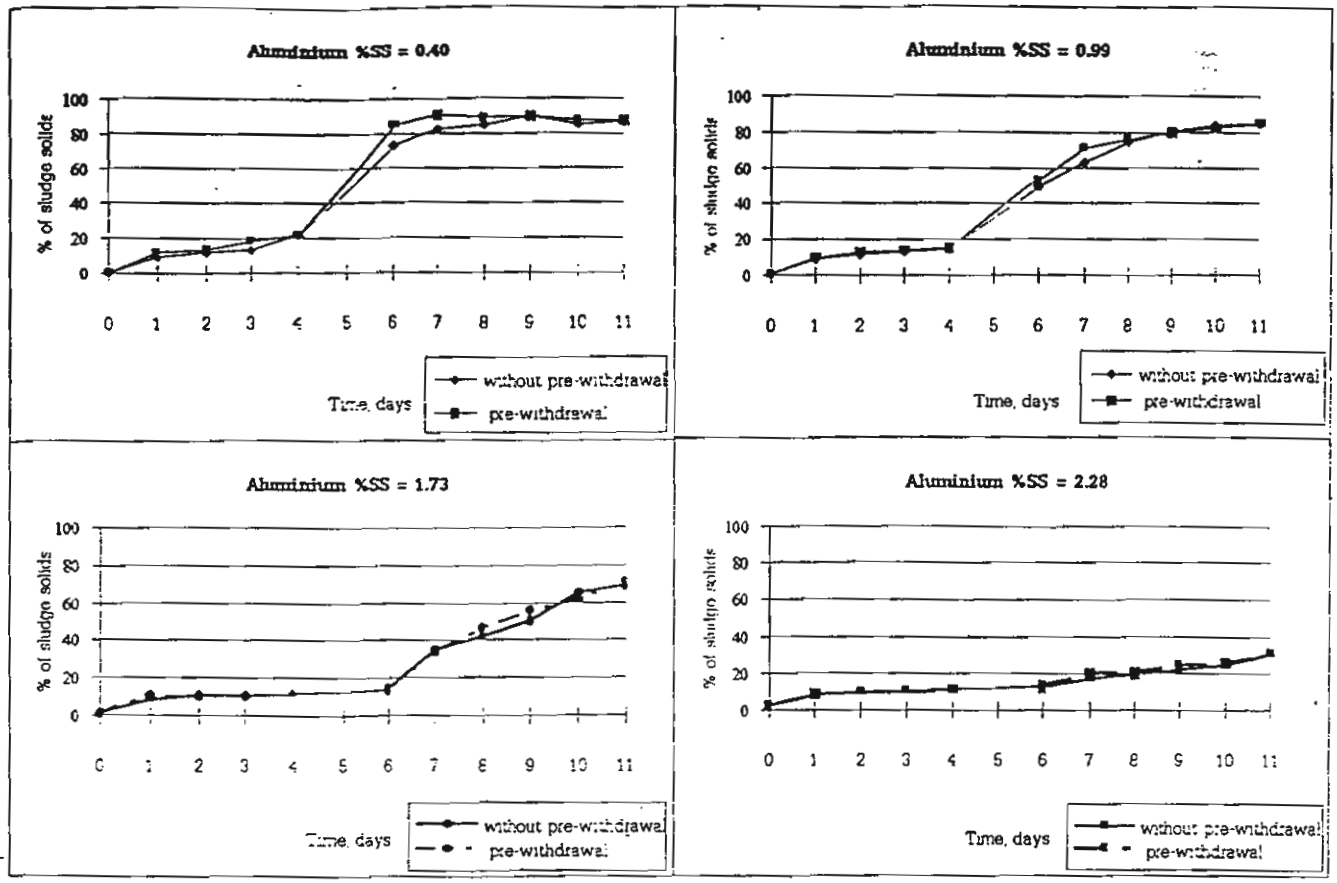

Fig 8 % of sludge solids vs. times (Aluminium)

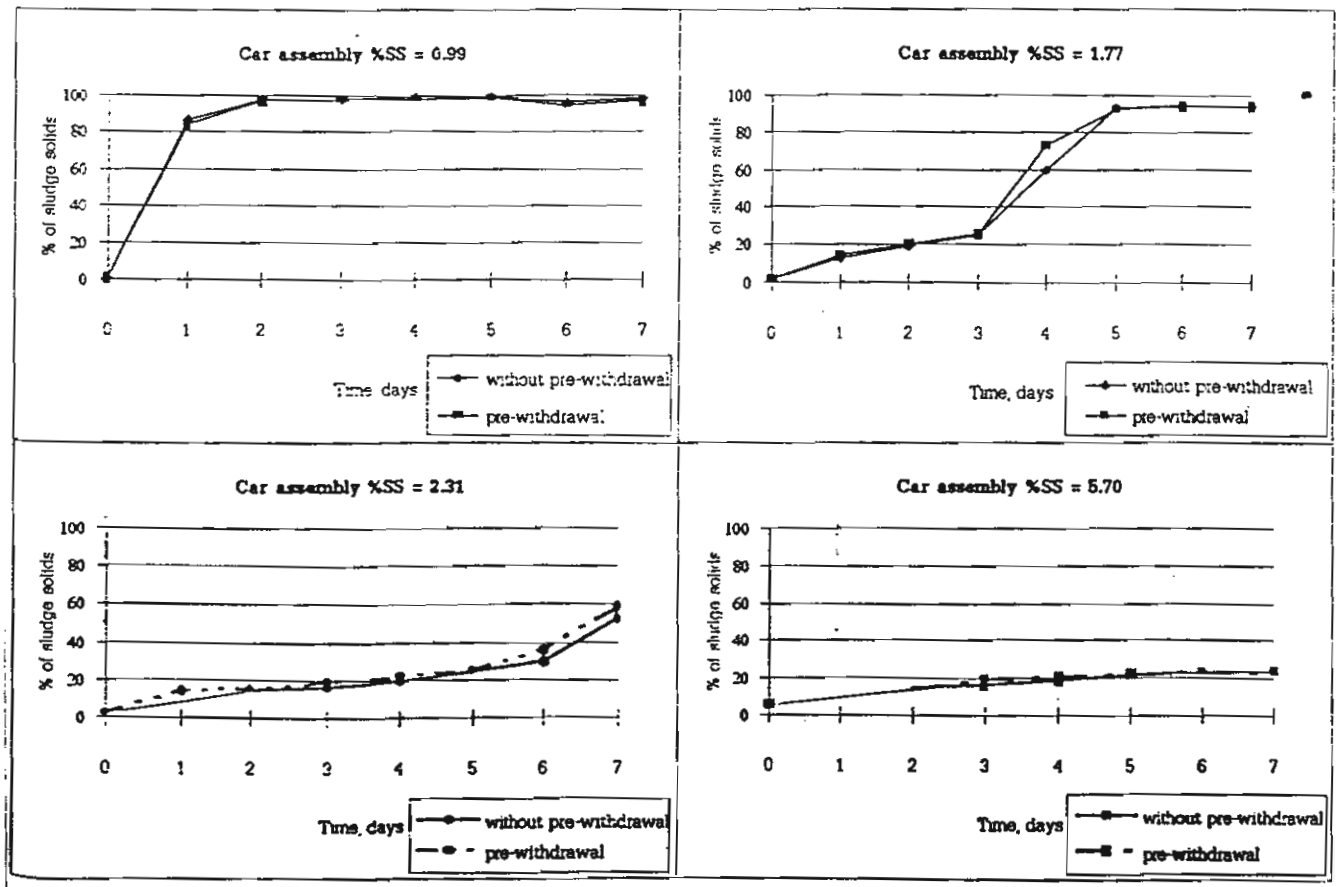

Fig 9 % of sludge solids vs. times (Car assembly)

2). Effect of sludge solid concentrations to the volume of siphonable supernatant.

Table 1 shows the volume of supernatant that was siphoned off as a function of sludge type and percent of sludge solids. It can be seen that the amount of supernatant was inverse proportional to the solids concentrations of the sludge. Low solid concentration sludge yield more supernatant than higher ones. However, if sludge contained very low solid concentrations, the water was drained so rapidly that supernatant could not be discharged in time (as marked with asterisk in table 1)

Table 1 volume of supernatant vary with type of sludge and different % of sludge solids.

Sludge source	Type of sludge	initial solid concentration (%)	volume of supernatant, litre	
			without pre-withdrawal	pre-withdrawal
1) dye wastewater	AS/ bio sludge	0.49	0.0	1.04
		0.51	0.0	18.30
		0.91	0.0	4.90
		1.36	0.0	0.00
2) dye wastewater	aerobic digestion/ bio sludge	0.34	0.0	22.00
		0.49	0.0	15.10
		0.68	0.0	8.70
		0.93	0.0	0.00
3) canned-fruit	AS/ bio sludge	0.64	0.0	39.60
		1.33	0.0	22.50
		1.75	0.0	5.80
		2.95	0.0	0.00
4) brewery	UASB/bio sludge	2.09	0.0	42.40
		2.89	0.0	49.10
		5.30	0.0	29.30
		7.83	0.0	14.20
5) tanning wastewater	AS	0.63	0.0	39.80
		0.96	0.0	35.00
		1.56	0.0	20.50
		2.35	0.0	0.00
6) elastic rubber	AS/bio sludge	0.63	0.0	33.30
		0.97	0.0	31.80
		1.32	0.0	26.70
		1.87	0.0	13.10
7) aluminium	chemical sludge	0.40	0.0	0.00*
		0.99	0.0	9.70
		1.73	0.0	7.40
		2.28	0.0	0.00
8) car assembly	chemical sludge	0.99	0.0	0.00*
		1.77	0.0	20.10
		2.31	0.0	3.40
		5.70	0.0	0.00

3). Effect of initial solid concentrations on drying time.

Sludge with higher solid concentration consumed more drying time than the lowers. This could be explained by the fact that sludge with high solid concentration will be thicker than sludge with low solid concentration when placed in drying beds, and thus it will be more difficult for moisture to move to the top surface of the layer and be evaporated.

CONCLUSION

The comparison of sludge drying time between 2 sets of experiments with (test unit) and without (control unit) pre-withdrawal of supernatant after sludge settling indicated little differences. It took only a few days for sludge in control unit to achieve similar solid concentration percentages.

ACKNOWLEDGEMENT

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presented paper

SALT TOLERANCE OF CARBON AND NITROGEN BACTERIA IN AN ANAEROBIC/ANOXIC/AEROBIC PROCESS

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ABSTRACT

Clean fresh water is becoming more and more scarce in certain cities, especially around the coastline areas. The sea water can be used for toilet flushing, resulting in the save of domestic water consumption in cities. However, the salt content in the flushed sea water can be harmful to any subsequent biological treatment processes. Certain industrial wastewaters from, for example, pickle and tanning factories are also high in salt content. A synthetic wastewater with the ratio of COD:N:P of 500:25:15 and the sodium chloride dose of 0, 5, 10, 20, and 30 g/l was fed to an anaerobic/anoxic/aerobic process with HRT of 2, 2 and 12 hours, respectively, so that the salt effects on the removal efficiency of the organics as well as nutrients of the said system could be determined. The experiment was done with and without inoculation of NaCl-acclimated bacteria (Models A and B) and the pH was kept at about 7.2 to 7.6 in the three tanks by daily adjustment with $Ca(OH)_2$ while the sludge age was 10 days for both units. The COD and T-N removal efficiency of Model A decreased from 97 to 60 percent and from 88 to 68 percent with the increase of salinity from 0 to 30 g/l, respectively. With NaCl acclimated Model B, the COD reduction capability improved to 71 percent at 30 g/l NaCl, showing some adaptability to high salinity of carbon bacteria. The total nitrogen removal efficiency of this latter model however did not increase appreciably because nitrogen concentration in the effluent was already very low (2.8-7.8 mg/l) and only a little improvement was possible. The tolerating capability to salinity of carbon and nitrogen bacteria was 20 and 30 g/l NaCl, respectively. The luxury P uptake was not apparent in this study and the salt tolerance of phosphate accumulating bacteria was therefore not possible to be determined.

KEYWORDS

Salinity effects, salt tolerance, BNR, EBPR, nitrogen bacteria, A_2/O process.

INTRODUCTION

Problems on water shortage experienced by many arid areas have forced planners and administrators to look for alternative water sources. Hong Kong has been using seawater for toilet flushing for at least 10 years, resulting in high salt content in the sewage. Certain industries eg. tanning, pickle, canned seafood etc. also discharge large amount of salt with their wastewaters. These high salt wastewaters can pose harmful effects on biological treatment processes Lawton and Eggert (1957), Stewart *et al* (1962), Kincannon and Gaudy (1966), Tokuz and Eckenfelder (1979), Matsuo and Hosobora (1988) and Hamoda and Al-Attar (1995) have reported the salt effects on several traditional treatment systems, but these researchers did not specifically work on BNR processes and it is the intention of this study to fill this gap by investigating the salinity effects on an Anaerobic/Anoxic/Aerobic (AA/A) unit

MATERIALS AND METHODS

Two separate AA/A models (A and B), details of which are shown in Figure 1 and Table 1, were parallelly run at room temperature of 27 to 33 °C with a synthetic wastewater with the composition as tabulated in Table 2. Excess sludge was manually drawn from the aerobic tank at 1.34 l/d to keep θ_c of 10 days and after 30 min of settling the supernatant was put back to the tank so that the effect on HRT was minimized. Models A and B which were with and without inoculation of NaCl-acclimated bacteria (taken from a local salt-rich tannery wastewater treatment plant), respectively, were tested with various dosages of salt, ie., 0 (control), 5, 10, 20 and 30 g/l NaCl. It is noted that while Model A was stopped after the study on the system performance on one run was accomplished and then re-started anew with new and non-acclimatized seed for the next run. Model B was continuously operated, ie., higher NaCl doses were simply added to the system between runs and additional acclimation took place with increase of time. However, in Model A, a certain degree of acclimatization was also possible during each individual run long before the steady state condition of that run was reached. The system pH was daily adjusted with $Ca(OH)_2$ to keep the pH of the anaerobic, anoxic and aerobic tank at about 7.2, 7.5 and 7.6, respectively. Samples were analyzed according to Standard Methods (APHA, 1995) except for COD test in which high dose of $HgSO_4$ (at the ratio of $HgSO_4$:Cl of 10:1) was added to the samples so that the chloride interference could be eradicated. Nitrate and Nitrite were analyzed by SHIMADZU UV-1201 Spectrophotometer and MERCK SQ118, respectively. Some samples were filtered with 1.2 μ GF/C paper before the analysis, these were refered to as filtrate or filtered samples.

RESULTS AND DISCUSSION

There was a slight drop in MLSS of Model A from 2385-2426 to 2015-1695 mg/l for the case of 0-10 to 20-30 mg/l NaCl, respectively. The related number for Model B

was however relatively constant (around 2000-2200 mg/l MLSS) for the whole range of NaCl concentration.

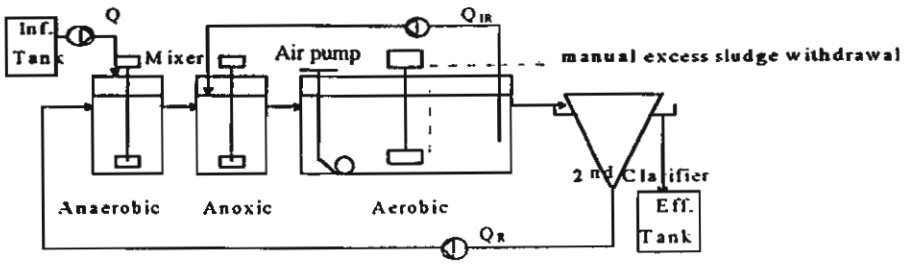

Figure 1. Experimental setup for this study.

Table 1. Operational parameters and tank size of test units.

Q	20 lpd	Anaerobic	volume	1.7 l
Q_{IR}	40 lpd		HRT	2 hrs.
Q_R	8 lpd	Anoxic	volume	1.7 l
Sludge age (θ_c)	10 days		HRT	2 hrs
$MLSS_{Initial}$	2500 mg/l	Aerobic	volume	10 l
			HRT	12 hrs

Table 2. Composition of the synthetic wastewater.

Constituents	Concentration (mg/l)	Constituents	Concentration (mg/l)
1. Sugar	301 (350 mg/l as COD)	5. $MgSO_4 . 7H_2O$	31.25
2. Acetic acid	0.133 cc./l (150 mg/l as COD)	6. $FeCl_3$	1.0
3. KH_2PO_4	65.8 (15 mg/l. as P)	7. $Ca(OH)_2$	85
4. $(NH_2)_2CO$	53.6 (25 mg/l as N)		

Note: NaCl concentration = 0 (control), 5, 10, 20 and 30 g/l.

Figure 2 illustrates the COD concentration in the effluent as well as the COD removal efficiency of both models. Bearing in mind that there might be some error in COD analysis due to the salt interference, the COD removal of Systems A and B dropped from 97 to 60 percent and from 90 to 71 percent for the 0 to 30 g/l and 10-30 g/l NaCl cases, respectively. It is obvious that the high salinity had adverse impact on the systems and such impact was to some degree less severe when the process had been earlier acclimated to the salt, both by pre-inoculation of acclimatized seed and by continuing exposure to the salt, as shown by Model B results. This less severeness was however not so distinct, indicating that the adaptation to salinity of carbon bacteria was not so good and already reached the maximum limit. For good effluent quality based on COD, the salt limitation to carbon bacteria should be less than 20 g/l NaCl. Most of the COD was apparently removed in the anaerobic step, especially when the system had not been exposed to high NaCl concentration before (Figure 3 a).

However, when the NaCl concentration increased to 20-30 g/l there was an increase of some 170 mg/l in the COD concentration at the anaerobic stage, indicating higher salt adverse effect on the facultative anaerobic carbon bacteria. Phosphate accumulating bacteria (PAB) were not in abundance in this case (see later discussion) and the effects on or by PAB could be neglected. Further COD reduction in the anoxic and aerobic steps was nil in the non or low (5-10 g/l) NaCl scenarios but was evidently noticeable in other cases. For Model B, the system performance on COD reduction in the anaerobic step was similar for all NaCl doses, indicating less impact of salt at the specific stage if the carbon bacteria were allowed ample time for adaptation to salinity (see Figure 3 b), especially when the NaCl concentration was lower than 20 g/l.

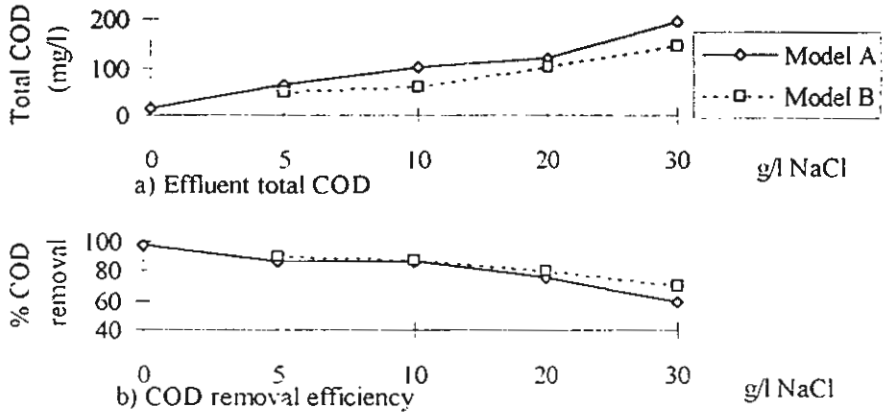

Figure 2 Effluent total COD and removal efficiency at steady state for both models.

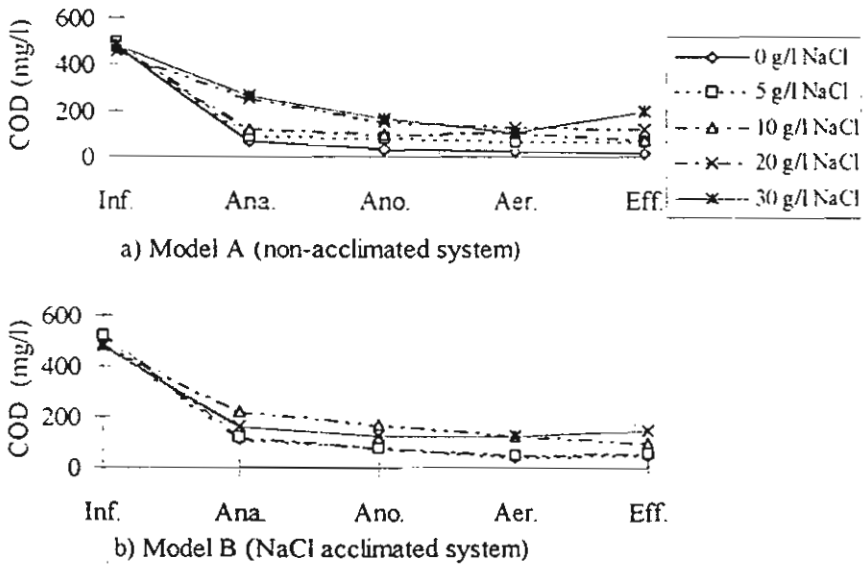

Figure 3 COD profile at different stages in the systems.