

# High rate biological/chemical treatment based on the moving bed biofilm process combined with coagulation

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## Abstract.

A high-rate wastewater treatment process for secondary treatment has been investigated, consisting of a highly loaded moving bed biofilm reactor directly followed by a coagulation/floc separation step. It is demonstrated that the biofilm in such a process mainly deals with the soluble organic matter while coagulation deals with the particulate/colloidal matter. The bioreactor may, therefore, be designed based on the soluble COD loading only, resulting in a very compact plant, especially when a compact separation reactor (i.e. filtration) is used as well.

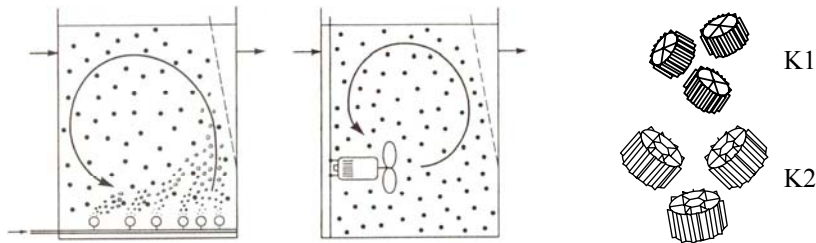
## 1. Introduction

Compact wastewater treatment processes are being looked for by cities all over the world as land available for treatment plants is becoming scarce. In many of the cities secondary treatment standards are to be met. Direct coagulation/-flocculation/floc separation, results not only in a very substantial removal of suspended and colloidal matter, but good removal of organic matter, bacteria and viruses and micropollutants as well [1,2,3]. These compounds are either associated with particles and colloids or soluble, high molecular weight organic substances

However, in order to meet secondary treatment standards it may be necessary, to remove low molecular weight, soluble organic matter as well. For this purpose biological processes are most suitable from an economic point of view. In the ambition to make compact treatment plants, biofilm reactors are especially suitable. However, many of the most compact biofilm reactors (like the granular media biofilters) cannot accept a high load of particulate matter, since this would easily clog the filter and result in too frequent filter washing. It is, therefore, normal to use a two-step process comprised of a pre-coagulation step followed by a biofilter step. By utilising lamella-separators for floc separation in the primary step, such treatment plants can be made very compact [4]. Nevertheless one is reluctant to use a very high organic load because of the fear of clogging of the biofilters. An alternative will be discussed in this paper, namely to combine the new moving bed biofilm process with direct coagulation. This process can accept both a high particulate load as well as a high soluble organic load.

## 2. The moving bed biofilm reactor (MBBR)

In the moving bed biofilm reactor (Figure 1) the biomass grows on carriers that move freely in the water volume of the reactor, kept within the reactor volume by a sieve arrangement at the reactor outlet. In aerobic processes, the biofilm carrier movement is caused by the agitation set up by the air, while in anoxic and anaerobic processes a mixer keeps the carriers moving (Figure 1a and b). The biofilm carrier (K1) is made of high density polyethylene (density  $0,95 \text{ g/cm}^3$ ) and shaped as a small cylinder with a cross on the inside of the cylinder and “fins” on the outside (Figure 1c). The cylinder has a length of 7 mm, and a diameter of 10 mm (not including fins). Lately a larger carrier (K2) of similar shape (length and diameter about 15 mm) has been introduced as well, intended for use in plants with coarse inlet sieves and especially for upgrading of activated sludge plants.



a. Aerobic reactor    b. Anoxic and anaerobic reactor    c. The biofilm carriers

**Fig. 1.** The moving bed biofilm reactor principle and shape of biofilm carriers

One of the important advantages of the moving bed biofilm reactor is that the carrier filling-fraction (% of reactor volume occupied with carriers in empty tank) may be subject to preferences. The standard filling fraction is 67 %, resulting in a total, specific carrier area of  $465 \text{ m}^2/\text{m}^3$  with the K1 carrier. Since the biomass grows primarily on the inside of the carrier, the effective specific surface area is  $335 \text{ m}^2/\text{m}^3$  for the K1 carrier and  $210 \text{ m}^2/\text{m}^3$  for the larger K2 carrier at 67 % filling fraction. It is recommended that filling fractions should be below 70 %. One may, however, use as much as needed below this, which is convenient, especially when upgrading plants – for instance from activated sludge to moving bed reactors.

The moving bed biofilm process has been used for many different applications [5],[6]. In this paper we shall only discuss the high-rate moving bed biofilm process. It has been shown that the carriers K1 and K2 performs equally well when comparisons are made based on the effective biofilm surface area [8].

The high-rate moving bed process was also investigated earlier [7], but then an alternative process scheme based on the use of aluminium-nitrate as coagulant and anoxic biodegradation was focused on.

### 3. The high-rate Kaldnes moving bed biofilm process

The fate of particles in biofilm reactors is not totally clear, but it is obvious that particulate matter to a far lesser extent is being degraded in a high-rate biofilm process than in a standard activated sludge process. When operating at such a high organic loading that the maximum COD degradation rate prevails, the COD-removal will primarily be due to consumption of soluble organic matter. The idea behind the high-rate Kaldnes moving bed biofilm process (Figure 2), is that the biofilm is supposed to deal with the soluble organic matter, while the coagulant is supposed to deal with the separation of the particulate matter, including colloids. If the new larger carrier (K2) is used, pre-treatment can consist of pre-sieving (3-4 mm sieve) only and the flow scheme shown in Figure 2 results then in an extremely compact plant, especially when a compact separation method (i.e. filtration) is employed.

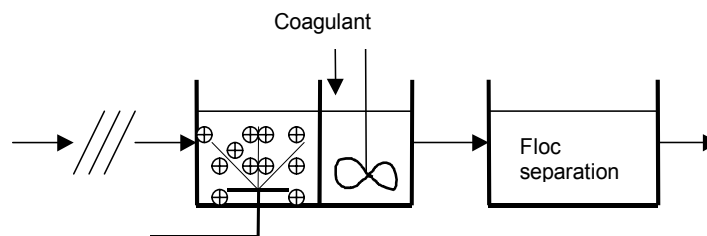


Fig. 2. Schematic flow diagram for the high-rate process discussed

When used for BOD-removal only, the process has traditionally been designed for a volumetric loading rate of 4-5 kg BOD<sub>7</sub>/m<sup>3</sup>d at 67 % carrier filling fraction (effective specific surface area: 335 m<sup>2</sup>/m<sup>3</sup>) and 15°C. This corresponds to ca 15 g BOD<sub>7</sub>/m<sup>2</sup>d, which is somewhat higher than for other biofilm processes (like RBC) for the same purpose. In the experiments that are reported here, the performance of the process at much higher organic loadings was investigated.

### 4. Biodegradation of soluble organic matter

The biodegradation experiments were primarily carried out in order to investigate the influence of carrier size and shape in the moving bed process. This is published elsewhere [8]. Here we shall concentrate on the influence of organic matter loading on biodegradation and only report the results with the carriers K1 and K2. The experiments were conducted in pilot plants in two parallel lines each, consisting of one moving bed reactor and one settling tank. The volumes of the bioreactors were 20 l. The moving bed reactors were operated at organic loads in the range of 10-120 g COD/m<sup>2</sup>d and 5-45 g SCOD/m<sup>2</sup>d. The experiments were carried out in three different periods. In the first one both reactors were given the same volumetric load at a filling fraction of 60 %, while in the second period the filling fraction was varied to give the same effective area load at constant flow.

The third period was devoted to a comparison between the two Kaldnes carriers (K1 and K2) at 70 % filling fraction. The two lines were operated in four sub-periods at close to constant flow in each period (e.g. the same residence time) and hence the same volumetric loading rate. The flow of the four periods corresponded to average residence times of 380, 52, 27 and 18 min. In Table 2 the wastewater characteristics for the various experimental periods are given.

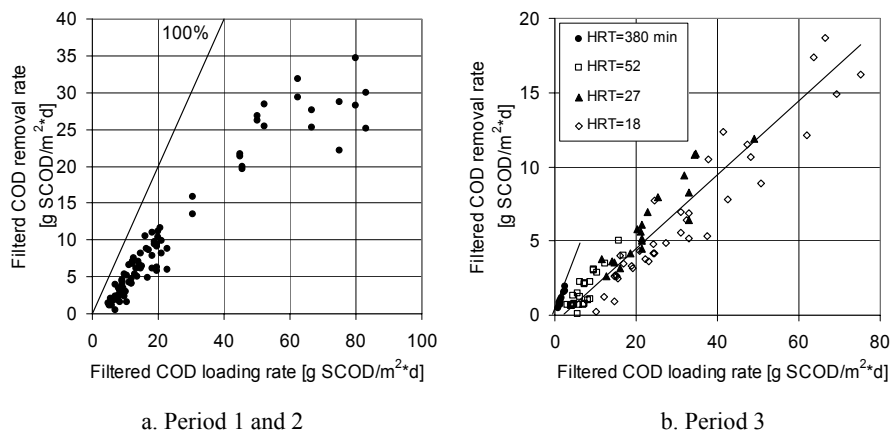
**Table 1.** Average, maximum and minimum influent values for the pilot plant

	Period 1			Period 2			Period 3		
	Ave. $\pm$ st.dev.	Max	Min	Ave. $\pm$ st.dev.	Max	Min	Ave. $\pm$ st.dev.	Max	Min
SS	136 $\pm$ 98	505	53	152 $\pm$ 55	232	58	88 $\pm$ 18	136	53
COD	323 $\pm$ 166	893	139	498 $\pm$ 235	915	125	219 $\pm$ 66	435	119
SCOD	123 $\pm$ 39	236	69	219 $\pm$ 128	431	36	100 $\pm$ 38	211	42
pH	7.3 $\pm$ 0.2	8.0	7.0	6.7 $\pm$ 0.3	7.1	6.5	7.5 $\pm$ 0.1	7.8	7.4

The raw water temperature was in the range of 10-15°C and the oxygen concentration in the range of 4-6 mg O<sub>2</sub>/l. Within this range, variations in O<sub>2</sub>-concentration are not expected to have any influence on the rate of COD-removal.

#### 4.1. Results from biodegradation experiments

In order to evaluate degradation of organic matter independent of the biomass separation step, one may look at the removal rate of soluble/filtered COD (SCOD) versus the loading rate. It is demonstrated in Figure 3 that the maximum removal rate in this wastewater was found to be around 30 g SCOD/m<sup>2</sup>d. This maximum rate was reached at a loading of around 60 g SCOD/m<sup>2</sup>d. A line through data points up to this loading is close to linear, indicating that the degradation rate was limited by the availability of biodegradable organic matter at organic loads lower than this.

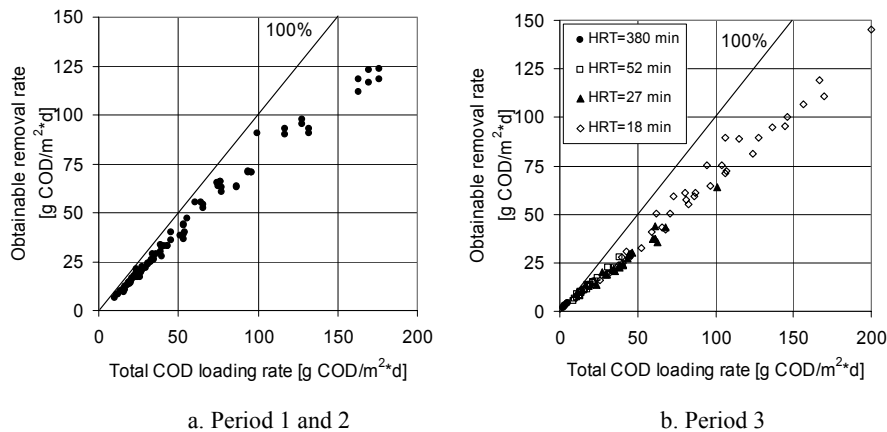


**Fig. 3.** Soluble COD removal rate versus soluble COD loading rate

The difference between this line and the 100 % removal line represents, therefore, the soluble COD that could not be biodegraded in this water within the actual residence time. The reason for the somewhat poorer removal in the second part of the experiment (Figure 3b and 4b) stems mainly from the fact that the water was more dilute then with a greater portion of the total COD that was not biodegradable. Most of the results are also obtained at very low residence times.

We can see, however, from the results of period 3, that there is not much difference between the results at 18 or 27 min residence time as compared to those at 52 min. This indicates that the removal of the biodegradable COD is rapid and that hydrolysis does not play an important part at these short residence times. It is interesting to note from Figure 3b, however, that at a very low load and long residence time (380 min - indicated by the line), the slope of the removal/loading rate relationship is significantly higher than that at the lower residence times (52, 27 and 18 min). This indicates that hydrolysis takes place at this long residence time.

Above we have concentrated on the removal of soluble (filtered) COD. It is not easy to analyse the total COD removal rates in the bioreactor, since the characteristics of both the soluble and the particulate organic matter change through the reactor by hydrolysis, assimilation etc. In order to be able to take the particulate matter into account, we have analysed what one may call the "obtainable" COD removal rate defined as:  $(\text{COD}_{\text{influent}} - \text{SCOD}_{\text{effluent}}) * Q/A$  where Q is the flow and A is the effective surface area of the carrier. This term illustrates the removal rate of organic matter if all particles larger than 1  $\mu\text{m}$  was removed in a downstream separation step. Figure 4 shows that 85-90 % removal of COD could have been obtained all the way up to a loading rate of 100  $\text{g COD}/\text{m}^2\text{d}$ , if the biomass downstream the bioreactor had been completely removed.



**Fig. 4.** "Obtainable" removal rate versus total COD loading rate

The results demonstrate that a much higher design load than normally used for secondary treatment may be accepted if efficient biomass separation is assured. In highly loaded plants clarification of the biomass may, however, represent a problem. In order to investigate this, another pilot study was carried out.

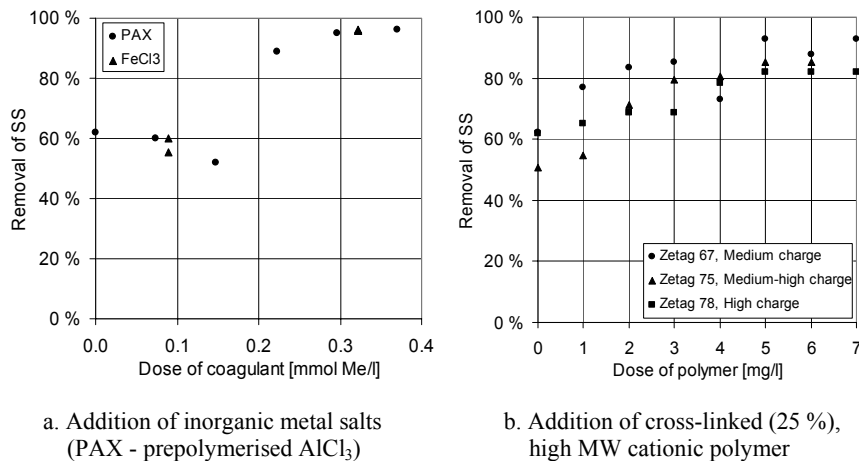
## 5. Separation of the biofilm by settling

Settleability experiments have been carried out in both jar- and pilot scale on effluents from high-rate moving bed reactors. In the jar-test experiments an apparatus based on continuous in-line mixing of the coagulants and pipe flocculation were used [9]. The pipe flocculators were operated at decreasing G-values ( $385 \text{ sec}^{-1}$  for 21 sec and  $135 \text{ sec}^{-1}$  for 47 sec) and the flocculated suspension was introduced to the 2 l jars from the bottom at a flow of 1 l/min. All experiments were performed with 15 min of slow mixing (25 rpm) and 15 min settling except for the ferric chloride tests where 60 min of settling was applied.

The pilot plant experiments were carried out in the same pilot plants as the biodegradation experiments (see above). When a coagulant was used, however, two flocculation chambers (with decreasing paddle speed and 40 min residence time) were introduced between the bioreactor and the settling tank. The settling tanks had a diameter of 0.38 m and a settling depth of about 1 m. Hydraulic bioreactor retention times in the range of 18 to 380 min were used, resulting in settling tanks overflow rates in the range of 0.05 to 1 m/h.

### 5.1. Results of settleability experiments

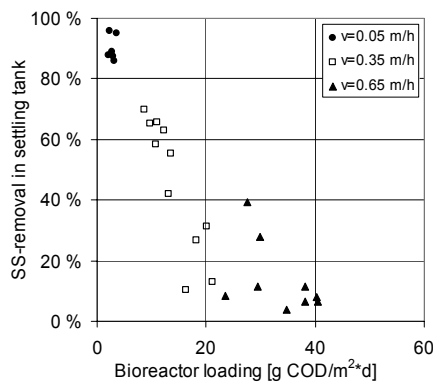
The effect of coagulation on separation of biofilm from a high rate moving bed reactor is demonstrated in Figure 6, showing results from the jar-test experiments



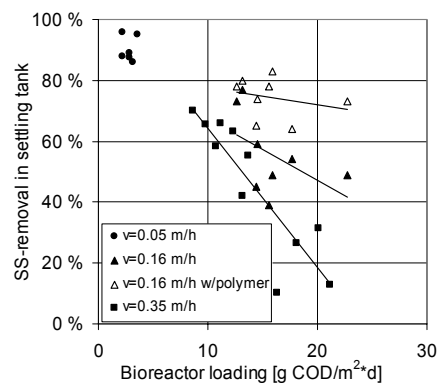
**Fig. 5.** Removal of suspended solids in jar tests of a high-rate MBBR effluent

It is demonstrated in Figure 5a that even a relatively small amount of metal, (about 0,2 mmol), gave dramatic improvement in settleability as compared to no coagulant addition. Figure 5b demonstrates that reasonably good SS-removal (85 %) could also be obtained by the use of a relatively low dosage (2 mg/l) of a cross-linked, medium charged, cationic polymer.

In Figure 6 and 7 settleability results from the pilot plants are shown. Figure 6 shows the influence of organic loading on the settleability of the biomass for the case that no coagulant was added. The SS-removal efficiency is given versus the total COD loading rate on the bioreactor at different overflow rates on settling tank. One should be careful in interpreting the actual removal percentages since these settling tanks were very small, but the general picture with respect to influence of loading rate can be considered to be correct. The different overflow rates correspond to three levels of flow and consequently different levels of organic loading on the bioreactor. This makes the analysis a little complicated since both loading rate and surface overflow rate varied at the same time.



**Fig. 6.** Influence of organic loading rate in bioreactor on settleability



**Fig. 7.** Influence of polymer addition on settleability

It is obvious, however, that not only the overflow rate but also the organic loading on the bioreactor has a pronounced effect on settleability. At a given overflow rate (for instance 0,35 m/h in Figure 6), there is a decrease in settleability with increasing organic load. This was the case both when total COD and soluble COD was considered, indicating that the organic loading regime experienced by the microorganisms, does influence on the settleability of the biofilm that is sloughed off the carriers. The consequence of this, from a practical point of view, is that settling ought to be enhanced by coagulation when one is operating this flow scheme at high organic loading rates. In Figure 7, results from experiments where two pilot plants have been run in parallel, are compared in order to study the effect of adding a cationic polymer coagulant at a given overflow rate. 1,5-2 mg/l of the medium charged, high MW polymer was added. The regression lines in Figure 7 are drawn for illustration purposes only.

Even though there is considerable scatter in the data, it is demonstrated by the results at the same overflow rate that; a) settleability is better with polymer addition and b) settleability is less influenced by the organic loading on the bioreactor when a polymer is added. The reason for this is believed to be that the polymer is able to flocculate the smallest particles that are more abundant the higher the organic loading is. The draw-back of poorer settleability at higher organic loading is therefore compensated for by the polymer coagulation.

## 6. Separation of the biofilm by direct filtration

Since the amount of sludge that is to be separated, is quite low in the actual process scheme, direct filtration may be an alternative to flocculation/settling. If phosphate removal is required, a metal coagulant (Al or Fe) will be necessary, but if only SS- and BOD-removal is looked for, a cationic polymer alone, or in combination with a low dose of metal salt, may be used. This would minimise sludge production and might make the use of direct filtration possible.

A filter for such an application should, of course, be built with a high sludge retaining capacity in order to achieve acceptable filter run times. In this project we have used an up-flow filter with expanded clay aggregates (Filtralite) as filter medium with a wide range of grain sizes (1.5-4 mm). The filter grain sizes will arrange itself from coarse to fine in the direction of flow after backwashing. This is ideal in terms of storage capacity and even distribution of particle deposits throughout the entire filter bed depth. Because of the low density of the lightweight expanded clay aggregates, this filter is operated with low filter backwash rates.

The filter of the pilot-plant used in our experiments (Figure 8) had a bed depth of 1.2 m, a 10-cm gravel layer (5-10 mm) as support layer, and 1.5-4 mm crushed Filtralite as filter medium. The inner diameter of the filter column was 85 mm and the total available head 1.8m. The head-loss could be determined by a pressure sensor at the base of the filter column.

Experiments were performed without coagulant addition as well as with the addition of various coagulants - ferric chloride at a dosage of 10 mg Fe/l, the low MW, highly charged polyDADMAC (Magnafloc 368) at a dosage of 1 mg/l and the high MW, highly charged polyacrylamide (Floerger FO4440SH) at 1 mg/l. The organic loading on the bioreactors varied somewhat throughout the experiments but was on average 8.0 kg COD/m<sup>3</sup>\*d (36.5 g COD/m<sup>2</sup>\*d) and 4.5 kg SCOD/m<sup>3</sup>\*d (20.5 g SCOD/m<sup>2</sup>\*d).

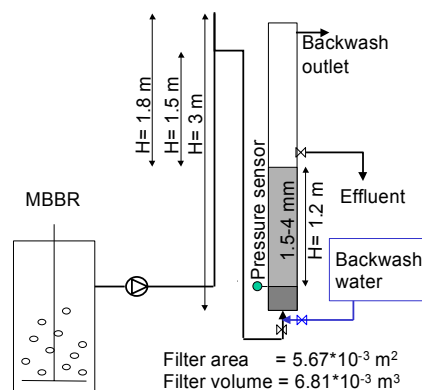
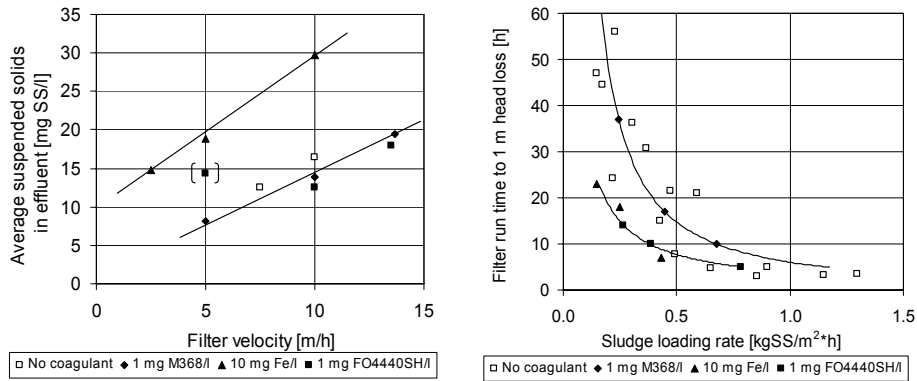


Fig. 8. Filter pilot plant

### 6.2. Results of filterability experiments

Figure 9 shows the effluent SS-concentration versus filtration rate. The lines are drawn for illustration purposes only. It is demonstrated that there is a close to linear relationship between the effluent SS-concentration and the filtration rate, no matter what kind of pre-treatment that has been used. The poorest results were obtained by the use of iron alone while the two different cationic polymers gave about the same results, which were almost 10 mg SS/l lower than with iron alone





**Fig. 9** Effluent SS-concentration versus filtration rate **Fig. 10.** Filter run time to 1 m head-loss versus sludge loading rate

Even at filtration rates as high as 20 m/h, the effluent SS-concentration could be kept under 20 mg/l while only 5 m/h could be used in order to obtain the same effluent SS-concentration with iron alone. This must be caused by the fact that iron alone produce small particles through coagulation in addition to those already present in the water, that escape the filter more easily than those bound together through the action of the polymer. Iron alone caused weak flocs leading to filter break-through as the determining factor for filter run time.

The filter run was terminated either when the maximum allowable head-loss (set at 1m) was reached or when breakthrough occurred (when maximum allowable effluent SS concentration, set at 30 mg SS/l, was reached). In all runs with iron alone as coagulant, break-trough determined the length of the filter run (max allowable effluent concentration of 30 mg/l was reached before max allowable head-loss) while maximum head-loss determined the length of filter run when a polymer was used as coagulant.

In Figure 10 the filter run time versus the sludge loading rate is presented. The lines are drawn for illustration purposes only. It is demonstrated that the cationic polymer with the lower molecular weight (Magnaflow 368) gave longer filter runs at a given sludge loading rate than the one with high molecular weight (FO440SH) at the same dosage. This is caused by the fact the low MW polymer acts as a pure coagulant, resulting in relatively small, compact flocs, while the high MW polymer acts according to the bridging mechanism as well, resulting in larger flocs that can not penetrate equally far into the filter. The fastest head-loss build-up was experienced with iron alone as a result of the higher amount of sludge to be separated as a consequence of metal hydroxide precipitation.

At the addition of only 1 mg/l of the low MW, cationic polymer, about 16 hours filter run time was obtained at a sludge loading rate of about 0,5 kg SS/m<sup>2</sup>h. A sludge loading rate as high as 0,75 kg SS/m<sup>2</sup>h could be used if a filter run time of 10 hours was acceptable. The latter loading equals a filtration rate of about 5 m/h

in a case without presettling ( $SS_{in\ filter} \sim 150\ mgSS/l$ ). At filtration rates below 10 m/h, the effluent SS-concentration could be expected to be  $\leq 15\ mg\ SS/l$ .

## 7. Conclusions

The following conclusions can be drawn from this study:

1. The combination of a high-rate moving bed reactor, possibly without primary settling, followed directly by a coagulation step will result in an extremely compact wastewater treatment plant, especially when a high-rate separation method (i.e. filtration) is also employed.
2. If only SS- and BOD-removal is looked for (secondary treatment), a cationic polymer alone can be used, minimising sludge production. In this case, a low molecular weight, highly charged cationic polymer should be chosen. The necessary dosage can be expected to be in the range of 1-2 mg/l.
3. If phosphate removal is required as well, a metal coagulant has to be used, but a low dosage is needed, normally  $< 0,2\ mmol/l$ .
4. If direct filtration is used for floc separation, a sludge loading rate of about 0,5 kg SS/m<sup>2</sup>h would result in a filter run time of about 16 hrs with the actual filter
5. At filtration rates between 7 and 10 m/h, the effluent SS-concentration could be expected to be  $\leq 15\ mg\ SS/l$  and  $\leq 10\ mg\ SS/l$  at filtration rates  $< 7\ m/h$ .

## 8. References

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