# REPORT

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# Enhanced Nutrients Removal in Membrane Bioreactor

# Project acronym: ENREM+

by

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# Abstract (English)

Within the 3.5 year ENREM project (Enhanced Nutrient REmoval in Membranebioreactors) in Berlin-Margaretenhöhe a novel and patented process was investigated to demonstrate the feasibility of a semi-decentralised solution reaching high effluent requirements set by the water authority of Berlin. This novel process could be a solution for suburban areas of Berlin which are not connected to central sewer system. The biological process combines enhanced biological phosphorus removal (EBPR) with post denitrification in MBR technology without dosing of any carbon sources. The process configuration of this demonstration plant enables advanced wastewater nutrients removal (C, P and N) and could be a promising option for wastewater treatment wherever high effluent qualities are required. A second prototype MBR system was operated in parallel, applying a different biological process, e.g. without biological phosphorus removal, enabling a comparison of these different technological approaches.

The demonstration plant showed high elimination rates for COD (>95%), phosphorus (>99%) and nitrogen (up to 98%) when operated within the appropriate range of design conditions. The operational experience within the first years showed that there is a possibility for process stabilisation by changing the ratio of the process steps. For this reason the volume of the anoxic zone was enlarged by reducing the aerobic volume in Feb 2008. The positive effects could be seen on the basis of the effluent concentrations after a short period of adaptation.

The membrane filtration performance was very reliable with a new cleaning strategy: Two membranes were operated alternating with an operational flux of 15 – 20 L/m²/h and a maintenance cleaning with low chemical concentration. Different cleaning agents were used in order to evaluate the cleaning efficiencies.

An economical evaluation of the demonstration plant was performed in comparison to the existing wastewater treatment costs of app. 7 €/m$^3$ by trucking away and the prototype MBR plant. Operated on the same site, the two MBR systems were used to calculate the actual costs, in relation to the effluent quality, and to perform a scale-up up to 5000 pe considering four different effluent quality classes. The results showed that the ENREM process applied in the demonstration plant is economically an alternative for plant sizes of 5000 pe and larger. For plant sizes smaller than 5000 pe, the prototype MBR system equipped with precipitation and a downstream adsorption filter for enhanced phosphorus removal proofed to be the more viable solution.

# Abstract (German)

Im Rahmen des ENREM Projektes (Enhanced Nutrient REmoval in Membranebioreactors) am Standort Berlin-Margaretenhöhe wurde ein neuartiger und patentierter Prozess über 3.5 Jahre untersucht. Der Einsatz dieses Prozesses in der dezentralen Abwasserreinigung bei sehr hohen Ablaufgüten sollte bewiesen werden. Ein solcher Prozess wäre eine mögliche Lösung für Wohngebiete am Rande Berlins, die nicht an das zentrale Abwassersystem angeschlossen sind. Der Prozess kombiniert die biologische Phosphorentfernung mit einer nachgeschalteten Denitrifikation in einem Membranbioreaktor ohne die Zugabe externer Kohlenstoffquellen. Die Prozessführung dieser Demonstrationsanlage ermöglicht eine hohe Nährstoffentfernung (C, P und N) und ist damit eine mögliche Anwendung, wenn hohe Ablaufgüten erforderlich sind. Ein zweiter Prototyp einer MBR-Anlage mit einem anderen biologischen Prozess, z.B. ohne biologische Phosphorentfernung, wurde ebenfalls betrieben und ermöglichte einen Vergleich dieser unterschiedlichen Lösungsansätze.

Bei planmäßigen Betrieb zeigte die Demonstrationsanlage sehr hohe Eliminationsraten für CSB (>95%), Phosphor (>99%) und Stickstoff (bis zu 98%). Die Betriebserfahrungen der ersten Jahre zeigte, dass eine Optimierung der Reaktorverhältnisse eine Stabilisierung des Prozesses erwarten ließ. Aus diesem Grund wurde der anoxe Anteil am Reaktorvolumen erhöht, indem das aerobe Volumen im Februar 2008 verringert wurde. Die Ablaufwerte verbesserten sich deutlich nach einer kurzen Phase der Anpassung an die neuen Milieubedingungen.

Die Membranfiltration zeigte eine zuverlässig gute Leistung und eine neue Reinigungsstrategie wurde erfolgreich getestet: Zwei Membranmodule wurden abwechselnd bei einem Flux von 15-20 $L/m^2/h$ betrieben und eine Reinigung mit niedriger Konzentration wurde monatlich durchgeführt. Verschiedene Reinigungsmittel wurden getestet, um deren Effizienz vergleichen zu können.

Eine wirtschaftliche Betrachtung der Demonstrationsanlage wurde durchgeführt, um die anfallenden Kosten mit den bisherigen Entsorgungskosten durch Abfuhr von ca. 7 €/$m^3$, sowie den Kosten der Prototypanlage zu vergleichen. Die Anlagen wurden genutzt, um die anfallenden Kosten in Abhängigkeit der Reinigungsleistung zu bestimmen. Des Weiteren wurden die ermittelten Ergebnisse verwendet um die Anlagen bis zu einer Größe von 5000 EW maßstabgerecht zu vergrößern. Die dabei gewonnenen Ergebnisse zeigten, dass die Demonstrationsanlage ab einem Einzugsgebiet von 5000 EW bei hohen Ablaufqualitäten wirtschaftlich konkurrenzfähig ist. Bei kleineren Einzugsgebieten war der Prototyp, zur vermehrten Phosphorentfernung ausgestattet mit einer Fällung und einem Adsorptionsfilter, die wirtschaftlich vertretbare Lösung.

# Abstract (French)

Le projet ENREM (Enhanced Nutrient REmoval in Membranebioreactors) s'est déroulé à Berlin-Margaretenhöhe pendant 3 ans et demi. Un nouveau procédé breveté a été testé pour démontrer la faisabilité d'un traitement semi décentralisé capable d'atteindre un effluent de haute qualité selon les critères établis par les autorités de l'eau de Berlin. Ce nouveau procédé pourrait être une solution pour les banlieues de Berlin qui ne sont pas connectées au système central de traitement des eaux usées. Le procédé biologique combine une amélioration de l'abattement du phosphore avec une post dénitrification dans la technologie BRM sans ajout de source de carbone. Dans cette configuration, le pilote de démonstration permet un abattement avancé des nutriments de l'eau usée (C, P et N) et peut être une option prometteuse pour le traitement des eaux usées lorsque des effluents de hautes qualités sont nécessaires. Un deuxième pilote, le prototype BRM a fonctionné en parallèle, appliquant un procédé biologique différent, par exemple sans abattement de phosphore biologique, permettant une comparaison des différentes approches technologiques.

Le pilote de démonstration a démontré un haut taux d'élimination de la DCO (>95%), du phosphore (>99%) et de l'azote (plus de 98%) lors de son fonctionnement sous des conditions appropriées et optimisées. Les essais ont montré lors de la première année que le procédé peut être stabilisé en changeant la proportion de chaque étape du procédé. Pour cette raison, le volume de la zone anoxique a été agrandi en réduisant le volume aérobique en février 2008. Les effets positifs ont été observés sur la base des concentrations de l'effluent après une courte période d'adaptation.

Les performances de filtration étaient très satisfaisantes avec l'implémentation d'une nouvelle stratégie de nettoyage : deux membranes fonctionnaient en alternant un flux opératoire de 15-20 L/m²/h et un nettoyage de maintenance à faible concentration chimique. Différents agents de nettoyage ont été utilisés dans le but d'évaluer leurs efficacités.

Une évaluation économique du pilote de démonstration a été effectuée en comparaison avec le coût actuel du traitement des eaux usées par collecte en camion qui est d'environ 7 €/$m^3$ et le coût du prototype BRM. Les 2 pilotes BRM, installés sur le même site, ont été utilisés pour calculer les coûts actuels en fonction de la qualité de l'effluent. Ces résultats ont permis de calculer et prévoir le coût du procédé pour des installations allant jusqu'à 5000 EH. Les résultats montrent que le procédé ENREM (pilote de démonstration) est une alternative économique pour une taille d'installation supérieure ou équivalente à 5000 personnes. Pour des installations plus petites que 5000 EH, le prototype BRM prétraité par précipitation et équipé d'un post filtre à adsorption dans le but d'améliorer l'abattement de phosphore apparaît comme la solution la plus fiable.

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# Glossary

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|---|---|
| A | Acids |
| AE | Aerobic reactor |
| AN | Anaerobic reactor |
| AOX | Absorbable Organic Halogen |
| AX | Anoxic reactor |
| AST | Activated Sludge Treatment |
| BB | Building Blocks |
| bEPS | Bound Extracelullar Polymeric Substances |
| BFM | Berlin Filtration Method |
| BOD | Biological Oxygen Demand |
| BSP | Bench-Scale Plant |
| CAS | Conventional activated sludge |
| CCTV | Closed Circuit Television |
| CIP | Cleaning In Place (curative cleaning) |
| COD | Chemical Oxygen Demand |
| Deox | Deox pot (oxygen release) |
| DIN | Deutsches Institut für Normung e.V. |
| DNR | Denitrification rate |
| DOC | Dissolved Organic Carbon |
| EBPR | Enhanced Biological Phosphorus Removal |
| EPS | Extracelullar Polymeric Substances |
| GAO | Glycogen Accumulating Organisms |
| IMF | Immersed Membrane Filtration |
| HOC | Hydrophobic part of organic carbon |
| HPLC | High Performance Liquid Chromatography |
| HRT | Hydraulic Retention Time |
| MF | Micro Filtration |
| MR | Membrane Reactor |
| MBR | Membrane Bioreactor |
| N | Neutrals (for DOC fractionation) |
| N | Nitrogen |
| NR | Nitrification Rate |
| $oPO_4$-P | Orthophosphate |
| P | Phosphor |
| p.e. | population equivalent |
| PAO | Polyphosphate Accumulating Organisms |
| PR | Proteins |
| PRR | Phosphorus Release Rate |
| PUR | Phosphorus Utilisation Rate |
| PS | Polysaccharides |
| PP | Prototype Plant |
| $P_x$ | Refractory Phosphorus Concentration |
| rbCOD | readily biodegradable Chemical Oxygen Demand |
| sbCOD | slowly biodegradable Chemical Oxygen Demand |
| sEPS | Soluble Extracelullar Polymeric Substances |
| SH | Humic Substances |
| SRT | Sludge Retention Time (= “sludge age”) |
| T | Temperature |

| | |
|---|---|
| TMP | Transmembrane Pressare |
| TN | Total Nitrogen |
| TP | Total Phosphor |
| TS | Total Solids |
| UF | Ultra Filtration |
| UPS | Uninterrupted Power Supply |
| VFA | Volatile Fatty Acid |
| wd | weekday |
| we | weekend |
| WWTP | Wastewater Treatment Plant |

# Chapter 1

# Background and motivation for project extension

With the end of the ENREM project which was performed from January 2004 to June 2007 some elementary questions were still open because of lack of time and necessary requirements for reliable plant operation (Gnirss *et al.* 2007). For example the primary filtration system was unsatisfying because of hydraulic problems in the membrane bioreactors and was changed in May 2007 to enhance the operation of filtration. Furthermore the expected high nutrients removal efficiency could only be shown within short periods because of plant overloading and operational problems within the first years. Also the question of the unknown carbon source which is assumed to be responsible for the high denitrification rates in the process could not be answered.

Because of the still ongoing plant optimization after the ENREM final report (Gnirss *et al.* 2007) and the well-founded expectation of gaining valuable information while continuing the research within the next months, the project was extended by one year up to December 2008. The extension has not only given the opportunity to complete the successful operation but also further investigations on decentralized wastewater treatment plants were possible. In the end a better understanding of the used process and its capacities and limits were achieved.

## 1.1 Objectives and adopted strategy of the project

It was proposed to extend the project by one year in order to match the initial outcomes of the project while allowing sufficient time to implement the required compensatory measures and to undertake the appropriate assessment. This included

- Long-term evaluation of another MBR filtration system (technology from A3 water solution was built in May 2007)
- Installation of a MBR prototype unit side-by-side to operate and enable the assessment of the demonstration unit under design load
- Identification of best long term and durable operation conditions for BWB, including membrane cleaning strategy, foam handling and aeration control

The second MBR unit was to be rent over at least 12 months. In contrary to the existing demonstration plant, it was built with a single aerated reactor for carbon removal and full nitrification only. This enabled therefore, as additional outcome to the project, to evaluate the performances of such simple systems, also in terms of operation costs (energy, maintenance, cleaning). This provided a full picture of the costs of MBR units for decentralised applications, with or without advanced treatment for nitrogen and phosphorus removal.

## 1.2 Project organisation

The project was subdivided into four work packages according to the objectives mentioned above. They were carried out simultaneously as described below.

**Work package 1: Demonstration plant operation**

The focus of the work package 1 was on the stable operation of the demonstration plant under design conditions (about 11 m3/d). This required a reduction of the incoming nutrients load that was reached mainly by installing a second MBR unit treating the excess of about 5 m3/d (see Work package 4).

Task 1.1: Performance assessment under design load

In this first part of the work package the nutrients load to the demonstration plant was reduced to the design parameters. This was reached through the treatment of part of the wastewater in a second MBR plant and removal of excess wastewater by trucks. The statement was to be made within which nutrients load the process can be run reliably.

Task 1.2: Sedimentation

Sedimentation trials investigated the possibility of nutrients load reduction and the usefulness of installing a sedimentation tank to relieve the plant biology.

Task 1.3: TS measurement

The plant operation should be simplified by using the TS concentration as control parameter (instead of “sludge age”). This assumes reliable TS measurements that were not reached with the initial TS sensor. New probe constructors were investigated with if possible plant operation under set TS.

Task 1.4: VeoliaLink

The prototype software of VeoliaLink was tested in 2006, but interaction problems with the local PC resulted in deactivating the software. In 2008, the commercial version of VeoliaLink, which should be more stable than the prototype, was installed by Veolia on the demonstration plant for long term usage and testing.

Task 1.5: Foam

Foam was one of the major problems for a reliable plant operation within the last 2 years and the experiences are to be investigate more in detail: Foam appearance in combination with high / low TS concentrations, high / low aeration levels and inflow conditions as well as foam destroying by stirring or aeration.

Task 1.6: Transfer to operation department

The on site work and presence / maintenance had to be reduced to promote the operation transfer to the operation department. This task was strongly linked with the process optimisation to decrease the manual settings. One operation manual was prepared for plant take over by the operation department.

Task 1.7: Economical evaluation

The economical comparison between the decentralised solution (plant extension for long term operation) and the connection to the public wastewater system was finalised within the project.

**Work package 2: Biology and process optimization**

The investigations on the biological processes provided important information about the prospective design parameters, the control strategy and the every day operation. The planned activities can be divided in the following tasks.

Task 2.1: Internal carbon source used for post-denitrification

The internal carbon source used for denitrification was unknown and was also a topic of this year research interests. Therefore a lab scale sequential batch membrane bioreactor was designed in cooperation with the Technische Universität Berlin. The start up and the adaption to a synthetic monosubstrate took place in Berlin whilst the investigations to identify the carbon source used was carried out in Lisbon in cooperation with the Universidade Nova Lisboa. These investigations helped to understand the microbiological processes involved and might lead to the identification of the microbiological community responsible for the enhanced nutrient removal with post-denitrification without external carbon dosing.

Task 2.2: Biological extension / Process optimisation

Due to the experiences of the first months of operation, the process in the present conditions seemed to be sensible to disturbances. For instance, change of influent concentrations and ratios might lead to insufficient denitrification, which results, besides high effluent concentrations, in a recirculation of nitrate to the anaerobic chamber. Denitrifying bacteria will compete with phosphorus accumulating organisms (PAOs) for substrate. This leads to a handicapped biological phosphorus removal and in the long term to the loss of PAOs.

Changes in the process control, sometimes requiring plant modifications, were investigated to increase the performance of the process.

(1) Extension of the anaerobic zone

(2) Prevention of aerated sludge backflow to the anaerobic zone

(3) Extension of the anoxic zone (or simultaneous nitrification / denitrification in AE2)

(4) Enhancement of aeration control / Reduction of O2-carryover

The planned changes of the set up of the existing plant provided the ability to provide complete anaerobic conditions to improve the performance of the biological phosphorus removal. These activities changed the ratios between anaerobic-aerobic-anoxic phases, which stabilized the biological process.

Task 2.3 Carbon addition

Once an unbalanced inflow has caused a disturbance of the biological process, certain ways of operation have to be implemented to recover the overall performance. Both, the biological phosphorus removal and the post-denitrification, rely on anaerobic conditions in the first step of the process. To reduce the effect of nitrate-recirculation to the anaerobic chamber organic acids are fed to support Bio-P. Different organic acids are going to be tested in order to identify the approach to a disordered system. In a first attempt, propionate was added to foster the post-denitrification and the biological phosphorus removal, later acetate was added to the system.

Task 2.4: Phosphorus precipitation

The demonstration plant was running up to now with acetate addition and without any precipitation. The use of a ground precipitation to support and stabilise the P-removal was investigated to ensure low and stable effluent concentrations.

**Work package 3: Fouling, membrane filtration performance and cleaning**

The already running investigations on membrane performance and cleaning strategies was further carried out and the results provided helpful hints for operational guidelines.

Task 3.1: Investigations on fouling

Weekly monitoring of several sludge characterisation parameters thought to be fouling indicators were performed on the 2 MBR units, using different techniques to identify soluble / colloidal organic substances as well as the examination of various parameters characterising the sludge.

Critical flux measurements were performed different devices and protocols. A comparability study was therefore possible.

Task 3.2: Cleaning strategies

As mentioned above there was an ongoing investigation on cleaning strategies with different chemical agents (H2O2 pH 11 vs chlorine solution). In addition to these strategies different cleaning protocols have been used for recovering the membranes when significant fouling occured.

In the end of this project, the performance of the different cleaning strategies was assessed and recommendations were issued for long term operation.

**Work package 4: Parallel MBR prototype unit**

The installation of the second MBR prototype unit did not only allow to solve the problems faced so far, but provided the opportunity to compare two processes used for decentralized wastewater treatment.

On the one hand, the already existing installation achieves the stringent effluent qualities using a recirculation of the activated sludge through anaerobic, aerobic and anoxic chambers implementing biological phosphorus removal with post-denitrification without external carbon addition.

On the other hand, the new installation is designed for carbon removal and full nitrification, (i.e. full aeration) and features low-cost and low-energy equipment. The volumetric capacity is planned to be 4 - 5 m3/d, the unit therefore treated about 30-40% of the flow entering the demonstration plant.

The filtration and treatment performances were assessed and compared with the demonstration plant. A comparison of capital and operation costs (energy requirements etc) was also performed.

The operation of two different processes on the same site, feeding the same influent provided the rare opportunity to compare those processes with the same surrounding under realistic conditions. The comparison of the activated sludges cultivated provided also information on how to maintain the microbiological community.

Task 4.1: Operation of prototype MBR plant

As a total different control scheme is used, the start up phase was required to gain experience to operate the prototype plant. Once stable conditions were achieved, i.e. targeted TS-concentration, the following tasks was executed.

Task 4.2: Nitrification / Denitrification

Although the plant was designed for nitrification only, easy adjustments within the set up and the control scheme were tested to identify the denitrification capacity of the plant. At the same time this provided the possibility to optimize process parameters as pH-value and dissolved oxygen concentration and therefore the energy consumption.

Task 4.3: Membrane performance, economical evaluation

This plant gives two options to control the permeate flow. Beside the use of a pump to assure a defined flux, the possibility to use gravity as driving force is provided. The use of gravity has obviously some advantages, just to mention energy savings or membrane protection, but has to be tested for our purposes.

The capital and operation costs of the “low tech” MBR system was ascertained and compared with those of the demonstration unit.

# Chapter 2

# Demonstration plant operation

## 2.1 Process description

The ENREM process combines EBPR and post-denitrification without carbon source addition together with a membrane filtration for separating sludge and treated wastewater. This process is the result of the previous demonstration project IMF (Adam and Kraume 2003; Lesjean *et al.* 2004) which demonstrated the advanced biological phosphorus and nitrogen removal with 2 recirculation loops and sludge ages > 25d. The drawback of this process is a larger anoxic volume (+ at least 50%) compared to conventional processes with pre-denitrification due to lower denitrification rates, which is generally not an issue for small units. On the other hand the benefits are lower possible nitrogen effluent values and biological plant volume reduction because of higher possible TS concentrations due to the membrane filtration. The flow scheme is given in Figure 1.

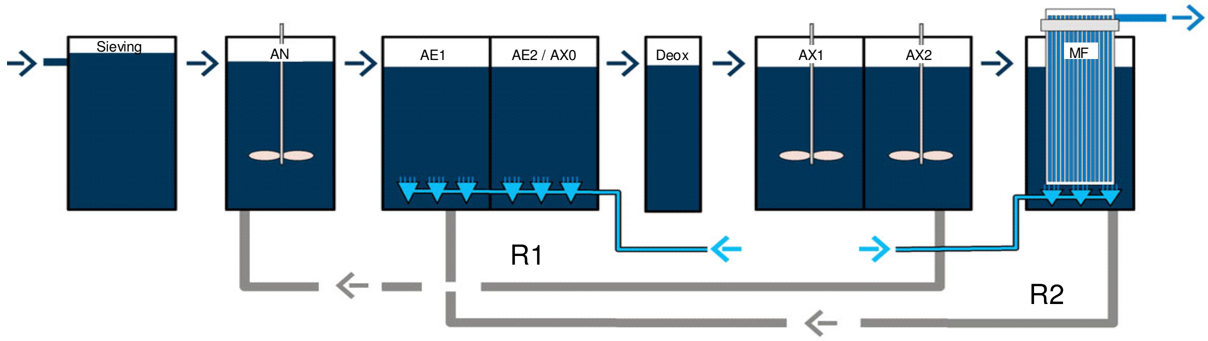



**Figure 1: Schematic of the ENREM process**

## 2.2 Plant description

The raw wastewater is first collected in household buffer tanks equipped with grinding pumps and then transported in a low-pressure drainage system to the buffer tank in front of the MBR plant. An influent valve which is installed before prevents overflowing of the buffer tank. The buffer tank homogenizes both the irregular wastewater flows and the different incoming nutrients concentrations in order to reach even feeding of the MBR plant (see (Gnirss *et al.* 2007)). A second buffer tank is installed for excess sludge, screenings, resultant waste water from membrane cleanings and later for excess wastewater.

The flow scheme is shown in Figure 2.

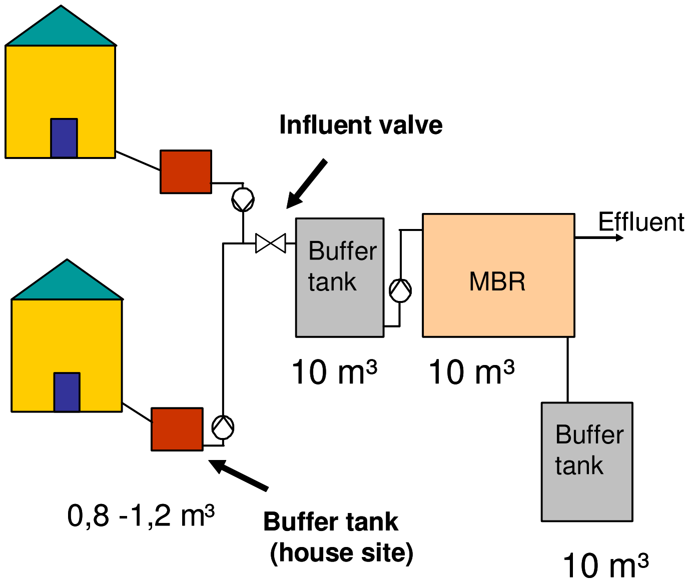
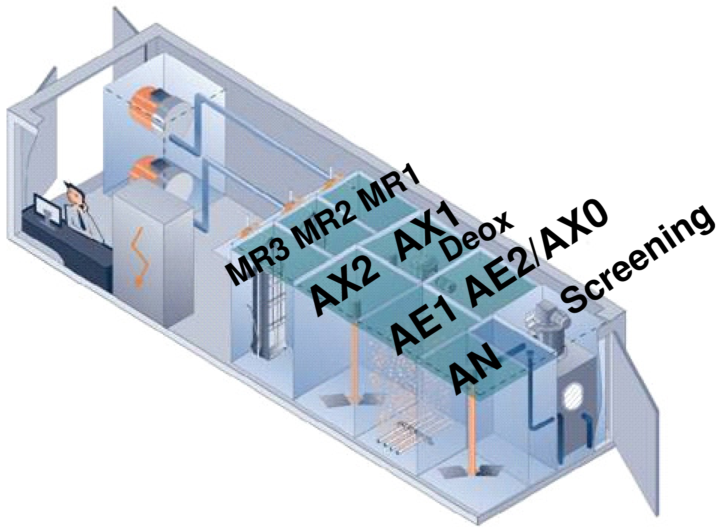



**Figure 2: Flow scheme of network and MBR plant**

Two raw water pumps are operated for constant feeding of the plant according to the water level in the buffer tank and in AX2. The screening is set up directly in the container where also the biological reactors are located. The screenings tank has a volume of 0.56 m3.

The dimension of the MBR prototype plant is about 10 $m^3$ (15 up to 30h HRT) and consists of a 2.5m deep rectangular shaped tank which is divided in 1 anaerobic reactor, 2 aerobic reactors (one converted to an anoxic reactor later on), 1 de-aeration zone, 2 anoxic reactors and 3 parallel membrane reactors. Collection channels before and after the three membrane units distribute the flow equally. The configuration of the reactors is given in Figure 3 and the size of the reactors is given in Table 1. The biological reactors were requested to suit the full design capacity and the filtration unit should have redundancy due to cleaning and/or maintenance. Two sludge recirculation loops (done by eccentric screw pumps) are necessary for the ENREM process (see Figure 1). For optional water reuse (cleanings) the filtrate is collected before discharge in a 0.98 $m^3$ filtrate tank which is located between the screening and the anaerobic zone.

In order to warrant the maximum volume for the biological reaction before filtration, a design constraint was to keep the volume of the filtration reactor smaller than 10% of the entire mixed liquor volume (Gnirss *et al.* 2008).




| Zone | Volume |
|---|---|
| AN | 0.70 m³ |
| AE1 | 1.86 m³ |
| AE2 / AX0 | 1.85 m³ |
| Deox | 0.15 m³ |
| AX1 | 1.80 m³ |
| AX2 | 1.91 m³ |
| MR (1 line) | 0.69 m³ |
| Total | 8.96 m³ |

**Figure 3: Configuration of the MBR plant**

**Table 1: Reactor sizes**

Two air blowers with a capacity of 60 $Nm^3/h$ each are set up in parallel sustained the air requirement of the biological and membrane units (each one in redundancy for the other). An air conditioning system is required for the cooling of the dry area inside the container, especially because of the heat of the blowers.

## 2.3 Performance assessment under acceptable overload conditions

The plant was designed to handle with expected 4 – 10 $m^3/d$ within the first 1-2 years which was the relevant period for the trials of the ENREM project. Experiences from other decentralized areas showed that there were household connection rates of at maximum 80% within 2 years which caused the quite careful plant design dealing with low expected amounts of incoming wastewater. The design parameters of the plant are shown in Table 2.

**Table 2: Design concentrations and loads for plant operation**

| Parameter | Concentration | 50%-tile Daily Volume Load (@ min flow of $4m^3/d$) | 50%-tile Daily Volume Load (@ max flow of $10m^3/d$) |
|---|---|---|---|
| BOD5 | 493 mg/L | 0.25 $kg/m^3/d$ | 0.62 $kg/m^3/d$ |
| COD | 986 mg/L | 0.49 $kg/m^3/d$ | 1.23 $kg/m^3/d$ |
| TS | 356 mg/L | 0.18 $kg/m^3/d$ | 0.45 $kg/m^3/d$ |
| TKN | 108 mg/L | 0.05 $kg/m^3/d$ | 0.14 $kg/m^3/d$ (peak 0.15) |
| TP | 15 mg/L | 0.009 $kg/m^3/d$ | 0.019 $kg/m^3/d$ |
| VFA | 94 mg/L | 0.005 $kg/m^3/d$ | 0.012 $kg/m^3/d$ |

After commissioning it came apparent that the connection rate as well as the nutrients concentrations and flows of the wastewater exceeded the assumptions considerably (Gnirss *et al.* 2007). The plant was running after approx. 6 month almost all the time with high overload conditions especially for nitrogen and phosphorus although some of the incoming wastewater was trucked away.

Within the extended project of ENREM+ the plant load should be reduced by installing a second MBR plant. This second plant should handle the excess wastewater in order to demonstrate the long-term performance of the ENREM plant within its design loads.

Unfortunately the second MBR plant was not able to treat all of the excess wastewater. The capacity of the plant did not reach the expected throughput of around 5 $m^3/d$ (see Chapter 5) and the incoming wastewater increased slightly from 2007 to 2008 and was too much for both plants especially during weekends (see Section 2.3.2). It has to be noted that the throughput of the second MBR plant was also reduced because of necessary denitrification trials because of restrictions of the Water Authority.

To minimize the costs for trucking away services it was decided to run both plants near their maximum capacities with the restriction of achieving permanent high nutrients removal rates. The optimal operation parameters should be found out with the given reactor volumes and optimizing the process engineering. It came apparent in the former project phase that the treatment restrictions are related to the volume of the biology and not to the membrane filtration.

Because of the high influent flows on weekends (see Figure 6) the plant throughput was increased by approx. 20% for 2-3 days per week by guaranteeing good treatment performance. The throughput restrictions were caused only by the biological reactor volume and not through the filtration performance (see Section 4.1). In the circumstances explained before the plant was not operated under design load but under acceptable overload conditions in the range of its maximum capacity.

### 2.3.1 Trials program and operation conditions

The period of examination (July 2007 – December 2008) is subdivided into 3 periods:

Period 1 Operation with a new filtration system (replacement of Martin Systems technology -> A3 technology because of hydraulic problems with the first technology, (Gnirss *et al.* 2007)

Period 2 Conversion of aerobic Zone 2 into anoxic zone 0 from February 08

Period 3 Start of co-precipitation with low amount of ferric precipitants from August 08

Continuous ·operation of the MBR plant started in March 2006 with seeding of sludge from a large WWTP with EBPR. After the period up to June 2007, which is described in the (Gnirss *et al.* 2007), three trial periods from July 2007 to October 2008 can be described. In the first period the previous main hydraulic problems were fixed up and steady state conditions were reached with modules from A3 (Period 1, 6 months), but still overloading the plant by 20 – 50% for nitrogen and up to 100% for phosphorus compared to the design (see Figure 4 and Figure 5). In February 2008 a process optimization was realized with changing the aerobic zone (AE2) into an anoxic zone (AX0) which resulted in a more stable process and lower effluent concentrations (Period 2, 6 months). The overload of the plant was still 20 – 50% (100% for phosphorus). Up to August 2008 there were several carbon dosings (see Section 3.1) but no ferric precipitation at all. From August 2008 on (Period 3, 4 months) a co-precipitation with low amount of ferric precipitant was implemented for a further stabilization and reduction of the effluent values.

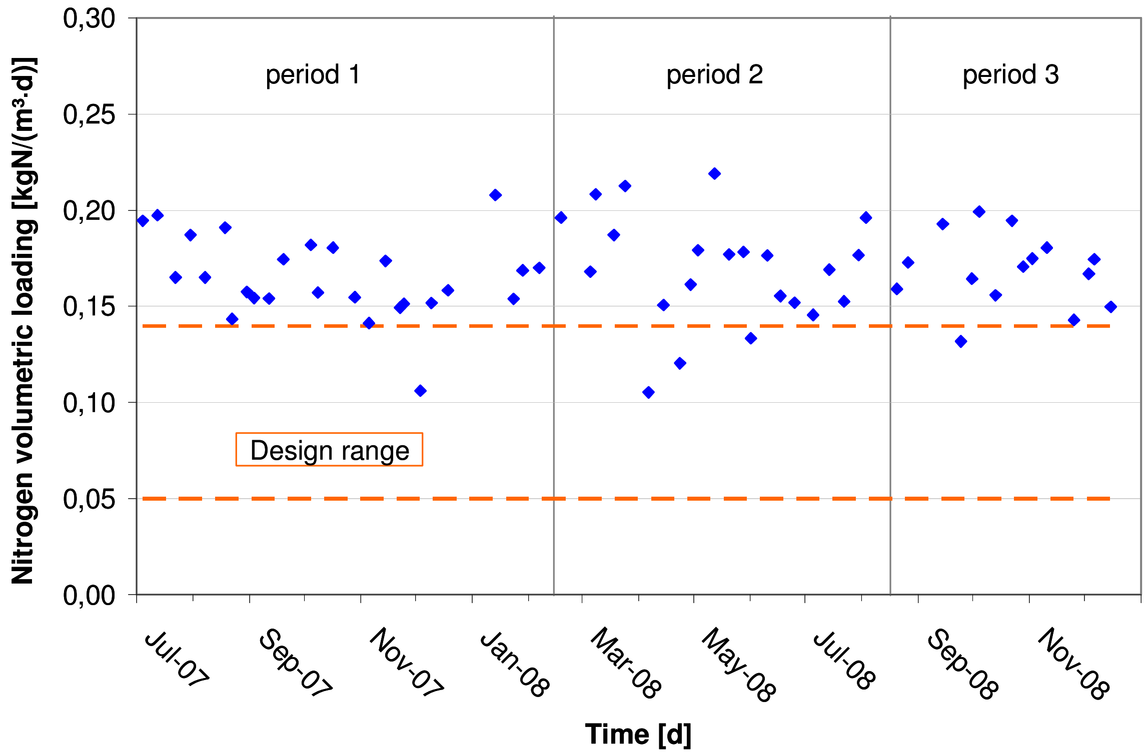



**Figure 4: Nitrogen volumetric loading over time (24h average samples)**

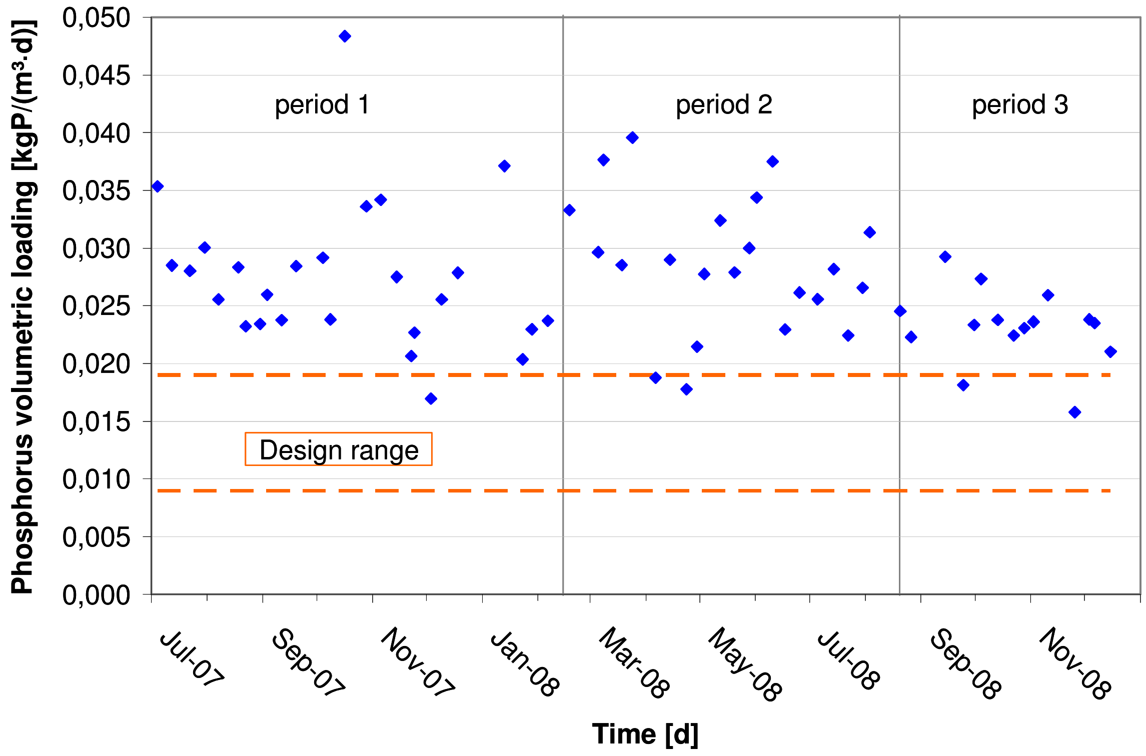



**Figure 5: Phosphorus volumetric loading over time (24h average samples)**

The sludge return and recycle ratios were set up at 150% from the anoxic to the anaerobic reactor (R1) and 400% from the membrane reactor to the aerobic zone (R2). The recycle rate leads to a contact time of 45-60min in the fully mixed anaerobic reactor and a sludge mass in the anaerobic reactor of only 4.8 % of total sludge in the MBR plant. Usually, at least 10% are required for sufficient EBPR performance. The volume ratio between anoxic and aerobic zone was before converting an aerobic zone into an anoxic zone 51:49 and afterwards 75:25. Especially in the ENREM process a larger anoxic volume is necessary because no carbon source is dosed and lower denitrification rates occur. The chamber conversion resulted in a much more stable process. During the trials, the pH-values were usually between 7.9 and 8.1 throughout the reactors.

The most important operation parameters for each period are given in Table 3. Some parameters in this table are calculated for both operating modes, weekday and weekend plant operation where weekday operation is defined with plant throughputs up to 11 $m^3/d$ and weekend operation above.

**Table 3: Average operational parameters of biological system**

| Parameter<br>Period | Unit | 1 | | 2 | | 3 | |
|---|---|---|---|---|---|---|---|
| Content | | **Start running A3 modules** | | **Conversion AE2 – AX0** | | **Start Precipitation** | |
| Time | | **1.7.2007-17.2.2008** | | **18.2.2008-20.08.2008** | | **21.8.2008-31.12.2008** | |
| Anaerobic reactor volume | $m^3$ | 0.7 | | 0.7 | | 0.7 | |
| Biological reactor volume<br>(Vax + Vae)<br>(Vax + Vae + Vm) | $m^3$ | <br>7.6<br>8.3 | | <br>7.6<br>8.3 | | <br>7.6<br>8.3 | |
| Volume ratio $V_{ax}$ : $V_{ae+m}$<br>$V_{ax}$ : $V_{ae}$ | % | 47:53<br>51:49 | | 69:31<br>75:25 | | 69:31<br>75:25 | |
| Temperature range (AE2) | °C | 9.5 – 24.0 | | 11.4 - 26.1 | | 11.9 – 24.7 | |
| Sludge concentration (AX2) | gTS/L | 11 (8*) -15 | | 12 (10*) - 17 | | 12 - 15 | |
| Sludge age | d | 20 - 60 | | 20 - 50 | | 18 - 30 | |
| Air flow biology (mean) | $Nm^3/h$ | 36 | | 46 | | 40 | |
| DO (AE1) mean / (goal) | mg/L | 3.4 (2.0) | | 5.7 (2.0) | | 2.4 (2.0) | |
| weekday (wd) / weekend (we) | | wd | we | wd | we | wd | we |
| Total retention time | h | 21.5 | 16.6 | 19.9 | 16.6 | 21.5 | 17.9 |
| Total contact time | h | 3.4 | 2.6 | 3.1 | 2.6 | 3.4 | 2.8 |
| Net flow | L/h | 420 | 540 | 460 | 540 | 420 | 500 |
| COD load | kg/d | 10.0 | 11.8 | 9.4 | 13.6 | 8.4 | 9.8 |
| N laod | kg/d | 1.20 | 1.41 | 1.17 | 1.49 | 1.23 | 1.42 |
| P load | kg/d | 0.20 | 0.23 | 0.20 | 0.25 | 0.17 | 0.20 |
| Mass organic load<br>(based on $V_{ax}$ + $V_{ae}$) | kgCOD/ (kgTS·d) | 0.114 | 0.118 | 0.097 | 0.117 | 0.083 | 0.098 |
| Mass nitrogen load<br>(based on $V_{ax}$ + $V_{ae}$) | kgN/ (kgTS·d) | 0.013 | 0.014 | 0.012 | 0.013 | 0.012 | 0.014 |
| Volume organic load<br>(based on $V_{ax}$ + $V_{ae}$) | kgCOD/ ($m^3$·d) | 1.32 | 1.55 | 1.24 | 1.79 | 1.11 | 1.29 |

* TS after foaming event

### 2.3.2 Evolution of inflow

The area of Margaretenhöhe contains approx. 230 persons in 90 households, whereas 20% of the households are inhabited only during the summer period.

The daily inflow to the MBR plant is shown in Figure 6 separated in weekday and weekend inflow. The plant operation started with around 4 m³/d in 2006 and a full connection of the sewer area at the end of 2006 resulted in an average inflow of 12-16 m³/d on weekdays and 16-20 m³/d on weekends during winter season from 2007 on. The designed max. inflow of 10 m³/d was already reached in June 2006, therefore wastewater handling with other opportunities was necessary (see 0). The assumption in designing the plant was that in 2006 the wastewater will not exceed 10 m³/d and in the following years the throughput could increase up to 20 m³/d while keeping the same organic and nutrients daily load (i.e. only increase of daily water use per capita). In reality not only the discharge increased but also the nutrients concentration exceeded the layout conditions and did not decrease with higher discharge. The decreased inflow in winter was mainly related to occupation of some parcels only in summer.

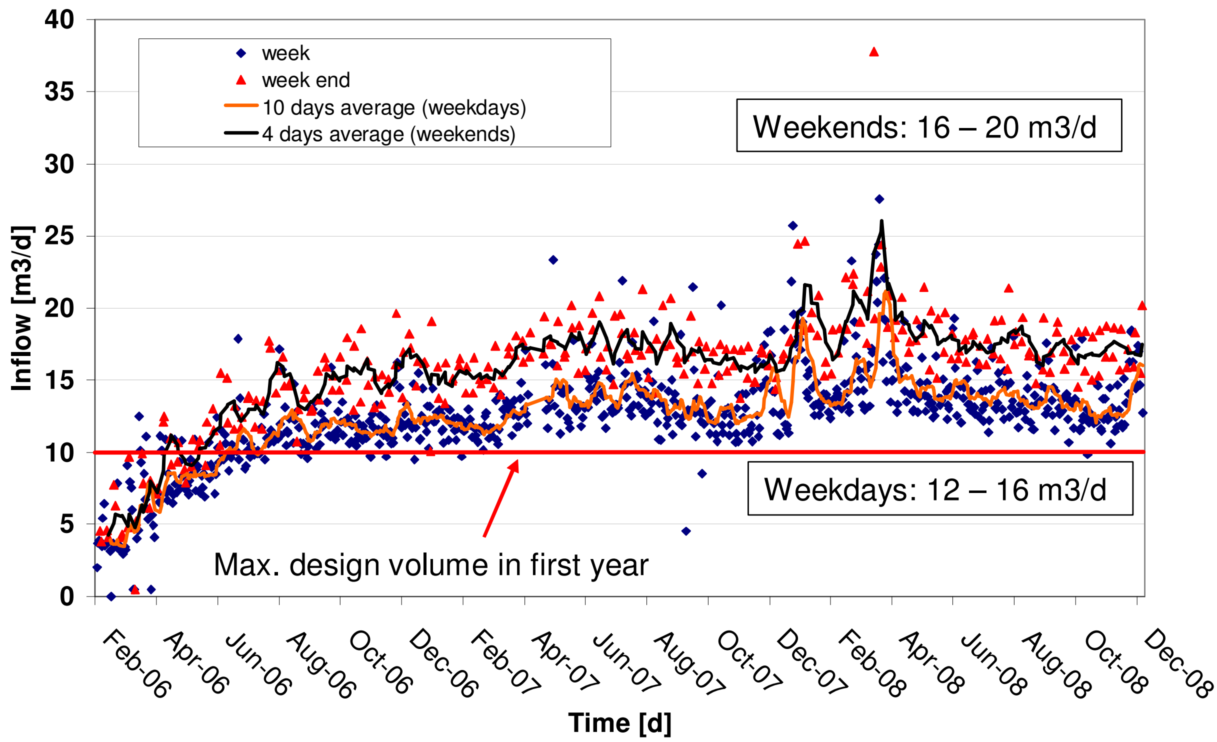



**Figure 6: Daily inflow to the MBR plant over time**

Single heavy discharges higher than 20 m³/d were observed at irregular intervals. It is assumed that rain events (some septic tanks were converted into rain water tanks) and a high groundwater level (basement drainage) are responsible for these events. The occurrence of infiltration water can be excluded because there is often no inflow during night hours, no air sewer flushing and no relation between inflow peaks and heavy rain water events. In March 2008 there were heavy rain water events which caused discharges up to 38 m³/d.

### 2.3.3 MBR plant throughput

The MBR plant was able to clean around 70 % of the incoming wastewater due to limitation of the biological capacity. The filtration could have managed all the incoming wastewater, only one of three filtration lines was in use, still with capacity reserve. Therefore a throughput regime was fixed to minimize the excess wastewater for the other disposal strategies and to investigate the plant characteristics near its capacity limit. The throughput regime is shown in Figure 7. From 2007 to June 2008 the throughput was limited to10 -11 m³/d weekdays and 13 m³/d weekends because of the higher incoming wastewater at the weekend. Under those conditions the plant was still overloaded by 20 – 50% compared to the maximum design layout. The three days duration with higher throughput at weekends were identified as the limit for the plant capacity as seen by slightly increasing effluent values at Mondays. Limitation of dissolved oxygen concentration during summer time made it necessary to adapt the throughput regime and TS concentration. In summer 2008 the throughput was minimized to 9 -10 m³/d at weekdays and 12 m³/d at weekends that was the maximum for a sufficient oxygen supply. The throughput below 9 m³/d are mainly related to temporal inflow stops because of membrane cleanings, screening problems (see Section 2.8.2) and low level in the buffer tank. The decreased throughput in November and December 2007 was related to a foaming event with sludge loss in the system which reduced the cleaning capacity of the plant.

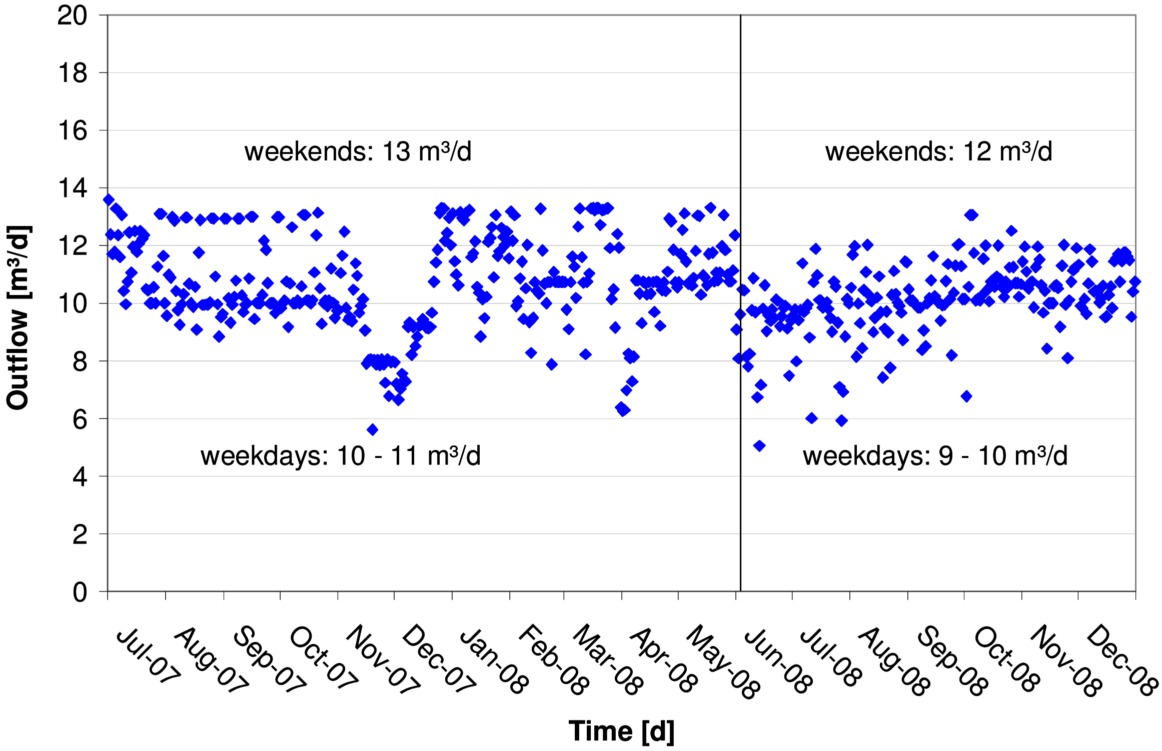



**Figure 7: Daily outflow of the MBR plant**

### 2.3.4 Operation conditions and evolution of total solid

Because of the high hydraulic and nutrients loadings (see Section 2.3.1) total solids (TS) concentration in the biological system was set as high as possible to guarantee good effluent values for a maximum plant throughput. The limitation of the TS concentration was the air supply to reach a minimum dissolved oxygen concentration of 2 mg/L, thus ensuring full nitrification, and to guarantee a good filtration performance. The experience in 2007 and 2008 showed that with sludge temperatures in summer above 20 °C the TS concentration was limited to 11- 12 g/L because of the DO concentration (period 1). In winter the TS concentration could be increased up to 16 g/L in order to warrant full nitrification and denitrification due to lower kinetic rates. On the basis of this experience a TS plant operation mode was implemented from summer 2008 onwards as shown in Table 4.

**Table 4: Adjustment of the target TS concentration in AE – AX depending on the sludge temperature**

| T in °C | < 12.5 | 12.5 – 15 | 15 – 17.5 | 17.5 - 20 | 20 – 22.5 | 22.5 – 25 | 25 – 27.5 |
|---|---|---|---|---|---|---|---|
| TS* in g/L | 15 | 14.5 | 14 | 13.5 | 13 | 12.5 | 12 |

*in Berlin, TS (g/L) ~ MLSS (g/L) + 1 g/L

Additional measures were taken:

- High temperature > 20°C: installation of two aerators above each other in AE (in order to achieve 2 mg DO/L)
- Low temperature < 12 °C: backconversion AX0 in AE2, to warrant enough aerobic volume for full nitrification

Because of installing two additional aerators in the aerobic zone in summer 2008 the TS concentration could be increased to more than 12 g/L during the summer period. The evolution of the total solids concentration (TS) and the solid retention time (SRT) is shown in Figure 8.

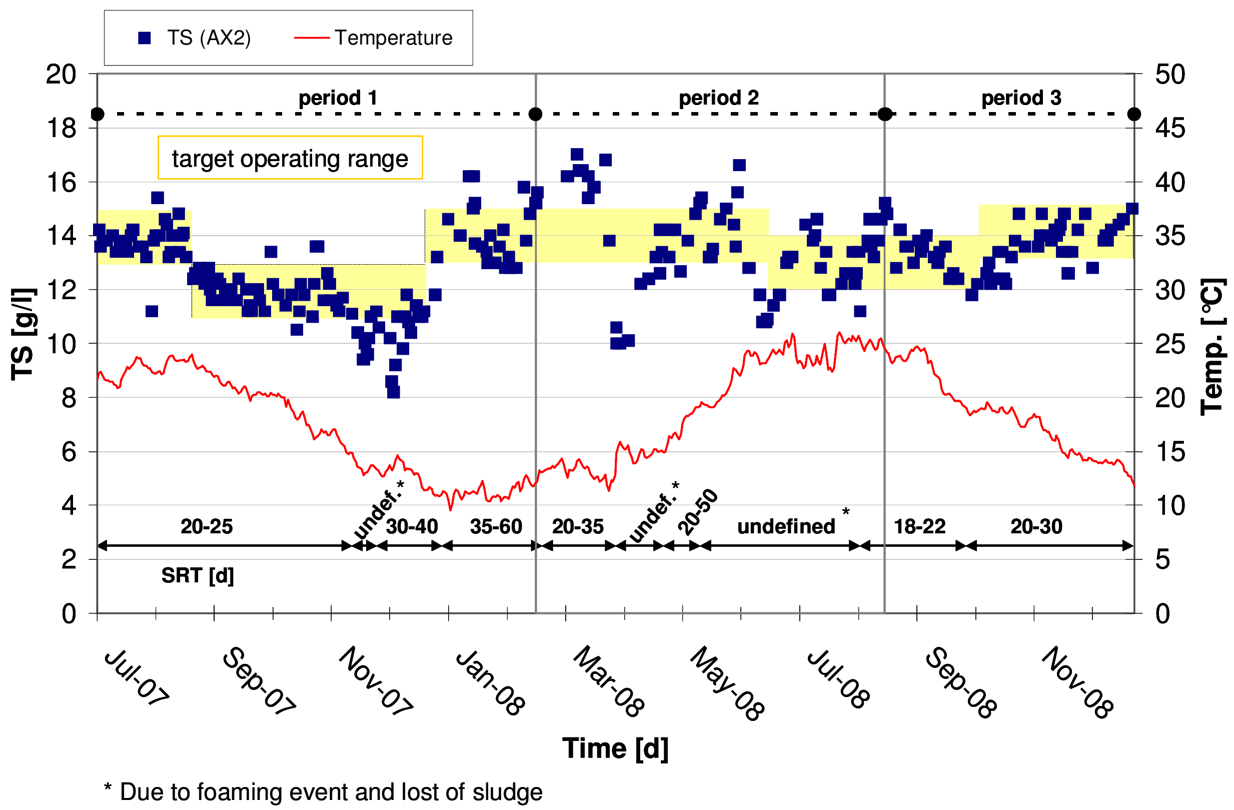



**Figure 8: TS concentration, sludge temperature and SRT over time**

In the first period (1.7.2007-17.02.2008) the TS concentration was within the range of the target operating range of 13 – 15 g/L and in late summer within 11 – 13 g/L because of the DO restrictions. End November 2007 there was a foam event with an unscheduled sludge loss and a TS fall down to 8 g/L. To increase the TS concentration the SRT was set up to 60 days. Also in the second period (18.2. – 20.8.2008) there were several foam events with sludge withdrawals resulting in immediate dropping of the TS concentration and undefined SRT. In period 3 (21.8 – 31.12.2008) no foaming occurred likely due to the ferric precipitant. For more information regarding the foaming events see Section 2.6.

TS control was implemented end of August 2008 to run the plant on a target TS concentration with slow varying SRT (see Section 2.5). The SRT in the 3$^{rd}$ period was between 18 and 30 days.

Volatile suspended solids (VSS) were stable between 70% and 75% of TS and the mean value was 9.7 g/L during period 3.

The sludge production rate of the plant was before the process optimization and therefore at high phosphorus and nitrogen effluent concentrations in the range of 0.23 kg VSS/kg $COD_{eli}$. Afterwards the sludge production rate was in the range of 0.41 kg VSS/kg $COD_{eli}$. With similar SRT as before: 22 to 27 days.

## 2.4 Average removal performance

This section presents the results of the weekly analysis performed on 24h-sample of influent and treated water by the accredited laboratory of Berliner Wasserbetriebe.

### 2.4.1 Average nutrients removal

The average value (24h-samples) of the substrate and nutrients concentration are calculated for the three representative periods and given in Table 5. The influent COD-concentration varied between 600 and 1700 mg/L, for the third period the concentration did not exceed 1000 mg/L. For all periods the average effluent concentrations of COD were below 45 mg/L and removal rate was about 96% but there were some samples exceeding 50 mg/L what is the guideline of the Water Authority. The Berlin wastewater is rich in natural refractory humic substances (here about 4% of COD in wastewater) so that no further COD reduction can be expected given the high influent concentration values.

During stabilized conditions (periods 2 and 3), nitrogen removal was very high with in average 93 - 95% and average effluent values for nitrate of 3.0 and 5.2 mg/L showed the high potential of the process, The total nitrogen (ammonia) concentrations of the influent were 125 mg/L (100) in average and 145 mg/L (110) in maximum. The refractory nitrogen fraction amounted to about 2 - 3% of the mass present in wastewater.

EBPR shows very satisfactory results after stabilization of the process in period 2 and an average effluent total phosphorus concentration of 0.23 mg/L could be reached without chemicals (99% elimination, about 0.5 to 1% of entering influent load being assimilated as refractory fraction). After start of low precipitation the average total phosphorous concentration in the effluent was 0.1 mg/L and 0.03 mg/L for the orthophosphate. For details regarding the precipitation and the refractory fraction see 3.4.

The suspended solids in the effluent with values of 1.8, 1.9 and 1.4 mg/L seem very high for membrane filtration. The suspended solids analysis was not made conform with DIN by using a 0.45 µm-Filter. Recent analysis showed great differences between this method and the DIN method where values far below 1.0 mg/L were measured. Also the bacterial analysis showed a very good reduction of E.coli and even phages which made the high data implausible. Moreover a contamination of the filtrate tank can not be surely excluded.

The pH value was in all periods in average between 7.9 and 8.1 with no drift after start of precipitation.

It has to be mentioned that during one check of the Water Authority the effluent from the filtrate tank was very dirty which could be attributed to a contamination because of a former foam event. The foam was getting through the cable hole on the top of the container and was not detected so far. Therefore it is highly recommended to separate the filtrate collection and discharge part from the sludge and waste water part spatial for future designs.

**Table 5: Average influent and effluent concentrations of the MBR plant for the three representative periods (24h samples)**

| Parameter | COD | SS | TN | $NH_4$-N | orgN | $NO_3$-N | TP | o-$PO_4$-P |
|---|---|---|---|---|---|---|---|---|
| Units | (mg/L) | (mg/L) | (mg/L) | (mg/L) | (mg/L) | (mg/L) | (mg/L) | (mg/L) |
| | Period 1 (1.July 2007 - 17.February 2008) 26 samples, (AE/AX = 51/49) | | | | | | | |
| **Influent** | 1031 | 276 | 123 | 98 | 24.5 | **-** | 20.5 | 12.7 |
| **(min - max)** | (618 - 1630) | (110 - 996) | (104 - 142) | (73 - 112) | (13.4 - 43.8) | - | (14.2-36.4) | (8.9 - 14.6) |
| **Effluent** | 45 | 1.8 | 20.9 | 0.05 | 2.5 | 18.4 | 6.2 | 5.9 |
| **(min - max)** | (38 - 64) | (0.6 - 4.3) | (9.6 – 35.6) | (0.01-0.39) | (0.7 - 4.6) | (7.6 - 32.7) | (0.12-15.1) | (0.02 - 14.8) |
| (Elimination) | (96%) | (99%) | (83%) | (100%) | | | (69%) | |
| | Period 2 (17.February - 20.August 2008) 22 samples, (AE/AX = 75/25) | | | | | | | |
| **Influent** | 994 | 279 | 118 | 94 | 23.7 | | 19.9 | 12.1 |
| **(min - max)** | (640 - 1730) | (106 - 790) | (85 - 132) | (67 - 108) | (13.9 - 40.9) | | (12.6-28.8) | (7.7 - 14.0) |
| **Effluent** | 43 | 1.9 | 8.5 | 0.05 | 2.9 | 5.2 | 0.23* | 0.12* |
| **(min - max)** | (31 - 58) | (0.2 - 5.6) | (2.9 – 17.0) | (0.01-0.39) | (0.5 - 6.7) | (0.1 - 14.5) | (0.11-0.51)* | (0.03-0.34)* |
| (Elimination) | (96%) | (99%) | (93%) | (100%) | | | (99%)* | |
| | Period 3 (21.August - 31.December 2008) 14 samples, (AE/AX = 75/25 + 4gFe/m³) | | | | | | | |
| **Influent** | 848 | 190 | 124 | 99 | 24.5 | | 16.8 | 12.2 |
| **(min - max)** | (698 - 992) | (120 - 280) | (111 - 145) | (92 - 109) | (19.1 - 48.9) | | (14.8-19.1) | (10.6 - 13.4) |
| **Effluent** | 44 | 1.4 | 5.6 | 0.07 | 2.7 | 3.0 | 0.10 | 0.03 |
| **(min - max)** | (36 - 55) | (0.1 - 8.0) | (0.2 – 15.5) | (0.02-0.13) | (1.6 - 4.1) | (0.1 - 12.7) | (0.06-0.16) | (0.01 - 0.07) |
| (Elimination) | (95%) | (99%) | (95%) | (100%) | | | (99%) | |

* representative period for EBPR performance from 14. April to 20. August 2008

The mass nitrogen loads were nearly constant for the 3 periods in the range of 0.012 and 0.014 kgN/(kgTS·d) in average.

Figure 9 shows the effluent values of ortho phosphate and total phosphate over the three periods. In period 1 the EBPR process was very unstable. This was mainly related to the high nitrate effluent concentrations during this period with recycling of a large amount of nitrate into the anaerobic zone which is disturbing the EBPR process (Gnirss *et al.* 2007). After increase of the anoxic volume and a short adaptation in period 2 the EBPR process was stabilized and the phosphate effluent values were at a low range even without carbon and precipitant addition.

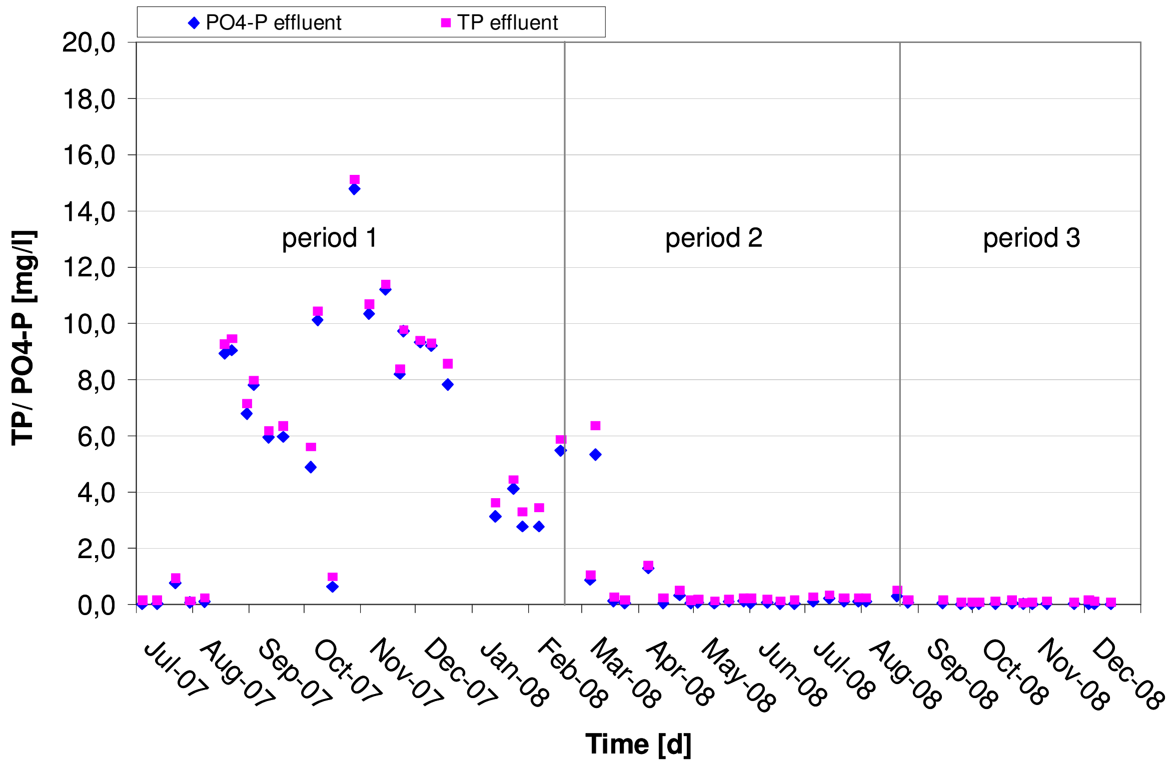



**Figure 9 Total phosphorus and oPO4-P effluent values (24h samples)**

The nitrate effluent concentrations in period 1 were very unsteady with some values above 30 mgN/L (during foaming event). However, comparing the total nitrogen elimination rate with the average elimination rate of the wastewater treatment plants in Berlin equipped with pre-denitrification (Figure 10, yellow bar) which were between 82 and 86% in 2008, the efficiency in Margaretenhöhe is within the same range. After addition of another anoxic zone (Period 2) the inorganic nitrogen dropped down to clearly below 1 mgN/L with elimination rates with more than 98%. On the other hand there are still single effluent concentrations for inorganic nitrogen up to 15 mgN/L which do exceed the target value of 10 mgN/L. Further actions to reduce these peaks are ongoing especially with a better acetic acid dosage control (see 2.8.6.5). It was observed that there was no significant increase of phosphorous in the last anoxic zone and the entire nutrients removal process was reliable with the present ratio between anoxic an aerobic volume in the usual range of temperature (12 – 25℃) and the TS control strategy described in Table 4.

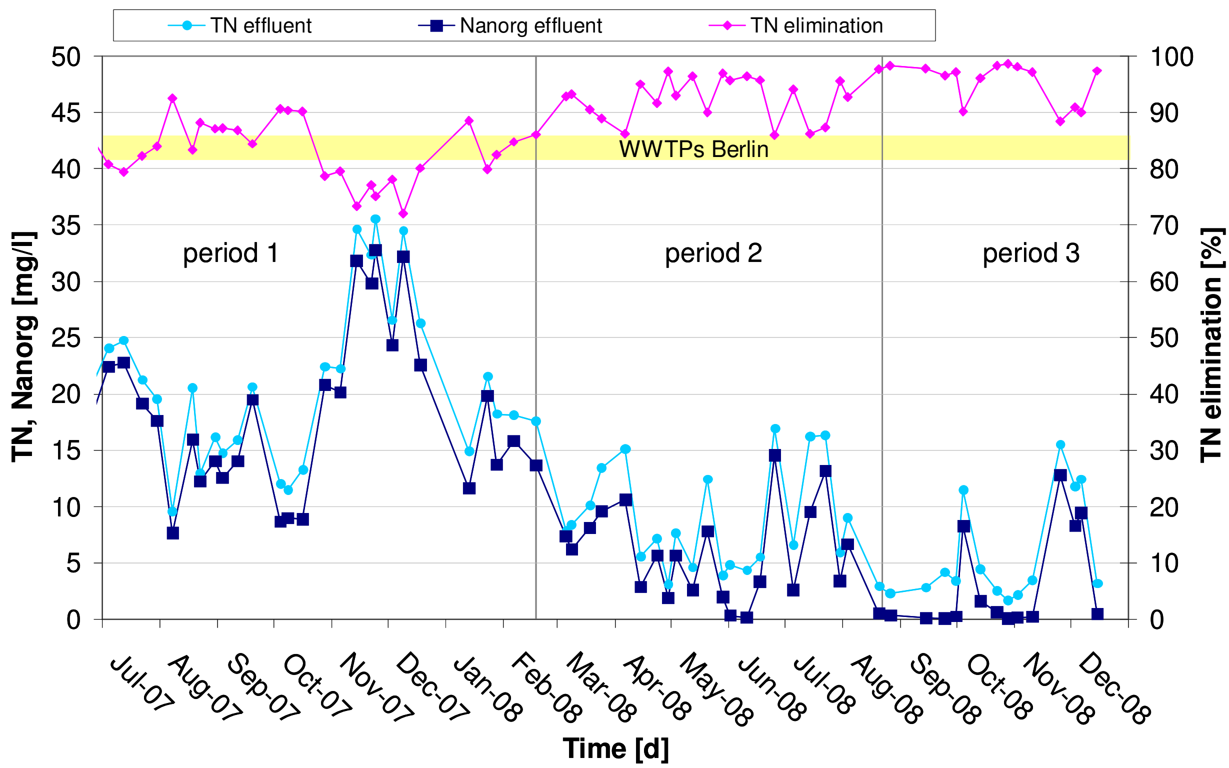



**Figure 10 Total nitrogen and NO3-N effluent values and TN-elimination rate (24h samples)**

### 2.4.2 Metals and trace organics

The metal concentrations were measured 5 times in 2008 only in the effluent in accordance to the protocol of the water authority. The measurements in 2007 were already finished by starting period 1 that no data are available for this period. The effluent concentrations of all metals were always far below the monitoring values (see Table 6). The increase of ferric effluent concentration by factor 3 in period 3 is related to the ferric precipitation.

**Table 6: Average metal effluent concentrations of the MBR plant (24h samples)**

| **Parameter** | **Cd** | **Cr** | **Cu** | **Fe** | **Hg** | **Ni** | **Pb** |
|---|---|---|---|---|---|---|---|
| Unit | µg/L | µg/L | µg/L | µg/L | µg/L | µg/L | µg/L |
| | Period 2 (17.Feb - 20.Aug 2008) 3 samples, (AE/AX = 75/25) | | | | | | |
| **Effluent** | < 1 | < 5 | < 11 | 53 | < 0.2 | < 10 | < 15 |
| **(min – max)** | < 1 | < 5 | < 10 -13 | 38 - 71 | < 0.2 | < 10 | < 15 |
| | Period 3 (21.Aug - 31.Dec 2008) 2 samples, (AE/AX = 75/25 + 4gFe/m³) | | | | | | |
| **Effluent** | < 1 | < 5 | < 10 | 145 | < 0.2 | < 10 | < 15 |
| **(min - max)** | < 1 | < 5 | < 10 | 130 - 160 | < 0.2 | < 10 | < 15 |
| **Detection limit** | 1 | 5 | 10 | - | 0.2 | 10 | 15 |
| **Target value** | 1.0 | 30 | 50 | - | 0.8 | 30 | 30 |

* no measurements in Period 1

The concentrations of AOX are shown in Table 7. Influent AOX concentrations ranged from 40 – 170 µg/L with an average of 92 µg/L and were reduced to 17 – 68 µg/L which was below the target level. The average effluent concentration was 37 µg/L. There in one high effluent value of 210 µg/L that was measured directly after a membrane cleaning with chlorine. It is assumed that there was no sufficient flushing of the membrane reactor before recommissioning. The membrane cleaning protocol was adapted (see Appendix E). Many studies already assumed that high AOX concentrations are produced in households with cleaning detergents.

**Table 7: Average influent and effluent AOX concentrations of the MBR plant (24h samples)**

| Parameter | AOX |
|---:|:---:|
| Unit | µg/L |
| **Number of samples** | 32 / 34 (influent / effluent) |
| **Influent (min - max)** | 92 (40 – 170) |
| **Effluent (min - max)** | 37 (17 – 68 / 210*) |
| **Monitoring value** | 80 |

* Value measured after membrane cleaning with chlorine, not taken into account for average calculation

### 2.4.3 Disinfection results

Every month two grab samples were analyzed for E.coli, Enterococcus and Coliphage according to DIN EN ISO 9308-3 (E.Coli, MPN method), DIN ISO 7899-1 (Enterococcus, MPN method) and in-house method (Coliphage, related to Federal Environment Agency method). The results over time are presented in Figure 11. During the trials no disinfection of the membrane (no CIP cleaning with chlorine!) was carried out. The samples with the modules 1 – 3 (Martin Systems PES 900 C high flux UF membrane, 37 nm) showed that bacteria and viruses were eliminated down to the detection limit. Therefore, the imperative values and even the guide values of the old EU-bathing water directive for the aforementioned bacteriological parameters could be matched over the trials period. Coliphage - as a surrogate organism for enterovirus - were completely eliminated. As these organisms are generally very well adsorbed by solids, their almost complete elimination is expected due to the high retention of the solids during membrane filtration. The two high values of E.coli may be due to recontamination after the membrane, but no clear statement can be given, as the modules went out of operation in April, and replaced by another technology of filtration system.

The samples with the modules 4 and 5 (A3 MX-020 MF membrane, 200 nm) cleaned every month with chlorine and hydrogen peroxide (micro filtration, 200 nm) showed for E.coli and Enterococcus values in the range of $10^1$ and $10^2$ what is still within the imperative and guide values of the old EU-bathing water directive but higher than with the ultra filtration system. Also the values for Coliphage are much higher up to $10^3$. The high values for bacteria in Jun 2007 and Feb 2008 is most likely caused by recontamination after the membrane. A subsequent cleaning of the discharge system with chlorine showed appropriate values afterwards.

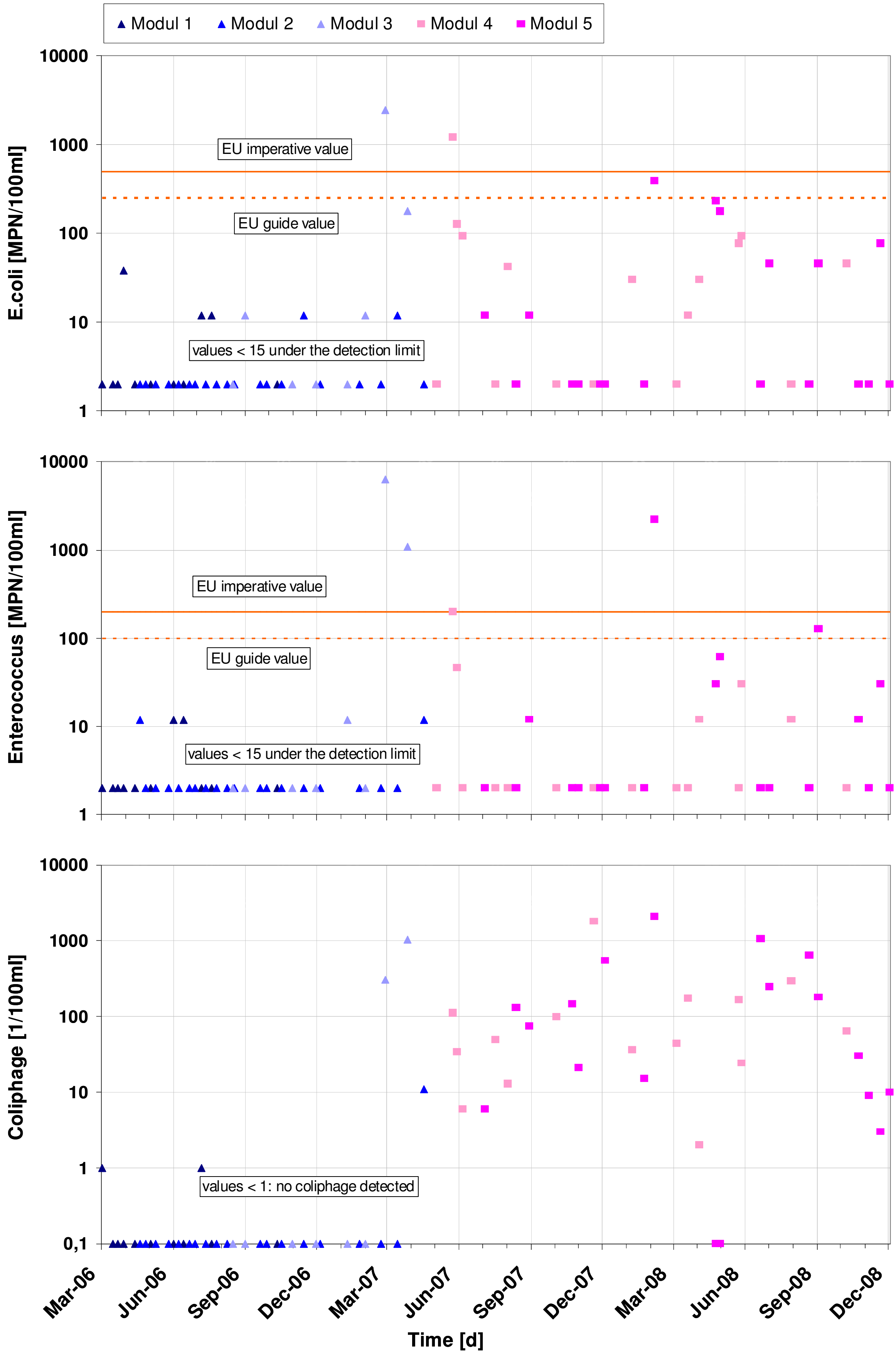



**Figure 11: Effluent concentrations of E.coli, Enterococcus and Coliphage**

## 2.5 TS measurement and control

The TS concentration is one of the most important parameter for plant adjustment and plant operation check. Therefore a reliable TS measurement is a basic requirement especially for the technician who is in charge of the plant operation. Furthermore an automation regarding the excess sludge withdrawal is helpful to hold the TS concentration within a certain range reducing the personnel time on site.

To get a reliable and fast TS concentration on site a microwave quick test was developed with the following sequence of action:

**Table 8: Protocol of the microwave quick test for TS measurement**

| | |
|---|---|
| 1 | Dry of a 500 ml beaker glass at max. heat of the microwave (600 W) for 5 minutes and following cooling phase in an exsiccator. Weighing of the beaker. |
| 2 | Filling 50 ml of sludge in the beaker |
| 3 | Preheating the sludge in the microwave at unfreezing level for 20 minutes to prevent heavy foaming |
| 4 | Heat the sludge at maximum level for 10 minutes |
| 5 | Cooling down of the beaker in an exsiccator |
| 6 | Weighing of the beaker and TS calculation |

The reliability of this field test is absolutely satisfactory for plant operation and can be easily and quickly done by technicians on site. The TS results are available within 1 hour. Comparison with DIN tests showed a good correlation (see also Figure 13, Section 2.6 )

For the TS online measurement in the first periods a sensor form Endress & Hauser was installed (Turbimax W CUS 41). The experience with this sensor was not satisfying because of many differences not only between the absolute online and quick test data but also because of different trends. Therefore it was decided to install a new TS system in spring 2009 (WTW Visolid 700 IQ) for measuring the TS concentration in the anoxic zone 2 (AX2). The experience with this system was satisfactory for plant operation but with some restrictions to keep in mind:

- The stirrer frequency in AX2 has an influence to the absolute values of the sensor, so that a new calibration is necessary when changing the stirrer frequency.
- The measurement is more reliable if the sensor is installed with a 90° break and parallel to the water level, with the head contrary to the stream.
- From time to time there is a drift of the online values according to the quick test results which is assumed to be related to changing of the sludge properties – therefore from time to time (if the drift is too strong) a new calibration is necessary

With a reliable online TS measurement installed it was decided to install a TS automation which is steering the excess sludge pump in accordance to the online TS values:

The control is based on a linear slope which is relating the pause time of the excess sludge pump to the online TS concentration value. The linear slope can be defined with a set point and a gradient which has to be chosen based on the plant operation specifications and adjusted with the experience of the operator. For TS control 8 hours mean values are calculated in order to avoid too short excess sludge pump reactions by online measurement fluctuations. To avoid unfavorable pump cycle because of implausible TS online values a minimum and maximum SRT can be chosen. A closed loop control for total automatic excess sludge was not installed because of the expected difficulties in PID controller settings.

Furthermore a TS concentration alarm will be installed to give an SMS to the responsible operator if the TS online value exceeds a defined range. In this case the operator can check the plant for further actions. This alarm is also very helpful for detecting foaming events as soon as possible (see Section 2.6).

Figure 12 shows the operation of the TS control in a 4 month period: The accordance of the online measurement values and the quick test results are satisfactory for daily plant operation. In case of high variations, a calibration was necessary to adjust the values (pink triangles). The sludge retention time is following reversely the TS concentration to hold a given TS range which demonstrates a satisfactory control. However, from time to time an adjustment by the operator based on his experience is necessary. Strong decreasing online values (end November 2008) indicate changing of sludge parameters and often resulted from foaming events. These measurement characteristics are also very helpful for early detection of foaming events.

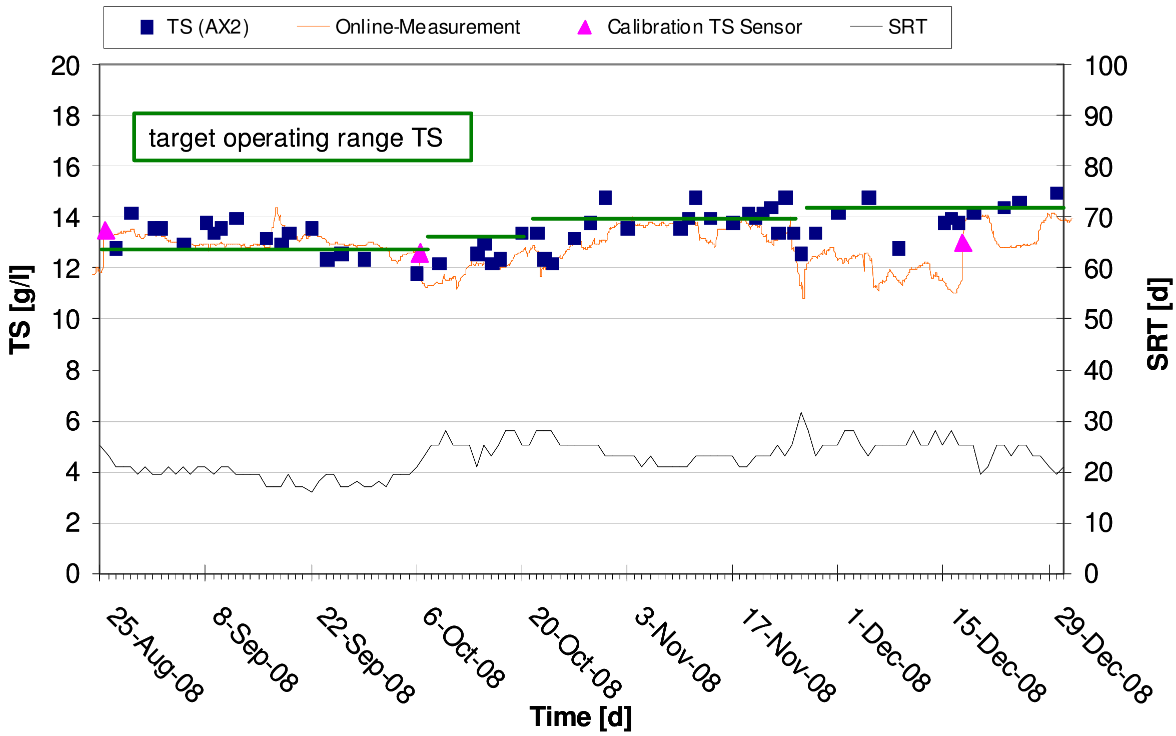



**Figure 12: TS control: SRT adjustment relating to the TS online measurement**

## 2.6 Foam

Foaming of activated sludge is a known and widely investigated phenomenon in conventional activated sludge systems (CAS), sludge treatment and membrane bioreactors (MBR) for wastewater treatment (Stratton *et al.* 1998; Schade and Lemmer 2002). It represents a major risk during operation for many types of processes due to various reasons:

- Foaming in the aerated reactors does affect the biological process leading to high sludge loadings, because a major fraction of the biomass will be immobilized within the foam. Overflow of this high concentrated foam leads to a loss of biomass.
- Reaching the anaerobic or anoxic reactors where engines ensure mixing, the rising foam might result in electrical or mechanical failure, respectively to an early wear out of equipment.
- The overflow of foam is also a local environmental burden and leads to additional man hours and costs for clean up.

There are numerous reasons for foaming and almost as many approaches to act against it. A change of the microbiological biocenosis, e.g. an increase of filamentous bacteria, is often identified to be the reason for foaming. Changing ambient temperatures is also linked to the formation of foam. Another possible reason is an influent contamination with toxic substances or shock loads of tensides.

Foaming within the demonstration plant was the major risk in operation and led to increased man hours on site, damage of equipment and occasional collapse of biological performance thus violation of discharge limits. Although the above mentioned reasons for foaming were investigated, the cause could not be definitely identified for the demonstration plant. Foaming due to the presence of filamentous bacteria is unlikely to be the only reason, as periods of foaming started within few hours. Microscopic investigations showed the presence of filamentous bacteria, nevertheless two serial studies showed that the amount during stable operation were higher than during a foaming event, even when monitoring the foam fraction.

Due to the fact that the later discussed prototype MBR, see Chapter 5, fed with the same influent showed foaming events too, it was concluded that shock loads or toxic substances in the inflow were most likely to be the origin of foaming. Tensides were extensively monitored but as the residual time in the buffer tank was less than one day, and often foaming events appeared on the weekends, no samples of the suspected contaminated wastewater could be collected and analyzed. Anyway, the chemical analysis to identify the substance of question would be time-consuming and expensive.

To be able to operate the plant despite foam occurrence, different actions addressing foam events were tested:

- The dosage of anti-foam products suitable for wastewater treatment plants were tested with positive short term results, but were not implemented in the long term, as the effect decreased within 24 – 48 h and stronger foaming appeared afterwards. Additionally, the dosage of anti-foam products showed some negative side effects such as decreased oxygen mass transfer and increased COD load. It is also assumed that anti-foam detergents are suitable food sources for filamentous bacteria. Polymers and powdered activated carbon (PAC) were tried also, but without any significant effect.

- Adaptation of aeration rate in the aerobic reactor was tested in order to ensure sufficient turbulent conditions for mixing. It was shown that according to the sludge properties high aeration rates could help reducing the formation of foam due to mixing. For long periods of operation, the aeration was therefore notably higher than required for the biological process. This action showed good results, but had to be adapted manually thus leading to increased man hours on site and required operational experiences.
- Recently the spraying of activated sludge on top of the forming foam showed promising results. This approach is another way to ensure mixing and helps keeping the biomass suspended. The loss of biomass showed to be the main problem, as the biological process took a long time to recover and lower throughputs were required leading to increased costs for trucking. The spraying should be placed in the aerobic reactor to reduce oxygen introduction to the anaerobic or anoxic reactors.

Foam events led to 4 major unintended sludge losses in 18 months. Figure 13 shows the evolution of total solids in the plant and the drop of biomass concentration in respect to foam events. In period 2 the unintended sludge withdrawals due to foaming are indicated and show the immense impact on the biomass concentration. A loss of biomass up to 40 % within few hours led consequently to a corresponding increase of the sludge load and a following collapse or disturbance of the biology.

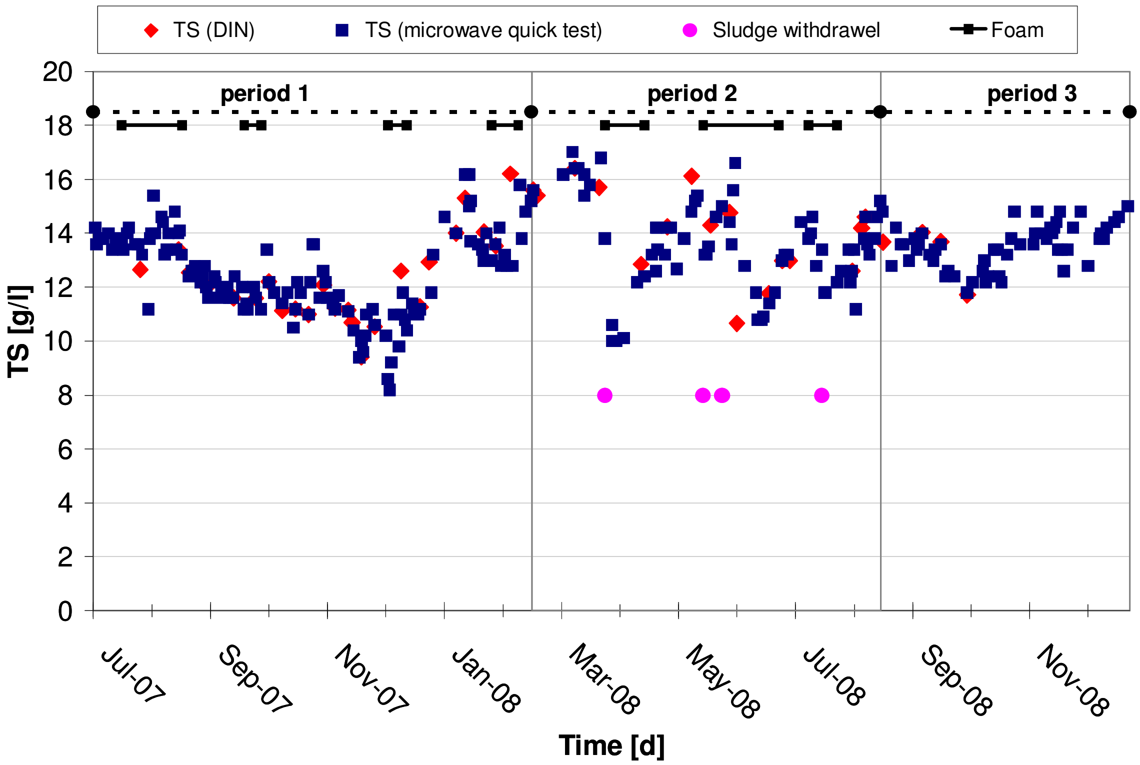



**Figure 13: Foam events and accidental sludge loss within the period of trials**

Besides these actions addressing directly the foam production, it was of major interest to install an alarm indicating a foam event just in time, thus giving the possibility to act immediately and avoid loss of sludge. As mentioned above, often foam events occurred on weekends thus no work force was present on site. During stable operation it was not required to be on site every day, so due to the quick formation of foam within few hours it was necessary to install an automated alarm informing the plant operators. Biomass loss should be prevented under any circumstances.

A liquid level probe was installed in the aerobic reactor, the origin source of foam production, approximately 40 cm above the surface. Rising foam induced the trigger for the alarm and the operators were informed by the installed SMS system. After a period of sensitivity optimization the probe showed good results. Nevertheless signal induced by sludge sprays due to heavy aeration led to false alarm. Therefore a new alarm regime was implemented, combining the measures of the total solid probe with the signal of the level probe. During foam events the loss of the biomass is recorded in the downstream installed TS probe. A rapid change in TS concentration of i.e. 1 g/L is also an indicator for foam events. Combining the two signals decreases the possibility of false alarm and indicates heavy foaming. An automated immediate response will be implemented reducing throughput and aeration. This will give the operators some time to reach the plant and act manually.

## 2.7 Microscopy (activated sludge)

Within the project a total of 6 investigations were carried out during and after foaming events for activated sludge characterization:

In 2006 only few filamentary bacteria were found such as *Microthrix parvicella* and bacteria of the type 0092. There was only cross linking within the flakes and not between them. In 2007 and 2008 also bacteria of the type *Nocardia specc* became apparent which causes cross lining between flakes. *Microthrix parvicella* did also cross link the flakes, whereas the type 0092 was only within the flakes. Filamentary bacteria were found in all samples from 2007 onwards.

The occurrence of these filamentary bacteria was changing a little but there was no clear correlation between foaming events and the number of bacteria. Also during periods with no foam there was heavy cross linking. It is assumed that a high TS concentration and a high air flow is advantageous for foam preventing.

## 2.8 Operation of mechanical and electrical system and trouble shooting

### 2.8.1 Buffer tank

The buffer tank's mode of operation – flattening both hydraulic and nitrogen loads - was very satisfying in order to reach even influent flow of the MBR plant. A reliable and steady treatment process could be achieved by storing the waste water during the day peaks and treatment during the night hours with low or no inflows. The membrane filtration and the blowers could also be operated more steadily and with lower throughputs, as desired in the plant design.

Depending also on the inflow of illegal rain water (see section 2.3.2) or maintenance work at the household buffer tanks (twice in 5 years, including flushing the deposit to the plant) it is recommended to completely clean the buffer tank at least once a month. Otherwise the deposit will be covering the inflow pumps which can cause pumping and screening problems resulting in inconstant plant feeding.

The excess sludge tank was not equipped with a level sensor in the design which was installed afterwards in order to automate the waste water shifting from the buffer tank. Not only for this application a level control of the excess sludge tank is recommended, e.g. to schedule the emptying of the tank without always looking into it from time to time.

### 2.8.2 Screening

The Martin Systems drum screen achieved efficient and reliable screening performances. The screen ran successfully without much manual intervention. The automatic screen cleaning with a rotating brush worked well, in case of heavy dirt and no success in flushing the screening chamber the screen had occasionally to be removed for better cleaning (see Picture 1). Reducing the follow-up time of the brush from 60 to 10 seconds (after running of the inflow pumps) to extend the durability of the brush caused a worsening of the cleaning performance. Therefore the time was set back to 60 seconds again. In 2008 one loose screw (which was not detected for a long time) resulted in an irregular rotating of the brush and caused cleaning problems too. Therefore stops in plant feeding through high water level alarms during the 10 second pumping of the inflow pumps occurred which reduced the daily throughput of the plant.

We have to keep in mind that both the grinding pumps at the households and in the buffer tank grind the solids to pieces smaller than 7 mm. A brush change is necessary every 1 to 1.5 years according to the screening performance and should be done immediately if the screen is frequently getting dirty in series.

The manual emptying of the screenings tank was necessary every 2 weeks that means 10 minutes work and 600 L inflow to the excess sludge tank which represents the excess sludge volume of 2 days. The extension to every 4 weeks is not recommended for reliable plant operation. It also came apparent that during illegal rain water inflow much more sediments came into the plant which reduced the emptying frequency to every week.

**Picture 1: Dirty screen to be cleaned**

### 2.8.3 Mixed liquor hydraulic distribution and mixing

Except for the two recirculation pumps, the hydraulic distribution of the mixed liquor throughout the unit occurs per gravity. The overall hydraulic head between top and bottom water level is by construction only a few centimeters. This caused severe problems of the mixed liquor flow in the plant, especially when foaming occurred. Hydraulic problems were particularly observed at the following locations (see Figure 14):

• Deaeration pot **(1)**: due to the narrow diameter, the foam accumulated and prevented the flowing of the mixed liquor. The water level rose in the previous zones. Elevation of the screen drum **(2)** to the maximum possible (+ 24cm) prevents the backflow of sludge from the anaerobic zone to the screen and the following screen blockage. To guarantee a discharge of foam from the previous zone, the deaeration pot was permanent flushed with sludge by a small pump.

• The distribution channel **(3)** to the membrane reactors: due to low hydraulic height, a bad distribution occurred, resulting occasionally in a thickening of the sludge in one of the reactors when operating with 2 filtration lines. By using only one filtration line one part of the channel was thickening without turbulence of the sludge what causes in little foaming after a while. The elevation of the 3 membrane reactor inlets with small cylinders (about 6cm), as well as the permanent aeration of the channel (to avoid sedimentation), solved this trouble.

• The collection channel **(4)** from the membrane reactors: foam tended to accumulate, rising the water level in the membrane reactor and inducing a bad distribution of the fluid between the reactors. Alternative spraying of permeate could solve the problem, but reduced the plant throughput much because of the spraying water. The extension of the aeration from the distribution channel to the collection channel improved the situation, but in presence of heavy foaming in the membrane reactors the aeration could not prevent the disturbance in sludge distribution. This problem still remains in case of heavy foaming.

• The normal water level in the aerobic zone 1 was by construction so high, that the inflow from the anaerobic zone **(5)** and the membrane reactors **(6)** were below the water level. By the occurrence of foam, blockages and back pressure of the sludge distribution were observed. A lowering of the water level in the biology could enhance this problem, but would also reduce the plant capacity.

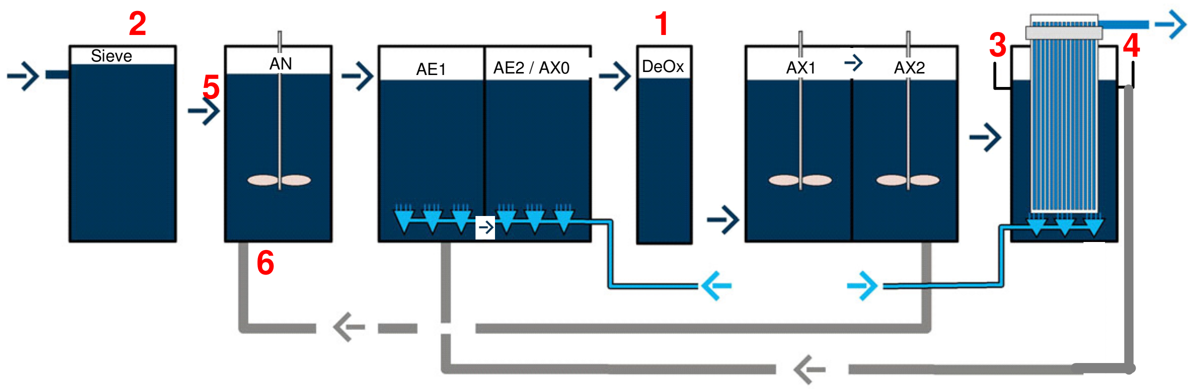



**Figure 14: Constructual hydraulic plant problems**

During foaming events the stirrer motors in the not aerated zones were covered with foam which caused many motor damages, especially in the anaerobic zone. In the year 2007 at least three motors had to be replaced. For this reason it was decided to change the stirrer system in the anaerobic zone to a mixing system with a pump. The experience with this system was very satisfactory in the anoxic zone 0 (see Chapter 3.2) which was equipped with a pump before. With these experiences it is recommended to use pumps for mixing because the appearance of foaming events can not surely be eliminated. The energy demand of the installed pumps (250 Watt) is approx. 40% higher than the demand for the stirrers (180 Watt). (Keep in mind that an energy optimization is difficult for this kind of plant size because of missing of appropriate aggregates).

### 2.8.4 Nutrients overload reduction

In order to overcome the scenario of permanent heavy overloads (see Chapter 2.3.2 and 2.3.3), the following measures were taken:

- The load was reduced by diverting the part of incoming flow which exceeded the capacity of the MBR plant. It was stored in the excess sludge tank and then trucked away approx. twice a week (from November 2006 onwards). For ease of operation a pump-automation for pumping the wastewater from the buffer tank to the excess sludge tank considering the water levels in both tanks was installed. For this reason it was necessary to install a level sensor in the excess sludge tank.
- From January 2008 onwards, part of the incoming wastewater was cleaned by a second MBR system (Busse plant, see Chapter 5). This plant was able to clean up to 3.5 to 4.5 $m^3$/d. As the plant could not treat the target amount of 5 $m^3$/d (in order to optimize the effluent concentrations) and there were still many irregular inflows (see 2.3.2), trucking away of waste water at least once per week before the weekend was necessary. In this case the buffer tank was often completely emptied in order to get the maximum buffer capacity for the weekend resulting in inflow stop to both MBR plants for some hours.
- The TS concentration in the anoxic reactors was elevated to 14-15g/L in winter to have more biomass in the system. Since the oxygen transfer is strongly influenced by the TS concentration the amount of aerators (and oxygen) were exceeded in the aerobic reactor in period 2 and 3. This way, complete nitrification was ensured in the aerobic reactor (from summer 2008 onwards).

All these actions finally helped to handle the incoming waste water from the sewer and to run the MBR plant with satisfying treatment performances despite the conditions of overloading compared with the maximum design load.

### 2.8.5 Air supply

The flow rate of the two blowers and the DO concentrations in AE1 for the 3 periods are shown in Figure 15. Except for the start of period 1 where the blower for biology was run by DO control, the blower performance had to be increased due to avoid flotation and foaming.

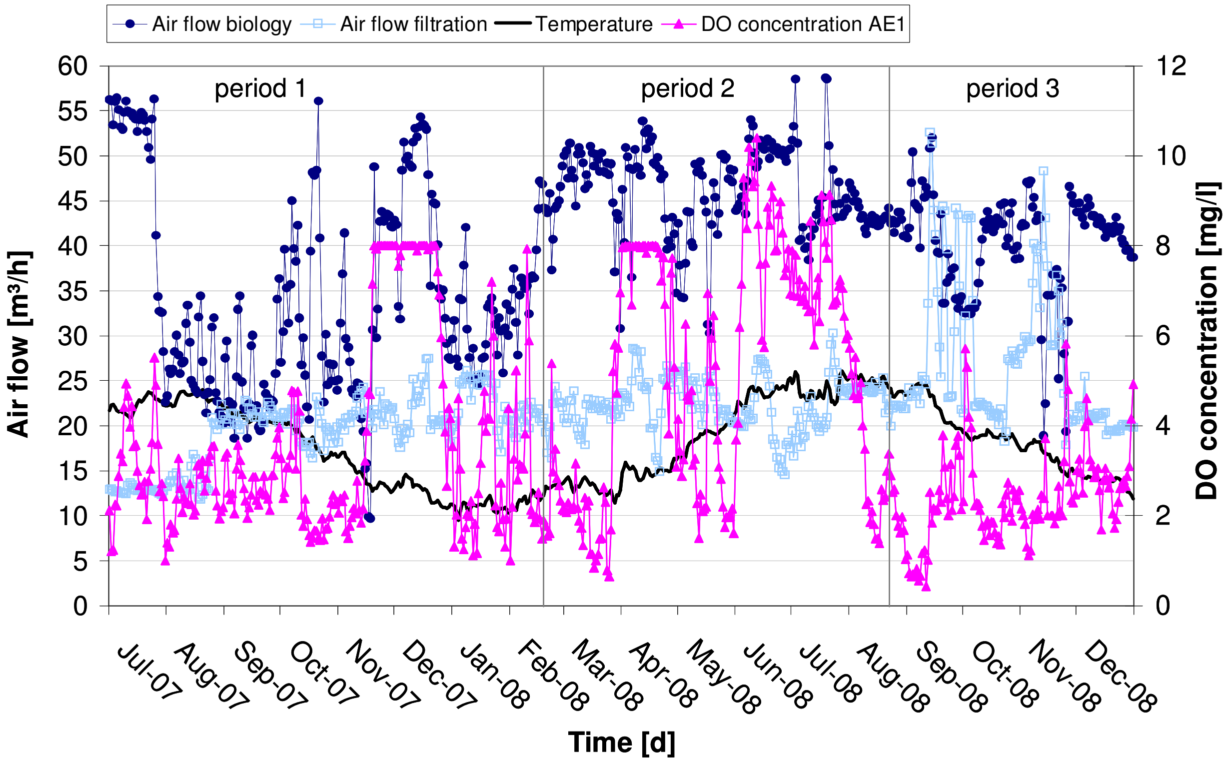



**Figure 15: Air flow of biology and filtration, sludge temperature and DO concentration over time (24h average)**

In March 2008 the DO concentration decreased below 1 mg/L because of TS concentrations up to 16 g/L and more followed by an increasing of DO concentration in April 2008 as a result of sludge loss during foaming (see Figure 13). A TS concentration of more than 15 g/L is identified as critical for oxygen transfer to the sludge in the given plant configuration.

In August and September 2008 the DO concentration of 2 mg/L could not be reached due to higher temperatures, TS concentration of around 14 g/L and low dissolved oxygen concentration due to large bubbles caused by the high blower performance. Therefore two additional aerators were installed in AE1 and supplied by the blower for the filtration (which increased in the graphic the air flow of the filtration). DO concentrations of 2 mg/L were reached with the additional low diameter bubble aeration. This experience is approving the work of (Cornel *et al.* 2003) who has detected the lower solubility of oxygen related to increasing sludge viscosities at higher TS concentrations.

In summary the biological volume of one zone was most of the time not limiting for nutrients removal (especially ammonia) though it was near the maximum limit. In summary the limitation came from the oxygen supply because of temperature, TS concentration and diameters of the air bubbles, whereas in winter (T < 12℃) the volume of the aerobic reactor was indeed limiting due to low nitrification kinetics, and two aerated zones were required.

### 2.8.6 Automation

2.8.6.1 Process control and remote monitoring

A constantly running PC is essential for the remote and alarm function as well as data acquisition. The installed industrial PC (Beckhoff) had many failures during the plant operation. Within the project period the CD device, the hard drive and the entire PC were replaced and failures occurred with data loss, remote and alarm transportation loss and much work with the reinstallation of the plant software. For that reason a cheaper desktop PC was installed at the end of 2008 which can be easily replaced if failures occur. An uninterrupted power supply unit (UPS) was installed in 2008 to minimize the stress for the PC at sudden power failure and to guaranty PC based SMS alarm function. The UPS is able to supply the PC and On-line analyzers for approx. 30 minutes with power and to do a controlled PC shut down at longer power failures. Within the project there were at least three power failures lasting some minutes and longer.

2.8.6.2 Feed water and filtration control

The control of the influent pumps and the filtrate pumps was mainly dependent on the water level in the buffer tank. A 3 step open loop control was implemented where 3 different pumping and pause periods can be preset for a specified buffer tank level. Switch off constraints (over-, under filling of the plant) are given by the water level in the reactor AX2. The installed feeding pumps with a minimal capacity of 6 $m^3/h$ made the use of frequency inverters for PID control difficult and caused irregular plant feeding. The feed pump had to be operated e.g. with 10 seconds running followed by a 3 – 4 minutes pause. Longer stops of feeding, which was caused through reaching the upper water level in AX2 and waiting to reach the lower water by running only the filtration, could be reduced down to about 15min every 6 – 8 hours by adjusting a good pause time of the pumps related to the filtration throughput (which was considered, due to the contact time in the anaerobic zone of at least 30min, to have minor impact on the EBPR significant mechanisms). The regular influent breaks caused drops of oxygen demand in the aerated reactors, and therefore impacted severely the oxygen control.

To minimize the effort of adjusting manually the feed pump pause set point by change of the filtration pump parameters, a control was installed to calculate the pause time of the feed water pumps automatically depending on the filtration pump set point. So the level in AX2 could be held as long as possible in the favored operation range. This was realized through a linear slope control with an adjustable fix point and gradient.

2.8.6.3 DO control

The problems with the DO sensors mentioned in the former report (lack of sustainable PID settings, lack of reliability of DO sensors and sludge flotation and foaming with either very low or high air flow rates during the continuous operation, did not occur in the observed periods. The set point of 2.0 mg/L DO concentration in AE1 was reliably achieved with the blower control. However, many times the blower had to be run at maximum capacity because of foam prevention (see Section 2.6).

General problems are seen in controlling the DO concentration of two different zones only with one blower and without controllable valves. For example in period 1 there were many manual adjustments of the hand valves of both air tubes (AE1 and AE2) necessary to set up the DO concentration especially in the not controlled zone AE2. From period 2 onwards, where only the zone AE1 was aerated, the DO control helped to reduce the blower throughput and the energy demand mainly during decreasing incoming nutrients loads and decreasing temperatures.

In summary DO control is recommended for future installations for energy savings because of the varying nutrients loads over time. The control should be implemented for every zone without manual settings for different zones, e.g. to realize with one blower for each zone or a controllable air supply system.

2.8.6.4 Sludge recycle control

The recycle ratio from the membrane reactor to the aerobic zone (R2) was automatically adjusted in accordance to the flow rate of the filtration pumps and was set to 400%. This ratio guaranties an acceptable TS concentration in the membrane reactor in consideration of the recycle pump power.

The sludge return and recycle ratio from the anoxic to the anaerobic reactor (R1) was fixed and only manually changeable. Because of the frequent changing of throughput during weekend and weekday this recycle ratio was also equipped with a control and fixed at 150% of the filtration pump flow rate. This ratio was identified as the best setting for the process and an outcome of computer biological modeling work and prior projects investigating the biological capacity of this process scheme. This improvement of the control strategy also reduced the manual work of the operator.

2.8.6.5 Control of acetic acid dosing

For fast stabilization of the process in case of excess of the phosphorous effluent values an automatic acid dosing was installed: If an adjustable effluent concentration is exceeded, a pump will dose acid into the anaerobic zone as long as the concentration is going below another adjustable effluent concentration. See Section 3.1 for detailed information. In 2009 the control will be extended by additionally monitoring the nitrate effluent concentration too.

2.8.6.6 Foam emergency shut down

In order to prevent big losses of biomass as happened during heavy foam events the plant will shut down automatically when foam and sludge loss is detected. The shut down will be activated if two alarms will be detected at the same time: 1) The foam detection probe will indicate foam and 2) the TS concentration probe will indicate a leaving of a preselected TS concentration range. The shut down will be send as SMS alarm. The plant has to be started manually, there is no automatic plant starting after alarm cleaning to force investigating the situation on site.

### 2.8.7 Instrumentation and online-analyzers

It was decided to equip the demonstration plant with much more instrumentation and on-line analyzers than what would be required for a commercial unit. The intention was to facilitate the evaluation but also to identify which devices would be helpful for routine operation. Most of the equipment was provided by the German company Endress + Hauser. The implementation and maintenance of these equipments were very time consuming and costly. At the time of the redaction, the following evaluation can be done on the different equipments:

- Oxygen sensors (1 per aerobic zone, about € 2,000 each): Quite unstable in the first months, then with reliable results. They finally enabled to control the aeration level through a

PID and are recommended for future installations. For good results, the probes have to hang free in the middle of the reactor, 50 cm below water level and have to be checked at least every 3 month.

- Nitrate analyzer (about € 5,000): Reliable, easy and low-cost maintenance and would enable on-line monitoring and control of a crucial parameter for the biology. However, often calibration was necessary for reliable data acquisition. Within the operation time the sensor had 2 serious damages (failure in optical system) which caused high repair cost up to > €1000 each. Recommended even for container installations with EBPR process to monitor the biology performance and enable a nitrate based acetic acid dosing control.

- Phosphate analyzer (about € 15,000): threshold value of 0.01 mgP-PO4/L and precision value of 0.05 mgP-PO4/L, but require regular maintenance (change of piping + chemicals, about € 500 per year). Recommended only for plants above 5,000 p.e. or for control of metal salt or carbon addition when strict values are required at grab-sample level.

- Sludge concentration probe (about € 5,000, low maintenance): was intended to help remote plant monitoring and excess sludge control. The signal appeared not being reliable for the old system (E&H) even with weekly calibration. Probes from other suppliers (WTW) delivered satisfactory results which enabled an automation of excess sludge withdrawal (see Section 2.5).

- Turbidity probe (about € 5,000, low maintenance): was planned in permeate for monitoring of membrane integrity. It was however poorly mounted by Martin Systems (not enough free space around the sensor) and the calibration of real absolute value was not possible. It was however monitored that the relative value reacted quickly when the water was slightly turbid. It is not recommended for commercial units, unless strict requirements of disinfection are specified (water reuse, bathing water guidelines). Alternatively, microbiological measurements at start-up and at regular interval may also provide evidence of the system integrity. A simple cartridge filter with pressure sensor (for hollow fiber systems, can be installed on backwash circuit as supplementary protection) may be also a good indicator of system integrity.

- Redox probe measured in anoxic zones (about € 2,000, low maintenance): Not recommended at this stage as the signal drifts much, rendering the interpretation or utilization difficult.

- pH probe (about € 2,000, low maintenance): Not recommended for hard water, as the pH appeared to be stable without requirement of pH control. Weekly manual measurement may be sufficient.

- Electromagnetic air flow meters (about € 6,000 each, no maintenance): were built on the biology and membrane aeration lines. They were reliable and useful for the evaluation but may not be required for commercial applications, although the information is advantageous for diagnosis and trouble-shooting. Visual air flow meters may be sufficient in commercial applications.

- Electromagnetic sludge flow meters (about € 6,000 each, no maintenance): were built on each sludge recirculation loop. Would be always recommended for setting and/or control of the sludge recirculation rates (crucial parameters for the biological performances)

- Foam detection sensor (about € 200, no maintenance): was installed in aerobic zone to detect foaming events, very important for early and reliable detection of foaming. Foaming events can strongly disturb the plant operation and quickly reduce the plant capacity by loss of biomass.

- Water level sensor (about € 500, no maintenance): were installed in buffer tank, excess sludge tank and anoxic zone. Crucial for plant automation (especially overflow detection in buffer tank and excess sludge tank) and remote control plant operation.

### 2.8.8 Air conditioning

Air conditioning of the operation room inside the container was necessary to avoid equipment breakdowns because of heat. During running of the air conditioning especially the blowers did not fail as it happened the month before at summertime where temperatures of more than 40-50 °C were measured inside the container. On the other hand the electrical power consumption in summer increased significantly.

### 2.8.9 CCTV

After installing a closed circuit television (CCTV) no more burglaries took place. In 2006 there were 3 burglaries at the container with e.g. PC robberies. The CCTV sends alarms to the WWTP Schönerlinde when movements in the observed area take place. The appearance of false alarms because of snow, rain, animals or falling leaves keeps within a limit.

## 2.9 Veolink®

The filtration monitoring software Veolink® was installed in summer 2008 to assist the monitoring of the filtration performance. With adjusting the relevant parameters it was possible to detect filtration disturbances much earlier (just in time) than it would be possible with a plain pressure limit value. By comparing the pressure gradient within the filtration and relaxation phase with values for normal operation, the filtration performance can be observed immediately, what is helpful especially during and after membrane cleanings. Even changing in sludge properties or biological process is detectable by interpreting the data.

Unfortunately, the described computer breakdown during evaluation phase prevented collecting of long term experience. Data monitored by Veolink® are presented in Figure 16 and alarm settings for filtration performance observation in Figure 17.

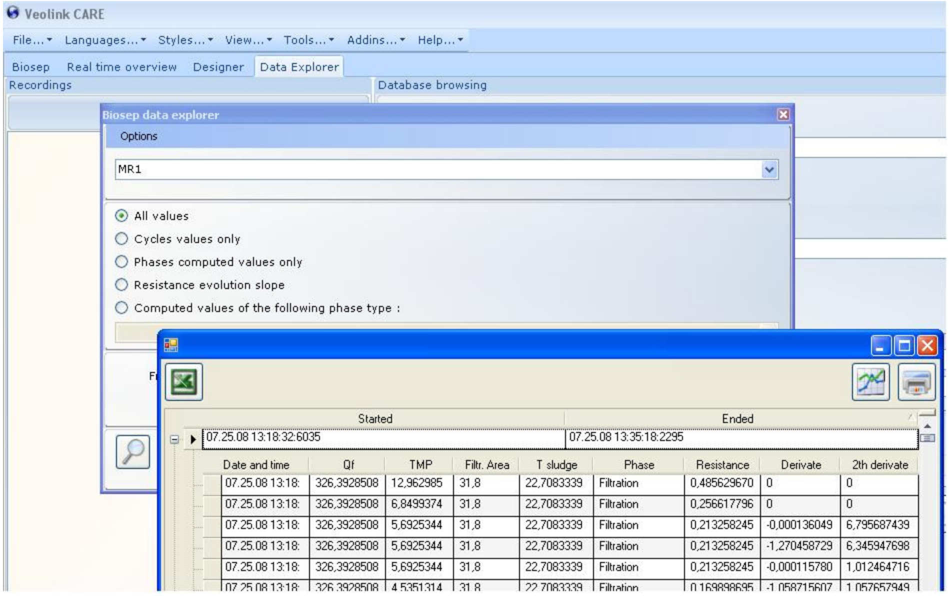



**Figure 16: Monitored filtration data from the software Veolink®**

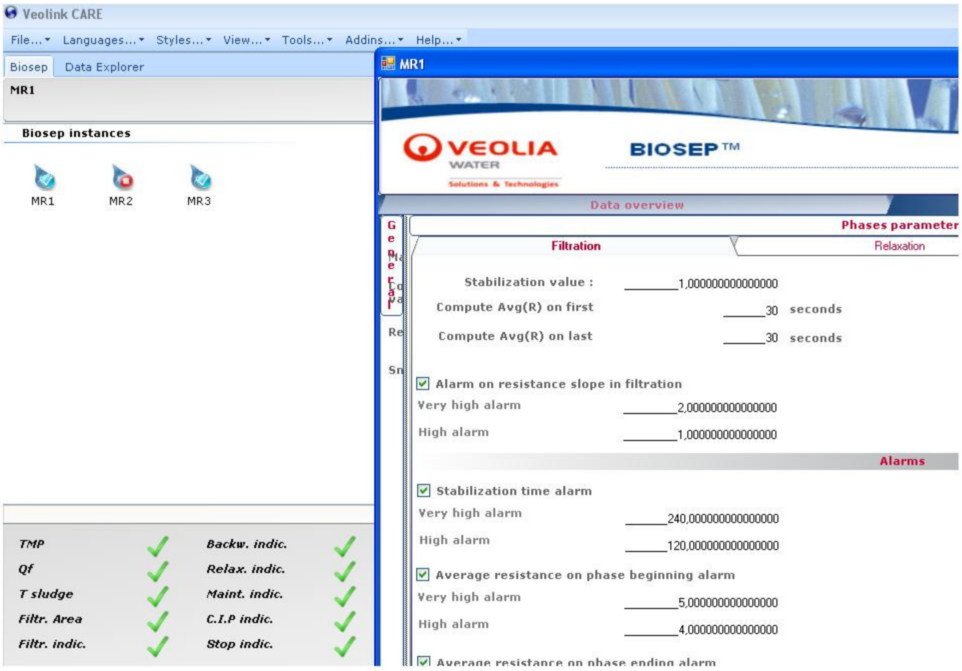



**Figure 17: Filtration performance observation with alarm settings (Veolink®)**

## 2.10 Sedimentation trials

Sedimentation trials started in September 2008 to investigate the possibility of load reduction in order to increase the MBR plant throughput. Provided that the nutrients reduction through sedimentation would deliver acceptable results, the implementation of a sedimentation step after the buffer tank would be taken into account to relax the overload situation in Margaretenhöhe.

**Materials and methods**

For representative samples from the sewer a sampler has taken a constant amount of wastewater before the buffer tank within 12 hours every 30 minutes. Afterwards the volume of the sediments was measured within 2 hours sedimentation time according to DIN 38409-H9-2 in an Imhoff cone (see Picture 2). The trials showed that mainly after 20 – 25 minutes the sedimentation was completed so that the sedimentation time was afterwards reduced to 30 minutes during the trials.

**Picture 2: Trials of sedimentation volume in Imhoff cone**

To investigate the possibility of load reductions, COD, TN, TP, Org. acids (with Hach Lange cuvette tests) and TS (with quick test, see Table 8) were measured in the mixed raw sample as well as in the supernatant after 30 minutes of sedimentation and compared. The data collection was taken within a 5 month period between September 2007 and January 2008.

### Results and discussion

The sedimentation volume after 30 minutes were in average 27 ml/L and within the range of 4 to 195 ml/L (see Appendix B). The values for Margaretenhöhe were slightly higher than average sedimentation volumes of other decentralized waster water treatment plants which are in the range of 1 to 20 ml/L. The elimination rates of the investigated parameters are given in Table 9.

**Table 9: Elimination rates of several parameters through sedimentation**

| Sample | COD [mg/L] | TN [mg/L] | TP [mg/L] | VFA [mg/L] | TS [g/L] |
|---|---|---|---|---|---|
| average (before sediment.) | 1357 | 178 | 20,9 | 348 | 2,6 |
| average (after sedimentat.) | 935 | 168 | 19,5 | 334 | 2,4 |
| range (before sediment.) | (830 - 1965) | (125,2 - 342,0) | (16,3 - 25,0) | (219 - 475) | (1,4 - 3,6) |
| range (after sedimentation) | (754 - 1198) | (118,6 - 328,0) | (15,2 - 22,4) | (208 - 447) | (1,2 - 3,2) |
| range (elimination [%]) | (9 - 58) | (0 - 25) | (0 - 19) | (0 - 41) | (0 - 39) |
| No of samples | 10 | 11 | 10 | 10 | 13 |
| ***average elemination:*** | ***27,9*** | ***6,8*** | ***5,5*** | ***5,2*** | ***17,2*** |
| ***standard deviation:*** | ***17*** | ***9*** | ***7*** | ***5*** | ***12*** |

The results of the mixed primary samples showed average concentration of 1350 mg/L COD, 180 mg/L TN, 21 mg/L TP, 360 ml/L organic acids and 2.6 g/L for TS, respectively. Significant elimination rates were only measured for COD with in average 28% (down to 1000 mg/L) and for TS with 17% (2.4 g/L). For the other parameters (TP, TN and org. acids) there were often no reduction measured or in a low range (see details Appendix A). These parameters are mainly dissolved in the wastewater and therefore not that much reduced through sedimentation.

The outcomes of the sedimentation trials could show that no load reduction of the critical parameters P and N could be achieved with a pre-sedimentation process step without chemical addition. In addition the experiences with the screening showed that the COD concentration after the 1mm screen was in the range of the data measured after sedimentation. The addition of a sedimentation step was therefore not perceived as a valid option to tackle the issue of overloading.

## 2.11 Transfer to operation department

One task of the project was to lay the foundations for the transfer of the MBR plant to the operation department in the WWTP Schönerlinde after the end of the project. To achieve this goal the following actions were undertaken:

- Enhancement of process control to minimize manual setting actions for the operators (see 2.8.6)
- Formulation of operation recommendations (see Chapter 6)
- Formulation of membrane cleaning recommendations with maintenance cleanings for reliable filtration (see Appendix E)
- Teaching of staff from operation department
- Project report

These points will help operators to handle the plants, but there will be no transfer to the operation department in the future. The situation of the plant overload with trucking away service and running a second MBR plant prompted the Berliner Wasserbetriebe to organize a connection of the Margaretenhöhe area to the public sewer network within the next two years. This is the most cost-effective solution according to the outcome of an economic feasibility study which compared the three options (i) plant expansion, (ii) trucking away service and (iii) status quo. The plant will be run by the project team of the Berliner Wasserbetriebe with support from the operation department until the connection is finished. The experiences made within the project showed that especially because of the irregular inflows from the sewer area and the identified hydraulic problems of the plant more care and maintenance is required than targeted (4h/week), which makes the operation not economically viable.

## 2.12 Plant design guidelines

To fulfill the treatment of all wastewater from the decentralized area without trucking away service another plant design would be necessary. The criteria for this new design can be taken from the loadings characteristics of the people living in this area and the actual cleaning capacity of the plant (see Figure 18 and Figure 19).

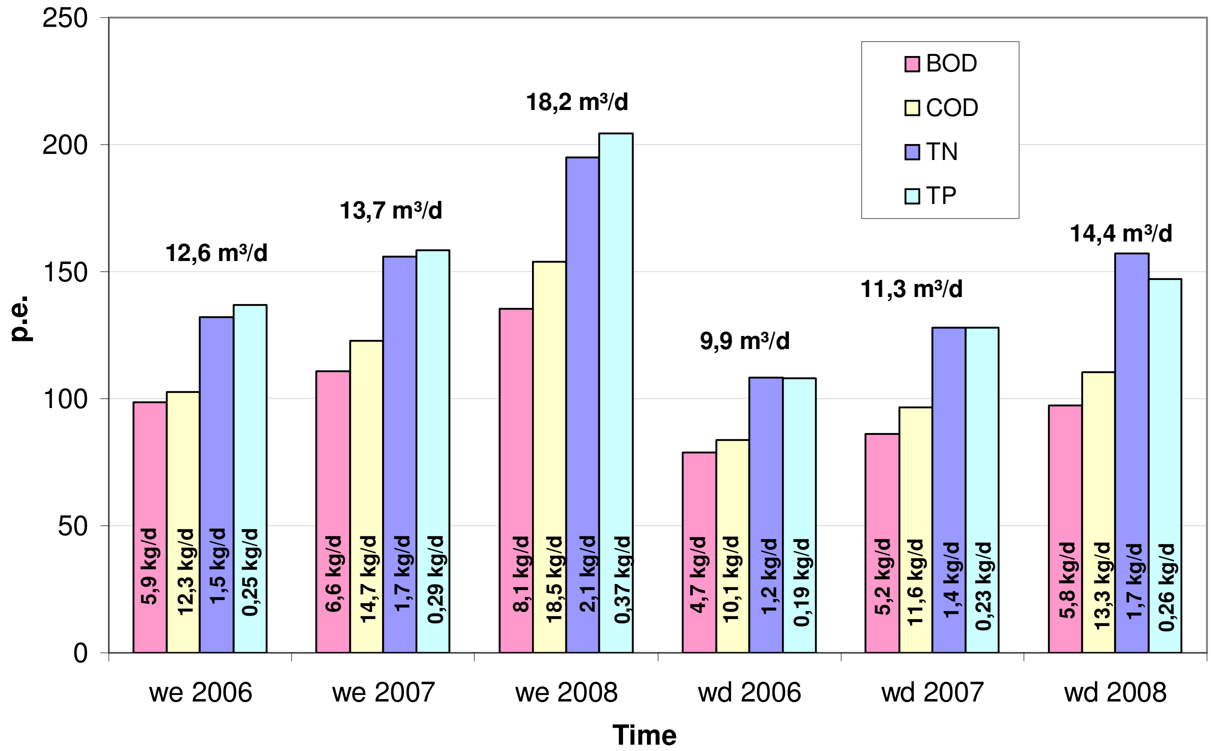



**Figure 18: Yearly average nutrients loadings from the decentralised area**

The yearly average nutrients loadings are calculated with the 24 h mixed samples and the related throughputs to the buffer tank, divided into samples taken on weekdays and weekends. The calculation into p.e. was made in relation to the values given for instance in (Henze and Ledin 2001) (see Table 10).

**Table 10: Calculation parameter from loadings to p.e.**

| Parameter | g / p.e. / d |
|---|---|
| COD | 120 |
| BOD | 60 |
| TN | 11 |
| TP | 1,8 |

It can be seen that the throughputs and the loadings are increasing every year though the connection of the people was finished end of 2006. The increasing throughputs could be explained by changed behavior of the people after connection from cesspits to central waste water discharge. The average throughput to the buffer tank in 2008 was 18.2 m³/d on weekends and 14.4 m³/d on weekdays. Against the assumptions in the beginning of the project the loadings did not stand on the same level through diluting but increased significantly.

There is also some discrepancy between COD loadings and the nutrients TP and TN: the TP and TN loadings are approx. 30-50% higher as the average wastewater composition. For 2008, 230 people connected on site produced the p.e. of around 200 people for TP and TN and 140 people for COD on weekends. On weekday the buffer tank throughput was around 150 p.e. for TP and TN and 100 for COD.

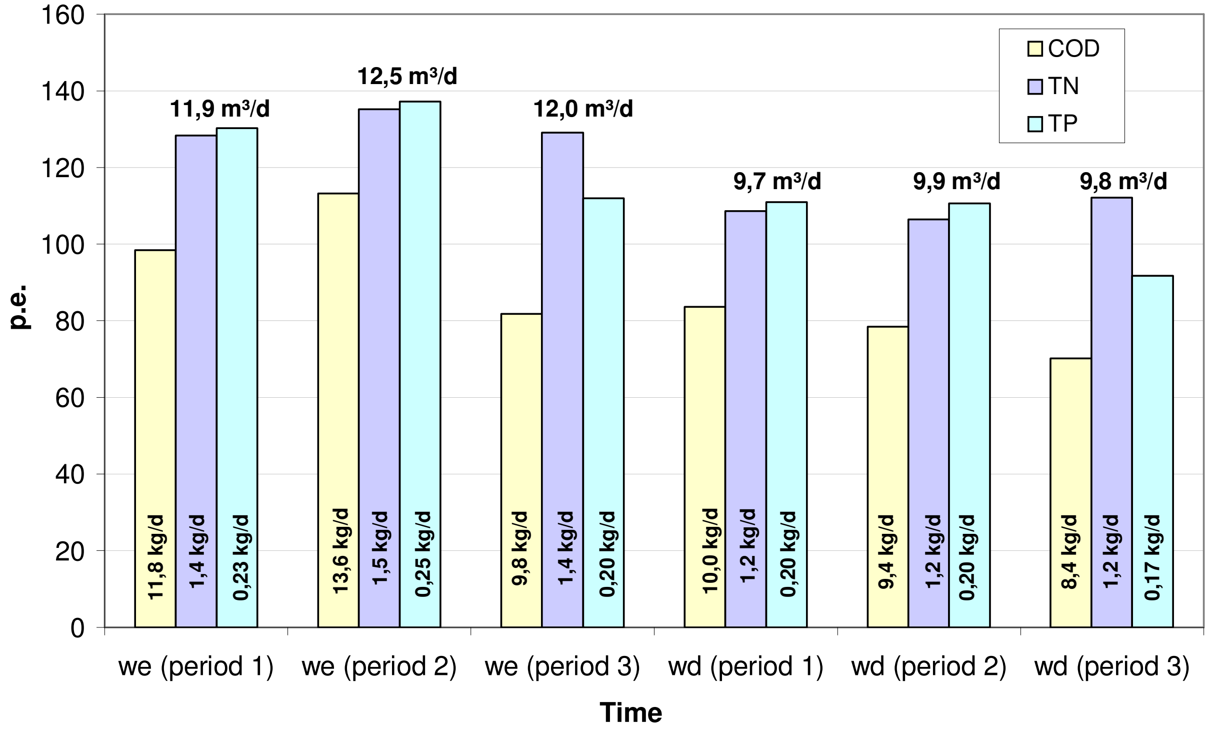



**Figure 19: Average nutrients loadings of the three periods treated by the demonstration plant**

The plant's treatment capacity was divided into the three periods given in this report also using the same 24 h mixed samples data. Compared to the calculation of the buffer tank throughput the loadings calculation of the plant throughput was made differently:

The definition of a weekend-day is a day where the plant throughput was 11 m³/d or higher. This could also be a weekday (e.g. public holyday, rain water event etc.). On the other hand not every weekday-day (< 11 m³/d) has to be on a real weekday. This calculation was made because of the different plant operations and throughputs (e.g. screening problems and fewer throughputs on some weekends, extending of high throughput settings to weekdays etc.).

The plant treatment capacity was on weekdays at around 10 m³/d and on weekends 12 - 12.5 m³/d. For determination of the maximum p.e. capacity the TN loading was chosen: The actual plant is able to treat around 130 p.e. with the given characteristics of the catchment area.

To design a new plant for real 250 p.e. under the given conditions, the plant's reactor volumes has to be enlarged by factor 3. This is because of the double loadings compared to the actual situation and to decrease the volumetric loadings into the design range with no dealing above the upper limit (see Figure 4). The nitrogen volumetric loading in case of enlargement of the reactor volumes by factor 3 and for 250 p.e. is shown in Figure 20.

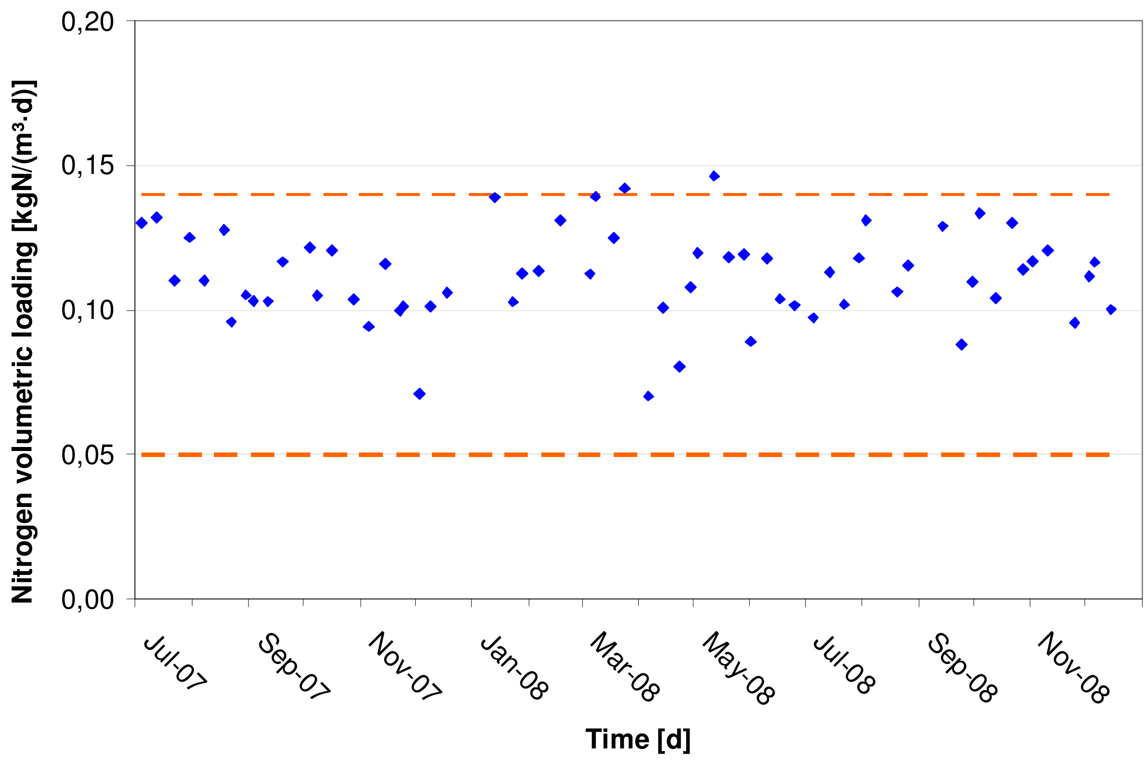



**Figure 20: Nitrogen volumetric loading for a new designed 250 p.e. plant**

With the experience of the actual plant it is also worthwhile to think about increasing the ratio of the anaerobic reactor, to enhance the hydraulic conditions in the reactors, to separate the filtrate tank from the sludge reactors and to renounce the deox-pot.

The proposed plant design explained above is the basis of the calculations given in Section 2.13 and Chapter 6

## 2.13 Energy consumption

The measured specific energy consumption related to the plant throughput is shown in Figure 21. For better interpretation of the data the daily average air flows of the blowers for the biology and for the membrane is added. An air conditioning system which was installed in April 2007 was operated automatically related to the indoor temperature and was also used occasionally for heating in winter.

The energy consumption in this figure was estimated from the measured overall energy consumption of all containers in Margaretenhöhe (demonstration, prototype and laboratory container) and the measured consumption of the prototype plant.

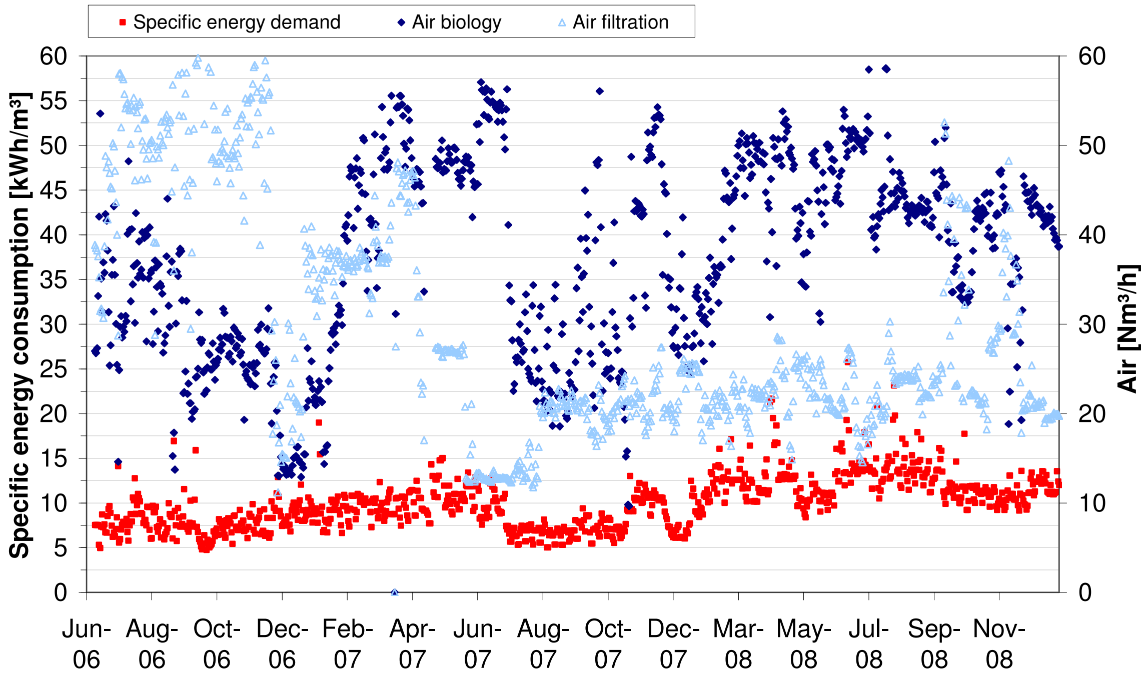



**Figure 21: Specific daily energy demand of the demonstration plant over time**

As seen in the figure, the specific energy demand is strongly related to the biology blower performance (dark blue diamonds). In periods with foaming because of low aeration (see Section 2.6) the blower had to work in a high level (up to 50 Nm³/h and higher) which was not necessary for biological nutrients removal (for example in Jun 07, spring 08). The energy demand increased up to 15 kWh / m³ and higher during these periods.

The light blue diamonds represent the daily air flow for the membrane filtration: Up to April 2007 most of the time two filtration lines were in operation whereas from May 2007 onwards, after the change of membrane system, only one line was in use. With the exception of some aeration tests around June 2007 with less than 15 Nm³/h the membrane aeration was operated with around 20 Nm³/h constantly. The flow exceeding 20 Nm³/h in summer and autumn 2008 was separated from the membrane aeration and used to support the oxygen transfer in the aerobic zone.

The estimated overall specific energy demand within the 2.5 years period (Jun. 2006 – Dez. 2008) was in average 10.4 kWh/m³, in the ENREM+ period at 10.8 kWh/m³ and before Jul. 2007 at 8.9 kWh/m³. It is important to know that the higher demands in 2008 are also related to the high biology blower performance because of foaming prevention and the conversion of several zones from mixing with stirrers into mixing with pumps. Additionally the higher performance of the filtration blower in summer and autumn 2008 which also supplied the aerobic zone (see Section 2.8.5) increased the demand.

Ideal operation conditions were observed from Jul. to Oct. 2007 where no increased blower requirements because of foaming were necessary and no stirrers were converted into mixing with pumps which had a higher energy demand. The specific energy demand during this period was in the range of 5 – 8 kWh/m³ and in average 6.8 kWh/m³. The demand increasing in Nov. and Dec. 2007 was caused through foaming, TS loss and plant throughput reduction.

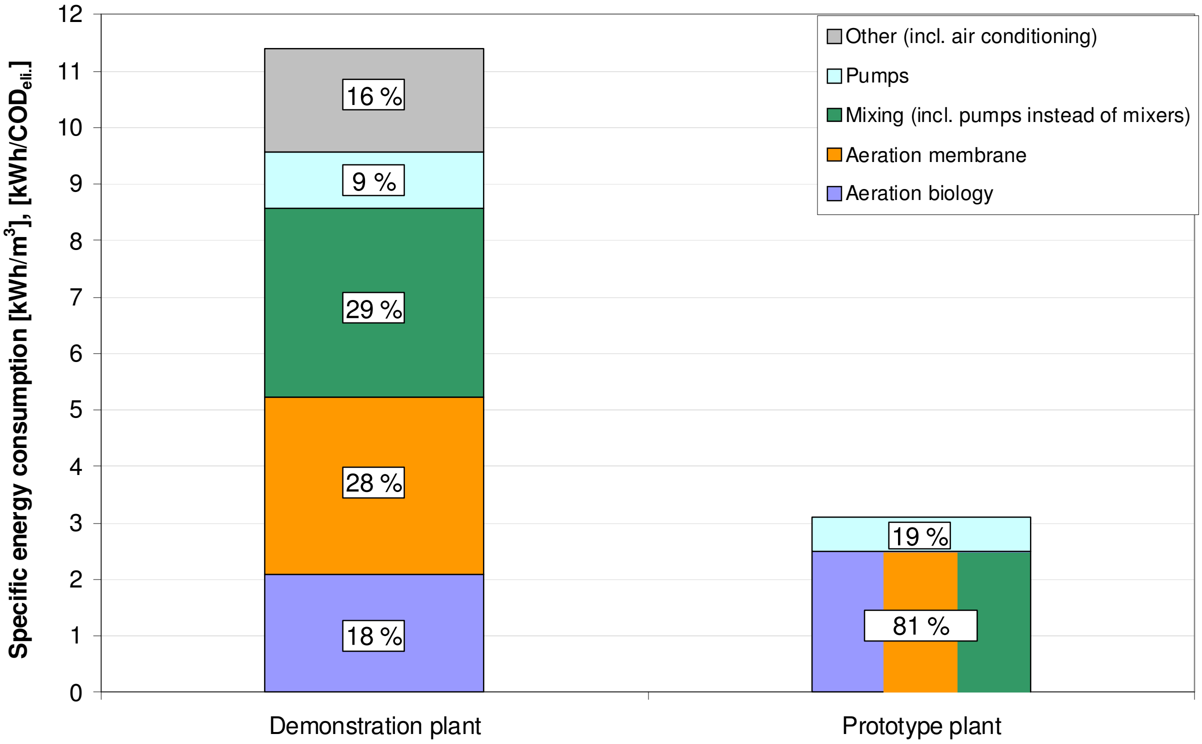



**Figure 22: Specific energy demand of demonstration and prototype plant (data 2009)**

A detailed investigation on the energy demand of the two plants was performed over a period of several months in 2009, using electric meters for each single consumer.

Figure 22 shows the measured specific energy consumption of the demonstration and the prototype plant. The data represents long time measurements from the overall energy consumption of both plants and from the blowers of the biology and the membranes. The data for other aggregates were estimated on the basis of the power ratings. The specific energy consumption is calculated with the average plant throughput during the measuring period (> 6 month for the demonstration plant, > 12 month for prototype plant). To represent the plant performances considering the high inflow concentrations the specific energy consumption is also related to the eliminated COD which is around 1 kg COD / m³ (see legend: kWh/$COD_{eli}$).

The average energy consumption for the demonstration plant under real working conditions was 122 kWh/d or 11.4 kWh/d with an average throughput of 10.5 m³/d. This value includes the running of the aeration blower in automatic (for oxygen supply only, not for foam prevention), the partly working air conditioning system and the running of several pumps for mixing instead of stirrers.

The aeration for the filtration requires around 28% of the total energy demand, for the biology around 18%. Another 29% are necessary for the mixing. The mixing energy demand could be reduced by installing better stirring systems. The exchange of two stirrers against pumps because of breakdowns related to foam resulted in an increase of around 1 kWh/m³. Notice: The increasing of the blower performance for foam prevention would further raise the energy demand by 3 kWh/m³.

The section “others” with 16% of the overall demand includes for example the air conditioning system, the permanently running computer system and the instrumentation. The running of the air conditioning system was depending on the outside temperature and required at hot days for cooling (or at cold days for heating) around 1.5 kWh/m³.

The average energy consumption for the prototype plant was 3.1 kWh/m³ or 3.1 $kWh/COD_{eli.}$ Because of the compact system where the aeration is responsible for both mixing and oxygen supply a differentiation between energy demand for membrane or biology aeration or mixing was not possible. For this system the aeration required around 81% and the pumping (feed pump, filtrate pump) around 19% of the total energy demand. The demand for other devices was negligible because of the lack of instrumentation and complex control.

The big gap between the energy demands of both plants could be partially explained because of the different working processes: The EBPR and Post-DN process in the demonstration plant require additional pumps and zones which had to be mixed permanently. The prototype plant is running in batch process which has periods with very low energy demand. Furthermore the aggregates for the demonstration plant are far away from energy optimization because of the unfavorable plant size between bench-scale and industrial scale.

## 2.14 Cost evaluation

As explained in the introduction, see Chapter 1, one of the goals of the ENREM project was to investigate the feasibility of decentralised wastewater treatment for suburban areas in Berlin which are not connected to the central sewer system. The current costs of app. 7 €/m³ for wastewater handling with tight cesspits compared to app. 2.50 €/m³ in the rest of Berlin made alternative solutions for these settlements relevant. Due to the regulations of the water authorities the effluent quality requirements were high despite the small plant size. It was assumed that the high instrumentation and the novel biological process would lead to increased specific treatment costs compared to central sanitation or decentralized treatment with lower effluent quality, but the exact figure had to be determined.

Therefore the operated plants (demonstration and prototype plant, see Chapter 5) were used to evaluate the costs for decentralized wastewater treatment with MBR systems. These plants were designed for diverse purposes in respect to the targeted effluent quality. The upgrade of the prototype plant (precipitation and adsorption filter) and the information gained during the long term operation of the demonstration plant allowed a detailed economical evaluation according to plant size and effluent quality. Furthermore a scale up was performed in order to estimate the costs for plant sizes up to 5000 PE in relation to the achieved elimination rates. The results can be used as a decision tool and help to define the ecologically required and economically feasible solution.

The costs were defined in €/m³ to give the possibility to compare the achieved results with published data, although this value does not include the wastewater constitution. Therefore the costs were also calculated in respect to the eliminated nutrient load. Considering the high concentrated wastewater of the catchment area Margaretenhöhe, this value describes the performance with a complementary perspective.

### 2.14.1 Material and methods

**Effluent quality definitions**

The economical analysis was performed for targeted treatment qualities achieved by different investigated wastewater treatment processes according to the following four elimination classes.

1. Minimum requirements:
    - Chemical oxygen demand (COD) removal and full nitrification
2. Additional nitrogen removal:
    - Total nitrogen (TN) elimination > 80 %
3. Additional nitrogen and phosphorus removal:
    - TN elimination > 80 %; Total phosphorus removal (TP) > 90 %
4. Additional nitrogen and enhanced phosphorus removal:
    - TN elimination > 80 %; TP elimination > 99 %

These elimination classes were achieved with the operated plants with different technologies and biological processes and were related to size, energy consumption and costs. This way a precise evaluation of the arisen costs could be determined and the actual benefits and efficiency of the applied processes could be quantified.

**Economical evaluation and scale up**

The experience gained during set up and operation of the plants were used to estimate the costs for decentralised, semi-decentralised and small scale treatment facilities. The sewer system was not included in the economical evaluation.

The assumptions for the economical evaluation were as follows:

- Period of validity: 25 years
- Wastewater constitution: Berlin, Margaretenhöhe
- Without VAT

*Investment and re-investment costs*

The evaluation of the investment and re-investment costs considered the actual costs for the two plants. Designed for different throughputs and effluent quality classes, the prices vary significantly for the investigated plants:

- Demonstration plant (TP eli > 99%; 130 PE): 288.000 € net
- Prototype plant (Minimum requirements, 50 PE): 32.200 € net
- 100% external finance with an interest rate of 5%

The installation of a buffer tank equalizing the throughput and a tank to collect the excess sludge and rough waste material is required for decentralised operation. In the presented case study both tanks contained 10 $m^3$ and were constructed of concrete. Furthermore the engineering costs were included to cover the construction and design costs of the units (20% engineering costs).

Also the purchase of the real estate where the plants are located was not included, as the costs for real estate vary significantly and a comparison of the treatment technology was in the focus of these investigations. The lifetime of the installed equipment was assumed with the experience of the financial department of the Berliner Wasser Betriebe, e.g. the membrane lifetime was given with 11 years.

*Energy demand*

The following data were assessed on the two investigated units.

- Demo (TP elimination > 99%): 8.5 kWh/m3 (10.5 m3/d)
- Prototype (minimum requirements): 3.1 kWh/m3 (4.2 m3/d)
- Heating and air condition not included
- 0.12 €/kWh

*Operational costs*

Following costs were used to calculate the operational costs:

- Energy costs
- Price increase for operation: 2%/y (included in the specific mean operational costs)
- Personal cost according to plant size e.g.:
  - 2 h/week 50 p.e.
  - 4 – 6h/week 130 p.e.
  - 14 h/week 5000 p.e.
- Sludge treatment:
  - < 1000 p.e.: 20 €/$m^3$ trucking
  - > 1000 p.e.: on site sludge treatment (Investment costs)
- Maintenance: according to existing maintenance contract
- Telecommunication: only for high tech plants
- Chemicals, e.g. Coagulant, cleaning detergents etc.
- Wastewater tax
- Membrane cleaning for the prototype plant is done by the supplier (Busse IS) app. twice a year. The company replaces the modules by clean ones and cleans the used modules using a cleaning protocol on site of Busse IS. The costs for this maintenance contract are included in the operational costs. For the scale up the specific cleaning costs of the demonstration plant are used to estimate the costs for chemicals.

**Scale up**

Semi-decentralised treatment facilities in the range of 50 – 5000 p.e. are promising application for the investigated processes. Based on the experience, research, measurements of energy consumption and used chemicals the operational costs for higher plant classes were assumed. The different aggregates, e.g. blowers and pumps, were chosen according to the new size resulting in more efficient aggregates, thus the specific energy demand decreases. This new equipment was also used to determine the new investment costs.

### 2.14.2 Results and Discussion

The results of this study are displayed showing the specific investment costs, specific energy demand, the specific mean operational costs and the specific overall costs in relation to the achieved effluent quality and plant size. The figures presented in the following sections are based on the two investigated plants, scale up and published data (Lesjean *et al.* 2008a). The presented figures are given within a range of +/- 20%.

**Investment costs**

The investment costs are given in Figure 23. The demonstration plant (High Tech) was scaled up to 250 p.e. and the results show the comparable high investment costs due to the high instrumentation (fix costs). With increasing plant size, the process of the demonstration plant becomes more attractive, as most of the instrumentation could be used also for larger plant sizes. For the prototype plant it is noticeable that higher effluent qualities increase the investment costs significantly only for the step from minimum requirements to required nitrogen elimination of 80%. The additional phosphorus removal up to 99% using precipitation and a downstream adsorption filter does not require expensive equipment. Therefore the Low tech system shows advantages for plant sizes up to 1000 p.e. considering only the investment costs.

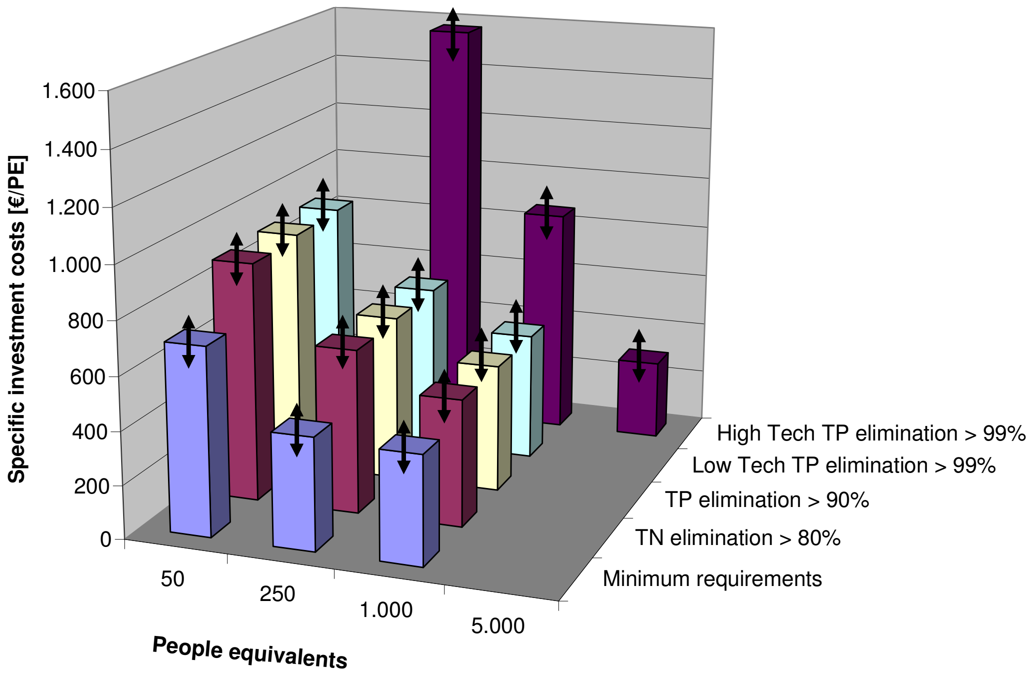



**Figure 23: Investment costs according to plant size and effluent quality**

**Energy demand**

Especially in decentralised wastewater treatment applications the energy demand is a crucial issue as electricity might be not available, is insecure or expensive due to decentralised production. Therefore emphasis has been put on this issue. The energy consumption was recorded and analysed during different states of operation, e.g. summer and winter, and over a sufficient period of time (minimum one year). Aeration is the main energy consumer for wastewater treatment, especially for MBR systems due to required membrane aeration. Several groups investigated this issue and their findings show the significant impact on the overall energy consumption (Verrecht et al. 2008).

*Demonstration plant*

Due to the small size of the demonstration plant, the blowers run often within a disadvantageous spectrum of operation. Therefore it was assumed that a scale up to the size of 1000 – 5000 would be beneficial. This is illustrated in Figure 24: The energy demand for the high effluent quality and 250 p.e. is comparably high. The scale up assuming more efficient equipment shows a significant influence. The energy demand for high effluent quality and 1000 p.e. is in the same range as for mid quality presented in literature (Lesjean *et al.* 2008a). For plant sizes of 5000 p.e. and larger the energy demand decreases further more showing the competitiveness of the ENREM process.

*Prototype plant*

Discussing the energy demand of the prototype plant, it is remarkable, that an increased effluent quality does not show a significant increase in the energy demand. This can be explained by the fact that in this set up the additional cleaning performance is not due to higher aeration or better mixing, but to the implementation of precipitation and adsorption. The additional energy consumption, e.g. additional dosage pump for the precipitant, is negligible. The results confirm however the assumptions that the concept of the demonstration plant is a feasible option for catchment areas of 5000 p.e. and larger. The energy demand is as high as for lower effluent qualities, thus the benefit of a higher effluent quality does not cost more energy.

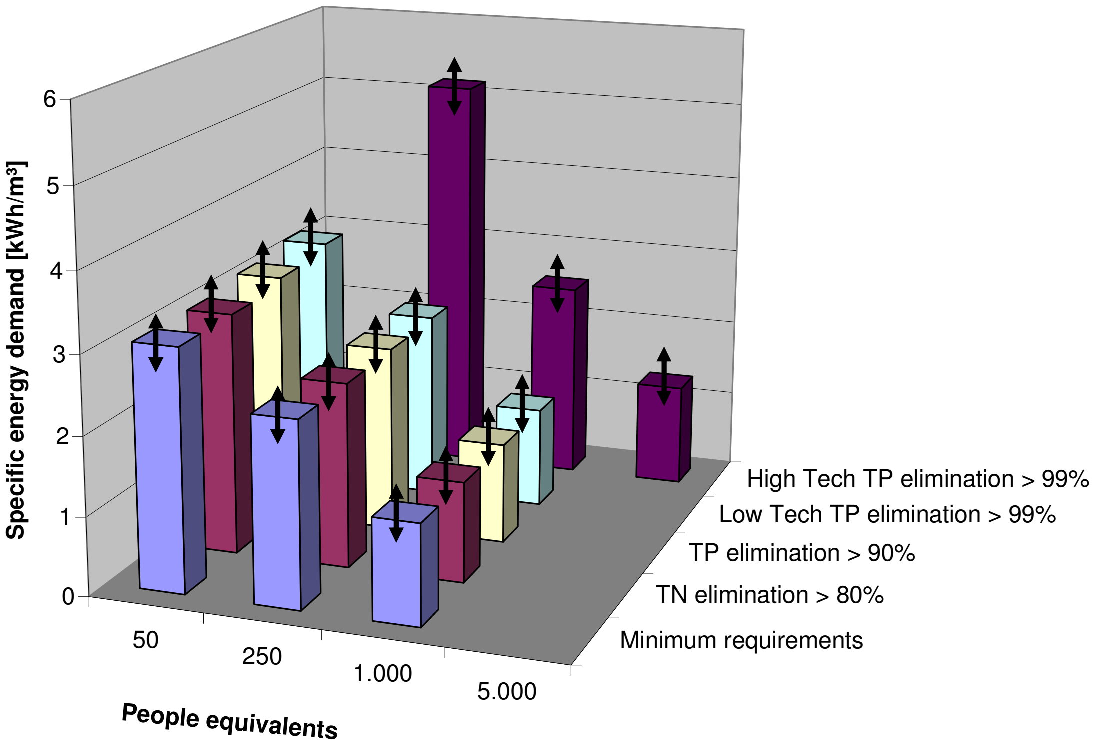



**Figure 24: Energy demand according to plant size and effluent quality**

### Operational costs

The advantage of the prototype plant set up for small scale applications was shown for both, investment and energy costs, with high effluent qualities and can be explained by the applied technology. The usage of relatively high amounts of precipitant and adsorption material is the additional price that has to be paid in comparison to the technology of the demonstration plant. The effect of these additional costs is shown in Figure 25. Comparing the operational costs of both technologies for high effluent qualities and plant sizes larger than 1000 p.e., the competitiveness of the High Tech solution becomes obvious. Especially the adsorption material for the Low Tech solution increases the operational costs significantly.

This also shows the great effort that has to be made to reach very high effluent qualities in decentralized wastewater treatment. The last step reaching 99% phosphorus elimination requires either high energy (High Tech) or additional physical treatment (Low Tech). Both options are expensive and play a major role in the operational costs.

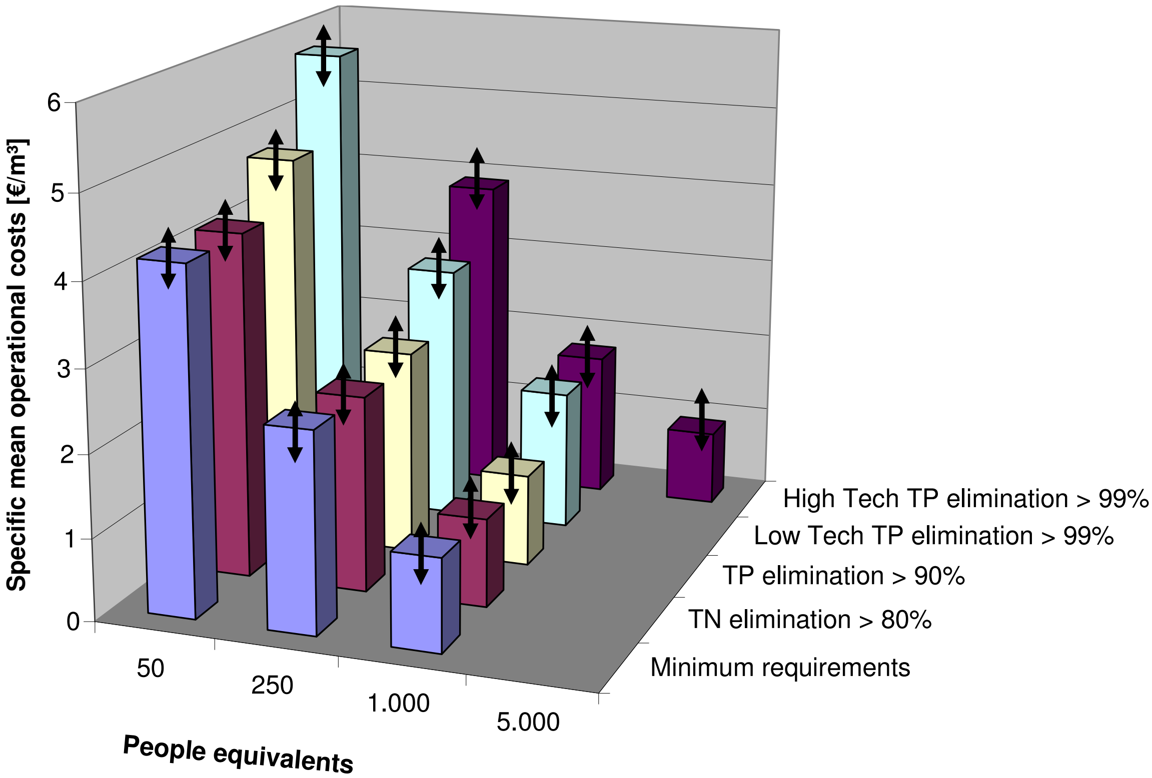



**Figure 25: Operational costs according to plant size and effluent quality**

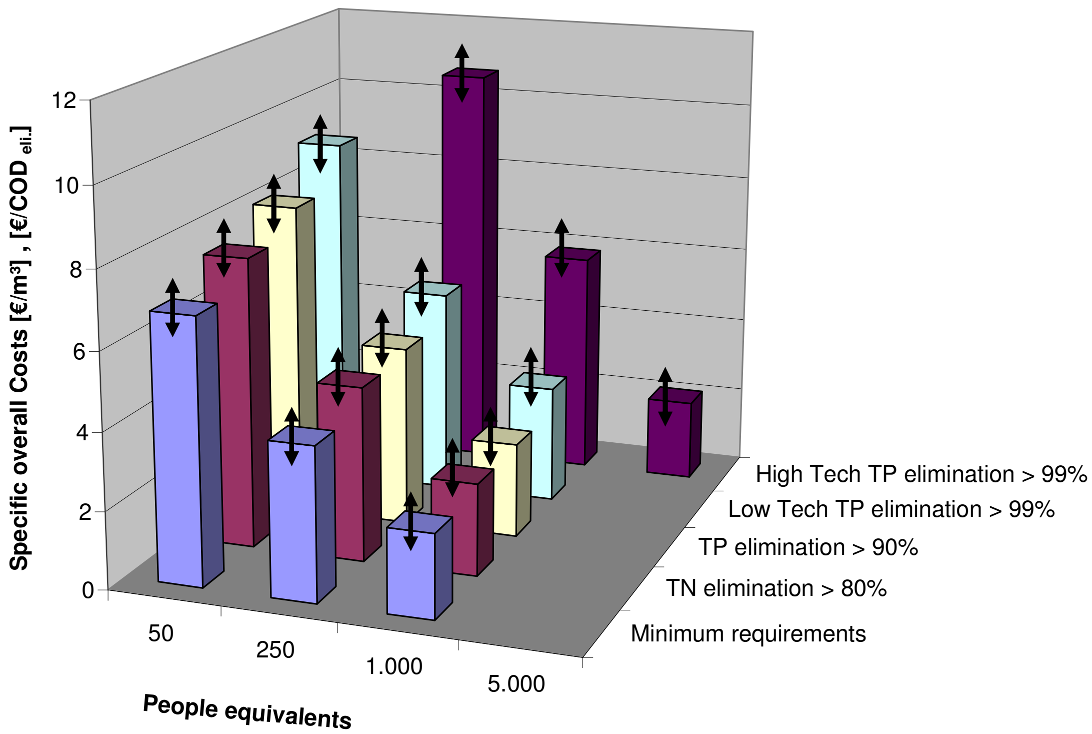



**Figure 26: Specific overall costs according plant size and effluent quality**

### Specific overall costs

The specific overall costs combine the different aspects of the economical evaluation and are given in Figure 26.

The High Tech solution shows relatively high costs in the smaller plant classes and thus reflects the unfavourable size for the installed equipment. Both, the specific investment and the specific operational costs, are high for the smaller plant classes. But the benefits of the biological process are shown for the larger plant classes. In Table 11 the numbers for the different plant sizes are given. The given figures show again the high costs for high effluent qualities in decentralized wastewater treatment plants, to be compared with the benchmark cost of 7 €/m3 for trucking away the wastewater produced by the household from tight septic tanks to the closest WWTP.

**Table 11: Specific overall costs for the investigated technologies**

| | **Demonstration plant** (TP elimination > 99%) | **Prototype plant** (TP elimination > 99%) |
|---|---|---|
| Plant size p.e. | Overall costs in € / m3 wastewater (net) | |
| 50 | | 7.5 – 10.5* |
| 130 | 16 – 17** | |
| 250 | 8.5 – 12.8 | 4 – 6 |
| 1000 | 4.6 – 6. 8 | 2.5 – 3.8 |
| 5000 | 1.8 – 2.7 | |

* assessment on operated prototype plant; ** assessment on operated demonstration plant

### 2.14.3 Conclusions

High effluent qualities in decentralized wastewater treatment applications are achievable with the presented technologies, but the high specific costs have to be considered and the overall benefits of the applied technologies have to be evaluated carefully.

The calculated energy demand for the different technologies was:

- High Tech: 2.5 – 5.2 kWh/m3 for 250 – 1000 p.e. and < 1 $kWh/m^3$ above 5000 pe
- Low Tech: 1.4 – 2.4 kWh/m3 for 250 – 1000 p.e.

The reason why the demonstration plant (High Tech) is not economically viable for smaller decentralized wastewater treatment plants is the unfavourable size of equipment and the high degree of instrumentation. But the scale up to 5000 p.e. proved to be a promising option when high effluent qualities are required. The energy consumption is not higher than those of existing plants of the same size achieving lower effluent qualities. Therefore the ENREM-process is a feasible option for these kinds of applications.

The operated prototype unit (Low Tech) showed to be a good solution for catchment areas of 50 - 1000 p.e. The effluent quality could be easily upgraded in terms of phosphorus removal implementing precipitation and a downstream adsorption filter. To assure sufficient nitrogen removal, anoxic conditions have to be adequate, thus the plant size has to be carefully planned and operated.

Carrying out a complete LCA for the prototype plant, the precipitant and the adsorption material have to be included. The impact on the operational costs was demonstrated and a LCA would also consider the production, shipment and disposal or recycle of the adsorption material, helping to evaluate the overall impact on the environment. The sustainability of the complete process cycle has to be evaluated in order to define a suitable application for wastewater treatment thus the energy demand is of great interest and has to be considered in the life cycle analysis.

# Chapter 3

# Biology and process optimization

The ENREM process scheme combining enhanced biological phosphorus removal (EBPR) with post-denitrification without addition of an external carbon source is capable to reach high elimination rates of nutrients, thus leading to low effluent concentrations. To achieve these high effluent qualities specific conditions for the activated sludge are mandatory. For EBPR processes the anaerobic stage is necessary to enrich the activated sludge biocenosis with phosphate accumulating organisms (PAOs). To ensure the advantage of competition of PAOs over "pre-denitrifiers" nitrate recirculation to the anaerobic reactor has to be minimized. The presence of nitrate and readily biodegradable COD at the same time will lead to the predominant existence of "pre-denitrifiers". In consequence to this disturbance not only the EBPR dynamics are affected, due to lack of carbon source in the anaerobic reactor, the denitrification rates in the downstream anoxic reactors decline too. This was found in the previous investigation period from January 2004 till June 2007. The experience gained during this project showed the importance of the biological conditions. Therefore to ensure an adequate biological environment by process optimization was a major task throughout this project. The applied measures as well as the associated results are presented in the following sections.

A response strategy to disturbances on the biological process was identified and implemented in the process control scheme (addition of carbon source in the anaerobic reactor). The outcomes are shown in Section 3.1.

The former chosen ratios of the different environments seemed not to be appropriate. Therefore the reactor conditions were adapted enabling more anoxic conditions volume for denitrification as explained in 3.2.

Despite the numerous experiments to identify the carbon source used for post-denitrification the metabolisms were not fully understood. To be able to investigate this further, a laboratory plant was constructed and operated with a synthetic wastewater containing acetate as the sole carbon source. This defined synthetic wastewater helped to gain more information on the post-denitrification metabolisms, see Section 3.3.

To ensure the process stability and increase the operational safety with regards to the effluent quality the dosage of $FeCl_3$ in a low concentration was implemented as a co-precipitation step. The effect on the biological performance was also monitored by batch test experiments in a period with and without co-precipitation, Section 3.4.

## 3.1 Carbon addition

The recurrent operational difficulties under constant overloading conditions regarding the nitrate recirculation and the following loss of EBPR dynamic required a response strategy to reduce the time needed to recover the phosphorous elimination capacity. Besides the later discussed process optimization, a quick and automated response should help to reduce the negative impact of shock loads. The possibility of an automated response was expected to help while acting at the first indication of disturbed operation.

The enrichment of the activated sludge with phosphorous accumulating organisms (PAOs) can only be successful when providing adequate conditions, see (Mino *et al.* 1998). The loss of phosphate elimination capacity was correlated to the recirculation of high amounts of nitrate thus leading to pre-denitrification. Readily biodegradable COD (rbCOD), e.g. volatile fatty acids (VFAs), present in the wastewater were quickly consumed and depleted by this pre-denitrification step thus letting no sufficient carbon source available for the EBPR dynamic. In order to avoid this pattern during periods of nitrate recirculation to the anaerobic reactor, the dosage of acetate into the anaerobic reactor was tested. The increased amount of VFAs, coming from the wastewater and the additional acetate dosage, was expected to be sufficient for both processes, the temporary unavoidable pre-denitrification step and the necessary phosphorous release and VFAs uptake by PAOs.

The initial amount of acetate dosed to the anaerobic reactor was calculated according to the theoretically needed amount for denitrification of the recirculated nitrate. It was necessary to avoid an overdosing of acetate as the leaking of high amounts of VFAs to the aerobic reactor was known to lead to an unfavorable change of the biocenosis. During operation the amount of acetate identified to improve the EBPR performance was between 2 – 2.5 L/d, 40 % acetic acid or 80 – 90 g $CH_3COOH$ / $m^3$ inflow.

During the tests for this approach the acetate amount has been adapted. At the same time the quick respond of the process optimization described in Section 3.2 regarding the nitrate recirculation helped to recover the EBPR performance. Since the conversion of the reactors (Period 2) acetate dosage was used to strengthen the fraction of PAOs in times of nitrate recirculation or insufficient phosphorus removal performance.

The online monitoring of effluent phosphate concentration was used to automate the acetate dosage. Following control scheme was identified:

- Start of dosage when phosphate effluent concentrations reach 0.08 mgP/L
- Stop of dosage when phosphate effluent concentration falls below 0.06 mgP/L

In addition, acetate dosage was manually activated when nitrate in the effluent was above 10 mgN/L.

Figure 27 shows the recorded concentration of phosphate and the time when acetate dosage was active. The phosphate effluent concentration is shown by the red curve in mgP/L. A period of one month is presented. The black bars indicate the period of acetate dosage.

Peaks of phosphate in the effluent appear during stable operation possibly due to a stop of inflow or peak loads in the influent. During the presented period of operation the nitrate recirculation was always below 10 mgN/L, thus this was not the reason for increased phosphorus effluent concentration. The automated dosage of carbon starts once the threshold was reached and the quick response of the EBPR performance is clearly demonstrated. In comparison to periods of operation when the carbon addition was not installed the phosphorus peaks could be reduced in both, the maximum value and duration, thus decreasing significantly the discharged amount of phosphorus.

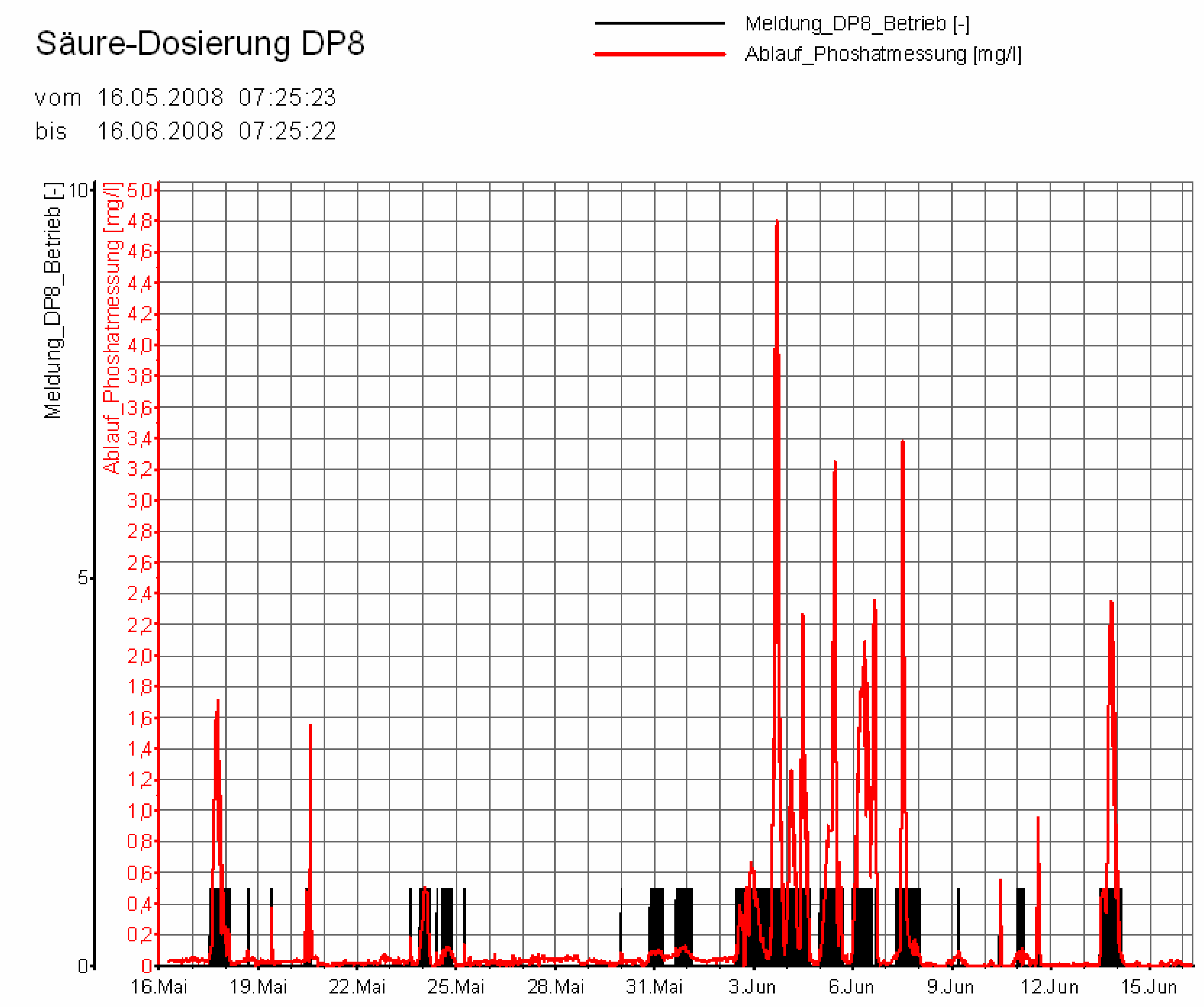



**Figure 27: Acetate dosing depending on the online PO4-P effluent value**

The implementation of an automated acetate dosage into the anaerobic reactor when phosphate peaks in the effluent appear was successfully tested. Due to the low amount of acetate fed the costs for acetate could be identified with 0.28 €/m$^3$ during periods of dosing. In 2008 the overall consumption of 60% acetic acid was 140 liters with overall specific costs of 0.07 €/m$^3$ at 3854 m³ annual plant throughput. The given price is calculated for the purchase of small amounts of acetate as the storage of high amounts of chemicals on site should be reduced to the minimum with regards to safety standards.

Further optimization of the control scheme is implemented in 2009, introducing the online nitrate signal. The improved control scheme will activate the dosage according to both effluent concentrations, phosphate and nitrate. Thus peak loads of nitrogen causing nitrate recirculation will be recognized before the EBPR dynamic is disturbed. The immediate response is expected to stabilize the whole process.

## 3.2 Biological / Process optimization

Facing operational difficulties during winter 07/08 with low sludge temperatures, the denitrification ability was insufficient and led to high nitrate content in the recycle stream from the last anoxic reactor. In consequence of this long period the biological phosphorus removal collapsed thus high nitrate and phosphate concentrations in the effluent were registered. To counter this vicious circle an increase of the anoxic volume was identified to be the best option to decrease the nitrate concentration without adding an external carbon source directly to the anoxic chamber. The direct dosage of an external carbon source might have helped to regain the biological phosphorous removal activity, but it would have been contradictory to the actual target of implying the post-denitrification without addition of an external carbon source.

The increase of the anoxic volume was done by switching the second aerobic reactor as first anoxic. The nitrification was identified to be complete after the first aerobic reactor. With the low temperatures in winter and spring, the oxygen solubility and therefore the oxygen mass transfer was sufficient. A carry over up to 2.0 mgN/L of ammonia was acceptable, as the membrane reactor was following and functioned as additional step for nitrification. But it was obvious that longer periods with a slight overload of ammonia will lead to an accumulation of ammonia, finally reaching the threshold concentration for inhibition of nitrification. Therefore the ammonia concentration was the crucial parameter to be monitored. As the aerobic reactor was mixed by the aeration no agitator was installed in the first place in the new anoxic reactor. After conversion of the reactors mixing was ensured by a submerged pump. The high dissolved oxygen concentration in the first aerobic reactor led to some carry over of dissolved oxygen to the new first anoxic reactor. But the great surface area helped in terms of degassing. Figure 28 shows the flow sheet after the conversion of the reactors.

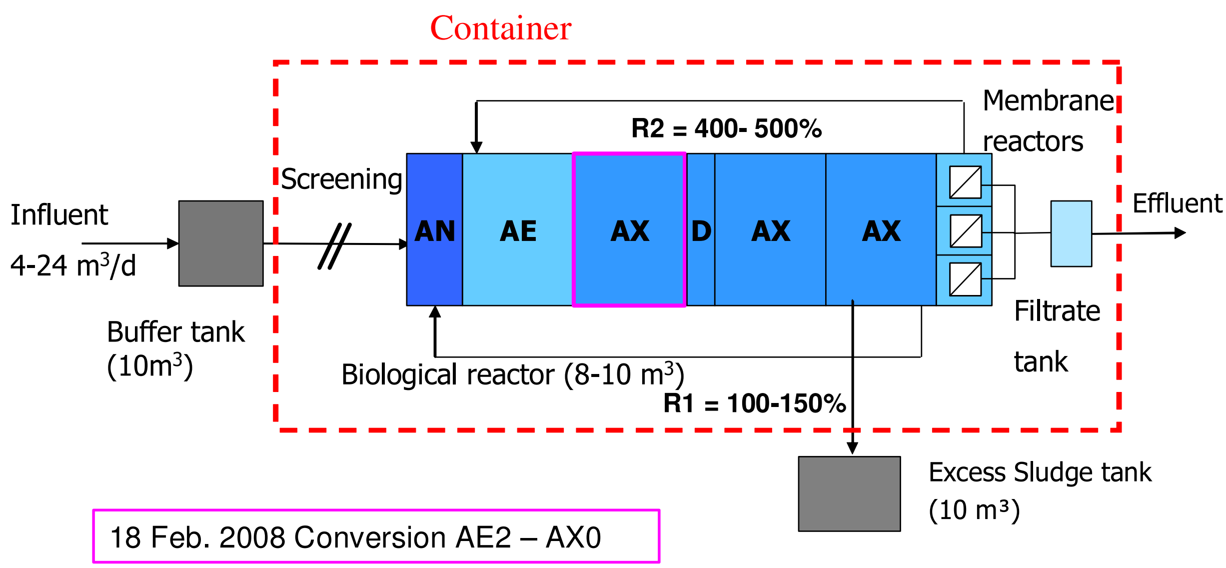



**Figure 28: Flow sheet of the MBR plant with process optimization**

This increased anoxic volume showed immediate effects on the denitrification capacity and the nitrate effluent concentration, see Figure 10, and a long term effect on the phosphorus dynamic and phosphate effluent concentration (Figure 9). Figure 29 shows a cycle study before the conversion.

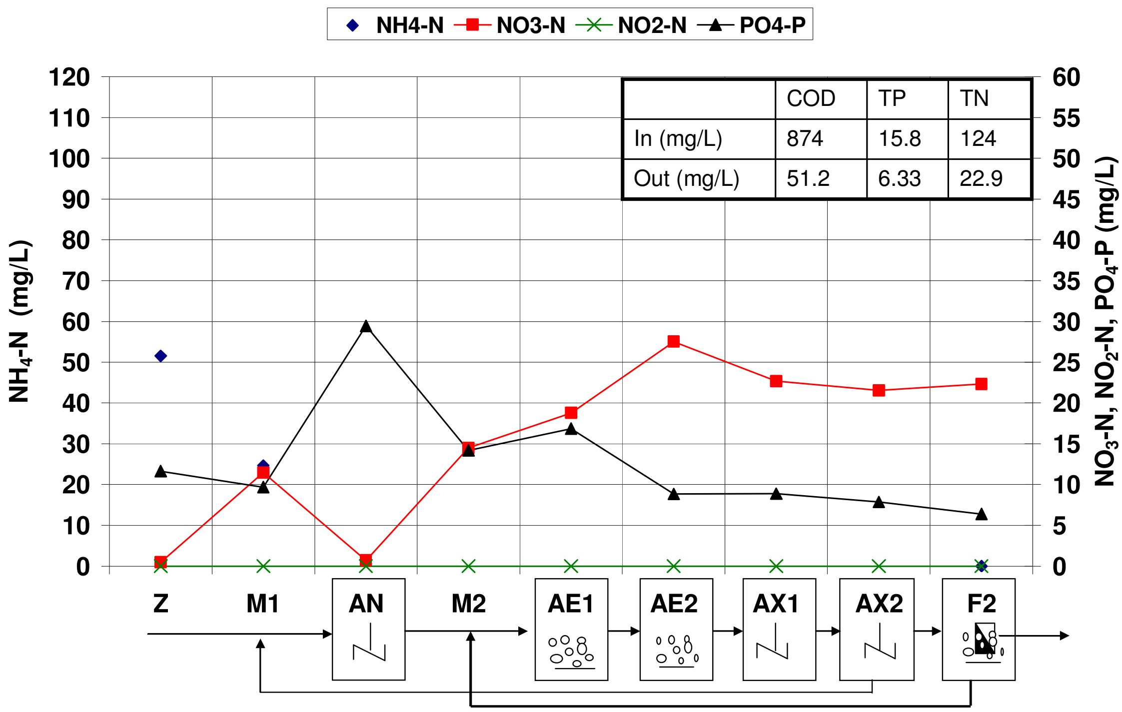



**Figure 29: Cycle study (February 7th 2008)**

The nitrate recirculation is shown by the carry over to the anaerobic reactor and the rapid denitrification there. Nitrification occurred as expected in the aerobic reactors, but the following post-denitrification was too slow to avoid the high effluent concentration.

At the same time, readily biodegradable COD present in the wastewater was used for denitrification, therefore missing for the uptake by PAOs. The phosphate release was consequently low as well as the uptake in the aerobic reactors leading to high phosphate effluent concentrations.

After the conversion a quick response on the denitrification capacity was seen, but the phosphate dynamic needed more time to adapt to the changed conditions. The long period of raised nitrate recirculation weakened the PAOs fraction. About five weeks after the conversion and adaptation the phosphate dynamic was established again as seen by higher phosphate release in the anaerobic reactor, see Figure 30. This cycle study shows clearly the recovered EBPR dynamic and the post-denitrification capacity leading to low effluent concentration for nitrate and phosphate. The denitrification rate was with 0.98 mgNO3-N/(gVSS*h) in the first anoxic reactor AX0 above endogenous rates. Further investigations on the achieved enhanced denitrification rates are presented in Section 3.3.

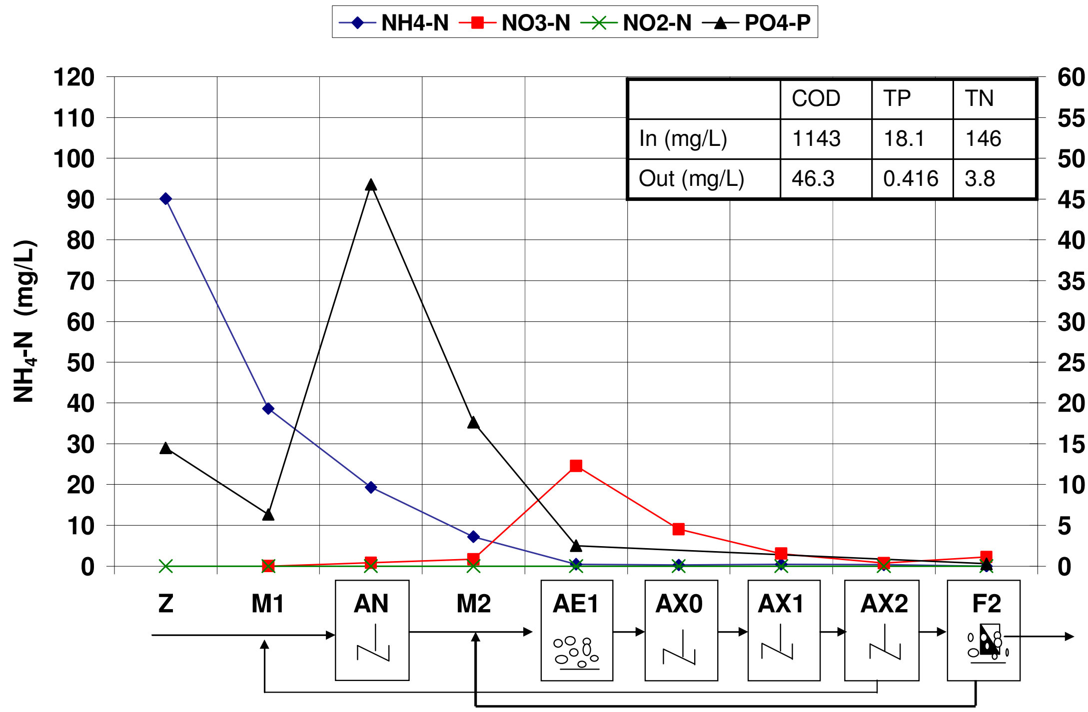



**Figure 30: Cycle study (March 12th 2008)**

Despite the successful optimization with the increased denitrification zone the demonstration plant was still overloaded and thus on the edge of the biological capacity.

## 3.3 Internal carbon source used for post-denitrification

Several investigations on the post-denitrification capacity of the ENREM process scheme showed denitrification rates higher than the expected endogenous ones (Vocks *et al.* 2005; Lesjean *et al.* 2008b; Vocks 2008). The measured enhanced DNRs raised the question which carbon source is used for denitrification. In experiments addressing this issue it was possible to rule out external and soluble carbon sources such as extracelullar polymeric substances (EPS) or lysis/hydrolysis products (Vocks 2008). Furthermore it seems to be unlikely that adsorption of carbonaceous substances could serve as a carbon source, due to the presence of the aerobic zone before the denitrification step. Acetate showed a low tendency for adsorption in batch test experiments (Bracklow et al. 2007). Easily biodegradable substances, e.g. volatile fatty acids (VFA) are either stored as internal compounds in the anaerobic reactor or consumed within the aerobic reactor hence these sources should not be available for post-denitrification. To be able to exclude the possibility of slowly biodegradable (particulate or soluble) COD being utilized as a carbon source for post-denitrification a laboratory scale MBR plant was constructed and operated in cooperation with TU Berlin.

### 3.3.1 Experimental set up and methods

This lab plant was operated with a substrate containing acetate as sole carbon source for five months to show the independence of this process with respect to particulate and slowly biodegradable COD.

The lab scale plant had a volume of 6 liters and was set up as a sequenced batch MBR (SBMBR). The sequences were arranged according to the ENREM process scheme, first anaerobic, typical for EBPR processes followed by an aerobic phase, necessary for phosphorus uptake and finally the anoxic phase for post-denitrification. Filtration was induced during the aerobic phase in order to minimize the overall cycle time, thus increasing the sludge load.

A small flat sheet membrane module was assembled using ultra filtration (UF) modules designed for laboratory purposes by Microdin-Nadyr©.

Control of the pH-value was crucial to ensure ideal conditions for the activated sludge. Therefore an automatic pH control was implemented keeping the pH-value between 7.2 and 8.4.

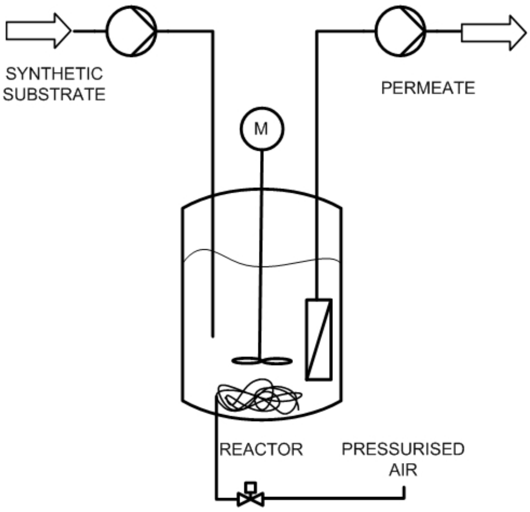



**Figure 31: Laboratory scale plant**

Table 12 shows the operational parameters of the lab scale plant. Solid retention time (SRT) was held constant at 25 days. The contact time in the different phases was similar than those of the full scale plant. The long hydraulic retention time compared to real wastewater treatment plants can be explained by the experimental set up, as the limited space in the small sequenced batch reactor did not allow implementing a membrane module providing sufficient surface area. The short filtration sequence, during aerobic phase leads to a high membrane surface area requirement. To be able to operate the sequenced batch reactor, a longer hydraulic retention time was accepted.

**Table 12: Lab scale plant parameters**

| Lab scale MBR | |
|---|---|
| Reactor volume [L] | 6 |
| Flow rate [L/h] | ~0.072 |
| Solid retention time [d] | 25 |
| Hydraulic retention time [h] | ~83 |
| Anaerobic contact time [min] | 40 – 50 |
| Aerobic contact time [min] | 50 – 60 |
| Anoxic contact time [min] | 90 |
| Anoxic contact time during long cycle experiments [min] | 200 – 415 |
| **Synthetic wastewater** | |
| COD [g/L] | 2.4 – 6.0 |
| Total nitrogen [mgN/L] | 90 – 230 |
| Total phosphorus [mgP/L] | ~109 |

The used substrate consisted of sodium acetate as the sole carbon source and a mineral media with the following components: 0.255 mg/L CaCl2, 840 µg/L MgSO4 7H2O, 9 mg/L EDTA, 135 µg/L CoCl2 6H2O, 27 µg/L CuSO 6H2O, 810 µg/L FeCl3 6H2O, 135 µg/L H3BO3, 162 µg/L KI, 54 µg/L MnCl 4 H2O, 54 µg/L Na2MoO4 2H2O and 108 µg/L ZnSO4 7 H2O.

Table 13 shows the resulting mass organic loadings of the lab plant. The mean feed to micro organism ratio (Mean F/M) is calculated as the COD sludge load over a day while the initial F/M is calculated for each single cycle. Both parameters are necessary to describe the operation condition sufficiently. The initial F/M was important, as an overload at the beginning of the cycle led to an incomplete substrate uptake, thus acetate could be carried over to the aerobic phase. This led to a shift of the microbial community and the kinetics of interest could not be determined. At the same time the mean F/M had to be sufficient for the metabolism of biological phosphorus removal and post-denitrification.

**Table 13: Mass organic loads of prototype unit**

| | Lab scale MBR |
|---|---|
| Mean F/M [g COD/g VSS.d] | 0.177 |
| Initial F/M [g COD/g VSS] | 0.027 |
| Mean VSS [g/L] | 6.9 |

Ammonia chloride and nitrate were fed as nitrogen sources. Ammonia chloride was fed as growth nutrients in a low concentration and nitrate was added to the anoxic phase in the right amount to ensure full denitrification, ~ 4 - 6 $mgNO_3\text{-}N/L_{Reactor}$, thus no carry over of nitrate into the next cycle and the anaerobic phase occurred. Figure 32 shows an operation cycle with the moments of feedings.

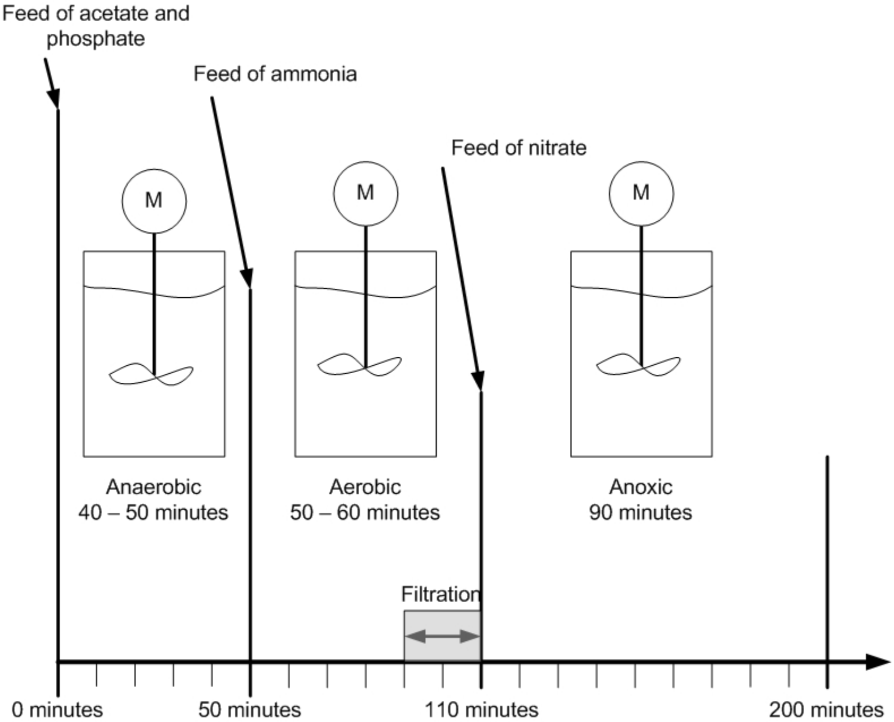



**Figure 32: Operation cycle with the biological conditions and times of dosage**

K2HPO4 and KH2PO4 were fed as phosphate sources in varying concentrations (0 - 480 mg/L).

An Ion Chromatograph (Dionex DX 100; IonPac 4a column) was used to analyze nitrate, nitrite, phosphate and ammonium of samples the lab scale MBR. COD, total nitrogen (TN) and total phosphorus (TP) were measured with Dr. Hach-Lange cuvette test kits. All cuvette tests complied with ISO 8466-1, DIN 38402 A51 and DIN 32645 in calibration, detection and quantitation limits.

In order to calculate the rates and therefore being able to describe the biological performance of the system, samples of the mixed liquor were collected in intervals of five to fifteen minutes. These samples were filtered immediately using 0.2 µm cellulose acetate membranes.

Two different procedures were applied:

I. For cycle studies during continuous operation the operational parameters, as sludge load and contact times, were kept constant. These experiments represent the exact conditions during operation.

II. To demonstrate the denitrification capacity, long cycle experiments were carried out in which the initial sludge load and the nitrate concentration were increased. In addition the anoxic phase was extended to show the evolution of the DNR.

### 3.3.2 Results and discussion

Table 14 shows the mean DNRs obtained for the lab scale plant during operation with a synthetic monosubstrate. The DNRs were above 0.6 [mgNO3-N/h gVSS] indicating the above threshold for endogenous DNR. This demonstrated that the enhanced denitrification of the ENREM process observed with real wastewater could be observed also with monosubstrate, and therefore that particulate or slowly biodegradable COD of the raw water was not the source of carbon for denitrification. The DNRs of Table 14 were calculated without the long cycle experiments (discussed later), which were carried out with a higher initial sludge load, leading to higher DNRs. The reason for these lower mass organic loads during continuous operation was the prevention of acetate residual leaking to the aerobic phase. Therefore the mass organic load during continuous operation was set to a lower value than the biological system was actually capable of. In previous studies it was observed that the permanent carry-over of acetate in the aerobic phase led to a fast change of the biocenosis resulting in a break down of the biological phosphorus removal.

**Table 14: DNRs obtained during continuous operation (without long cycle experiments)**

| | **Lab scale MBR** |
|---|---|
| Number of measurements | 9 |
| Temperature in °C | 21.2 |
| Mean Total solids in g/L | 11.52 |
| Mean VSS in g/L | 6.95 |
| Mean DNR [mgNO3-N/(gVSS*h)] | 0.79 |
| Min DNR [mgNO3-N/(gVSS*h)] | 0.65 |
| Max DNR [mgNO3-N/(gVSS*h)] | 1.06 |

Further experiments were carried out to investigate the evolution of the DNR during a long anoxic phase. Therefore the anoxic phase was extended and nitrate concentration was manually raised to prevent a limitation of the DNR. Figure 33 shows the results of one cycle study. Starting with high DNR above endogenous rates, the DNR reached values in the highest range of endogenous denitrification after about 2 hours. The decrease of the denitrification rate with time indicates the depletion of the used carbon source. This is in agreement with the results of Vocks et al. (2005), although in the current results no clear elbow could be distinguished in the DNR evolution. This discrepancy might be due to the different experimental setting e.g. Vocks et al. (2005) used real wastewater activated sludge, while the experiments discussed in this article were conducted with activated sludge adapted to a mono substrate under laboratory conditions.

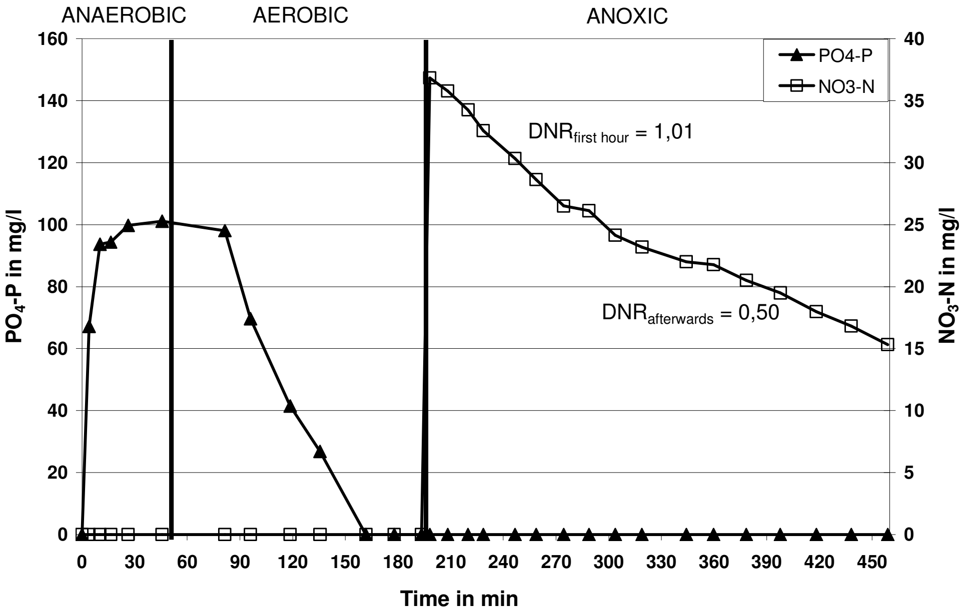



**Figure 33: Phosphate and Nitrate evolution during long cycle experiment 7**

Table 4 shows that similar results were consistently observed over 7 long cycle measurements performed in different days over several weeks of operation with the lab scale MBR, with clear differences of the rates determined in the beginning and the end of each experiment.

**Table 15: DNRs obtained during long cycle experiments**

| Experiment | $DNR_{first\ hour}$ [mgNO3-N/(h*gVSS)] | $DNR_{afterwards}$ [mgNO3-N/(h*gVSS)] | Initial F/M [gCOD/gVSS] |
|---|---|---|---|
| 1 | 0.99 | 0.31 | 0.033 |
| 2 | 0.97 | 0.57 | 0.034 |
| 3 | 1.12 | 0.69 | 0.031 |
| 4 | 1.09 | 0.05 | 0.045 |
| 5 | 0.89 | 0.63 | 0.035 |
| 6 | 1.04 | 0.81 | 0.052 |
| 7 | 1.01 | 0.50 | 0.041 |
| Mean | 1.02 | 0.51 | 0.039 |

A correlation between DNRs and initial F/M can be noted, see Figure 34. The importance of the mass organic load was previously discussed in literature (Bracklow *et al.* 2008; Lesjean *et al.* 2008c; Vocks 2008). The results obtained during the present study show also a relationship to the initial F/M. This is not trivial as this denotes a correlation of the denitrification rate and the carbon load in the anaerobic zone (i.e. before aeration) and not in the anoxic zone as commonly accepted in case of dosing of external carbon source.

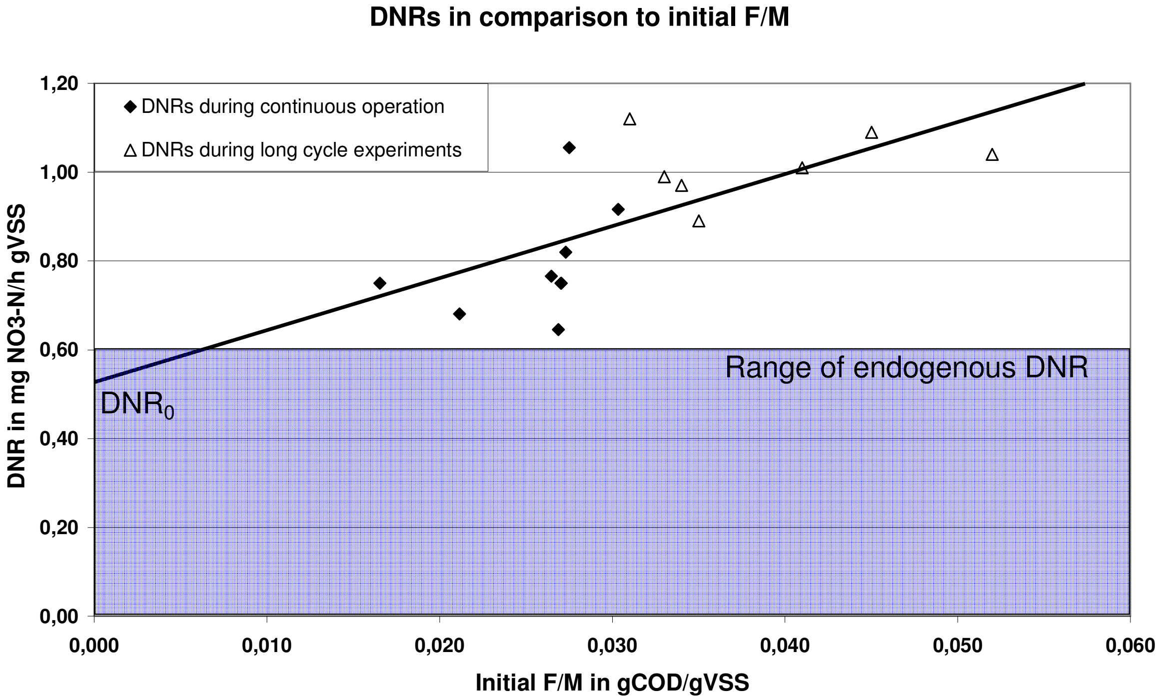



**Figure 34: Correlation of DNRs to F/M**

The range of endogenous denitrification rates is shown in the highlighted field up to 0.6 mg NO3-N/(gVSS*h) and the linear regression of all collected results points to an $DNR_0$, defined as the DNR at no initial F/M.

The use of the synthetic mono substrate excludes the usage of particulate COD or sbCOD from the feed water for denitrification, as acetate is the only carbon source present in the substrate. In addition, the long operation period of lab scale MBR of five months with mono substrate, being capable of reproducing the high DNRs shows the independence of this process with regards to sbCOD or particulate COD. One can therefore conclude that the particulate or sbCOD present in the raw water cannot explain the observed enhanced denitrification rate. The question of the carbon source remains therefore open. Previously it was demonstrated that sources such as extracelullar polymeric substances (EPS) present in the wastewater or lysis/hydrolysis products could not be accounted for the unknown carbon source (Vocks 2008).

It can be therefore hypothesized that the carbon source used for denitrification must be coming from the biomass itself. Two hypotheses are proposed:

(i) an internally stored carbon source is used for the denitrification, as formerly discussed by Vocks et al. (2005) and Lesjean et al. (2008), or

(ii) bound EPS produced by the biomass is used for the denitrification.

Further investigations are required in order to validate one of these hypotheses. However, we can already note that if the first hypothesis could clearly explain the observed correlation between mass organic load in the anaerobic zone and DNR in the anoxic zone, the second hypothesis fails to directly describe this correlation.

### 3.3.3 Conclusions

The upper threshold for endogenous denitrification rates can be set at 0.6 mg NO3-N/(gVSS*h) (Kujawa and Klapwijk 1999; Vocks 2008). Higher denitrification rates are assumed to be only possible if a readily biodegradable carbon source is present during the anoxic phase. To identify the used carbon source for post-denitrification and to be able to identify the role of slowly biodegradable COD and particulate COD in this process, a lab scale MBR was operated with a synthetic wastewater containing only acetate as the sole carbon source. By achieving higher rates than the endogenous rate, it was proven that sbCOD or particulate COD can not be accounted for the enhanced rates.

Further experiments to investigate the pathway of carbon in this process focusing on the carbon mass balance for the anaerobic phase and the conversion of acetate using the nuclear magnetic resonance (NMR) technology were carried out in cooperation with the New University Lisbon. The results obtained suggested that denitrifying PAOs (DPAOs) were very likely the denitrifying microorganisms in the post-denitrifying step. The carbon source used by these bacteria is under anaerobic conditions internally stored PHAs. Therefore further experiments focusing on PHAs in correlation to the DNR are proposed. Additionally the newly developed FISH probes for *A. phosphatis clade I* should be used to link the kinetics to the microbiological community.

## 3.4 Phosphorus co-precipitation

### 3.4.1 Results of trials on demonstration plant

After process optimization in period 2 the phosphorus effluent concentrations were stable in a low range with an average concentration of 0.23 mgP/L for total phosphorus and 0.12 mgP/L for orthophosphate without addition of metalsalts. The TP peaks went up to 0.5 mgP/L which exceeded the project goal of 0.2 mgP/L. In order to reach further minimization of the effluent values, a low precipitation was started end of August 2008. The start was nearly at the maximum sludge temperature for better comparison of the results with and without precipitant in the same range of $T_0$. Ferric chloride [$Fe(III)Cl_3$] was chosen as precipitant and continuously added into the anoxic zone 2. The amount of ferric dosing relates to the feed and was in the very low range of 3 – 4 $gFe/m^3$. The results are shown in Figure 35:

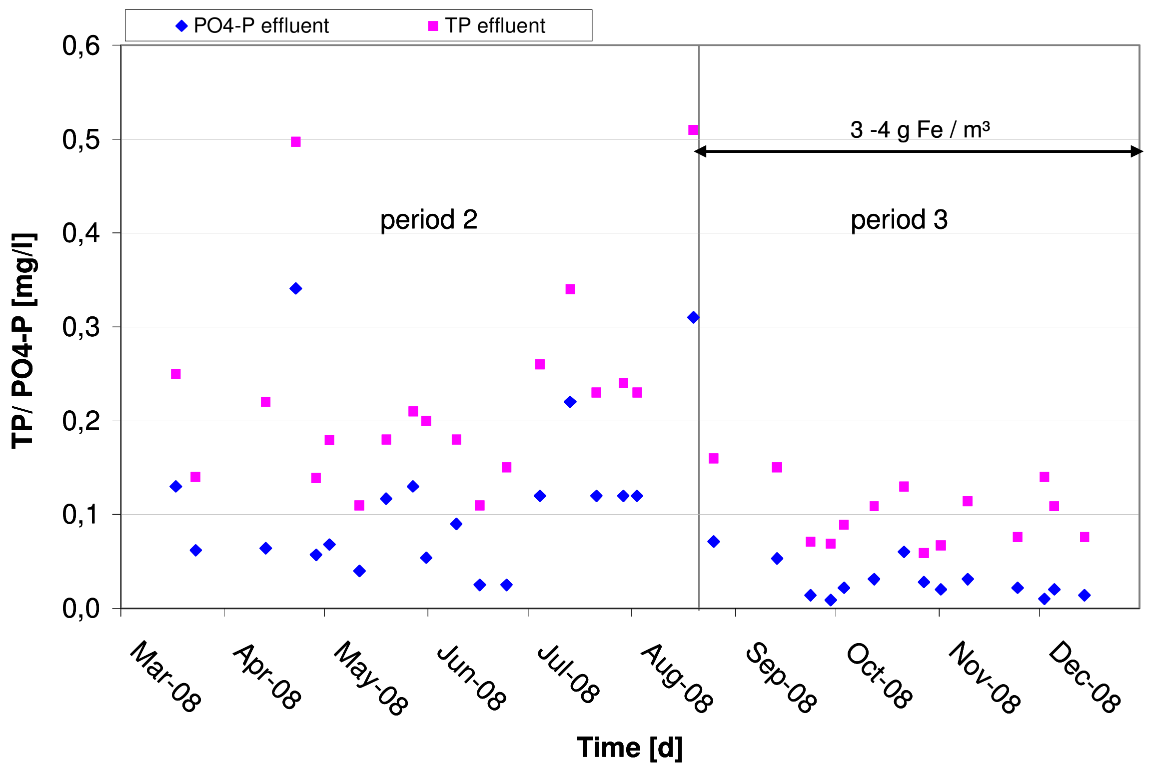



**Figure 35: Phosphorus effluent concentrations before and after precipitation (24 h samples)**

After start of precipitation there was a significant decrease of the effluent values. The average concentration for total phosphorus was in this period 0.10 mgP/L and for orthophosphate 0.03 mgP/L. Even the threshold of 0.2 mgP/L for total phosphorus could be reached for every sample what is a very good performance with the background that the average refractory fraction of phosphorus was around 0.07 mgP/L in the 3rd period.

The pH-value in the aerobic zone did not change between the 2nd and 3rd period and remained in average at 8.1.

The refractory phosphorus fraction – the difference between total phosphorus and orthophosphate in the membrane filtrated effluent – is shown in Figure 36. In period 1 with unsteady phosphorus removal the average refractory phosphorus fraction was around 0.29 mgP/L. The precipitation provided a reduction of this fraction during period 2 (stable phosphorus removal without precipitation) and period 3 of 0.04 mgP/L from 0.11 mgP/L to 0.07 mgP/L in average.

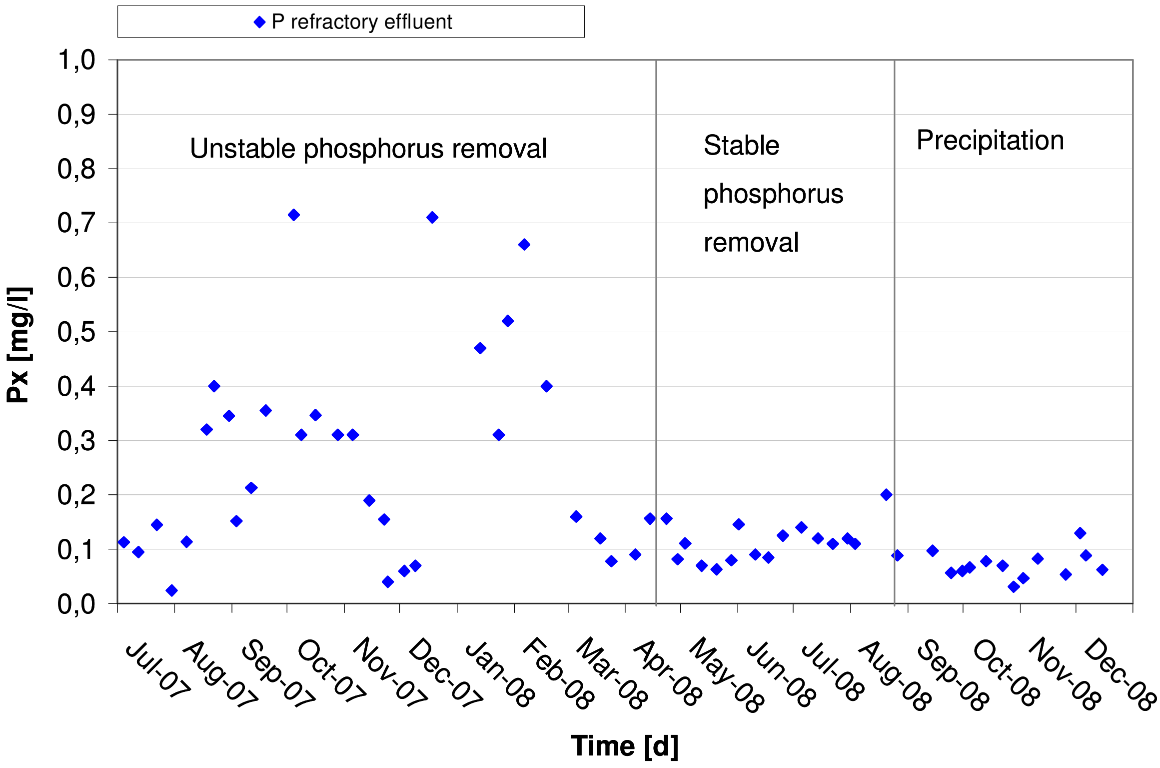



**Figure 36: Refractory phosphorus fraction of the MBR effluent**

The costs for the described precipitation are in the range of around 0.35 – 0.45 €cent/m³ with precipitant supply via the large waste water treatment plants (bulk buying). This means for Margaretenhöhe additionally costs for chemicals of around 15 € yearly. For individual order the costs might be slightly higher.

### 3.4.2 Accompanying batch test experiments

The dosage of a low amount of precipitant was assumed to have a twofold effect on the overall phosphorous elimination performance:

i. The precipitation of excess phosphate that has not been taken up by PAOs during aerobic conditions. This additional cleaning capacity was supposed to help to reduce further the effluent concentration, especially during shock load events.

ii. The outcomes of a prior study showed a significantly increased EBPR dynamic when dosing a low amount of $FeCl_3$ leading to low effluent phosphorous concentration (Adam and Kraume 2003). The increase of the phosphate uptake rate (PUR) with precipitation was shown to be almost twice as high as without precipitation. Due the low amount, it was assumed that the dosage of $FeCl_3$ did have a positive effect on the biological kinetics.

In order to confirm this second hypothesis and to be able to evaluate the effects of co-precipitation on the EBPR performance, batch test experiments accompanying the introduction of precipitation were carried out. The influence of precipitant dosage on the EBPR dynamic can be best described comparing the phosphorous release and uptake rates (PRR and PUR). The phosphorous uptake rates were determined for both conditions, aerobic ($PUR_{AE}$) and anoxic ($PUR_{AX}$). Experiments with sludge taken from the demonstration plant during a period without (EBPR) and with dosage of precipitant (EBPR+co-precipitation) were compared. The batch test experiments were carried out in a two liter batch reactor aerated with nitrogen gas for anaerobic and anoxic conditions and with pressurized air during aerobic conditions ensuring a minimum dissolved oxygen concentration of 2.0 mg/L. The batch tests phases followed the ENREM process scheme, anaerobic, aerobic and anoxic, one hour each. In the beginning acetate as a carbon source was fed targeting an initial feed to micro-organism ratio (F/M) of approximately 0.07 gCOD/gVSS. As the VSS value was measured after the experiment, this F/M value was set according to the expected VSS value and calculated afterwards. Ammonia was fed at start of the aerobic phase, recording the nitrification rate. Nitrate formed during the aerobic phase was denitrified during the following anoxic phase. The pH-value was kept constant between 7.0 and 8.0 and the temperature was approximately 20℃. For both periods ten batch test experiments were analyzed.

3.4.2.1 Effects on EBPR dynamics

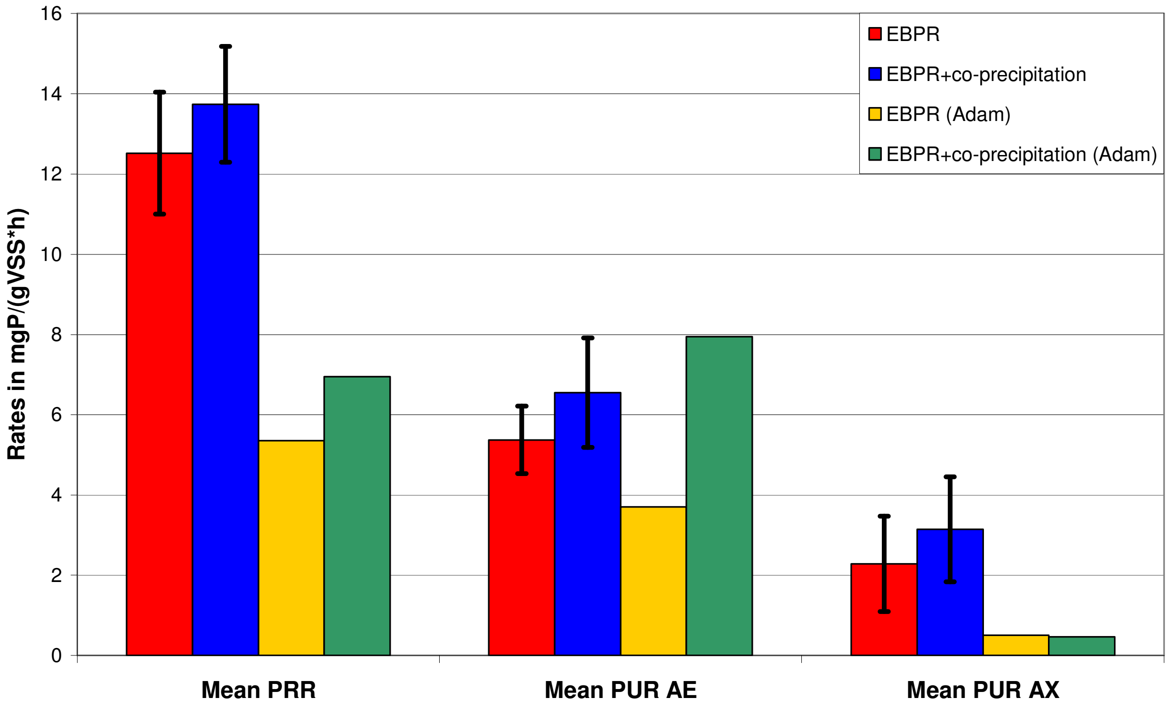



**Figure 37: Comparison of achieved rates (EBPR dynamics)**

Figure 37 shows the obtained rates for the current study and (Adam 2004).

Higher PRR values compared to the presented values by (Adam and Kraume 2003; Adam 2004) can be explained by the high phosphorous content in the activated sludge,~ 4%, due to stable EBPR performance of the demonstration plant, whilst (Adam 2004) obtained a stable TP/TS concentration of around ~ 3%.

An increase of the $PUR_{AE}$ with co-precipitation could be determined in both studies, though the increase in the present study was with below 20% significantly lower than those measured by (Adam 2004), who could record a raise of more than 50%. For the $PUR_{AX}$ during the EBPR+co-precipitation period, higher rates could be determined compared to the EBPR phase. It has to be stated that during stable operation of the demonstration plant the phosphorus uptake was usually completed within the aerobic reactor, thus the anoxic phosphorus uptake did not contribute to the overall phosphorous elimination performance.

In comparison to (Adam 2004) the $PUR_{AX}$ are clearly higher what can be explained by the low phosphorous concentration after the aerobic phase in the batch test experiments carried out by (Adam 2004).

Table 16 shows the obtained EBPR dynamics obtained during the present study in comparison to (Adam 2004). The β-value is correlated according to the inflow total phosphorus concentration in molFe/molP. The mean inflow values for period 2 and 3 are 19.9 mgP/L and 16.8 mgP/L respectively, see Section 2.4.1. A ratio is given correlating the rates obtained for EBPR with the ones for EBPR+co-precipitation. This shows the effect on the compared rates.

**Table 16: Comparison of EBPR dynamics**

| | **This study** | | | **(Adam 2004)** | | |
|---|---|---|---|---|---|---|
| mgP/(gVSS*h) | **EBPR** | **EBPR+co-precipitation** | **Ratio %** | **EBPR** | **EBPR+co-precipitation** | **Ratio %** |
| **PRR** | 12.52 | 13.74 | 91.1 | 5.35 | 6.95 | 77 |
| **$PUR_{AE}$** | 5.27 | 6.55 | 82.0 | 3.7 | 7.95 | 47 |
| **$PUR_{AX}$** | 2.29 | 3.14 | 72.8 | 0.5 | 0.46 | 109 |
| **β (Fe/P)** | 0.087 – 0.137 | | | 0.15 | | |

According to the present investigations only a slight improvement of the EBPR dynamics correlated to the dosage of precipitant could be observed. The dosage did not show any massive impact on the EBPR dynamic within the demonstration plant. Nevertheless the elimination ratio was increased and the total effluent concentrations were lowered by the co-precipitation und helped to stabilize the phosphorus elimination performance in the demonstration plant and avoid infrequent peaks due to short feeding interruptions, membrane cleanings etc.

#### 3.4.2.2 Effects on nitrogen kinetics

The rates for nitrification (NR) and denitrification (DNR) were as well determined during these batch test experiments to be able to register any influence of the nitrogen metabolism and removal efficiency. Figure 38 shows the results for the experiments. No clear trend can be stated. During the EBPR phase without precipitation the nitrification rate seemed to be slightly higher and more stable. The F/M ratio was in the beginning lower than targeted, so the first experiments were not used for the calculation of the mean nitrification and denitrification rates. Only experiments where the F/M was above 0.027 gCOD/gVSS were considered. The rates were similar for each phase, thus no influence of co-precipitation can clearly be determined.

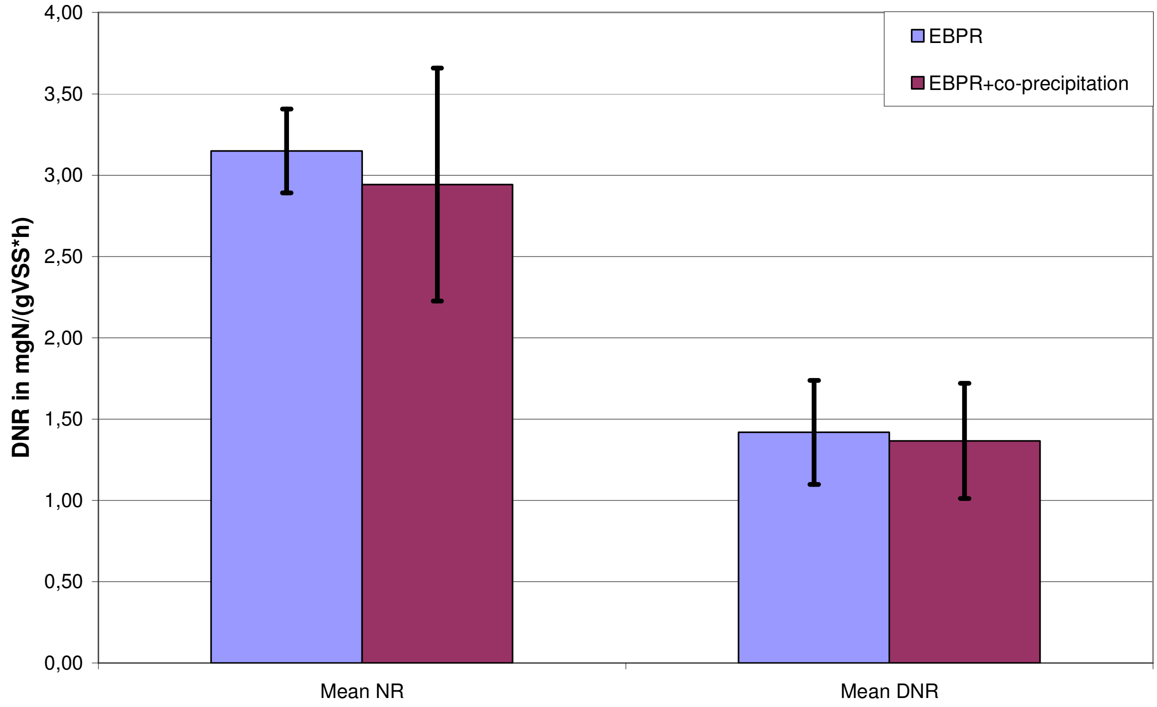



**Figure 38: Comparison of achieved rates (nitrogen kinetics)**

# Chapter 4

# Fouling, membrane filtration performance and cleaning

A key factor for the success of membrane based wastewater treatment is the reliability of the membrane modules performance. The demonstration plant operated in the frame of the ENREM-project was therefore object of further investigations addressing two main subjects in membrane development:

- The applied cleaning strategy for membranes has an enormous impact on the economical success of the containerized MBR system. This impact is even more severe for decentralized treatment plants as the cleaning process is always connected to man hours needed on site and downtime of the module. Additionally the need of chemical detergents during the cleaning raises the question of storage, handling and disposal in respect of safety and environmental issues.
- The demonstration and the prototype plants were embedded in a monitoring program on MBR fouling conducted by the Berlin Center of Competence for Water together with the Technical University of Berlin investigating sludge characteristics and fouling potentials.

It is obvious that these two research topics have to be seen in interconnection. Sophisticated fouling monitoring and control can reduce the necessity of high grade cleaning occasions. Vice versa a sufficient cleaning strategy might lead to less fouling events and higher availability of the installed membranes.

The installed modules and the overall filtration evolution are described in Section 4.1.

The use of two different cleaning detergents applied with a novel cleaning strategy suitable for decentralized wastewater treatment plant is shown in Section 4.2.

The outcomes of the MBR fouling monitoring program are briefly shown in Section 4.3.2. The detailed results of these investigations are presented by (De la Torre 2009).

Throughout the ENREM project, the membrane modules were replaced, leading to the nomenclature of rising numbers for the modules, e.g. module 1 – 5. In the following sections only the modules after replacement are presented and discussed, thus these modules were renamed and labeled as modules 1 and 2.

## 4.1 Filtration performance

Table 17 shows the key characteristics of the installed membrane modules and operational parameters.

**Table 17: Details of membrane and filtration parameter**

| | | |
|---|---|---|
| Reference | MX-020 (A3 water solutions) | |
| Surface area | 2 *15.9 $m^2$ | |
| Material | Polyvinylidenfluorid (PVDF) | |
| Pore size | 0.20 µm Micro filtration | |
| | ***Recommended by A3*** | ***As operated*** |
| Operational pressure difference | 20 - 300 mbar | < 100 mbar |
| Instantaneous flux | 15 – 25 L/($m^2$*h) | 15 – 20 L/ ($m^2$*h) |
| Net flux | / | 14 – 19 L/($m^2$*h) |
| t filtration / t relaxation in min | / | 12 / 3 |
| pH during cleaning | 2 - 11 | 2 – 11 |
| pH during operation | 5 - 9.5 | 7.2 – 8.1 |
| Temperature range | 1 - 50 °C | 9 - 27 °C |
| Cleaning agents | Base; oxidant; tenside, acid | $H_2O_2$, NaOCl, citric acid |
| Cleaning interval | 3 – 12 months high grade | monthly medium grade |

The plant was equipped with 2 parallel filtration reactors, each installed with two filtration modules and an autonomous filtration system. The two modules, manufactured by A3 Water Solutions (Germany) are assembled one upon another which leads to a reduced footprint of the membrane reactor. The height of the two modules plus aeration system is about 2.2m. Only one of these modules is in operation at a time, thus leading to two different permeability evolutions shown in Figure 40.

The permeabilities were calculated as follows:

The values for instantaneous flux and trans-membrane-pressure (TMP) are recorded as 2 h mean values. Only values recorded during filtration after flow stabilization are used to calculate the mean values, therefore the flux and the permeability can be calculated according to (Trussel *et al.* 2005):

$$J = \frac{Q_{Permeate}}{A_{Membrane}}$$

where J = membrane flux (L/$m^2$*h); $Q_{Permeate}$ = membrane permeate flow (L/h); and $A_{Membrane}$ = membrane surface ($m^2$).

The permeability L can be calculated through

$$L = \frac{J}{TMP}$$

where TMP = trans-membrane-pressure.(bar).

The permeability has been normalized to 20° C as described by (Trussel *et al.* 2005) using the following equation:

$$L^{20°C} = L * e^{(-0,0239\ (T-20))}$$

where T = water temperature (°C).

In addition clean water flux tests showed a pressure loss in the permeate system and before the pressure sensor due to turbulent flow, as the Reynolds number (Re) lies above 7500 for the applied fluxes.

The pressure loss due to a turbulent flow should be considered for the calculation of the permeability. The theoretical equation to determine the pressure loss in cylinder pipes is as follows:

$$\Delta p = \zeta * \frac{\rho}{2} * w^2 * \frac{L}{d}$$

where $\Delta p$ = pressure difference (Pa); ζ = drag coefficient; ρ = density ($kg/m^3$); w = velocity (m/s); L = pipe length (m); d = diameter (m).

As the drag coefficient is constant for high Re values and the ratio of length to diameter does not change, the following pressure corrections as a function of the flux and density were identified for the two modules:

Module 1:

$$\Delta p_{headloss} = 1.0368 * 10^{11} * \rho * J^2$$

Module 2:

$$\Delta p_{headloss} = 0.7776 * 10^{11} * \rho * J^2$$

where $\Delta p$ = pressure difference (Pa); ρ = density ($kg/m^3$); J = membrane flux ($L/(m^2*h)$).

This equation has been determined experimentally by recording the evolution of the TMP during a clean water test for different fluxes, as exemplary shown for module 1 in Figure 39. It was realized, that the two modules were installed differently, in respect of tube fitting and pipe connections, which had an impact on the real flow conditions thus on the pressure loss due to turbulent flow. Therefore different factors for each module are given above.

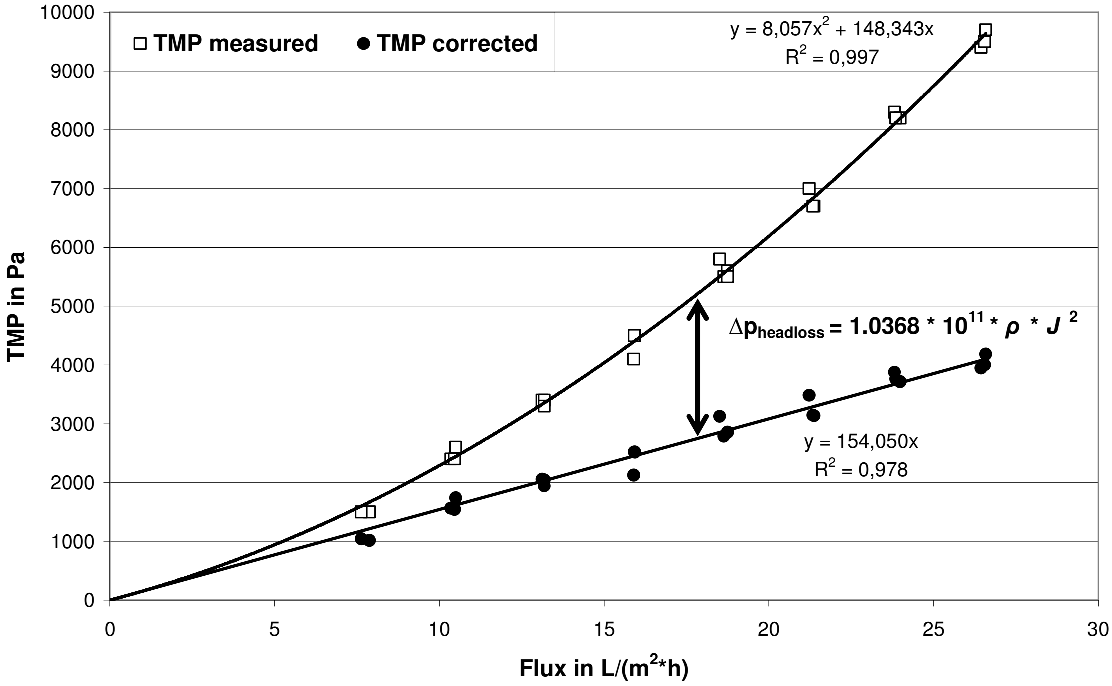



Figure 39: Head loss determination for module 1

The pressure difference has therefore been reduced by $\Delta p_{headloss}$ and used for the calculation of the normalized permeability, i.e. corrected for temperature and pressure:

$$TMP \quad = \quad \Delta p_{sensor} - \Delta p_{headloss}$$

The recorded pressure difference caused by the head loss can be as high as 60% of the total pressure difference. This shows the necessity to take the head loss into account when calculating the permeability for systems operating in the range of a turbulent flow.

Figure 40 shows the evolution of the permeability for both modules throughout the investigation. Module 1 was cleaned with $H_2O_2$ shown in blue, whereas module 2 was cleaned with NaOCl displayed in red.

The membrane modules were operated with an air scour between 0.6 and 0.9 $Nm^3/(h*m^2)$ except of the period between June and August 2007 as indicated. The reduced air scour was approximately 0.4 $Nm^3/(h*m^2)$. This reduction was expected to minimize energy costs for aeration, but the permeability of both modules decreased quickly. The permeability of module 1 decreased to an amount that was not tolerable therefore the air scour was increased again. Both modules recovered afterwards showing the strong influence of the air scour on the performance and discarding the possibility that inappropriate sludge characteristics led to the decline of permeability.

An event of heavy fouling is related to a reduction of the sludge filterability in November 2007, see results of the fouling monitoring program in Section 4.3. Module 1 was heavily affected by these conditions and only an intensive cleaning helped to recover the permeability, see Section 4.2.

The permeability evolution shown in Figure 40 demonstrates the stable operation throughout the investigated period of time. Lower permeabilities could be directly linked to insufficient operational conditions or low filterability of the activated sludge. The permeability sustained at a tolerable level and reliable operation was secured.

A sophisticated online-monitoring tool as Veolink, see Section 2.9, could help to avoid even periods of heavy fouling, indicating transient increase of the TMP during filtration. A rapid increase could be detected within few filtration cycles. Measures such as increased aeration, lower fluxes and increased relaxation time could be carried out manually or automatically. Additionally, the next planned cleaning could be adapted, i.e. higher concentrations or contact times implemented. This early response would prevent irreversible damage of the modules thus probably could increase the module life time and operation reliability.

## 4.2 Cleaning strategy

As described above, two of these double deck modules were assembled within the demonstration plant in two independent filtration reactors. Only one of these modules was in operation at a time, and the other was soaking in the cleaning solution, with a switch approximately every month. Therefore, two different cleaning agents could be compared for the soaking solution:

1. H2O2 used for module 1 with a concentration of 1000 ppm
2. NaOCl used for module 2 with a concentration of 500 ppm chlorine

For both agents the following cleaning procedure was applied:

1. After withdrawal of the activated sludge the module was rinsed with permeate. Strong aeration was exercised in order to detach remaining sludge. To increase this effect and to reach surface areas assumed to be less affected by aeration e.g. corners and edges, the membrane pockets were filled with permeate up to a maximum pressure of 50 mbar. Afterwards permeate was withdrawn from the membrane reactor, which was then filled with tap water.
2. The chemical agent was added to the membrane reactor. When cleaning with $H_2O_2$ the pH was increased to approximately 11 adding sodium hydroxide.
3. Filtration and aeration was performed for 15 minutes to ensure a well mixed reactor. Using NaOCl special attention had to be paid, as heavy foaming appeared, thus aeration was shortened according to foam production.

The module was soaked within the cleaning agent for one month until the next cycle of operation was started, with disappearance of the chemical agent with time (few days).

Occasionally (every 3-4 cleanings), an additional cleaning step with citric acid (5000 ppm, 1h) was performed before and after the main cleaning to attack the inorganic fouling. It has to be pointed out, that both chemical agents have to be of high quality, e.g. stored adequately as wrong storage way lead to reduced concentrations.

The efficiency of each cleaning was distinguished using the permeability values calculated during operation. As described above, each module was in operation for approximately one month. During this time a decrease of the permeability could usually be monitored and the efficiency of the cleaning could be calculated using following equation:

$$R = \frac{L_i - L_{i-1}}{L_0 - L_{i-1}} * 100$$

where R = recovery in percentage; $L_0$ = clean membrane in sludge at the beginning of operation: ~2,200 L/($m^2$*h*bar) for module 1 and ~ 1,600 L/($m^2$*h*bar) for module 2; $L_i$ = permeability after cleaning; $L_{i-1}$ = permeability at the end of the filtration run (before cleaning).

The average permeability value of two days at the start and the end of the filtration run was used to calculate the recovery rate. Figure 40 shows the permeability throughout the year, indicating the number of cleaning for each module. This way it was possible to compare the cleaning efficiency for the modules operated for the same period of time. The cleaning efficiencies are shown in Figure 41 by the recoveries.

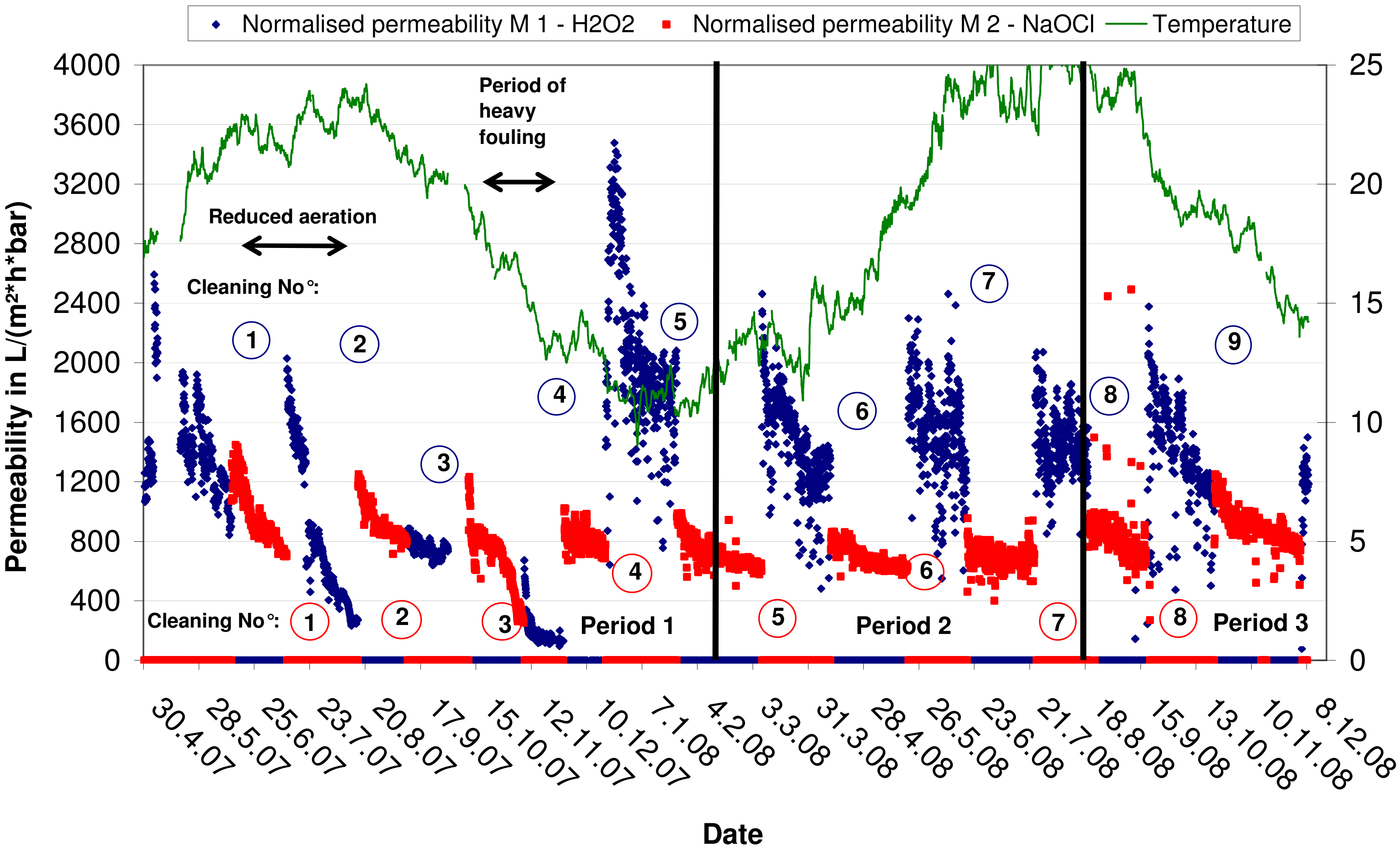



**Figure 40: Permeability evolution indicating the cleaning events**

The event of heavy fouling mentioned above is shown in November 2007. While module 2 was decommissioned during this period of fouling and the permeability could be recovered using citric acid before and after the cleaning step with a higher grade of NaOCl, module 1 was operated within this sludge for about four weeks. This caused a dramatic decrease of the permeability of this module, which could not be recovered only with the soaking with the planned $H_2O_2$ (1000 ppm) cleaning. That is why an intensive chemical cleaning was required. Using a higher grade of NaOCl (2000 ppm at ph 11) and citric acid (2000 ppm at pH 2) while increasing the contact time to 48 h for NaOCl and 24 h for citric acid finally recovered the permeability.

The recovered permeability was even higher than the initial one. This demonstrates that the intensive cleaning is a good possibility to recover a strongly fouled module, but might change the membrane characteristic in a way that could lead to a reduced life time of the module. However, this did not affect significantly the subsequent filtration behavior of this module, nor its disinfection performances.

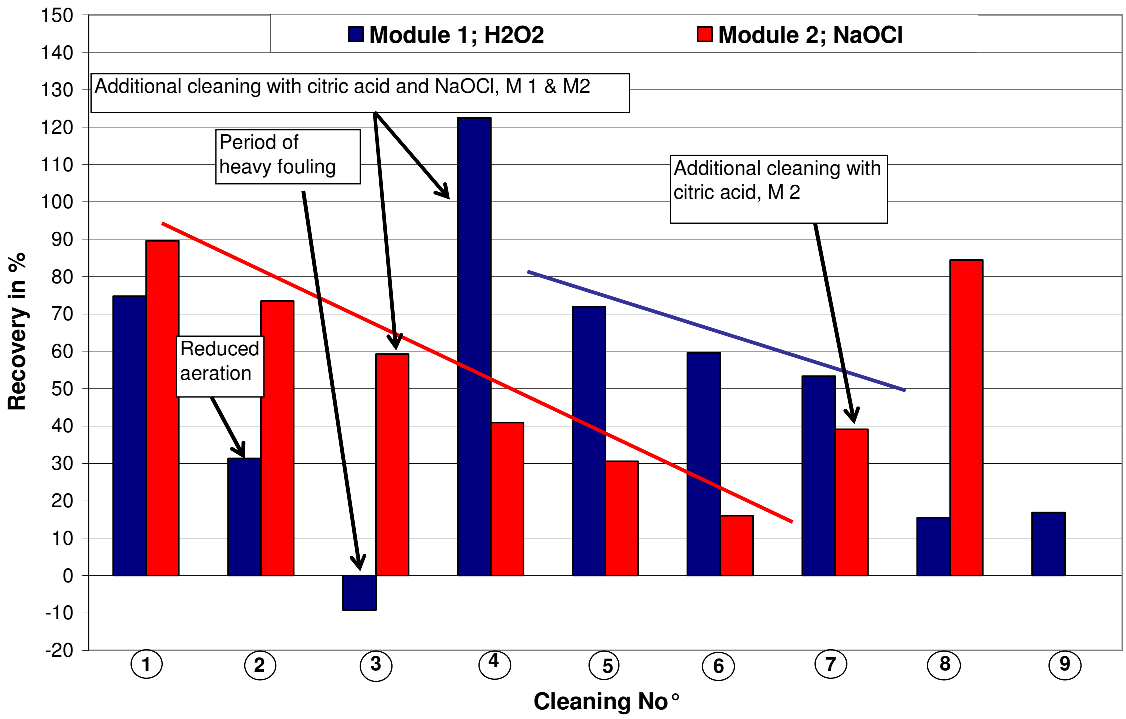



**Figure 41: Recovery of permeability after cleanings**

Figure 41 shows the recovery evolution for each monthly cleaning with both cleaning agents. Module 1 cleaned with $H_2O_2$ is displayed in dark blue and module 2 NaOCl in red. The numbers refer to the cleanings as indicated in Figure 40. The earlier discussed events of instable operation are also reflected in the cleaning results. When module 1 was commissioned during July 2007, the permeability fell due to the reduced air scour. The cleaning was not representative of the rest of the period with moderate fouling. Similarly, the event of heavy fouling in November 2007 led to an interfact in the calculation of the recovery: Cleaning N° 3 shows a negative recovery, due to the strong fouling within the first day of operation. As two day mean values are used for calculation of the permeability, a rapid decline within the first day of operation makes it impossible to determine a cleaning effect. As described above the bad performance during this fouling event led to an adapted cleaning protocol and the recovery of cleaning No° 4 was over 100 %.

The cleaning strategy included also the use of citric acids according to the performance of the filters (as indicated in Figure 41 for cleanings N° 3 and 7). Cleaning No° 7 shows the importance of citric acid, to remove inorganic foulants that might have accumulated over time and contribute to irreversible fouling.

As chlorine is considered in Germany and other countries to be hazardous to water bodies and decentralized waste water systems do usually not have the capacity to store toxic agents in a safe way, an alternative agent being easier and safer to handle showing the same cleaning results is an improvement in the field of decentralized waste water treatment with membranes. The results accumulated over these 22 months of operation show that $H_2O_2$ is a feasible option for semi-decentralized MBR plants, and confirms the experience reported by (Wedi *et al.* 2007) on the Monheim MBR plant (9,700 p.e.; 288 m3/h), cleaned with maintenance cleaning ($H_2O_2$ 2000 ppm; pH 9.7) on a 2-week basis. The cleaning with NaOCl should still be possible in case of heavy fouling events.

## 4.3 Investigations on fouling

Incorporated in the fouling monitoring program carried out for ten months in 2007 and 2008, sludge characteristics as well as operating conditions were investigated in order to define crucial parameters describing the fouling propensity. Combined with the collected operational experience this data enables to give valuable suggestions for MBR plant operation. Incorporating these guidelines in the operating instructions helps to reduce fouling events, thus increasing operational availability and decreasing expenses due to man hours and cleaning detergents.

Several indicators used for activated sludge characterization in wastewater treatment were investigated and evaluated. A screening of sludge indicators used in conventional waste water treatment plants was conducted in the attempt to define a universal fouling indicator. The detailed description of these trials can be found in (De la Torre 2009).

In the following sections the outcomes of the trials relevant for the ENREM demonstration and the prototype plant are further discussed.

### 4.3.1 Material and methods

#### 4.3.1.1 Critical flux measurements

(Field *et al.* 1995) introduced the theory of Critical Flux ($J_c$). He stated:

*The critical flux hypothesis for MF is that on start-up there exists a flux below which a decline of flux with time does not occur; above it fouling is observed. This flux is the critical flux and its value depends on the hydrodynamics and probably other variables.*

In order to measure this $J_c$ and its evolution, two methods were applied:

Berlin Filtration Method (BFM)

A mobile filtration unit consisting of a small UF membrane module was constructed in a compact way that allowed the measurements to be carried out on site of the MBR unit. This gave the opportunity to measure the $J_c$ within the plant, thus changing parameters due to transport and storage can be excluded. Figure 42 shows the set up of the test cell and the applied flux-step protocol (see (De la Torre *et al.* 2008)). This protocol gives the most information regarding the TMP evolution and irreversible fouling.

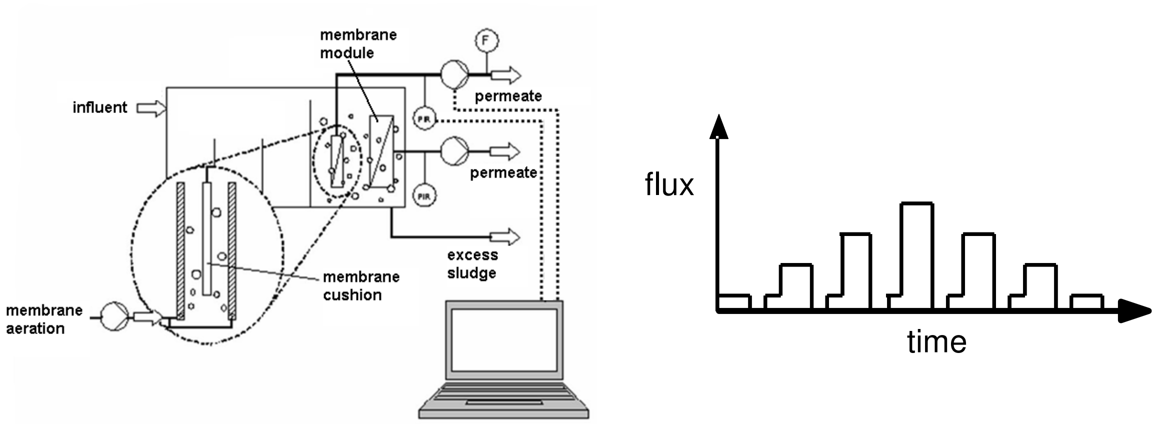



**Figure 42: Scheme of the *in situ* BFM test cell and $J_c$ pre-step protocol**

The pre-step lasted 2 minutes followed by 5 minutes of filtration and another 2 minutes for relaxation. An increase of 3 Lmh for each step was applied.

As this test cell was constructed in a compact way, it fitted easily in both MBR units introduced in this report. The membranes of this test-cell could be replaced and cleaned easily therefore it was possible to characterize the sludge filterability independently from the effects of long term operation.

$J_c$ of the installed modules

The demonstration plant gave the opportunity to investigate the $J_c$ evolution with the installed modules. Therefore the air scour was reduced to 5 $Nm^3/h$ and a flux step protocol similar to the pre-step protocol mentioned above was applied. The only difference was an additional flux step after the investigated one (see Figure 43). During these measurements the plant operation was influenced in terms of the achieved throughput, therefore the measurements were conducted without reducing the flux again. The pre-step lasted one minute with ~ 8 Lmh, followed by 5 minutes of filtration, 1 minute of the post-step at ~ 5 Lmh and 3 minutes of relaxation. The flux was also increased by steps of approximately 3 Lmh.

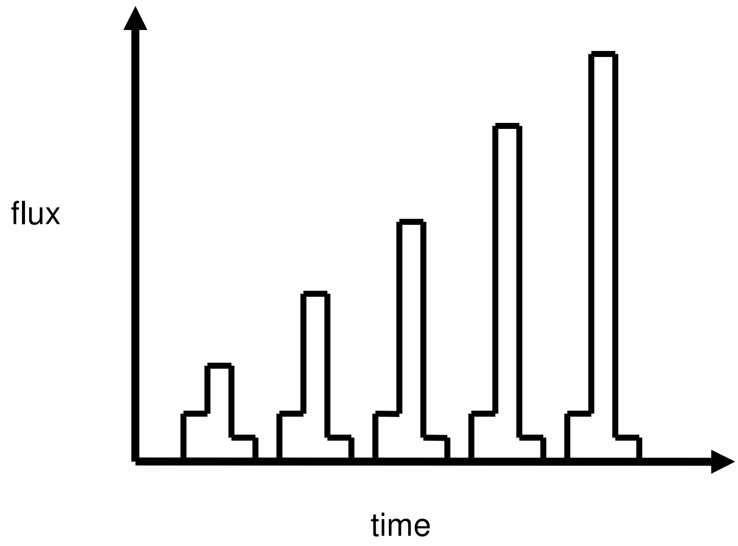



**Figure 43: $J_c$ protocol applied for the installed modules**

4.3.1.2 Sludge characterization

Several indicators describing the sludge characteristics are known and used in conventional wastewater treatment plants. In the quest for an universal fouling indicator these parameters were measured and correlated to the $J_c$ by (De la Torre 2009).

Physical measurements

The capillary suction time (CST) was measured using a standard device of Triton Electronics.

Another fast and easy way to determine the filterability of activated sludge is to measure the time needed to collect a specific amount of filtrate through a paper filter. Time to filter (TTF) was measured as the time to filter in dead end filtration 25 ml out of 250 ml activated sludge through a black ribbon filter paper (Whatman, pore size between 12 and 25 µm, diameter 90 mm).

The COD of the collected filtrate (supernatant) was measured using Dr. Hach-Lange test kits complying with ISO 8466-1, DIN 38402 A51 and DIN 32645 in calibration, detection and quantization limits.

The sludge volume index (SVI) was measured diluting the activated sludge in a ratio of 1:5 with permeate and using 2 L Imhoff funnels to determine the SVI. Due to floating and non-settling of the MBR-sludge, this parameter was excluded from this campaign later on.

Chemical measurements

*Transparent Exopolymer Particles* (TEP) was identified to be a possible indicator for fouling propensities. The following measurement protocol was kindly provided by (De la Torre 2009):

Before the analysis, the mixed liquor supernatant was obtained by filtration. For that purpose, filter papers (Schleicher and Schuell / Whatman, black ribbon Ø 90 mm, Germany) were rinsed with 200 mL deionized water. After that, 50 mL of sample was filtered to obtain the filtered mixed liquor and filtered influent. Duplicates of all analytical measurements were performed in order to minimize the random error. The analysis method used for the determination of the TEP concentrations (De la Torre *et al.* 2009b) is based on the protocol developed for TEP quantification in sea water (Arruda *et al.* 2004). The former consists of mixing 5 mL of pre-filtered sample with 0.5 mL of 0.055% (m/v) alcian blue solution and 4.5 mL of 0.2 mol/L acetate buffer solution (pH 4) in a flask. The flask is then stirred for 1 min and then centrifuged (Centrifuge MR23i Jouan GmbH, Germany) at 15,300 rpm (23,292 x g) for 10 min. TEP react with the alcian blue solution yielding a low solubility dye–TEP complex. The concentration of the alcian blue in excess is determined by reading the absorbance at 602 nm (UV-vis spectrophotometer, Analytic Jena, Germany). Xanthan gum is used for the calibration, and the results expressed in mg/L xanthan gum equivalent.

Both soluble and bound *Extracellular Polymeric Substances* (EPS) were analysed. The bound EPS were extracted using a cation ion exchange resin after the method described by (Frølund *et al.* 1996). After extraction, bound EPS measurement followed the same procedure as SMP, using (Dubois *et al.* 1956) for the determination of sugars and (Frølund *et al.* 1995) for the determination of proteins. The sum of these two terms was taken as total EPS. The polysaccharides were corrected for Nitrate and Nitrite as it is described in (Drews *et al.* 2007) using the formula:

$$C_{PS\ corrected} = C_{PS} - 0.099\ C_{NO3} - 1.99\ C\ N_{O2}$$

Where $C_{PS\ corrected}$ is the concentration of PS corrected, $C_{PS}$ is the concentration without the correction, and $C_{NO3}$ and $C_{NO2}$ are the concentrations of nitrate and nitrite in the paper filtered activated sludge, respectively.

### 4.3.2 Results and discussion

#### 4.3.2.1 Critical flux measurements and filtration resistance evolution

The measurements according to the BFM were carried out on a weekly basis by students of the TU Berlin under the supervision of Mrs. de la Torre. The evolution of the temperature corrected $J_c$ obtained with the BFM is shown for both plants in Figure 44.

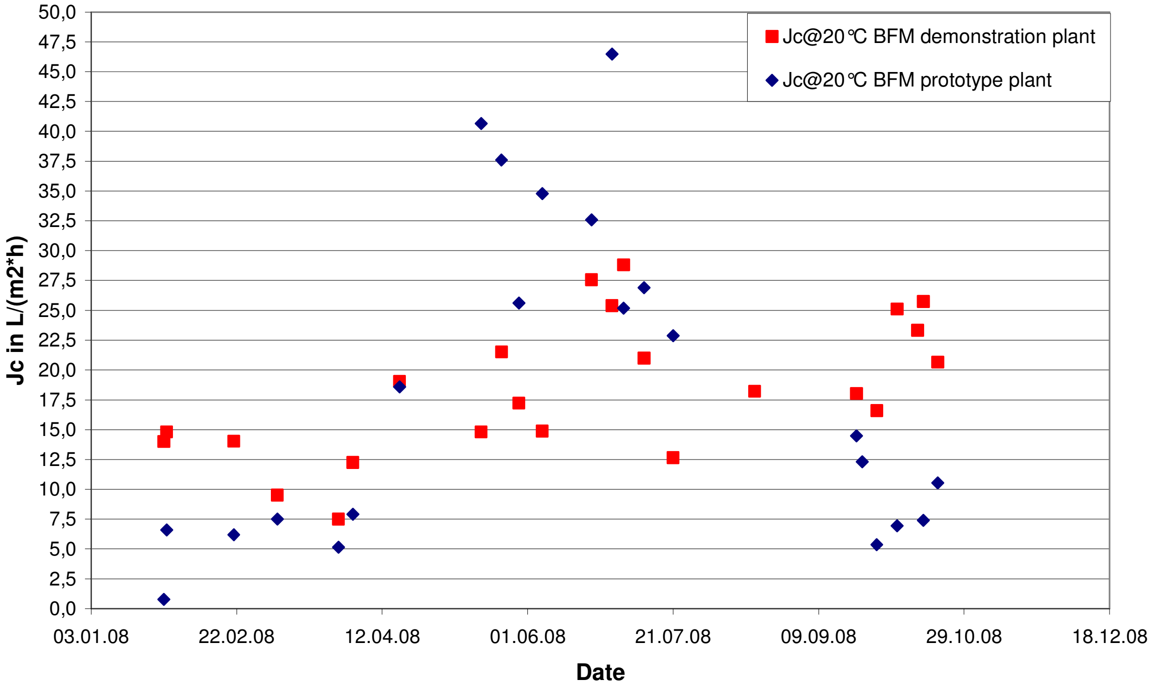



**Figure 44: Jc@20°C obtained with the BFM for both plants**

The $J_c$ for the demonstration plant was most of the time between 10 and 25 L/($m^2$*h) but no significant correlation to the permeability data gathered online during operation could be determined.

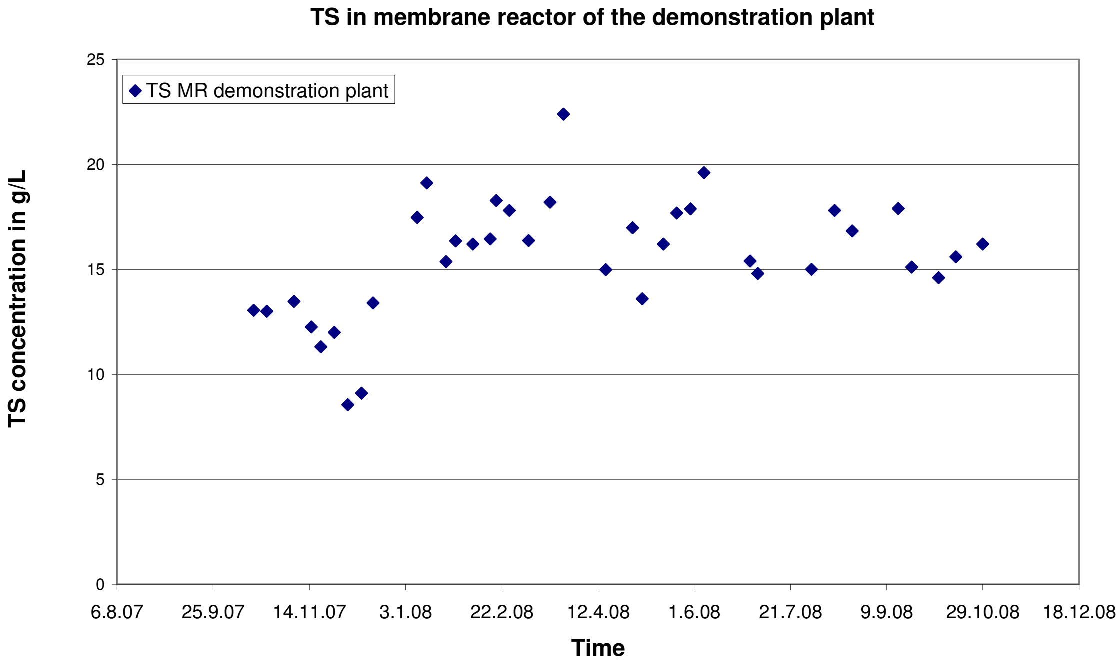



**Figure 45: Evolution of TS in the membrane reactor of the demonstration plant**

The results for the prototype plant were more scattered. Two fouling incidents during this measurement campaign might help to explain these results:

In February 2008, just after commissioning, incomplete nitrification led to foaming and heavy fouling. Afterwards the plant was commissioned again and the good filterability is shown by the high values identified for the $J_c$. The same reason, incomplete nitrification, led to the second fouling incident in September 08, indicated with a critical flux as low as 5 L/($m^2$*h). Even though the incomplete nitrification was recognized at an early stage, and the throughput was reduced as a response, the filterability needed ~ 4 weeks to recover. Only a complete halt of operation for 24 h and a slow commissioning afterwards helped to recover the system, see Chapter 5.

The critical flux obtained by the measurements with the installed modules has to be evaluated differently than those identified with the in situ test cell used for the BFM. In opposite to the BFM the installed membrane modules were analyzed directly, thus the results reflect both, the sludge characteristic at that moment as well as the history of the membrane module in operation. Figure 46 shows the evolution of the measured critical fluxes corrected to 20 °C. Due to the use of the installed modules and filtration pumps, the maximum flux was limited. Therefore in some of the measurements no critical flux could be determined. These experiments are indicated with an arrow.

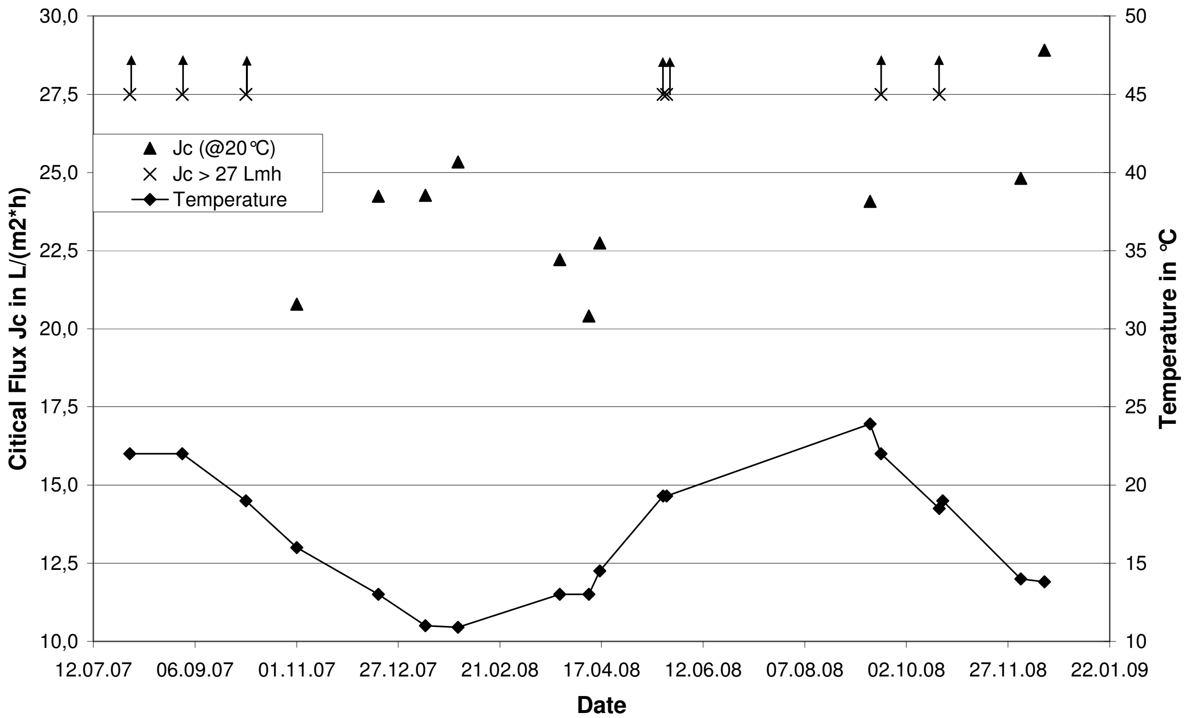



**Figure 46: Critical flux measurements (installed modules)**

The obtained $J_c$ with the modules installed in the demonstration are generally higher than those measured with the test cell, see Figure 47, which can be explained by the different set up, e.g. place of measurement or aeration intensity. Only the period when both measurements took place is shown.

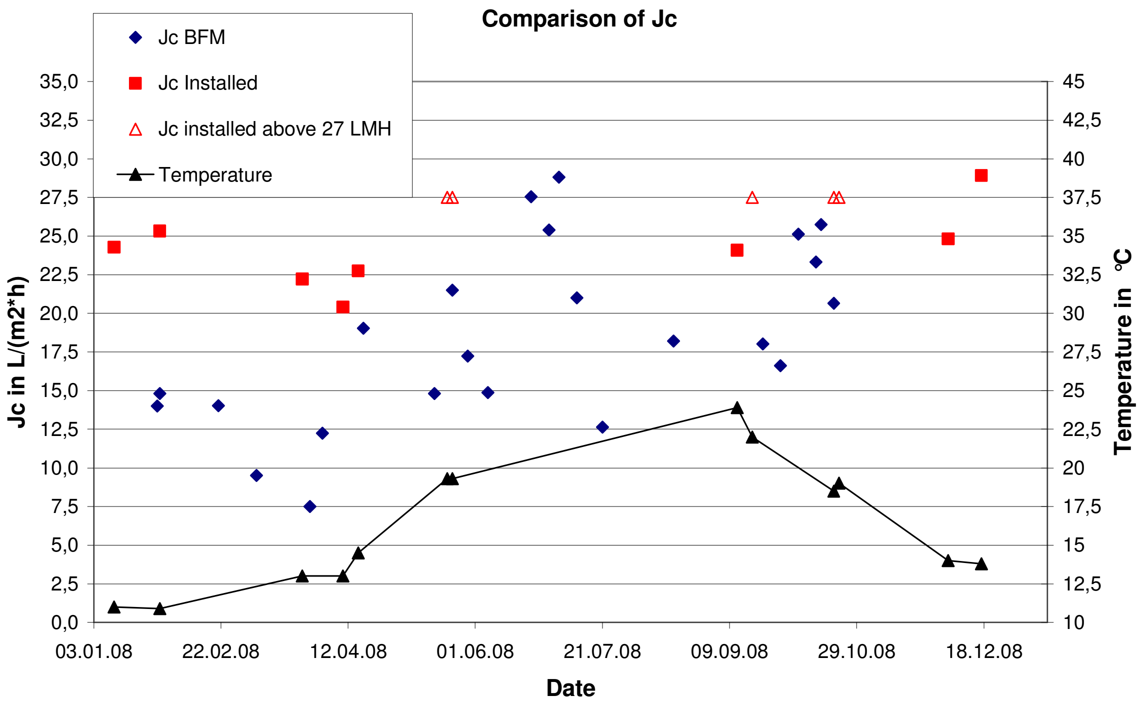



**Figure 47: Comparison of Jc obtained with BFM and installed modules**

Nevertheless, the trend for higher $J_c$ during warmer temperatures in summer and lower $J_c$ with colder temperatures in winter is shown by both measurement methods. The viscosity was corrected to 20 °C using the equation described in Section 4.1. The results obtained in the monitoring program indicate therefore that the temperature effect on the filtration is not only due to the change of viscosity but is an intrinsic characteristic of the sludge. As discussed in (De la Torre 2009) the temperature probably impacts the flocculation patterns of the mixed liquor.

The filtration resistance evolution was investigated using the data collected online. The total resistance, the reversible and irreversible fouling rate was calculated in the period of 14$^{th}$ of October 2007 till the 4$^{th}$ of December 2008, see Figure 48. The irreversible fouling rate is calculated over one day, which can explain the negative values: The resistance might decrease over two days, even though the rise in the long term is obvious. The accuracy of the pressure signal has an enormous impact on the calculated values. The reversible fouling rate is calculated per hour, thus the sensibility is a lot higher, as the displayed rates are within the same dimension.

At the same time the resistance increases very slowly over the period of operation (4 - 6 weeks), which explains the small value of the irreversible fouling rate. The irreversible fouling rate as the slope of the resistance over one day only shows a significant value during the period of heavy fouling in Nov. 2007. The maximum irreversible fouling rate was with 10 E(-11) /m/d twenty times higher than the rate during stable operation.

The reversible fouling rate accounted for 95 – 99 % of the overall fouling rate during the fouling event and it is therefore recommended to monitor this value online in order to detect changing sludge propensities. During stable operation, the reversible fouling rate was zero, which shows also the advantages of this parameter in terms of online fouling control.

The TMP slope in each filtration cycle was also calculated for different periods of operation and proved to be a sensible parameter for fouling events. This parameter is a valuable benefit, if sophisticated operational software such as Veolink® monitors the plant. Immediate or even automized action can help to protect the membranes and might help to reduce the costs for cleaning detergents and increase operational safety.

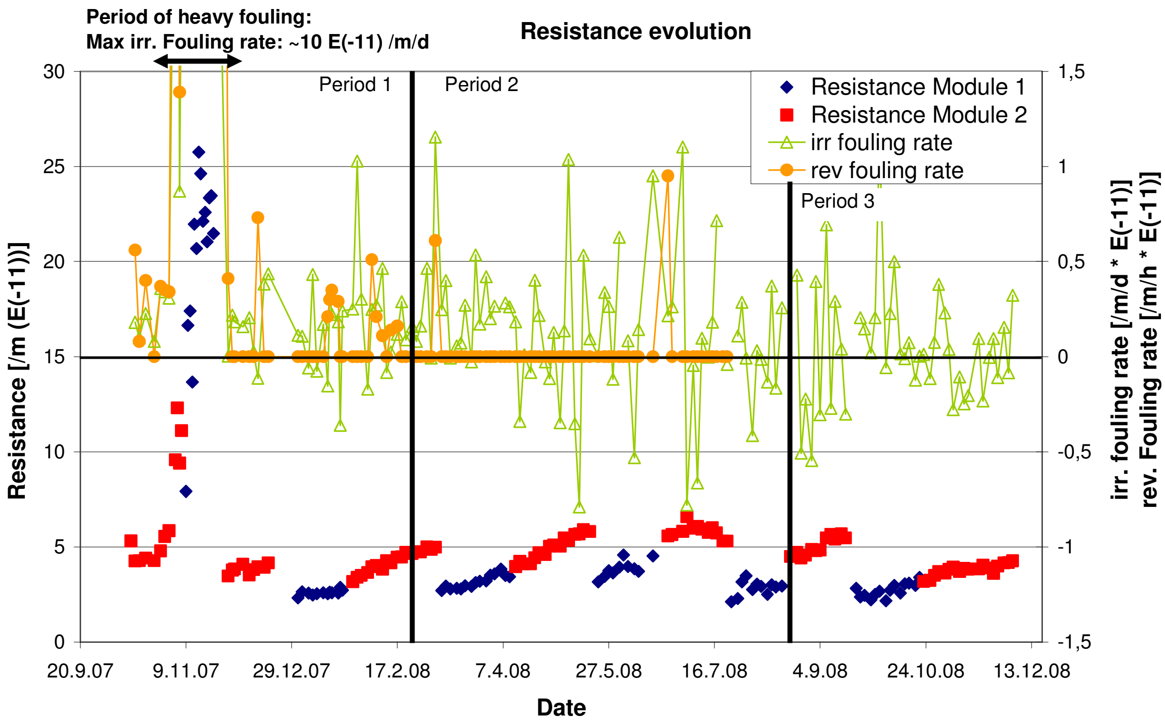



**Figure 48: Evolution of the filtration resistance, irr. and rev. fouling rate for the demonstration plant**

4.3.2.2 Time to Filter

The measurement of the TTF showed to be an easy, quick and cheap method to characterise the activated sludge. The evolution of TTF values are shown for both plants in Figure 49 and Figure 50. Fouling events of the prototype plant are indicated in the figure to show the correlation between TTF measurements and the filterability.

The correlation of $J_c$ and TTF for both plants is shown in Figure 51.

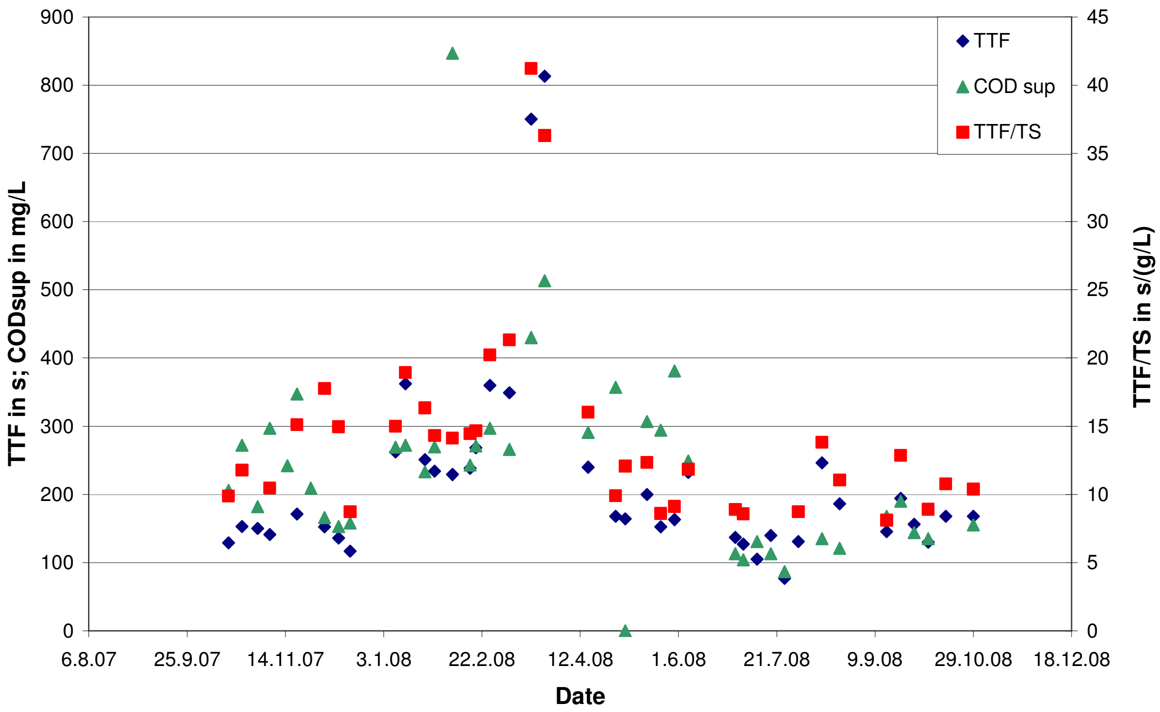



**Figure 49: TTF, CODsup and TTF/TS for the demonstration plant**

The benefit of TTF measurements can be better demonstrated with the data collected for the prototype plant shown in Figure 50. Two events of heavy fouling were registered during operation, the first in February and the second in September 2008. Both events are clearly indicated by the high values for TTF and CODsup. Consequently the permeability dropped for the first incident, as no measures were implied due to a late recognition of the disturbed system, see Figure 66. The second event of heavy fouling has been recognized earlier because of TTF monitoring, thus the throughput was reduced by shortening the filtration time which showed a good effect in respect to the permeability which did not decrease as drastically as in the first fouling incident. The taken measures are further explained in Section 5.1.2.

During stable operation the TTF value did not vary more than 20% compared to the value measured a week before. The slight changes in TTF can be explained by higher TS-concentrations. When a change of more than 30% occurred in relation to the previous week, further measurements have been applied such as determination of ammonia and nitrite concentrations, which might indicate incomplete nitrification due to poor aeration or shock load of ammonia.

Applying the TTF measurements on a daily or every other day basis, a change in the fouling propensities is recognized very early and active measures can be implemented reducing the fouling effects on the membranes. Therefore collecting the TTF values as an operational parameter periodically can help to improve and stabilize the operation.

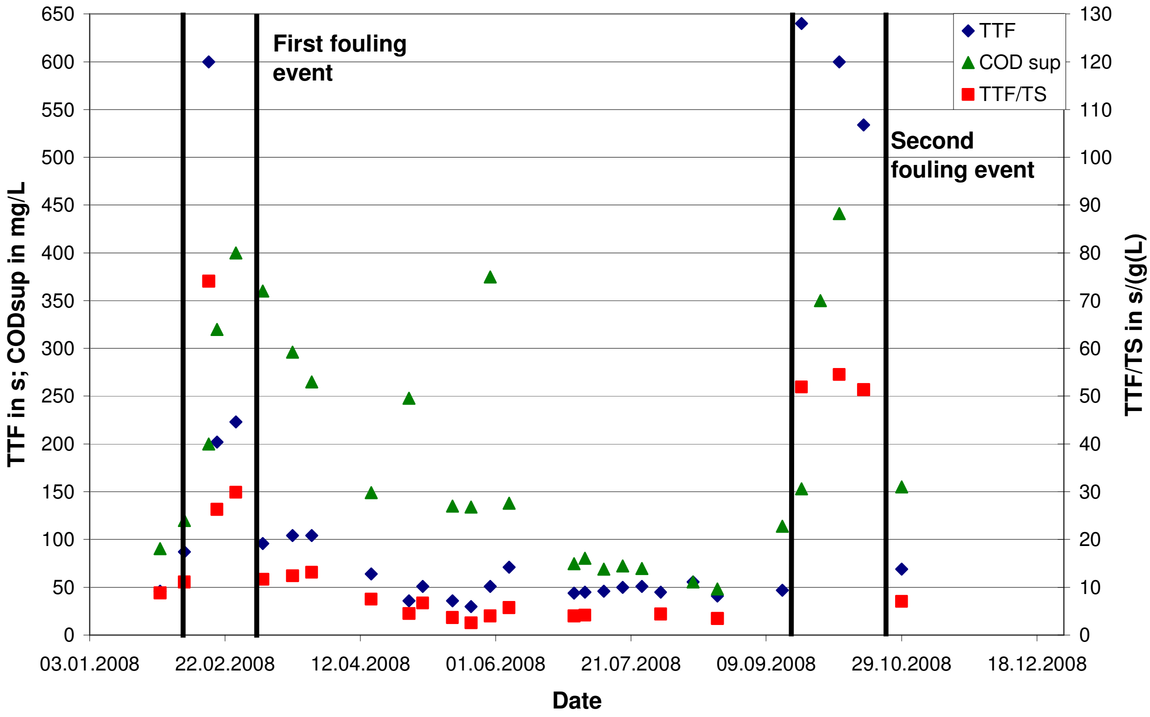



**Figure 50: TTF, CODsup and TTF/TS for the prototype plant**

Additionally to the TTF measurements the COD of the supernatant was analyzed and in both fouling events a retarded reaction in correlation to the TTF measurements can be seen, see Figure 50. Also, the CODsup values did decline slower than the TTF values, once the filterability was recovered. Therefore the TTF measurements seem to be the more sensible data regarding the fouling propensities. The CODsup is valuable information regarding the biological performance, but to be able to act early in case of increasing fouling propensities, the TTF measurements are recommended.

Figure 51 shows the relation of $J_c$ to TTF for both plants. A logarithmic trend line is plotted showing a good correlation between the two parameters for the two units, except of two runaway points which were measured at commissioning of the prototype plant with low TS. TTF seems therefore be a good indicator for mixed liquor filterability in the TS range of 10 g/L and it is recommended to perform this simple and quick analysis on a regular basis in the operation of decentralized MBR systems.

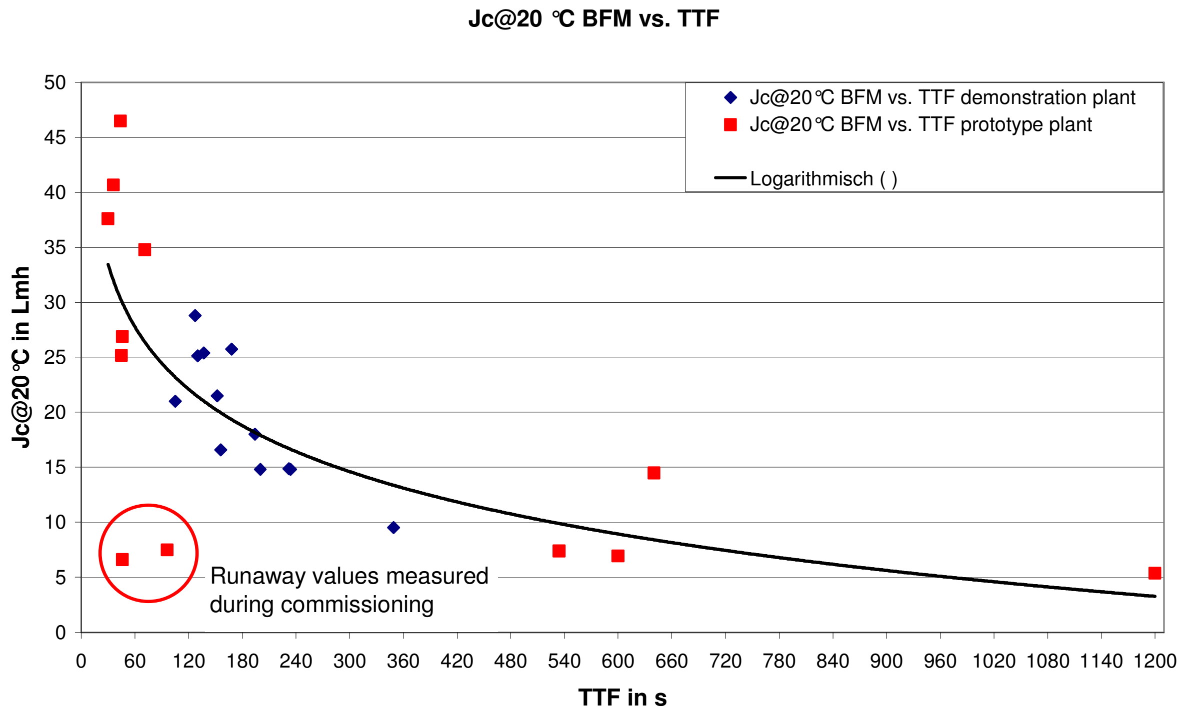



**Figure 51: Critical flux in relation to TTF for both plants**

4.3.2.3 EPS measurements

The parallel operation of the prototype plant gave the possibility to compare the EPS formation of two plants under similar operation conditions applying different biological processes. Various investigations on EPS formation according to operational conditions showed numerous influences, such as dissolved organic carbon (DOC), solid retention time (SRT), organic loading rate, substrate type and COD:N:P ratio, temperature, stress situation and reactor type (Lu *et al.* 2001; Rosenberger *et al.* 2006). Incorporated in the fouling monitoring program, EPS was measured as the sum of polysaccharides (PS) and proteins (P) in the demonstration and the prototype plant. Both, the soluble (sEPS) and the bound EPS (bEPS) fractions were determined.

The period from April 22$^{nd}$ till August 1$^{st}$ 2008, when stable operation was achieved in both plants, was compared. During this time the measurements were carried out on the same day giving the opportunity to investigate the EPS concentration with identical influent constitution and temperature. Due to operating condition not all parameters were exactly the same. Table 18 gives some important parameters describing the way of operation.

**Table 18: Plant parameters**

| | HRT in h | SRT in d | Organic load in kgCOD/(kgTS*d) | Total solids in g/L |
|---|---|---|---|---|
| Demonstration plant | 19.9 – 16.6 (wd – we) | 20 – 50 | 0.083 – 0.118 | 12 – 17 |
| Prototype plant | 48 – 34 | ~ 35 | 0.077 | 7.8 – 12.8 |

Table 19 shows the values obtained for both plants. bEPS is also correlated to the TS in order to be able to compare the results. The outcomes give an indication on the influence of the biological process on EPS formation.

**Table 19: EPS values for both plants**

| | Mean sEPS in mg/L / No° of values | Mean bEPS in mg/L / No° of values | Mean bEPS/TS in mg/g / No° of values |
|---|---|---|---|
| Demonstration plant | 51.1 / 10 | 770.3 / 10 | 53.3 / 7 |
| Prototype plant | 25.8 / 10 | 358.3 / 10 | 36.1 / 7 |

The values of sEPS, bEPS and bEPS/TS for the demonstration plant with EBPR are significantly higher than those obtained in the prototype plant. Twice as high absolute values can be explained by the generally higher concentration of total solids in the demonstration plant, but even the specific concentration of bEPS is 1.5 as much as those of the prototype plant.

A decisive difference between the two plants was the biological process applied. The demonstration plant combined the biological phosphorus removal with post denitrification as described above. The prototype plant was designed as a sequential batch membrane bioreactor (SBMBR), thus introducing the possibility of pre-denitrification. The implementation of an anaerobic stage in the demonstration plant led to a distinctive difference in the biocenosis of the plants. For instance, PAOs were most likely not present in decisive numbers in the prototype plant, as true anaerobic conditions did not take place under stable operation. Additionally different types of denitrifiers established themselves in the two systems. These differences in the biocenosis could lead to the significant higher values of EPS in the demonstration plant. The varying operational parameters such as organic load and hydraulic retention time are also known to play an important role on EPS formation.

Incorporated within the monitoring program the additional parameters of TEP was measured and also related to fouling affinity. The evolution of both parameters, EPS and TEP, for the demonstration plant is shown in Figure 52, subdivided in sEPS, bEPS, sTEP and bTEP.

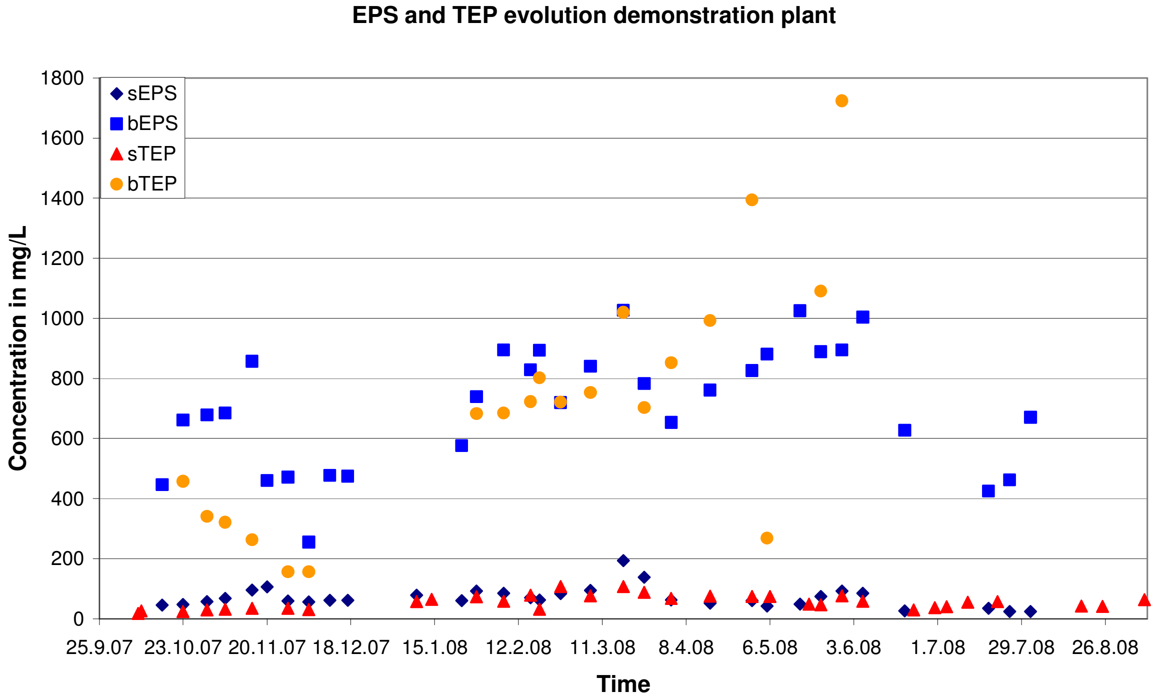



**Figure 52: EPS and TEP evolution of the demonstration plant throughout the monitoring program**

Interpretation of these parameters becomes more obvious looking on the ratios of bEPS/sEPS and bTEP/sTEP as depicted in Figure 53. The fouling event in November 2007, as discussed in 4.1, is clearly indicated by the decline of both ratios. Afterwards the ratios recover slowly when filterability was recovered quickly, thus more attention should be paid on the evolution of these parameters than on the absolute number, which is typical for each plant and its operational environment. The ratios represent the flocculation-deflocculation kinetics and emphasize the importance of this dynamic parameter in the quest of a full description of the fouling phenomena.

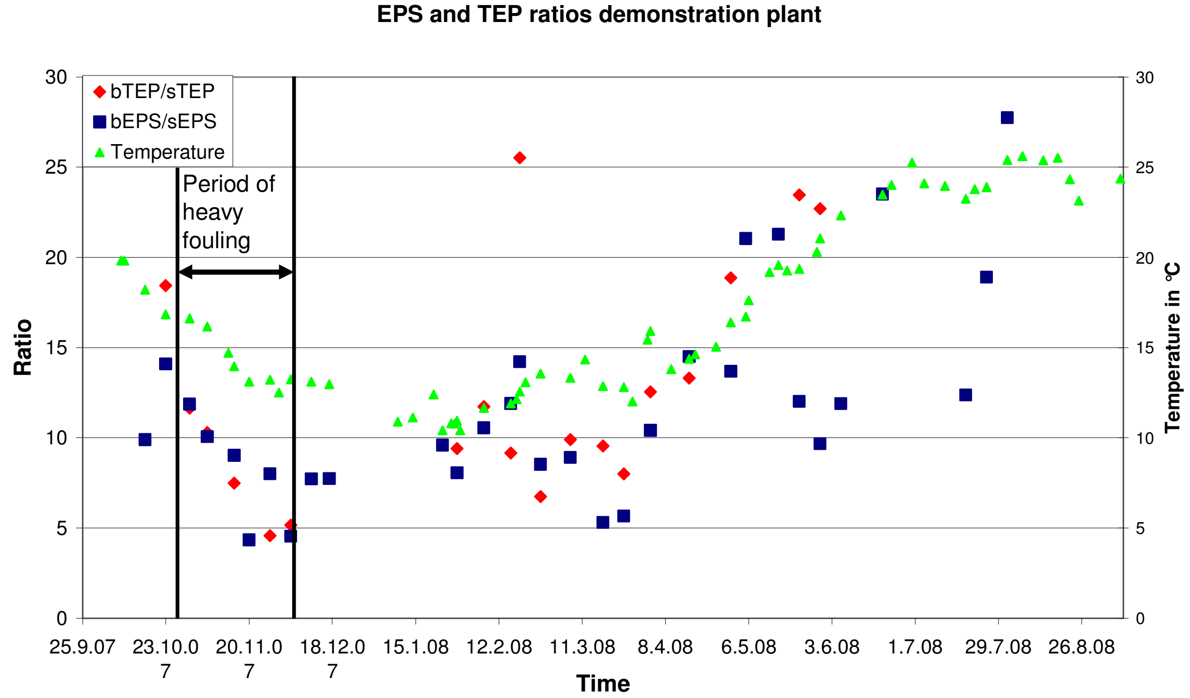



**Figure 53: Ratios of EPS and TEP in the demonstration plant**

*DNR*

The high bEPS concentration shows the possibility of bEPS being used as the carbon source for post-denitrification.

These presented values were measured with sludge samples from the membrane chamber only, thus no direct relation to the usage for denitrification can be drawn. This high amount of carbon available by EPS formation supports this hypothesis. The evolution of bEPS within the demonstration plant and during batch test experiments should be carried out to prove this hypothesis. Nevertheless, the usage of EPS as a carbon source for denitrification can not explain the observed correlation between F/M and DNR.

#### 4.3.2.4 Liquid Size Exclusion Chromatography – Organic Carbon Detection (LC/SEC/OCD)

Samples of the influent, sludge supernatant and permeate of the demonstration plant were measured in regard of organic carbon fractioned present in the process. These measurements were carried out at and with the kind support of Anjou Recherché in Paris. Five samples were analyzed in 2007 and 2008. For each sample the DOC concentration and fractions were analyzed for the raw wastewater, the supernatant of the activated sludge and permeate. Raw wastewater and sludge were paper filtered with black ribbon paper filter.

*DOC removal*

Figure 54 shows the DOC concentrations for the samples as well as the removal efficiency. Striking is the decline of the DOC concentration with time. As the effluent concentration stays within the same range, the removal efficiency decreases synchronically. The significant decline of the influent concentration, 142 mg C/L in July 2007 to 37 mg C/L in May 2008, can not only be explained by further dilution of the incoming wastewater with time. The dilution is shown by the mean concentration of COD, TN and TP in Table 5. (Rosenberger *et al.* 2006) showed that the DOC measured in the effluent of MBR systems consists most likely of organic matter that was produced by the biomass itself rather than fractions that were present in the raw wastewater. Therefore the stable effluent concentrations can be explained although the influent concentrations declined.

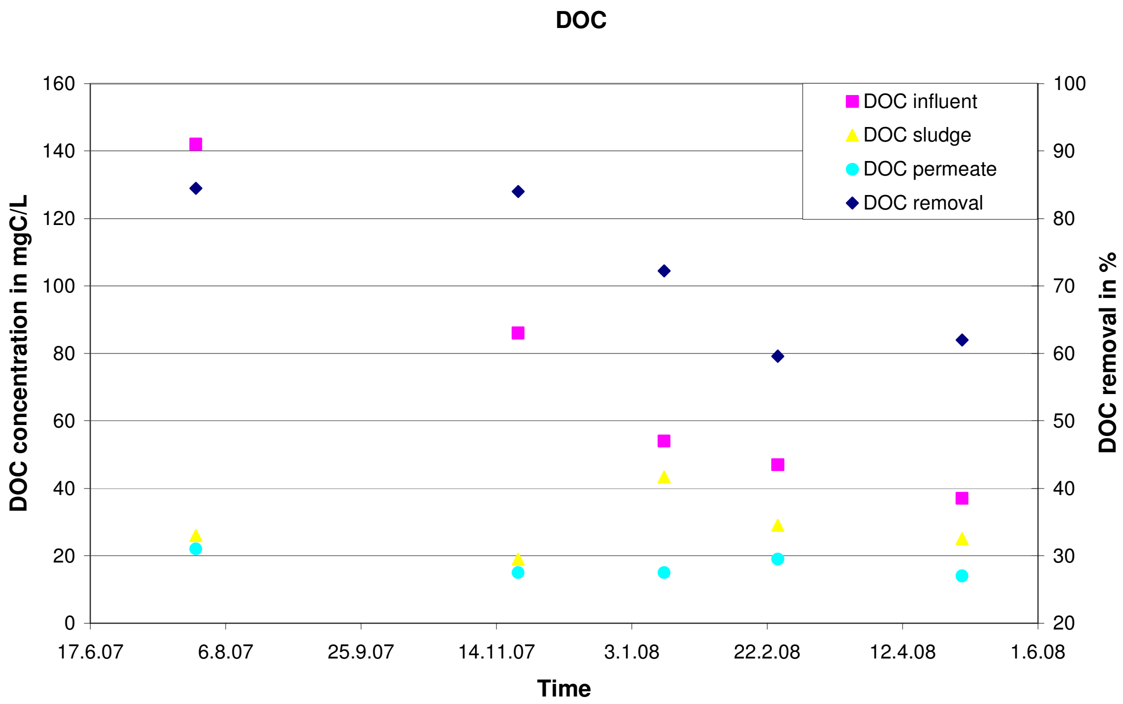



**Figure 54: DOC concentrations and removal efficiency**

The following diagrams show the fractions of DOC for the influent, sludge supernatant and permeate.

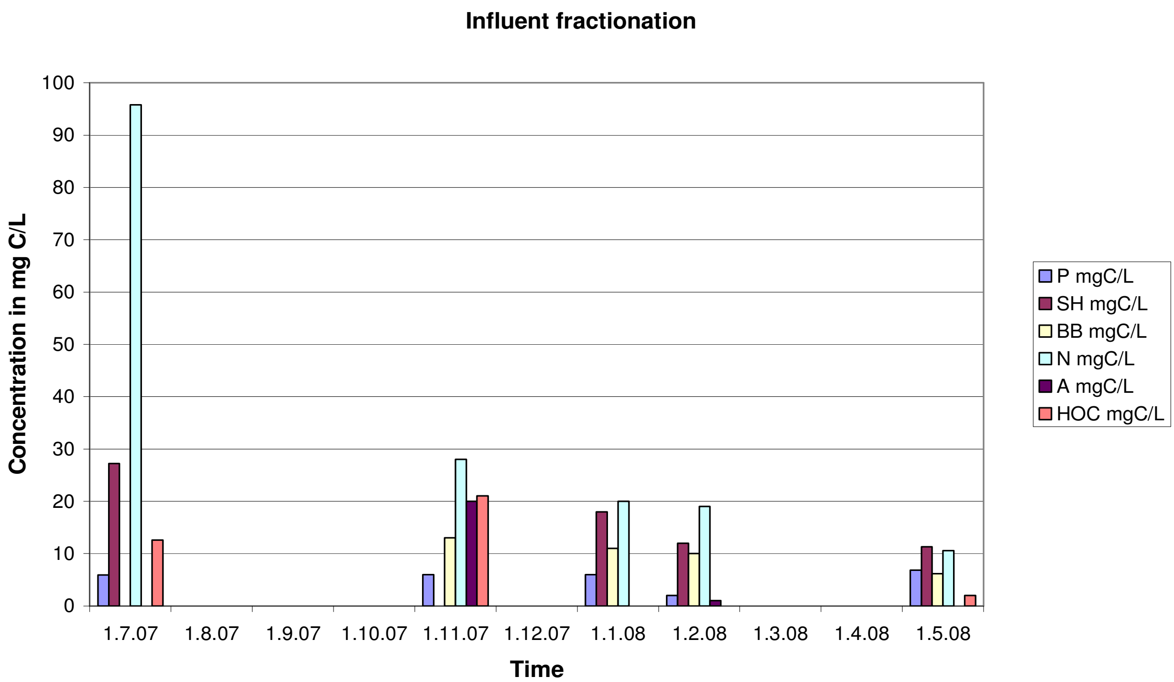



**Figure 55: Influent DOC fractionation**

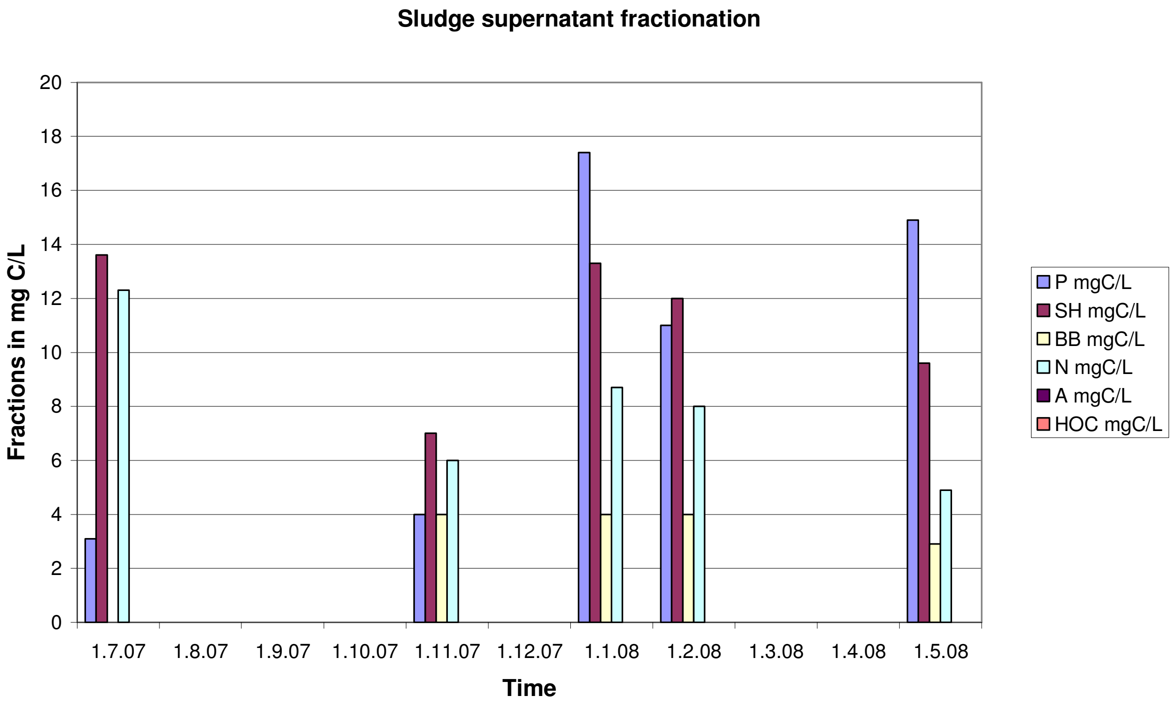



**Figure 56: Sludge supernatant DOC fractionation**

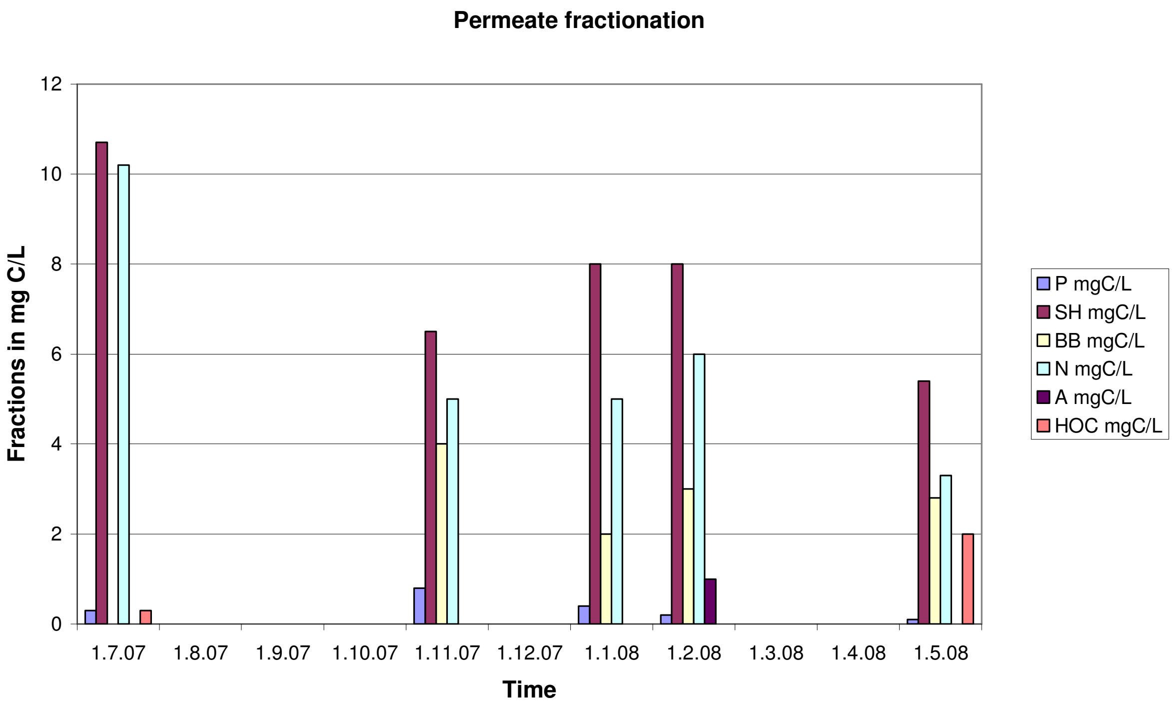



**Figure 57: Permeate DOC fractionation**

The abbreviations are explained in Table 20.

**Table 20: Abbreviations of DOC fractionation**

| Abbreviation | Meaning |
|---|---|
| P | Polysaccharides, Proteins, Aminosugars, Colloids |
| SH | Humic substance |
| BB | Building blocks – mostly breakdown products of humics |
| N | Neutrals |
| A | Acids – Summary value for for monoprotic organic acids < 350 Da |
| HOC | Hydrophobic part of organic carbon |

4.3.2.5 Conclusions

Membrane fouling reduces the efficiency and availability of MBR systems due to reduced fluxes and increased cleaning efforts. The results of the prior presented investigations suggest implementing an online tool, e.g. Veolink®, identifying changes in sludge filterability quickly. Due to the online availability of the crucial parameters, the plant operation can be adjusted and an alarm can inform the operators. Especially for decentralized treatment plants the possibility to inform the operators with the first indicators of reduced filterability is an important and helpful tool. At the same time, a high automized plant can run different procedures to prevent damaging the membrane reducing the effort to recover the membrane once it is fouled. For instance membrane aeration can be increased to reduce cake formation or the throughput can be adapted by implementing longer relaxation periods.

Time to filter (TTF) showed to be an easy, quick and cheap way to characterize the sludge filterability for TS concentration > 10 g/L and it is recommended to perform this analysis once or twice a week in order to monitor the sludge characteristics. A significant change of TTF indicates a change in the biological process and further measurements, such as ammonia and nitrite concentrations or pH value, are recommended to identify the reason of this behavior, as e.g. incomplete nitrification is known to cause fouling.

According to the outcomes of the investigations presented in this report and also by others researchers, on fouling propensities considering sludge characteristics, flocculation-deflocculation kinetics seem to be an important aspect and should be investigated further (De la Torre *et al.* 2009a; van der Graaf *et al.* 2009). The impacts on these kinetics are not understood yet and a clear understanding why EPS/TEP is formed and released to the surrounding media could help to define operational guidelines. Microbiological investigations on this phenomenon are therefore recommended.

Further on the nature of fouling causing materials has to be identified in order to give operational as well as design guidelines and to be able develop cleaning detergents attacking these fractions of NOM. The source of the foulants has to be identified and actions preventing membrane blocking could be defined in a precise way.

# Chapter 5
# Parallel MBR prototype unit

High inflow volumes which exceeded the treatment capacity of the demonstration plant made it necessary to truck away the additional wastewater by the Berliner Wasserbetriebe. In order to reduce these extra costs, a second MBR unit was rented from the German company Busse IS. Designed for carbon removal and full nitrification, this plant was planned to treat up to 5 $m^3$ per day.

Technical descriptions as well as results of commissioning and operation of this prototype plant are given in section 5.1.

Even though this plant was designed for carbon removal and full nitrification only, the denitrification capacity of this system was investigated and optimized through process control adjustments. Furthermore the possibility of enhanced phosphorus removal applying the combination of precipitation and a downstream adsorption filter is presented in section 5.2.

Incorporated in a fouling monitoring program carried out by the Berlin Center of Competence for Water, the filterability evolution and the membrane performance was investigated, section 5.3.

The energy requirements of the prototype plant are described in section 5.4.

An economical evaluation is described in section 5.6.

## 5.1 Operation of prototype MBR plant

### 5.1.1 Plant description

The prototype plant is constructed within a standard cargo 20” container with two reactors, each providing approximately 2.5 $m^3$ of reaction volume. The feed enters the activated sludge reactor which is equipped with three aerators for oxygen supply and mixing. An overflow leads to the membrane reactor equipped with four Kubota M-Box membrane-modules combined for 20 $m^2$ of membrane surface. Recirculation is enforced with two airlift beet pumps. Aerators for membrane cleaning and oxygen supply are installed providing both fine and coarse bubbles. To assure permeate flow, a permeate pump is installed besides the possibility to use gravity flow.

The plant is equipped with one floater in the membrane chamber indicating the liquid level in the plant. Due to this *high/low* level process scheme the plant is designed as a *sequential batch membrane bioreactor (SBMBR).* The possibilities as well as the limits of this set up are further described in section 5.1.3.

Figure 58 shows a standard filtration cycle with a filtration time of 30 minutes followed by 25 minutes of relaxation. Due to the set up with a floater indicating *high* and *low level*, the reactor was filled twice or three times per filtration period with fresh wastewater. After optimization of the control scheme, the relaxation period took always place in the state of *high level*, see Section 5.1.3.

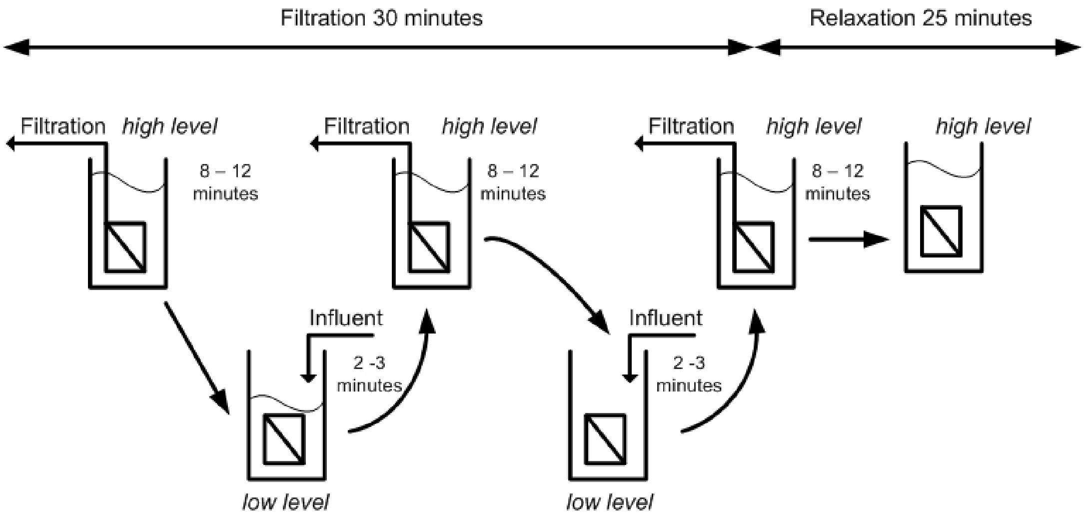



**Figure 58: Filtration cycle of prototype plant**

### 5.1.2 Commissioning and operation

Figure 59 shows the throughput within the investigated period of time. During commission and operation it became obvious that the biological capacity was the limiting factor. Due to the filtration capacity a higher throughput would have been possible, but as complete nitrification is arbitrary to meet the disposal limits, the throughput was adjusted according to the biological performance. When a stable operation was reached, the control scheme was changed to a week-weekend scheme, treating 3.5 $m^3/d$ and 4.5 $m^3/d$ on weekdays and weekends respectively. This increased throughput helped to handle the added wastewater during the weekends, when most residents spend more time at home. Therefore the prototype plant helped to reduce the costs for trucking and made it possible to operate the demonstration plant under stable conditions.

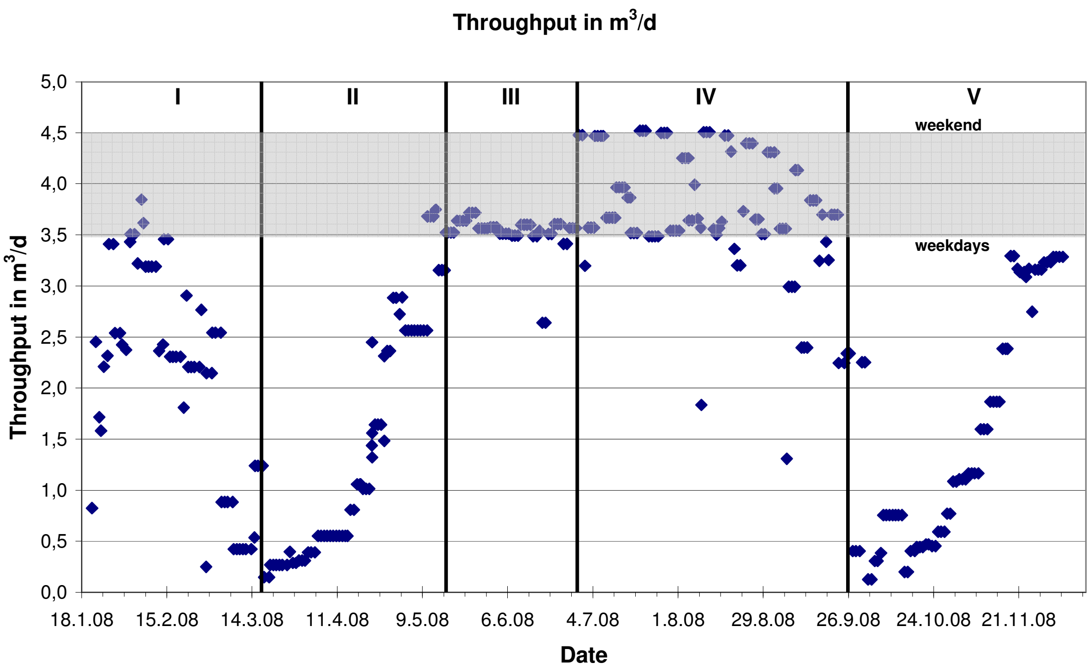



**Figure 59: Evolution of throughput of the prototype MBR unit over time**

Five phases of operation can be identified according to the throughput:

I. First commissioning 18.01.08 – 15.03.08
II. Second commissioning 15.03.08 – 13.05.08
III. Constant throughput of 3.5 $m^3$ per day 13.05.08 – 29.06.08
IV. Adjusted throughput week / weekend 24.06.08 – 18.09.08
V. Third commissioning 18.09.08 – 31.12.08

**Phase I**: First commissioning: 18.01.08 – 15.03.08

After a few days of foaming during commissioning due to a period of adaptation and stress for the biomass, the throughput could be increased to about 3.5 $m^3/d$. Foaming as a response to changing milieu conditions is a known and reported behavior of activated sludge systems (Stratton *et al.* 1998). The activated sludge used for inoculation was taken from the demonstration plant, which is operated with a different biological process scheme, and therefore explains the foaming.

COD and $NH_4$ elimination were sufficient during this first phase of operation and the TS concentration increased, as no excess sludge was withdrawn. Before the targeted TS concentration of 11.0 g/L was reached, heavy foaming appeared. At the same time the filterability of the sludge decreased and accumulation of nitrite was noticed. It can not be fully explained why this strong foaming and fouling appeared, but the demonstration plant showed the same behavior in terms of foaming at the same period, so it can be concluded that the influent contained a toxic substance.

A seeding with sludge from the demonstration plant, in order to enhance the nitrification/denitrification capacity and to recover the biological system, did not have any positive effects neither for nitrification nor for denitrification. In opposite, the seeding caused further foaming, similar to the period of foaming during first commissioning.

Fouling as a result of incomplete nitrification reduced the throughput, because at a certain TMP the low pressure valve opened and the TMP was held constant. In order to reduce the nitrite concentration, a complete interruption of operation was necessary. The plant was not operated for about 24 h during which aeration and filling did not take place. Because of this long anoxic period the biomass was able to denitrify the nitrate and nitrite content.

A slow commissioning following this stop of operation represents Phase II.

**Phase II**: Second commissioning: 15.03.08 – 13.05.08

During the second commissioning the throughput was increased by 25 – 35 % every 2 - 3 days, as long as the nitrite concentration was below 0.1 mg-N/l. Nitrite was the key parameter deciding whether the throughput could be increased or not. This approach showed good results and with higher throughputs the TS concentration rose to the targeted 11.0 g/L. Also the permeability recovered and achieved a sufficient value. The permeability results are further described in section 5.3.

**Phase III**: Constant throughput of 3.5 $m^3/d$: 13.05.08 – 24.06.08

A constant throughput of approximately 3.5 $m^3/d$ was reached since 13[th] of May. During this period, the elimination rates reached the expected range and first trials to increase nitrogen and phosphorus elimination were conducted.

Since the 2nd of June excess sludge was withdrawn according to the measured TS concentration. Approximately 1.0 $m^3$ was withdrawn throughout a week, leading to a sludge age of ~ 35 days. Due to this approach the TS concentration in the reactor could be held within a range of 8.5 – 12.5 g/L, see Figure 60.

**Phase IV**: Adjusted throughput week / weekend: 24.06.08 – 18.09.08

In order to achieve the main goal set for this prototype plant, treatment of additional waste water and reduction of the throughput of the demonstration plant, the throughput of the prototype plant was increased during the weekend.

Even though the TS concentration was kept stable, the higher temperatures and higher throughputs led to low oxygen concentrations. Therefore the used blowers (3 * 60 L/Min) in the activated sludge reactor were replaced by stronger ones (3 * 100 L/Min). In addition, the blowers installed in the membrane reactor were turned on permanently. Oxygen concentrations above 3 mg/L could be achieved in the membrane reactor, ensuring full nitrification. This aeration regime had the positive side-effect, that the membranes were aerated not only during filtration, but also during filtration break enhancing the cleaning effect.

During this period the trials for enhanced phosphorus removal started on 1st of August. The set up, goals and results of these tests are presented in section 5.2.3.

**Phase V**: Third commissioning: 18.09.08 – 31.12.08

A pH value of ~ 5.5 in combination with a high ammonia load is thought to be responsible for an incomplete nitrification. The inhibition of full nitrification resulted in an accumulation of nitrite. This led to an increased TTF (see Figure 67) and a decreased filterability of the sludge. Once this was recorded, the throughput was decreased to approximately 2.5 $m^3$/d unfortunately this reduction did not show the expected result and the nitrite concentration rose further on. The prior observed effects, foaming and an increase of the TMP to the maximum of 200 mbar, appeared again. The identified action to recover the system, an interruption of operation for 24 – 48 hours and a slow commissioning according to the nitrite content was successfully applied again.

The development of the total solids concentrations is shown in Figure 59. After the start of excess sludge withdrawal the TS concentration scatters around 11 g/L. The cloudy data points can be explained by the excess sludge regime explained earlier.

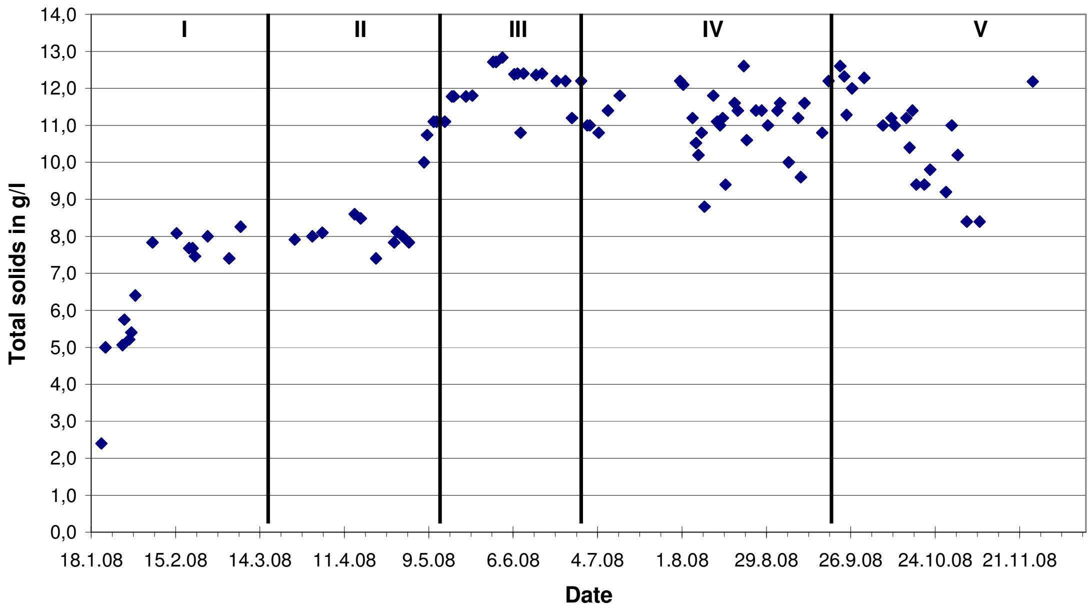



**Figure 60: TS evolution of the parallel MBR unit over time**

A TS concentration of about 11 g/L in the activated sludge reactor was the target in order to fulfill the set goals. Full nitrification, $NH_4$-N concentration in the effluent below 0.1 mgN/L and a COD concentration in the effluent below 50 mg/L was achieved most of the time.

TS concentration includes salts concentration of app. 1g/L, therefore the TS = MLSS + 1 g/L.

### 5.1.3 Process control scheme

Optimization of the control scheme for nutrient removal was carried out using a *stored program control* (SPC) device from ABB. As mentioned before, the usage of one floater as the only back loop control parameter leads to a *high/low level* control scheme. Table 21 shows the different control program applied during the different phases of operation. These changes did influence the biological process and therefore the overall performance. Section 5.2 will further focus on the impact of the control scheme on the nutrients removal and the optimization of elimination capacity.

The set up using a floater as a level indicator reduces the instrumentation to a single persistent device. The floater is installed within the membrane reactor and indicates *high* and *low level*. This level information is used to start/stop various control timers. Two different approaches to the control scheme were used during this year of operation:

- *Feed pump controlled* (phase I and II):

  The feed pump is steered by a timer, which activates the pump according to the planned throughput. During an active period the reactor is filled till high level is indicated. Afterwards filtration takes place as long as high level is stated. Once low level is reached, filtration stops and refill takes place, or, in case the feed pump is inactivated, a pause of operation leads to a period of relaxation and denitrification.

- *Filtration pump controlled* (phase III – V):

  This control program emphasizes the importance of a sufficient relaxation time for the modules enabling the operator to define exactly the time for filtration and relaxation (30 minutes of filtration and 15 – 25 minutes of relaxation during stable operation). A filtration pause shortly after refilling is possible this way, providing substrate for denitrification. This effect is explained further in Section 5.2.2.

The disadvantage of the *feed pump controlled* scheme is that under specific conditions a permanent filtration is possible, most likely prompted for shorter cleaning intervals. Therefore the *filtration pump controlled* scheme was implemented from 25$^{th}$ of April onwards and used for further optimization according to the nutrients elimination.

**Table 21: Applied control program**

| | *Control scheme* | *Effect* |
|---|---|---|
| Phase I: First commissioning and operation | Feed pump controlled | Permanent filtration possible |
| Phase II: Second commissioning | Feed pump controlled<br>(reduced throughput) | Permanent filtration possible |
| Phase III: Operation | Filtration controlled<br>Excess sludge withdrawal<br>Increased denitrification | Possibility of pre-denitrification due to availability of substrate during anoxic phases |
| Phase IV: Third commissioning | Filtration controlled | Long anoxic phases for denitrification, in order to recover the system;<br>low throughput |
| Phase V: Operation | Filtration controlled<br>Excess sludge withdrawal<br>Optimized for denitrification | Possibility of pre-denitrification due to availability of substrate during anoxic phases |

## 5.2 Nutrients removal

The concentration limits for discharge set by the water authority (see Table 22), made it necessary to optimize the process and to reduce the effluent concentration of phosphorous. Even though the nitrogen elimination would have been just about sufficient to achieve the discharge limits for 2008 under stable conditions, the process was also optimized according to the nitrogen elimination in order to identify the capacity of the tested plant. Also a reduced nitrogen effluent concentration helped to ensure the correct operation of both plants, decreasing the sensibility to disturbances, e.g. shock loads.

**Table 22: Concentration limits for the overall discharge of both MBR plants**

| | Until 2007 | 2008 |
|---|---|---|
| Chemical oxygen demand in mg/L | 50 | 50 |
| Inorganic nitrogen in mgN/L | 10 | 35 |
| Ammonia in mgN/L | 5 | 5 |
| Total phosphorous in mgP/L | 0.1 | 0.5 |

The concentration limits were set for the blend of both plants. According to the planned throughputs threshold concentrations for both plants could be calculated for the effluent concentrations for 2008, see Table 23.

**Table 23: Calculated maximum effluent thresholds in 2008**

| | Demonstration plant | Prototype plant |
|---|---|---|
| Throughput in $m^3/d$ | 10 | 3.5 |
| Chemical oxygen demand in mg/L | 50 | 50 |
| Inorganic nitrogen in mgN/L | 10 | 105 |
| Total phosphorous in mgP/L | 0.2 | 1.3 |

Elimination rates for the nutrients carbon, nitrogen and phosphorus are shown in Figure 61 and further described in the following sections.

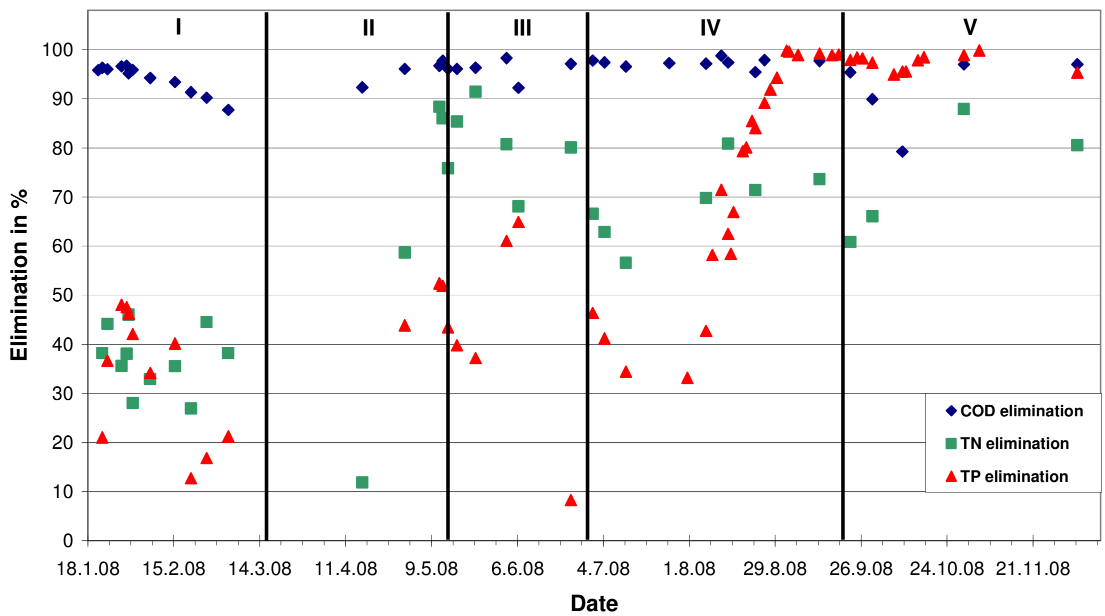



**Figure 61: Elimination rates for COD, TN and TP over time**

Table 24 shows the mean elimination rates and mean effluent concentrations for the nutrients of interest. Phase III and IV are highlighted in grey being the representative period of operation.

**Table 24: Key parameters for nutrients elimination and plant operation**

| | Phase I | Phase II | Phase III | Phase IV | Phase V | Overall |
|---|---|---|---|---|---|---|
| | 18.01.08 – 15.03.08 | 15.03.08 – 13.05.08 | 13.05.08 – 24.06.08 | 24.06.08 – 18.09.08 | 18.09.08 – 31.12.08 | 18.01.08 – 31.12.08 |
| **Chemical oxygen demand (COD)** | | | | | | |
| Mean Effluent concentration in mg/L | 60.8 | 34.8 | 43.1 | 34.8 | 81.4 | 49.9 |
| Mean elimination in % | 94.1 % | 95.7 % | 96.0 % | 97.4 % | 91.7 % | 95.2 % |
| **Total nitrogen (TN)** | | | | | | |
| Mean effluent concentration in mgN/l | 78.1 | 38.9 | 25.6 | 45.3 | 36.6 | 50.8 |
| Mean elimination rate in % | 37.13 % | 61.2 % | 80.3 % | 69.4 % | 73.8 % | 60.2 % |
| **Total phosphorus (TP)** | | | | | | |
| Mean effluent concentration in mgP/l | 11.6 | 10.2 | 10.0 | 5.6 | 0.5 | 6.2 |
| Mean elimination rate in % | 33.4 % | 27.6 % | 42.4 % | 74.6 % | 97.5 % | 65.2 % |
| **Total solids (TS)** | | | | | | |
| Mean value in g/L | 6.4 | 9.3 | 12.1 | 11.1 | 10.5 | 10.2 |
| **Throughput** | | | | | | |
| Mean value in $m^3/d$ | 2.2 % | 1.3 % | 3.5 % | 3.8 | 1.5 | 2.5 |

### 5.2.1 Chemical Oxygen Demand elimination

Chemical oxygen demand (COD) removal is one of the main targets of wastewater treatment and is usually achieved through utilization by micro organisms, either for growth or for respiration. Wastewater treatment plants using activated sludge benefit from this natural metabolisms and process schemes providing sufficient oxygen were designed.

The investigated prototype plant achieved the COD discharge limits most of the time. The mean elimination rate for COD removal was above 95 % during the whole year. Only during periods of disturbed operation, the effluent concentration was higher than the set discharge limits of 50 mg/L, see Table 24. These periods of disturbed operation were accompanied by low pH-values and heavy fouling and foaming. The high mean effluent concentration of the last phase of operation is due to a short period of time that was investigated, so only few measurements were carried out. The system recovered afterwards at the end of 2008.

Due to the successful and robust operation with regard to the COD elimination, the main focus was put on the reduction of nitrogen and phosphorous.

### 5.2.2 Nitrogen elimination

Full nitrification is arbitrary to reach the discharge concentration limits for ammonia of 1 mgN/L. Since nitrogen is needed by the biomass for growth, an elimination of ~ 30% of nitrogen is achieved through growth and discharge of excess sludge (Mudrack and Kunst 1991). Further reduction of the nitrogen effluent concentration can be achieved through periods of anoxic conditions providing phases that can be used for denitrification. The formed nitrate is converted to elementary nitrogen that is released to the atmosphere.

Anoxic phases were implemented through the above mentioned pauses induced by the control programme. During phase I and II, operating with the *feed pump controlled* scheme, these anoxic phases took place at the end of a cycle, ending with the low level signal. This also means that a long aerated phase just ended, leading to a low concentration of readily biodegradable substances present in the reactor. This deficiency of readily biodegradable substances resulted in a low denitrification capacity. The denitrification rates were not measured within the plant, as the reactors are not equipped with stirrers for mixing during these phases, so settling took place and collecting samples for analysis of both nitrate evolution and biomass determination could not be done in a representative way. Considering the fast utilization of easily degradable substances during aerobic conditions, denitrification rates in the range of endogenous rates are most likely to take place afterwards.

An optimization that was successfully tested was the implementation of a long denitrification break once or twice day. Figure 62 shows the nitrate evolution in the effluent for one cycle over 24 hours. These effluent samples were collected when the control program was set to cycle duration of 55 minutes, implementing a break of two hours every 22 hours. The graph shows clearly the trend of nitrate accumulation over one day. The pause of two hours led to a low nitrate concentration of approximately 22.5 mg $NO_3$-N/L at the beginning of the measurements. In the end a value of almost 38 mg $NO_3$-N/L was reached. The drop to 15.0 mg $NO_3$-N/L is probably due to an invalid sample and can be neglected. This measurement shows the range of the effluent nitrate concentration during stable operation with a throughput of ~3.5 $m^3$/d between 20 and 40 mgN/L.

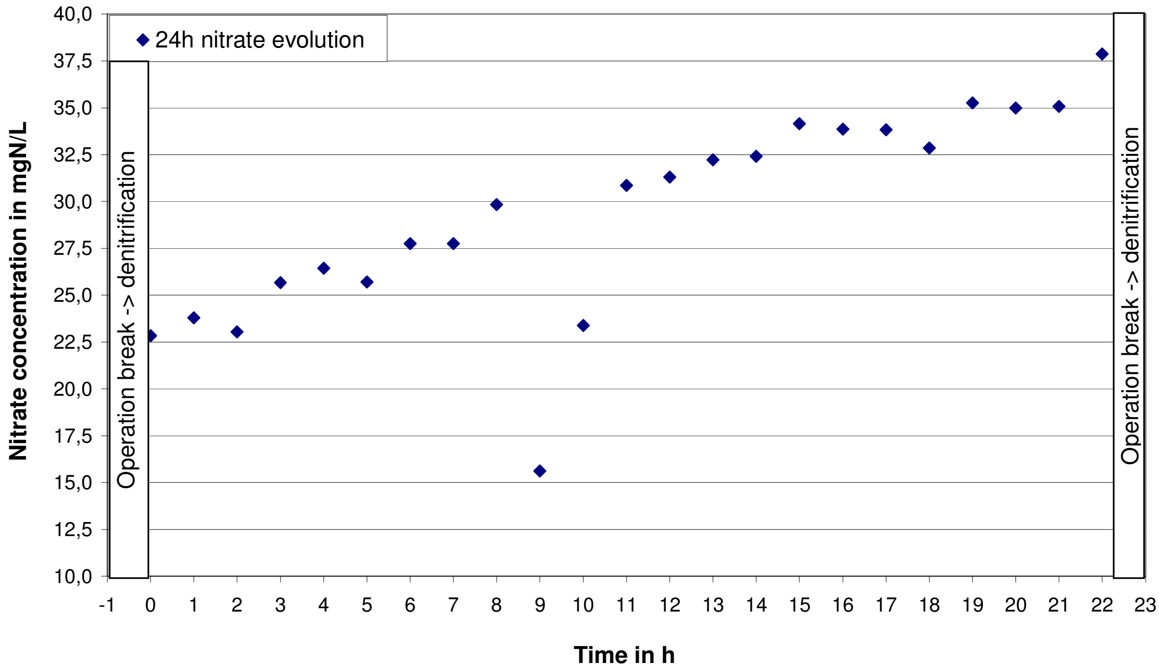



**Figure 62: Nitrate concentration in permeate collected in a 24 hours test**

With the change of the control scheme to *filtration controlled* operation, it could be assumed that at least some readily biodegradable substrate was available for denitrification which was depleted within a short time and endogenous denitrification rates occur afterwards. In addition the break of two hours was split up to one hour breaks twice a day. Thus higher denitrification rates due to availability of readily biodegradable substrate led to higher nitrogen elimination.

The evolution of nitrogen elimination is shown in Figure 61 and the low elimination range is clearly shown in the beginning of phase I. The elimination rate of nitrogen varied between 25 and 45 %. These rates can be explained by the prior mentioned biomass growth. However, the effluent concentration was between 65 and 105 mgN/L and thus just below the defined threshold of 105 mgN/L. As mentioned above, reducing the nitrogen effluent concentration increased the operation safety and lowered the influences of shock loads, thus optimizing the denitrification capacity was an important goal.

Phase II was characterized by an event of heavy foaming and fouling, leading to a halt of operation and thus a reduced throughput afterwards. Therefore fewer measurements have been carried out. Once a throughput of more than 1.5 $m^3$/d was reached, the process scheme was changed to the prior explained *filtration pump controlled process* scheme. Using this scheme nitrogen elimination increased to a peak elimination of over 90 % with a throughput of 3 $m^3$/d. This can be explained by the fact that using this control scheme filtration interruption, respectively anoxic phases, take place in the state of *high level*. This means that in opposite to the *feed pump controlled* scheme, readily biodegradable substrate might be present at the beginning of the anoxic phase. This enhances the denitrification ability similar to plants operated with pre-denitrification. Still, the availability of substrate depends on surrounding parameters, e.g. filtration performance or coarse aeration that influences the floater.

During the operation phases III to V the nitrogen elimination rate was between 60 and 90 % and thus significantly higher than before. A control program that ensures the availability of substrate during anoxic phases by feeding wastewater at the beginning of an anoxic phase has been programmed but could not be tested in 2008 due to operational problems. This program might result in a more stable elimination rate between 80 and 90 %.

### 5.2.3 Phosphorus elimination

In opposite to the demonstration plant, the prototype plant was not designed for biological phosphorus removal. As anaerobic conditions took only place at the end of a long pause and thus did not occur during stable operation, the enrichment of the activated sludge with phosphorous accumulating organisms was not very likely. In order to be able to reduce the phosphorous effluent concentration to the defined threshold, the plant was additionally equipped with precipitation and a downstream adsorption filter. The results of this test period are presented in the following sections.

#### 5.2.3.1 Precipitation

Orthophosphate present in wastewater can be removed by precipitation using metal ions. Different ions, e.g. $Fe^{3+}$ or $Al^{3+}$, are used in wastewater treatment plants according to process configuration and economical aspects. The precipitant used in the prototype plant was ferric chloride ($FeCl_3$) which was directly fed to the activated sludge tank.

A drop of pH due to the dosage of precipitant might lead to an inhibition of full nitrification or denitrification and thus to a collapse of the biological performance. This is why an overload of precipitant had to be avoided. As the prototype plant is set up as a SBMBR the dosage of precipitant was connected to filtration periods. This ensured that only precipitant is dosed when raw wastewater was fed. Marble plates with a low solubility were mounted in the activated sludge reactor and functioned as a pH buffer.

To identify the right amount of precipitant the dosage was increased slowly starting from a β-value of 0.5. The targeted effluent concentration of total phosphorus after precipitation should be between 1 – 2 mgP/L. Therefore a slow adaptation of the activated sludge was important to avoid a sudden pH drop. Also this additional time should help to gain enough experience with the set up and the handling of the precipitant. The β-value was slowly increased until 27.08.09 up to an end value of approximately 1.6, equivalent to a concentration of 37 $gFe/m^3$ of wastewater. Figure 63 shows the influent and effluent concentration of total phosphorous. The reduction of the total phosphorus concentration during the first period of operation is clearly demonstrated.

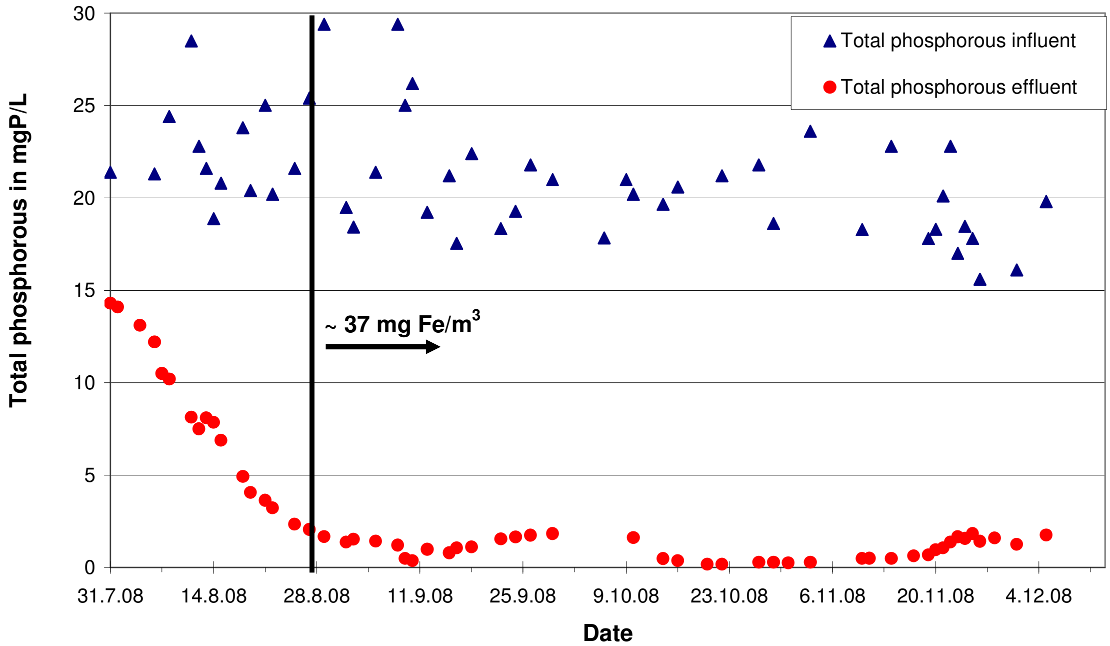



**Figure 63: Evolution of the total phosphorus concentration after start of precipitation**

After the targeted effluent concentration of 1 – 2 mgP/L was achieved, the adsorption filter was commissioned.

5.2.3.2 Adsorption filter

To identify the parameters for a successful implementation of a downstream adsorption filter, a small column was tested for almost three months. The geometrical parameters of this test filter are given in Table 25 and Picture 3 shows the used column. To assure sufficient contact of the inflow and the adsorbent and to minimize the possibility of a shortcut flow, the filter was designed as an upflow column.

**Table 25: Adsorption filter geometry**

| | |
|---|---|
| Diameter in m | 0.18 |
| Height of packed bed in m | 1.1 |
| Volume in $m^3$ | 0.028 |
| Ratio of height to diameter H/d | 6.11 |
| Volume flow in $m^3/s$ (L/min) | $9.17 \cdot 10^{-5}$ (5.5) |
| Superficial velocity in m/s | ~ 0.0036 |
| Rising velocity in m/s | ~ 0.0144 |

Ferric hydroxide ($Fe(OH)_3$) was used as the adsorbent produced by HeGo Biotech GmbH and distributed under the name FerroSorp® Plus. Once this material is depleted it can be recycled by HeGo Biotech GmbH or disposed with the domestic waste.

The physical properties are shown in Table 26 (Product specifications supplied by HeGo Biotech GmbH).

**Table 26: Physical properties of FerroSorp® Plus**

| | |
|---|---|
| Component | Ferric hydroxide |
| Formula | $Fe(OH)_3$ |
| Ferric hydroxide content | Min. 70 % |
| Bulk density | Max. 0.65 +/- 0.05 g/cm$^3$<br>(0.15 – 1.0 mm particle size) |
| Porosity | Min. 70 % |
| Maximum load with phosphorous | 10 g/kg |

Due to the set up as a SBMBR, the impact of the membrane performance on the flow rate and the different throughputs applied during this test phase, the contact time varied significantly. Typically for MBRs, filtration pauses were regularly implemented leading to changing contact times. To minimize this impact on the measurements, samples were always taken at the end of filtration phase. This represented the shortest contact time, thus the highest concentration in the effluent. Calculated with a porosity of 70 % and an adsorbent volume of approximately 28 liters the contact time during filtration was approximately 3.6 – 3.9 minutes. Due to the nature of a packed bed column the exact contact time and contact efficiency can not be determined without further experiments.

Figure 64 shows inlet and outlet concentration of the adsorption filter. The filter was designed for short term trials, therefore the stand time was comparatively short. The adsorbent was changed, when no cleaning effect could be determined, meaning that the inlet concentration was as high as the outlet concentration. This was usually around 1.0 mgP/L. Under stable operation the adsorbent had to be changed every four to five weeks. The load of phosphorous was determined with 4.5 g/kg, which lies below the theoretical maximum load given by the manufacturer of 10 g/kg. This can be explained by the short contact time during filtration.

The usage of this adsorption filter helped to achieve the defined effluent concentrations and functioned as a safety step in case a shock load of phosphorous reached the prototype plant.

The combination of precipitation and a downstream adsorption filter was successfully implemented in the prototype plant. The effluent concentration of total phosphorus was lowered to a mean value of 0.35 mgP/L since start of dosing precipitant. Therefore the discharge limits set by the water authority were achieved.

**Picture 3: Adsorption filter**

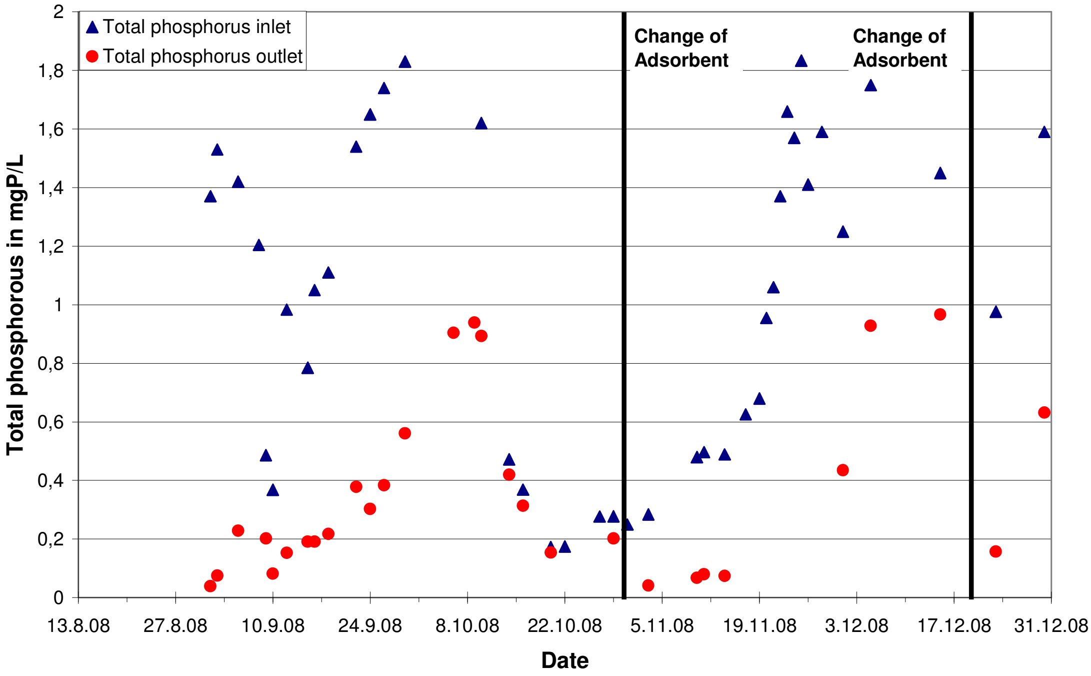



**Figure 64: Inlet and outlet concentrations of the adsorption filter**

## 5.3 Membrane performance

The filtration performance was of special interest, as this prototype plant gave the opportunity to investigate two MBRs fed with the same wastewater but using different biological processes. Also the cleaning strategies of the two plants were different:

- The modules of the demonstration plant were cleaned every month.
- The modules of the prototype plant were not cleaned on site. This means no chemicals were used during the year of operation to recover the modules. Twice, mid July and mid December the modules were replaced by Busse IS. The modules were cleaned by Busse IS and, if approved, reinstalled in MBR systems.

The measurements carried out at the prototype plant were mainly tests to characterize the sludge. The modules themselves were not object of tests such as critical flux determination or increased fluxes during periods of high inflow. Nevertheless the filtration performance of the modules were recorded and analyzed regarding the achieved flux and permeability.

The technical equipment did not allow a permanent record of all data, therefore the throughput and filtration was recorded manually approximately twice a week. The pressure values were registered by a small data logger once a minute. Daily mean pressure values obtained during filtration were used to calculate the resulting permeability. The temperature was not recorded permanently throughout the year of investigation, so sporadic values were measured directly in the activated sludge and missing data was interpolated. In addition, the values were compared to those recorded in the demonstration plant.

Figure 65 shows the instantaneous and the net flux throughout the year. The flow was kept constant at approximately 5.5 L/min, thus the instantaneous flux was constant at approximately 16.5 L/$m^2$h during stable operation. A lower instantaneous flux was possible, when heavy fouling occurred and either

i. the increased transmembrane pressure (TMP) led to a lower flow by the filtration pump, or

ii. the low pressure valve opened reaching a TMP of more than 200 mbar. The drawn air reduces the permeate flow, thus leading to a lower flux.

The two periods of heavy fouling are shown through the reduced Net flux which is in correlation to the adapted throughput as discussed in Section 5.1.2.

In opposite to the first period of heavy fouling in phase II, the instantaneous flux was not influenced by the period of fouling during phase V. This can be explained by the immediate actions introduced with the first recognition of disturbed operation, e.g. increased TTF values and a notable nitrite concentration.

At first the throughput was reduced to approximately 2.5 $m^3/d$, assuming that this reduction will help to decrease the amount of nitrite in the reactor, therefore the filtration cycles were reduced, giving the membrane longer relaxation pauses thus more time to recover. As this reduction showed to be insufficient, a complete stop of operation was initiated. Afterwards the plant was slowly commissioned again, as described prior. In addition to the fewer filtration cycles, the filtration time per cycle was reduced from 30 to 15 minutes. This way the TMP did not increase to the fullest and therefore the instantaneous flux and the permeability were not heavily affected by the unfavorable sludge characteristics.

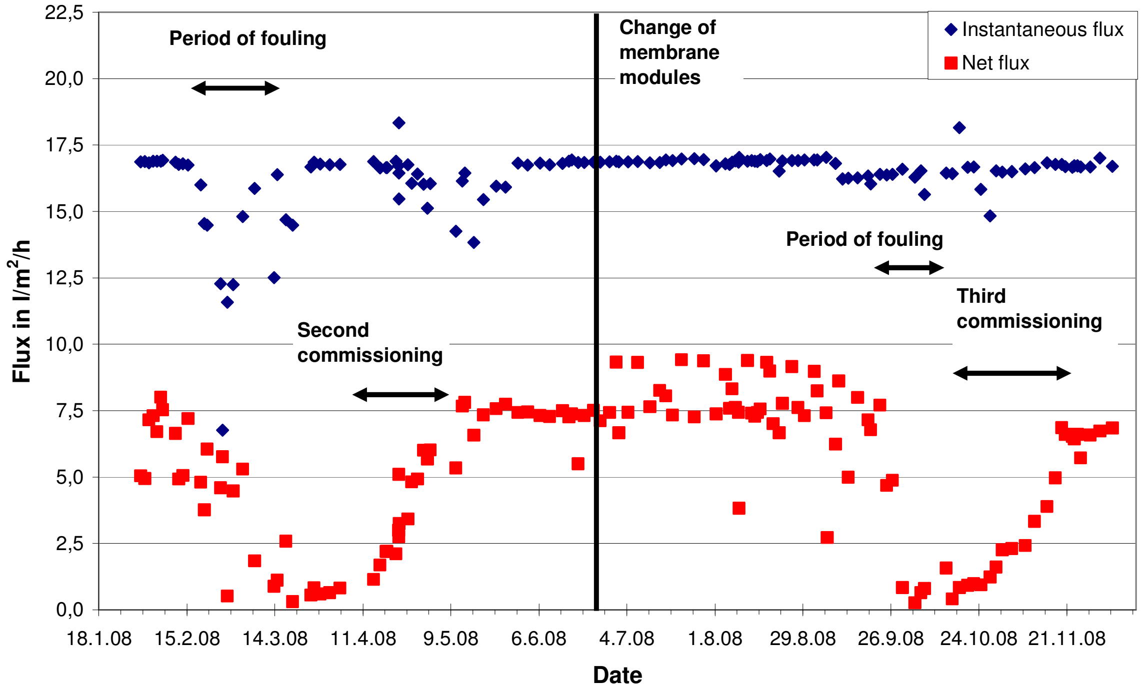



**Figure 65: Instantaneous and Net flux**

The normalized permeabilities are shown in Figure 66. The initial permeability was just below 300 L/($m^2$*h*bar). Within the first weeks of operation the permeability decreased rapidly. This is explained by the fact that during commissioning the TS increased according through the higher throughput. Before a stable operation was achieved the permeability decreased further to a minimum of ~ 45 L/($m^2$*h*bar). This period represented the first event of heavy fouling. During the slow commissioning afterwards it is noticeable that the permeability increased as a result of improving sludge characteristics, but was still scattering. It has to be noted that the recorded pressure values during filtration were not very precise thus the calculated permeabilities show rather a trend than an accurate absolute value. This can be explained by the vibration caused by the filtration pump and the collection of only one data point per minute.

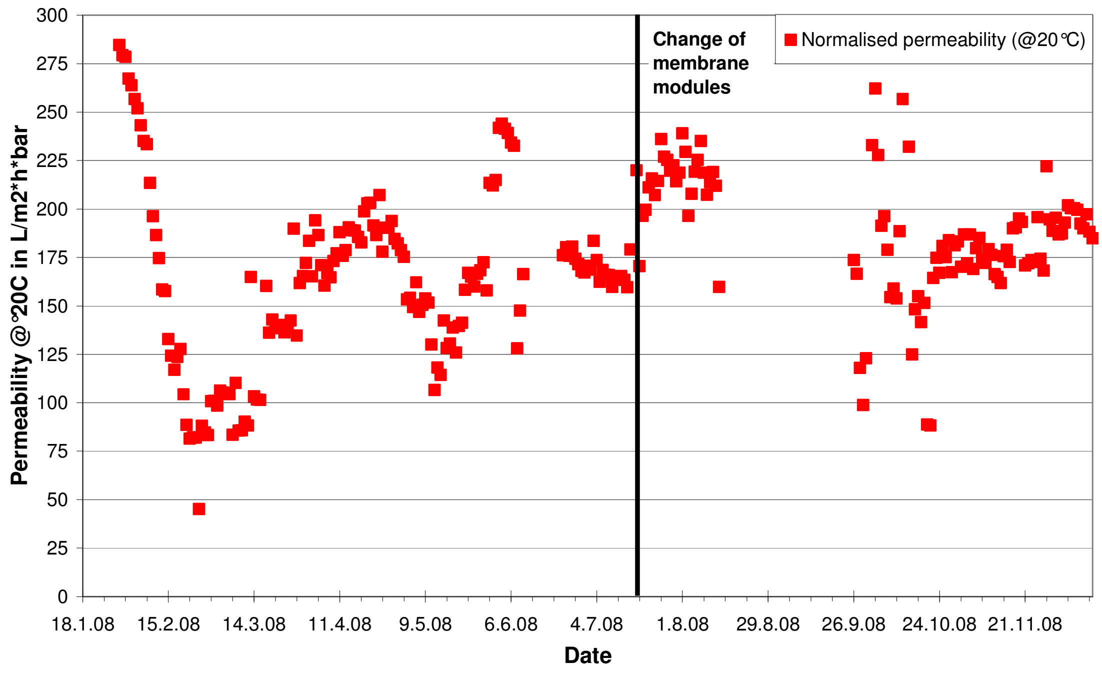



**Figure 66: Normalized permeability (@20°C)**

This showed the importance of an adequate commissioning after a collapse of the biological system. When recovering a MBR plant, both the biological system as well as the filtration performance has to be watched.

One key characteristic that showed the low filterability during the events of heavy fouling was the time to filter (TTF) which was measured as the time to filter in dead end filtration 25 ml out of 250 ml activated sludge through a black ribbon filter paper (Whatman, pore size between 12 and 25 µm, diameter 90 mm). TTF values gave similar information as the also measured capillary suction time (CST), but could be determined with less required equipment. Monitoring the values for time to filter gave quick information of the filterability of the activated sludge. Figure 67 shows the values for time to filter measurements throughout the year. Both events of fouling are represented in values more than tenfold to the average values obtained during stable operation. It is also shown that a daily measurement is necessary to be able to respond fast enough. In both cases between a regular value and a value indicating a major swift in sludge characteristics was just a week. Nevertheless at this point no other promising action was identified than the explained halt of operation followed by a slow commissioning. A daily measurement might help to record the upcoming fouling event earlier, giving the chance for immediate actions without a complete downtime of the plant.

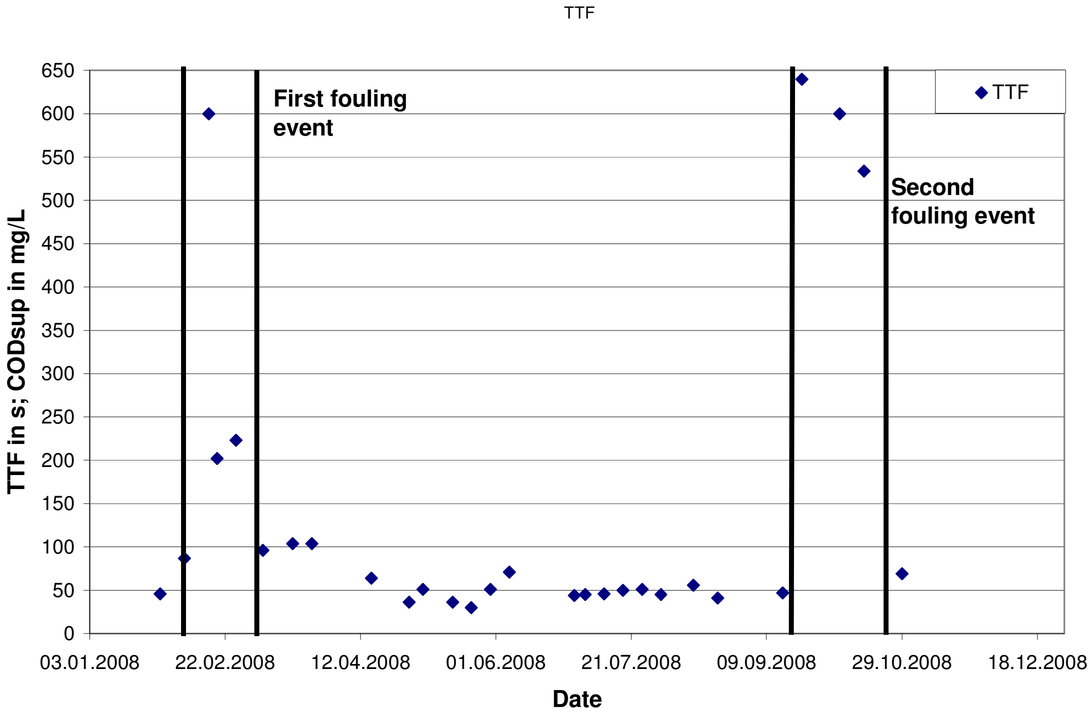



**Figure 67: Time to filter measurements**

## 5.4 Energy demand

Including all phases, the mean specific energy consumption was app. 4.9 kWh/m$^3$, due to periods of low throughput caused by operational difficulties.

The energy consumption is a key figure to describe the impact on the environment of any process. Great efforts to achieve the goal of sustainability, or at least to come as close as possible, are done in wastewater treatment plants. In the field of decentralized wastewater treatment the energy consumption is of high interest as the equipment is downscaled to the required size, leading to lower efficiencies. At the same time the effluent quality requirements can be as high as of centralized treatment plants depending on the sensitivity of the receiving water bodies. Therefore the energy consumption of the prototype plant was recorded and correlated to the achieved effluent quality.

To be able to distinguish between the energy required for wastewater treatment and the total energy needs, including heating and research related instrumentation, two data points are given net and total energy demand. This helps to determine the exact energy requirements for the treatment and enables the comparison to other decentralized wastewater treatment plants.

Figure 68 shows the specific energy demand throughout the year. In addition, the achieved throughput is shown to demonstrate this enormous impact, e.g. after the first event of heavy fouling just a minimum throughput was achieved, resulting in a high energy consumption. This demonstrates the ground level of energy that is consumed, no matter how much wastewater is treated. This ground level of energy is caused by periodical recirculation and aeration in order to provide enough oxygen for the biomass even though no wastewater is treated. Therefore, increasing the throughput to the planned 3.5 to 4.5 m$^3$/d led to a reduction of the specific energy demand. During stable operation of phases III and IV a mean value of 3.6 kWh/m$^3$ (net) showed the low energy demand.

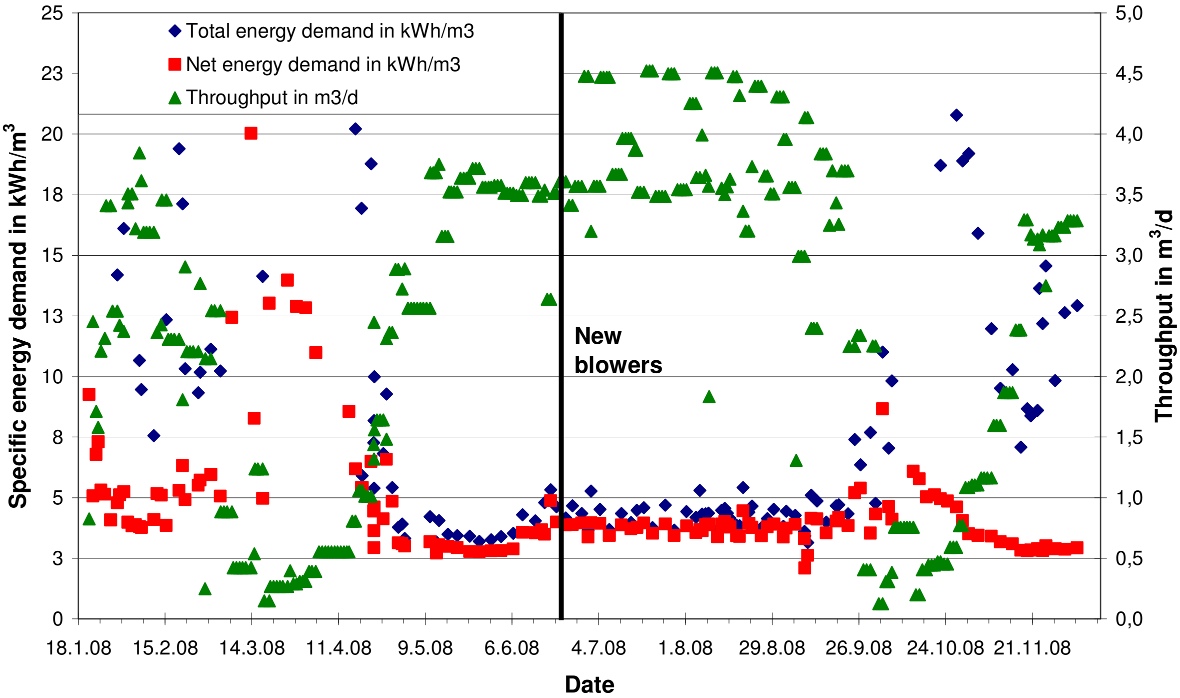



**Figure 68: Energy demand of the BUSSE prototype plant**

Table 27 shows the mean energy consumption in comparison to the achieved elimination rates. COD elimination was above 95 % despite the lower hydraulic retention time during the weekends. The total nitrogen elimination was above 74 %. As mentioned before, higher nitrogen elimination rates might be possible, but could not be tested within this year of operation. The total phosphorus elimination includes periods before precipitation and adsorption, therefore the rate is lower than the achieved eliminations rates mentioned in Section 5.2.3.

**Table 27: Energy consumption and elimination rates for phase III + IV**

| | **Prototype plant** |
|---|---|
| Energy demand in kWh/$m^3$ | 3.6 |
| Mean throughput in $m^3$/d | 3.7 |
| Mean COD elimination in % | 96.9 |
| Mean TN elimination in % | 74.1 |
| Mean TP elimination in % only precipitation | 76.0 |
| Mean TP elimination in % precipitation and adsorption filter | 98.1 |

Aeration for both, oxygen supply and membrane cleaning is the main consumer in this prototype plant, see Figure 22. During phase III dissolved oxygen concentration became a critical parameter. Higher temperatures led to an insufficient oxygen mass transfer using the same blowers and aeration regime. Therefore a new aeration regime was introduced on the 6th of June. The membrane chamber was permanently aerated and the activated sludge reactor was aerated during filtration. Few days later on the 16th of June stronger aerators were installed for the activated sludge reactor in order to increase the dissolved oxygen concentration during aeration. As the activated sludge reactor had to supply anoxic conditions during filtration pause, permanent aeration could not be implemented in this reactor. This changed aeration regime showed to be sufficient to ensure full nitrification. The achieved dissolved oxygen concentration within the membrane chamber was around 3.0 mg/L. The specific energy consumptions rose slightly from around 3.0 to 3.6 kWh/$m^3$.

Further optimization might be possible, but would most likely lead to increased automation requirements, demanding oxygen measurement combined with a feedback control. This might also lead to a more instable system.

## 5.5 Design recommendations

As discussed in Section 2.14 the experience gained during operation of the prototype plant were used to scale the plant. Additionally recommendations can be given in order to achieve higher effluent qualities with respect to nitrogen and phosphorus removal. The prototype plant design can be used for catchment areas between 50 and 200 pe. The following recommendations were defined for this range of installations.

- Full COD removal and complete nitrification:

  | | |
  |---|---|
  | COD sludge load: | 0.05 – 0.1 kg COD/kg TS*d. |
  | Nitrogen sludge load: | 0.006 – 0.014 kg N/kg TS*d |
  | TS: | ~ 11 g/L |
  | SRT: | app. 50 days |
  | HRT: | ~ 24 h |

  Permanent aeration of the membrane reactor assures full nitrification.

- Additional N removal of ~ 80 %

  Same process parameters, but the HRT should be increased up to 34 hours, implementing anoxic conditions sufficient for denitrification.

- Additional P removal of ~ 99%

  Precipitation with $FeCl_3$ in the activated sludge reactor was successfully tested and it was shown that effluent concentration as low as 1 – 2 mgP/L could be achieved. For further P removal a downstream adsorption filter was installed and reduced the furthermore, see 5.2.3.

## 5.6 Economical evaluation

The main goal of the parallel operation of prototype plant was to reduce costs for trucking away the additional wastewater. Table 28 shows the costs accumulated for operation during 2008.

**Table 28: Costs prototype plant for 2008**

| | **€/a** (net) |
|---|---|
| **Operational costs** | |
| Personnel | 4.000,- |
| Change of modules and maintenance by Busse IS (2 * 650.-) | 1.300.- |
| Energy | 1.168.- |
| Consumables (precipitant, adsorption granules) | 730.- |
| **Investment cost** | |
| Imputed costs external finance at interest loan of 4.5% | 3.258.- |
| **Sum** | **10.456.-** |

Personnel costs are expected to decrease, due to the experiences gained in this first year of operation. The trucking of the additional wastewater by an external company costs approximately 20.- €/$m^3$ (net). The prototype plant treated the amount of 791 $m^3$ in 2008 which would have cost 15.820 € (net). The savings achieved with the prototype plant are 5.364 €. As discussed prior, two events of heavy fouling and the slow commissioning phase afterwards reduced the throughput significantly. Considering a higher throughput due to a stable operation in the years to come, the savings will increase considerably.

## 5.7 Conclusions

The operation of the prototype plant provided by Busse IS helped to reduce the costs for wastewater treatment in Berlin-Margaretenhöhe. The trucking of wastewater during high inflow periods was reduced by the operation of the prototype plant. In addition this helped to operate the demonstration plant under more stable conditions.

The total sum of 791 $m^3$ of wastewater was treated from January to December 2008. The mean throughput of 2.52 $m^3/d$ was lower than expected, what can be explained with the low commissioning after events of disturbed operation. The reasons for these disturbances are not completely understood, but toxic substances in the influent are most likely to be the reason. It was shown, that a stable operation with a throughput of approximately 3.5 $m^3/d$ could be achieved. The limit of the achieved throughput was about 4.5 $m^3/d$ which was operated during the weekends. During these periods of higher throughputs dissolved oxygen concentration could be insufficient due to high loads of ammonia. Therefore it was of high importance to monitor this value, preventing an inhibition of nitrification by high ammonia concentrations.

Designed for COD removal and full nitrification, this prototype plant was optimized successfully in terms of further nutrients removal. Total nitrogen elimination could be increased by implementing secured anoxic phases with readily biodegradable COD present. The elimination of total phosphorus was increased by introducing chemical precipitation and a downstream adsorption filter.

The energy requirements with regards to the achieved effluent quality and investment costs are low in comparison to wastewater treatment plants of similar size.

The robust set up helped to ensure a stable operation regarding only few man-hours and little maintenance.

# Chapter 6

# Conclusion, technical recommendations and outlooks

The following chapter summarizes the results and experiences presented before, and gives the key recommendations for the technical operation and design. Furthermore prospective research activities are proposed according to the presented results.

## 6.1 Conclusions

### 6.1.1 Biological performance

The biological nutrient removal performance of the demonstration plant applying biological phosphorus removal combined with post-denitrification proved to be able to reach very high effluent qualities. In respect to the high influent concentrations the plant showed the highest elimination rates worldwide published so far.

**Table 29: Key effluent parameters in relation to plant size**

| | Requirements Waste Water Ordinance | | | | | Requirements Margaretenhöhe | Results w/o precipitation | Results With low precipitation |
|---|---|---|---|---|---|---|---|---|
| **Plant class** | 1 | 2 | 3 | 4 | 5 | | | |
| | mg/L | mg/L | mg/L | mg/L | mg/L | mg/L | | |
| **COD** | 150 | 110 | 90 | 90 | 50 | 50 | 43 | 44 |
| **BOS** | 40 | 25 | 20 | 20 | 15 | < 5 | < 3 | < 3 |
| **$NH_4$-N** | | | 10 | 10 | 10 | 1 | 0.05 | 0.07 |
| **$N_{anorg.}$** | | | | **18** | **13** | **10** | **5.3** | **3.1** |
| **TN** | | | | | | | 8.5 | 5.6 |
| **$PO_4$-P** | | | | | | | 0.12 | 0.03 |
| **TP** | | | | **2** | **1** | **0.1** | **0.23** | **0.1** |
| **Disinfection** | | | | | | New EU Bathing Water Directives | | |

Table 29 shows the achieved effluent concentration in comparison to the requirements. The outstanding effluent quality has to be seen in relation to the plant size. For the demonstration plant effluent qualities exceeding those of centralized treatment plants were demanded due to the sensible receiving water body. According tot the plant size alone, the required effluent quality is significantly lower.

### 6.1.2 Cost evaluation

Further investigation on decentralized wastewater treatment with MBR systems in respect to the achieved effluent qualities and the costs were implemented in the present study. The experiences and data collected over three years within the demonstration project ENREM for decentralized wastewater treatment was not only used to calculate the costs for this specific case, but also to estimate the costs for applications serving larger catchment areas, see Section 2.14. The main outcomes are presented in Table 30.

**Table 30: Specific overall costs for the investigated technologies**

| | **Demonstration plant**<br>**ENREM process**<br>(TP elimination > 99%) | **Prototype plant**<br>**NR/DN + precipitation + p - adsorption**<br>(TP elimination > 99%) |
|---|---|---|
| Plant size p.e. | Overall costs in € / $m^3$ wastewater (net) | |
| 50 | | 7.5 – 10.5* |
| 130 | 16 – 17** | |
| 250 | 8.5 – 12.8 | 4 – 6 |
| 1000 | 4.6 – 6. 8 | 2.5 – 3.8 |
| 5000 | 1.8 – 2.7 | |

* assessment on operated prototype unit

** assessment on operated demonstration plant

The results show that the operation of the ENREM process becomes economically competitive for plant sizes of 5000 p.e. and larger. For smaller catchment areas the prototype plant equipped with precipitation and a downstream adsorption filter is economically a competitive solution. It has to be noted that ENREM process achieves higher nitrogen elimination rates and might be therefore used for applications requiring this high effluent quality, e.g. sensible receiving water bodies.

## 6.2 Technical Recommendations

### 6.2.1 Operational experience and design recommendation

- Buffer tank: The equalization of the hydraulic flow and loadings was presented in the prior report by (Gnirss *et al.* 2007). The benefits for both the biological process and the required membrane surface area justify the buffer tank and it is therefore recommended. Since the buffer tank and the tank collecting grid and excess sludge have to be emptied occasionally the ground should be designed with a pump well, simplifying collecting the grid.
- Inflow regime: Due to the oversized feed pumps the inflow runs discontinuously and only for short period of time, app. 10 s every 2-3 minutes. This inflow regime does not have a negative effect on the biological cleaning performance but might enhance the wearout of the feeding pumps.
- Sieving: The sieving equipped with an 1.0 mm hole sieve and automatic brush is suitable for the raw wastewater characteristics received in Margaretenhöhe for larger plants. To reduce the man hours on site and to increase plant availability an automatic flush of the collected grit is recommended.

- Anaerobic reactor: Nitrate recirculation into the anaerobic reactor caused the break down of biological phosphorus removal due to consummation of VFAs for denitrification and insufficient anaerobic contact time, see Chapter 3. To maintain the ability of biological phosphorus removal true anaerobic conditions have to be ensured. The volume of the anaerobic reactor has to be designed with respect to the flows, expected nitrate concentrations, biological rates (e.g. phosphorus release rate or pre-denitrification rate) and economical considerations. For the demonstration plant it is recommended to double the size of the anaerobic reactor, thus increasing the overall ratio.
    - Additionally the possibility to feed a carbon source (e.g. acetate or propionate) into the anaerobic reactor would improve the overall performance and operational availability of the plant in case of disturbances. As explained in Chapter 3 the consummation of VFAs for denitrification leads to an insufficient carbon source for PAOs strengthening microbiological competitors and finally leading to the loss of PAOs in the system. The proposed feeding of a carbon source was successfully tested as explained in Section 3.1 supporting PAOs and ensuring sufficient phosphorus removal.

**Table 31: Reactor volume ratios of the demonstration plant**

| | Ratio of reactors in %<br>AN : AE : AX : MR |
|---|---|
| Demonstration plant | 8 : 21 : 63 : 8 |
| Recommended design | 15 : 25 : 50* : 10 |

* AE/AX reactor assumed to be AX

- Aerobic reactor: Due to unavoidable foaming events the aerobic reactor should be designed with a sufficient head space. Nitrification and phosphate uptake as well as sbCOD removal takes place under aerobic conditions.
- AE/AX switch reactor: To be able to react on changing ambient conditions, such as temperature or wastewater constituent, a flexible reactor should be considered for the transition between aerobic and anoxic zones. This AE/AX reactor has to be equipped with an aerator as well as with a mixing device. The advantages regarding the biological nutrient elimination of such a design are as follows:
    - Nitrification: The fraction of nitrifying organisms decreases with falling temperatures due to lower grow rate of this species compared to heterotrophic organisms. Therefore the ability to increase the aerobic contact time by activating aeration in the AE/AX reactor might be necessary to ensure sufficient nitrification. On the contrary with high temperatures the solubility of oxygen in water decreases thus the aeration might be insufficient without the AE/AX reactor.
    - Phosphate uptake: Peak loads of phosphorus, insufficient dissolved oxygen concentrations, poor fraction of PAOs in the system or deficient PHA build up by PAOs can lead to lower phosphorus uptake rates. Thus a longer aerobic contact time helps to increase the phosphorus removal.
    - Denitrification: In contradiction to the prior cases where a longer aerobic contact time leads to a better performance, the post-denitrification step might

be limited by the DNR. The importance of low nitrate concentrations recirculated to the anaerobic reactor was expressed in Chapter 3. Therefore the use of the AE/AX reactor with anoxic conditions is beneficiary for the overall process and the nitrogen elimination rate.

What condition in the AE/AX reactor should be used depends on the actual ambient circumstances, see 6.2.2. A flexible reactor enables the operators to react quickly on changing conditions and will therefore also increase the plant availability and stability. It gives also the possibility to operate this reactor with a very low DO concentration with low oxygen carry over to the anoxic zone. This reactor has to be prepared for both conditions, thus providing aeration devices and connections as well as the electricity and mounting possibility for an engine/stirrer.

- Deoxidation zone: The demonstration plant was equipped with a small reactor designed for deoxidation (deox) following the aerobic reactor. Since foaming led to blockage of this reactor causing intensified sludge loss due to failure of sludge distribution throughout the plant and the benefits were in no relation to the disadvantages, it is proposed to leave out this reactor.
- Anoxic reactors: Ideal mixing without oxygen transfer has to be ensured, thus the use of controllable engines via frequent converters should be installed. Two constraints have to be respected for the anoxic zone:

  1. No dissolved oxygen, and
  2. sufficient mixing.

  Oxygen carry over from the aerobic reactor can not totally be neglected but should be minimized. The design of a deoxidation reactor between the aerobic and anoxic zones has to be carried out carefully. Stirring speed has to be adjusted well, as high velocities lead to oxygen transfer and low velocities to insufficient mixing. Therefore the engines should be controllable by frequency converters.
- The membrane reactor has to be designed with sufficient head space as well as adequate space around the membrane module to ensure a proper air lift circulation. The reactor volume in relation to the combined volume should not exceed 10%, see Table 31.
- The recirculation and filtration pumps were reliable throughout the project and just the planned maintenance was carried out. Following recirculation rates according to the throughput are recommended:
  - Last anoxic to anaerobic reactor: 150%
  - Membrane to aerobic reactor: 400%
- Engines: The installed engines for mixers were mounted directly above each reactor (AN and AX) thus facing electrical failure during foam events. As the foam produced in the aerobic and membrane reactor was distributed over the whole plant, the foam had direct contact to the engines leading to their collapse. Therefore it is proposed to realize the mixing with either a separate engine compartment or waterproof mixing devices (e.g. submerged pumps).
- Stirrers: Since the first mounted propeller stirrers did not mix the reactors properly, H-shaped stirrers were installed and successfully operated. For larger plants, other stirrers could be more cost efficient.

- The possibility to truck away the excess sludge, grid and wastewater in cases of extremely high inflow volumes has to be considered for smaller plant sizes, when no sludge handling on site is feasible.
- Proper foam handling can be realized by leading the foam from the point of origin (aerated reactors) to a buffer tank where the foam/sludge can be stored and pumped back to the aerobic chamber. This way sludge loss will be minimized and the designed flow regime is kept.
- The sludge distribution through the reactors was realized with hydraulic flow only, reducing the number of pumps necessary and thus optimizing the energy demand. The disadvantage of such a flow scheme is the sensibility to level changes caused by e.g. foam formation. The overall height difference between the first (anaerobic) and last (anoxic) reactor was just a few centimeters. In cases of foam formation, when the foam level exceeded significantly the designed water/sludge level, a backflow of foam to the anaerobic reactor led to oxygen transfer and the wear out of the engine installed. Therefore it is recommended that a sufficient height difference is planned and foam formation as well as MLSS concentrations up to 20 g/L (increased viscosity) is considered, when designing a hydraulic flow.

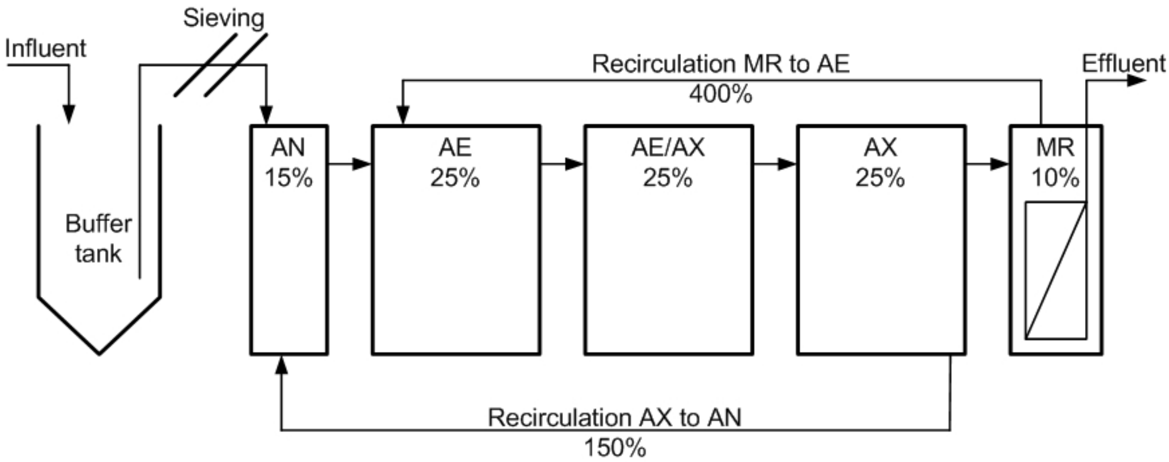



**Figure 69: Recommended plant design**

Since the nitrogen load is the crucial factor in this biological process it is recommended to design the plant within the following range:

- Minimum: 0.04 kgN/m$^3$
- 85% tile: 0.15 kgN/dm$^3$
- Peak load: 0.02 kgN/dm$^3$

The design guidelines with respect to the loads are explained in detail in Section 2.12.

### 6.2.2 Automation

- Controlling of the influent pumps and sludge recycling in accordance to the plant discharge (filtration pumps) are crucial for automatic plant operation.
- TS control: Controlling the excess sludge removal and the SRT against the TS concentration is favorable to adjust a constant TS concentration avoiding frequent

manual SRT settings. The control must have constraints to avoid problems because of implausible TS online values.

- Plant shut down at foaming: To minimize the influence of foaming events (heavy sludge loss) the plant will be shut down when terms are fulfilled: 1) The foam sensor is detecting foam and 2) the TS concentration given by the TS sensor is leaving a given range. In this case the plant will shut down, gives an SMS alarm and requires a contact of the operator.
- Different responses to changing effluent concentration of nitrate and phosphate were tested throughout the project and the following interconnection should be implemented in the control scheme as optional measures:
  - Phosphate:
    - The online value activates the carbon dosage according to an set threshold value, see Section 3.1
    - The online value is used to increase the dosage of precipitant either according to a set threshold or an implemented calculation
  - Nitrate:
    - Increased nitrate effluent concentration activates the carbon dosage in order to replace VFAs consumed for denitrification. Due to the point of measurement in the permeate, the value might be higher than the actual nitrate concentration in the recirculation stream, as the aerobic condition in membrane reactor allows nitrification.
- The implementation of the flexible AE/AX reactor should be accompanied with the introduction of an ammonia probe. The ammonia probe enables the process control to decide whether aerobic or anoxic conditions in the flexible AE/AX reactor are more beneficial for the total process. During the operation of the demonstration plant it was shown, that an ammonia concentration of 2 mgN/L entering the membrane reactor was tolerable, as full nitrification will take place there. Therefore it is recommended to use the AE/AX reactor with anoxic conditions, as long as the ammonia concentration is below 2 mgN/L, consequently switching on the aeration when insufficient nitrification is observed.

### 6.2.3 Filtration performance and recommendations

The filtration performance of the used MF flat sheet modules of A3 Water Solution was consistently more than sufficient and was never the limiting factor for the plant throughput. The chosen operation scheme with one module in use, whilst the second is preserved in cleaning agent, thus instantly available in case of an increased throughput demand, secured the required operation reliability and contributed its part to the fact that the demonstration plant was operated continuously since commissioning of the modules. Even during periods of heavy fouling when extraordinary low sludge filterability was recorded, see Chapter 4, the required throughput was achieved. The design and module type is therefore recommended for decentralized MBR systems as additionally the working hours accounted for filtration maintenance is acceptable.

A new cleaning strategy was tested, see Section 4.2, and can be recommended for decentralized treatment plants. The use of $H_2O_2$ with pH 11 as a substitute for chlorine was successfully demonstrated and should be considered as the main cleaning agent, due to the lower environmental hazard potential. Nevertheless it was also demonstrated that in periods of heavy fouling chlorine is the more effective cleaning agent and therefore the possibility to clean with chlorine should always be considered when designing a MBR system, in terms of e.g. handling, short term storage or housing materials.

The commercial filtration control software Veolink® was implemented in the control regime to test the benefits of such a sophisticated online monitoring tool, but unfortunately could not be tested sufficiently, see Section 2.9. Therefore it is recommended to carry out further tests with this tool to collect sufficient data allowing to decide whether such a tool is beneficial for MBR systems in general and for decentralized system in particular. The implementation of this control tool in decentralized MBR systems could allocate for significantly increased plant reliability, due to automatic measures in case of a decreasing filtration performance. Additionally the tool would help optimize cleaning intervals and record the cleaning efficiency. This all will help to run a MBR system in a more sustainable way, reducing chemicals, energy, working hours, operational and investment costs.

### 6.2.4 Operational risks

- The presented process of high nutrients removal and the operation with no staff on site make a high grade of instrumentation and automation necessary. The risk of plant operation trouble because of instrumentation failures is higher as in simple constructed plants. Plausibility checks of the online values are crucial for good cleaning performance. Though the automation level is quite high, manual intervention is recommended every second week for a plant size of 130 p.e.
- The sewer of this decentralised area has a high risk of irregular discharges. The behaviour of the inhabitants has a high influence on the process stability. Illegal rain discharges and disposal of for the biology harmful waste (e.g. high amounts of tensides, heavy metals) are big problems for the plant because small plant size and the missing of dilution as in larger plants. Some heavy foam events in the plant are strongly suspected to be related to illegal discharges.
- Considering the plant operation experiences of the last three years foaming can not totally be ruled out. Design-engineering, automatic alarms and organisational arrangements are necessary to reduce the impact of foaming
- Nutrient influent loads can vary in a wide range especially in a small area because of seasonal (vacation, summer houses) or social circumstances (less discharge on weekdays). The plant must be able to respond to changing loads in a proper time.

## 6.3 Outlooks

Despite the distinctive research activities on various fields within the ENREM project, there are still questions unsolved and further investigations focusing on the biological process as well as the filtration characteristics are proposed to give valuable information for future technological trends.

### 6.3.1 Investigations on post-denitrification

As presented in Section 3.3, the last experimental results addressing the kinetics of post denitrification and microbiological organisms involved strongly suggest the presence of denitrifying phosphorus accumulating organisms (DPAOs) in the ENREM process, utilizing internally stored PHAs as the carbon source for denitrification. Therefore it is recommended to run a series of batch tests using real wastewater sludge and acetate as the feed to correlate the consumption of PHAs to enhanced denitrification rates. Once the expected correlation is determined in a sufficient number of consecutive experiments (5 – 10 batch tests) the enhanced denitrification rates observed throughout the ENREM project can finally be explained. Recent results by research groups focusing on the microbiological organisms thought to be true denitrifying PAOs led to the development of a new FISH probe marking *A. phosphatis clade I* in activated sludge samples. The use of this probe is therefore also recommended in order to show the relation of enhanced denitrification rates and these organisms in real wastewater sludge sampled from an operating system.

### 6.3.2 Investigations on membrane fouling

The relation between soluble and bound TEP/EPS has been identified by different research groups to play a major role in membrane fouling. Two main directions for coming research activities can be derived from these results:

- The cause of changing bEPS/sEPS ratios: The identification of microbiological and operational parameters provoking a significant swift of sEPS/bEPS ratios could help to identify new operational guidelines for MBR systems, reducing the impact of fouling events thus leading to a more sustainable operation.
- The identification of substances responsible for fouling: Fundamental investigations with enhanced measurement techniques (e.g. MALDI TOF, LC-OCD) are proposed to determine the fractions of EPS/TEP that are involved in membrane fouling. As EPS/TEP is a composite parameter containing various, partially unknown substances, further investigation to break down this parameter into identifiable fractions is necessary. Once fouling can be directly correlated to one of these fractions, measures can be identified to reduce or avoid these substances, e.g. through optimized coagulants. Also cleaning agents that specifically attack the identified foulants can be development.

### 6.3.3 Perspectives

The ENREM process scheme is a technical solution for decentralized and semi-decentralized when the highest effluent qualities are required. Additionally it was shown in Section 2.14 that the process is economical competitive for plant sizes of 5000 p.e. and larger. The high effluent quality allows also reuse of the effluent for various purposes, which will be a great benefit for arid and semi-arid regions.

The prototype plant showed to be able to close the treatment gap between households systems of 4 – 8 p.e. and installations serving catchment areas of 1000 p.e. The modification of the MBR system with precipitation and a down stream adsorption filter increases the effluent quality in terms of phosphate effluent loadings and therefore can also be used when there is a sensible receiving water body.

# Appendix

- Appendix A: Elimination results of sedimentation trials
- Appendix B: Sedimentation volumes
- Appendix C: Recommendations for plant operation I (German)
- Appendix D: Recommendations for plant operation II (German)
- Appendix E: Membrane cleaning protocol (German)
- Appendix F: Action plan in case of foaming events (German)

# Appendix A

# Elimination results of sedimentation trials

| Date | Sample | COD | TN | TP | Org. acids |
|---|---|---|---|---|---|
| 19.09.2007 | incl. sediment [mg/L] | 1301,0 | 132,0 | 22,5 | 397,0 |
| | after sedimentation [mg/L] | 1004,5 | 135,0 | 22,1 | 338,0 |
| | elemination [%] | 22,8 | 0,0 | 1,8 | 14,9 |
| 19.09.2007 | incl. sediment [mg/L] | 830 | - | 16,34 | 219 |
| | after sedimentation [mg/L] | 754 | 100,4 | 15,18 | 208 |
| | elemination [%] | 9,2 | | 7,1 | 5,0 |
| 30.10.2007 | incl. sediment [mg/L] | 1623,0 | 158,1 | 20,8 | - |
| | after sedimentation [mg/L] | 788,0 | 118,6 | 18,0 | - |
| | elemination [%] | 51,4 | 25,0 | 13,4 | - |
| 31.10.2007 | incl. sediment [mg/L] | 1365,5 | 248,0 | 23,6 | - |
| | after sedimentation [mg/L] | 1080,0 | 190,0 | 21,4 | - |
| | elemination [%] | 20,9 | 23,4 | 9,3 | - |
| 07.11.2007 | incl. sediment [mg/L] | 1040,0 | 154,3 | 23,2 | - |
| | after sedimentation [mg/L] | 933,0 | 137,2 | 22,4 | - |
| | elemination [%] | 10,3 | 11,1 | 3,4 | - |
| 20.11.2007 | incl. sediment [mg/L] | 1964,5 | 144,8 | 17,1 | - |
| | after sedimentation [mg/L] | 829,0 | 137,5 | 17,0 | - |
| | elemination [%] | 57,8 | 5,0 | 0,6 | - |
| 21.11.2007 | incl. sediment [mg/L] | 1663,0 | 40,2 | 12,4 | - |
| | after sedimentation [mg/L] | 995,0 | 144,6 | 17,8 | - |
| | elemination [%] | 40,2 | | | - |
| 22.11.2007 | incl. sediment [mg/L] | 1080,0 | 182,8 | 22,2 | - |
| | after sedimentation [mg/L] | 778,0 | 171,0 | 17,9 | - |
| | elemination [%] | 28,0 | 6,5 | 19,3 | - |
| 11.12.2007 | incl. sediment [mg/L] | 1468,0 | 154,8 | 25,0 | 485,0 |
| | after sedimentation [mg/L] | 1198,0 | 167,4 | 22,4 | 287,0 |
| | elemination [%] | 18,4 | 0,0 | 0,0 | 40,8 |
| 11.12.2007 | incl. sediment [mg/L] | | 125,2 | 19,5 | 390,0 |
| | after sedimentation [mg/L] | | 132,0 | 19,7 | 376,0 |
| | elemination [%] | | 0,0 | 0,0 | 3,6 |
| 04.01.2008 | incl. sediment [mg/L] | - | - | - | 475,0 |
| | after sedimentation [mg/L] | - | - | - | 447,0 |
| | elemination [%] | - | - | - | 5,9 |
| 11.01.2008 | incl. sediment [mg/L] | - | - | - | 242,0 |
| | after sedimentation [mg/L] | - | - | - | 267,0 |
| | elemination [%] | - | - | - | 0,0 |
| 11.01.2008 | incl. sediment [mg/L] | - | - | - | 350,0 |
| | after sedimentation [mg/L] | - | - | - | 350,0 |
| | elemination [%] | - | - | - | 0,0 |
| 11.01.2008 | incl. sediment [mg/L] | - | - | - | 332,0 |
| | after sedimentation [mg/L] | - | - | - | 321,0 |
| | elemination [%] | - | - | - | 3,3 |
| 15.01.2008 | incl. sediment [mg/L] | 1233,0 | 342,0 | 19,2 | 305,0 |
| | after sedimentation [mg/L] | 988,0 | 328,0 | 19,2 | 268,0 |
| | elemination [%] | 19,9 | 4,1 | 0,0 | 12,1 |
| 15.01.2008 | incl. sediment [mg/L] | - | 170,2 | - | 429,0 |
| | after sedimentation [mg/L] | - | 170,0 | - | 400,0 |
| | elemination [%] | - | 0,1 | - | 6,8 |
| 15.01.2008 | incl. sediment [mg/L] | - | 150,6 | - | 339,0 |
| | after sedimentation [mg/L] | - | 156,0 | - | 363,0 |
| | elemination [%] | - | 0,0 | - | 0,0 |
| | ***average elemination:*** | ***27,9*** | ***6,8*** | ***5,5*** | ***5,2*** |
| | ***standard deviation:*** | ***17*** | ***9*** | ***7*** | ***5*** |

# Appendix B
# Sedimentation volumes

| sample | sample volume [mL] | Sediment [mL/L] | | |
|---|---|---|---|---|
| | | 10 min | 20 min | 30 min |
| 1 | 1000 | 41,00 | 35,00 | 34,00 |
| 2 | 1000 | 35,00 | 34,00 | 34,00 |
| 3 | 1000 | 48,00 | 37,00 | 36,00 |
| 4 | 1000 | 26,00 | 22,00 | 22,00 |
| 5 | 1000 | 20,00 | 19,00 | 19,00 |
| 6 | 1000 | 21,05 | 17,89 | 17,89 |
| 7 | 1000 | 46,49 | 38,92 | 37,84 |
| 8 | 1000 | 36,22 | 30,27 | 30,27 |
| 9 | 1000 | 88,89 | 72,22 | 71,11 |
| 10 | 1000 | 134,33 | 107,46 | 105,97 |
| 11 | 1000 | 10,34 | 9,48 | 9,48 |
| 12 | 1000 | 9,50 | 9,50 | 9,50 |
| 13 | 1000 | 33,57 | 28,67 | 27,97 |
| 14 | 1000 | 20,57 | 18,29 | 18,29 |
| 15 | 1000 | 9,00 | 11,00 | 10,50 |
| 16 | 1000 | 26,32 | 21,05 | 21,05 |
| 17 | 1000 | 23,00 | 21,50 | 21,50 |
| 18 | 1000 | 20,98 | 19,58 | 19,58 |
| 19 | 1000 | 70,86 | 62,86 | 54,86 |
| 20 | 1000 | 23,00 | 20,00 | 19,00 |
| 21 | 1000 | 7,37 | 7,89 | 7,89 |
| 22 | 1000 | 24,21 | 21,05 | 21,05 |
| 23 | 1000 | 8,00 | 9,00 | 9,00 |
| 24 | 1000 | 66,00 | 62,00 | 60,00 |
| 25 | 1000 | 44,00 | 37,00 | 37,00 |
| 26 | 1000 | 21,00 | 20,00 | 20,00 |
| 27 | 1000 | 15,00 | 14,50 | 14,50 |
| 28 | 1000 | 25,00 | 22,00 | 22,00 |
| 29 | 1000 | 61,1 | 44,2 | 44,2 |
| 30 | 1000 | 9,0 | 9,0 | 10,0 |
| 31 | 1000 | 40,0 | 36,0 | 35,0 |
| 32 | 1000 | 6,0 | 9,0 | 10,0 |
| 33 | 1000 | 55,0 | 46,0 | 42,0 |
| 34 | 1000 | 8,0 | 8,5 | 9,0 |
| 35 | 1000 | 13,0 | 13,0 | 13,0 |
| 36 | 1000 | 65,0 | 58,0 | 50,0 |
| 37 | 1000 | 2,7 | 3,8 | 4,6 |
| 38 | 1000 | 22,0 | 21,0 | 21,0 |
| 39 | 1000 | 5,4 | 6,5 | 6,8 |
| 40 | 1000 | 8,5 | 9,5 | 10,0 |
| 41 | 1000 | 28,0 | 25,0 | 25,0 |
| 42 | 1000 | 40,0 | 33,0 | 33,0 |
| 43 | 1000 | 5,0 | 3,0 | 4,0 |
| 44 | 1000 | 27,0 | 25,0 | 25,0 |
| 45 | 1000 | 96,0 | 76,0 | 70,0 |
| 46 | 1000 | 81,0 | 64,0 | 59,0 |
| 47 | 1000 | 290,0 | 225,0 | 195,0 |
| 48 | 1000 | 12,0 | 12,0 | 12,0 |
| 49 | 1000 | 120,0 | 96,0 | 93,0 |
| 50 | 1000 | 5,5 | 6,5 | 7,0 |
| 51 | 1000 | 5,5 | 6,0 | 7,0 |
| 52 | 1000 | 26,0 | 23,0 | 23,0 |
| 53 | 1000 | 56,00 | 48,00 | 48,00 |
| 54 | 1000 | 8,00 | 8,25 | 8,50 |
| 55 | 1000 | 11 | 11 | 11,5 |
| 56 | 1000 | 105,0 | 78,0 | 74,0 |
| 57 | 1000 | 34,0 | 32,0 | 31,0 |
| 58 | 1000 | 44,0 | 42,0 | 41,0 |
| 59 | 1000 | 16,0 | 16,0 | 16,0 |
| 60 | 1000 | 12,0 | 14,0 | 14,0 |
| 61 | 1000 | 19,0 | 18,0 | 17,0 |
| 62 | 1000 | 17,0 | 15,0 | 15,0 |
| 63 | 1000 | 22,4 | 22,4 | 22,4 |
| 64 | 1000 | 8,6 | 10,3 | 10,3 |
| 65 | 1000 | 18,1 | 16,9 | 16,9 |
| 66 | 1000 | 25,0 | 22,0 | 22,0 |
| 67 | 1000 | 16,0 | 16,0 | 16,0 |
| 68 | 1000 | 46,0 | 35,0 | 34,0 |
| | **average:** | **31,44** | **27,40** | **26,69** |
| | **standard deviation:** | **25,89** | **20,16** | **18,82** |

# Appendix C

## Recommendations for plant operation I (German)

| | Bei jedem Besuch | | ca. 1x pro Woche | | ca. 1x pro Monat | | Nach Bedarf | |
|---|---|---|---|---|---|---|---|---|
| **Allgemein** | A11 | Bei Betreten und Verlassen des Geländes Schaltwarte benachrichtigen | A21 | B1 bzw. B2 absaugen lassen (bevorzugt freitags) | A31 | B1 komplett absaugen (Rechengut am Boden) | A41<br>A42<br>A43<br>A44<br>A45<br><br><br>A46 | Eisenchloridlösung aus der OWA Beelitzhof abholen<br>Probenehmer stellen<br>Reagenzien und Materialien bestellen<br>Messungen am Zulauf<br>EINMAL JÄHRLICH:<br>- Wartung Klimagerät<br>- Überprüfung Drehkran<br>Überprüfung des Zulaufmessschachtes auf Wassereintritt |
| **Klärchen** | K11<br>K12<br>K13<br>K14 | Anlagenfunktion überprüfen<br>Zonen auf Schaum-entstehung überprüfen<br>Messwerte der Sonden überprüfen<br>Schichtbuch schreiben (PC) | K21<br>K22<br>K23<br>K24<br>K25 | Füllstand Fällmittellösung und Essigsäure überprüfen und ggf. neu ansetzen<br>TS messen, Vergleich mit Sonde, in Datei „Handdaten Klärchen" notieren<br>Filtratpumpe, die nicht in Betrieb ist, kurzzeitig einschalten (FP1 bzw. FP3)<br>Stromzählerstände notieren<br>Reagenzien für $PO_4^{3-}$-Analysator überprüfen/ggf. nachfüllen | K31<br>K32<br>K33 | Membranen reinigen und Membran-kammerwechsel<br>Schläuche des $PO_4$-P-Analysators begutachten, ggf. reinigen/ austauschen<br>Funktion der Schaumsonde überprüfen | K41<br>K42<br>K43<br>K44<br>K45<br>K46 | Sonden reinigen<br>Messungen am Ablauf<br>Bei Unregelmäßigkeiten: Messungen in den Kammern (besonders $NH_4^+$ in AX0/AX1)<br>$O_2$-Sonden überprüfen, ggf. kalibrieren (alle 3 Monate)<br>Grobstoffzelle ablassen (ca. alle 2 Wochen)<br>Luftfilterkontrolle Gebläse & Zusatzgebläse für Belüftung der Ablaufrinne |
| **Busse-Anlage** | B11<br>B12<br>B13 | Anlagenfunktion überprüfen<br>ÜSS manuell ablassen (ca. 1-3 x pro Woche) und notieren<br>Anlagenschichtbuch führen (handschriftlich) | B21<br>B22<br>B23<br><br><br><br>B24<br>B25 | TS messen<br>Füllstand Fällmittellösung überprüfen und ggf. neu ansetzen<br>Messungen<br>- $O_2$ während Belüftungsphase<br>- pH<br>- Pt vor und nach Adsorption<br>Stromzählerstände notieren<br>Durchfluss Filtrat und Betriebs-stunden notieren | B31 | Austausch Adsorbens (je nach Ablauf-qualität) | B41<br>B42<br>B43<br>B44 | Membranwechsel, Wartung (Fremdfirma)<br>Messungen am Ablauf<br>Luftfilterkontrolle Gebläse<br>Dosierschlauch für Fällmittel kontrollieren (Verstopfungen,..) |

# Appendix D

## Recommendations for plant operation II (German)

ERLÄUTERUNGEN / KOMMENTARE

A11: Telfonnummer: 8644-90669
A21: Vorher die Kanalreinigung benachrichtigen (so früh wie möglich)
A41: Leere Kanister zum Auffüllen und blaue Säcke o. ä. als Transportschutz mitnehmen
A42: Probenahme auf 1 Tag vor Probenabholung stellen (24h-Probe)
A43: z.B. Dr.Lange Testkits, Pipettenspitzen, etc.
A46: Bei stehendem Wasser im Zulaufmessschacht muss dieses abgepumpt werden (Gefährdung der elektronischen Messgeräte)

K12: Bei Schaumauftritt nach Aktionsplan handeln, Messungen vornehmen
K13: Bei Abweichungen von Sollwerten Messwerte mit Testkits bzw. mit O2 oder pH-Handmessgeräten überprüfen
K21: Füllstände in Datei „Handdaten Klärchen" notieren,
zum Nachfüllen des Fällmittels 40%ige $FeCl_3$-Lösung (gelagert bei Busse) und Wasser im Verhältnis 1: 9 mischen, Fasspumpe benutzen,
zum Nachfüllen der Essigsäure-Lösung 60%-ige Essigsäure (gelagert bei Klärchen) und Wasser im Verhältnis 1:1 mischen, Fasspumpe benutzen
K22: Mikrowellenkochmethode, ggf. Sonde neu kalibrieren (3-Punkt-Kalibrierung)
K23: Saugseite geschlossen, ohne Medium laufen lassen. Vor-Ort-Kontrolle ob Welle dreht. Mit Inbetriebnahme soll Festsetzen verhindert werden
K24: Inklusive des Straßenverteilers außerhalb des Geländes; Werte in die Datei „Handdaten Strom" eintragen
K25: Nachbestellung der Reagenzien jeweils, wenn letzte Flasche angebrochen wird
K31: Reinigungsprotokoll führen, in Betrieb sind alternierend Kammer 1 und 3
K33: Optische Kontrolle vor Ort: Sind Verkrustungen an den Fühlern? ggf. Reinigung
K41: Trübungsmesser, Nitratsonde, TS-Sonde, ...
K43: Proben dazu erst filtrieren (erst grob, dann fein)
K44: Kalibrierung der $O_2$-Sonden laut Ereignisprotokoll vom 28.8.2007

B12: In Abhängigkeit vom TS-Wert. Überschussschlamm in den Fäkalienspeicher B2 pumpen (über Schlauchanschluss an Außenseite des Containers, anschließend Leitung mit Wasser spülen, Schlauchanschluss im Winter nach jeder Spülung mit ausreichend Frostschutzmittel füllen!)
B21: Mikrowellenkochmethode
B22: Füllstand notieren (Datei „Handdaten Busse"), zum Nachfüllen 40%ige $FeCl_3$-Lösung (gelagert bei Busse) und Wasser im Verhältnis 1: 1 mischen
B23: Werte in Datei „Handdaten Busse" eintragen
B24: Stromzählerstand im Baustromverteiler und bei der SPS notieren, Werte in Datei „Handdaten Strom" eintragen
B25: In die Datei „Handdaten Busse" eintragen
B31: Wechsel bei [P] > 1 – 1,5 mg/l, je nach Zulaufkonzentration vor Adsorptionssäule
B41: ca. alle 4 – 6 Monate
B43: ca. alle 3 – 4 Monate

# Appendix E

## Membrane cleaning protocol (German)

| Nr. | Thema |
|---|---|
| 1 | Reinigungsstrategie A3-Filter |
| 2 | - Die 2 A3-Module sollen im Wechsel jeweils 2 Monate im Betrieb bleiben, anschließend gereinigt werden und darauf 2 Monate in stand-by verbleiben<br>- Die Filter verbleiben zur Reinigung in den jeweiligen Membranreaktoren<br>- Filter 4 befindet sich in MR1, Filter 5 in MR3<br>- Die Reinigung soll einstufig (nach Bedarf zweistufig mit Zitronensäure) mit niedriger Konzentration und ohne Ablassen der Reinigungslösung durchgeführt werden, der Filter verbleibt bis zur nächsten Inbetriebnahme in der Reinigungslösung |
| 3 | Reinigungsprotokoll ab dem 20.08.2009<br>1. Vor Außerbetriebnahme: Ruhedruck sowie weitere Betriebswerte im Schlamm bei 5,3 l/min und 10,6 l/min notieren -> siehe Datenprotokoll Reinigung<br>2. MR über Ausspiegeln mit 2. MR und Beschickungspumpe entleeren; komplett in AE zurückpumpen (Schlammalter einhalten!!), mit Trinkwasser nachspülen und in B2 leiten; Beim Entleeren und Befüllen Schlauchleitung Saugseite Filter für Be- und Entlüftung trennen!<br>3. MR mit Trinkwasser füllen<br>4. 10 min starke Belüftung ohne Filtration<br>5. (Einmaliges Rückspülen über in-situ-Reinigung bis 50 mbar Überdruck erreicht sind, anschließend) optional!!!<br>6. 10 min starke Belüftung ohne Filtration und ohne erneutes Rückspülen optional!!!<br>7. a) Filter 4 (MR1) mit 1000 ppm H2O2 ansetzen: 1,75 L H2O2 35% auf 675 L pH 11 einstellen: in mehreren Schritten ca. 2500 ml 1M NAOH in MR geben, dabei zwischen jedem Schritt pH-Wert oben im MR nachmessen, nachdem die Belüftung zur Durchmischung kurzzeitig angestellt war (Schaumbildung!)<br>!!!!!!!!!pH-Wert 11 nicht überschreiten!!!<br>b) Filter 5 (MR3) mit 500 ppm Cl ansetzen: 2,13 L NaOCl 13% auf 675 L<br>8. Belüftung (nicht bei NaOCl, schäumt stark!) und Filtration zur Durchmischung ca. 15 min ein (Rezirkulation), Drücke im Filtrat aufnehmen (siehe Punkt 1)<br>9. <u>Nur Reinigung H2O2:</u><br>ca. 200-400 ml 1M NAOH nachdosieren, um pH-Wert 11 erneut einzustellen (nach Filtration ca. pH 10,7 in MR, 10,3 in Filtrat)<br>10. Nach Absprache: nach 1 Woche Reinigungslösung ablassen und Reinigung mit Zitronensäure (5000 ppm), anschließend siehe Punkt 8<br>11. Am Folgetag pH- und Betriebswerte im Filtrat erneut aufnehmen (siehe Punkt 1)<br>12. IDM reinigen (durch Spülen oder Ausbau und mech. Reinigung), um Fehlmessungen zu verhindern<br>13. Anschließend Filtratpumpe mit Wasser spülen und absperren, ggf. konservieren, um Korrosion zu minimieren<br>14. Modul ohne Belüftung in Reinigungslösung aufbewahren<br>15. Filtratpumpe 1x Woche anfahren, um Festsetzen zu verhindern |
| 4 | Vor / Bei Inbetriebnahme:<br>16. pH-Wert und Betriebswerte in Reinigungslösung aufnehmen (siehe Punkt 1)<br>17. Besonders bei der Chlorreinigung den MR und die Membran ausreichend mit Wasser spülen, um Chloreintritt in Anlage und damit AOX-Bildung zu verhindern<br>18. Werte im Schlamm aufnehmen (siehe Punkt 1) und Datenprotokoll abschließen |

# Appendix F

# Action plan in case of foaming events (German)

Folgenden Aktionen sind, je nach Schaumaufkommen, zu empfehlen:

1) Belüftung aus, Schlamm sich setzen lassen, Deox-Topf prüfen: wenn dieser mit Schaum gefüllt ist, Deox-Pumpe überprüfen

2) Schaum auf den Kammern mit Nasssauger absaugen und nach B2 verwerfen (kann Giftstoffe und schaumbildende Bakterien enthalten)

3) Übergelaufenen Schaum entfernen (nach B2), Flächen reinigen/abspritzen (Schaufel/Schippe/Saugewagen; Nasssauger eignet sich nur, wenn er nach jeder Befüllung gereinigt wird → Schwimmerschalter)

4) Schaum mit so wenig Wasser wie möglich nieder spritzen (Beeinträchtigung der Durchsatzmenge)

5) Tauchpumpen in schäumende Behälter einbauen, um Durchmischung und Verrieselung des Schlammes zu gewährleisten (dadurch Niederschlagen des Schaums mit Eigenmedium)

6) Zulaufstrom reduzieren; RLS-Pumpe ausschalten, wenn Anlage außer Betrieb genommen wird (ansonsten Eindickung AN)

7) Messungen durchführen (besonders Nitrit und Ammonium sollten überwacht werden, wenn viel Ammonium vorhanden ist, muss ausreichend Belüftung gewährleistet werden → zusätzliche Zone belüften

8) Belüftung variieren, ggf. Zoneneinteilung verändern (belüftet/unbelüftet, abhängig von Messwerten)

9) Abdeckungen auf Membrankammern legen (Styrodur), Schaum gezielt in MR2 (leere Kammer) laufen lassen

10) TS überwachen (fällt durch Überschäumen meist stark ab)→ es gibt Alarm durch die Schaumsonde bei Abweichungen vom Sollwert, bei gleichzeitigem Auslösen der Schaumsonde wird Anlage automatisch ausgeschaltet → Alarm „Anlagenstillstand Schaum“

11) ÜSS aus Busse-Anlage (bei Normalbetrieb ohne Probleme) zu Klärchen leiten, um TS zu erhöhen

12) Übergelaufenen Schlamm (z.B. in MR 2) verwerfen, wenn er älter als 24 h ist, sonst über RLS-Pumpe nach AN fördern

13) Bei geringer Schlammtemperatur Containerheizung einschalten → Erhöhung Schlammtemperatur um ca. 1-3 °C, je nach Außentemperatur

14) Auf die Anwendung von Anti-Schaummittel verzichten, da dadurch zeitlich versetztes verstärktes Schäumen auftritt

15) Zulauf analysieren (CSB, Schwermetalle,…), gibt es Hinweise auf eventuelle Einleitungen?

16) Bei Verdacht auf schädliche Einleitungen das Abwasser aus B1 direkt nach B2 leiten und nicht in die Anlage gelangen lassen → Abfuhr

17) Tabellen (u.a. Tabelle Schwimmschlamm) in Datei „Handdaten Klärchen“ aktualisieren

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