

ADVANCED COMPACT WASTEWATER TREATMENT BASED ON COAGULATION AND MOVING BED BIOFILM PROCESSES

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ABSTRACT

Advanced compact wastewater treatment processes are being looked for by cities all over the world as effluent standards are becoming more stringent and land available for treatment plants more scarce. In this paper it is demonstrated that a very substantial portion of the pollutants in municipal wastewater appears as particulate and colloidal matter. Pre-coagulation, therefore, gives very efficient pre-treatment that results in considerable saving in the total space required by the plant, especially when combined with a biofilm process for the removal of the soluble matter. A new biofilm process for this purpose is described. The moving bed biofilm process is based on plastic carriers, that move in the reactor, on which biomass attach and grow. The carriers are kept within the reactor by a sieve arrangement and biomass that is sloughing off the carriers is separated before effluent discharge. In addition to combining the moving bed biofilm process with pre-coagulation, the paper discusses also the use of a high-rate moving bed process combined with coagulation directly after the biofilm reactor in order to enhance separability. This results in very compact treatment plants for secondary treatment and possibly phosphate removal.

KEYWORDS

Wastewater, characteristics, coagulation, biofilm reactor, organic loading, performance, kinetics

INTRODUCTION

There are several factors that have to be taken into account when evaluating different treatment methods for wastewater treatment, such as; (1) treatment efficiency, (2) cost, (3) area requirement, (4) sludge production and (5) sustainability (e.g. ecological impact and energy use).

With respect to treatment efficiency, one has so far been focusing on oxygen consuming matter (organic matter and ammonium) and nutrients (P and N). This has led to more interest being devoted to removal of soluble matter than particles. In the future, however, it is reason to believe that more focus will be on microbial contamination (such as helminth eggs, parasitic protozoa, bacteria and viruses) and micropollutants (heavy metals as well as organic micropollutants). A very considerable fraction of these pollutants, as well as of organic matter, are present in municipal wastewater as particles or colloids or strongly associated to such. This ought to lead to a shift in the research interest from removal of soluble matter to removal of particles.

The cost of wastewater treatment is dependent on local circumstances. In the big cities, however, the space required for the plant is very determining for the investment cost. Compact treatment alternatives methods are consequently more and more being favoured. Biofilm systems are replacing activated sludge systems and high-rate separation techniques are replacing traditional settling tanks.

The disposal of sludge represents an increasingly higher portion of the total cost of wastewater treatment. Methods are looked for, therefore, that minimises the amount of sludge produced. The sludge does represent a resource, however. It contains both an energy potential that can be utilised as well as specific compounds that are limited in the world, such as phosphorous. It is a requirement today that treatment of wastewater should be sustainable. This means that the treatment should consume as little materials, chemicals, energy and land as possible and make reuse of the water, the nutrients, the energy and the chemicals.

The development of wastewater treatment has to a very large extent been hampered by tradition. After the discovery of the activated sludge process, biological treatment based on this process has become the dominant one for a number of purposes. From the start the main objective was removal of oxygen consuming material such as organic matter and ammonium. Later the process has successfully been developed also to be able to remove the nutrients phosphorous and nitrogen. Traditionally the activated sludge step has been following a primary treatment step based on settling. In many countries, however, (Denmark, Holland etc) extremely low-loaded systems without pre-treatment have been preferred, partly because of the ambition to minimise sludge production by in-reactor aerobic stabilisation. It can be questioned, however, if low-loaded activated sludge plants are sustainable. They consume a lot of materials (e.g. concrete) because of their large tank volumes, they use quite a lot of energy for aeration, they convert most of the organic matter to useless CO₂ and the resources of the wastewater can not optimally be reused

In this paper we shall discuss wastewater treatment in view of my basic wastewater treatment philosophy (see Figure 1). According to this philosophy, one should first remove as much of the particulate matter as economically feasible (by enhanced primary treatment, e.g. pre-coagulation) and then deal with the soluble matter afterwards in a compact bioreactor (e.g. a biofilm reactor). The organic concentrate should be made use of by producing energy (in the form of biogas) and carbon source that can be used in the soluble removal part of the process, if necessary. Other valuable resources of the sludge should be recycled such as phosphorous and possibly metals.

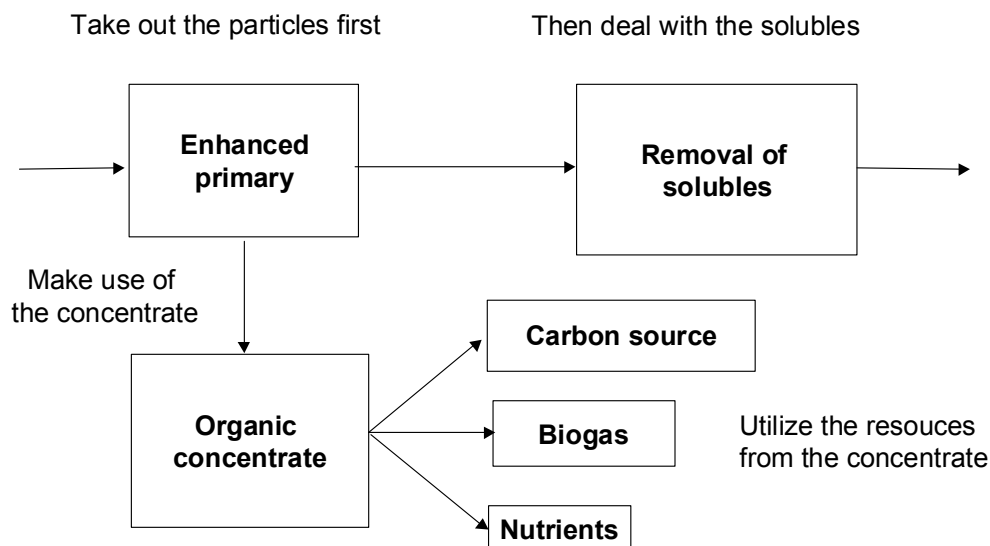


Figure 1 My basic wastewater treatment philosophy

WASTEWATER CHARACTERISTICS AND ITS INFLUENCE ON TREATMENT

Already in the fifties there were studies carried out in order to divide the organic contaminants in wastewater in different size fractions and demonstrate the difference in biodegradability of these fractions (Balmat, 1957; Heukelekian and Balmat, 1959; Richert and Hunter, 1971; Munch et al, 1980). Generally it was found that about 25 % of the COD was on a soluble form (defined as compounds/particles with size < 0,08 µm). About 15 % of the organic matter was found to be appearing as colloidal (0,08–1,0 µm) particles, about 25 % as supracolloidal (1-100µm) and about 35 % as settleable (> 100 µm) particles.

Levine et al (1985) concluded that organic contaminants in municipal wastewater could effectively be classified as being either greater or smaller than 0,1 μm based on a thorough study of particle size distributions in the primary effluents of some US plants. It was found that 63-70 % of the TOC was associated with particles less than 0,1 μm . Neis and Thiem (1996) carried out size distribution analysis in primary effluents in some German plants. They found that among particles $> 0,1\mu\text{m}$, 25-30 % was $< 1 \mu\text{m}$, 70-85 % $< 8 \mu\text{m}$, 85-95 % $< 32 \mu\text{m}$ and 100 % $< 100\mu\text{m}$. In primary effluents 48-69 % of the COD was found to be associated with particles $> 0,1 \mu\text{m}$ while 7-18 % was associated with particles $> 8 \mu\text{m}$.

Bacteria and viruses are particles in the size range of 0,01-10 μm . Even though the viruses are very small (0,01-0,1 μm), several investigations have demonstrated that 60-100 % of the viruses in sewage are adsorbed to particles (Ødegaard, 1987). Parasitic protozoa (like Cryptosporidium and Giardia) range in size from 2-60 μm while helminth eggs (like Ascaris and Taenia) are 10-100 μm (Asano, 1998). It has been shown that the particulate fraction of heavy metals in municipal wastewater range from 50-85 % (depending on the metal) and that metals as well as organic micropollutants like PCB and PAH have a high affinity to particulate matter (Ødegaard, 1987).

When evaluating data from Scandinavian plants, Ødegaard (1999a) reported that the suspended fraction (1 μm filter) in raw water samples was typically 60-80 % of the total COD and total BOD. This is shown in Figure 2 where each point represents data from one full-scale plant in one of the various countries.

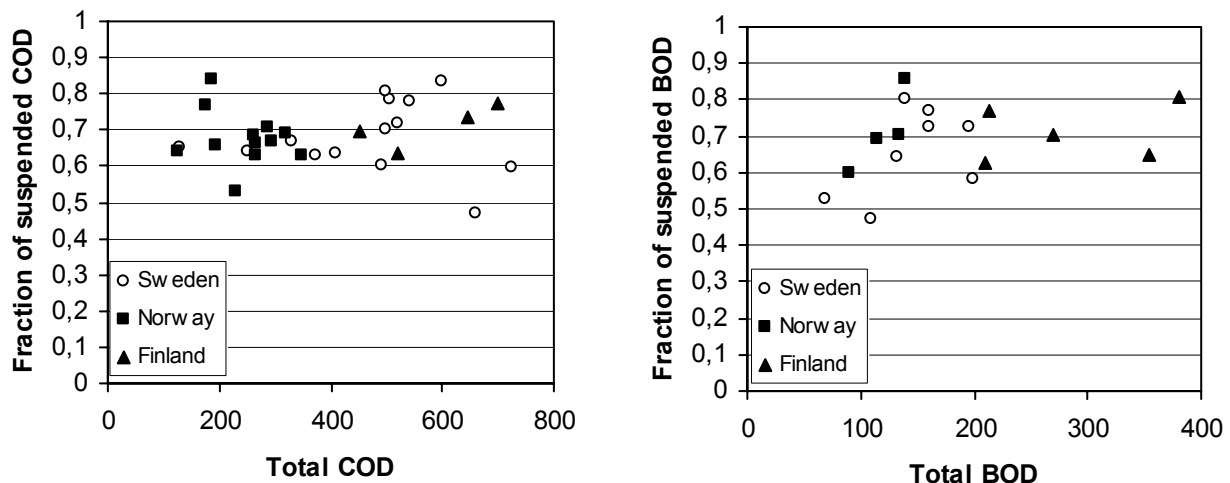


Fig 2 Fraction of suspended COD and BOD versus total COD and BOD (Ødegaard, 1999a)

It is quite remarkable that even though the concentrations of organic matter is vastly different from one plant to the other, as well as from one country to the other, the fractions of suspended COD and BOD are generally quite high and on the same level in the three countries. The data only include the suspended fraction. When including the colloidal fraction (0,1-1 μm) that normally represents 10-15 % of the COD, it can be estimated that only a fraction of 15-20 % of the total organic matter content is truly soluble.

How large the fraction of biodegradable organic matter is in a given case, is dependent upon several factors, such as; (1) origin of the wastewater - for instance influence of food industry wastewater, (2) influence of leakage water - storm water and infiltration water and (3) characteristics of sewer system. Particular attention should be given to the transformations of wastewater composition in the sewer network. The processes taking place may be physical (settling), chemical (precipitation) and biological (biodegradation). Of special importance with respect to organic matter and nutrients are the biodegradation processes and consequently the availability of oxygen in the sewer system. If the wastewater carries enough oxygen, aerobic biodegradation will take place in biofilms and in bioflocs. The most easily biodegradable matter will rapidly be converted into bacterial cells, i.e. into particulate, less biodegradable matter. On the contrary, if anaerobic conditions prevail in the sewer system, biodegradation of soluble organic matter is very slow. Hydrolysis of the particulate fraction of the organic matter may be significant, however, and results in an increase in the soluble fraction and a decrease in the particulate fraction (see Figure 3).

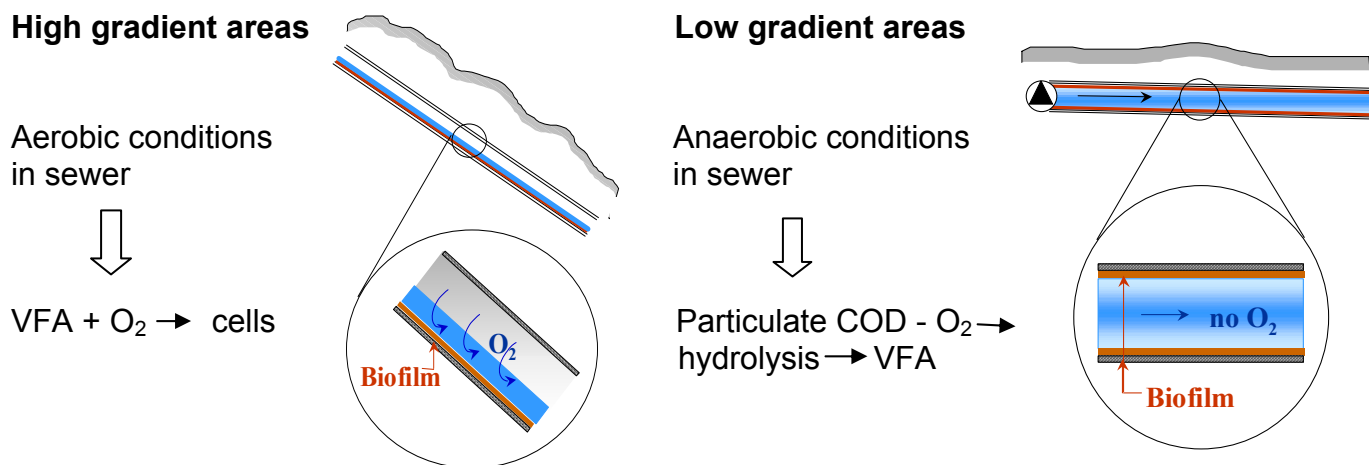


Fig. 3 Schematic of processes that are taking place in the sewer network

This may explain some of the regional differences. In Norway, for instance, there are many wastewater systems that carry oxygen-rich wastewater (caused by rapid flow, many pumping stations, large water surface-to-volume ratio etc). In most of the Norwegian plants reported, the soluble COD-concentration is well below 100 mg/l. It is experienced that the heavily biodegradable or inert soluble COD is in the order of 30-40 mg/l. Different investigations in the three countries that took part in the evaluation mentioned above (Ødegaard, 1999a), indicate that this is typical in Scandinavia. In Figure 4 the "BSCOD"-concentration for the different plants are plotted, assuming that the non-biodegradable soluble COD-concentration is 35 mg/l in all cases. The average BSCOD-levels calculated this way was found to be $122,4 \pm 79,2$ for the Swedish plants, $129,3 \pm 22,0$ for the Finnish plants and only $46.7 + 28.3$ for the Norwegian plants.

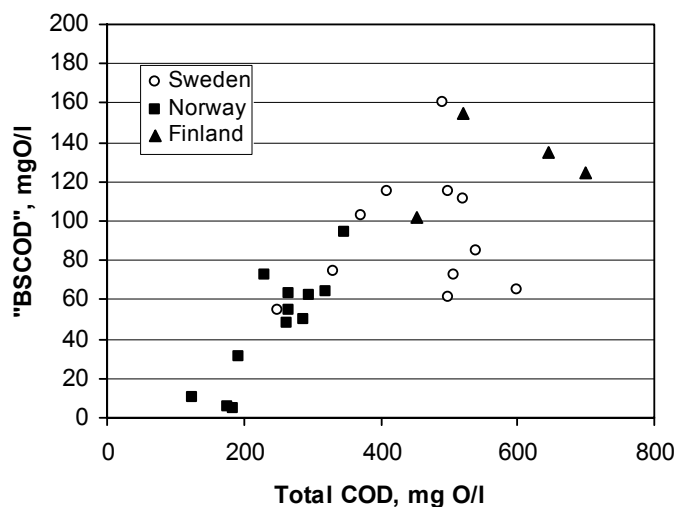


Fig. 4 Calculated biodegradable, soluble COD (BSCOD)-concentration versus total COD-concentration (Ødegaard, 1999a)

From what has been discussed above, it is quite clear that a very substantial reduction of organic matter and also other pollutants, can be removed by enhanced particle separation in the raw sewage. The smaller organic fractions biodegrade more rapidly than the larger fractions. This means that not only will the load of organic matter on the biological processes be reduced by enhanced primary treatment, but the rate at which the bio-process will perform, will be increased as well.

An objection to the use of pre-coagulation can arise from the fact that you remove organic matter that could be used as carbon source in biological nutrient removal processes. When organic matter is to be usable as carbon source in a compact biofilm process, however, it should be readily biodegradable since the residence time is normally too short for significant particulate organic matter hydrolysis. It may be better, therefore, to take out the organic-rich concentrate from an enhanced primary treatment (see Figure 1) and process this (by hydrolysis) so that the necessary carbon source is provided for in a controlled manner.

ENHANCED PRIMARY TREATMENT BY COAGULATION/FLOCCULATION

Traditionally settling alone has been used for particle separation in primary wastewater treatment. With the overflow rates normally used (around 2 m/h) and the densities of typical wastewater particles, it can be calculated from Stokes law that particles down to around 50 μm will settle. This means in practice that around 50 % of the suspended solids and 30 % of the organic matter is removed by primary settling. If particles down to around 0,1 μm could be removed, considerable improvement would be gained.

The most well-known and commonly used method for enhancing particle separation in primary treatment, is the addition of coagulants and/or flocculants. In primary coagulation (sometimes also referred to as primary precipitation) normally a metal salt is added to the raw wastewater resulting in destabilisation of colloids. The small aggregates of primary destabilised particles are flocculated and separated by settling (most common), flotation or filtration. The coagulant is normally based on aluminium or iron resulting in precipitation of phosphate as well as coagulation of colloids. In countries where primary precipitation plants are frequently used, the primary treatment goal has, in fact, been phosphate elimination. In situations where aerobic conditions prevail in the sewer network, primary coagulation is especially favourable, since a major part of the organic matter appears in the form of particles/colloids. This is the case in Norway as demonstrated in Table 1 where the average treatment result from Norwegian primary precipitation plants, taken from two different investigations of larger and smaller plants respectively (Ødegaard, 1992 and Ødegaard and Skrøvseth, 1995) are given.

It is shown that very good efficiencies in SS-removal removal (with corresponding good removal of organic matter) can be obtained even in small plants with large variations in flow, demonstrating the operational stability of the process. In a pre-coagulation plant the COD-load on proceeding biological processes can be reduced by more than 75 %. When a metal salt is used as the coagulant (as in this case) excellent phosphate removal is obtained as well. These results are contingent upon optimal design of reactors for mixing of coagulants, flocculation and settling.

Table 1. Average treatment results in 23 larger (>2.000 pe)(Ødegaard, 1992) and 35 smaller (<2.000 pe) (Ødegaard and Skrøvseth, 1995) primary precipitation plants in Norway

Parameter	Average inlet concentration		Average outlet concentration		Average treatment efficiency	
	Large plants	Small plants	Large plants	Small plants	Large plants	Small plants
SS (mg/l)	233 \pm 171	226 \pm 150	17,3 \pm 10,0	22,3 \pm 16,6	92,5	90,1
BOD ₇ (mg/l)	187 \pm 143	-	25,4 \pm 11,7	-	86,4	-
COD (mg/l)	505 \pm 243	494 \pm 90	108 \pm 40	121 \pm 72	78,6	75,5
Tot P (mg/l)	5,40 \pm 3,01	5,33 \pm 2,26	0,28 \pm 0,14	0,50 \pm 0,46	94,8	90,6

There are, however, two items that should be addressed when discussing pre-coagulation of wastewater; (1) how to minimise the sludge production and (2) how to minimise the space requirement.

Minimising sludge production

The downside of traditional chemical primary coagulation is the increased sludge production as compared to primary settling only, mainly as a results of improved SS-removal but partly also due to precipitated material. The sludge produced during chemical coagulation consists basically of the suspended solids removed and the coagulated/precipitated matter, as described below (Ødegaard and Karlsson, 1994):

$$SP = SS_{in} - SS_{out} + K_{prec.} * D$$

where SP = sludge production (g SS/m³)
 SS_{in}, SS_{out} = suspended solids concentration in influent and effluent respectively (g SS/m³)
 K_{prec.} = sludge production coefficient (g SS/g Me), typically around 4 for Fe and 6 for Al
 D = dosage of metal coagulant (g Me/m³)

One can only reduce sludge production by reducing K or D or both. The level of the dosage in primary precipitation plants in practice is very much determined by the need for phosphate removal, since this is primarily governed by the pH of coagulation. Plant operators have experienced that the best phosphate removal takes place at pH around 6 and they add enough of the acid metal coagulant to get down to this pH. This results in overdosing and a considerable precipitation of metal hydroxide, i.e. excessive sludge production. If, on the other hand, phosphate removal is not the important issue, but rather particle removal, one could lower the dosage without ruining coagulation efficiency by replacing part of the metal cation with an organic polymer cation. This cation will not result in precipitation and would only add little extra sludge production caused by coagulation. The least sludge production would be obtained by coagulation with a polymer cation alone. The K_{prec.}-value would then be close to zero. Experience has shown, however, that it is difficult to obtain equally good particle removal with polymer alone as with the use of metal coagulant. An evaluation of this matter can be done from Figure 6 where jar-test results from various dosing situations are compared (Ødegaard, 1998).

In Figure 5 a, c and d the ratios between the amount of SS produced and the amount of SS removed (SS_{in} - SS_{out}) as well as removal efficiency (1-SS_{out}/SS_{in})*100 are given versus dosage of coagulant. In Figure 5a only FeCl₃ was used as coagulant. The ratio of sludge production relative to the amount of removed SS, increases with the iron dosage from 1 to about 2 since matter is precipitated. The SS-removal efficiency increases with dosage to its maximum at dosages around 15 mg Fe/l or above. Figure 5b shows that the difference between the amount of sludge produced and the amount removed, increases with dosage, demonstrating the amount of sludge production caused by precipitation. It can be noted that a curve through the points seems to cross the x-axis at around 5 mg Fe/l under which no precipitation seems to occur.

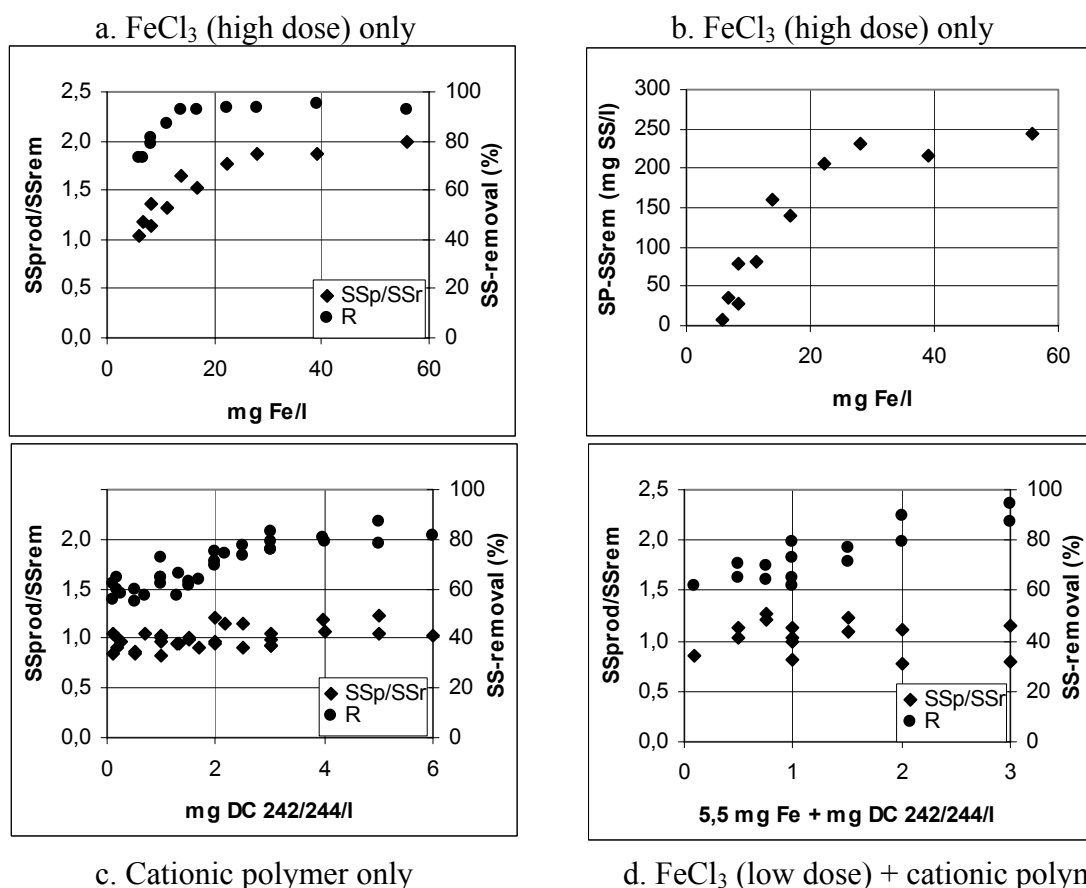


Fig. 5. Comparison of primary particle separation at different dosage scenarios (Ødegaard, 1998)

In Figure 5c the results from the use of a low molecular weight, strongly cationic polymer, as the only coagulant, are shown. Two different polymers were tested (Sepco DC 242 and Sepco DC 244). They showed little difference in performance, so the results from the use of both are included in the same figure. Now the sludge production is demonstrated to be close to equal to the amount of SS removed independent upon coagulant dosage. This means that close to nothing is precipitated. The SS-removal efficiency increases with dosage but reaches a maximum of about 80 % at dosages over 4 mg/l.

High SS-removal efficiency, combined with low sludge production could, however, be obtained when combining a low dosage of metal coagulant with a relatively low dosage of cationic polymer, as demonstrated in Figure 5d. From this figure it seems that the low iron dosage does not result in precipitation but rather in enhancement of the coagulation of particles.

Minimising the space requirement

Settling tanks for primary wastewater treatment have been designed for an overflow rate of 2-2,5 m/h while those for primary coagulation have been designed for 1-1,5 m/h, somewhat higher than typical for separation of activated sludge in secondary treatment plants (0,8-1,0 m/h). The acceptable overflow rate is given by the floc settling velocity that is primarily dependent upon floc size and density. Both these can be influenced by the addition of a flocculant, normally in the form of a synthetic, organic, anionic polymer. This is demonstrated in Figure 6 (Ødegaard et al, 1992) that is taken from an experiment where the floc settling velocity was determined in a small scale floc blanket reactor at two chemical dosage situations. These include iron chloride (JKL) followed by an anionic polymer (Praestol 2540) and prepolymerised aluminium chloride (PAX 60) followed by the same anionic polymer. Without the anionic flocculant, the settling velocity was in the order of 2-2,5 m/h in both cases, well in accordance with what is experienced as maximum acceptable overflow rates in practice. It is demonstrated, however, that the floc settling velocity could be considerably improved by increasing dosages of the anionic flocculant at a given metal dose (0,85 mmol Fe/l and 0,9 mmol Al/l) without ruining effluent particle concentration.

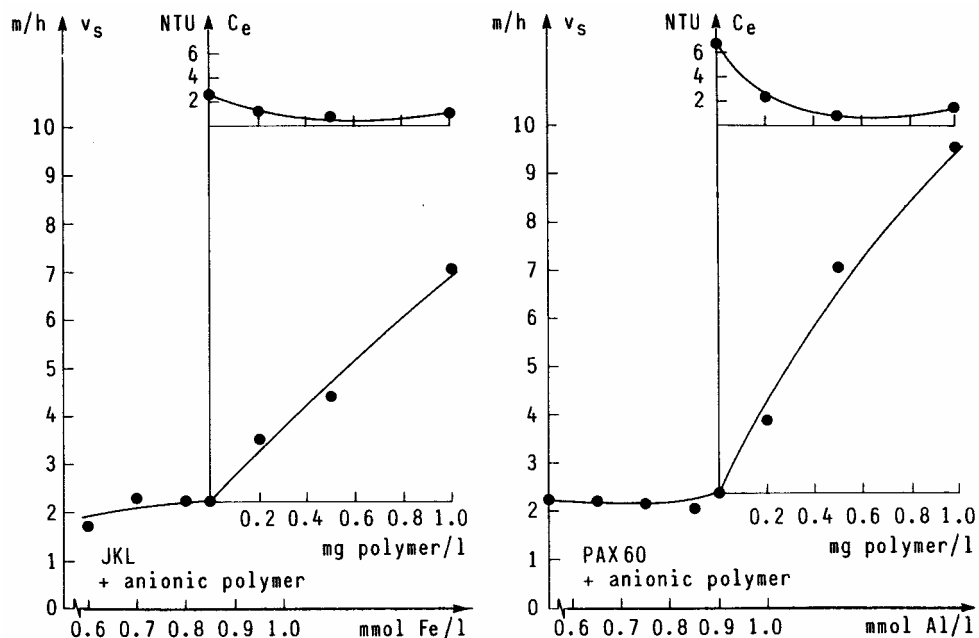


Fig. 6 Floc settling rate (v_s) versus dosage of metal salt and anionic polymer (Ødegaard et al, 1992) (C_e - turbidity of settled effluent)

It can be summarised from this section that very good particle removal, and consequently organic matter removal can be achieved by optimal use of chemically enhanced primary treatment. Especially when phosphate removal is not an objective, sludge production can be minimised by replacing part of the metal cation for coagulant by a organic polymer cation. The addition of an anionic organic polymer as flocculant can improve the settleability of the flocs dramatically with corresponding reduction of the space required.

THE MOVING BED BIOFILM REACTOR

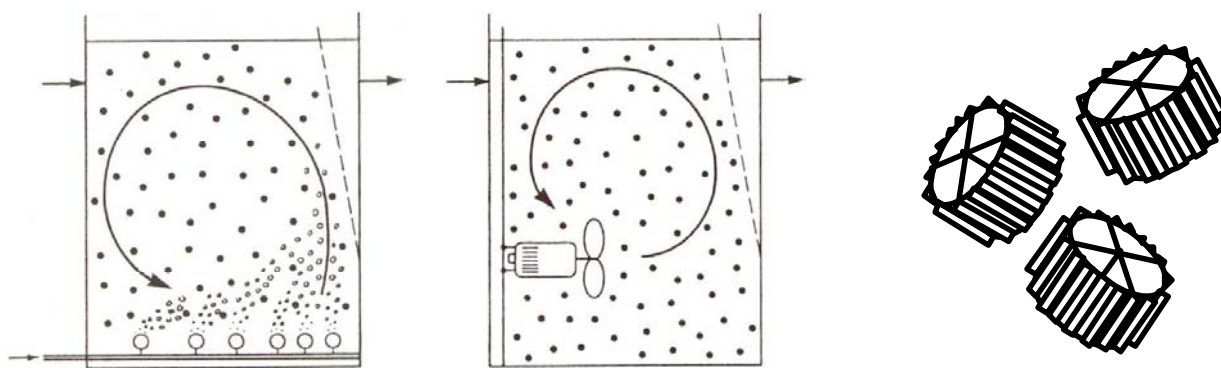
Over the last decades there has been a growing interest in biofilm processes for wastewater treatment because:

- Biofilm plants require less space than activated sludge plants
- The treatment result is less dependent on biomass separation since the biomass concentration to be separated is at least 10 times lower than in activated sludge
- The attached biomass becomes more specialised (higher concentration of relevant organisms) at a given point in the process train, because there is no sludge return

There are already many different biofilm systems in use, such as trickling filters, rotating biological contactors (RBC), fixed media submerged biofilters, granular media biofilters, fluidised bed reactors etc. They have all their advantages and disadvantages. The trickling filter is not volume-effective. Mechanical failures are often experienced with the RBC's. It is difficult to get even distribution of the load on the whole carrier surface in fixed media submerged biofilters. The granular media biofilters have to be operated discontinuously because of the need for backwashing and the fluidised bed reactors show hydraulic instability. For these reasons the moving bed biofilm process (Eur. pat. no. 0575314, US pat. no. 5,458,779) has been developed in Norway during the last 10 years (Ødegaard et al, 1994, Ødegaard et al, 1999b). There are presently more than 100 treatment plants based on this process in operation or under construction in 17 different countries all over the world. They are used for many different purposes for municipal as well as industrial wastewater treatment, like organic matter removal, nitrification and nitrogen removal.

Description of the moving bed biofilm reactor (MBBR)

The idea behind the development of the moving bed biofilm process was to adopt the best from both the activated sludge process and the biofilter processes without including the worst. Contrary to most biofilm reactors, the moving bed biofilm reactor utilises the whole tank volume for biomass growth, as does also the activated sludge reactor. Contrary to the activated sludge reactor, it does not need any sludge recycle, as is also the case in other biofilm reactors. This is achieved by having the biomass grow 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. Since no sludge recirculation takes place, only the surplus biomass has to be separated - a considerable advantage over the activated sludge process. The reactor may be used for both aerobic, anoxic or anaerobic processes, see Figure 7.



a. Aerobic reactor b. Anoxic and anaerobic reactor c. The biofilm carrier (K1)

Fig.7. The principle of the moving bed biofilm reactor and the shape of the biofilm carrier

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 (normally a horizontal shaft mounted banana mixer) keeps the carriers moving. In the aerobic reactors a special coarse bubble aeration system has been developed. The carriers are kept within the reactor by an outlet sieve. Most of the plants are designed with vertically mounted, rectangular mesh sieves, but it is sometimes shaped as a cylindrical bar sieve, vertically or horizontally mounted. 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 (see 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.

One of the important advantages of the moving bed biofilm reactor is that the filling fraction of carrier in the reactor (% of reactor volume occupied with carriers in an empty tank) may be subject to preferences. The standard filling fraction is 67 %, resulting in a total, specific carrier area of 465 m²/m³ with the K1 carrier. Since the biomass is growing primarily on the inside of the carrier, one is calculating with an effective specific surface area of 335 m²/m³ for the K1 carrier and 235 m²/m³ for the larger K2 carrier (at 67 % filling fraction). In order to be able to move the carrier suspension freely it is recommended that filling fractions should be below 70 % (corresponding to 350 m²/m³ effective specific area for K1 and 220 m²/m³ for K2). 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. It was developed at the time when nitrogen removal became in focus and most of the scientific data has been gathered from this application (Hem et al, 1994; Ødegaard et al, 1994; Rusten et al, 1995a and 1995b). Later, however, organic matter removal has been more investigated (Ødegaard et al, 1999c). At this time research is being conducted in order to evaluate the process for biological phosphate removal (Helness and Ødegaard, 1998). In Figure 8 various process schemes for different applications are outlined. Among those where the moving bed reactor is combined with chemical coagulation, are Figure 8b where a moving bed reactor is followed directly by a coagulation step. This process is frequently used in Scandinavia for BOD- and P-removal. It can also be used in high-rate systems for only BOD-removal without pre-settling tank if the larger carrier (K2) is used. In Figure 8c a scheme based on pre-coagulation, frequently used in nitrification plants, is shown. Figure 8h shows a typical pre-coagulation/post-denitrification process while Figure 8i shows a solution used in several Swedish plants where an activated sludge plant with post-precipitation has been upgraded to a pre-coagulation/post-denitrification plant with moving bed reactors for denitrification. Figure 8j shows the most frequently used solution in Norway based on coagulation and combined precipitation.

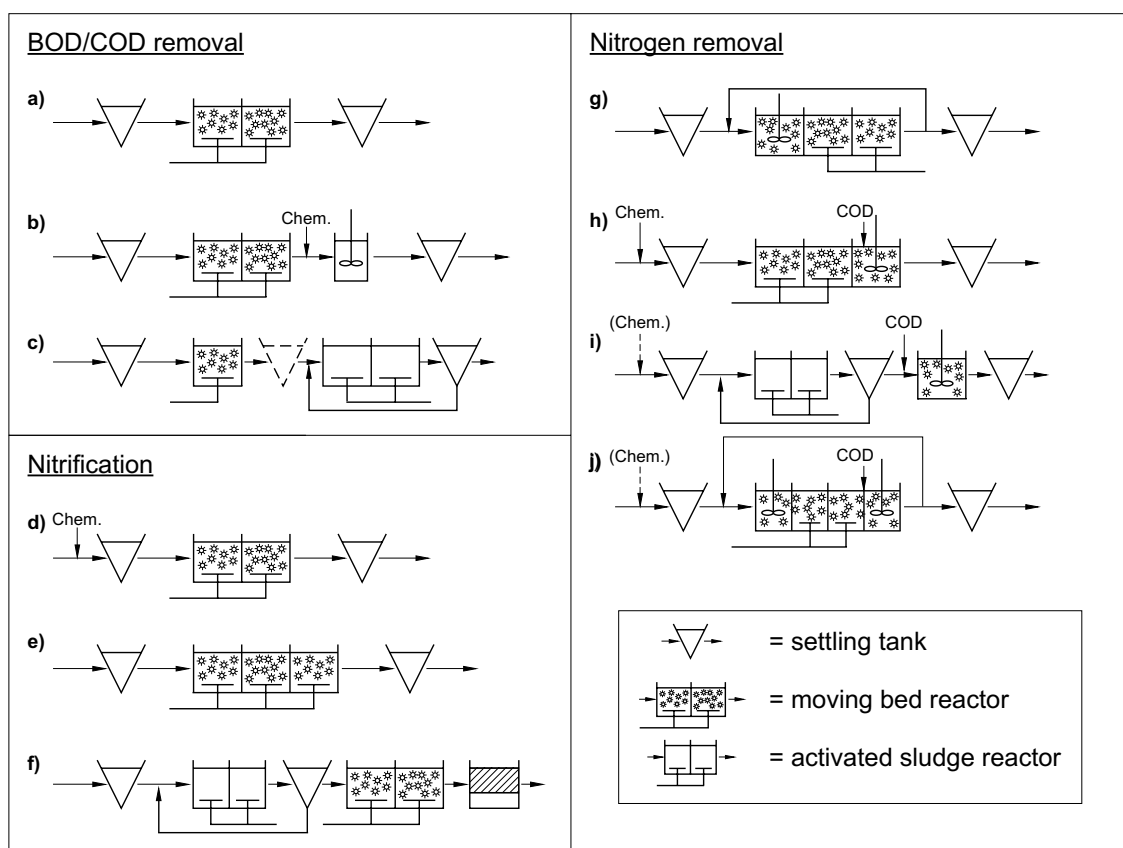


Fig. 8 Flow schemes combining moving bed biofilm processes with coagulation processes (Ødegaard et al, 1999b).

MOVING BED BIOFILM PROCESSES IN COMBINATION WITH COAGULATION

The fate of particles in biofilm reactors is not totally clear, but it is obvious that particulate matter to a much lesser extent than in activated is being degraded in a biofilm process, especially in a high rate process. In some biofilm reactors, for instance the trickling filter, the time in which the water (carrying particles) is in touch with the biomass is very short - a few minutes. In a moving bed reactor it is a bit longer, 0,5 - 3 hrs depending on the objective of the process. The fact that the biofilm is primarily active in biodegradation of readily biodegradable soluble matter and that the particles pass through the reactor more or less unchanged, makes it natural to combine the moving bed reactor (for the removal of soluble matter) with coagulation (for the removal of particulate matter). This can be done according to two main principles :

1. The particles are allowed to enter the moving bed reactor, they pass it more or less unchanged and they are being removed in a subsequent coagulation step that can be placed directly after the MBBR.
2. The particles are removed ahead of the MBBR, thus reducing the organic load on this

The first alternative may be of special interest in a high rate plant for SS- and COD-removal while the second alternative is more of interest in a low-rate situation where nutrient removal is required.

High-rate moving bed biofilm process combined with coagulation (see Figure 8b)

For BOD-removal only, the process has normally been designed for a volumetric loading rate of 4-5 kg BOD₇/m³d at 67 % carrier filling (335 m²/m³) and 15°C. This is corresponding to a loading based on area of ca 15 g BOD₇/m²d, which is somewhat higher than what RBC-processes have been designed for the same purpose. The process for high-rate BOD/COD-removal has, however, been evaluated. The main objective of that investigation was evaluation of the carrier size and shape in the moving bed process (Ødegaard et al, 1999c). Here, some of the data shall be used to demonstrate the kinetics of organic matter removal.

In order to evaluate degradation of organic matter independent of the biomass separation step, one may look at the removal rate of soluble COD (SCOD). In Figure 9 and 10 the results from pilot plant investigations are shown in which carriers of two different sizes are shown, here called K1 and K2 (early version). It is demonstrated that there is no significant difference between the removal rates of the smaller K1 carrier (effective area 410 mm²/piece) as compared to the larger K2 carrier (810 mm²/piece) as long as they are compared on the basis of surface area removal rate (g/m²d). Of course, the smaller carrier will need a smaller reactor volume at a given loading rate (as g/m²d) when the carrier filling fraction is the same. An objective for development of a larger carrier (K2) was to use a media retention sieve of 10 mm opening, giving greater operator confidence when using 6 mm inlet screens.

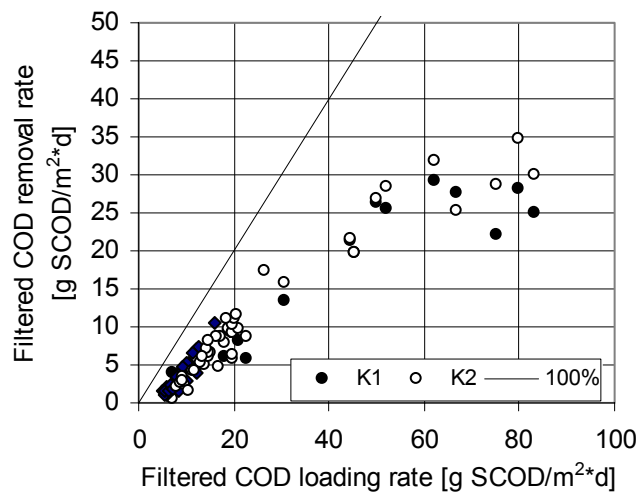


Fig 9. Soluble COD removal rate versus soluble COD loading rate

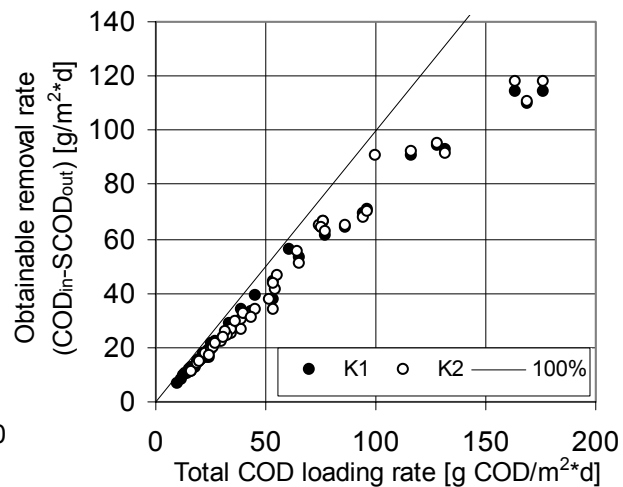


Fig 10 “Obtainable” COD removal rate versus total COD loading rate

It is demonstrated in Figure 9 that the maximum removal rate in a dilute municipal wastewater was found to be around 30 g SCOD/m²d. This maximum rate seemed to be reached at a loading of around 60 g

SCOD/m²d. The data point line up to this loading is close to linear, indicating that the availability of biodegradable organic matter was limiting the degradation rate at loads lower than about 50 g SCOD/m²d. 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.

This does not, however, give the true picture since biodegradable, soluble organic matter is produced in the process by hydrolysis. An alternative is to evaluate the so-called “obtainable” removal rate, meaning the removal rate of total COD at 100 % biomass separation. Figure 10, that gives the "obtainable" removal rate versus total COD loading rate, demonstrates that high removal efficiencies may be obtained even at extremely high loading rates if good biomass separation can be assured. The curve levels off very slowly and very high loading rates (> 100 g COD/m²d) can be used without losing much on the removal rate of filtered COD. The data show that 85-90 % removal of COD could have been obtained all the way up to a loading rate of 100 g COD/m²d, if the biomass downstream the bioreactor had been completely removed..

It is also demonstrated that the design load for COD/BOD-removal normally used (25 - 30 g COD/m²d) is very conservative, and that a much higher design load may be accepted when an efficient biomass separation method is employed. In highly loaded plants, clarification of the biomass does, however, represents a problem. This is demonstrated in Figure 11 where the influence of organic loading on the settleability of the biomass is shown. The SS-removal efficiency in small pilot plants is given versus the total COD as well as SCOD area loading at different overflow rate. 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 may be considered to be correct. The different overflow rates correspond to the 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 was varied at the same time.

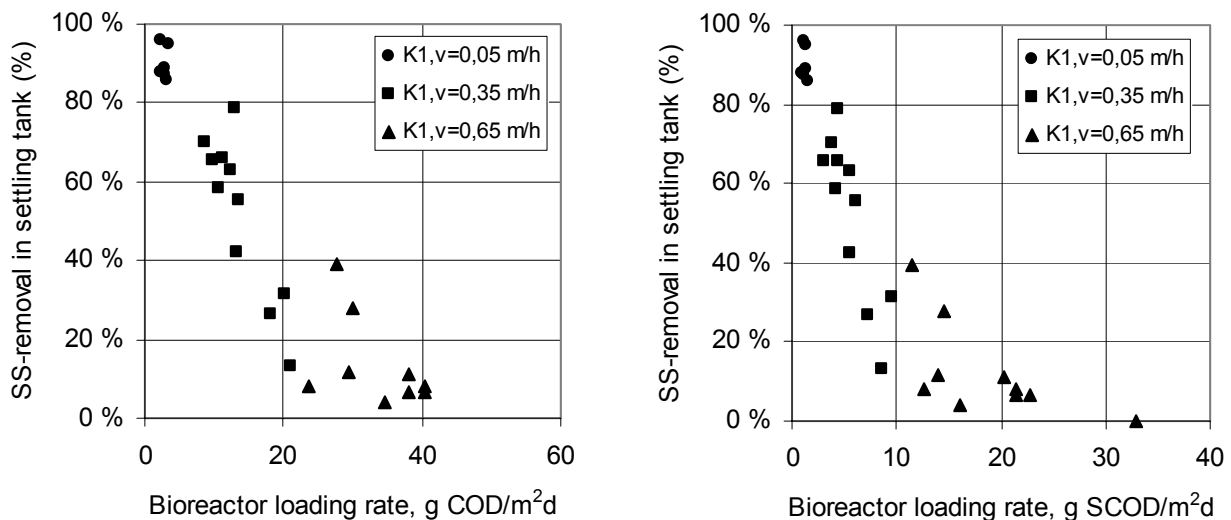


Fig 11. Influence of organic loading rate in bioreactor on settleability

It is obvious, however, that both the settling tank overflow rate and the bioreactor organic loading rate have a pronounced effect on settling. At low organic loading rates and low overflow rates, the separation of SS is very good, but it is poor at the high overflow rates and high bioreactor loading rates. For each overflow rate, however, there is a clear decrease in settleability when the organic load is increasing. One cannot judge correctly the influence of organic loading rate of the bioreactor load on settleability by analysing the total COD diagram alone, since a considerable part of the removed SS is coming in with the raw water. When taking into account that the same influence is seen in the SCOD diagram, it can be concluded without doubt that increasing the organic loading rate on a moving bed reactor decreases the settleability.

The natural consequence of this is that settling should be enhanced by coagulation (by adding metal salts or organic polymers) or that separation techniques other than plain settling should be applied. The effect of such direct coagulation is shown in Figure 12 where the improvement of SS-removal in the effluent from a highly loaded MBBR is demonstrated

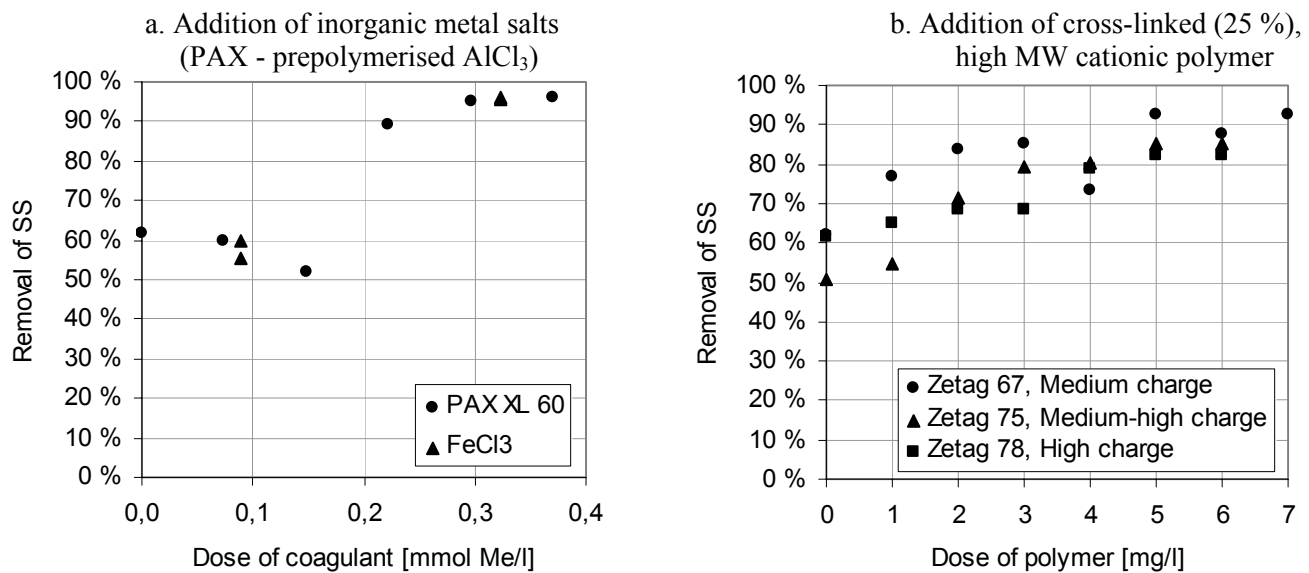


Fig 12 Removal of suspended solids in jar tests of a high-rate MBBR effluent

It is demonstrated in Figure 12a that even relatively small amounts of metal, for instance 0,22 mmol of polyaluminium chloride (< 7 mgAl/l) gave dramatic improvement in settleability. Figure 12b demonstrates that reasonably good SS-removal could also be obtained by the use of a cationic polymer alone at relatively small dosages (85 % removal at 2 mg/l of a cross-linked, medium charged cationic polymer).

It can be summarised that the combination of a high-rate moving bed reactor, possibly without primary settling, followed directly by a coagulation step (see Figure 8b) will result in an extremely compact wastewater treatment plant, that is easy to operate. If only SS- and BOD-removal is looked for, a cationic polymer alone, or in combination with a low dose of metal salt, can be used, minimising sludge production. If phosphate removal is required as well, a metal coagulant has to be used.

Low-rate process for nutrient removal

The moving bed biofilm process has very favourably been used also in low-rate systems, for nitrification and nitrogen removal (Hem et al, 1994, Ødegaard et al, 1994 and Rusten et al, 1995a and b), with chemical coagulation as pre-treatment (see Figure 8d, h, i, and j). Three factors, the load of organic matter, the ammonium concentration and the oxygen concentration primarily determine the nitrification rate. The influence of these parameters is schematically shown in Figure 13. It is demonstrated that the organic load is a key factor and should be as low as possible. At organic loading rates over about 4 g BOD₇/m²d, high oxygen concentrations (> 6 mg O₂/l) are required in order for nitrification to take place. An enhanced pre-treatment by coagulation is, therefore, very favourable.

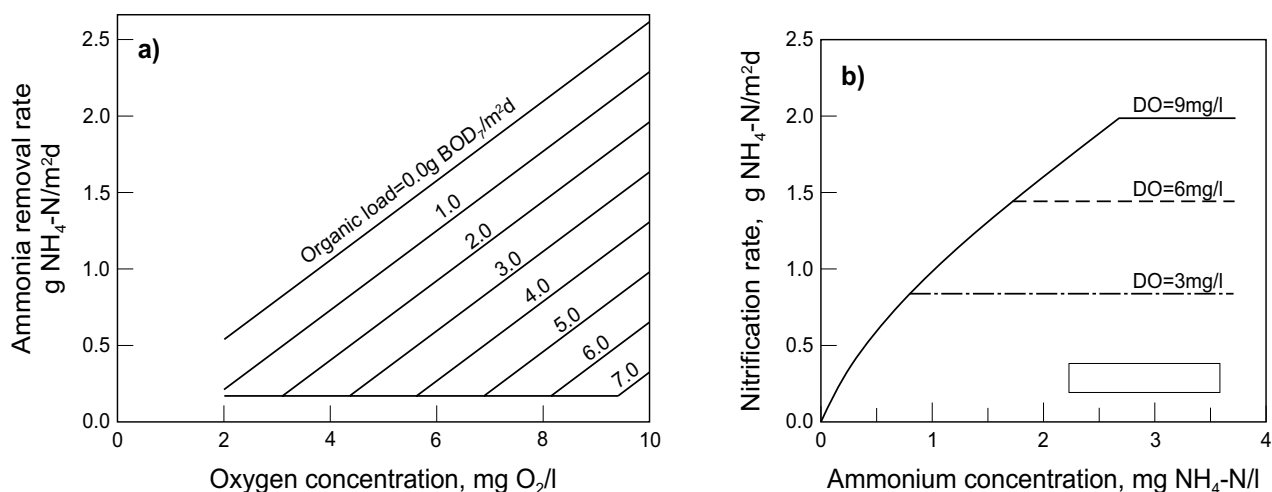


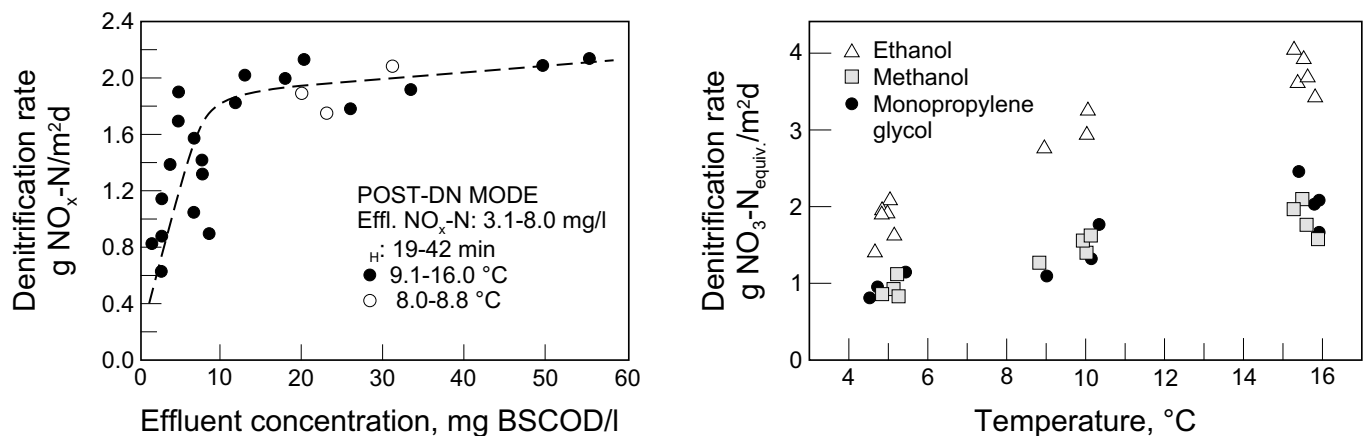
Fig. 13 Influence of BOD₇, oxygen and ammonium on nitrification rate (Ødegaard et al, 1999b)

As shown in Figure 13b, the ammonium concentration is only limiting the nitrification rate at low ammonium concentrations ($< 3 \text{ mg NH}_4\text{-N/l}$). Far more important is the influence of oxygen concentration that may limit the nitrification rate even at relatively high concentrations. It is experienced that an oxygen level above $2\text{-}3 \text{ mg O}_2\text{/l}$ is needed in order to initiate nitrification. The nitrification rate has been found to be close to linearly dependent upon the oxygen concentration, up to more than $10 \text{ mg O}_2\text{/l}$ (Ødegaard et al, 1994, Æsøy et al, 1998). Above an ammonium concentration of $3\text{-}4 \text{ mg NH}_4\text{-N/l}$, the nitrification rate is primarily governed by the oxygen concentration and the organic load.

Three different flow schemes are given in Figure 8d-f. In Figure 8d is shown a process scheme frequently used in Scandinavia, where coagulation is used as pre-treatment, removing phosphate as well as particulate and colloidal organic matter. In this case the load of particles (biomass) on the nitrifying reactors will be less than in the process of Figure 8e. This results in higher nitrification rates

Nitrogen removal in moving bed biofilm plants may be achieved both by pre-denitrification (Figure 8g) and post-denitrification (Figure 8h and 8i) and a combination of the two, the so-called combined denitrification process (Figure 8j). The latter is particularly useful when combining the moving bed process with pre-coagulation.

The denitrification rate may be limited by the nitrate concentration, the biodegradable organic matter concentration or by the oxygen concentration (or rather the presence of oxygen). At $\text{NO}_3\text{-N}$ concentrations above about $3 \text{ mg NO}_3\text{-N/l}$, the denitrification removal rate will be completely governed by the type and availability of easily biodegradable carbon source (see Figure 14a).



a. DN-rate versus biodegradable, soluble COD concentration (Rusten et al, 1995b)

b. DN-rates versus temperature with different carbon sources (Rusten et al, 1996)

Fig. 14 Denitrification rates reactors determined in moving bed pilot plant reactors

If oxygen is supplied to the reactor with the inlet water or recirculated water, biodegradable organic matter will be consumed for oxygen respiration and thus reduce the available amount for denitrification. This is the problem with the pre-denitrification process (Figure 8g), since it requires a high recirculation ratio in order to give high nitrogen removal. The available easily biodegradable carbon source in the raw water is often very limited (as discussed above), and consequently a carbon source has often to be added even in pre-denitrification systems .

In post-denitrification systems, one has to add carbon source and since this will be easily biodegradable, a very high denitrification rate may be expected, as demonstrated in Figure 14b. The post-denitrification mode has several advantages over the pre-denitrification mode. It may lead to considerably lower total bioreactor volumes (40-50 %) and it gives much better process control. The major drawback with the post-denitrification process, is the need for higher carbon source addition. In order to minimise the use of carbon source, the flow scheme of Figure 8j - combined denitrification - has been preferred in several Norwegian plants. The combination process offers greater flexibility with respect to operation of the plant. It can be used with or without pre-coagulation. In the latter case phosphate is also removed if a metal salt is used, preferably a pre-polymerised aluminium chloride in order to minimise alkalinity consumption.

The organic matter in the coagulated wastewater will partly consist of low molecular weight, easily biodegradable organic matter (20-40 mg BSCOD/l), that has the capacity to denitrify a certain amount of nitrate (4-8 mg/l). This is brought to the pre-denitrification step by a moderate recirculation flow (0,25-0,5 times Q) thus minimising oxygen recirculation. The rest of the nitrate is removed in the post-DN step where the measured effluent nitrate concentration controls the carbon source addition.

In the general process scheme that was presented in Figure 1, it was shown that the carbon source that has to be added in the post-denitrification step, may be produced by hydrolysis of the primary sludge. Both biological hydrolysis (Æsøy and Ødegaard, 1994) and thermal hydrolysis (Barlindhaug and Ødegaard, 1996) may be used. Biological hydrolysis, achieved by keeping the sludge under anaerobic conditions for about 2 days (for instance in a thickener), can give a yield (defined as $COD_{solubilised}/COD_{tot}$) of 10-15 %. Thermal hydrolysis, achieved by keeping the sludge under high temperature (150-180 °C) for about 30 min, will result in a yield of 25-30 %. In experiments with a process similar to the one presented in Figure 15, it was found that 67 % of the filtered COD released by the thermal hydrolysis was biodegradable and utilised for denitrification (Barlindhaug and Ødegaard, 1996). In order to avoid carbon source limitation the necessary C/N-ratio ($COD_{filtered}/NO_3\text{-equiv.}$) had then to be above 6,5-7. The maximum denitrification rates was found to be f 3-4 g $NO_3\text{-equiv.}/m^2\cdot d$ at 20 °C, close to what has been found with external carbon sources (see Figure 14b). Thermal hydrolysis leads also to release of ammonium. It has been demonstrated, however, that at a typical raw water COD/N-ratio of 15 and a fraction of readily biodegradable COD of 10 %, a removal efficiency of 70 % can be obtained without nitrogen removal from the hydrolysate (Ødegaard et al, 1996). An increase of about 5% may be obtained by ammonium removal in the hydrolysate, for instance by ammonium stripping, as shown in Figure 15.

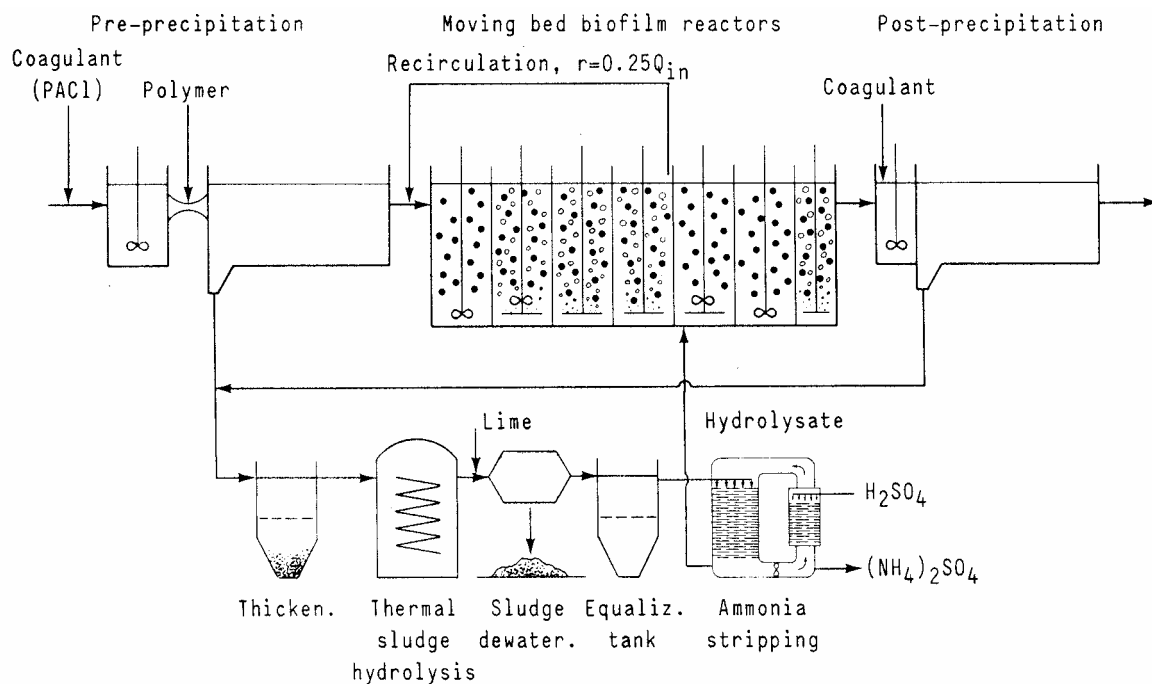


Figure 15 A flow diagram for a process based on pre-coagulation, nitrogen removal in moving bed reactors by combined denitrification and internal production of carbon source by thermal hydrolysis

CONCLUSIONS

It has been demonstrated in this paper that a process scheme based on a combination of moving bed biofilm reactors and coagulation will lead to very compact treatment plants usable for a number of purposes.

1. If only BOD- removal is aimed for, high-rate moving bed reactors followed directly by a coagulation step, with cation polymer addition, will result in a very cost-effective alternative. If phosphate removal is required, a metal coagulant has to be used.

2. If nitrification is aimed for, a process based on pre-coagulation followed by moving bed biofilm reactors will be very area-efficient especially if filtration is used as the final separation step.
3. If nitrogen removal is aimed for, a process based on pre-coagulation, followed by moving bed reactors in a combined denitrification process, will be compact and easy to control. Carbon source for this alternative may be provided by thermal hydrolysis of the enhanced primary sludge.

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