Effluent values and compliance with requirements

As from 1 December 1996, the Damhusåen WWTP had to comply with the effluent requirements stated in table 3 which correspond to the standards laid down in the Danish Action Plan on the Aquatic Environment. However, non-compliance with the effluent requirements would be accepted in a running-in period of 1 year if such was attributable to technical operating problems at the plant.

Parameter	**Damhusåen WWTP**	
	mg/l	ton/year
COD	75	3,000
BOD	15	600
Total N	8	320
Total p	1.5	60

Table 3: Effluent requirements for the Damhusåen WWTP

Since the beginning of September total nitrogen in the effluent from the plant has been at a level of 6.1 mg/l, which means that the limit value has not been exceeded.

During the period with biological phosphorus removal only total phosphorus in the effluent has been at a level of 2.2 mg/l. After 1 December 1996 when both biological and chemical phosphorus removal was performed, total phosphorus has been at a level of 1.3 mg/l, and consequently the limit value has been observed.

The limit value for COD and BOD has been observed during the period, as the concentrations in the wastewater discharged have on average been 40 mg/l and 4 mg/l, respectively.

Running in of the Lynetten WWTP

In the following, a brief description incl. status is given of the running in of the Lynetten wastewater treatment plant. The plant is in principle designed as the Damhusåen wastewater treatment plant, with biological phosphorus tanks which are the existing old aeration tanks in converted form. The treatment to remove nutrients takes place in new aeration tank sets of which there are 10 sets. The existing settling tanks have been supplemented with a set of new settling tanks, thereby increasing the settling tank capacity by 45%.

Plant data

Table 4 below shows the key plant data for the Lynetten WWTP.

Plant data	**Lynetten**
Catchment area	76 km^2
Design basis PE	750,000
Qmax. plant	41,500 m^3/h
Qmax. biology	23,000 m^3/h
Biological phosphorus tanks	24,300 m^3
Aeration tanks	147,000 m^3
Final settling tanks	62,000 m^3
Digesters	18,000 m^3
Primary sludge production	30 t TS/d
Biological sludge production	27 t TS/d

Table 4: Selected plant data for the Lynetten WWTP.

Floc formation and sludge build-up

Due to conversion of the existing plant, the capacity of the plant in the running-in period is 50% for the primary tanks, 50% for the biological phosphorus tanks and 60% for the final settling tanks.

The plant was put into operation on 6 August 1996. The activated sludge from the existing plant was pumped into the new plant. In the period from 9 August to 22 August, nitrifying sludge was pumped into the Lynetten WWTP from the Damhusåen WWTP to seed the plant. This was possible because the

Damhusåen WWTP used to function as a physical pretreatment plant for the Lynetten WWTP before the extension to nitrogen removal. A total of 100 t TS was pumped from the Damhusåen WWTP to the Lynetten WWTP.
To prevent sludge escape from the final settling tanks, the hydraulic load on the new biological plant was reduced from 23,000 m^3/h to 16,000 m^3/h. In the beginning of January 1997, the converted existing plant will be put into service again and the Lynetten WWTP will thus again have the full hydraulic capacity.

Nitrogen and phosphorus removal

In the beginning of September 1996, i.e. after approx. 28 days of operation, the ammonia concentration in the effluent from the plant was less than 1 mg/l and full nitrification had been established. The oxygen set-point in the nitrification phases has been 1.5 mg O_2/l during the entire running-in period.

The plant was then readjusted to phased operation with a 25% denitrification period and a 75% nitrification period. The nitrate level in the effluent dropped within a few days from 10 mg/l to approx. 5 mg/l.

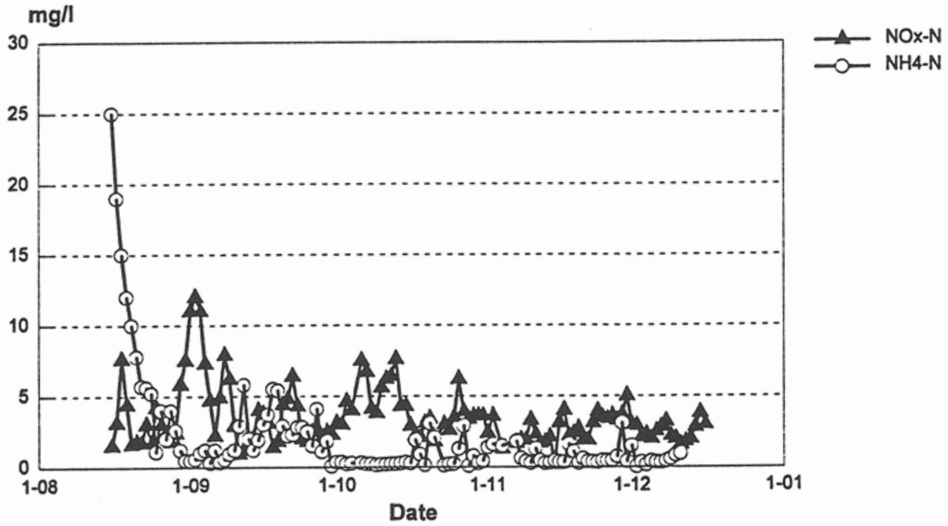

Figure 6: Ammonium N and nitrate N in the effluent from the plant.

To denitrify more nitrate, the plant was readjusted to a 35% denitrification period and a 65% nitrification period on 31 October.

As the biological phosphorus tanks do not have the full capacity, the running in of the biological phosphorus removal will not take place until in the first quarter of 1997. The Lynetten WWTP must first meet the new effluent requirements, shown in table 5, from 1 July 1997, and there is thus time for optimisation of the nitrogen and phosphorus removal when the full hydraulic capacity of the plant has been provided.

Effluent results and compliance with requirements

As from 1 July 1997, the Lynetten WWTP must comply with the effluent requirements stated in table 5 which correspond to the standards of the Danish Action Plan on the Aquatic Environment.

Parameter	Lynetten WWTP	
	mg/l	t/year
COD	75	6000
BOD	15	1200
Total N	8	640
Total P	1.5	120

Table 5: Effluent requirements for the Lynetten WWTP.

In the period from 1 November 1996 the nitrogen removal has functioned satisfactorily and the amount of total nitrogen in the effluent has varied around 7 mg/l. The limit value of 8 mg/l has thus been observed.

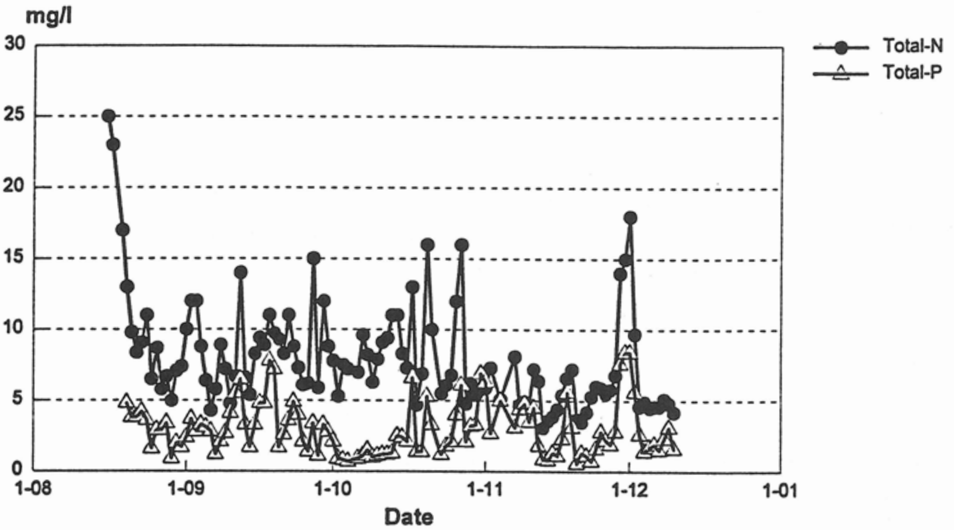

Figure 7: Total N and total P in the effluent from the plant

The few increased values of total nitrogen and total phosphorus in the effluent are caused by sludge escape and overflow of primary wastewater directly into the outlet from the final settling tanks during rain.

The biological phosphorus removal has also functioned satisfactorily considering that the capacity of the biological phosphorus tanks has only been 50% and that there has been a considerable recirculation of phosphorus with the return flow "filtrate" during the running-in period due to phosphorus stripping from the biological sludge in the thickeners. This recirculation will cease when the final plant for treatment of biological sludge has been established in the middle of 1997. The concentration of total phosphorus in the effluent has varied around 2.5 mg/l

The limit value for COD and BOD has been observed in the period as the effluent concentrations have been 70 mg/l and 4 mg/l, respectively, on the average.

Continued optimisation of the plants

In 1997 focus will be on the following fields of activity:

Continuous process optimisation of both wastewater treatment plants especially in the cold period from December 1996 to March 1997 according to which they are designed. At this process optimisation the following conditions will be included, among others:

- Observance of the limit values
- Minimisation of the energy consumption
- Minimisation of the chemical consumption and consumption of other auxiliary materials,
- Possibility of production of easily degradable carbon through hydrolysis of primary sludge for use in connection with the biological phosphorus and nitrogen removal

Forces will be joined with the participating municipalities to find and eliminate the sources of heavy metal discharge in the catchment area of both wastewater treatment plants. Besides, activities relating to the mapping of xenobiotic substances will start.

Plans are being made for establishment of a permanent pilot plant at the Lynetten WWTP. This plant will, among other things, be used for tests involving a risk of increased effluent values, especially considering the new Danish wastewater charges.

Summary

It has been possible to establish effective biological nitrogen and phosphorus removal at the Damhusåen WWTP so that the effluent requirements applying from 1 December 1996 are fulfilled. Thus, it has only been necessary to apply an insignificant dosing of iron sulphate to comply with the effluent requirements for total phosphorus.

At the Lynetten WWTP, the nitrogen removal has been provided within approx. 35 days. This can be ascribed to the fact that the seeding with nitrifying sludge from the Damhusåen WWTP has had a positive effect on the establishment of full nitrification. If seeding with nitrifying sludge had not taken place, full nitrification would not have been obtained until after approx. 60 days of operation, as demonstrated at previous pilot tests conducted at the Lynetten WWTP.

The amount of total nitrogen and total phosphorus in the effluent from the Lynetten WWTP has been 7 mg/l and 2.5 mg/l, respectively, in the period with nitrogen removal. The running in and optimisation of the nitrogen and phosphorus removal will continue in the first quarter of 1997 when the full hydraulic capacity of the plant has been established.

UPGRADING OF A WASTEWATER TREATMENT PLANT BASED ON A SITE-SPECIFIC TEST APPROACH

B Andersson*, H Aspegren*, U Nyberg*, J la Cour Jansen, H Ødegaard*****

* Malmö Water&Sewage Works, S-20580 Malmö, Sweden
** Water Quality Institute, Agern Allé 11, DK-2970 Hörsholm, Denmark
*** Dep.Hydraulic and Environmental Engineering, NTH, N-7034 Trondheim, Norway

Abstract

The upgrading process for enhanced nutrient removal at the Sjölunda wastewater treatment plant in Malmö has been based on a site-specific test approach where the goal was an optimal use of existing facilities. The entire wastewater flow could be nitrified in the existing trickling filters provided a high loaded activated sludge process preceeded the filters. Denitrification can be achieved in a fixed-film process implemented after the trickling filter plant after addition of an external carbon source. Sludge treatment with digestion is used for sludge stabilisation and energy recovery. Nitrification of the supernatant is used to save some of the external carbon source and decrease the ammonia load on the trickling filters. The upgrading scheme could be introduced with a minimum of new construction works and less investments and total costs than for an alternative based on a low-loaded activated sludge process approach. This paper demonstrates the importance of considering the site specific conditions and the local competence when a plant is to be upgraded for increased demands.

1. Introduction

Advanced treatment for high removal of BOD and phosphorus was introduced in Sweden as a general concept during the 1970's. The general policy prescribed combined biological and chemical treatment. Primary and secondary treatment were in most cases followed by tertiary treatment with post-precipitation in a separate stage. Secondary treatment in a high loaded activated sludge process was the dominating biological method but trickling filters were also used in some instances. The design rules did not take into account the need for further upgradings and the need for increased nitrogen removal was completely overlooked.

Large wastewater treatment plants in the south of Sweden have to be upgraded for future effluent standards of less than 10 mg BOD_7/l, 8 mg N/l and 0,3 mg P/l. The effluent nitrogen standard will be based on annual averages and the BOD and phosphorus standards will be based on monthly averages.

The standards and the general experience, expressed in the design rules, may suggest a process scheme at a wastewater treatment plant based on a low loaded activated sludge process designed for nitrogen removal and tertiary treatment in a filter to meet the stringent phosphorus standard. However, it has to be recognized that there is a strong economic incentive to investigate alternative process schemes which may not be examplified in the general design rules.

At the Sjölunda wastewater treatment plant in Malmö, an on-site test approach has formed the basis for the upgrading process. The approach can be defined as an upgrading based on optimal utilization of the existing facilities by alternative use of the existing reactors and on-site testing with the processes in question. The overall objective has been to find a concept that could meet the new standards at the lowest total cost. The paper will describe this work and the selected upgrading concept for enhanced nitrogen removal at the Sjölunda plant.

2. The Sjölunda Wastewater Treatment Plant

The Sjölunda wastewater treatment plant has already been upgraded twice. Large investments were made at the plant during the last upgrading for extended BOD and phosphorus removal in the end of the 1970's. The plant is designed for 550 000 population equivalents corresponding to an average wastewater flow of 1650 l/s. The present effluent standards are 15 mg BOD_7/l and 0,5 mg P/l.

The raw wastewater at the plant can be characterized as dilute. The concentrations during the last few years have on the average been around 200 mg BOD_7/l (90 g /cap*d), around 380 mg COD/l (170 g /cap*d), around 24 mg N/l (11 g /cap*d) and about 5 mg P/l (2.2 g /cap*d). The relatively high specific BOD and COD concentrations and the high ratio of BOD to COD can probably be attributed the fact that a number of food processing industries are connected to the wastewater treatment plant.

A schematic flow sheet of the existing plant is shown in figure 1. The plant is built in three stages with primary, secondary and tertiary treatment. The primary treatment comprises screening, grit removal, pre-aeration and primary clarification. Ferrous sulphate is added for phosphorus removal in the primary treatment. The flow is divided into two parallel treatment lines after the primary treatment, one treatment line with an activated sludge process and one treatment line with plastic-packed trickling filters. The two lines were designed for the same capacity. The biological treatment is followed by post-precipitation with flocculation and flotation. Alum is added to the effluent from the trickling filters. The effluent standards are reached for the water from the activated sludge process without coagulant addition in the precipitation plant. The sludge treatment comprises thickening, digestion and dewatering. Sludge digestion stabilizes the sludge and the produced methane gas is recovered in a gas motor plant for electric energy production. The dewatered sludge is used in agriculture.

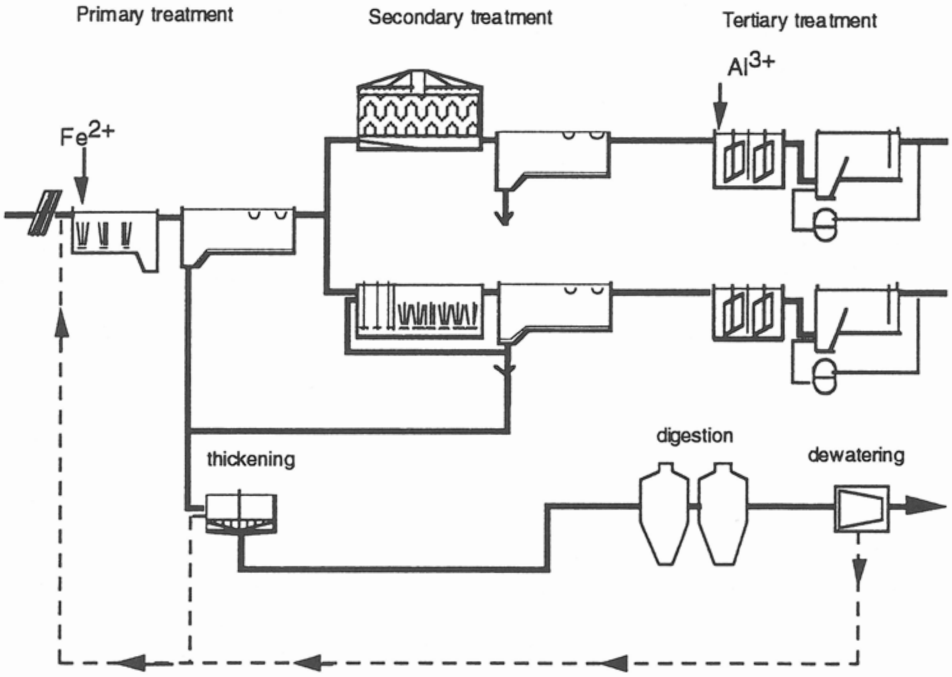

Figure 1. Schematic layout of existing plant at Sjölunda.

3. Upgrading Concept

A flow scheme of the upgrading concept at the Sjölunda plant is shown in figure 2. Ferrous sulphate is used for phosphorus removal and is added ahead of the primary clarifier. The high loaded activated sludge process is operated with a small anoxic zone which is used for denitrification of nitrified supernatant from the sludge treatment system. The trickling filters are used for nitrification and denitrification takes place in a separate post-denitrification reactor based on a moving bed process. Ethanol or methanol is used as carbon and energy source for the process. The sludge treatment system includes thickening, anaerobic digestion and dewatering in centrifuges. The ammonia in the supernatant from the sludge treatment system is nitrified and then returned to the head of the works.

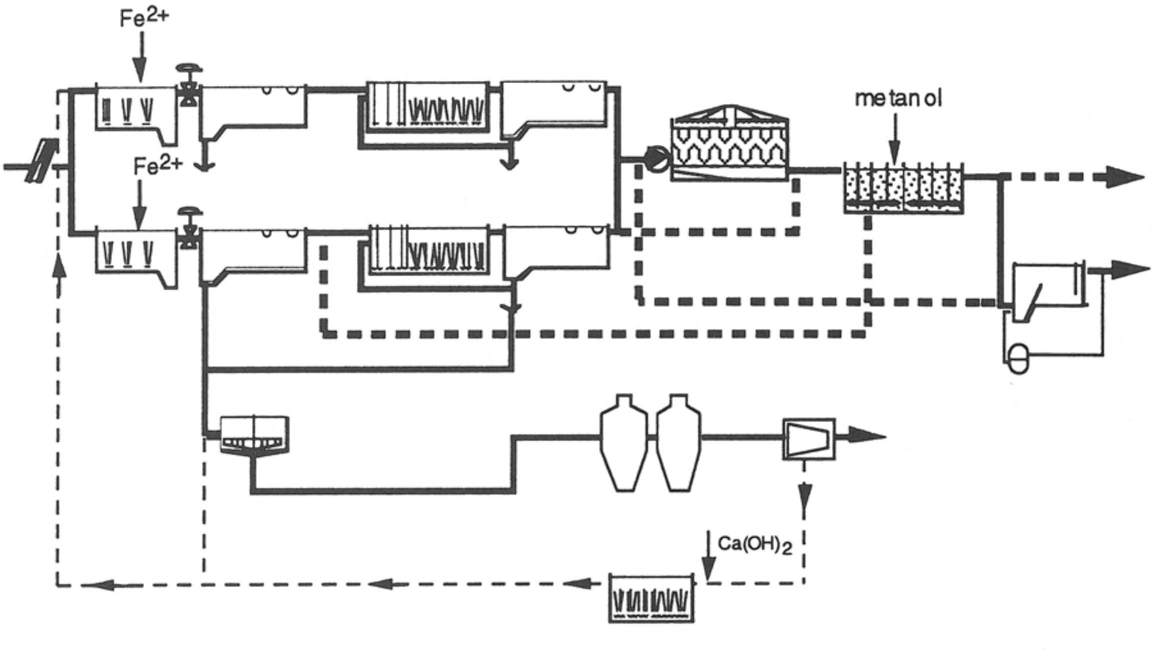

Figure 2. Upgrading concept at the Sjölunda plant.

The only new basin volumes that have to be erected, are a bioreactor volume of about 10000 m^3, a fixed-bed process for denitrification and a reactor for the nitrification of the supernatant.

In the following, the process scheme will be presented in some more detail and the rationale for selecting the different processes will be discussed.

3.1 Carbonaceous Removal and Nitrification

The starting point for the upgrading process was to investigate how to integrate existing plastic packed trickling filters in a concept for nitrogen removal. The trickling filter has an excellent oxygenation capacity. This implies that the filters should be used in an aerobic application and that especially the nitrification capacity of the filters has to be assessed. The nitrification capacity of the filter is to a large extent determined by the amount of organic material in the influent water. Consequently, a key feature in the use of trickling filters for nitrification is how to produce a water suitable for a process scheme with nitrifying trickling filters (NTF). At the Sjölunda plant, a NTF concept requires that high nitrification rates and low effluent ammonia concentrations can be maintained simultaneously.

In a preliminary study, the nitrification capacity in trickling filters with combined carbonaceous removal was investigated (Andersson, 1990). A high nitrification level was obtained when the wastewater temperature was above 16 °C. The nitrification level decreased however at lower temperatures. One of the conclusions was that the organic load on the filters has to be decreased in order to reach high nitrification levels even during low temperature periods.

The primary clarifiers at the Sjölunda plant are shallow resulting in poor performance, and completely new clarifiers are needed in order to improve the results. Consequently, a decrease of the organic load at the Sjölunda plant can only be achieved by two stage biological treatment where the carbonaceous removal takes place in the activated sludge plant and nitrification takes place in the filters operated in a NTF mode (Parker et al., 1990).

The activated sludge process produces a secondary effluent with consistently low concentrations of BOD7 and suspended solids. The concentration of suspended solids in the secondary effluent during the last years is presented in figure 3. The concentration is on the average 12 mg/l and although some peak values can be noticed the process must be deemed as being quite stable resulting in a quality suitable for further treatment in NTF:s.

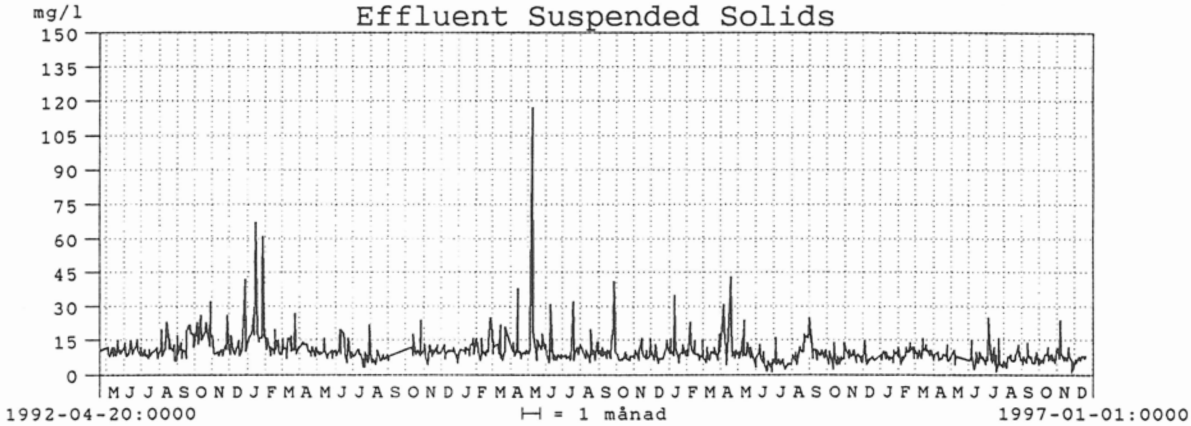

Figure 3. Concentration of suspended solids in secondary effluent.

As experiences with NTF operation were very limited, pilot plant tests had to be performed. Certain key factors were identified to be of importance for the overall application. Factors which required further attention were the effect of the influent concentration of suspended solids, the wastewater temperature and methods for biofilm and predator control. Biofilm control implies design features or operational procedures that secure the developement of a uniform biofilm without patchiness throughout the whole depth of the filter. A suggested way to get a more uniform biofilm was to operate two filters in an alternating series mode. Predator control implies design features or ways of controlling primarily filter fly larvae. The results from this intensive test phase have been presented previously by Andersson et al. (1994) and in the following only a short summary will be given.

The NTF operation mode was a very stable process configuration. The operation could be further improved by operating the filters in an alternating two stage filtration mode. This mode of operation seems to be an interesting way in order to maximize the use of a specific filter media volume. Very low effluent concentrations were reached even at ammonia loads of more than 2 g N/m^2*d.

Sludge loss from the secondary clarifiers in the activated sludge process affected the rate in a short term perspective in especially the upper parts of the filter. The nitrification capacity, however, recovered quite quickly from these incidents. The temperature also affected the nitrification rate but high rates and low effluent ammonia concentrations could be reached even during low temperatures.

Predators could not be shown to have any effect on the nitrification performance. The presence of filter fly larvae was limited throughout the whole study period. Filter fly larvae probably require a thicker biofilm than the one found in a NTF application. The filter fauna was dominated by Naididae worms which are well adapted for the NTF environment. Large number of snails appeared when the water temperature was high but they did not affect the performance and disappeared when water temperature decreased.

Finally, it could be concluded that predator control methods such as flooding or using different flushing intensities for controling the number of filter fly larvae, are not necessary design features of the NTF process.

To indicate the everyday stability of the process, the time series of the ammonia concentration in the influent and effluent water to the NTF system and the nitrification rate based on daily composite samples during 1994 are shown in figure 4. The filters were operated during this period without the previous intensive control. The hydraulic load on the filters during the first half of the year was equal to the total load on the existing trickling filters at design flow in full scale. During the second half of the year, the load was increased with 50 %. The supernatant from the sludge treatment system had been redirected away from the wastewater stream that enters the pilot plant. It can be seen from the figure that the influent ammonia concentration during summer was very low due to nitrification in the full scale activated sludge plant. The NTF performance however recovered very quickly when the ammonia load again increased.

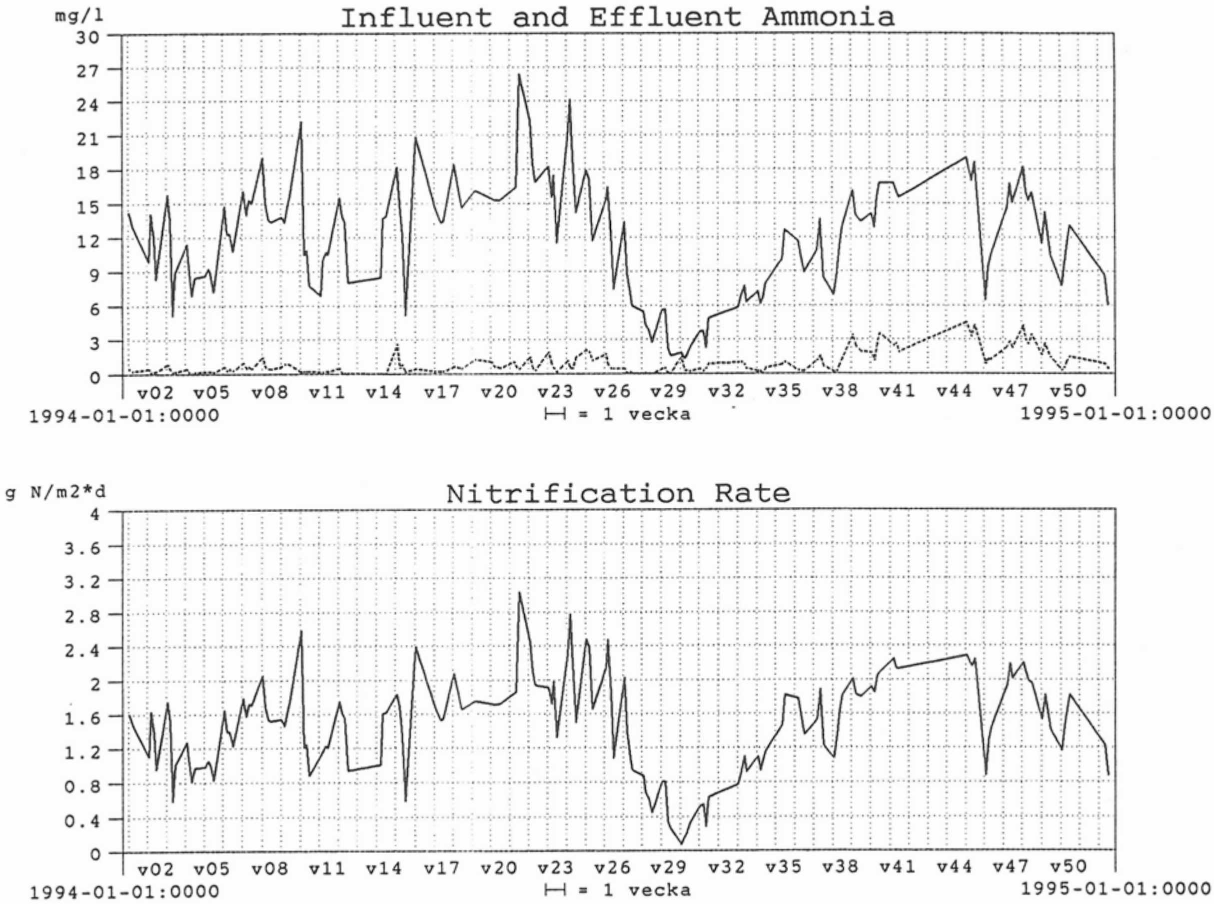

Figure 4. Ammonia in influent and effluent water and nitrification rate during the NTF pilot tests.

About 50 % of the wastewater flow is treated in the existing activated sludge plant and the load cannot be increased to any greater extent. When the entire flow is to be treated for BOD removal, the capacity of the activated sludge plant needs to be doubled. However, only new reactor volumes are required as the secondary settlers in the trickling filter plant can be used for this application when the operation of the plant is changed to a NTF mode.

3.2 Denitrification

In order to utilize the organic material in the influent water for denitrification, a process scheme where the water from the NTF:s is circulated back to the activated sludge process is often suggested (Kraut and Roth 1991, Lyngå and Balmer 1992, Maisch and Schwentner 1994, Schleypen and Nordmann 1994, Bever et al. 1994). In such a process the load on the secondary clarifiers will increase as a result of the circulation of nitrified water from the NTF:s.

Furthermore, the nitrified water will be almost saturated with oxygen which in turn will consume organic material. These inherent drawbacks affect the maximum recirculation ratio. A maximum level of around 100 - 150 % seems plausible and consquently it can be anticipated the overall nitrogen removal is in the range of 50 - 60 %. Apart from the fact that the requirement for nitrogen removal at Sjölunda (75 %) cannot be met with security with this process, new secondary clarifiers would also have to be built in order to accommodate the increased flow rate. Furthermore, water transition within the plant would prove to be rather complicated and expensive.

Utilizing the organic material in the influent water for denitrification is associated with large investments which have to be compared with the cost for a post-denitrification process where an external carbon source is used. In this case either methanol or ethanol can be used as carbon source. At the Sjölunda plant, a post-denitrification alternative is preferred both with respect to economy and technical feasibility.

Post-denitrification with an external carbon source is preferably performed in a biofilm reactor. The sludge production is comparatively low in a post-denitrification process and the control of the sludge inventory can in some cases be quite awesome as sludge loss from the clarifier cannot be avoided (Andersson and Aspegren, 1990).

A biofilm reactor can be introduced ahead of the the existing post-precipitation plant. Full-scale experiments have indicated that the combination of a fixed film process for denitrification and flotation for separation of suspended organic particles has great potential (Andersson and Aspegren, 1990). One of the more interesting biofilm carriers in this context is the use of suspended plastic carriers as used in the KMT moving bed biofilm process (Ødegaard et al., 1994).

A biofilm process at the Sjölunda plant can be introduced in a post-biofilm reactor based on a sand filter. Experiments in half technical scale with denitrification of secondary effluent in an upflow sand filter have demonstrated the viability of this process (Andersson et al., 1991). With this type of process, the operation of the flotation plant will not be required during normal flow conditions.

3.3 Phosphorus Removal

In order to be able to comply with the rather stringent requirements for phosphorus removal, it is necessary to remove phosphorus down to a concentration of around 0.5 mg/l upstream of the post-denitrification process. This can either be accomplished by means of chemical precipitation or enhanced biological phosphorus removal (EBPR) in the high loaded activated sludge process. EBPR operation in a high loaded activated sludge plant has been studied at the Sjölunda plant in pilot scale since 1991. Considering the overall effects, the precipitation process will in this case be the more simple and economic alternative at the Sjölunda plant.

3.4 Nitrification of Supernatant

As about 50 - 70 % of the volatile solids are degraded during digestion, the supernatant from the sludge treatment system contains high concentrations of ammonia. This ammonia, which represents some 15 % of the influent nitrogen load to the plant, can be nitrified in small volumes as both the temperature and concentration are high. Two basic alternatives for nitrification of the supernatant at the Sjöunda plant are discussed - a chemostat alternative and a sludge circulation alternative. In both cases, a SRT of around 2 days is considered. In the chemostat alternative, the SRT represents the actual detention time. The alkalinity in the supernatant covers only about 60 % of what is needed for complete nitrification and alkalinity has to be supplied to the process.

With an anoxic zone in the high loaded activated sludge process, the amount of organic material which has to be supplied in the post-denitrification process decreases. Furthermore, the ammonia load on the filters will decrease as well. This in turn will increase the stability and secure that low ammonia concentrations consistently are obtained in the effluent water from the filters.

4 Discussion

It could be argued from a technical standpoint that the suggested concept is complicated with a great number of processes. However, when looking at the present plant upgraded twice already, it is not very complicated. The process scheme comprises well known "workhorses" in Sweden like the high loaded activated sludge process for BOD removal, chemical precipitation for phosphorus removal, trickling filters and anerobic digestion. The only new process is actaully the post-denitrification plant where an external carbon source, ethanol, is to be added to the wastewater. The fact that the different processes have been divided on different reactors may actually enhance the understanding of the overall treatment scheme. Furthermore, the fact that ferrous sulphate is used for phosphorus removal and that an external carbon source is used for denitrification makes the plant operation rather insensitive to the inherent wastewater variations and a stable operation can be expected.

This upgrading will be the third since the plant was built in 1963. The fact that the demands and the load constantly change, stress the importance of having a flexible plant structure which allows future upgradings and load expansions. The wastewater may be said to reflect the infrastructure of the city and when the structure changes, the wastewater also will change.

There are primarily two infrastructural changes in Malmö which can be foreseen that may alter both the wastewater composition and the load. During the last few years it has been observed that the food processing industry in Malmö, which probably contributes with a good portion of the easily degradable organic material in the wastewater, is subject to structural changes. This may have a severe impact on a process scheme where the raw wastewater is to be the sole source of organic material. Furthermore, the bridge shortly to be built between Malmö and Copenhagen also will influence the entire infrastructure in Malmö.

Another uncertain area is the sludge handling possibilities in the future. In this perspective a high loaded activated sludge process combinbed with EBPR has potential. EBPR can easily be introduced in the main stream process in the upgrading concept.

The upgrading concept was tested in an independent competition during 1994 when the city of Malmö decided to test the efficiency of the Malmö Water Sewage Works. They consequently asked the Malmö Water Sewage Works and a number of Swedish and international companies for tenders for the operation of the entire Water Sewage Works and for the upgrading of the Sjölunda wastwater treatment plant for nitrogen removal. In all, seven consortia responded to the invitation and participated in the competition. An independent consultant was comissioned by the city to evaluate the tenders and the upgrading alternatives. The evaluation showed that the upgrading scheme suggested by the Malmö Water & Sewage Works was the most cost effective upgrading alternative.

One important aspect gained from this experience is that the site specific conditions and a local competence are very important ingredients when a plant is to be upgraded for increased demands. The NTF concept suggested for the Sjölunda plant is based on a profound knowledge about the wastewater quality and the existing plant. Operational experiences with the suggested processes have been obtained either in full scale or in half technical scale within a context where the local standards have been contemplated and included.

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The VEAS-concept, a system for N and P removal at a total retention time of 4 hours. Status by the end of 1996.

Paul Sagberg, Kirsti Grundnes Berg, Pia Ryrfors og Eivind Kvamme Jensen, Vestfjorden Avløpsselskap-VEAS, Norway.

1. Introduction.

The VEAS-WWTP, serving most of the city of Oslo and three neighboring communities, was originally a pre-precipitation plant for phosphorus removal, put in operation in 1982, serving approx. 650,000 person equivalents. The plant removed by the end of the 1980's 97% P, at a total retention time of 3 hours. Because of the low oxygen levels in the deep waters of the inner Oslofjord, VEAS started tests with volume effective biofilm techniques in 1986. The first goal was the removal of all biodegradeable organic material, later extending this to include nitrification. After the Norwegian signing of the North Sea Agreement, the goal was extended to at least 70% nitrogen removal. The plant is located inside rock caverns with small possibilities for extention. It was decided to try to fit the new processes into the existing sedimentation tanks, only extending the depth. During the development process, for each change, mass balances have been recalculated for the entire process to try to understand the full impact of each change.

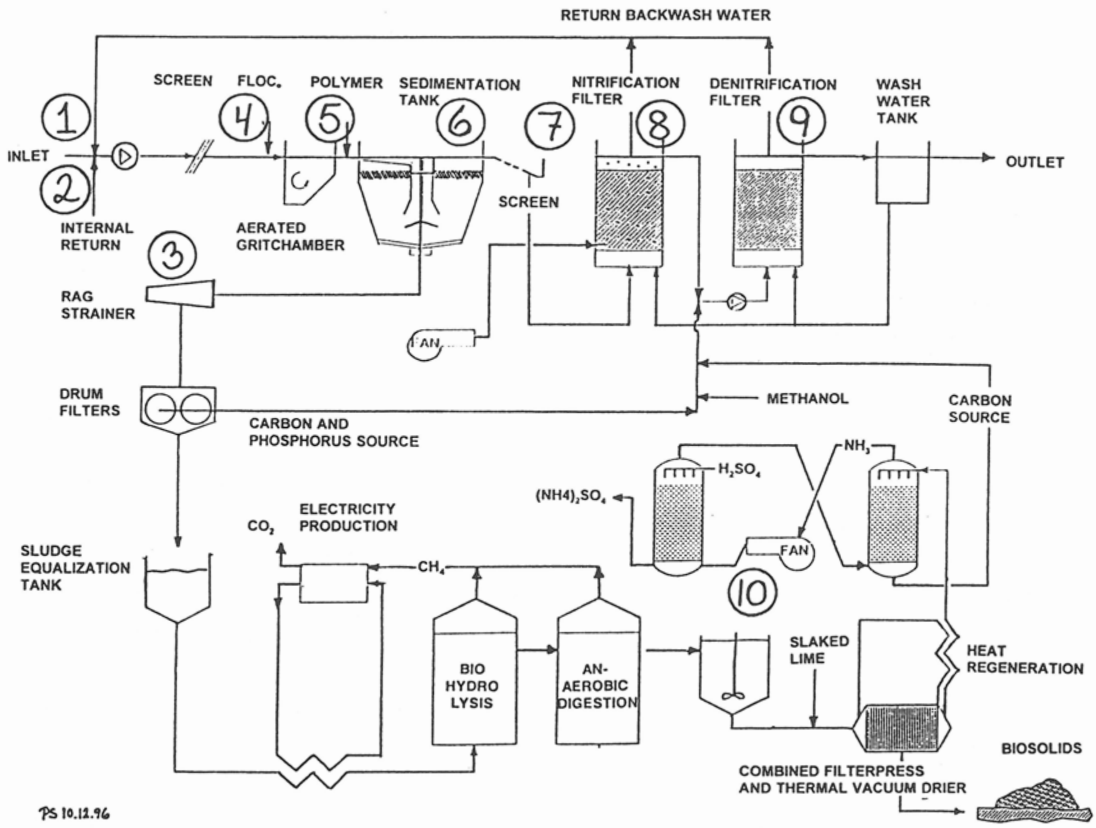
The funding and political decisions took a long time forcing the research and development work to run in parallel with the planning and building phase. Status of this work has been presented to this symposium earlier [1,2,3]

2. The VEAS concept for nitrogen removal by the end of 1996.

Figure 1 presents the VEAS concept as it is believed to end up. Most of the process is already implemented in full scale. The numbers in the figure are refering to the subnumbers of paragraph 3. The process consists of pre-precipitation with prepolymerized aluminum chloide in 11 meter deep, short tanks, followed by upflow BIOFOR biofilters for nitrification and post-denitrification, with both internal and external carbon sources for denitrification.

Sludge from sedimentation tanks is thickened in drum filters. The thickened sludge is heated to 37°C and hydrolyzed for 2 days prior to digestion for methane production, used for electricity and heat production. The digested solids are dewatered and possibly vacuum dried in a filterpress after lime addition and used as soil conditioner. Ammonia in filtrate is stripped off and absorbed in hydrosurfuric acid. The stripped filtrate, rich in dissolved organic material is used as carbon source in denitrification. Filtrate from drum filters, rich in phosphate and organic material could possibly also go to the denitrifying filters.

THE VEAS CONCEPT



Figur 1. The VEAS-concept for nitrogen removal by Desember 1996, as it seem to end up.

3. Status for the development, what is achieved and what is to be investigated.

3.1 Incoming sewage.

The diurnal flowpatterns have been studied. Peak shaving or equalization can be achieved by regulating pumping stations and gates in the 200,000 m3 tunnel reservoirs. This is still not implemented as it would be achieved at the cost of storing capasity for heavy rainfall. Regulating the discharge pattern from major industrial contributors has not yet been investigated.

3.2 Internal return streams.

Initially the decant from the gravitational thickeners and the filtrate water from the final sludge presses were returned to the headworks. Later they were collected in a 1500m3 tank and dosed into the inlet stream. The load of suspended solids were as much as 40% of the incoming sewage, it also contains a high load of free ammonia and dissolved carbon.

These return streams are now going to be handled in a different way. The ammonia in the filtrate water from the sludge presses, at high pH after the lime conditioning, is air stripped to a hydrosulfuric acid absorption tower. 10-15% of the incoming nitrogen to the plant will be removed as ammonia sulphate. The stripping plant will be started in january 1997 reducing the ammonia load on the nitrifying filters. The treated filtrate, rich in dissolved organics, is directed to the denitrifying filters as carbon source, thereby also reducing the load of organic material on the lower part of the nitrifying filters.

In the old gravitational thickeners, fermentation of sludge started. Dissolved organics and a large amount of particulate matter in the decant disturbed the performance of the sedimentation tanks and the nitrifying filters. To reduce this effect drum thickeners were installed in line between the sedimentation tanks and one gravitational thickener. With the addition of 1-1.6 kg of the same anionic polymer used in the pre-precipitation process, the sludge could be thickened to 8-9% dry solids. The drum filters have given several benefits, the retention time in the digestors is increased, more organic material give more methane, less energy is needed for heating the sludge and easier dewatering is achieved in the filterpresses. On the sewage treatment side less particulate material need to be separated in the sedimentation tanks and less dissolved organic material to be degraded in the nitrifying filters. The filtrate from the drum filters will probably be routed to the inlet of the denitrifying filters to avoid the return load earlier in the prosess. A benefit could also be to increase the load of phosphorus which can often be the minimum factor for the denitrification process.

In addition to these return streams the washwater from the nitrifying and denitrifying filters are important. Approximately 7% of the incoming hydraulic load is today recirculated to the headworks. We hope to reduce this further to 5% or lower. The washwater contains 12-14% of the Tot-P, Tot-N and TOC of the incoming sewage mainly in particulate form. Other return streams such as spill from the lime slaking and hydrochloric acid from washing of filterpresses are still collected in a 1500 m3 tank and dosed into the process.

3.3 Rag problems.

Earlier, rags from the gravitational thickeners were floated away with the decant water to the equilibration tank and returned to the inlet, ahead of the 3 mm step screens. After introduction of the drum thickeners, the rags follow the sludge through the digesters causing problems for the filter press piping and pumps. In- line strainers are being mounted ahead of the drum filters to remove rags and fibers. The equipment will be put in service in February 1997 and will hopefully reduce the problem significantly.

3.4 Flocculants for preprecipitation.

Through an extensive research project funded by the Norwegian industrial development fund, Ferriklor, a company in the Kemira group and VEAS, more than a hundred new flocculants were developed and tested. The aim was to achieve good partical separation without removing too much ortho-P and alkalinity. The first needed in both the nitrification and the denitrification processes, the latter sometimes a limiting factor in the nitrification process. In addition the sludge produced should be at a minimum, thicken easily and be subject to good dewaterability. The process should work well in the temperature range from 2-16°C. Prepolymerized aluminum with high OH^- to metal ratio, with addition of small amounts of calsium, performed superb with the sewage quality at the time of testing. After all the process alterations have been implemented, a new round of tests should be staged. The aim should be to reduce the metal dose and increase the level of ortho-P for the biofilm steps. The dosage is regulated empirically according to the effluent turbidity level, trying to keep it at 8-10 NTU. A good dosing strategy to optimize the flocculant dosing is still missing. The flocculant is mixed with air and added to the sewage in a mist form from the bottom of the inlet channel to the grit chamber.[4,5,6]

The agitation in the zone just following the initial dosing is important, as well as the energy of flocculation along the aerated grit chambers. Several changes will be implemented throug the spring of 1997. The sewage speed into the gritchamber should be stopped and the aeration intensity increased. Grease removal in the grit chambers, earlier moved to the sedimentation tanks, should be reestablished, sending the grease directly to the anaerobic digestion.

3.5 Polymer addition.

Anionic, high molecular weight polymers are added to help floc formation. The dosing point has been shown to be very important. Dosing at the outlet of the gritchamber from a perforated crossbar has been a very successful technique at surface loads up to 8-9 m/h. Floc destruction appeared at higher hydraulic loads. By moving the dosing point closer to the inlet to the sedimentation tanks it was possible to increase the load on the sedimentation tanks to 10-13 m/h. The best method could be to have two or three different dosing points and let the computer select the best point with help of on/off valves according to hydraulic load.

3.6 Sedimentation tank design and performance.

The sedimentation tanks are 15.7m x 17.7m, 11 m deep with short (63cm) lamellas. They now function well up to more than 11m/h, when load of suspended solids from returns are kept low. Today functioning well means effluent turbidity below 8-10 NTU. Our goal has been to keep this turbidity with loads up to 12 m/h, but we have now lifted the goal to 15 m/h by the end of 1997. All the 6 sedimentation tanks have different designs for historical reasons. This gives us usefull information. It will take a long time before we have implemented the best designs in all tanks. The volume beneath the lamellas seem to be too large, causing strong currents. Pilot tests with units 1.5m x1.5m did not have the same problems. A solution could be to introduce vertical sectioning elements below the lamellas.

3.7 Inlet screens in front of nitrification.

To protect the nozzles at the bottom of the nitrification filters from clogging, the effluent from the sedimentation tanks passes through perforated plates with 2.5 mm round holes. Cleaning these plates is today too time consuming. We hope the problem will vanish as a result of other actions described in this presentation, among them the grease removal from the gritchambers.

3.8 The nitrification filters.

The filters are of the Degremont BIOFOR design. 24 filters 87m^2 each with 4 meter media thickness and aerated by coarse aerators 50 per m^2, a few with 25 per m^2. The aerators are located in a 30 cm thick layer of gravel above the nozzles.

Through a project financed by Norsk Leca, Degremont, VEAS and the Norwegian industrial development fund, very effective expanded clay aggregates were developed and tested. High density, high strength, porous, crushed media were superior [3]. Partly as a result of this development the number of filters in the plant could be reduced from 32 to 24. 16 of the 24 filters now have these best materials.

The performance of the filters can be limited by many factors. The load of suspended materials, the dissolved oxygen level, the load of dissolved organic material, the alkalinity, the phosphorus concentration, the temperature etc. The filters perform better than expected at the time the system was selected. The bacterial culture seem to adapt to the conditions, but the adaptation takes time. Seeding the youngest filters with washwater from the oldest help. Still after 2 years the biofilm seem to develop further. The biofilm is only 10-30 um thick and oxygen transportation is mainly the limiting factor. Enriching the air with pure oxygen gave immediately higher removal rates near proportional to oxygen concentration [2]. The filters seem to function better at higher hydraulic loads. This fact can be utilized although we do not think it is neccessary at VEAS.

The filters will have higher capacity than the load at our plant, at least when we add some alkalinity. The removal rate can at least reach as high as 0.8 kg NH4-N/m^{3}d. The average needed at VEAS to achieve 70% removal is probably around 0.45 kg NH4-N/m^{3}d. The effect of low temperature is uncertain. The lowest temperature is seen during snow melting. At that time we have high hydraulic load and low ammonia concentration. The air dissolves better at low temperature. During the last two snow melting seasons the percentage of nitrification has not been reduced.

Our two major challenges are to regulate the aeration to minimize the energy cost and minimize the uptake of oxygen after the water leaves the filtermedia on its way to the denitrifying filter. Excess oxygen must be removed by the help of a carbon source at high cost. We plan to change the design of the top of one of the nitrifying filters in 1997, but further development will take time. It should be possible to reduce the influent oxygen concentration to the denitrification-filters by 2 mg/l.

The washing frequency of the nitrification filters is today once every 26 hours, this can probably be reduced to once every 36 hours within a few months as a result of other changes in the process. We have, however, a couple of times encountered problems with formation of a jellylike substance clogging the bottom of the filters. This can perhaps be attributed to excretion of extracellular polysaccarides by stressed bacterias or be a function of our polymer usage.

The low effluent concentration of total phosphorus from the nitrifying filters is a problem. The average concentration is well below 0.1 mg total-P/l causing unstable denitrification.

3.9 The denitrification filters.

The plant has 24 denitrifying upstream BIOFOR filters, each 65 m2 with 3m media thickness without gravel. The filter material from norsk Leca is of the rounded high density, high strength expanded clay type.

The denitrification prosess in full scale has suffered from problems even though all pilot testing ran smoothly. The problem seem to be lack of phosphorus. For long periods the process ran well, but then the denitrification rate slowly decreased. This could be stopped and reversed by adding small amounts of phosphoric acid. The effect seem to be more important at lower water temperatures. The cost of adding small amounts of phosphoric acid is not large, but being a plant for phosphorus removal we try to avoid adding external phosphate. The aim is to increase the amount of internal phosphate to the denitrification- filters. When phosphorus is not the limiting factor, the filters can remove all the nitrate present with very good margins. Probably we could have reduced the the size of the filters significantly or reduced the number.

The neccessary amount of methanol in gram methanol pr. gram nitrate-N equivalents removed, is approx. 2.2-2.6. With the low phosphate load the bacterias hydrolyze dead cells to liberate phosphate and dissolved organic material thereby causing low sludge production. The ammonia consumption in the denit-filter for cell production is approx. 1 mg/l.

The washing frequency of the denit-filter is today once every 42 hours.

5-10% of the total incoming nitrogen leaves the plant as non-biodegradeable or very slowly biodegradeable nitrogen compounds.

3.10 Anaerobic sludge handling.

A lot of work has been carried out to see if thermal hydrolyzis of the anaerobic sludge could be beneficial in the total process. High investment cost, corrosion and odour problems have reduced the interest. Instead a drying process in the existing filterpresses is being studied. An American company, DryVac, has patented a process where they press the sludge before they apply vacuum to the filtrate side and hot water to membranes on both sides of the sludge cakes. Within a cycle time similar to normal filter cycle times, they can achieve close to complete dryness. Methods for recovering the heat of evaporization is studied. Successfully recovering this heat, the technique will have a lot of positive effects. Cells will be hydrolyzed and the dissolved carbon can be returned to denitrification. Dissolved ammonia is separated from the sludge and can be removed in the stripping process. Odour will probably not cause problems neither in production nor in storage or spreading. Transportation will be significantly reduced. The storage capasity in the plant will be increased. A high grade of hygienization is achieved. Limedosage and washing frequencies will probably be reduced. If we select this method, it could be fully installed before the end of 1997.

4. References

The research work at VEAS has been documented through more than 60 reports, mostly in the Norwegian language. They can be ordered from VEAS at printing costs.

[1] Sagberg, P. et al. (1991) *Nitrogenrensing med biofilmprosesser.* Seminar, Oslo januar 1991, Nordiska Vattengruppen, Nordiska ministerrådet.

[2] Tandberg, I., Ydstebø, L. and Sagberg, P. (1992) *VEAS-konseptet- et kompakt nitrogenfjerningsanlegg.* Seminar, København september 1992,

Driftsättning och driftstrategier vid Henriksdals reningsverk

Lars-Gunnar Reinius och Johan Ståhl
Stockholm Vatten AB

Bakgrund

Utloppsledningarna från Stockholm Vattens tre reningsverk Henriksdal , Bromma och Loudden mynnar ungefär i samma punkt i den inre delen av Saltsjön. Istället för reningskrav för respektive anläggning är villkoret för utsläpp formulerat som ett gemensamt sammanslaget krav för de tre anläggningarna. Därmed har investeringarna för att uppfylla de skärpta kraven kunnat göras på de anläggningar där det gett mest effekt. För Henriksdalsverket, som är den största anläggningen i regionen, har de största investeringarna gjorts med utbyggnad av biosteget och en ny filteranläggning. På Bromma verket har endast filter byggts ut. Det biologiska steget har endast kompletteras med anoxiska zoner för kväverening. På Loudden har det biologiska steget uppgraderats med nytt luftningssystem och mekanisk omrörning i de anoxiska zonerna.

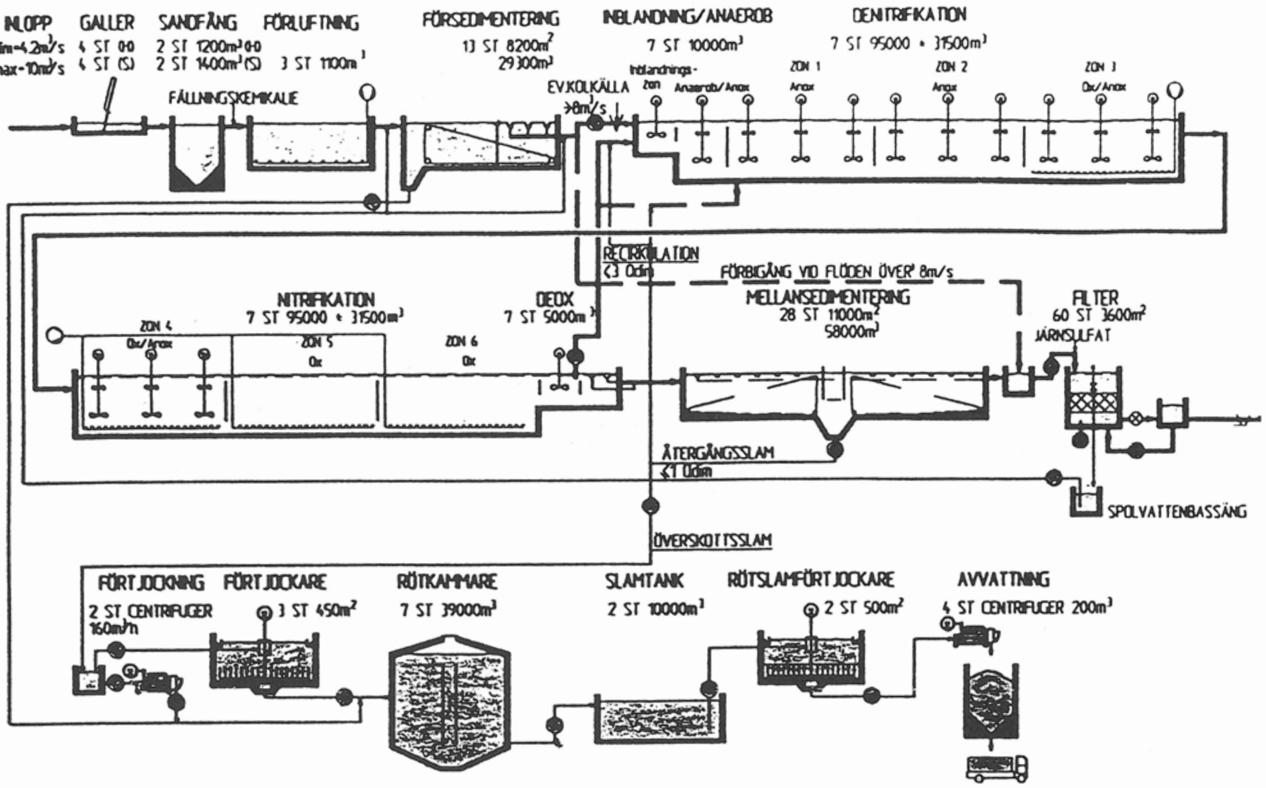
De tillfälliga reningskrav som gäller fram till att de slutgiltiga villkoren fastställs är för fosfor 0,3 mg/l och för BOD_7 10 mg/l räknat som kvartalsmedelvärde. För kväve är kravet 15 mg/l som årsmedelvärde och för ammonium kväve 3 mg/l under perioden juli till oktober

Henriksdals reningsverk är det största i Stockholm. Det betjänar ca 600 000 personer boende i Stockholms centrala delar, huvuddelen av de södra förorterna samt delar av kommunerna Nacka, Tyresö och Huddinge.

De äldsta delarna av Henriksdal är från 1941. Det biologiska steget som tagits i drift i början på 1970-talet var hårt slitet med ett mycket dåligt luftningssystem. Även andra delar i anläggningen såsom elförsörjning, ventilation m m var i stort behov av renovering. Försök på anläggningen med kvävereduktion startade 1985 med mål att utvärdera ett nytt luftningssystem och möjligheter till kvävereduktion. Försöken visade att processen nitrifikation/denitrifikation krävde större volymer än de tillgängliga 68 000 m^3 i de befintliga luftningsbassängerna. Dessutom utgjorde de hårt belastade sedimenteringsbassängerna en flaskhals eftersom slamhalten i luftningsbassängerna inte kunde höjas utan risk för slamflykt.

En jämförande kalkyl över olika processalternativ gav vid handen att en utbyggnad av volymerna i det biologiska steget samt en säkerställd avskiljning av partiklar genom filtrering var fördelaktigast. Förutom reningskraven och kostnaderna för investering och drift togs särskild hänsyn till driftfrågor såsom en stabil och resurssnål process.

Utbyggnad i berg sätter stora begränsningar i planlösningen av bassängerna. Det tillgängliga utrymmet i Henriksdal tillät endast att fem nya linjer med bergrum fick plats. Den slutliga designen stannade på en fördjupning av de 11 befintliga luftningsbassängerna, från 5 m till 12 m, en utbyggnad av tre nya 12 m djupa bassänger och 60 st tvåmedia nedströmsfilter.

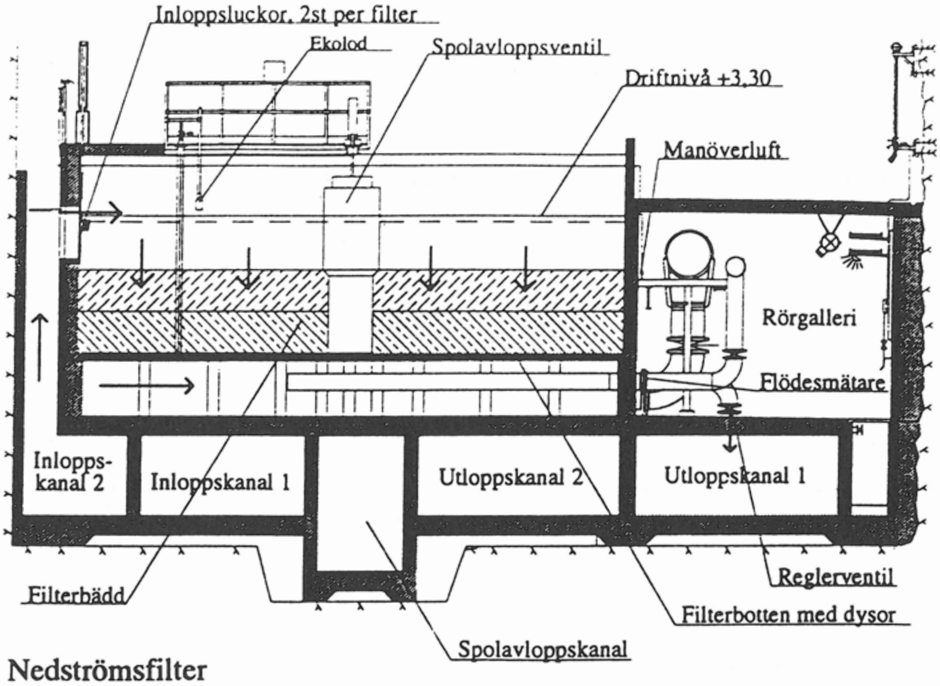


Figur 1: Processchema

De 14 bassängerna är indelade i sju block om vardera två biologiska reaktorer och två sedimenteringsbassänger. Varje block är delat i sex zoner. De två första är alltid anoxiska. De fyra efterföljande är försedda med luftningssystem. Två av de luftade zonerna är även försedda med mekanisk omrörning. Nitrat kan återföras genom pumpning från slutet av luftningen och med retursslam.

Luftningssystemet kan ge 60 000 Nm^3/h med fyra kompressorer i drift. Ytterligare en femte maskin finns i reserv. Kompressorerna håller ett tillräckligt tryck i en manifolder för att distributionen skall kunna ske ut till varje zon. Varje luftad zon är försedd med en flödesmätare, reglerventil och syremätare. På detta sätt hålls god kontroll över syrehalt och luftfördelning.

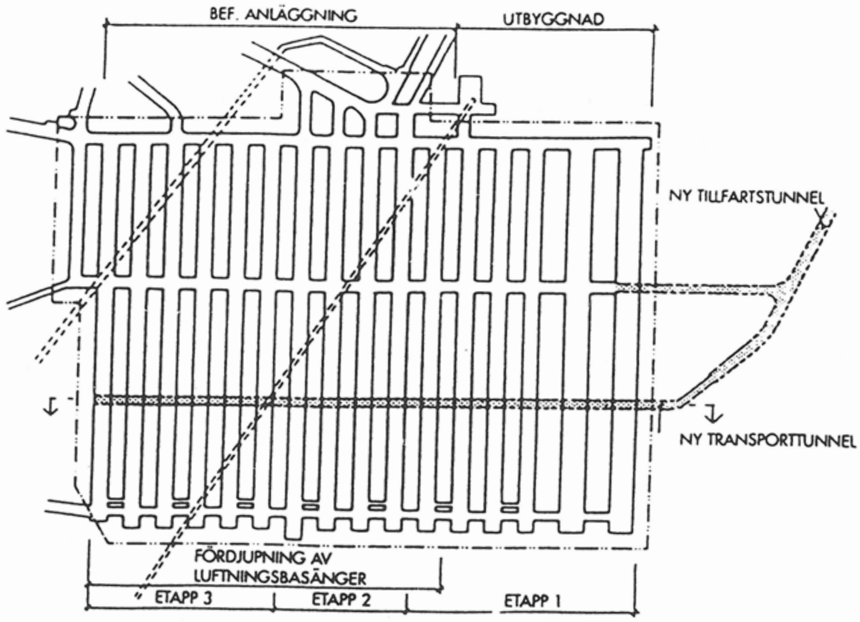
Filtren är av nedströmstyp med två medier. Ett 0,5 m djupt lager av sand i botten med korndiameter 1,2–2,0 mm . Det övre lagret är 1 m djupt och består av ett lättare grövre material av expanderad lera, diameter 3–4 mm. Filtrens uppbyggnad framgår av figur 2 på nästa sida.



Figur 2: Nedströmsfilter

Utbyggnad i etapper

Utbyggnaden startade 1992 och har skett parallellt med att övriga anläggningen varit i full drift. Byggnadsarbetena har därför uppdelats i tre etapper, se figur 4. I den första etappen byggdes filtersteget samt de tre nya biologiska reaktorerna. Dessutom togs en av de gamla luftningsbassängerna ur drift och fördjupades. Denna etapp driftsattes under perioden november 1994--mars 1995. Under etapp 2 fördjupades ytterligare 4 bassänger. Under denna period var både nya och gamla anläggningsdelar i drift. Etapp 2 togs i drift under vintern 1995/1996. När etapp 2 driftsatts kunde de återstående bassängerna tas ur drift och fördjupas.



Figur 3: Etappindelning av utbyggnaden

Dessa bassänger kommer att tas i drift under våren 1997. Under utbyggnadstiden har flera begränsningar i anläggningen påverkat driften. Den största begränsningen har varit den låga kapaciteten på eftersedimentering. Eftersom sedimenteringsbassängerna hör ihop med motsvarande biologiska reaktorer har även dessa varit avstängda under utbyggnaden. Detta har lett till en strategi med att endast ta in ett flöde till biosteget som kan hanteras och låta det övriga gå direkt på filtren. Därigenom har slamflykt och urspolning av biosteget förhindrats.

Driftsättning

Driftsättning av de nya anläggningsdelarna sker, liksom utbyggnaden, i tre omgångar. Driftsättningen har ingått i projektet och en speciell organisation har inrättats för denna. I organisationen ingår både egen personal och externa konsulter. En viktig del i driftsättningen har varit kontrollen och verifieringen av prestanda och funktion. Vid dessa kontroller som finns beskrivna i förfrågningsunderlagen är även entreprenören närvarande. Kontrollen består av

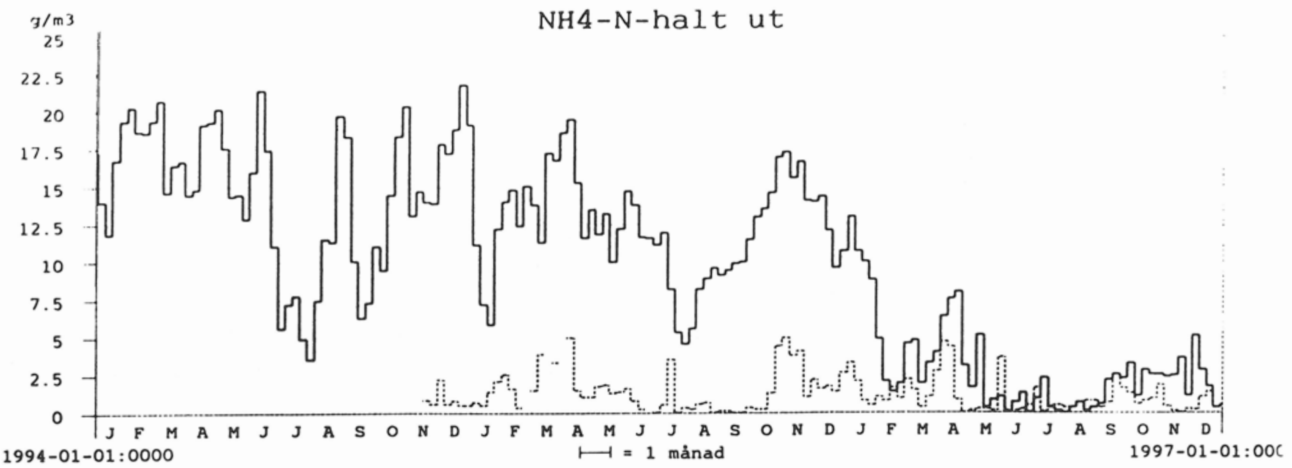
* Egenkontroll
* Samordnad egenkontroll
* Funktionskontroller
* Slutbesiktning
* Driftsättning
* Funktionskontroller, prestandaprover
* Funktionsbesiktning

Vid varje kontroll skrivs protokoll så att det finns ett dokumenterat underlag för framtida behov.

Driftsättningen har planerats i detalj och samtliga berörda har fått en pärm med alla aktiviteter beskrivna. Här finns redovisat vem som är ansvarig, vilka som skall medverka, hur aktiviteten skall utföras och när den skall ske.

Dagens anläggningar som är uppbyggda med en hög grad av automatisering och stora möjligheter till reglering kräver en mycket noggrann kontroll av alla funktioner. Fel i grundläggande funktioner kan i efterhand skapa många problem som är svåra att spåra orsaken till.

Praktiskt har driftsättningen skett med ett block i taget. Hösten 1994 togs det första biologiska blocket i drift. Aktivt slam pumpades i form av överskottslam från de 10 gamla bassängerna som fortfarande var i drift. Slammet i den gamla delen var delvis nitrifierande så nitrifikationen kom igång nästan omgående. Blocket belastades under den inledande fasen med ett lägre flöde än det framtida. Ett bekymmer som genast uppstod var att luftningssystemet ej kunde hantera de låga luftflöden som krävdes när endast en bassäng var i drift. Kompressorerna, som är fem i antal, har en kapacitet på 15 000 Nm^3/h vardera. De kan regleras ner till ca 40% av maximal kapacitet. Detta var dock fortfarande för mycket. En friblåsningsventil fick monteras så att trycket kunde hållas inom tillåtna gränser.



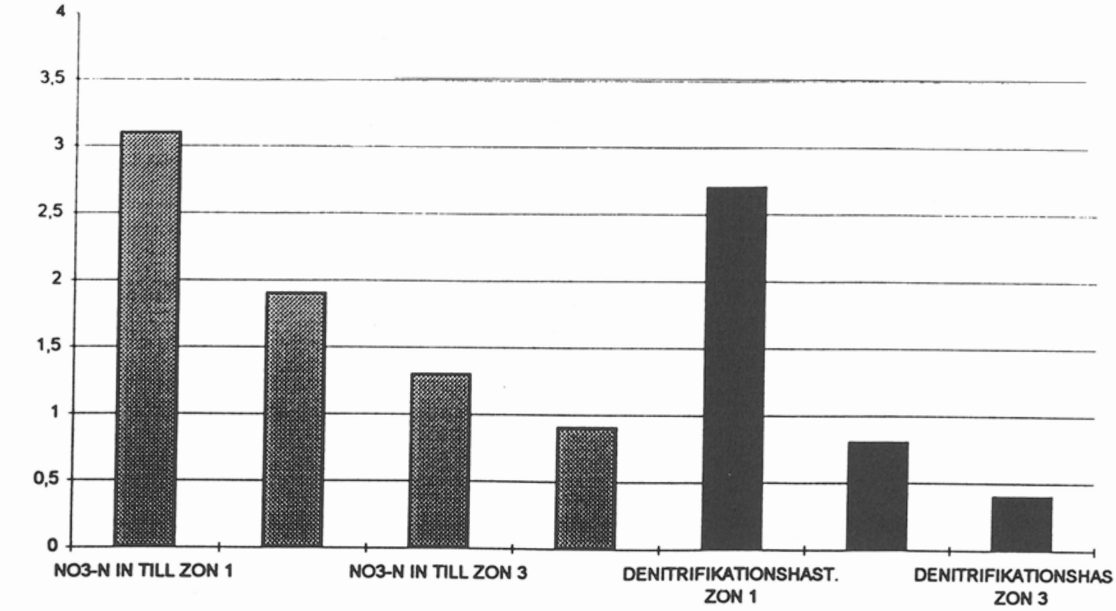
Figur 4: Driftstart av block 7 under november 1995, ammonium ut från block 7 och övriga anläggningen.

Driftsättningen av de tre efterföljande blocken har skett på motsvarande sätt. Slam har inympats från angränsande bassänger. Processen har därvid kommit igång snabbt.

Filtren togs i drift under våren 1995. Fällning med järnsulfat direkt på filtren startade under hösten 1995.

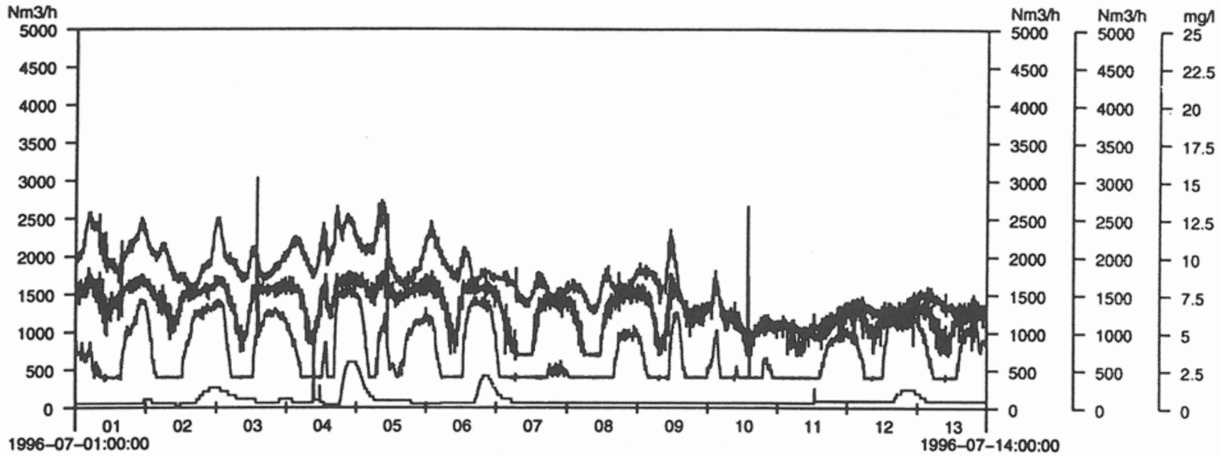
Driftstrategier

Driftstrategierna på Henriksdal har som mål att med god marginal klara kraven på kväve och fosfor samtidigt som kostnader för kemikalier och energi skall minimeras. För att upprätthålla en hög kvävereduktion är det viktigt att bibehålla nitrifikationen i anläggningen. Detta har högsta prioritet vid driften av biosteget. Nitrifikationen kontrolleras kontinuerligt genom mätning av ammoniumhalten ut från biosteget. Om halten är högre än 3 mg/l som dygnsvärde och trenden är stigande kan i första hand syrehalten ökas för att tillfälligt öka kapaciteten. Om halten fortsätter att stiga kan ytterligare luftade zoner tas i drift. Slutligen kan slamåldern höjas genom en minskning av överskottslamuttaget. Strategin är att alltid driva processen med minimal slamålder och minsta antal luftade zoner. En för hög slamålder tycks öka risken för tillväxt av trådformiga bakterier. En luftad zon förbrukar ca 500 Nm^3 luft/h vilket förbrukar ca 500 kWh per dygn. Den mekaniska omrörningen förbrukar ca 160 kWh per dygn. Dessutom sker en denitrifikation i en mekaniskt omrörd zon även om denitrifikationshastigheten är låg på grund av låg tillgång på organiskt material



Figur 5: Nitrat(mg/l) och denitrifikationshastighet(g N/kgVSS.h) i olika zoner

Med hjälp av on–line mätning av ammoniumhalten och luftflöden i olika zoner kan aktiviteten i olika delar av bassängen uppskattas. Ammoniumhalten visar nivån på nitrifikationen. Luftflödena ger en indikation på var i bassängen som nitrifikationen sker. I figuren nedan visas luftflödena och ammoniumhalten i block 6 under två veckor.



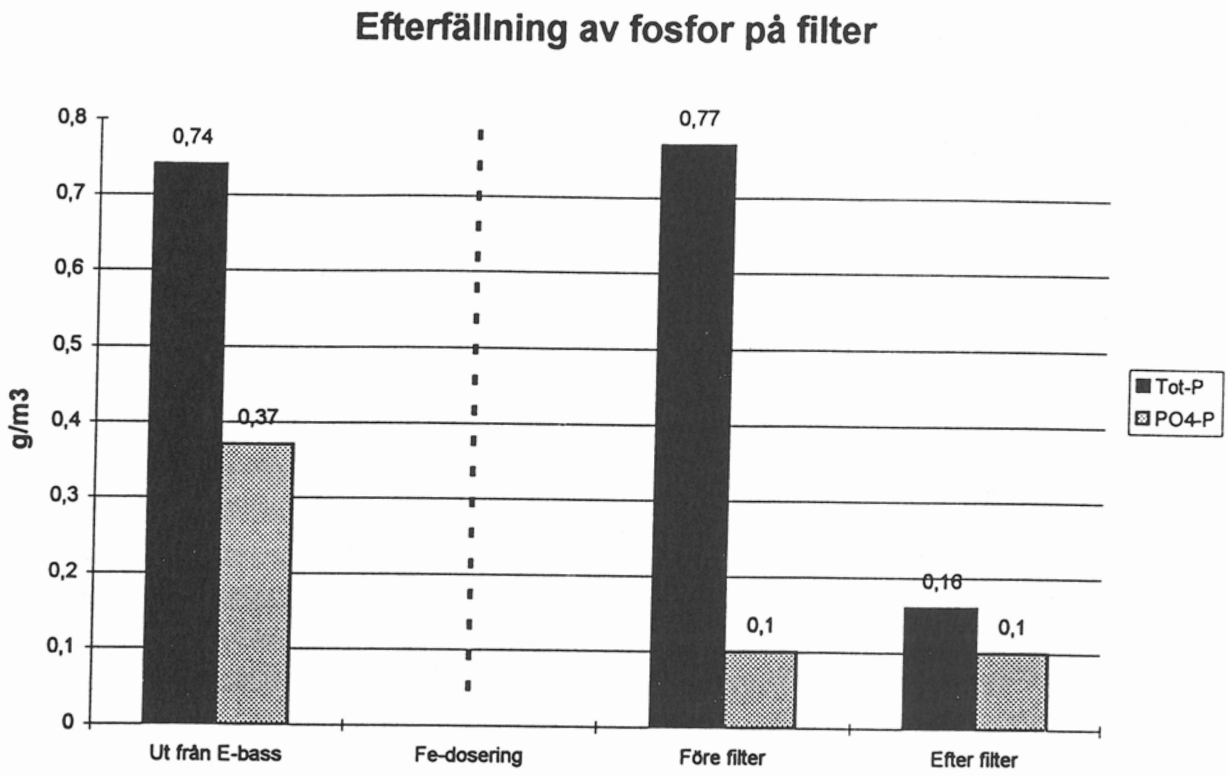
Figur 6: Luftflöden i olika zoner och ammoniumhalt i utgående vatten från block 6.

Energi för luftning

Luftningssystemet består av ett bottentäckande system av membranluftare. Dessa är indelade i fyra zoner i varje block med separat reglering av luftflödet till varje zon. Genom att endast tillföra den luft som förbrukas i processen kan systemet drivas mycket effektivt. Luftförbrukningen ligger idag på mellan 20 000–30 000 Nm^3/h med en energiförbrukning per dygn mellan 20 000 och 30 000 kWh. Det gamla luftningssystemet förbrukade drygt 30 000 kWh/d och hade dessutom inte kapacitet överföra tillräckligt med syre för nitrifikation.

Fällning på filter

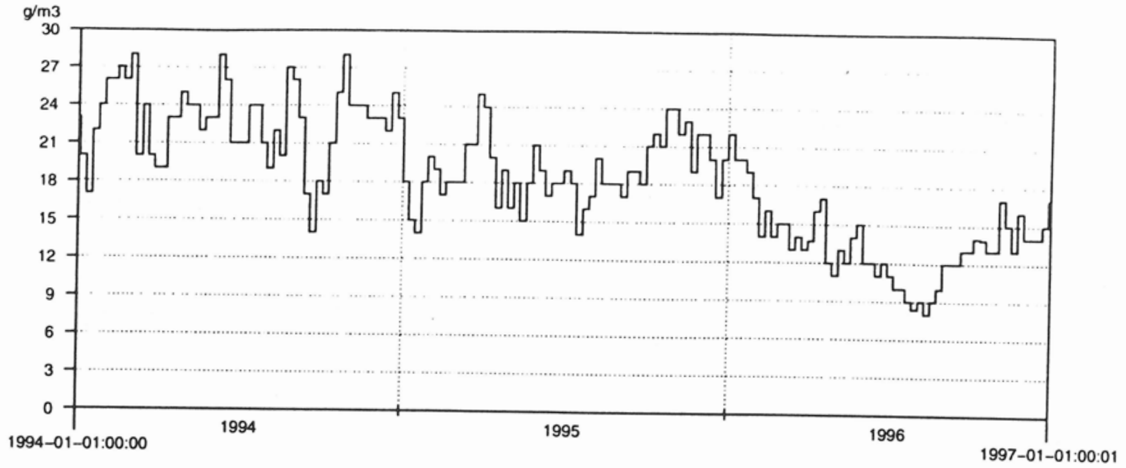
Den lösta fosforhalten ut från den biologiska reningen har ökat till under långa perioder över 0,3 mg/l efter införandet av kvävereduktion. För att fälla ut denna fosfor med enbart förfällning måste dosen höjas kraftigt. Genom att tillsätta 2–4 g Fe/m^3 före filtren kan fosforn fällas ut och avskiljas . Fördoseringen blir då inte lika viktig utan kan sänkas. På Henriksdal har den totala dosen sänkt med ca 4 g Fe/m^{3}. Figur 9 visar fosforhalten ut från biosteget , efter dosering och efter filtrering.



Figur 7: Totalfosfor och löst fosfor ut från biosteget, efter dosering och efter filtrering.

Resultat

Hela utbyggnaden är ännu inte klar. Ytterligare tre block kommer att driftsättas under våren 1997. Någon optimering av processen har därför ej kunnat ske eftersom sedimenteringen varit hårt belastad under perioden. Figur 8 visar totalkvävehalten under perioden 1994–1997. I November 1994 togs två bioblock i drift. Under april 1995 togs filtren i drift. Två biologiska block driftsattes under våren 1996. Samtidigt togs de kvarvarande gamla bassängerna ur drift.



Figur 8 Utgående halt av totalkväve.

Kostnader

Kostnaden för utbyggnaden budgeterades totalt till 740 miljoner kronor i 1991 års priser. Idag är slutprognosen 655 miljoner kronor i löpande priser. Efter justering av index och valuta hamnar projektet på ca 100 miljoner kronor under budget. I hög grad beror detta på det gynnsamma läget för upphandling av byggentreprenader som rått under perioden.

SAMMANFATTNING

Utbyggnaden av Henriksdal för kvävereduktion och förbättrad fosforavskiljning är klar under våren 1997. Resultaten hittills från de drifttagna anläggningsdelarna pekar på att de uppsatta kraven på fosfor och kväve kommer att klaras. Den valda processen ger stora möjligheter till en drift med minskad tillförsel av energi och kemikalier. Elförbrukningen och kemikalieförbrukningen är idag lägre än innan anläggningen byggdes ut.

Holbæk WWTP - Experiences with operation of SBR-reactors for nitrogen removal

Søren Andersen, Holbæk WWTP, DK

Historical outline

Holbæk had its first wastewater treatment plant in 1934, an Emscher tank. In 1937 the plant was extended to a physical wastewater treatment plant with a digester.

In 1976 a completely new physical and biological treatment plant was built. The biological method applied was based on the activated sludge principle.

In 1987, the municipality decided to extend the plant for nitrogen and phosphorus removal. The extension started in 1989. The commissioning of the new physical, biological and chemical plants was initiated at Easter in 1991 and by the end of 1992 the final extension of the wastewater treatment plant was ready. The plant was from now on capable of fulfilling the requirements made in the Danish Action Plan on the Aquatic Environment.

Plant configuration (see drawings attached)

The physical treatment part

The untreated wastewater from the catchment area is introduced into the plant via two ø 1000 pipes. First, the wastewater passes an overflow structure from where amounts > 3,600 m^3/h are led into the Holbæk bay via two "coarse screens". Downstream of the overflow structure, the water flows through a chamber which retains large objects.
From there the water continues through two fine screens with a bar spacing of 3 mm (Aqua Guard) and an aerated grit chamber. Just upstream of the inlet to the screens, a gas and oil alarm is installed. If alarm is given, the screens and the inlet pumps stop. Between the screens and the grit chamber, a sampler is located.

The screenings are pressed and the grit dewatered in a grit dewatering unit before being led to a drained container. Screenings and grit are subsequently transported to a controlled tip.

From the grit chamber, the wastewater is conveyed to the main pumping station. Just before the main pumping station a new overflow structure is located where amounts > 2200 m^3/h are led into the Holbæk bay.
The main pumping station is equipped with 5 submerged centrifugal pumps with a capacity of 2200 m^3/h. The pumps are level controlled. One pump is frequency controlled. In the pumping station, a temperature/pH meter is installed. The values

are collected in the Supervisory Computer and Data Acquisition system (SCADA). From the outlet chamber of the pumping station, the physically treated wastewater is led to distribution chamber 1. From there it is directed towards two of the SBR reactors or towards distribution chamber 3. From the distribution chamber 3, it is either led to another 3 SBR reactors or via distribution chamber 2 to equalising tank 1.

In equalising tank 1, a submerged centrifugal pump is mounted which can pump the wastewater back to the process via distribution chamber 1.

The biological treatment part

The biological treatment is based on the SBR technique (**S**equential **B**atch **R**eactor) with subsequent polishing in a continuously operated sand filter.

The SBR technique is a batch-wise treatment of the wastewater.
An SBR reactor functions both as activated sludge tank and settling tank.

There are 5 SBR reactors, 3 large and 2 small ones. The 3 large reactors each have a volume of 3000 m^3 and the 2 small ones each have a volume of 1450 m^3. The SBR reactors 2-5 treat the incoming wastewater and reactor 1 treats return "filtrate" flow from the sludge treatment and sometimes leachate from the municipality's tip.

In the 4 SBR reactors, the wastewater is subjected to a final treatment with a treatment cycle of 4-6 hours comprising the steps FILLING, REACTION, SETTLING, DECANTATION and perhaps RESTING (see appendix 1).

It is possible to select the following cycle periods: 3.0, 4.0, 4.4, 4.8, 5.2, 5.6 and 6.0 hours in the SCADA system (see appendix 2).

Within the fixed cycle period, the time intervals for filling, aeration (nitrification), mixing (denitrification), settling, decantation and perhaps resting are later chosen. There is only a rest period in those cases where the reactors have not been filled by the end of the filling period, or if the preset decantation period is longer than the time it takes to empty a full reactor. It is possible to choose more aeration and mixing periods within the same cycle period (see appendix 1).

The control system is made in such a way that it is possible to connect a reactor whenever needed (see appendix 3). A reactor can also be taken out of operation whenever necessary, e.g. due to repair works, poisoning or oil leaks.

SBR-1 pretreats the return "filtrate" flow and leachate, if any, before it is pumped back into the process unit. This water is returned to distribution chamber 1.

A reactor is started up when the start conditions for an operational reactor are available, i.e. all valves for wastewater, air and chemical dosing are closed and all pumps stopped.

Each SBR reactor has a max. volume which is divided into a basic volume and a filling volume.

Table 1: Reactor volumes (per reactor)

Reactor	Total volume	Basic volume	Filling volume
1 and 2	1450	900	550
3,4 and 5	3000	1900	1100

Below, a brief description is given of the individual sequences forming part of SBR operation:

Filling
Each cycle starts from a certain basic volume selected in the SCADA system. The filling period normally constitutes 30-50% of the cycle period.
The inlet valves open when filling is started and close at the end of the filling period or before if the max. level in the reactor has been reached.

The filling period can begin either with an anaerobic period or an aeration period dependent on what is desired.
After for instance the anaerobic period, alternating aerobic periods (aeration and perhaps mixing) and anoxic periods (mixing and perhaps aeration) follow.

The aerobic period (oxygen content 1.5-3 mg O_2/l) is controlled through the positions of the air valves (PI control)/blowers (see appendix 4) and the mixing, if any, with floating mixers.

In the anoxic periods (oxygen content 0 mg O_2/l), only mixing takes place.

Reaction time
When the filling period ends, a reaction period begins. During this, additional removal of organic matter, nitrogen and phosphorus takes place. This is also effected through alternating aerobic and anoxic periods.

Settling
After the reaction phase we have a settling period during which the sludge formed settles (formed through decomposition of the organic matter, the nitrogen and biological/chemical phosphorus removal).

During the last 10 minutes of the settling period, the transparency of the water is measured. This is done by means of an on-line turbidimeter, submerged approx. 50 cm into the water.

If the turbidity of the water is not satisfactory (< 25 mg SS/l, optional in the SCADA system), it is not possible to let water out of the reactor. The value is used to make sure that "sludge" is not admitted to equalising tank 2 before pumping into the sand filter (screw pumps).
Furthermore, the depth of visibility and the SS content in the reactors (in the summer the SS content is approx. 3500-4500 mg/l and in the winter approx. 4000-6000 mg/l) are measured manually twice a week by means of a portable SS meter. The SS content is measured in a fully aerated reactor (basic level).

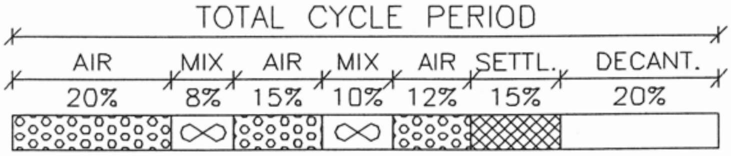
Decantation
The decantation period completes the cycle. During this period, the treated wastewater is led through a decanter and outlet valve into the decantation well where submerged centrifugal pumps pump the water to distribution chamber 2, and from there it is led to the sand filter via the equalising tank 2 and the outlet structure. In the outlet structure, two on-line turbidimeters are placed, one before and one after the sand filter. In case of a too high turbidity, alarm is triggered and the sand filter screw pumps stop.

The decantation starts and stops according to a preset time or when the minimum level in the SBR reactor has been reached.

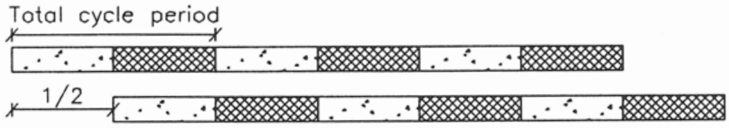
When the basic level has been reached or the decantation period has ended (max. 1 hour), the outlet valve closes. When this end stop has been acknowledged, the decanter closes. It is then possible in the SCADA system to choose whether one wants to start a new cycle or whether there is to be a rest period.

Resting
As mentioned earlier, there is only such a period in those cases where the reactors have not been filled by the end of the filling period, or if the preset decantation period is longer than the time it takes to empty the reactor.

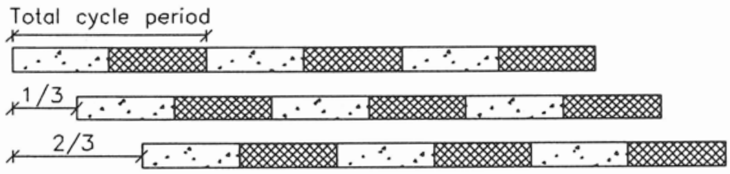
Examples of operation
A cycle period of 4.8 hours can for instance comprise the following:



Reactor operation in case of 2 reactors in operation (serial filling):



Reactor operation when 3 reactors are operating (overlapping inflow):



The chemical part

Phosphorus removal is obtained by simultaneous precipitation. The chemical process can be performed through precipitation at several points, iron sulphate can be added after the main pumping station and in the 5 SBR reactors (see appendices 1 and 5).

Dosing of iron sulphate
Normally, it will be sufficient to add 20-25 mg/$FeSO_4$/l after the main pumping station.

1. *$FeSO_4$ dosing after the main pumping station*
   If one or more wastewater pumps are operating, $FeSO_4$ is added. The amount to be dosed is adjusted manually on the pumps. The pump capacity is 0-150 l/h.

2) The amount of $FeSO_4$ fed to the SBR reactors can be controlled by the connection period of the pump. The connection period varies on a linear basis with the amount filled into the reactor in addition to basic volume.

   The $FeSO_4$ dosing is set to take place in the beginning of the first aeration period of the SBR cycle, but can be changed to any aeration period. In order for the motor valve in the reactor in question to open, the oxygen level in this tank must be > 1.5 mg O_2/l. This set point can be freely chosen in the SCADA system. The motor valves open immediately before the dosing pump starts. Only one motor valve **can be** open at a time, and it closes as soon as the dosing period is over.

   The dosing period to reactors 1 and 2 is adjusted on the basis of the volume with the factor $(V_{max}-V_{tank})_{1-2}/(V_{max}-V_{tank})_{3,4,5}$.

Dosing of ethanol

This has never been in operation (See however appendices 1 and 5)

This dosing is performed in the exact same way as the dosing of $FeSO_4$ to the reactors. It takes, however, place in the beginning of the **last** anoxic period. The operation of the two dosing pumps alternate.

Excess sludge/sludge treatment

Excess sludge (purely biological) is removed in the settling period and in some cases in the decanting period (see appendix 1).
The pumping period for excess sludge is preset and the pumping is carried out by submerged centrifugal pumps, placed in the reactors (See appendix 6.)
The on and off periods are adjusted in the SCADA system on the basis of the depth of visibility and the SS values measured in the reactors.

The excess sludge is pumped to a 350 m³ thickener and after being thickened to 2-4% it is pumped to the 900 m³ digester by means of an eccentric screw pump. The digester is operated at thermophilic conditions (approx. 58°C). The excess sludge gravitates from the digester to a sludge store before it is pumped to a chamber filter press by piston pumps.

The return 'filtrate' flow from the thickener and the chamber filter press flows to the SBR reactor 1 as mentioned previously. It also receives leachate from the municipal landfill.

At first the digester was mesophilic. Due to the short retention time in the digester the mode of operation was changed to thermophilic conditions.

In order to ensure an energy-neutral operation (avoid to buy gas/electricity for heating) of the thermophilic digester one more sludge/sludge heat exchanger has been installed.

An average of approx. 80 m³ sludge/d with a TS content of approx. 3% is pumped in, which gives a gas production of approx. 580 m³/d. At mesophilic conditions approx. 300 m³/d were produced when the same amount was pumped in.

In order to make a mass balance for the digester, flow meters and TS meters have been installed.

The gas produced is used for power production. The excess heat is used for heating the digester and the shop buildings.

The power produced in the gas motor is used elsewhere in the plant, and during periods when the load is low, **it can** be sold to the local power supply plant.

Operating experience

As previously stated, the running in of the plant was commenced at Easter in 1991 and completed at the end of 1992.
It took 1½ year because of the programming of the SCADA system, as the original programming of the reactor operation was made more operator-friendly.

Preliminary treatment plant

The plant has worked without problems. It should, however, be mentioned that the container for screenings and grit has been replaced by a more well-working draining system.

Biological treatment part

The plant operates as a simultaneous denitrification plant.

Since the start-up, the plant has mainly been operated with a cycle period of 4.8 hours (the design time), which has shown satisfactory results. In 1996 a cycle period of 4 hours was tested to increase the wastewater flow through the plant. This has been possible without impairing the operating results. During the summer only the 3 large SBR reactors need to be in operation. During the winter 4 and preferably 5 reactors must be in operation.

The blowers (3) supplying the reactors are HV-Turbo blowers (max 4,400 m^3/h/unit, min. 1,980 m^3/h/unit). The blowers supply all the reactors. This is not a good choice in connection with SBR operation, as it is the blowers which control the process and not the process which controls the blowers. It would be better if there was a blower for each reactor.

In the reactors fine-bubble membrane diffusors (Sanitaire) are installed, which has worked well. They have run for 4 years without being replaced (they were replaced in 1992 (one year old) due to an error in the design of the lower casing).

To reduce the power consumption the SBR reactors have been operated without mixing during the anoxic periods (however, a very brief period of aeration has been used to keep it mixed).

After innumerable tests with biological phosphorus removal we have to note that at Holbæk WWTP it has not been possible to obtain any large effect (practically none). But it has turned out to be better to dose iron sulphate immediately after the main pumping station than in the reactors as first intended.

If the biological and chemical treatment in the SBR plant is considered separately, approx. 1.0-1.2 kWh has been used to remove 1 kg of BOD, whereas if considering the entire plant operation, approx. 1.6-2.0 kWh has been used to remove 1 kg of BOD.

The sand filter has been the plant's security against sludge flight, and in this way contamination of the recipient has been avoided. Sludge flight to the sand filter has only occurred in connection with errors in the control program and incorrect operation of valves in manual operation.
The on-line meters in the outlet (Amm-N and NO_3-N) have been a great help as they make it possible to determine from which reactor the poorly treated wastewater comes. Consequently quick action to correct operating parameters and set points has been possible. Also on-line temperature and oxygen meters have been installed.

As regards **sludge treatment** at Holbæk with thermophilic digester operation a considerably more environment-friendly sludge with salmonella below 10/ml and faecal streptococcus below 100/g has been obtained. The subsequent dewatering is consequently also more environment-friendly and has been more reliable than previously (no problems in the winter). The sludge is well-hygienised, however not controlled as prescribed by the Danish EPA. As a result of the increased gas production (approx. a doubling), the amount of sludge to be removed has decreased.

More generally the following can be said:

- Control, regulation and monitoring of the plant has been an advantage
- Each reactor needs its own blowers, operating parameters and set points
- Good with someone on call after working hours
- Calibration and cleaning of measuring instruments as required, however at least once a week.
- Own analyses have to be carried out (twice a week) with a view to operational optimisation.
- In case of poisoning only one reactor will usually be affected, and if the reactors have their own operating parameters and set points it is possible to treat the poisoning separately (e.g. 24-hour operation).
- Equalisation before and especially after the SBR reactors is an advantage, as usually a large amount of wastewater has to be discharged in a short time.
- It has been an advantage to be able to treat the 'filtrate' flow from the sludge treatment in a separate reactor, especially in Holbæk, where there has been an increase of the nitrogen amount in the filtrate (from approx. 300 to 500-700 mg/l) due to the thermophilic operation of the digester.

Design load

PE	44,000	
$Q_{diurnal,\ average}$	9,600	m^3/d
$Q_{time,\ des}$	550	m^3/h
BOD_5	2,640	kg/d
SS	2,820	kg/d
Total N	525	kg/d
Total P	180	kg/d

Effluent standards

BOD_5	< 15 mg/l
SS	< 20 mg/l
Total N*	< 8 mg/l (winter)
	< 5 mg/l (summer, 1 May to 1 November)
Total P	< 1 mg/l
pH	6.5-8.5

* from 1 Jan. 1997 < 6 mg/l (winter) and 4 mg/l (summer) are expected

Operating results

Actual loading

PE	30,000	
$Q_{diurnal,\ average}$	7,600	m^3/d
BOD_5	1,800	kg/d
Total N	340	kg/d
Total P	75	kg/d

Effluent concentrations

Table 2: Results from analyses made at a food laboratory

Parameter		1993	1994	1995	1996 *
Flow	m^3/d	5599 (22)	9163 (11)	7688 (12)	5550 (13)
pH		7.9 (22)	7.7 (11)	7.7 (12)	7.7 (13)
SS	mg/l	< 5.0 (22)	< 5.0 (11)	< 5.0 (12)	< 5.0 (13)
BOD_5	mg/l	< 3.0 (22)	< 3.0 (10)	< 3.0 (11)	< 3.0 (13)
COD	mg/l	47.6 (20)	35.8 (11)	36.3 (12)	35.8 (13)
AMM-N	mg/l	2.5 (13)	1.0 (11)	0.5 (11)	1.1 (01)
NO_3-NO_2-N	mg/l	3.3 (12)	2.3 (11)	2.7 (12)	2.1 (01)
Total N	mg/l	6.0 (22)	4.1 (15)	4.4 (21)	4.3 (13)
Total P	mg/l	0.7 (22)	0.7 (15)	0.6 (19)	0.7 (13)

* The period 1 Jan.-1 Oct. The figure in brackets indicates the number of analyses.

Table 3: Results from analyses made at the plant's own laboratory.

Parameter		1993	1994	1995	1996 *
Flow	m^3/d	6623 (255)	9203 (321)	8878 (363)	6101 (274)
COD	mg/l	27.6 (215)	27.5 (122)	16.4 (035)	16.0 (004)
AMM-N	mg/l	2.2 (253)	1.3 (231)	0.8 (199)	2.3 (102)
NO_3-NO_2-N	mg/l	3.0 (249)	2.8 (231)	3.6 (200)	2.9 (098)
Total N	mg/l	5.5 (231)	5.6 (149)	5.2 (107)	3.9 (059)
Ortho-P	mg/l	0.74 (251)	0.88 (230)	0.97 (197)	0.98 (108)
Total P	mg/l	0.76 (231)	0.90 (146)	1.00 (146)	1.01 (030)

* The period 1 Jan.-31 Oct. The figure in brackets indicates the number of analyses.

A comparison of the two tables shows a certain degree of disagreement, but as it will appear from the analysis results, an improvement of the operation on the basis of the analyses made at the plant would mean that the plant will be able to comply with the effluent standards without problems. However, the COD values should not be used as a guiding parameter for the treatment efficiency as it is always much lower than the value of the authorised laboratory.

It appears from the effluent nitrogen concentrations that it is not a problem to comply with the future effluent standards.

Figure 1: The power consumption in 1993, 1994, 1995 and 1996 (½ year).

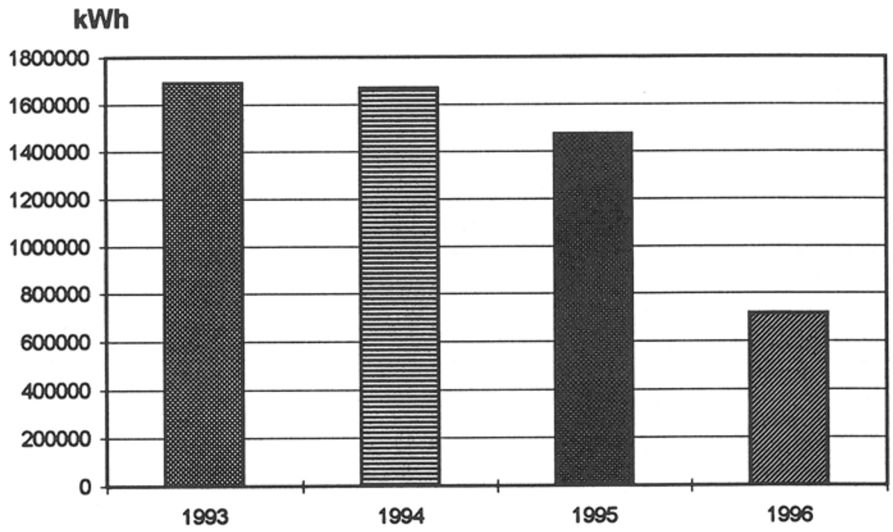


Figure 2: Removed dewatered sludge in 1993, 1994, 1995 and 1996 (½ year).

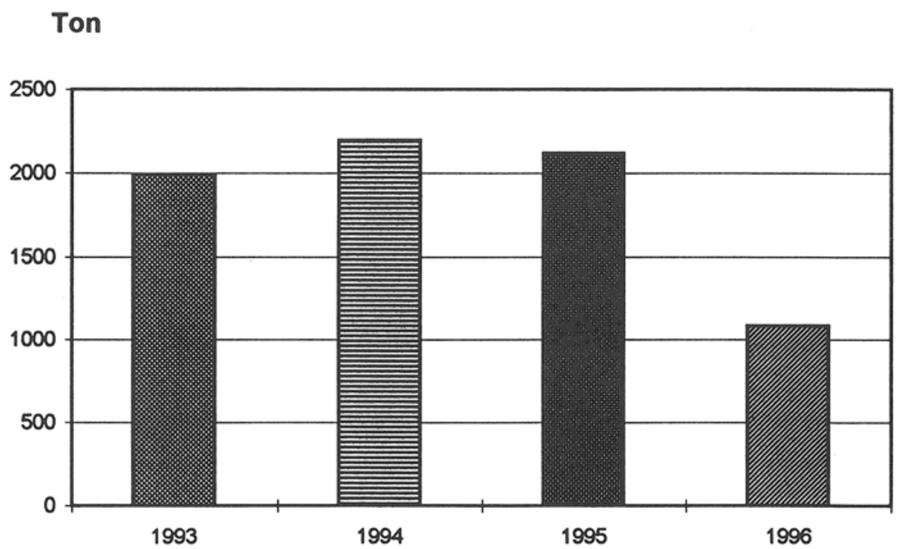
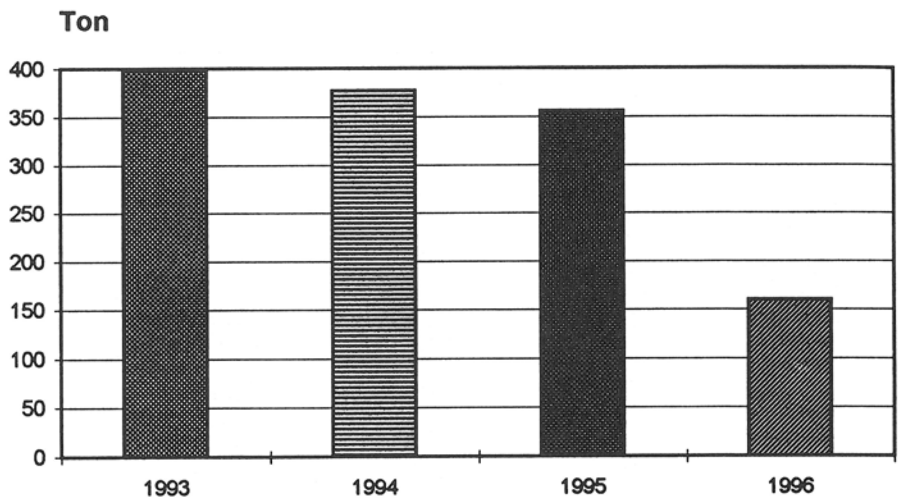
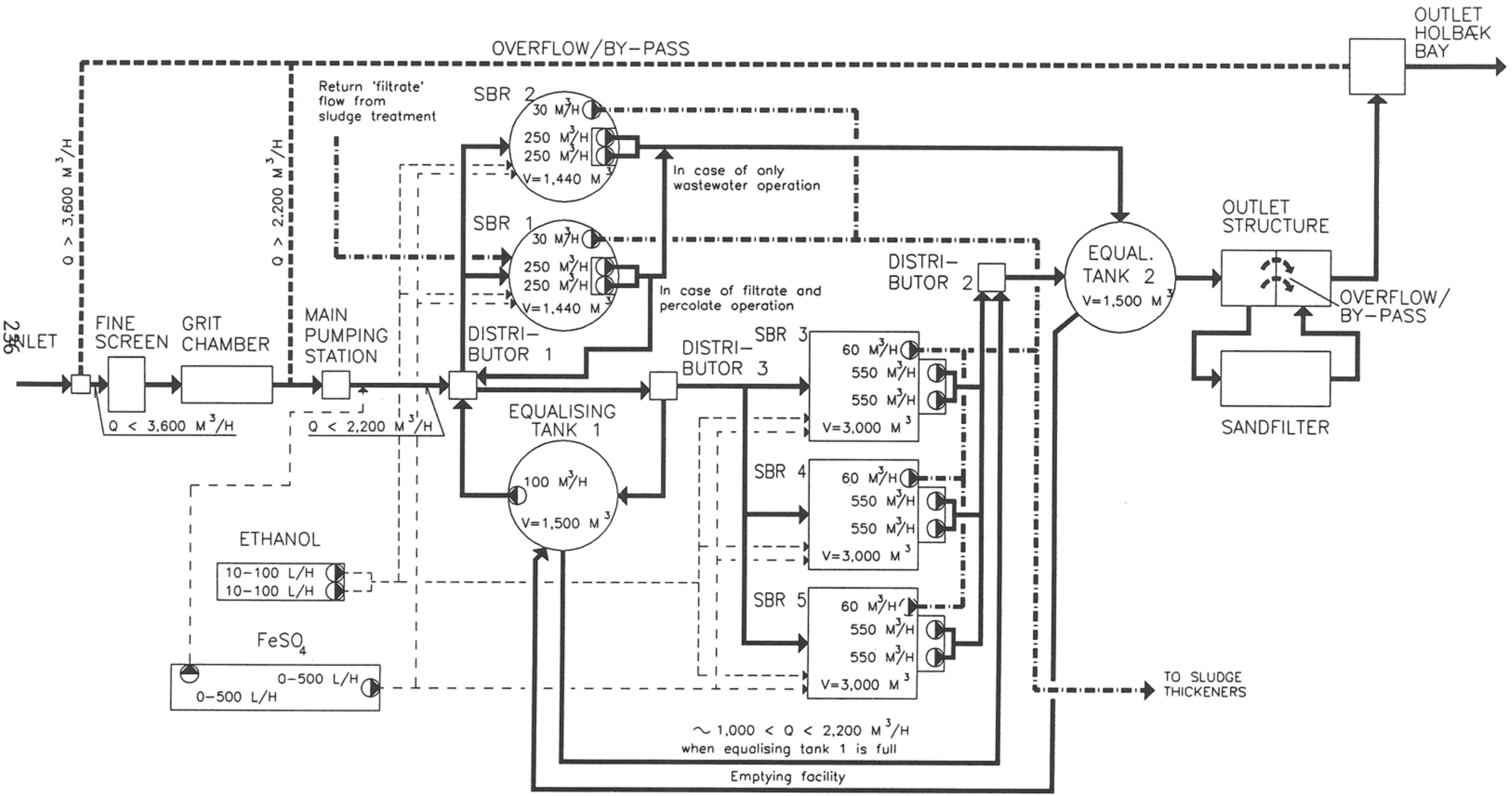


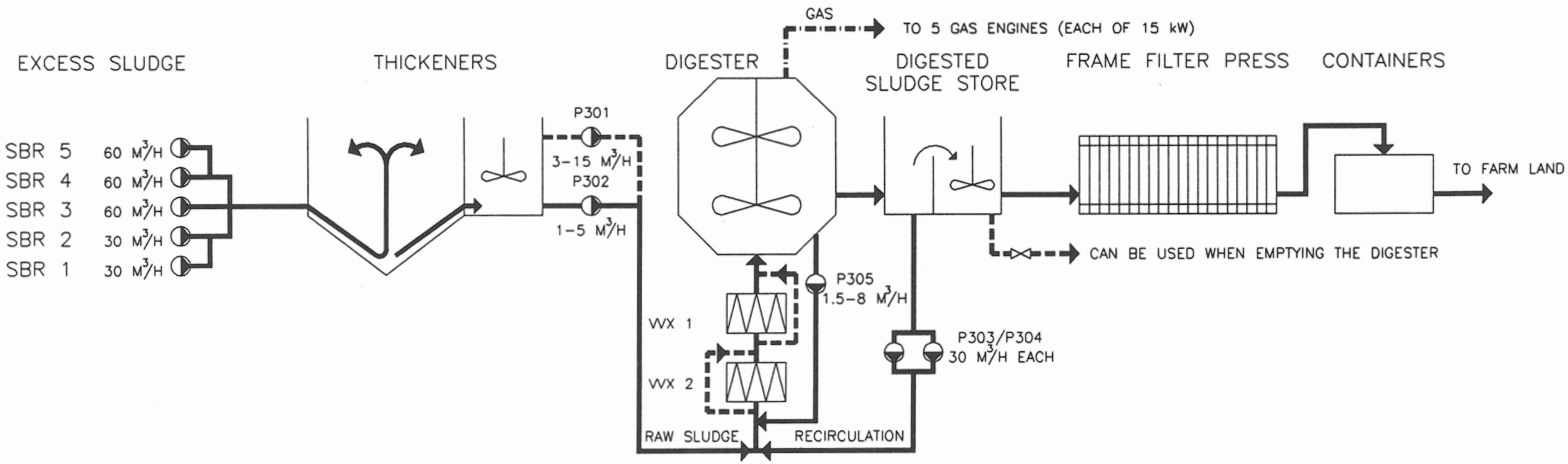
Figure 3: Removed screenings and sand in 1993, 1994, 1995 and 1996 (½ year).





FLOW DIAGRAM

?K: FLOWD4GB



SLUDGE DIAGRAM

HOLB?K: FLOWD5GB

*** DRIFTSPARAMETRE FOR SBR-TANKE, FÆLLES *** 96-11-13 13.35

CYKLUSTID = 4,0 t (= 240 min)

CYKLUSTID / REAKTORVALG

BELUFTNING

Pause fra tid 0: 0 min

1. Ox-periode..: 51 min
1. Anox-periode: 59 min
2. Ox-periode..: 11 min
2. Anox-periode: 119 min
3. Ox-periode..: 0 min
3. Anox-periode: 0 min
4. Ox-periode..: 0 min
4. Anox-periode: 0 min
5. Ox-periode..: 0 min

OMRØRING (fra tid 0)

1. Driftsperiode: 0 min

1. Pauseperiode : 65 min
2. Driftsperiode: 40 min
2. Pauseperiode : 135 min
3. Driftsperiode: 0 min
3. Pauseperiode : 0 min
4. Driftsperiode: 0 min

METHANOL-DOS.(P271/P272)

Pause fra tid 0: 0 min

DEKANTERING

Pause fra tid 0: 190 min
Driftsperiode..: 50 min

Turbiditetsgrænse
for dekantering: SS/1

OVERSKUDSSLAMPUMPNING

Pause fra tid 0: 185 min

FeSO4-DOS. (P262/P263)

Pause fra tid 0: 1 min

(PARAMF40)

Fyldningsperiode: 85 min.

ÆNDRING AF DRIFTSPARAMETRE, SBR 1-5 96-11-04 15.16

Følgende perioder og set-punkter er fælles parametre for SBR-reaktorerne.
Disse er bundet til cyklustiderne.

Følgende kan ændres: Beluftnings-, omrørings- og dekanteringsperioder
Pausetid for overskudsslampumpning
Set-punkt for jernsulfat / methanol dosering

Fæl. param., cyklus = 3,0 t (192 min)

Fæl. param., cyklus = 4,0 t (240 min) Fæl. param., cyklus = 5,2 t (312 min)

Fæl. param., cyklus = 4,4 t (264 min) Fæl. param., cyklus = 5,6 t (336 min)

Fæl. param., cyklus = 4,8 t (288 min) Fæl. param., cyklus = 6,0 t (360 min)

Følgende perioder og set-punkter er individuelle parametre for hver enkelt

SBR-reaktor. Disse er ikke bundet til cyklustiderne.
Følgende kan ændres...: Niveauer (basis-, fyldnings- og højniveau)
SS-, ilt- og redox-grænser
Drifttid overskudsslampumpning

Individuelle parametre Start/stop niveau dekanteringspumper

(PARAÆNDR)

OVERSIGT SBR   VALG AF SBR-REAKTORER OG CYKLUSTID   96-11-13 15.12

Her vælges hvilke reaktorer der ønskes i drift. Ligeledes kan deres aktuelle forløb i cyklus'en vælges. Denne indsættes i segmenter, hver cyklustid er inddelt i 240 seg. Når vi kører med 2, 3 eller 4(5) reaktorer skal der være segment spring på henholdsvis 120, 80 og 60 mellem reaktor start/forløb.
Ved følgende cyklustid (min): 240, 264, 288, 312, 336 og 360; SKAL DET AKT. MINUTTAL DELES MED FAKTOR...: 1.0, 1.1, 1.2, 1.3, 1.4 og 1.5 => ny segment.
AKT CYKLUSTID: 240   Manuel drift: ↑↓240

SBR/VALG	AKTUEL/NY SEGMENT		
	minut =	seg =	seg
1 ↑↓1	236	236	236
2 ↑↓1	236	236	236
3 ↑↓1	56	56	56
4 ↑↓1	116	116	116
5 ↑↓1	176	176	176

** PUMPESTATION **

CYKLUSTID (t)	NIV. (cm)	AUTO/MAN
4,0 (240 min)	240	MAN
4,4 (264 min)	220	AKT. NIV
4,8 (288 min)	180	155
5,2 (312 min)	140	
5,6 (336 min)	100	
6,0 (360 min)	60	

Betydning: 0 = Ude af drift,
1 = Husspildevand,
2 = Perkolatforsøg (SBR 1)

(REAKVALG)

INDIVIDUELLE PARAMETRE FOR ILTREGULERING 96-11-04 15.15

PARAMETER	SBR 1	SBR 2	SBR 3	SBR 4	SBR 5
SETPUNKT ILT mg/l	3.5	3.5	3.5	3.5	3.5
SAMPLINGSTID........sek.	10	10	10	10	10
MAX. STYRESIGNAL....msek	2500	2500	2500	2500	2500
STARTÅBNING VENTIL %....	35	35	35	35	35
MAX. ÅBNING VENTIL %....	55	60	70	70	70
KD......................	1	1	1	1	1
KP......................	1	1	1	1	1
B	10	10	10	10	10
MIN. ÅBNING VENTIL %....	15	15	28	28	28

(ILT_REG)

DRIFTSTIMER DRIFTSPARAMETRE FOR KEMIKALIEPUMPER TIL BIOBLOKKEN 96-11-04 15.29

Placering : **FeSO4 og Methanol-pumperne er placeret i kemi.bygn.**

Driftsform: **Automatisk - tidsstyret i henhold til fyldningsniveau i tankene.**

Manuel - kontinuerlig drift

	SANDFANG	SBR-REAKTORERNE		SBR-REAKTORERNE	
Status	FeSO4 P261	FeSO4 P262	FeSO4 P263	Meth. P271	Meth. P272
Drift..................	STANDSET	STANDSET	KØRER	STANDSET	STANDSET
Pumpehastighed.........	0 %	0 %	0 %	0 %	0 %
Hastighed ved 0-100 m³.	40 %				
Hastighed ved ≥550 m³.	60 %			**Niv. Methanol** 21 %	
Omskifter i tavle......	0 el. MAN	0 el. MAN	0 el. MAN	AUTOMATIK	AUTOMATIK
Anden driftsfejl.......	INGEN FEJL	INGEN FEJL	INGEN FEJL	INGEN FEJL	INGEN FEJL
Trykvagt..............	INGEN FEJL	INGEN FEJL	INGEN FEJL	INGEN FEJL	INGEN FEJL
Tørløbssikring.........	INGEN FEJL	INGEN FEJL	INGEN FEJL	INGEN FEJL	INGEN FEJL
Styreskab K1...........	INGEN FEJL	INGEN FEJL	INGEN FEJL		

Doseringstid	SBR 1		SBR 2		SBR 3		SBR 4		SBR 5	
	max	**min**	**max**	**min**	**max**	**min**	**max**	**min**	**max**	**min**
Jernsulfat.(FeSO4).min.	0	0	0	0	0	0	0	0	0	0
Methanol...........min.	0	0	0	0	0	0	0	0	0	0
Aktuelt niveau .....cm.	360		355		414		441		353	

Vedligeholdelse. Aktuelt flow i sandfanget: 271 m³ **(KEMIPUMP)**

*** DRIFTSPARAMETRE FOR SBR-TANKE, INDIVIDUELLE *96-11-13 13.35

INDIVIDUELLE PARAMETRE	SBR1	SBR2	SBR3	SBR4	SBR5
Basisniveau .... cm	290	290	355	355	355
Fyldningsniveau ....cm	350	355	485	485	485
Højniveau-alarm ....cm	390	400	515	515	515
Lavt alarmniveau ilt ....mg/l	0.0	0.0	0.0	0.0	0.0
Højt alarmniveau turbiditet ....mgSS/l	25.0	25.0	25.0	25.0	25.0
Højt alarmniveau redox ....mV	1000	1000	1000	1000	1000
Forsinkelse for genåbning af dekanter..min	4	4	4	4	4
OVERfladebelufter / BUNDbeluftere	BUND				
Overskudsslampumpning driftsperiode ...min	10	10	15	25	20
Dekanteringspumper, niveau:start 2 .....cm	390	390	420	380	435
stop 2 .....cm	320	310	350	320	370
start 1 .....cm	280	280	300	300	300
stop 1 .....cm	260	260	260	260	260
højt niveau .cm	435	435	505	505	505

CYKLUSTID / REAKTORVALG   OVERSIGT SBR   (PARAMIND)

A wastewater and sludge treatment process integrating biofilms, wetlands and aerobic sludge digestion for nutrient recovery.

Arne Ingvar Helland[#], Roald Kommedal[£] and Rune Bakke[$]

[#]*Municipality of Klepp*
[£]*School of Science and Technology, Stavanger College*
[$]*Institute of Environmental Technology, Telemark College*

Abstract

The Øksnevad wastewater and sludge treatment plant, designed to solve local pollution problems in a rural community, has been in operation for two years. The wastewater is treated in a biofilm reactor with sedimentation, followed by a pond and wetland concept. Sludge produced in the wastewater treatment and in septic tanks throughout the municipality is treated in a three stage aerobic digestion process designed for maximum nutrient recovery. The sludge and wastewater treatment processes are closely integrated within the same building to maximise efficiency and minimise operation costs and avoid air pollution/loss of nutrients to the atmosphere. About 90% total nitrogen and phosphorous removal from the wastewater is consistently achieved. The treated sludge is stable, nitrified, and odour free. The pathogen kill is efficient and the nutrient recovery is complete; total nitrogen in the final product is the same as in the raw sludge but it is converted from organic and ammonia nitrogen to nitrate. The processes are robust and cost efficient.

Table of content:

1. Introduction

1.1 Background

Øksnevad biological treatment plant was established in 1994 as an integrated part of the Øksnevad Lake restoration program, initiated by the Øksnevad agriculture high school. The restoration program comprised a multi-disciplinary approach including sediment and vegetation removal and sewage divagation in the lake catchment area. A steering committee with participants from local industry, local environment authorities, Øksnevad agriculture high school (Rogaland county), Særheim agriculture research centre (planteforsk) and Stavanger College was established.

The principal idea of the program was to look into the possibilities for creating a system of optimal nutrient recovery inside the catchment area. The lake's free water surface was re-established through sediment removal and de-watering, including subsequent reuse on adjacent cropland. An advanced wastewater treatment plant was proposed for tertiary treatment of divagated sewage, which had to meet stringent quality standards to release the final effluent into a local stream. Concepts, developed at Stavanger College (by Prof. R. Bakke), of an integrated "zero discharge" wastewater treatment and total nutrient recovery sludge treatment, was chosen as a cost efficient solution to this challenge.

Tegle Redskapsfabrikk was in charge of construction and provided the necessary technology for a robust combined wastewater and sludge treatment plant, named the «Øksnevad concept». The wastewater treatment was designed for the total population of the Øksnevad area (including residentials, mechanical industry, dairy production, and public facilities). The sludge treatment was designed to also treat septic sludge from another 2-3000 residents. It was decided that the plant should also serve as a pilot and demonstration plant for sustainable solutions to sanitary problems. It was built with financial support from the national and regional governments; SFT, SND and Rogaland County.

1.2 Location

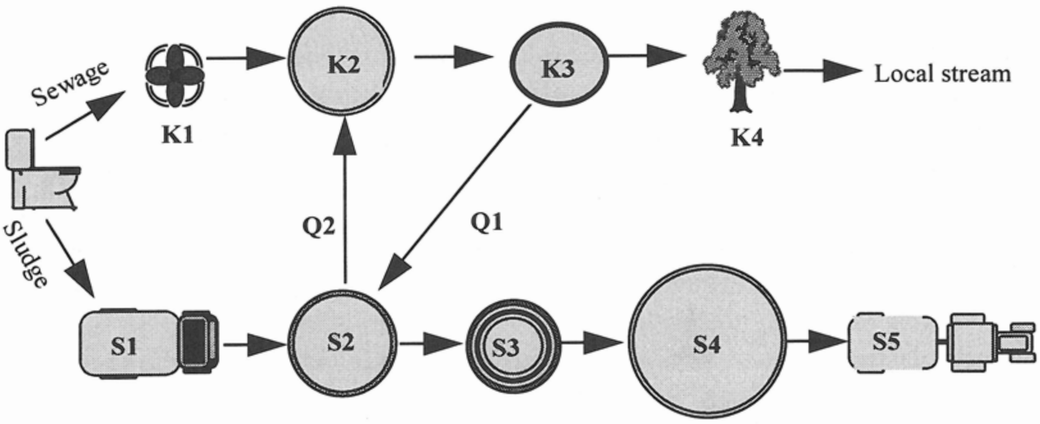
Location of the plant was chosen based on the following factors:

- minimise wastewater and sludge transportation costs
- local effluent disposal (to maintain stream flow and decentralise discharge)
- sufficient area for establishment of constructed wetlands
- easy access for research, and teaching at the agriculture high school
- an area where the quality of the transport system could be monitored
- possibilities of bypassing to another treatment facility in case of plant failure.

1.3 Integration

The project concept presupposed an integrated effort at several levels. Biological qualities, such as regeneration and construction of habitat, reversal of accelerated lake succession (a result of eutrophication) and securing biological diversity, were to be enhanced. Improved water quality, in and downstream of the Øksnevad area was also a main objective. This integration of both biological and water quality objectives form the basic philosophy of our solution. Combining intensive treatment processes with construction of wetlands provides an excellent foundation for ecological synergy, resulting in new and improved wetland habitats and effluent water quality similar to the quality in the recipient.

Integration of sludge and wastewater treatment is another feature evolving from our objective of maximising nutrient removal and recovery. Specifically, these goals have resulted in the production of reject water from the sludge treatment compartment which gives a net potential of phosphorous and nitrogen removal in the bioreactor. This is achieved by returning an organically strong, but nutrient poor reject water providing sufficient carbon source for both volatile fatty acid production and electron donor for denitrification. In addition, heat energy is dissipated back through the processes, rising temperature, especially the indoor air temperature, increasing nutrient recovery through plant assimilation processes.



K1 - Grinder pump
K2 - Biofilm reactor
K3 - Clarifier
K4 - Wetlands

S1 - Septic tanks
S2 - Import storage
S3 - Thermophilic/aerobic sludge reactor
S4 - Export sludge storage/ nitrification
S5 - Agriculture use

Q1 - Sludge transport
Q2 - Reject water

Figure 1 *Sewage and sludge flow chart indicating integration of the process lines. The sewage is collected from the local community and pre-treated by comminution, while sludge is collected from septic tanks throughout rural Klepp municipality.*

The sludge treatment is designed to minimise nutrient loss, especially nitrogen as ammonia. The principal idea is to keep all parts aerobic and adjust sludge retention time in order to secure complete nitrification. Optimising suspended solids concentration, considering transportation costs, nutrient loss and limiting wastewater treatment costs is also necessary in order to design an ecologically sound concept.

Although most of the individual components are not original or unique, the concept of creating an integrated habitat, treatment and resource recovery process is quite rare. The particular process solution chosen here, with its rather intricate process integration is, as far as we know, unique.

1.4 Goal

The main goal of this project was to develop an ideal, sustainable wastewater treatment solution for a rural community. Minimising costs, energy consumption and chemical additions, while maximising treatment quality and resource recovery were key objectives to meet this goal.

2. Processes

2.1 Wastewater treatment process

Biofilm reactorPretreatment is limited to comminution. The wastewater is then treated in a circular rotating aerobic biofilm reactor with sedimentation followed by non-intensive, natural treatment, including a pond with granular medium filtration and a wetland.

Wastewater, at an average hydraulic loading rate of 22 m^3/d (1996), is fed the biofilm reactor (53 m^3) through grinder pumps (comminution), with no other pre-treatment (no screening nor grit removal). The biofilm is immobilised on specially designed plastic elements, bringing the substratum area in the reactor to about 1400 m^2. Oxygenation is provided by a submersible pump equipped with an ejector, securing air supply and stirring power. Oxygenation is controlled by a temperature compensated algorithm, based on hydraulic loading.

The sedimentation basin is designed as an integral part of the cylindrical biofilm reactor. Wastewater enters this secondary clarifier (approx. 43 m^3) in the periphery at the bottom. Sludge is transported to a sludge well by a radial sludge scrape. Following hindered sedimentation in the sludge well a grinder pump transports compressed sludge to the sludge treatment process. Clarified wastewater leaves the sedimentation tank through a weir in the centre of the clarifier.

By careful process control, simultaneous nitrification and denitrification have been achieved in the biofilm reactor. In addition, non steady state operation led to enhanced phosphorous removal. This effect was achieved through implementation of computer modelled process control. This strategy of selectively controlling population dynamics in the biofilm will be further developed in a new research project.

2.2 Wetlands

Leaving the clarifier, wastewater pass a carbonate column providing the opportunity for alkalinity adjustment, before entering a stabilisation pond. The pond is in-house, adjacent to the biofilm reactor, and drain through a gravel matrix, surrounding the reactor, to a pump well. Percolate is returned to the pond, composing a re-circulated gravel filter. Further polishing is achieved by introducing floating aquatic plants in the pond, increasing biofilm area and removing nutrients by assimilation.

The wastewater is then led to a two stage wetland arranged as a buffer/polishing zone between the intensive treatment facilities and the recipient. Wastewater is fed the wetland through a distribution ditch, to enter a sub surface flow, reedbed filter. Depending on groundwater levels, parts of the wetland function more as a free water surface (FWS) constructed wetland. The combination of physio-chemical processes in the soil/sand matrix, and biochemical reactions at the rootzones of surface plants provide tertiary polishing for nutrient removal.

Effluent from both post-treatment units is then discharged into the Øksnevad brook / channel, a side-stream of the Figgjo-river.

The non-intensive constructed wetland system, at a total of 5 da, is not equipped with liners or membranes, causing hydraulic equalisation and mass exchange with groundwater. Tracer analysis, using chloride as a conservative parameter, have revealed an average of around 50% dilution by groundwater.

2.3 Sludge treatment

The sludge treatment processes are integrated with the water treatment process line, giving mutual advantages regarding energy and mass transfer.

Produced sludge from the biofilm reactor/sedimentation and sludge collected from septic tanks, enters the sludge treatment process at the first (a pre-treatment and inlet storage tank) of a three stage reactor configuration. The reactors are constructed as three concentric cylinders, all operated as sequential batch reactors, with different sequences. The centric cylinder serve as the main stabilisation reactor. It is surrounded by the pre-treatment and inlet storage tank. The outer cylinder serve as a post-treatment and storage reactor. This physical configuration minimise energy loss and pumping.

Stabilisation and massive reduction of pathogens are accomplished in the 60 m^3 central cylinder by a thermophilic aerobic culture, at an average solid detention time of 15 days. Respiratory energy release is retained by insulation, giving stabile operation in the thermophilic region (45 - 70°C). Typical reactor temperature varies from 50 to 60°C, at sludge loading of 3 - 4 m^3/d.

3. Results and discussion

3.1 Discharge quality

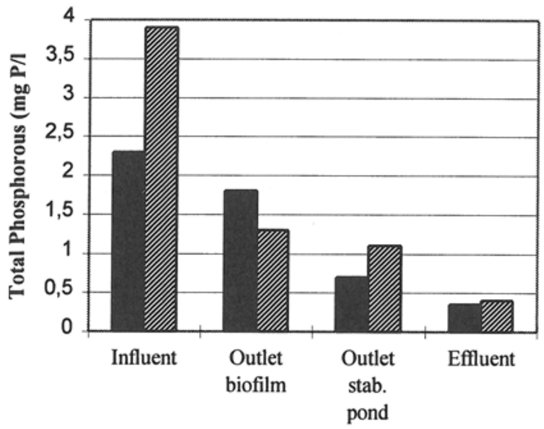
Fylkesmannen in Rogaland, the regional environmental authority, has set the following effluent limits in the discharge permit (Permit given 10.11.93):

Table 1 *Discharge limits Øksnevad Biological Treatment Plant. K1 and K2 values a respectively daily average and peak concentrations*

Parameter	Concentration after biofilm reactor		Concentration after wetland	
	K1 mg/l	K2 mg/l	K1 mg/l	K2 mg/l
Total Phosphorous	2,0	3,0	0,4	1,1
Diss. Org. Carbon	100	110	20	30
Total Nitrogen	min. 75% reduction			
Susp. Solids	to be monitored		to be monitored	

Effluent control and plant performance have been observed using computer logging of physio-chemical parameters (pH, temperature and flow measurements) and water analysis showing instantaneous and weekly averaged water quality values (Tot. P., Tot. N, PO_4^{3-}, NO_3^-, NH_4^+, DOC, COD, Cl^-, SS). Results of nutrient removal from Jul. 1994 (start up) till Jan. 1995 are shown as black bars in figure 2. Later results are presented as patterned bars.

Due to work on the electrical equipment and installation of remote operation systems, sufficient operational stability has not been achieved during the first half of 1996, resulting in poor data for the period. Frequent sampling and stable processes are now re-established, yielding results similar to those achieved in 95.



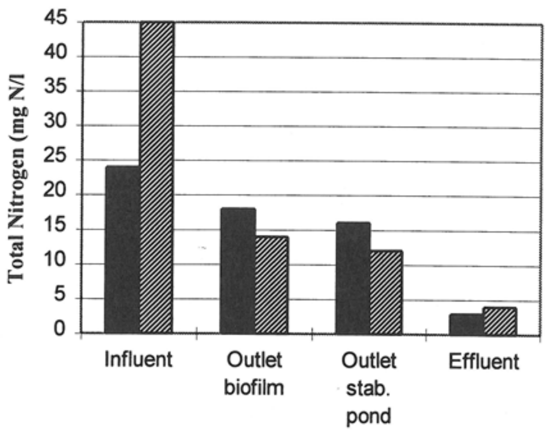
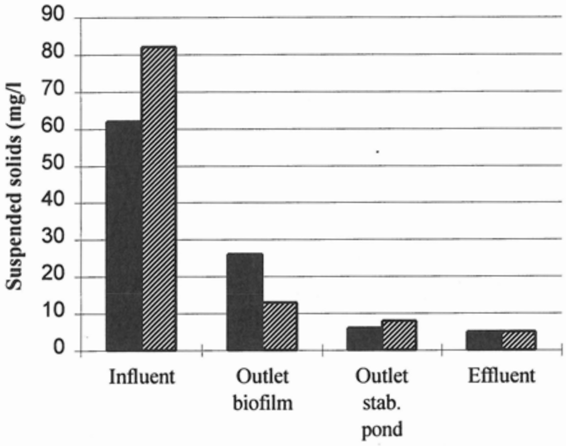


Figure 2 *Influent plus effluent concentrations following [a] the intensive biofilm reactor, [b] the semi-intensive stabilisation pond and [d] the non-intensive wetlands. Performance during the initial test period in 94 (black bars) and after (patterned bars) initiation of model based process control show a distinct increase in both phosphorous and nitrogen removal efficiencies. Effluent concentrations reported her have been compensated for groundwater dilution.(Based on a total of 301 samples)*

The observed increase in inlet concentrations are results of sewage divagation and reduced in-leakage to the sewage lines, increasing mass load, while reducing hydraulic loading. The other parameters show the same trend, also indicating higher inlet concentrations after divagation work (figure 3).

The total concept, including intensive and non-intensive unit processes, show approx. 90 % phosphorous removal and >90 % nitrogen removal. The biofilm reactor performance was improved from respectively 30 % (T-P) and 25 % (T-N), to efficiencies around 70 % total nitrogen and phosphorous removal, after introduction of the model based process control strategy.



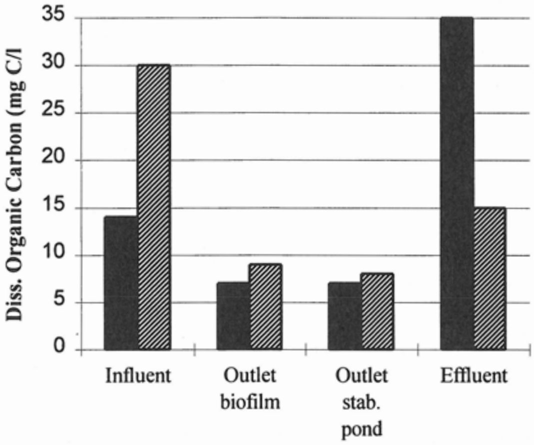


Figure 3 Influent and effluent concentrations following the intensive biofilm reactor, the semi-intensive stabilisation pond and the non-intensive wetlands. Performance before (black bars) and after (patterned bars) initiation of the nutrient removal control show a distinct increase in both suspended solids and DOC removal efficiencies. (Based on 170 samples).

The high effluent DOC concentrations are caused by humus wash out from the wetlands, historically being a fertilised pasture. This explanation is supported by the observed reduction from 1994 (black bar) to 1995 (patterned bar). As expected, most SS and DOC removal takes place in the aerated reactor and secondary clarifier. However, polishing in the wetlands contribute substantially.

3.2 Sludge quality

A total of 500 m^3 stabilised sludge has been exported for use on productive agricultural land during the last two years. Comprehensive chemical and bacteriological analysis were used to control sludge quality, in order to secure compliance to the national regulations. Table 2, summarise data.

Table 2 Chemical and bacteriological sludge quality exported from Øksnevad Treatment Plant, and regulative limit-concentrations. Amended values, following the 1996 revision, in brackets. (Based on 128 analysis, 8 samples per parameter)

Parameter	Average concentrations	Regulative limits	
Nutrients:			
Total Solids	10 g TS/kg	min. 200 g TS/kg	(200)
Total Carbon	0,74 g C/g TS	not regulated	(n.r.)
Kjeldahl Nitrogen	0,03 g N/g TS	not regulated	(n.r.)
Total Phosphorous	0,007 g P/g TS	not regulated	(n.r.)
Calcium	0,03 g Ca/g TS	not regulated	(n.r.)
Potassium	0,04 g K/g TS	not regulated	(n.r.)
Heavy metals:			
Cadmium	3,4 mg/kg TS	4 mg/kg TS	(2)
Lead	30 mg/kg TS	100 mg/kg TS	(80)
Mercury	1,4 mg/kg TS	5 mg/kg TS	(3)
Nickel	19 mg/kg TS	80 mg/kg TS	(50)
Zinc	1100 mg/kg TS	1500 mg/kg TS	(800)
Copper	260 mg/kg TS	1000 mg/kg TS	(650)
Chromium	17 mg/kg TS	125 mg/kg TS	(n.r.)
Pathogens:			
Total Coliforme Bact.	50 ind./g TS	2500 ind./g TS	(2500)
Salmonella. sp	not observed	0	(0)
Parasite eggs	not observed	0	(0)

All regulative requirements are met, accept for water content. However, Klepp municipality being the local authority governing sludge use, has exempted from the actual regulation in order to maximise recycling of nutrients. Analysis have shown that large parts of the nutrients are in fact dissolved or colloid, leading to nutrient loss during de-watering. Therefore, de-watering must be evaluated regarding transportation costs and nutrient value as the main opposing variables.

Even though the amended cadmium regulations do not enter force until the end of 1999, we face a serious challenge in reducing Cd-concentrations. However, preliminary studies have led our focus on internal heavy metals sources, and an extensive material replacement approach should bring both Cd and Zn levels down.

Further study and evaluation must be carried out to determine the systems maximum stabilising capacity. However, 14 separate (in time) samples show an overall reduction of COD from 190 g O_2/l (averaged sample from the import storage) to 6 g O_2/l in the exported sludge. Qualitatively the sludge satisfies the regulations, being free of offensive odours, and showing no signs of putrefaction in the soil.

Thermophilic aerobic sludge treatment provides several important advantages compared to tradition anaerobic digestion. At a minimum sludge retention time of 24 - 30 h, pathogen reduction is almost complete (see table 2 for details). In addition, high aerobic bacterial (thermophilic) metabolic rates increase stabilisation processes, resulting in shorter retention demand, thus lowering reactor volume. A third positive feature is the process' ability of reducing nitrogen loss. Whereas high temperatures and alkalinity often leads to increased pH in anaerobic reactors causing ammonia stripping, a continuos aerobic environment stabilise proton concentration between pH 6 and 7. Oxygenation may, however, cause ammonia stripping at high dosage of untreated sludge, thus requiring careful operation.

Retaining ammonia and reducing energy loss has also been achieved by installing the aerator suction side upstream the sludge flow. This means that air is sucked through the pre-treatment reactor before entering the high rate stabilisation process aerator. Further, the post-stabilisation/aerobic storage reactor draw its supplied air from the stabilisation reactor. This means that potential ammonia rich air is distributed in an ammonia free aerobic environment (as micro bubbles in nitrified stabile sludge) where hungry nitrifiers get energy from the ammonium. The oxygenation regime also reduce odour release resulting in a nearly completely nuisance free sludge treatment system.

Increasing solid detention time past the endogenous growth phase has revealed possibilities for achieving sludge nitrification. Although this is observed in the thermophilic reactor, complete nitrification is carried out in the third stage process, also functioning as sludge storage. Post-stabilisation in a separate reactor is combined with settling de-water function, creating an important carbon source for wastewater treatment processes and elevating suspended solids concentration before export. At the moment export sludge is de-watered to about 1 % TS, but this should be increased to 2-3 % TS.

The main source of sludge in the high rate thermophilic stabilisation reactor is septic from rural areas in the municipality, which is imported and treated parallel to internally produced sludge. The septic, normally transported to and disposed of at the regional landfill site, is thus converted from a problem and a potential environmental hazard, to a valuable soil fertiliser product. This practise is fully in accordance with the projects basic philosophy of ecological improvements and resource (nutrient) recovery and with the governments expressed sludge reuse goal.

3.3 Operation and maintenance

3.3.1 Wastewater treatment plant

Operation and maintenance procedures are approximately the same as for a normal size sewage pump station. Weekly inspections are done by a trained operator. All processes can be monitored and operated via a remote control system. In case of any complication in the process, alarms are transmitted to an operator on duty. Routines are made for periodic inspections of pumps and other installations. There are only two pumps and a sludge scrape involved in the process. Spare pumps are kept in stock and can be replaced within a couple of hours. Since the plant became operational in may 1994, one sludge pump has been repaired. Some problems with the telephone lines caused by lightening have been solved. Costs of operation are presented in table 3.

Table 3 O & M cost: Wastewater treatment plant

	Unit cost NOK	Annual cost NOK
Electricity	9127 kWh /year	4 563,-
Labour cost	39 hour/year	5 460,-
Water quality analysis		20 000,-
Maintenance cost; pumps etc.		6 000,-
Maintenance cost; building		2 000,-
Total		38023,-
Annual Cost per PE	300 P.e.	127,-

3.3.2 Sludge treatment plant

Septic sludge is imported three periods a year. During these periods there is a need for closer follow up of the process. Problems caused by foam in the process have been solved. Transfer of sludge from the pre-treatment reactor to the 2. and 3. reactors has been fully automated.

The process mechanically involves 3 aeration pumps and 3 sludge transport pumps. Also the sludge treatment plant can be monitored and operated via remote control. Costs of operation are presented in table 4.

Table 4 *O & M cost: Sludge treatment plant*

	Unit Cost NOK	Annual cost NOK
Electricity	40080 kWh /year	20 040,-
Labour cost	52 hour/year	5591,-
Sludge quality analysis		6000,-
Maintenance cost; pumps etc.		21000,-
Maintenance cost; building		6000,-
Other costs		2000,-
Treatment cost 1200 m3		60631,-
Treatment cost per m3		**50,-**
Transport of sludge to the plant		120,-
Transport of sludge to the farmer		40,-
Transport cost per m3		160,-
Total cost per m3		**210,-**

3.4 Cost analysis

Calculated cost efficiency for nutrient removal depends, to a large extent, on how you choose to perform the analysis. We have chosen to divide all operation and maintenance costs on phosphorus and nitrogen removal due to the fact that discharge limits are set for both for P and N.

P & N removal cost: 38.000,-/(31kg P + 358 kg N) * 90% = kr 108,- per kg.

The phosphorous removal process do not contribute to any increase in construction-, operation- or maintenance costs since it is completely integrated with the nitrogen removal process. We have looked at more sophisticated cost analysis also, but the general conclusion is the same: This is a very cost efficient treatment plant.

4. Conclusions

The Øksnevad wastewater and sludge treatment plant, which was designed to solve local pollution problems in a rural community, has been successfully operated for two years. Despite some technical problems, it has served the local environment well and it has served well as a pilot and demonstration plant for sustainable solutions to sanitary problems.

The wastewater is treated in a biofilm reactor with sedimentation, followed by a pond and wetland concept. About 90% total nitrogen and phosphorous removal from the wastewater is consistently achieved.

Sludge produced in the wastewater treatment and in septic tanks throughout the municipality is treated in a three stage aerobic digestion process designed for maximum nutrient recovery. The treated sludge is stable, nitrified, and odour free. The pathogen kill is efficient and the nutrient recovery is complete; total nitrogen in the final product is the same as in the raw sludge.

The sludge and wastewater treatment processes are closely integrated within the same building to maximise efficiency and minimise operation costs and avoid air pollution/loss of nutrients to the atmosphere. This integration has been a success, the processes are robust and cost efficient.