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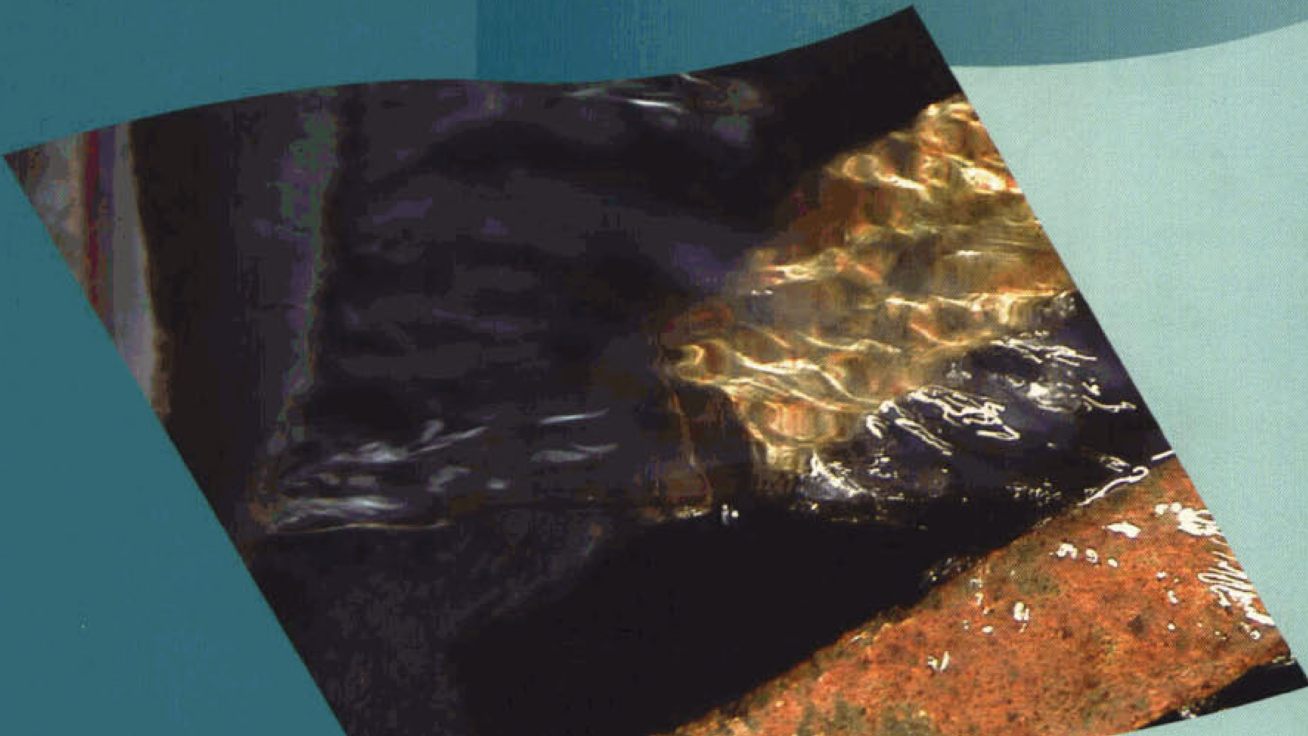
MBR for municipal wastewater treatment

Pilot plant research Beverwijk WWTP



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2002 11A

Publicaties en het publicatie-
overzicht van de STOWA kunt u
uitsluitend bestellen bij:
Hageman Fulfilment
Postbus 1110
3330 CC Zwijndrecht
tel. 078 - 629 33 32
fax 078 - 610 42 87
e-mail: hff@wxs.nl
o.v.v. ISBN- of bestelnummer en
een duidelijk afleveradres.
ISBN 90.5773.169.X

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TEN GELEIDE

De membraan-bioreactor (MBR) vormt een belangrijke vooruitgang voor de zuivering van huishoudelijk afvalwater. In vergelijking tot conventionele zuiveringstechnieken wordt op een aanzienlijk geringer oppervlak een betere effluentkwaliteit geproduceerd. Thans doet zich in Nederland de kans voor om deze technologie door te laten breken en verder te ontwikkelen tot een volwassen techniek. Daarvoor is binnen de branche een breed gedragen ontwikkelingstraject in gang gezet. Het ontwikkelingstraject bestaat uit de uitvoering van een vergelijkend onderzoek op pilotschaal, de realisatie van een demonstratie-installatie en twee tot drie grootschalige rwzi's. Het gaat hierbij om een opschaling die wereldwijd nog niet is gerealiseerd en een geweldige uitdaging betekent voor de betrokken waterbeheerders, adviseurs en leveranciers.

Voorliggend eindrapport beschrijft de resultaten van het vergelijkend pilot-onderzoek dat op de rwzi Beverwijk door DHV Water is uitgevoerd. Dit onderzoek is begin 2000 gestart door het Hoogheemraadschap van Uitwaterende Sluizen en na een half jaar overgenomen door de STOWA om zorg te dragen voor voldoende representativiteit voor de Nederlandse waterbeheerders. Onder de Nederlandse omstandigheden die spelen bij het zuiveren van huishoudelijk afvalwater, zijn vier membraaninstallaties (Kubota/Solis, Mitsubishi/Grontmij, X-Flow/Nuon en Zenon) vergeleken op hun prestaties. In het onderzoek is onder meer aandacht besteed aan de membraanwerking in relatie tot de biologische zuivering en optredende vervuiling, de haalbare effluentkwaliteit, en bedrijfsvoeringsaspecten. In een aantal deelonderzoeken is verder aandacht besteed aan de benodigde voorbehandeling, membraanvervuiling en -reiniging, energieverbruik, effluentkwaliteit en slibverwerking.

Aan de totstandkoming van dit tussenrapport hebben veel organisaties meegewerkt, welke in bijlage 1 zijn weergegeven. De goede samenwerking tussen al deze organisaties ligt mede ten grondslag aan de gedegenheid waarmee dit hoog ambitieuze onderzoek is uitgevoerd. De heer H.F. van der Roest van DHV Water B.V. was in deze de centrale spil. Het DHV projectteam bestond verder uit D.P. Lawrence, A.G.N. van Bentem, P. van der Herberg en R.M.W. Kraan. Voor de begeleiding van het project uit de STOWA zorgde een commissie bestaande uit ing. A.A.J.C. Schellen (voorzitter), ir A.H. Dirkzwager, ir K.F. de Korte, dr. J. Kruithof, ir J.W. Mulder, ir P.J. Roeleveld, ir R.M.W. Schemen, ir P.F.T. Schyns, ing. J.G. Segers, ir P.I.M. Walet en ing. H.J. Ellenbroek.

Naast de STOWA en Uitwaterende Sluizen hebben DWR/AGV, ZHEW, Regge en Dinkel en RIZA financieel bijgedragen aan het gehele onderzoekstraject op de rwzi Beverwijk. De STOWA is Uitwaterende Sluizen zeer erkentelijk voor het beschikbaar stellen van een onderzoekslocatie. Bijzondere dank gaat uit naar de medewerkers van de rwzi Beverwijk en de SDI Beverwijk voor hun gastvrijheid tijdens het onderzoek en de nodige excursies.

Utrecht, maart 2002

De directeur van de STOWA

ir J.M.J. Leenen

SAMENVATTING

Het membraan-bioreactor (MBR) concept is een biologisch proces gecombineerd met membraanfiltratie. Recente technologische ontwikkelingen en een significante daling van de membraanprijs heeft de inzetbaarheid van MBR-technologie voor de behandeling van huishoudelijk afvalwater aanzienlijk vergroot. Met name in Nederland, waar stedelijke en industriële gebieden vaak zijn gelegen nabij gevoelige oppervlaktewateren, toont MBR-technologie meerdere voordelen ten opzichte van het conventionele actiefslib proces, zoals de hoge effluentkwaliteit, het beperkte ruimtegebruik en de mogelijkheid voor een flexibele en gefaseerde uitbreiding van bestaande rioolwaterzuiveringsinstallaties (rwzi's).

In het kader van de ontwikkeling van MBR-technologie op mondiaal niveau, zijn meerdere haalbaarheidsstudies verricht naar de toepasbaarheid van het MBR-concept op rwzi's in Nederland. De hieruit voortvloeiende positieve conclusies zijn voorgelegd aan de Nederlandse waterbeheerders, op basis waarvan vorm is gegeven aan een onderzoeksprogramma naar de toepassing van het MBR-concept onder typisch Nederlandse omstandigheden wat betreft influentkwaliteit en debietvariaties. In eerste instantie in opdracht van het Hoogheemraadschap USHN en vervolgens van de STOWA, is een grootschalig pilotonderzoek uitgevoerd.

Het doel van het pilotonderzoek was voornamelijk de ontwikkeling van MBR-technologie. Tevens was de technische ontwikkeling en opschaling van het MBR-concept van grote importantie. Het kader van het pilotonderzoek was als volgt gedefinieerd:

- een vergelijking van verschillende membraansystemen;
- het samenwerken met membraanleveranciers;
- het verzamelen van de mondiaal aanwezige kennis;
- uitvoeren van een onderzoek representatief voor de Nederlandse omstandigheden;
- uitvoeren van een onderzoek op relevante schaal.

Op basis van dit kader zijn de volgende onderzoeksdoelen gedefinieerd:

- onderzoek naar de technische haalbaarheid van het MBR-systeem voor de Nederlandse situatie;
- vergelijken van verschillende beschikbare membraansystemen;
- vergelijken van het MBR-concept en een conventioneel systeem met betrekking tot ontwerpaspecten, kosten, energieverbruik en effluentkwaliteit;
- verder ontwikkelen van MBR-technologie en het elimineren van onzekerheidsfactoren;
- onderzoeken van de effecten van verschillende voorbehandelingsmethoden en verschillen in afvalwatersamenstelling op de membraanwerking;
- onderzoeken van langetermijneffecten;
- integreren van bestaande kennis van actiefslibtechnologie met membraantechnologie;
- onderzoeken van de effecten van lage temperaturen op de membraanwerking.

Gebaseerd op de mondiale MBR-markt en relevante ervaringen met MBR-technologie zijn vier membraanleveranciers geselecteerd om binnen het pilotonderzoek op de rwzi Beverwijk te participeren. De leveranciers zijn: Kubota (Japan), Mitsubishi (Japan), X-Flow (Nederland) en Zenon (Canada). Per leverancier is op het terrein van de rwzi Beverwijk een pilotinstallatie geplaatst die representatief dient te zijn voor opschaling naar een full-scale installatie. Deze pilotinstallaties zijn bedreven en geoptimaliseerd in de loop van het onderzoek om de beste procesvoering onder vier verschillende procesregimes te bereiken voor de lozingseisen: N_{totaal} en P_{totaal} van respectievelijk <10 mg/l en <1 mg/l.

In fase 1 van het onderzoek heeft pre-precipitatie in de voorbezinktanks plaats gevonden om de hoeveelheid aan colloïdaal materiaal in afvalwater te reduceren welke nadelig kan zijn voor de membraanwerking. In fase 2 is voorbezinking in combinatie met simultane fosfaatverwijdering toegepast en in fase 3 is overgegaan naar ruw influent (na een microzeef) gevolgd door simultane fosfaatverwijdering. In fase 4 is de chemicaliëndosering gestopt en is de mogelijkheid van ruw influent in combinatie met biologische fosfaatverwijdering onderzocht. In alle fases zijn de pilotinstallaties onderworpen aan normale proportionele debietsvariaties zoals die zich op Nederlandse rwzi's voordoen.

Van het initiële concept dat membranen en bioreactoren als twee onafhankelijk van elkaar opererende systemen worden gezien, is snel afgestapt. De beide procesonderdelen blijken in die mate interactief dat het hele MBR-proces als één integraal systeem dient te worden beschouwd. Vanuit dit oogpunt zijn verschillende relevante deelstudies uitgevoerd ter optimalisatie van parameters en randvoorwaarden die het MBR-proces beïnvloeden. Deze deelstudies zijn: voorbehandeling, membraanreiniging, α-factor en energieverbruik, effluentkwaliteit en slibverwerking.

De biologische prestatie van de vier systemen was afhankelijk van de influentkwaliteit in fase 1, 2, 3 en 4 en de biologische slibbelasting. Nadat de biologische werking van elke pilotinstallatie in kaart was gebracht, bleek dat de biologische prestaties van de MBR-systemen vergelijkbaar zijn met die van een conventioneel actiefslibsysteem bij eenzelfde belasting en debietvariaties. Veel van de uitgangspunten en ervaringen opgedaan met conventionele technieken blijken toegepast te kunnen worden op het MBR-actiefslibsysteem.

De membraanwerking is afhankelijk van de biologische prestatie en andersom. De actiefslibeigenschappen bepalen voor een groot deel de membraanwerking. Elk membraansysteem is tijdens het onderzoek geoptimaliseerd hetgeen voor elk systeem een gedefinieerd werkingsgebied heeft voortgebracht. De werkingsgebieden zijn opmerkelijk verschillend voor de verschillende systemen, maar elk systeem kon met goede resultaten worden toegepast.

De vooraf gestelde doelen met betrekking tot het pilotonderzoek op de rwzi Beverwijk zijn meer dan bereikt. Na bijna twee jaar intensief onderzoek en verdere ontwikkeling van de technologische haalbaarheid van de MBR voor de Nederlandse situatie, is het nu mogelijk de stap naar een demonstratie-installatie te zetten. De stap van pilotschaal naar toepassing kan worden gemaakt, waarbij verdere technologische optimalisaties mogelijk blijven.

Een vergelijking tussen de vier pilotinstallaties wordt gedetailleerd weergegeven in de evaluatie van dit rapport. Het is duidelijk dat de vier installaties op het gebied van biologische- en membraanwerking verschillen. Tijdens het onderzoek opereerden Kubota en Zenon overwegend op een proportioneel debiet, terwijl Mitsubishi en X-Flow hoofdzakelijk op constant debiet draaiden. Daardoor is een vergelijking tussen de systemen omtrent piekflux en continue flux moeilijk te maken. De bedrijfsvoering van elk systeem was omschreven binnen bepaalde grenzen met betrekking tot de sturing en het ontwerp. Binnen deze grenzen bleken alle systemen goed en consistent te presteren.

Veel van de onzekerheden binnen de MBR-technologie zijn tijdens het pilotonderzoek onderzocht. Van belang is met name de vervuiling van de membraanmodules door bijvoorbeeld haren en grove bestanddelen, en het effect van chemische reiniging op de biologische prestatie van de systemen. Daarnaast is inzicht verkregen in het effect van lage procestemperaturen op de membraanwerking. Alle risico's die vooraf aan het onderzoek waren gedefinieerd zijn onderzocht en geëlimineerd of gereduceerd tot een acceptabel niveau.

De voorbehandeling van het afvalwater heeft de belangrijkste bijdrage geleverd aan de goede bedrijfsvoering en werking van de membraansystemen. De vereiste voorbehandeling is beduidend intensiever dan benodigd bij een conventioneel systeem. Onjuiste voorbehandeling van het afvalwater heeft een negatieve invloed op het membraan. Met de juiste voorbehandeling kan verontreiniging van het membraan op macroniveau worden voorkomen, waarbij het membraan beter zal presteren. Het ontwerp van de voorbehandeling was gedefinieerd als één van de meest kritische ontwerpaspecten voor een MBR-installatie.

Zowel de langetermijneffecten van de biologische processen als van de benodigde reinigingsprocedures ten behoeve van het onderhouden van de membranen, zijn onderzocht. Deze onderzoeken hebben weinig inzicht verschaft in de verwachtte levensduur van membranen. Wel bleken de wijze en frequentie van chemische reiniging en de toegepaste concentraties kritische factoren te zijn. Indien werd afgeweken van de standaard membraanspecificaties leidde dit tot serieuze verslechtering van de membraanlevensduur. Daarom zijn gedurende het onderzoek de intensieve reinigingsprocedures omgezet in frequentere maar aanzienlijk minder geconcentreerde onderhoudsprocedures.

Op dit moment zijn Kubota en Zenon goed ontwikkeld en kunnen zij binnen een gedefinieerd werkingsgebied betrouwbaar worden toegepast voor de behandeling van huishoudelijk afvalwater. Ook de systemen van Mitsubishi en X-Flow hebben een aanzienlijke progressie doorgemaakt. Uitgaande van de verkregen resultaten is de verwachting dat bij constante permeaatonttrekking en optimale procesvoering en reinigingsprocedures, ook voor deze systemen betrouwbare installaties kunnen worden gerealiseerd, ofschoon verder onderzoek noodzakelijk wordt geacht om deze verwachtingen te kunnen onderbouwen.

De mondiaal aanwezige kennis op het vlak van huishoudelijke MBR-technologie was uitermate summier te noemen en vaak gericht op specifieke, voor een bepaald land geldende, soorten afvalwater en aanvoerkarakteristieken. De geteste procescondities aangaande de Nederlandse situatie zijn uniek in vergelijking tot andere onderzoeken en hebben geleid tot nieuwe inzichten en methodes voor de optimalisatie van MBR-systemen. Het onderzoek op de rwzi Beverwijk was 'state of the art' en een uitbreiding van de reeds beschikbare kennis verkregen uit onderzoek in andere landen.

De ontwikkeling van MBR-technologie kan alleen doorgaan wanneer de financiële haalbaarheid wordt gewaarborgd. Gebaseerd op het pilotonderzoek en de haalbaarheidsstudies kan worden geconcludeerd dat een MBR-installatie op dit moment al financieel haalbaar is als specifieke effluenteisen worden gesteld. Naar verwachting zal de markt voor de toepassing van MBR-installaties in Nederland en daarbuiten snel groeien. In de nabije toekomst zal, in aanwezigheid van een volwassen membraanmarkt, voor veel rwzi's het MBR-concept een haalbaar alternatief zijn met een duidelijke toegevoegde waarde.

PREFACE

The membrane bioreactor (MBR) forms an important advancement in the treatment of municipal wastewaters. In comparison to conventional treatment techniques the MBR is extremely compact and produces a better effluent quality. Presently in the Netherlands there exists the chance for this technology to breakthrough and develop further into a mature technique, therefore within the sector a broad and comprehensive development plan has been initiated. The development plan constituted the execution of a comparable research programme at pilot scale, the realisation of a demonstration installation and two or three large scale WWTP's. The latter proliferated a scale up criteria as yet not envisaged or realised, this supplied a great challenge to all parties involved, the water boards, advisors, and suppliers.

The following report describes the results from the comparative pilot research carried out by DHV Water on location at WWTP Beverwijk. This research was initiated begin 2000 by the Water board Uitwaterende Sluizen and after six months was taken over by the STOWA who represented the interests of all the Dutch Water boards. Under typical Dutch municipal wastewater conditions four membrane bioreactor installations were compared, Kubota/Solis, Mitsubishi/Grontmij, X-Flow/Nuon and Zenon. The research was amongst others directed towards the functionality of the membrane, the biological treatment, membrane fouling, achieved effluent quality, and system operability. In a number of side studies the required pre-treatment, membrane fouling/cleaning, energy usage, effluent quality and sludge processing were also addressed.

This final report was made possible via the co-operation of many organisations and individuals. The excellent co-operation between all parties was built upon a solid foundation of professionalism through which this highly ambitious research could progress. H.F. van der Roest of DHV Water was the axle in the programme, and supported in the execution by D.P. Lawrence, A. van Bentem, P. van der Herberg and R.M.W. Kraan. The direction of the project came from the STOWA and consisted of the following individuals: A.A.J.C. Schellen (Chairman), A.H. Dirkzwager, K.F. de Korte, J. Kruithof, J.W. Mulder, P.J. Roeleveld, R.M.W. Schemen, P.F.T. Schyns, J.G. Segers, P.I.M. Walet and H.J. Ellenbroek.

Alongside the STOWA and Uitwaterende Sluizen; DWR/AGV, ZHEW, Regge en Dinkel and RIZA have contributed to the financial feasibility of the research plan at WWTP Beverwijk. The STOWA is extremely grateful to Uitwaterende Sluizen for the availability of the research site, and special thanks must be extended to the personnel of the WWTP and SDI Beverwijk for their co-operation, participation and knowledge during the research programme and the necessary site visits.

Utrecht, March 2002 — The director of STOWA

ir. J.M.J. Leenen

SUMMARY

The term membrane bioreactor (MBR) defines a combination of a biological process and membrane separation. Due to recent technical innovations and significant cost reductions, the applicability for the MBR technology in municipal wastewater treatment has sharply increased. Especially in the Netherlands where urban and industrialised areas are frequently located near sensitive surface waters, the MBR technology displays several advantages compared to the traditional activated sludge processes, such as high effluent quality, limited space requirement and possibilities for a flexible and phased extension of existing waste water treatment plants.

In light of the developing MBR technology throughout the world a number of feasibility studies were carried out for the applicability of MBR for full scale municipal wastewater treatment plants in The Netherlands. The positive conclusions of these studies were relayed to the Dutch municipal authorities and a research programme was drawn up to represent the Dutch wastewater characteristic in terms of quality and flow pattern. Firstly under the direction of the Waterboard USHN, followed by STOWA, a large scale pilot study was executed.

The aim of the pilot research was chiefly to develop the MBR technology. The technical development and scaling up of the MBR concept was also of tantamount importance. A pre-requisition for the execution of the pilot study at the Beverwijk WWTP was as follows:

- comparison of different membrane systems;
- co-operation with manufacturers;
- inclusion of international know-how;
- representation for the Netherlands;
- execution of the research at suitable scale.

The above pre-requisitions defined the trials objectives as follows:

- to research the technical feasibility of the MBR system for the Netherlands;
- to make a comparison between the various membrane systems available;
- to make a comparison between the MBR and a conventional system concerning: design aspects, costs, energy consumption and effluent quality;
- to further develop the MBR technology and eliminate uncertainties;
- to investigate the effects of different types of pre-treatment and consequently different wastewater compositions on the membrane performance;
- to investigate the long-term effects;
- to integrate existing know how on activated sludge technology with membrane technology;
- to investigate the effects of low temperatures on the membrane performance.

Based on the world wide MBR market and relevant experiences attained, four membrane suppliers were chosen to be represented at the Beverwijk WWTP. The suppliers were Kubota (Japan), Mitsubishi (Japan), X-Flow (The Netherlands) and Zenon (Canada). Each supplier provided a pilot to represent a suitable scale up to full scale. These pilots were operated and optimised in the course of the research programme to achieve the best operating window under four separate operating regimes for a set discharge criteria of $N_{total} < 10$ mg/l and $P_{total} < 1$ mg/l. Phase 1 was pre-precipitation on the primary clarifier to eliminate colloidal material from the wastewater stream which could be detrimental for membrane performance, phase 2 was pre-clarification followed by simultaneous P removal. Phase 3 used screened raw influent followed by simultaneous P removal and in phase 4 the chemical dosing was stopped and the combination of raw influent with biological phosphorus removal was investigated. In all phases the pilots were subjected to the normal proportional flow pattern seen in typical Dutch WWTPs.

The initial concept of membranes and bioreactor as two separate unit operations was quickly revised during the first month of the trial. The process operations were so interactive that the entire MBR process was considered as a single integral process. From this basis a number of relevant side studies were carried out to optimise the peripheral parameters and conditions effecting the MBR process. These included, pre-treatment, membrane cleaning, α-factor and energy consumption, effluent quality and sludge handling.

The biological performance of the four systems was dependent on the influent quality according to phase 1, 2, 3 or 4 and the flow to treatment. Once the biological operating windows of each pilot was defined the biological performance achieved was comparable to that of a conventional activated sludge system for the same loading and flow regime. Many of the conventional treatment techniques and practises could be applied to the MBR biology.

The membrane performance was dependent on the biological performance and visa versa. The nature of the activated sludge often dictated the membrane performance. Each membrane system was optimised during the study and each yielded a defined operating window. The operating criteria were remarkably different for the various suppliers, but all could be operated with good results.

The objectives originally set for the pilot plant study at the Beverwijk WWTP were more than achieved. After almost two years of intensive research and further development of the technological feasibility of the MBR for the Netherlands, it is now possible to be extended to demonstration scale. The step from pilot scale to application can be made, with the potential for further technological optimisations.

A comparison of the four pilots has been made and is detailed in the report's evaluation. It was clear that the four pilots generated different results, regarding biological and membrane performance. Kubota and Zenon were run at predominately proportional flow and Mitsubishi and X-Flow were run at predominately continuous flow. As a result comparisons of peak flux and continuous operating flux could not adequately be made between the systems. Each system operation was defined within certain limitations of process and design, and within these defined boundaries the system's all performed correctly and consistently.

Many of the uncertainties involved with the MBR technology were considered and tested in the course of the pilot trial. The latter was of particular importance regarding the macro fouling of the modules due to hair and coarse debris and the effect of cleaning chemicals on the biological performance. Also the effect of low operating temperatures on the membrane performance was addressed. All the risks addressed at the onset of the pilot trial were pin-pointed and eliminated where possible or significantly reduced to acceptable operating levels.

The pre-treatment of the waste water was the most significant contributor to the correct operation and performance of the membrane systems. The required pre-treatment is significantly more intensive than for a conventional treatment plant. Poor pre-treatment reflected badly on a membrane, with the correct pre-treatment no macro fouling was seen during the trial, thus enhancing the membrane's availability and ability to perform. The design of the pre-treatment was defined as one of the critical design parameters of a MBR installation.

The long term effects of the biological process as well as the cleaning processes involved with membrane performance maintenance, were investigated. These assessments yielded very little in terms of a membrane life expectancy. What came to light as a critical issue however was the frequency of chemical cleaning and the concentrations used. If the latter was taken out sight of standard module specifications serious deterioration in membrane integrity was determined. Therefore, during the study the intensive cleaning procedures were metamorphosed into more frequent but considerably less intensive maintenance cleaning procedures.

At this moment both the Kubota and Zenon system are well-established and within a the defined operating window reliable for use in the municipal wastewater market. Also the Mitsubishi and X-Flow system have significantly progressed. From results obtained, it is expected that under constant permeate extraction conditions, optimised process control and cleaning procedures, also for these systems reliable installations will be realised, although further test work has to be carried out to prove the expectations.

The available world knowledge regarding municipal MBR was extremely limited and often very specific for a particular countries wastewater and wastewater flow characteristic. The conditions set by the Dutch situation were the first of its kind and lead to new methods of MBR optimisation. The work carried out at Beverwijk WWTP was 'state of the art' and could be used to enhance the work carried out in other countries.

The Dutch MBR development can only proceed if it remains financially feasible. Based on the pilot trial and feasibility studies it can be stated that a MBR installation is already feasible if specific effluent requirements are set. It is expected that the market for the application of MBR installations in the Netherlands and abroad will grow fast. In the near future, with the existence of a mature membrane market, for many WWTPs the MBR concept will be a feasible alternative with additional value.

1 INTRODUCTION

Motivation
The membrane bioreactor (MBR) is a combination of an activated sludge process and membrane separation. Due to further technical developments and significant cost reductions, the interest in the MBR technology for municipal wastewater treatment has sharply increased. The technology for full scale municipal wastewater treatment was developed some ten years ago most notably in Japan and Canada. Compared to the traditional tubular membranes applied in the past, hollow-fibre and sheet membranes require much less energy, and therefore promote a progression to larger installations.

Especially in the Netherlands where urban and industrialised areas are frequently located near sensitive surface waters, the MBR technology displays several advantages compared to the traditional activated sludge processes:
- high effluent quality;
- limited space requirements;
- possibilities for a flexible and phased extension of existing waste water treatment plants (WWTP's).

Development MBR technology in the Netherlands
Several feasibility studies have been carried out by the waterboards Hoogheemraadschap van Uitwaterende Sluizen in Hollands Noorderkwartier (USHN), Zuiveringsschap Hollandse Eilanden en Waarden (ZHEW), Waterschap Rijn & IJssel (WRIJ) and Dienst Waterbeheer en Riolering (DWR). The encouraging results of these studies has led to a full scale pilot study on MBR-technology in the Netherlands. This pilot study has been executed by DHV Water. The pilot study, testing four different membrane systems with a capacity of 2 - 10 m^3/h, can be seen world-wide as one the most extensive studies of its kind. Besides the specific expertise of the membrane suppliers, the contribution of Dutch and German universities as well as engineering companies were involved, all having specific expertise in the area of MBR-technology applied in municipal wastewater treatment. The above mentioned waterboards and the Foundation of Applied Water Research (STOWA) were also involved. The STOWA initiates research activities (as the MBR development) and is financed by the Dutch waterboards, provinces and the Directorate-General for Public Works and Water Management.

As can be seen from Figure 1, the study started at the Beverwijk WWTP in the beginning of 2000 and was the starting point of a development leading to the realisation of a demonstration plant (755 m^3/h) in the year 2002 and the expected upscale to full scale MBR-plants with a capacity of 2,500 - 10,000 m^3/h in the year 2003 - 2006. The pilot research will be continued in 2002.

The aim of the pilot research was to integrate the existing know-how on conventional technology with the membrane technology and to develop the MBR technology. The technical development, application and scaling up of the MBR concept will be the main goals of the demonstration installation. In the last phase, the realisation of full-scale MBR installations, the market will be further developed. Due to larger MBR installations, the membrane surface required will increase and prices will decrease. As a result, the MBR technology can be applied without additional financial support.

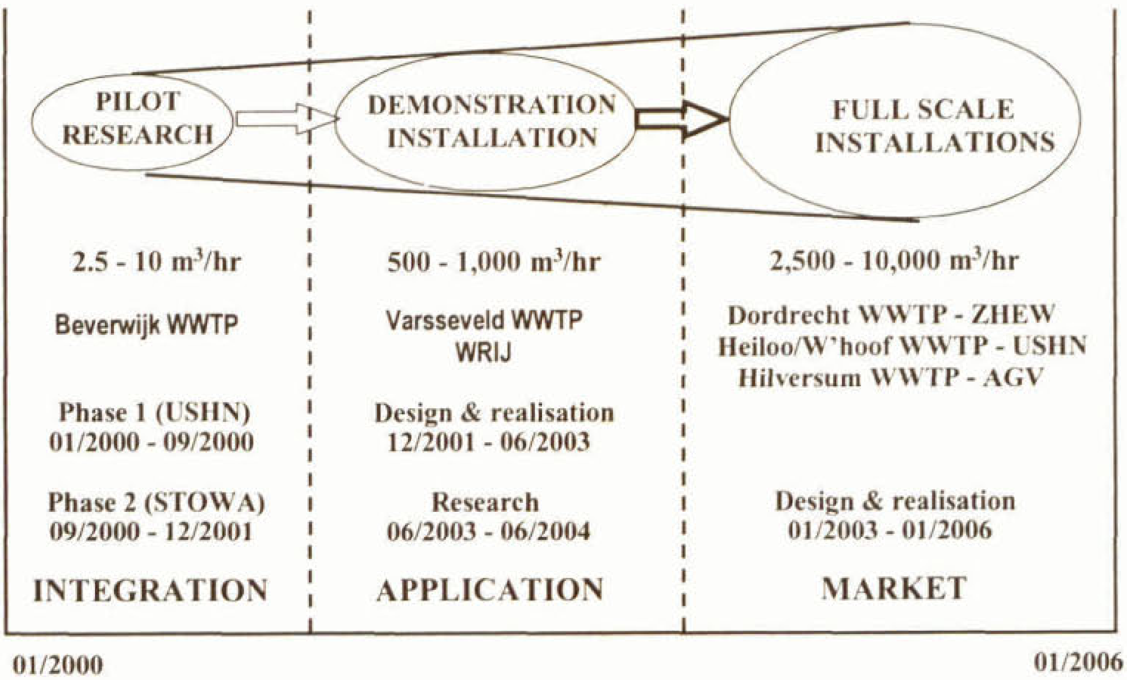


Figure 1 - Development MBR technology in the Netherlands

Objectives and prerequisites

Prerequisites for the execution of the pilot study at the Beverwijk WWTP were set as follows:

- to compare different membrane systems;
- to co-operate with manufacturers;
- to include international know-how;
- to be representative for the Netherlands;
- to execute the research at representative scale.

The objectives set for the pilot plant study at the Beverwijk WWTP were as follows:

- to do research into the technical feasibility of the MBR system for the Netherlands;
- to make a comparison between the various membrane systems available;
- to make a comparison between the MBR and a conventional system concerning: design aspects, costs, energy consumption and effluent quality;
- to further develop the MBR technology and eliminate uncertainties;
- to investigate the effects of different types of pre-treatment and consequently different wastewater compositions on the membrane performance;
- to investigate the long-term effects;
- to integrate existing know how with membrane technology;
- to investigate the effects of low temperatures on the membrane performance.

Publicity

To promote competition and consequently further technical developments and price reduction, all parties involved undersigned a secrecy agreement. By publishing the intermediate report in May 2001 this agreement lost its validity. The intermediate report described the results from the period of March 2000 till March 2001. This final report includes an integrated version of the intermediate report, a supplementary report with the appendices detailing the results of five additional research items, and a management summary.

Reader

The set up of this report is as follows. In chapter 2 the general principles and backgrounds of the MBR-technology are described and compared to the conventional activated sludge process. In order to gain a better understanding and readability of this report, often used concepts are explained. An introduction to the state of art will be introduced by a summary of a literature review. As a logic continuation the representative research items and points for attention will be made clear.

In chapter 3 the set up and execution of the research project at the Beverwijk WWTP will be explained. Besides the organisational and technical set up, also the planning and different phases will be introduced. The measuring programme and methodology will be briefly illustrated, but described in more detail in the supplementary report.

During the whole research period the Beverwijk WWTP has been used as a comparison. In chapter 4 the WWTP is described in terms of design, operation conditions and biological performance. To be able to make a simple comparison with the MBR pilot plants, in this report standard tables and graphs are used. In the chapters 6 until 9 the same approach has been applied for the MBR pilot plants of Kubota, Mitsubishi, X-Flow and Zenon. Besides the biological performance also the membrane performance is dealt with in separate sections. Before this, the membrane systems and suppliers involved in the pilot research are shortly introduced in chapter 5.

In chapter 10 a summary is given of the extra research items which were investigated in addition to the main pilot plant research:

1. pre-treatment
2. membrane fouling and cleaning
3. α-factor and energy consumption
4. effluent quality
5. sludge treatment

The reports on those side studies are added to the supplementary report.

In chapter 11 the research at the Beverwijk WWTP will be evaluated. The MBR systems are compared to each other and the results obtained are related to the goals set at the beginning of the project. Finally in chapter 12, the current status of the MBR technology will be described and recommendations will be made for further MBR development.

2 MEMBRANE BIOREACTOR

2.1 Introduction

The membrane bioreactor concept is a combination of conventional biological wastewater treatment and membrane filtration. The concept is technically similar to that of a traditional wastewater treatment plant, except for the separation of activated sludge and treated wastewater. In a MBR installation this separation is not carried out by means of sedimentation in a secondary clarification tank, but by membrane filtration. Technologically and biologically, the MBR system and the conventional system show great differences. In Figure 2 a schematic presentation of both the traditional wastewater treatment concept and three different configurations of the MBR concept are shown.

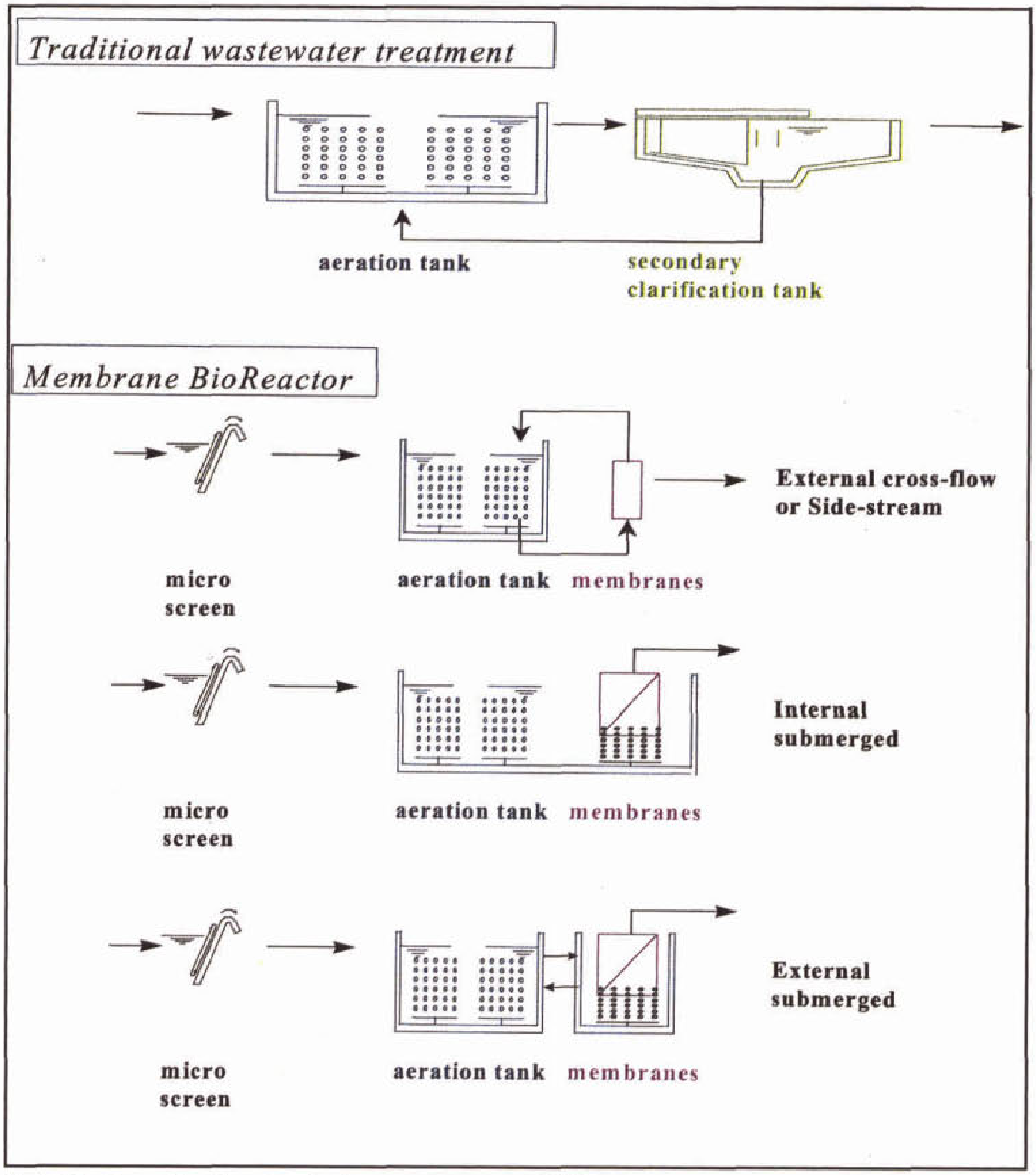


Figure 2 - Conventional wastewater treatment system and MBR

The first generation MBR's consisted of cross-flow operated membranes, that were installed outside the activated sludge tank. The cross-flow principle with its associated high flow velocity was used to prevent the build up of solids on the membrane surface, so called cake-layer formation. This method of cross-flow operation required large amounts of energy to generate the sludge velocity across the membrane surface, to maintain both the high cross-flow velocity for membrane cleaning and the required pressure drop necessary for permeation. Due to the energy requirements this concept was considered as non-viable for the applicability in municipal wastewater treatment. Furthermore the use of the cross flow re-circulation pump with its associated high pressure and excessive shear was supposed to be detrimental to the floc size and stability within the system.

An important development for membranes came when it was proposed to submerge the membrane in the aeration tank. To achieve permeation the technique utilised a reduced pressure as opposed to an external installation in pressure tubes and the necessity for high over-pressure.

This type of submerged membrane filtration in a biological system was referred to as submerged MBR (SMBR). Energy consumption was significantly reduced. The reduced pressure applied in permeate extraction was considerably lower than that required for cross-flow permeation. Furthermore an essential part of the cross-flow technique, the re-circulation pump, was absent in the SMBR configuration. Even in the SMBR configuration the cross-flow principle for maintaining a clean membrane surface was incorporated but in another form. The mechanism used to create the cross-flow stream over the membrane surface was low pressure air diffusion, which could be considered part of the activated sludge process. The air diffusion facilitates two processes: the cleaning of the membrane surface and the supply of oxygen to the biomass. The shear stress in the mixed liquor of the SMBR was much lower compared to that experienced in a cross-flow system, and as a result, sludge characteristics were much better.

2.2 Principles and backgrounds

In this chapter basic terminology and principles are explained which enables the reader to grasp the fundamentals being presented in this report.

Filtration is defined as the separation of two or more components from a fluid stream. In conventional usage, it usually refers to the separation of solid or insoluble particles from a liquid stream. Membrane filtration extends this application further to include the separation of dissolved solids in liquid streams and hence, membrane processes in water treatment are commonly used to remove various materials ranging from salts to micro-organisms. The most commonly employed membrane processes and the filtration ranges in which they operate, are presented in Figure 3.

Membrane processes can be categorised in various, related categories, three of which are: their pore size, their molecular weight cut-off, or the pressure at which they operate. As the pore size gets smaller or the molecular weight cut-off decreases, the pressure applied to the membrane for separation of water from other material generally increases. The water treatment objectives will determine the basis on which a process is selected.

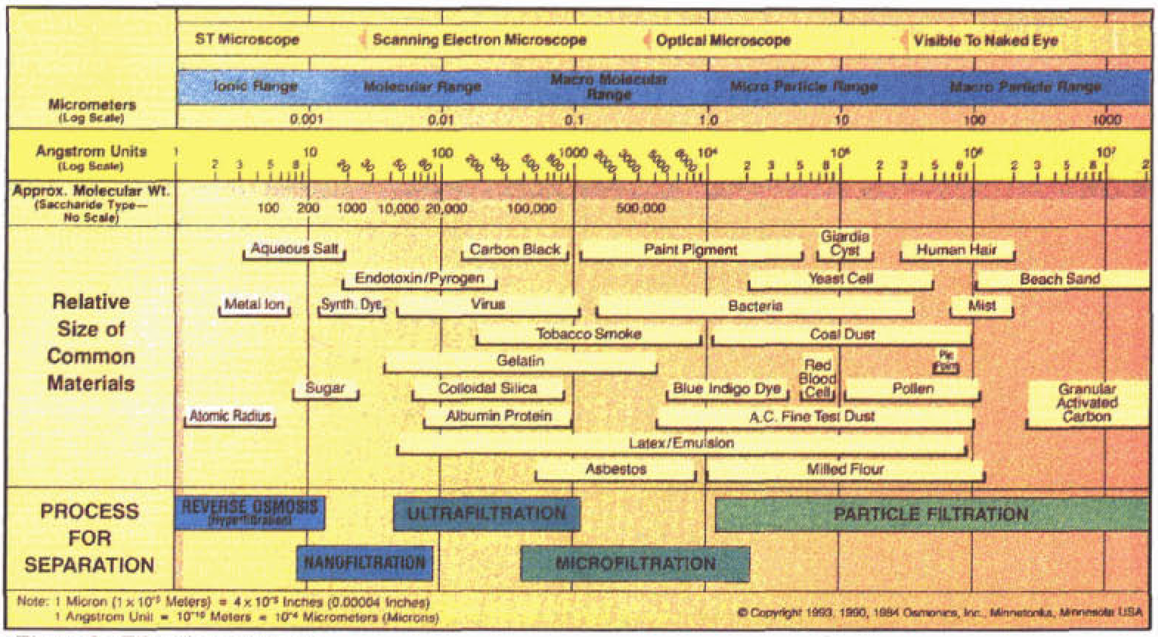


Figure 3 - Filtration spectrum

Micro-filtration (MF) and Ultra-filtration (UF) are processes that filter material on the basis of size and are generally applied in MBR concepts. In membrane separation, MF is typically used to separate or remove relatively large particles, such as emulsified oils, suspended solids and macromolecules with molecular weights greater than approximately 50,000. Pore sizes of MF membranes range from approximately 0.05 µm to 2 µm. UF and MF processes overlap to a large extent and the definition of each is vague. In general the UF membranes are able to achieve higher levels of separation, particularly regarding bacteria and viruses. UF can separate macromolecules to a molecular weight of greater than 5,000 and displays a pore size ranging from approximately 0.005 µm to 0.1 µm.

2.3 Membrane terminology

There are a number of important terms related to MBR membrane filtration processes and these can be split into two categories: membrane characteristics and filtration processes.

2.3.1 Membrane characteristics

MF and UF membranes can be made of an organic polymer, such as poly-ethylene or poly-sulphon or from ceramic material. In both cases the filtration principle is the same. The membrane can be manufactured on top of numerous support materials or be self supporting.
The MF or UF membrane is a device which allows the passage of certain components, but rejects others above a particular size or weight. This separation gives rise to two streams, the permeate or liquid stream which has passed through the membrane (clean side) and the concentrate stream which remains in the process tank (polluted side). The permeate is equivalent to the final effluent and is the product which is discharged after biological treatment.

The flow rate at which the permeate is made is dictated by the required throughput of the treatment process. This required flow must pass through the membrane barrier of a defined surface area. The flow of liquid through a specific membrane surface area is called Flux, and is expressed as:

$$Flux \quad [l/(m^2.h)] \quad = \quad \frac{permeate\ flow \quad [l/h]}{membrane\ surface\ used \quad [m^2]}$$

In Japan the unit of flux is expressed as $m^3/(m^2.d)$ or m/d, but in this report the European standard of $l/(m^2.h)$ will be used.

The membranes are in process mode most of the time. Depending on the membrane type, a relaxation and/or a back-pulse mode is required for cleaning purposes. These procedures effect the net flux of the system, which is therefore lower than the gross flux.

During relaxation mode (RLX) the membranes are recovered without extraction of permeate:

$$Fn \;\; [l/(m^2.h)] \;=\; Fg \;\; [l/(m^2.h)] \;\times\; \left\{ \frac{PR \;\; [s]}{PR \;\; [s] \;+\; RLX \;\; [s]} \right\}$$

During back-pulse mode permeate is pumped back through the membranes:

$$Fn \;\; [l/(m^2.h)] \;=\; Fg \;\; [l/(m^2.h)] \;\times\; \left\{ \frac{\left(PR\,[s] \times Q_{PR}\,[l/s] \right) \;-\; \left(BP\,[s] \times Q_{BP}\,[l/s] \right)}{\left(PR\,[s] + BP[s] \right) \;\times\; Q_{PR}\,[l/s]} \right\}$$

with:	Fn	net flux	$[l/(m^2.h)]$
	Fg	gross flux	$[l/(m^2.h)]$
	PR	process mode	[s]
	RLX	relaxation mode	[s]
	BP	back-pulse mode	[s]
	Q_{PR}	process flow	[l/s]
	Q_{BP}	back-pulse flow	[l/s]

In generating a flow through the membrane the liquid must have an associated driving force, a pressure drop. The latter gives rise to two pressure points, the static pressure at zero permeate flow and the dynamic pressure with permeate flow. Form these pressures the trans-membrane pressure (TMP) can be determined:

$$Trans-membrane\ pressure\ TMP[bar] = static\ pressure[bar] - dynamic\ pressure[bar]$$

The flux and TMP alone yield relatively little regarding the performance of the membrane, but define the operating range. If the flux is divided by the TMP the resultant is the specific flow rate through a specific surface area for a particular pressure drop. This is known as the permeability and is expressed as $l/(m^2.h.bar)$.

$$Permeability \;\; [l/(m^2.h.bar)] \;=\; \frac{flux \;\; [l/(m^2.h)]}{TMP \;\; [bar]} \qquad (at\ temperature\ T[°C])$$

This parameter is used to assess the performance of the operating membrane system and must be related to the operating temperature. During operation the membrane must process flow variations according to the dry weather flow (DWF) and rain weather flow (RWF) conditions. The permeability at a given time defines the condition of the membrane in operation. Comparisons can be drawn between the operating permeability at different times and under different conditions. If the temperature is relatively constant, the effect of peak loads can be directly seen on the membrane performance and the associated recovery. In the long term, permeability from different periods of time can be correlated via a standard temperature (15°C) and the durability and longevity of the membrane can be interpreted. Permeability is also used to establish the effect of cleaning on the membrane, be it chemical based or time/process based. Through the latter the membrane filtration process can be optimised. For a system operating at constant flow (constant flux) the permeability is used to establish the onset of required cleaning. The biological process as well as the processing conditions also reflect on the measured permeability. The permeability is a membrane characteristic and should not be confused with filterability which is a sludge characteristic.

The temperature of water also plays an important role in the assessment of the membrane performance due to the changes in the viscosity of the permeate and concentrate (biomass in the MBR). The pores of the membrane are very small and as the viscosity of water increases with decreasing temperature, the driving force or TMP needed to achieve the required flux will increase, thus reducing the permeability. To avoid the latter confusion of relating data at different temperatures, all data is corrected to a standard temperature of 15°C. In most other membrane applications this corrected temperature is set at 20°C or 25°C, but since the average yearly wastewater temperature at Beverwijk is 12°C with a slight increase to 15°C average in the bioreactor system the standard 15°C was chosen.

It must be noted that the permeability depicted in all graphical representations is the permeability at the operating process temperature at the time of the data sample, i.e. not temperature corrected. Curves are made that show the relationship between corrected and non-corrected permeability and show that the water viscosity plays a relatively minor role. The more pronounced decreases or increases in permeability are therefore caused by other factors.

A factor that has an effect on membrane performance is fouling. The fouling consists of two main types: surface fouling (macro) and pore fouling (micro). Surface fouling can be a build up of solids on the surface caused by a too high solids-flux toward the surface, thus blinding the pores and reducing the available surface area for filtration, or an inorganic scaling which forms a rigid non porous layer or scale over the surface. In both cases this fouling can lead to a substantial build-up of solid material around the membrane giving rise to the potential problem of sludging, i.e. the build up of biomass in between the membranes.
Pore fouling on the other hand occurs at microscopic level and involves the blinding of pores via soluble organic material such as surfactants, slime, extra-cellular polymeric substances (EPS) and soluble microbiological products (SMP). Scaling can also occur at microscopic pore level with the result that the scaling uses the pore as an active site for further, prolonged precipitation. In both cases the pore will be blocked, thus eventually reducing the available surface area for filtration. In extreme cases the pore scaling can cause the membrane to become brittle.

2.3.2 *Filtration Processes*

The method of extracting permeate from the bioreactor is referred to as the 'process' mode, this mode is interrupted with in-situ cleaning modes which vary depending on the membrane manufacturer and the extent of the fouling. During the process mode the membranes are often aerated with coarse bubbles to keep the solids from building up around the membrane.
Some membranes require a 'relaxation' mode to stabilise the surface solids flux before being returned to the process mode. This relaxation mode is a simple stop of the permeate flow for a short period of time, the membranes, which are basically elastic in nature, then return to their original relaxed state. During relaxation the aeration of the membranes often remains on to assist the renewal of the biomass solids in the vicinity of the membrane surface, and also has the effect of scouring the surface of the membrane thus removing any solids build up. Other membranes utilise the so called 'back pulse' mode. After a process mode period of operation the permeate produced exits the system via Clean In Place (CIP) tank. This tank stores enough permeate to allow the membrane to be flushed for a short period in the opposite direction of process filtration. The latter has the effect of flushing the membrane surface of solids build up and fouling before being returned to process mode. The modes are summarised in Figure 4. Some membranes are running at continuous process (permeation) mode, others require a regular back flush and / or relaxation mode.

All the membrane systems installed have the capacity to be cleaned with chemicals. The chemicals often used are: sodium hypochloride (NaOCl), sodium hydroxide (NaOH), citric acid, oxalic acid, hydrochloric acid (HCl), and detergents or combinations of these. The use of the chemicals depends strongly on the fouling and the type of membrane.

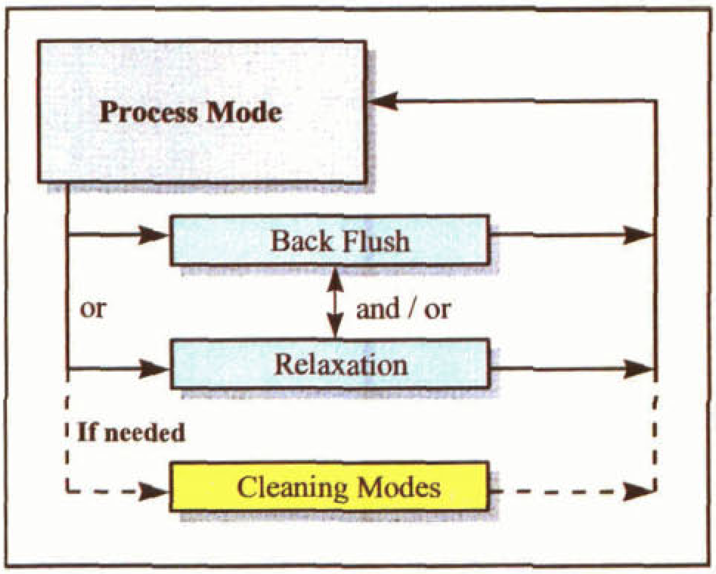


Figure 4 - Filtration and cleaning modes

The cleaning processes can be split into two distinctive categories: maintenance clean (MC) and intensive clean (IC). The MC is as suggested a preventative clean carried out with low chemical concentrations but at a higher cleaning frequency, thus prolonging the time between IC. The IC is simply a cleaning procedure established to return the membrane back to its original permeability after a long period in operation. The chemical concentrations used and the contact time are higher and longer respectively, hence intensive.

All the above mentioned modes of operation are carried out automatically with the exception of the IC, where some manual supervision is needed. Under the conditions of a full-scale treatment it is expected that this also can be fully automated. The combinations of the modes and the MC and/or IC vary greatly depending on the membrane supplier, the feed flow conditions (RWF or DWF), the season (summer or winter) and more so the performance of the bioreactor.

It should be noted once more that the membrane is not just a replacement of a secondary clarifier, but an integral part of the MBR unit operation.

2.4 Literature

2.4.1 History

Despite the high tech image of the MBR today the roots for this process were conceived from a humble beginning in the late 1970's as a simple concept of filtering biomass, anaerobic or aerobic, utilising available filtration techniques at that time. The filters of that time however, proved unreliable due to fouling and breakage. In the 1980's the development of membranes progressed on three fronts, North America/Canada, Europe and Japan. After considerable R&D the techniques used in reverse osmosis (spiral) were extrapolated into simple tube membranes, requiring a high cross-flow velocity over the membrane surface to keep it clean of solids. Many of the manufactured membrane types became specific for various food processing application before the step to biomass filtration was taken and were made from every conceivable material imaginable. The 1980's also saw numerous problems within the membrane world, as membrane reliability and module breakage yielded an overall bad impression for the membrane industry.

The late 1980's early 1990's saw the evolution of more reliable membranes, the tube membranes were optimised into a robust mature product with very little in the way of mechanical problems, and the prices reduced, opening the doors to new markets other than high value food, or drug processing. The filtration of biomass and the true concept of MBR was born.

Figure 5 - Tube membrane installation (X-Flow)

Numerous small-scale industrial and municipal systems were built but the market remained limited to low volume, high concentration wastewater streams. Operating costs were high, with particular emphasis on membrane replacement and energy consumption. This too, also limited the competitiveness of the MBR and was often found to be compared on investment cost and exploitation cost to conventional treatment systems. The expected boom in MBR references at this time did not come to fruition, despite the cries of better effluent quality, reduced space, 'zero' sludge and water reuse [ref. 1,2,3,4].

The realisation of the so-called submerged membrane from Japan and Canada in the mid-1990's saw a new opening in the MBR market. The need for high cross-flow over the membrane to maintain a clean operating surface was eliminated, but the higher associated cross-flow fluxes were sacrificed. However, more membrane surface could be built into a far smaller module footprint. The modules were cheaper to build and required far less energy in their operation. The tube based MBR systems found it impossible to compete with the new low pressure submerged membranes and were pushed back to the high value end of the market to food, drug, juice and very difficult wastewater streams.

By the end of the 1990's the cost of the submerged membrane had reduced significantly and therefore became a competitor in the high flow, low concentration market - large-scale municipal wastewater. A number of feasibility studies were carried out for full-scale municipal treatment systems and SMBR was found to be competitive, particular in sensitive water areas.

2.4.2 *Current Status*

Presently, over 1,000 MBR's are in operation world-wide, with many more proposed or currently under construction. MBR's have proliferated in Japan, which has approximately 66% of the world's total installations. The remainder can be found mainly in North America or Europe. Over 98% of the systems couple the membrane separation process with an aerobic biological process rather than to an anaerobic bioreactor. Approximately 55% of the systems have the membranes submerged in the bioreactor while the remainder have the membranes external to the biological process.

The first applications of MBR technology in Japan (in 1990) concerned small scale installations for domestic wastewater treatment and re-use and some industrial applications, mainly in the food and beverage industries where highly concentrated flows were common. The domestic applications often consist of so-called Johkaso or septic tank treatment and in-building (office or domestic) wastewater collection systems. The applications in Canada and Europe mainly focused on industrial applications and treatment of landfill leachate (percolation water).

Figure 6 - Johkaso tank treatment in Japan (Kubota)

There are a limited number of MBR's in operation in municipal wastewater treatment. These plants normally operate under different process circumstances than those common to Dutch municipal wastewater treatment. The differences can be expressed as: constant feed flow, elevated process temperatures, low sludge productions due to extremely low biological loading and debris-free influent. Furthermore, it must be noted that the plants currently in operation are relatively small in hydraulic capacity. Since the last few years a number of relatively large municipal MBR plants are in operation as shown in Figure 7 to Figure 10. The Swanage plant is in operation since the end of 2000 [ref. 5].

Figure 7 - Rödingen (Zenon, 135 m³/h, 1999)

Figure 8 - Markränstadt (Zenon, 180 m³/h, 2000)

Figure 9 - Porlock (Kubota, 80 m³/h, 1998)

Figure 10 - Swanage, (Kubota, 720 m³/h, 2000)

The majority of MBR plants in operation are solely designed for the removal of organic matter. In some cases nitrification and/or denitrification is required. The same can also be said for phosphorus removal. If required, it is solely achieved via precipitation with metal-salts. These design criteria can have large implications in the design and operation of a MBR installation.

2.4.3 *Research & Development*

In the last twenty years, the MBR process has been studied in numerous ways. The main goal of these studies was focused on exploring the feasibility of the MBR process and methods for process enhancement. Emphasis was on effluent quality achieved, higher sludge loading, system compactness, water re-use, and the potential of 'zero' sludge growth. However, most R&D was application based, and several shortfalls occurred in the MBR applicability to municipal wastewater streams.

The MBR process has been found to depend on several inter-related parameters, and to understand the process and pin-point recommendations for research, it was necessary to divide these processes into several fundamental categories, which are described in the following sections.

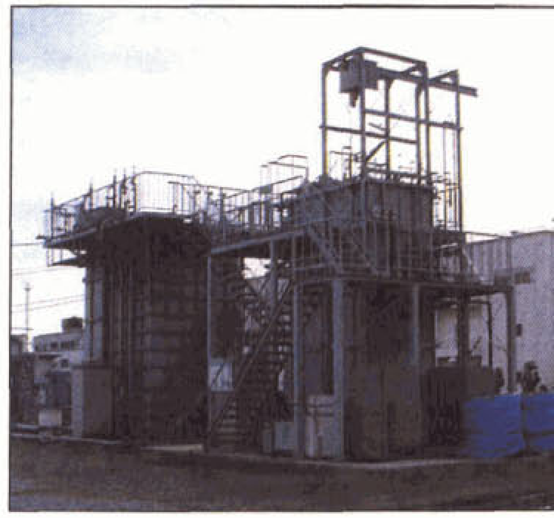


Figure 11 - MBR pilot in Misato Japan (Mitsubishi)

Pre-treatment

Pre-treatment is essential to reduce fouling and sludging on the membranes. Simple pre-screening is carried out on almost every MBR application, this more often being in the form of 1mm screens. However, recent experience has suggested that, depending on the membrane type, a one step pre-screening unit operation is inadequate, and can result in full scale sludging.

Screens that are applied to membrane systems are generally rotating drums, step, screw, static (with primary clarifier) and fine screens. There is no defined standard method and results and experiences are based on shortcomings in this critical area. The lack of performance data on any screen in this municipal application is seriously lacking and as a result maintains a high priority [ref. 6,7].

In conventional WWTP's it is common practice to dose precipitation chemicals to the primary clarifier to reduce phosphate. The main goal of pre-precipitation in MBR systems is not only phosphate removal, but also flocculation of colloidal material to reduce the potential fouling load on the membranes. According to references there has been little research in this technique. Simultaneous precipitation with Aluminium or Ferric Chloride is the most common way of phosphate reduction but has not been linked to membrane performance.
Conclusively it was recommendable to explore the potential advantages of pre-precipitation, one, to enable a comparison with simultaneous precipitation, and secondly the effect on the membrane performance.

Biological treatment

The specific biological treatment in a MBR system is a very complex and inter-related process. The difference between sludge composition of a conventional system and a MBR is significant. Much research has been carried out on the biological performance of the MBR system as documented in all references listed, however, a detailed understanding of the processes involved is lacking. The effect of sludge concentration on the alpha factor has been addressed [ref. 9] but not fully researched. It is theorised that External Polymeric Substances (EPS) and Soluble Microbial Products (SMP) may have an extended influence on the alpha factor value of MBR sludge and therefore recommended to investigate this influence further [ref. 8,9].

Other areas where much interest has been generated is the floc structure, its characteristics and potential effects on the performance of the membrane system, as well as related aspects to the alpha factor. A full understanding of the detailed biological make up is needed to push the MBR process to the next level. The influence of predators in MBR sludge has been seldom analysed, especially regarding the relation to membrane performance, this must also be addressed [ref. 10, 12, 13, 14, 15, 16].

The macro biological processes have been well documented [ref. 17,18,19] especially regarding nitrification and denitrification. However, the research carried out until now has been limited to industrial systems or municipal systems running at a relatively constant load. The Dutch situation has not been addressed and must be further investigated to realise the potential of the biological macro processes. Total nitrogen concentrations in the effluent of <10 mg/l are common, in some areas <5 mg/l and in very sensitive areas the so called MTR quality is required with a total N of <2.2 mg/l.

Membrane separation
All the references listed suggest various methods of membrane operation. However, the specific operating windows of each are not clearly defined. Each membrane system in combination with the biological process will have a different optimised operating window. This operating window relies strongly on the membrane suppliers, but leaves room for significant optimisation. The suppliers suggest an optimum operating window for the modules themselves as well as for the cleaning. Little is related to the optimised biological operating window to promote longer membrane life, longer periods without cleaning and optimised energy usage. Rather than just a membrane and just a biology considered as two separate entities, the MBR must be considered as one unit operation with numerous interactions between biology and membrane and visa versa.

Effluent quality
There is no question or doubt regarding the macro quality achieved from the permeate produced by MBR installations. Question do arise however, regarding micro pollutants, heavy metals, EOX, PAH, bacteria, colour and viruses. Only the surface has been touched regarding these items, but as the discharge criteria become more stringent, these micro pollutants will become a significant issue, and therefore must be addressed here.
Re-use of effluent may also require different regulations. For example effluent can be re-used as irrigation for farming, horticulture, public parks or golf courses. All these applications differ in the quality requirements of the effluent used for the specific irrigation purposes [ref. 25, 26, 27]. Irrigation water re-use is a small step from process water for industry and a slightly larger step to drinking water.

Sludge handling
Extremely little work has been carried out on this topic. The MBR concept is often related to extremely low loaded activated sludge systems with a very low sludge production. The latter is related to the size of the current installations. In Germany some larger normally loaded municipal MBR systems are in operation and the sludge production is comparable with conventional treatment plant. Also the thickening and dewatering results are comparable.
The various conventional handling steps for primary and secondary sludge treatment will be addressed regarding the treatment of MBR sludge. Digestion must also be considered.

Research items and points for attention
From the literature study summarised above and discussions with manufacturers, universities and consultants, the research items and points for attention for the pilot research at Beverwijk WWTP were formulated. Table 1 shows the research items following the comparison between the current status of experience and the possible future situation.

Table 1 - Research items and points for attention

Research item	Current situation	Future situation	Side study
Experience	Industrial wastewater	Municipal wastewater	-
Size	Small scale	Large scale	-
Process conditions	High process temperatures Constant feed flow Extremely low loaded	Low process temperatures Fluctuating feed flow (RWF) Low loaded	-
Pre-treatment	Little debris in feed	Debris (sand, hair, leaves, etc.)	1
Fouling and cleaning	Manually	Automatic, long life-time	2
Energy consumption	Not of interest	Optimisation required	3
Biological performance	Removal of organic matter	Nutrient removal (N/P) and MTR quality	4
Sludge treatment	Low sludge production	Normal sludge production	5

3 PROGRAMME APPROACH

3.1 Introduction

In this chapter the programme set up and execution is described. In section 3.2 all companies and organisations involved are mentioned, including a short description of their contribution. In section 3.3 the technical set up is presented and the execution of five side studies is further introduced. In section 3.4 the different research phases are described, while in section 3.5 the measuring programme and methods are shortly presented. The continuation of the project after May 2001 is described in section 3.6. This chapter gives a global overview, for more technical details chapters 4 and 6 to 9 are referred to.

3.2 Organisational set up

Based on the prerequisites and objectives mentioned in the introduction, many companies and organisations were involved in the pilot study. From the point of view of client and participants two periods can be distinguished as can be seen from Table 2.

Table 2 - Participants	**PERIOD**		
	March 2000 - September 2000	**September 2000 - December 2001**	
Client	Waterboard USHN	STOWA	
Execution	DHV Water BV	DHV Water BV	
Manufacturers / Representatives	Kubota - Solis Mitsubishi - Grontmij X-Flow - NUON Zenon	Kubota - Solis Mitsubishi - Grontmij X-Flow - NUON Zenon	
Project group	DHV Water BV Waterboard USHN University Aachen Berthold Günder S.A.G.	*Additional study 1:* DHV Water BV Waterboard USHN *Additional study 2:* DHV Water BV All manufacturers & representatives University Aachen University Delft TNO	*Additional study 3:* DHV Water BV BRCC (Inter)national experts *Additional study 4:* DHV Water BV *Additional study 5:* DHV Water BV Alfa Laval
Expert group / Steering committee	Waterboard WRIJ Waterboard ZHEW Water company PWN STOWA	Waterboard WRIJ Waterboard ZHEW Waterboard USHN Water company PWN STOWA RIZA GTD DWR Waterboard WSRL Waterboard WRD	

For contact persons and addresses see Appendix 1

The pre-selection for membrane suppliers was based on the following criteria:
- world-wide experience with (full scale) application of MBR technology;
- expected technical suitability for application of the membranes in municipal wastewater treatment systems under Dutch circumstances;
- future membrane production capacity and pricing;
- liability of the companies involved.

With the selected companies the following four different membrane systems were placed under investigation at the Beverwijk WWTP:

- Plate membrane system from Kubota, represented by Solis;
- Fibre membrane system from Mitsubishi, supported technologically by Grontmij;
- Tubular membrane system from X-Flow, in co-operation with NUON;
- Zeeweed membrane system from Zenon.

Between March and September 2000, the Universities of Aachen and Stuttgart (Günder) and consulting company S.A.G. were involved, all having specific expertise in the area of MBR technology in municipal wastewater treatment in Germany. They contributed their expertise through participation in a Project Group. Between September 2000 and June 2001 five side studies were defined and other scientific organisations from the Netherlands participated.

In the first period the pilot study was overseen by an Expert Group acting independently of the Project Group and other study staff. In early September, the Expert Group presented its findings and positive recommendations to the waterboard USHN and the STOWA, regarding the specific application of MBR technology at Beverwijk WWTP and its general application in The Netherlands. From the beginning of the second period, the Expert Group was converted into the STOWA steering committee and was extended with more participants as shown in Table 2.

3.3 Technical set up

The four pilot plants were designed as full self-supporting and independent WWTPs. All features necessary for automatic operation were incorporated, including data collection and -processing. The size of the pilot plants was according to the prerequisite of a representative scale and amounted to 2 - 10 m^3/h. In chapter 6 to 9 the technical design of the pilot plants is described in more detail.

During research phase 1 and 2 (see section 3.4) the influent for the pilot plants was taken from one of the six existing primary clarifiers (one reserve) of the Beverwijk Zaanstreek WWTP. The pre-settled wastewater from this clarifier was separately pumped from the overflow weirs to the four MBR pilot plants. In a latter phase the location of the MBR-influent was changed to the splitter-box just before the primary clarifiers, to feed the MBR pilot plants with coarse screened raw wastewater as is schematically presented in Figure 12.

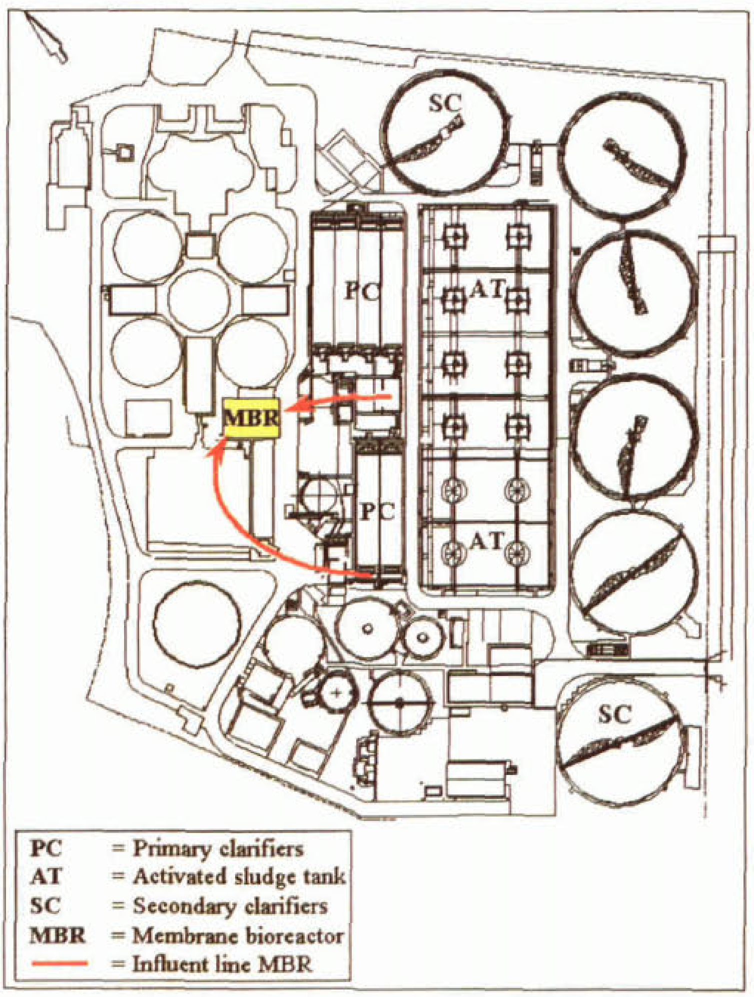


Figure 12 - Top view of Beverwijk WWTP and location pilot plants

The pilot results, especially regarding the biological performance and (sludge-) characteristics, were compared to those of the existing Beverwijk WWTP. To obtain an optimised comparison, the normal applied phosphorous-removal method i.e. simultaneous precipitation was changed to pre-precipitation for two of the six waterlines during the first research phase. This is further explained in chapter 4.

Since the applicability of MBR technology can be sensitive to hair, debris and sand, and practical experience regarding the necessary pre-treatment steps was not available at that time, it was chosen to incorporate different principles and techniques. The results and evaluation will be presented in the supplementary report and are summarised in paragraph 10.1 (side study 1 - pre-treatment).

The upscale of the MBR-technology consequently effects the procedures for membrane cleaning. The current procedures applied to existing MBR installations are not longer applicable for capacities over a few hundred m^3/h. Also the effect of chemical usage on membrane lifetime, costs and effluent quality can not be neglected. To get a complete overview of all important influencing factors, a separate study was undertaken. The results and evaluation will be presented in the supplementary report and are summarised in paragraph 10.2 (side study 2 - membrane fouling and cleaning).

One of the main investigated topics was the energy consumption of a MBR installation. Besides the energy used for coarse air in keeping the membrane modules clean, the reduction of the α-factor increases the energy consumption of the biological processes. The α-factors for the different MBR-installations and the reference waterlines were measured regularly and supported by a separate study to investigate α-factor and relationships with biological and physical parameters. The results and evaluation will be presented in the supplementary report and are summarised in paragraph 10.3 (side study 3 - energy consumption / α-factor).

An advantage of applying MBR technology is the effluent quality, which is superior compared to the quality achievable with traditional technologies. In a one-step treatment not only suspended solids are 100 % retained, but membrane separation also influences bacteriological and virological quality, as well as concentrations of heavy metals and organic micro pollutants. Furthermore it is possible to operate MBR systems at higher sludge concentrations, allowing high nitrogen and phosphorous removal efficiencies. The results and evaluation will be presented in the supplementary report and are summarised in paragraph 10.4 (side study 4 - effluent quality).

Finally MBR sludge characteristics have been investigated, not only at laboratory scale but also at semi full scale. Due to the limited sludge production of the pilot plants, tests at semi full scale could not be performed continuously. The results and evaluation will be presented in the supplementary report and are summarised in paragraph 10.5 (side study 5 - sludge treatment).

3.4 Research phases

The research programme was split into four phases in which the nature of the pre-treatment, and as a consequence the composition of the influent, and the phosphorus removal principle was changed. In this way the research results could be seen as representative for the Dutch situation. The four research phases including approximate planning are listed in Figure 13.

For the Kubota pilot the fourth phase started two months earlier, on March 12, 2001. The pilot plant was rebuild and an anaerobic compartment was incorporated, to improve the biological phosphorus removal process.

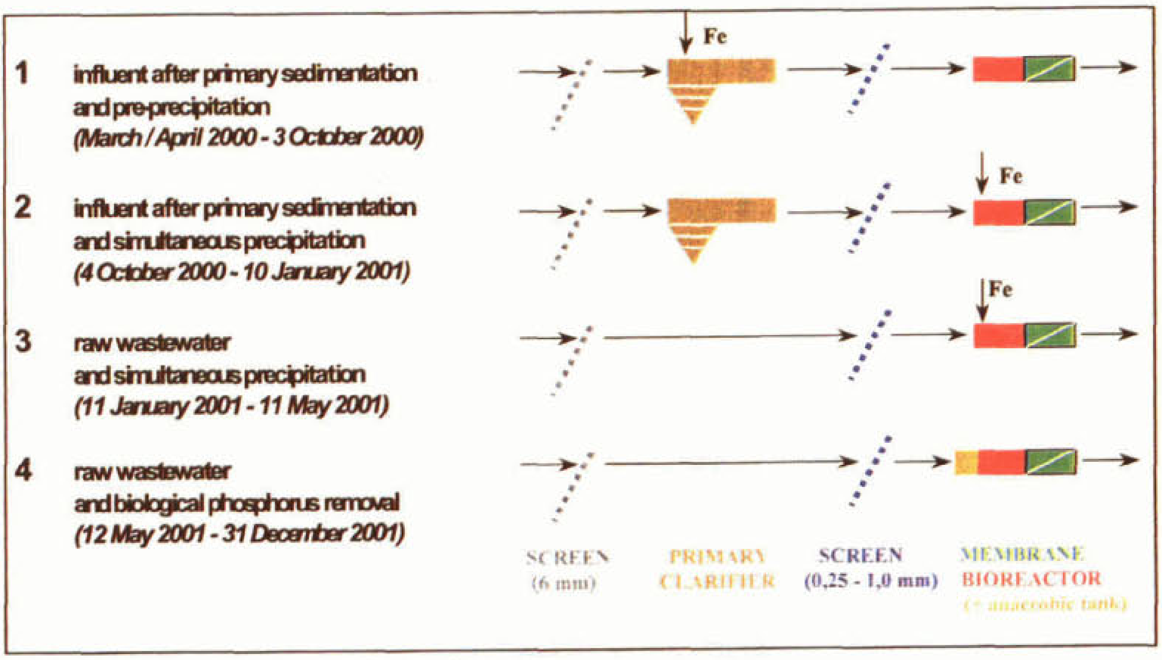


Figure 13 - Phases pilot research MBR development

For the Mitsubishi, X-Flow and Zenon system the ferric dosing was stopped around 11 May 2001. Since then the Mitsubishi and Zenon system are operated without any means of phosphorus removal, except the normal biological phosphorus uptake. In the new X-Flow plant (see chapter 8) an anaerobic compartment was included.
The chemical dosing for the pre-precipitation process took place on 2 of the 6 full-scale primary sedimentation tanks. The chemical dosing for simultaneous precipitation was introduced separately on all pilot plants.

Regarding the research programme during phase 1 the following parts can be distinguished:

- to set up and test the systems and to execute clean water tests;
- to seed the bioreactors allowing a quick start up;
- to establish a constant flow condition and eventually proportional flow related to the Beverwijk WWTP influent flow variations;
- to simulate RWF conditions and low process temperatures;
- to perform membrane cleaning procedures.

Based on the experiences in phase 1, during phases 2, 3 and 4 the pilot plants were more than once adapted for optimised operation and design. These adaptations are described in chapter 6 to 9. For more detailed information, the research programmes for phases 1, 2 and 4 are added in the supplementary report.

3.5 Measuring programme and methods

To control the overall performance of the MBR pilot plants and the reference water lines of the Beverwijk WWTP an extended analytical and technical research program was incorporated.

To determine the biological performance i.e. COD, nitrogen and phosphorous removal of each pilot plant, influent and permeate samples were taken on a daily basis. 24 Hour composite samples were analysed with a photometric analyser on site and controlled with NEN-analyses carried out by the laboratory of waterboard USHN.

The frequency of the analyses was based on the results achieved and reduced at a later stage of the study. It should be noted that the feed samples were always taken behind the micro-screens. The actual feed concentrations for COD and SS will be somewhat lower, depending on the screen type (paragraph 10.1).

With online analysers the NH_4-N, NO_3-N and PO_4-P concentrations in the permeate flow of all pilots were measured until the end of February 2001. The results of these analyses were used for daily operation. Several times a week the sludge concentration and inorganic part of the reactors was controlled to adapt the surplus sludge pumps and check the changes in sludge characteristics.

On a regular basis additional analyses and tests were performed, intended to relate the different parameters to the overall performance of the MBR pilot plant.

The CST (Capillary Suction Time) was measured to yield information on the filterability of the sludge. Besides this "dead-end" filtration measurement also a dynamic flow test was carried out to measure the dynamic viscosity, the so called Y-flow test. Also a standardised viscosity measurement using a laboratory viscosity meter was executed.
The α-factor was measured using both a surface aerator and a bubble aerator pilot system. For this measurement also the oxygen uptake rate of the sludge was measured. Both α-factor and oxygen uptake rate were measured to yield information on the sludge characteristic and the energy requirement for the system.

Microscopic sludge analyses were executed to gain information on the development of the sludge characteristics. Photographs were taken. The standard investigation for activated sludge, i.e. Eikelboom, is not fully applicable for the MBR sludge. Obviously the presence of filamentous organisms, protozoa and other organisms is monitored, but the floc structure is difficult to describe. In general the flocs present in a MBR system are much smaller compared to a conventional system, looking more or less like pin flocs.

Each pilot plant was designed with a number of critical analogue signals which were monitored on a continuous basis (10-15 s), like feed- and permeate flow, bioreactor oxygen concentration, pH-value and temperature. Also the permeate-suction and static pressure were monitored to be able to calculate the TMP and in combination with the flux, the permeability of the systems.

3.6 Continuation

This report describes the pilot results until December 2001. However, Mitsubishi, Kubota and Zenon have decided to continue the research in 2002. The main goals of this continuation are:

- further process and cleaning optimisation;
- investigation of long-term effects;
- investigation of cold temperature effects.

In 2002 also two new MBR pilot plants, belonging to two other membrane suppliers, will take part in the research programme in Beverwijk.

In May and November 2001 two 3 week MBR courses have taken place, on which in total 11 operators and 6 process engineers of the Waterboards involved participated. The course contained a one week theoretical part, a two week practical part, and was ended with an exame. Since these courses most of the trained operators are working in shifts on the MBR pilots in Beverwijk, and are responsible for the daily operation. The operation was still under DHV supervision, in co-operation with the suppliers.

4 BEVERWIJK-ZAANSTREEK WWTP

4.1 Introduction

In this chapter the performance of the Beverwijk-Zaanstreek WWTP is described. In section 4.2 the system configuration and design data are given. In section 4.3 the biological performance is presented and summarised.

The Beverwijk WWTP's are shown in Figure 14. In the photograph the waterline of the Beverwijk-Zaanstreek WWTP and the Beverwijk en Omstreken WWTP, the sludge treatment and the location of the MBR pilot plants are shown.

Figure 14 - Beverwijk-Zaanstreek WWTP

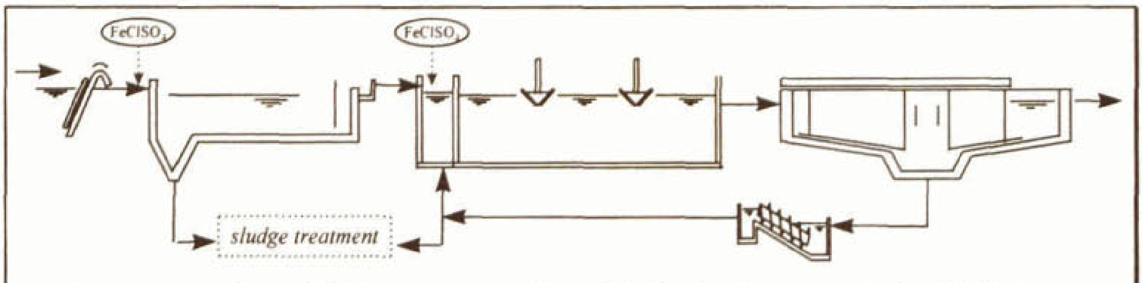
4.2 Configuration Beverwijk WWTP

The Beverwijk-Zaanstreek WWTP is a medium loaded activated sludge system with primary clarification. The design is based on a 20 mg N_{total}/l effluent requirement and the design sludge loading amounts 0.100 kg COD/(kg MLSS.d). The WWTP is overloaded; the current loading is 317,000 p.e. (à 54 g BOD), the design loading 226,000 p.e. The hydraulic design loading amounts 8,050 m^3/h.

The wastewater of the Beverwijk-Zaanstreek WWTP is transported by pressure mains from Castricum and Assendelft. Besides this, a part of the wastewater from the Beverwijk en Omstreken WWTP and the condensate of the sludge drying installation is also treated.

The configuration of the Beverwijk-Zaanstreek WWTP waterline and the main dimensions are schematically presented in Figure 15.

The aeration control is based on a combination of oxygen control and time switches. For precipitation purposes PIX-110 ($FeCl_3SO_4$, 12.3 w.% Fe, 1.500 kg/m^3) is used. Phosphorus precipitation can take place in the primary clarifier (pre-precipitation) or in the aeration tank (simultaneous precipitation).



screen			primary clarifier			aeration tank (incl. selector)			secondary clarifier		
number	2		number	6		number	6		number	6	
width	6	mm	loading	3.4	$m^3/(m^2.h)$	volume each	5.250	m^3	loading	0.84	$m^3/(m^2.h)$
			length	50	m	nr. of aerators	12		diameter	45	m
			width	8	m	capacity each	170	kg O_2/h	side depth	2.0	m
			depth	2.5	m						

Figure 15 - Beverwijk-Zaanstreek WWTP configuration of the waterline

4.3 Biological performance

4.3.1 Process conditions

The influent flow during the research period is presented in Figure 16.

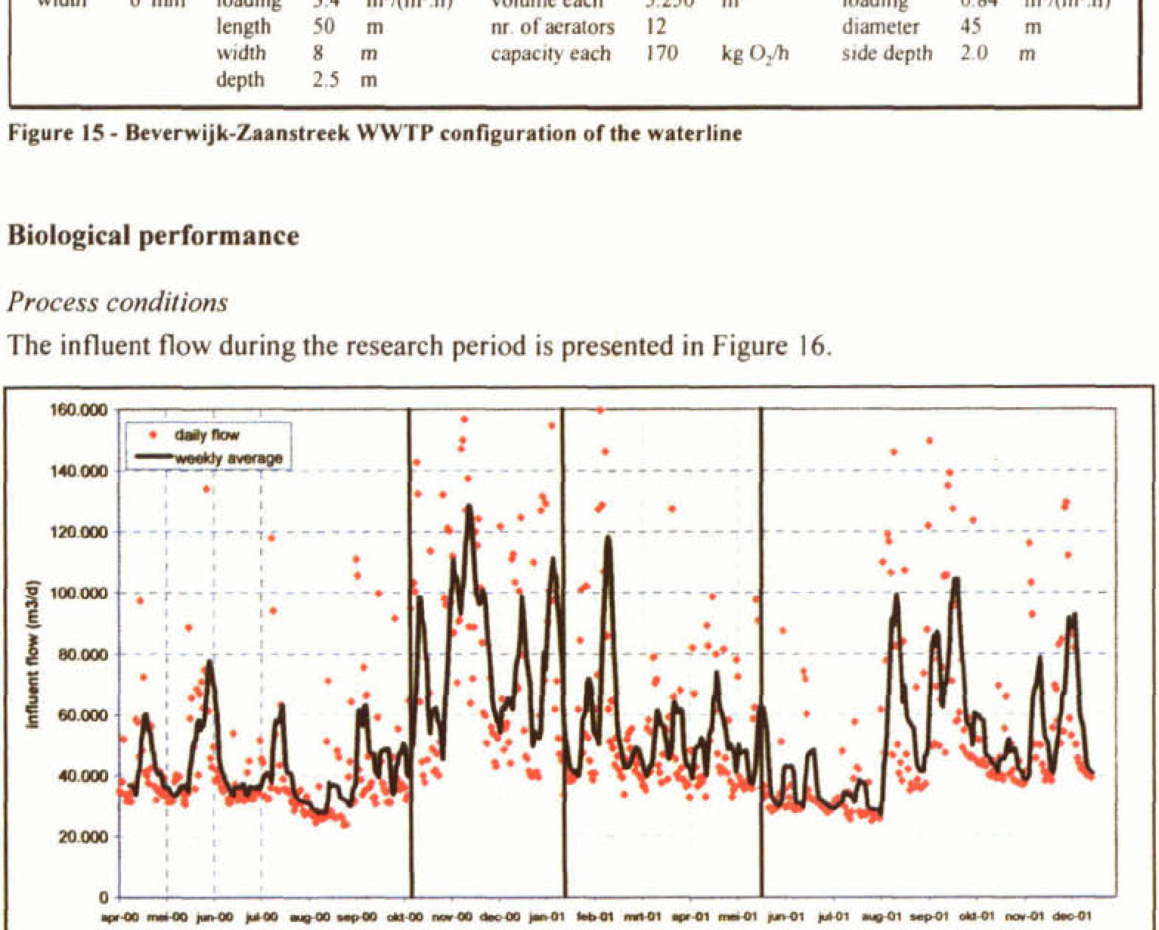


Figure 16 - Rain fall and daily flow on the Beverwijk WWTP

The design flow at DWF conditions of the Beverwijk-Zaanstreek WWTP is approximately 40,000 m^3/d. The average influent flow during the research period varied strongly due to rain weather conditions. The average flow during the whole research period (incl. rain) amounted to 54,000 m^3/d.

Two weekly flow patterns on the Beverwijk WWTP during DWF and RWF situations are presented in Figure 17.

The main average process conditions during the pilot research are presented in Table 3.

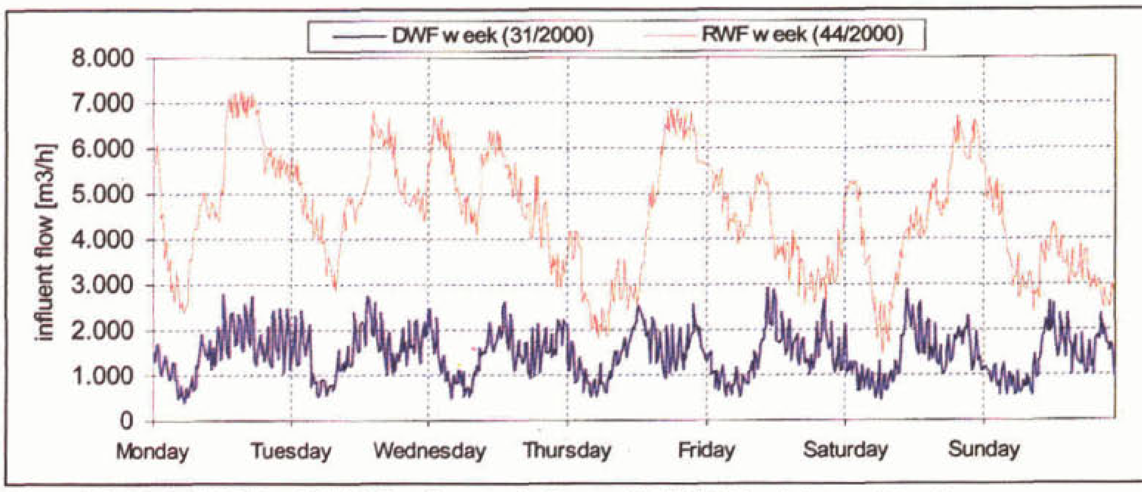


Figure 17 - Typical DWF and RWF situations on the Beverwijk WWTP (excl. internal flows)

Table 3 - Process conditions of the Beverwijk-Zaanstreek WWTP

parameter	unit		phase 1	phase 2	phase 3	phase 4
influent flow	m^3/d		42,300	80,800	52,900	50,700
process temperature	°C	average	21	16	16	21
		range	14 - 24	8 - 22	7 - 23	11 - 27
pH	-		7.3	7.3	7.3	7.3
biological loading	kg COD/(kg MLSS.d)		0.12	0.15	0.15	0.12
sludge concentration	kg MLSS/m^3		3.7	4.5	4.6	4.1
organic part	%		65	63	65	65
sludge production	kg MLSS/d		6,150	4,620	5,010	6.780
sludge age	d		19	31	29	19
ferric dosing AT	l/d		3,530	5,940	3.930	2.680
ferric dosing ratio	mol Fe/mol P		0.94	0.99	1.10	0.76

4.3.2 *Results*

The main biological parameters are presented graphically in Figure 18 to Figure 20. In Table 4 the average feed and effluent concentrations are presented.

Table 4 - settled influent and effluent concentrations

parameter		unit	phase 1	phase 2	phase 3	phase 4
COD	influent settled	mg/l	386	317	416	402
	effluent	mg/l	46	52	48	50
	efficiency	%	88	84	88	88
N_{kj}	influent settled	mg/l	62	41	58	63
	effluent	mg/l	6.5	9.2	7.1	9.1
NO_3-N	effluent	mg/l	8.6	5.6	7.2	5.7
N_{total}	effluent	mg/l	15.2	14.8	14.0	14.8
	efficiency	%	75	64	76	7.7
P_{total}	influent settled	mg/l	9.1	7.6	6.9	7.1
	effluent	mg/l	1.3	1.8	1.6	1.4
	efficiency	%	85	76	77	80

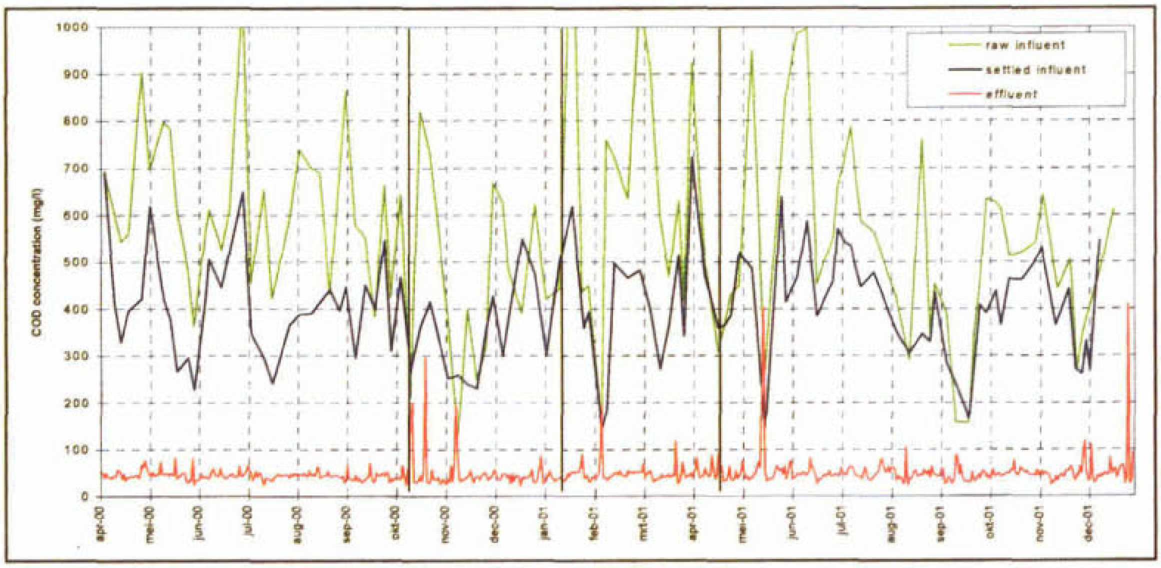


Figure 18 - COD removal on the Beverwijk WWTP

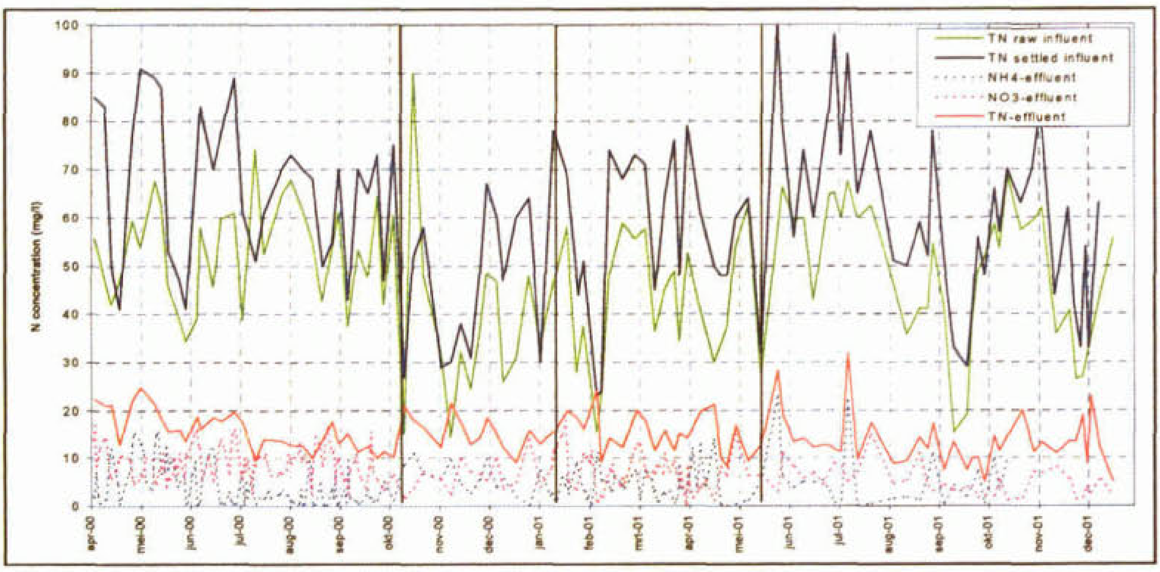


Figure 19 - Nitrogen removal on the Beverwijk WWTP

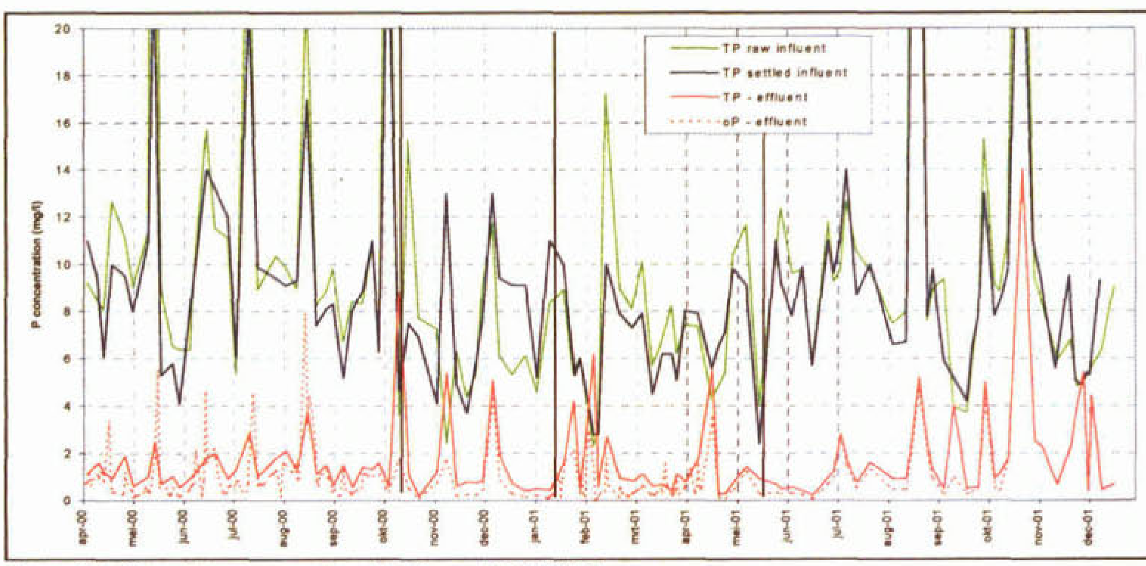


Figure 20 - Phosphorus removal on the Beverwijk WWTP

Nitrogen removal

As the figures show, the raw influent TN is lower than the settled influent TN. This is due to the fact that the nitrogen content of the reject water and the condensate of the sludge drying installation is monitored in the settled influent but not in the raw influent.

The nitrogen effluent concentration is relatively stable over the whole research period at a level around 15 mg N_{total}/l.

Phosphorus removal

On the Beverwijk WWTP temporarily high phosphorus loadings occur due to industrial discharges. As a reaction to this, the chemical dosing on the Beverwijk WWTP is manually adjusted. The response time is long and as a consequence high effluent concentrations are measured.

4.3.3 *Sludge characteristics*

The main sludge characteristics of the Beverwijk WWTP are presented in Table 5. In phase 4 no additional measurements on the sludge characteristics have been performed.

Table 5 - Sludge characteristics (sludge from AT3)

	unit	phase 1	phase 2	phase 3	phase 4
Sludge parameters					
DSVI	ml/g	240	210	150	160
Viscosity					
viscosity value	mPa.s	3.0	6.8	3.3	n.a.
shear rate	1/s	60	50	60	n.a.
α-factor					
surface aeration	-	n.a.	n.a.	0.72	n.a.
bubble aeration	-	0.72	0.80	0,83	n.a.

The flocs are compact with a medium floc size. The floc structure is both compact and lose. The main structure consists of compact flocs, with decreasing compactness to the outside of the floc. In this area of the floc, occasionally floc-bound filamentous organisms are present. During the first two phases some red-brownish coloured scale amoebas were observed in each microscopic sample. Since January no scale amoebas were present. Mono cultures are scarcely observed. Occasionally higher organisms like rotifers and free swimming ciliates were present. Normally stalked ciliates were present in the samples. During wintertime conditions their number decreased.

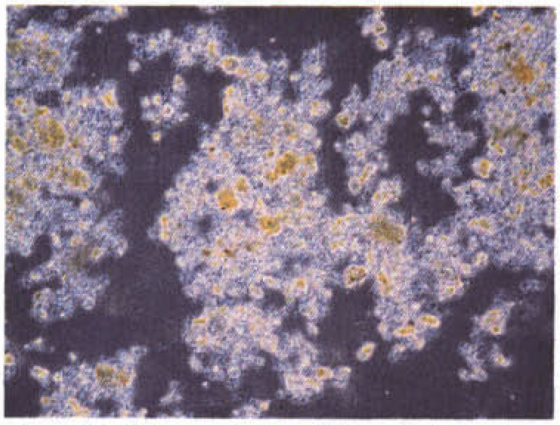


Figure 21 - Microscopic view of the activated sludge

5 MEMBRANE SUPPLIERS

5.1 Introduction

There are a multitude of membrane and membrane system suppliers in the world, from which many have developed their own MBR systems. A study was made of the most developed or matured products and the related experience in industrial and municipal wastewater treatment. Based on the prerequisites of maturity and experience four suppliers of membrane products were chosen to represent the MBR study at Beverwijk STP. Zenon, a Canadian based company wielded 20 years of product development and membrane applicability, with some of the largest reference plants in municipal wastewater treatment and numerous industrial MBRs. Kubota, a Japanese based company, wielded two decades of experience in MBR realisation on both the municipal and industrial side, they currently have the largest working MBR at 13.000 m^3/day. Mitsubishi, also a Japanese based corporation expressed vast diversity in both small municipal and industrial installations and currently have the most reference plants. The fourth supplier X-Flow, a Netherlands based supplier, displayed limited municipal experience but a wide variety of industrial systems of reasonable capacity. Four pilots were placed at Beverwijk STP and were all running by March 2000. Three from the four were of the submerged type, the X-Flow membranes were of the external (or side-stream) type.

5.2 Kubota

The Kubota filtration surface is a flat-plate 0.4 μm pore size (MF), strong polymeric membrane cast on the outside surface of a porous flat support medium. This flat surface is further mounted on a flat plastic plate with a spacer material. The plate is 1000 mm high, 490 mm wide and 6 mm thick, the plate has a total surface of 0.8 m^2 per plate. In the Beverwijk case, one unit consists of 150 plates. The 150 plates are assembled in a glass fibre reinforced plastic housing.

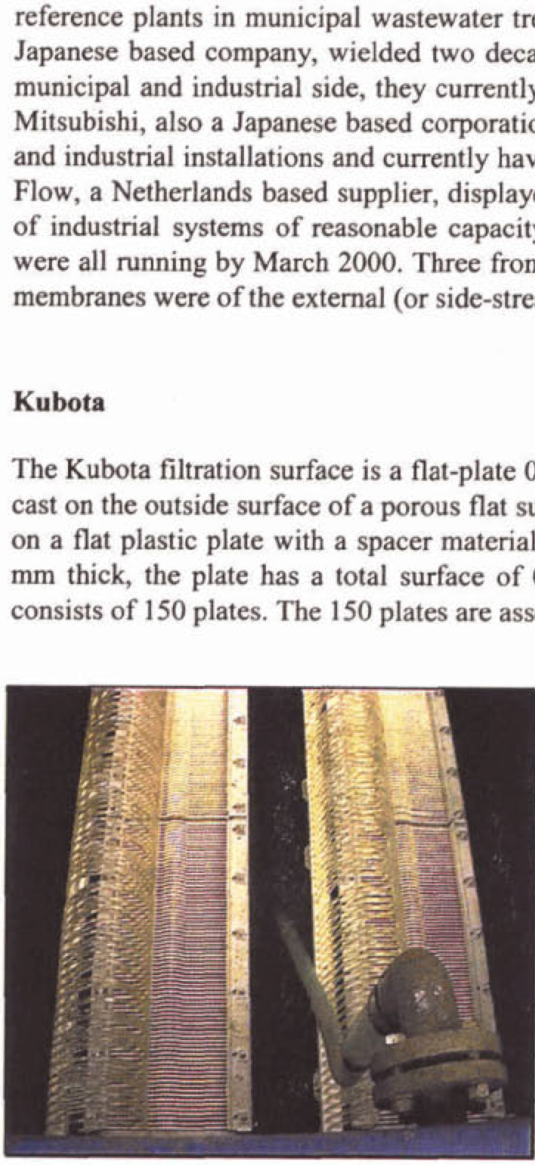


Figure 22 - Top view of two Kubota units

Figure 22 shows the top view of the Kubota units as installed in the Beverwijk Kubota pilot plant. Figure 23 shows a single Kubota unit and units installed in a tank. The plates can be constructed in many combinations, but for full-scale purposes the cartridges will be used in a single side by side configuration as shown in Figure 24, or as a Double Deck (DD) system with two or more units above each other. The water being treated passes through the membrane as clean filtrate (permeate), while the contaminants are rejected by the membrane. The treated water is discharged to drain, while the concentrated reject remains in the tank for further treatment.

The Kubota submerged micro-filtration system uses low-pressure air to maintain a turbulent cross flow pattern across the vertical membrane plates. This helps to keep the filtration surface clear of contaminant build-up, which would cause a reduction in the efficient operation of the unit.

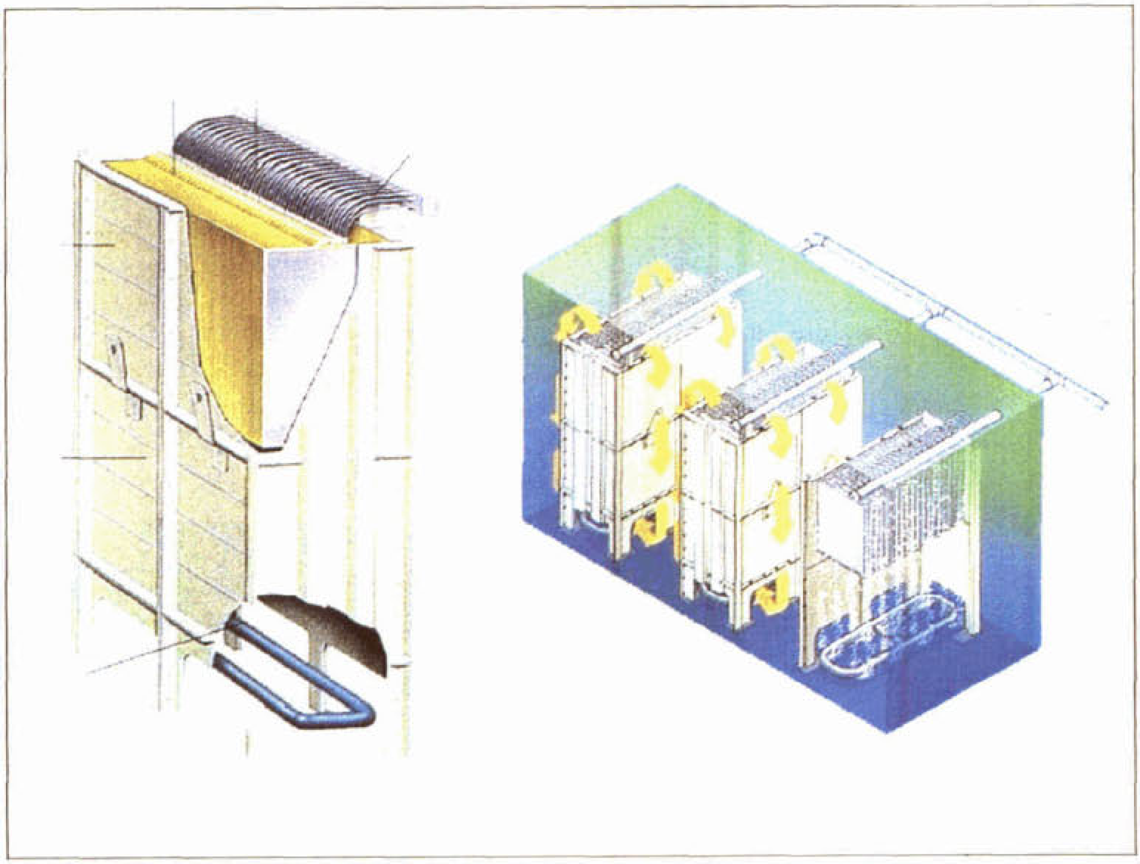


Figure 23 - Kubota unit and units installed in a tank

The filtrate (permeate) is drawn through the membrane from the bulk fluid by a partial vacuum applied within the membrane plate matrix (gravity flow is also possible), with an outside to inside filtration process. The plates can be washed from the inside at low pressure to facilitate chemical cleaning.

Figure 24 - Kubota DD system in Beverwijk

For the purpose of full scale development the DD system could be considered. This reduces the total membrane footprint required, reduces the biological volume consumed by the membrane system itself and reduces the operating cost associated with air injection to maintain a clean membrane surface (the air is effectively used twice in the DD configuration).

The DD configuration also yields a more controllable biological process and the possibility of short circuiting in the biological process is eliminated.

5.3 Mitsubishi

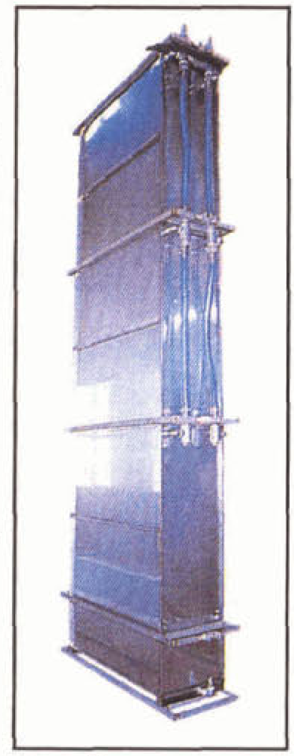


Figure 25 - Mitsubishi cartridge

The Mitsubishi filtration surface is a 0.4 μm pore size (MF) polyethylene self-supporting hollow fibre membrane (capillary). The water being treated passes through the membrane as clean filtrate (permeate), while the contaminants are rejected by the membrane. The treated water is discharged to drain, while the concentrated reject remains in the tank for further treatment.

The Mitsubishi submerged micro-filtration system uses low-pressure air to maintain a turbulent flow pattern across the horizontal membrane fibres. This helps to keep the filtration surface clear of contaminant build-up, which would cause a reduction in the efficient operation of the unit. The filtrate (permeate) is drawn through the membrane from the bulk fluid by a partial vacuum applied within the membrane fibre. The membranes are relaxed as the main recovery process, but are periodically back-pulsed by pumping a fraction of the permeate back into the membrane fibres, forcing flow in the reverse direction to remove fouling material from the surface of the membrane.

Figure 26 - Top view Mitsubishi element

The SUR234 is the latest membrane from Mitsubishi suitable for the wastewater market, but only the module configuration has changed, not the fibre. The fibre is 540 μm in diameter with a 360 μm hollow centre. The hollow fibres are built into an element of around 1000 mm x 400 mm with a double left and right sided permeate header, which in turn is built into a cartridge. Each element houses approximately 2000 flexible horizontal hollow fibres and can be built into almost any size cassette. The cassette of three layers at Beverwijk consisted of 315 m^2 membrane surface with an outside to inside filtration process. Figure 26 shows a top view of this cassette.

5.4 X-Flow

The X-Flow filtration surface is a 0.03 μm pore size (UF) PVDF membrane on a tubular support. The water being treated passes through the membrane as clean filtrate (permeate), while the contaminants are rejected by the membrane. The treated water is discharged to drain, while the concentrated reject is returned to the tank for further treatment.

The X-Flow external ultra-filtration system uses low pressure air and low pressure sludge circulation to maintain a turbulent flow pattern along the vertical membrane tubes. This helps to keep the filtration surface clear of contaminant build-up, which would cause a reduction in the efficient operation of the unit.

Figure 27 - X-Flow element

The filtrate (permeate) is drawn through the membrane from the bulk fluid by a partial vacuum applied on the outside of the tube bundle (space). Periodically, the membranes are back-pulsed by pumping a fraction of the permeate back through the membrane tubes, forcing flow in the reverse direction to remove fouling material from the surface of the membrane. Relaxation is also used to recover the permeability as the module is removed from 'operation' mode.

The Compact Element Type 38PRV/F-4385 has been tested at Beverwijk with a surface area of 29 m^2 in the 3 m high, 8 inch tube bundle as shown in Figure 27. Each element houses approximately 600 5.2 mm vertical tubes and the filtration process is from inside to outside. In the second pilot plant in Beverwijk, four F-4385 elements with 5.2 mm tubes and four F-8385 elements with 8.0 mm tubes have been tested.

5.5 Zenon

The ZeeWeed® filtration surface is a neutral 0.035 μm pore size (UF), strong polymeric membrane cast on the outside surface of a porous hollow support fibre. Total diameter of a single fibre amounts 2,0 mm. Bundles of fibres are assembled in a module, eight modules together form a cassette as shown in Figure 28. The water being treated passes through the membrane as clean filtrate (permeate), while the contaminants are rejected by the membrane. The treated water is discharged to drain, while the concentrated reject remains in the tank for further treatment.

The ZENON ZeeWeed® submerged ultra filtration system uses low pressure air to maintain a turbulent flow pattern along the vertical membrane fibres. This helps to keep the filtration surface clear of contaminant build-up, which would cause a reduction in the efficient operation of the unit.

Figure 29 shows a schematic of a ZeeWeed® membrane fibre in operation. The filtrate (permeate) is drawn through the membrane from the bulk fluid by a partial vacuum applied within the membrane fibre lumens, with an outside to inside filtration process. Periodically, the membranes are back-pulsed by pumping a fraction of the permeate back into the membrane fibres, forcing flow in the reverse direction to remove fouling material from the surface of the membrane.

Figure 30 shows a top view of the Zenon membrane pilot installation (ZW500a) at Beverwijk.

Figure 28 - Zenon Cassette

The ZW500a has been superseded by the ZW500c in 2001, but only the module configuration has changed slightly. The fibre is approximately 2,000 μm in diameter with a 1,000 μm hollow centre. The hollow fibres are built into an element of some 2,155mm x 820mm with a single top permeate header (ZW500c), which in turn is built into a cassette. Each element houses approximately 1800 flexible vertical hollow fibres and 22 elements make up one cassette. The unit càssette (ZW500c) for large systems houses 440 m^2 of membrane surface. The ZW500c promotes a higher packing density of fibres, thus more surface area in a set footprint. The air distribution has been improved to allow a more intense bubble penetration into the fibre bundle preventing the potential build up of sludge. Both the ZW500a and the ZW500c modules are tested in two separate pilots.

The ZW500c pilot is used to establish the non-sludging ability of the new module, whereas the original pilot is used as the MBR.

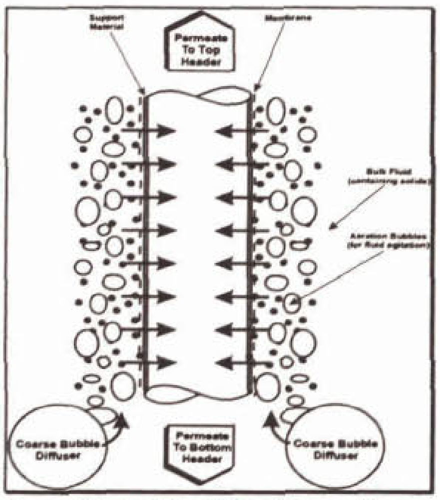


Figure 29 - ZeeWeed fibre in operation

Figure 30 - Top view Zenon membranes

6 KUBOTA

6.1 Introduction

Figure 31 - Photograph Kubota pilot plant

In this chapter the performance of the Kubota pilot plant is described. In section 6.2 the system configuration and design figures are given. In section 6.3 and 6.4 respectively the biological and membrane performance is presented. Finally in section 6.5 the conclusions are summarised.

The Kubota pilot plant was in operation since week 17 of the year 2000. In Table 6 the main activities for the Kubota pilot plant research are summarised.

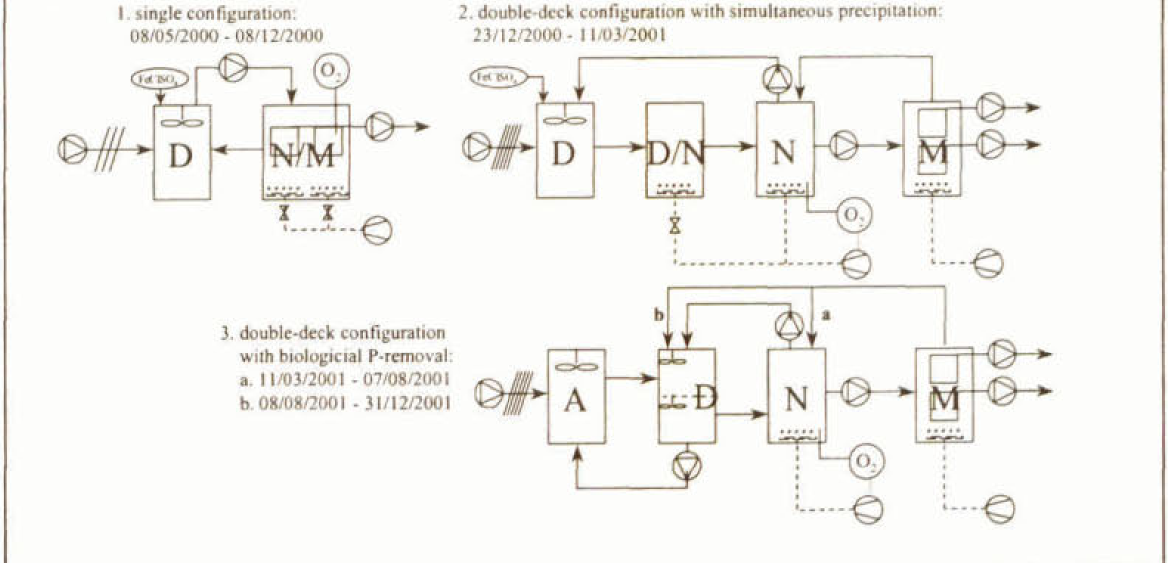
Table 6 - Research activities

phase 1	**primary clarification with pre-precipitation** 27/03/00 - 03/10/00 week 17/2000 - 40/2000
week 17-18	testing installation and seeding with sludge
week 19-24	constant flow
week 25-29	proportional flow
week 30-38	peak flow testing, cleaning and optimisation
week 39-41	low temperature test, incl. peak test
phase 2	**primary clarification with simultaneous precipitation** 04/10/00 - 10/01/01 week 40/2000 - 02/2001
week 42-44	sludging between membranes, followed by a mechanical cleaning
week 45-50	proportional flow
week 50-51	rebuilding to a double-deck configuration
week 52-02	bottom module at constant flow, adjustments to software
phase 3	**raw influent with simultaneous precipitation** 11/01/01 - 11/03/01 week 02/2001 - 10/2001
week 02-10	proportional flow, optimisation operational settings
phase 4	**raw influent with biological phosphorus removal** 12/03/01 - 31/12/01 week 11/2001 - 52/2001
week 11	rebuilding to a system with biological phosphorus removal
week 11-52	proportional flow, focus on fouling characteristics of DD configuration

6.2 System configuration

The configuration and design figures for both the single configuration and the double-deck configuration are presented in Table 7. The configuration and process settings have changed during the research period. The main changes are described in Table 7. The most important change was the rebuilding of the system from a single to a double-deck configuration, which was developed by Kubota in Japan for larger full-scale applications (see paragraph 5.2).

Table 7 - Design configurations and figures



process part	parameter	unit	single	double-deck	
			chemical P removal		bio-P
			< 08/12/00	*23/12/00 - 11/03/01*	*> 11/03/01*
influent pump	capacity	m^3/h	15		
	RWF design flow	m^3/h	7.8 → 10 [4]		
	design flow	m^3/d	39 → 50 [4]		
influent screen	type	-	basket filter	rotating drum	
	slot size	mm	2.0	0.25 → 1.0	
aeration tank	total volume (and depth)	m^3	30.8	34.0	42.8
	- anaerobic volume (A)	m^3 / (m)	-	-	13.5 (2.7[1])
	- anoxic volume (D)	m^3 (m)	12.0 (2.4[1])	10.1 (2.0[1])	8.4 (2.7[1])
	- fac. compartment (D/N)	m^3 (m)	-	5.7 (1.8[1])	-
	- oxic compartment (N)	m^3 (m)	-	5.7 (1.8[1])	8.4 (2.7[1])
	- membrane tank (M)	m^3 (m)	18.8 (3.0)	12.5[2] (4.05)	12.5[2] (4.05)
aeration biology	compressor capacity total	Nm^3/h	-	160	80
ferric dosing	type	-	$FeClSO_4$		-
	ferric content	%	12.3		
	capacity (at Me/P = 0.8)	ml/h	160		
membrane filtration	number of modules	-	2 (x 150 plates)		
	total surface	m^2	240		
	max. net flux (at RWF)	$l/(m^2.h)$	32.5 → 41.7 [4]		
aeration membranes	compressor capacity total	Nm^3/h	90 - 180	115	
	specific capacity	$Nm^3/(m^2.h)$	0.75	0.48	
re-circulation pumps	D → N/M	m^3/h	15	-	-
	N → M	m^3/h	-	10-30 (5:1) [3]	10-30 (5:1) [3]
	N → D	m^3/h	-	15	15
	D → A	m^3/h		-	9

1: the level in the anoxic and oxic tank is adjustable
2: excluding the membrane module volume
3: the re-circulation flow is related to the feed flow (ratio 5:1)
4: the hydraulic design load is increased during phase 1

As a consequence of the system rebuilding the biological / membrane aeration of the system was changed. In the single configuration system the membrane tank also functioned as a nitrification tank. The coarse bubble aeration required for the membrane operation was also responsible for the oxygen input for the biological process. The coarse bubble aeration capacity was based on the number of membrane modules in operation; 1 module in operation: 90 Nm^3/h, 2 modules: 180 Nm^3/h. In the case that one module was in operation (normal operation) and the oxygen concentration was below the required set point, the second modules aeration would also be switched on (large aeration excess).
In the double-deck configuration there was a separate compressor for the aeration of the sludge in the compartments N1 and N2. The aeration in compartment N1 could be shut down by the aeration control. The aeration control was based on the oxygen measurement in the second aeration compartment (N2). The coarse bubble aeration in the membrane tank depended on the membrane settings (see section 6.4) and was set at a fixed flow. In week 11/2001 the system was rebuilt into a system with an anaerobic tank. In the following period the feasibility of biological phosphorus removal is investigated. The mixing in the anoxic tank (D) is executed by intermittent aeration (10 seconds ON, 10 minutes OFF).

6.3 Biological performance

6.3.1 Process conditions

The main process conditions are presented graphically in Figure 32 to Figure 34. The overall process conditions during the phases of the pilot research are presented in Table 8.

During the first weeks after start-up the pilot was fed with a constant flow. Since week 25 the feed to the pilot is related proportionally to the feed of the full-scale Beverwijk WWTP. The proportional settings were 1:1,000 and were maintained during the whole research period. As a consequence the biological loading has increased step-wise in phase 2 and 3. The design flow at dry weather conditions amounts to 50 m^3/d. The average influent flow during the research period varies strongly due to rain weather conditions.

Table 8 - Process conditions of the pilot plant

parameter	unit		phase 1	phase 2	phase 3	phase 4
influent flow	m^3/d		48	72	51	51
process temperature	°C	average	21	13	12	19
		range	14 - 25	8 - 20	6 - 14	10 - 28
pH	-		7.4	7.4	7.3	7.4
biological loading	kg COD/(kg MLSS.d)		0.054	0.060	0.084	0.100 #
sludge concentration	kg $MLSS/m^3$		10.5	12.0	10.6	10.8
organic part	%		59	61	63	63
sludge production	kg MLSS/d		4.6	9.0	13.5	10.3
sludge age	d		70	41	27	30 #
ferric dosing AT	l/d		0	1.7	1.7	0
ferric dosing ratio	mol Fe/mol P		0	0.29	0.24	0

\# calculation based on the activated sludge volume excl. anaerobic tank (29.3 m^3)

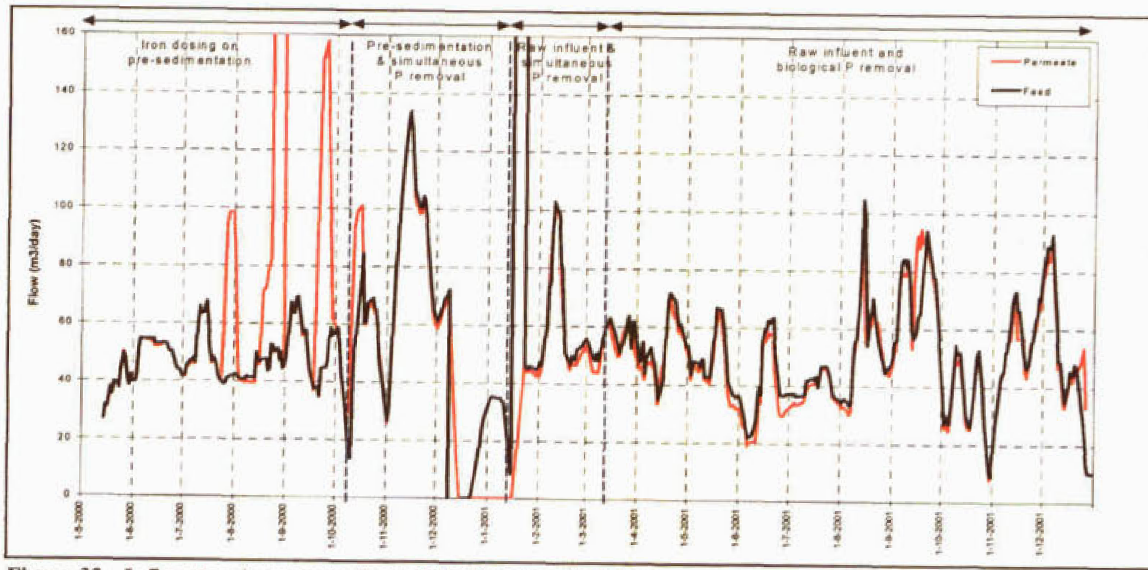


Figure 32 - Influent and permeate flow of the Kubota pilot plant

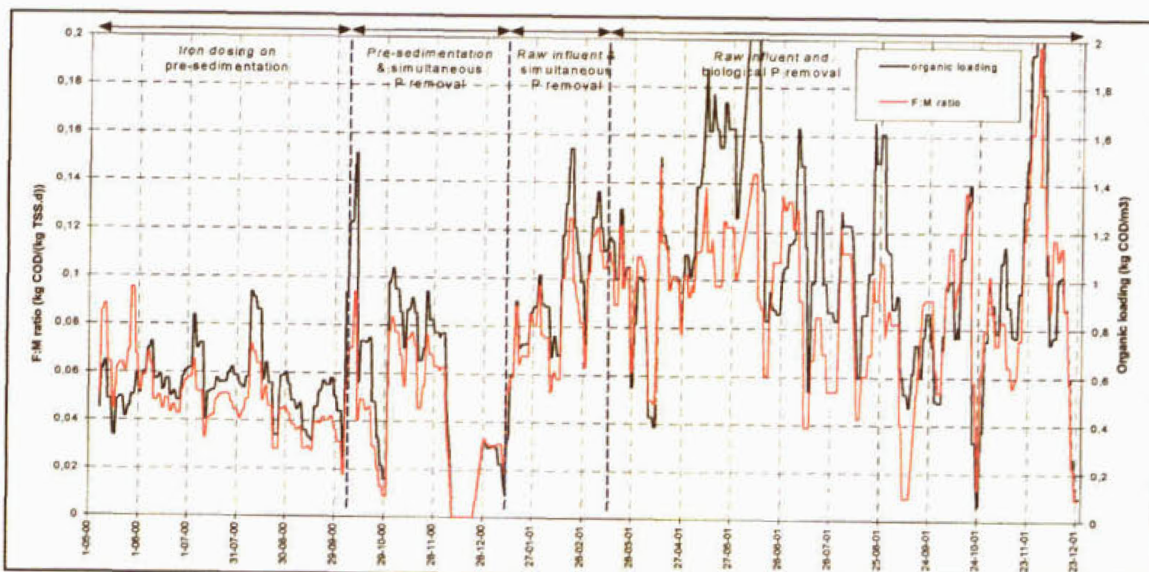


Figure 33 - Biological loading of the Kubota pilot plant

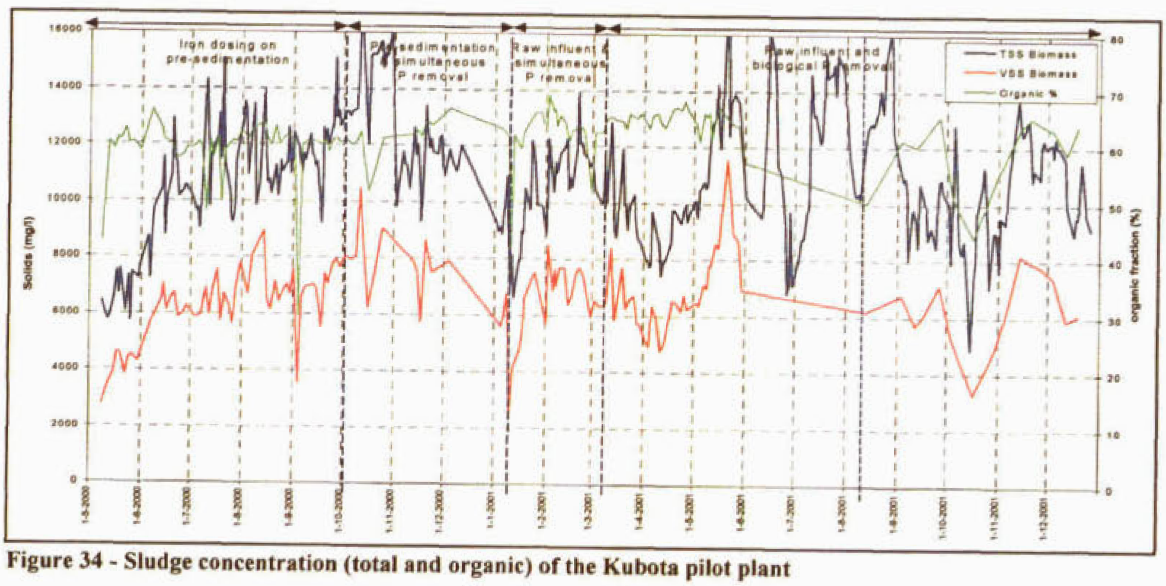


Figure 34 - Sludge concentration (total and organic) of the Kubota pilot plant

The original set-point for the sludge concentration was 12 kg MLSS/m^3. In October the set-point was increased to 15 kg MLSS/m^3 at Kubota's request. During this period the membranes clogged with sludge due to the blocking of the coarse bubble aeration system. After the mechanical cleaning on 30 October, the set-point was reduced to the original sludge concentration of 12 kg MLSS/m^3. To prevent clogging, a flushing system was installed. It should be stated that the blocking of the aeration system most probably was not a direct consequence of the sludge concentration increase. A typical sludge concentration for Kubota systems under Japanese conditions is 15-20 kg MLSS/m^3.

In the original single configuration, the coarse bubble aeration system of the Kubota membranes was also responsible for the complete oxygen input for the biological process. As a consequence, the control flexibility of the oxygen input (for biological processes) was limited and the oxygen concentration remained relatively high, especially during the night at low organic loadings. After introducing the double-deck configuration this situation was greatly improved.

The process conditions during the research periods differ significantly as shown in Table 8. Besides the large influent flow variations also the process temperature and the biological loading had changed. The ferric dosing ratio during the simultaneous precipitation period has been relatively low due to technical optimisation reasons.

6.3.2 *Results*

The main biological parameters are presented graphically in Figure 35 to Figure 37. In Table 9 the average feed and permeate concentrations are presented.

Table 9 - Feed and permeate concentrations

parameter		unit	phase 1	phase 2	phase 3	phase 4
COD	feed	mg/l	341	297	568	621
	permeate	mg/l	32	21	31	32
	efficiency	%	91	93	95	95
N_{kj}	feed	mg/l	61	40	61	58
	permeate	mg/l	2.7	1.5	1.6	3.0
NO_3-N	permeate	mg/l	8.6	7.1	8.0	7.9
N_{total}	permeate	mg/l	11.3	8.6	9.6	10.8
	efficiency	%	81	79	84	81
P_{total}	feed	mg/l	7.4	8.3	10.1	10.9
	permeate	mg/l	2.8	1.3	1.3	0.8
	efficiency	%	62	84	88	93

COD removal

The COD-effluent concentration was relatively low due to the fact that the non-soluble fraction had been completely removed by the membranes. The COD effluent concentration in phase 2 was only 21 mg/l, which was lower than the concentration in phase 1, 3 and 4. This is due to the dilution effect during long rain weather conditions.

Nitrogen removal

In phase 1 the nitrate concentration in the permeate was relatively high with an average concentration above 10 mg/l due to the pre-precipitation process. The COD/N ratio of the feed was 6 which is relatively low. In phase 2 this ratio increased to 8 and consequently nitrogen removal improved. In this respect also the RWF dilution effect must be taken into account.

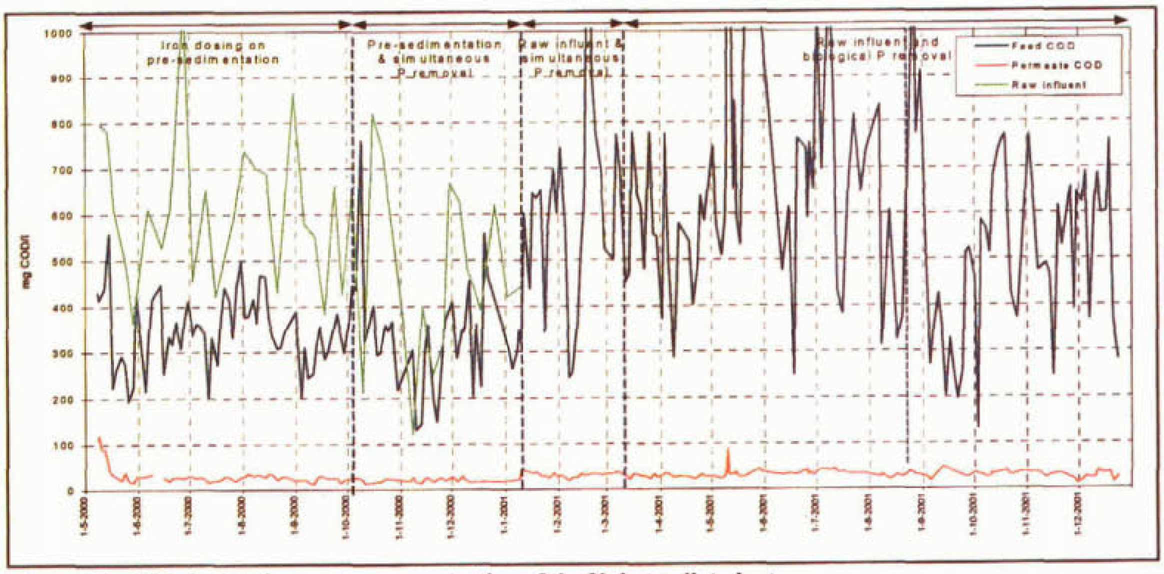


Figure 35 - COD feed and permeate concentration of the Kubota pilot plant

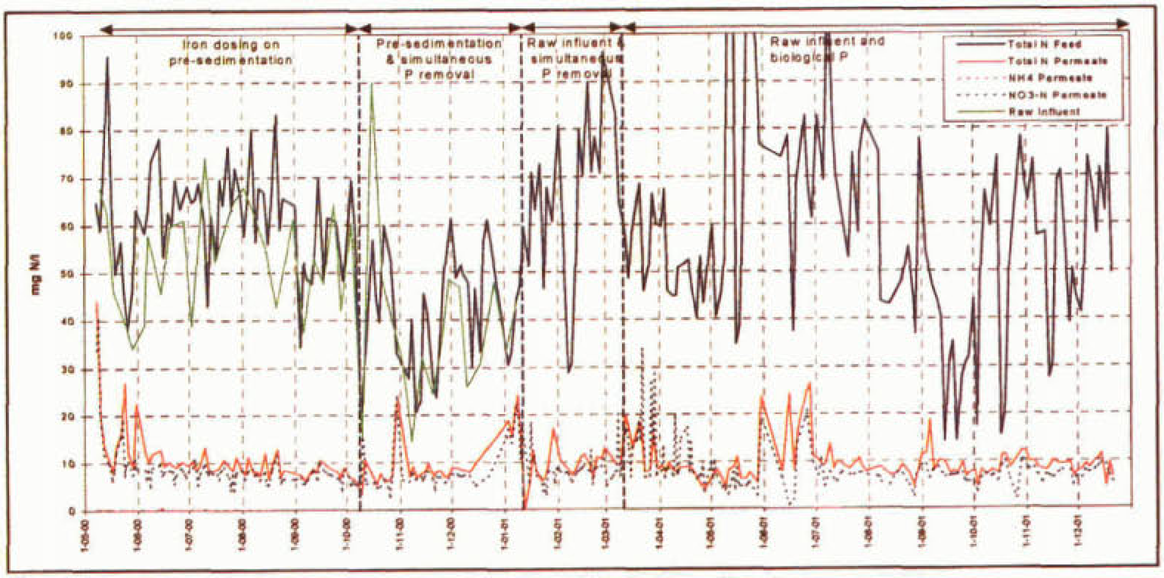


Figure 36 - Nitrogen feed and permeate concentration of the Kubota pilot plant

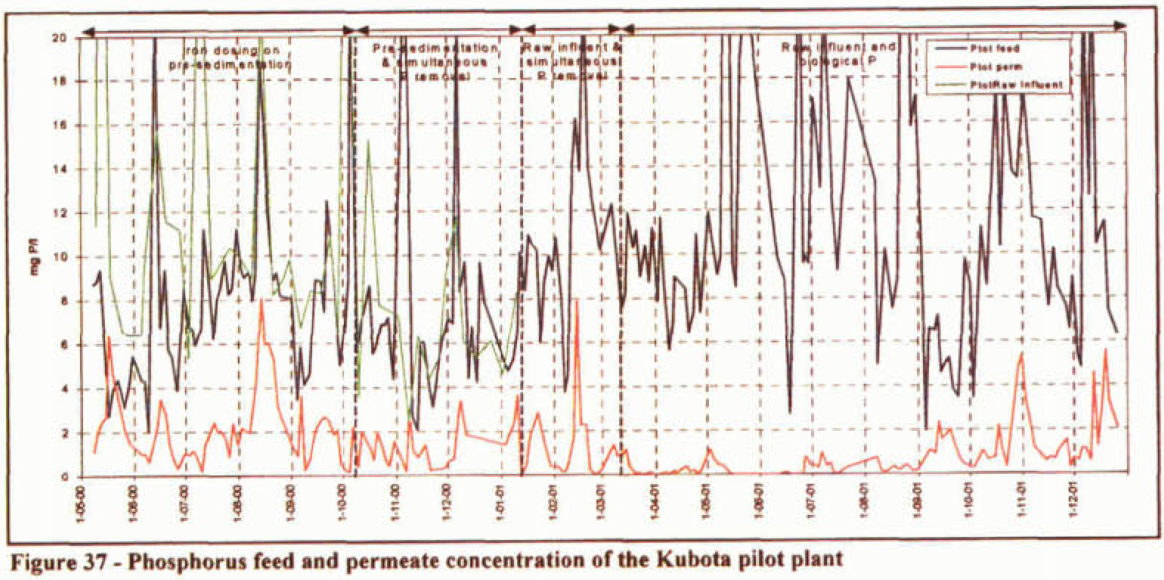


Figure 37 - Phosphorus feed and permeate concentration of the Kubota pilot plant

The nitrogen removal in phase 3, without primary clarification, performed well. The COD/N ratio was higher (around 10) and not limiting. In phase 4 effect of bringing the recirculation flow from the membrane tank directly into the anoxic compartiment, was investigated. The effluent concentration did not improve in this period. The limiting factor for the denitrification process in phase 3 and 4 was the internal recirculation flow. This will be further discussed in chapter 11.

Phosphorus removal

In phase 1 it was not possible to achieve a phosphorus effluent concentration < 1 mg P_{total}/l at a Me/P dosing ratio of 1.2. Both the removal efficiency of the primary clarifiers and the sludge production were too low. In phase 2 and 3a the phosphorus removal was improved. The average effluent concentration amounted to 1.3 mg P_{total}/l at a relatively low dosing ratio. The chemical dosing was constant and not related to the influent flow or the phosphorus concentration. As a consequence, the permeate phosphorus concentration varies.

In phase 4 the chemical dosing is stopped and biological phosphorus removal is introduced. The hydraulic retention time in the anaerobic tank was about 1 hour. The permeate phosphorus concentration decreases to an extreme low level (< 0.1 mg/l) and remains very stable even at high influent peak concentrations. Since September 2001 the phosphorus permeate concentration increased to an average concentration of 1.5 mg P_{total}/l.

6.3.3 *Sludge characteristics*

The main sludge characteristics are presented in Table 10.

Table 10 - Sludge characteristics

	unit	phase 1	phase 2	phase 3	phase 4
Sludge characteristics					
DSVI	ml/g	80	90	90	90
CST	s	50	70	60	50
Y-flow	s	240	540	n.a.	100
Viscosity					
viscosity value	mPa.s	7.8	8.3	8.9	5.8
shear rate	1/s	90	130	130	60
α-factor					
surface aeration	-	0.52	0.54	0.50	n.a.
bubble aeration	-	0.46	n.a.	0.54	n.a.
Gravity thickening					
settling velocity	cm/h	6.0	4.0	5.0	n.a.
thickened sludge concentration	%	2.8	3.4	2.8	n.a.
Mechanical thickening					
MLSS at 3900 rpm / 10 min	%	7.3	5.8	10.4	n.a.
MLSS at 1000 rpm / 3 min.	%	3.0	2.3	5.1	n.a.

The DSVI was relatively stable at a level of 80-90 ml/g, which was rather low. This is related to a good floc structure as described in the next section. Also the CST and the Y-flow were low. Both CST and Y-flow seemed to be related to the sludge concentration. The viscosity of the MBR sludge was 8 mPa.s on average, which was relatively low for the applied sludge concentrations. The viscosity of the MBR sludge was related to the sludge concentration.

The average α-factor value for surface aeration remained constant at 0.50-0.55 during the three research phases. With a surface aerator 10% higher values were achieved in phase 1 than with bubble aeration. The α-factor with bubble aeration increased during the research period from <0.5 in phase 1 to >0.6 in phase 3, and was eventually higher than with surface aeration.

The average settling velocity at the beginning of the gravity thickening test amounted to 4 to 6 cm/h on average, depending on the sludge concentration. The thickened sludge concentration (after 24 hours) varied from 2.8-3.4%. The achievable sludge concentration with mechanical thickening increased from 6-8% in phase 1 and 2 to 8-15% in phase 3.

During the start-up the flocs had a medium size with a compact structure, also some filamenteous organisms were present. Since the start-up, more than 10 scale amoebas per sample were observed. During the summer of 2000 the majority of the flocs showed an open structure, while during phase 2 the flocs became more compact and ciliates and mono cultures were present. The number of mono-cultures had decreased dramatically by the end of phase 2. Normally, some crawling ciliates (Aspidisca) and free swimming ciliates (Euplotus) were observed. The number of filaments had dropped below 1 according to Eikelboom, but seemed to be part of the microbial population.

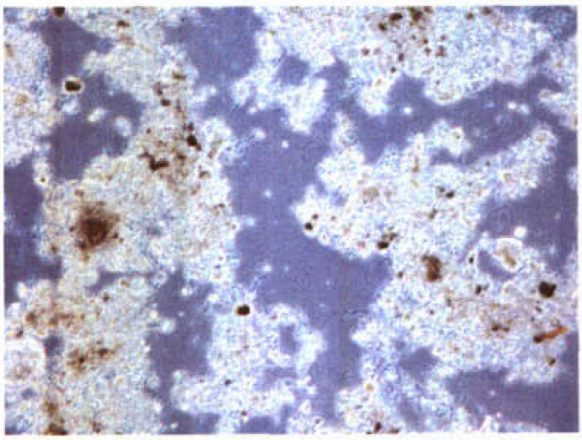


Figure 38 - Microscopic view of the activated sludge

6.4 Membrane performance

The results of the whole research period are presented graphically in the fold out page and further explanation is given in the text. The data presented is damped to assist readability and conclusions are made regarding the membrane performance and the relevant operating and design criteria.

It was expected that the maximum achievable flux would be 42 $l/(m^2.h)$ net (single configuration) for a period of 3 days continuously. The cleaning was estimated at twice per year and the cleaning method was set according to Kubota's specifications of 5,000 mg/l NaOCl followed by 1% Oxalic acid solution. Relaxation was not expected in the process operation mode.

Until 18/09/00 the pilot had no relaxation and the permeation rate was at 100% of the incoming flow, under these conditions the gross flux was the same as the net flux.
After 18/09/00 the membranes were periodically relaxed with air as a form of in-situ membrane surface cleaning (2 minutes relaxation after 8 minutes process mode). During this period the ratio between net flux amounted 80% of the gross flux.
From on 23/12/00 (double-deck mode) the frequency of the relaxation was kept the same but the onset was changed. On average the current efficiency of the new set up was 80% for RWF and 85-90% for DWF.

6.4.1 *Phase 1 - Primary settled wastewater and pre-precipitation*

Start up and definitions

To achieve a stable biological start up the system was run at the 24 hour average daily flow rate of 42 m^3/d. This flow was later increased to 55 m^3/d to increase the biological growth. The membranes followed the feed flow via a set level in the bioreactor and the modules switched in and out according on timers.

Electrical problems during start up the membrane system caused the biomass to thicken to 25 g/l, it was seen that the module 1 and module 2 were not aerated during permeate extraction and aerated only during relaxation. The latter caused the membrane permeability to drop rapidly to about 300 l/(m^2.hr.bar) even at very low hydraulic loading, the membrane were significantly fouled. In practise such an accident would be prevented by interlocking the suction pump and the blower.

The pilot displayed stable operation after six weeks with an acceptable sludge growth and the research programme proceeded to a proportional flow control at a ratio of 1000:1 to that of Beverwijk WWTP. The permeability of the Kubota pilot plant steadily decreased in time from 300 l/(m^2.h.bar) to 275 l/(m^2.h.bar). However, during the first real RWF onset, the permeability decreased to 175-200 l/(m^2.h.bar), and at DWF flow the permeability stabilised at a level of 200-250 l/(m^2.h.bar). Limited recovery of permeability was seen.

Peak tests without Relaxation

In July 2000 a peak tests was executed. During the peak test the influent pump was set on a constant flow, permeate re-circulation was used for increasing the hydraulic loading of the pilot plant but not the biological loading. The test was divided into two periods, which are described below. The permeability and the flux during and after the peak tests are presented in Figure 39.

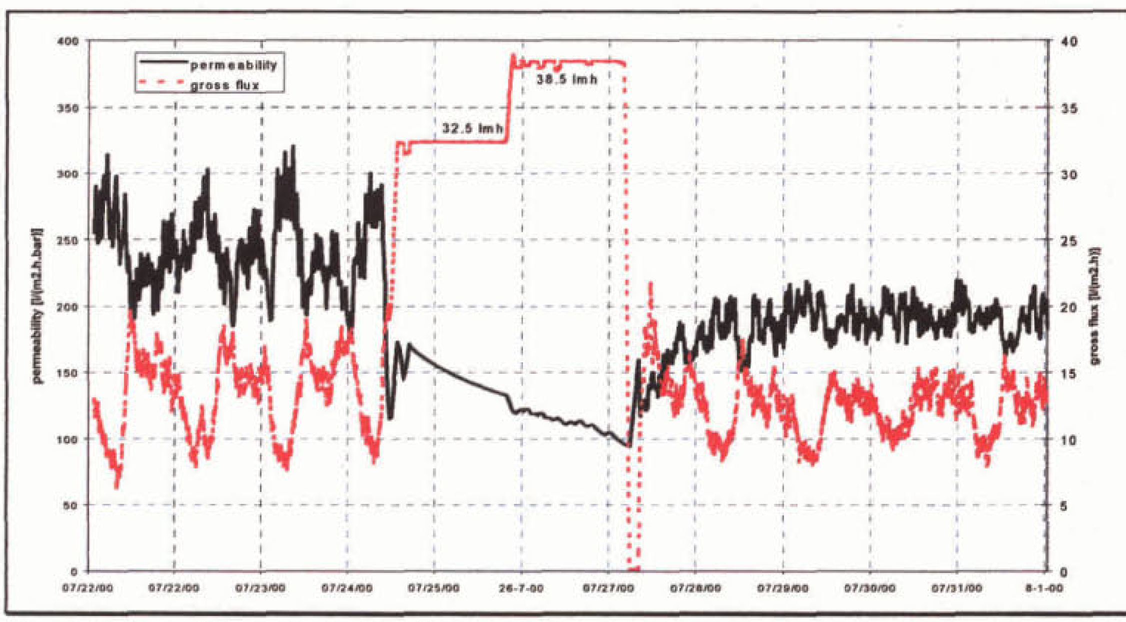


Figure 39 - Permeability and gross flux during and after peak testing

Before the peak test started, the permeability varied from 200-275 l/(m^2.h.bar), dependent on the flow variations of DWF. After starting the 32.5 l/(m^2.h) peak test, the permeability rapidly decreased to 170 l/(m^2.h.bar) and subsequently slowly decreased to 130 l/(m^2.h.bar). From the permeability curve it seemed that the permeability would have stabilised at a level > 100 l/(m^2.h.bar).

Table 11 - Peak test settings of the Kubota plant

period	duration [hours]	influent flow [l/h]	permeate flow [l/h]	total flow [l/h]	flux [$l/(m^2.h)$]
1	30	2,000	5,800	7,800	32.5
2	33	2,250	7,000	9,250	38.5

The 38.5 $l/(m^2.h)$ peak test was started immediately after the previous peak test. As a result, the permeability steadily decreased to 100 $l/(m^2.h.bar)$ and would have decreased further if the test had not been stopped. The permeability was not stable, and the peak flux expectation of 42 l/(m2.h) appeared unrealistic.

On returning to proportional DWF, the permeability recovered steadily to about 175 $l/(m^2.h.bar)$ within 24 hours. In the next days the permeability remained stable at this level, which was lower than the original permeability level before the peak test; further fouling had occurred. The above peak test without relaxation inferred that the Kubota membranes could not handle the 38.5 $l/(m^2.h)$ net peak flux under the circumstances described above.

Cleaning procedure

Since the start of the pilot research, one of the modules was partly fouled and thus had a large influence on the results, caused by a prior technical failure. During the whole research period until this point no chemicals were used. The first official Kubota cleaning procedure was executed on 15/8/00. The cleaning procedure was a standard procedure from Kubota Corporation and consisted of 2 separate chemical treatments; first a cleaning with a 5,000 mg/l NaOCl solution, followed a day later by a cleaning with a 1% oxalic acid solution, both steps were carried out in situ in biomass.

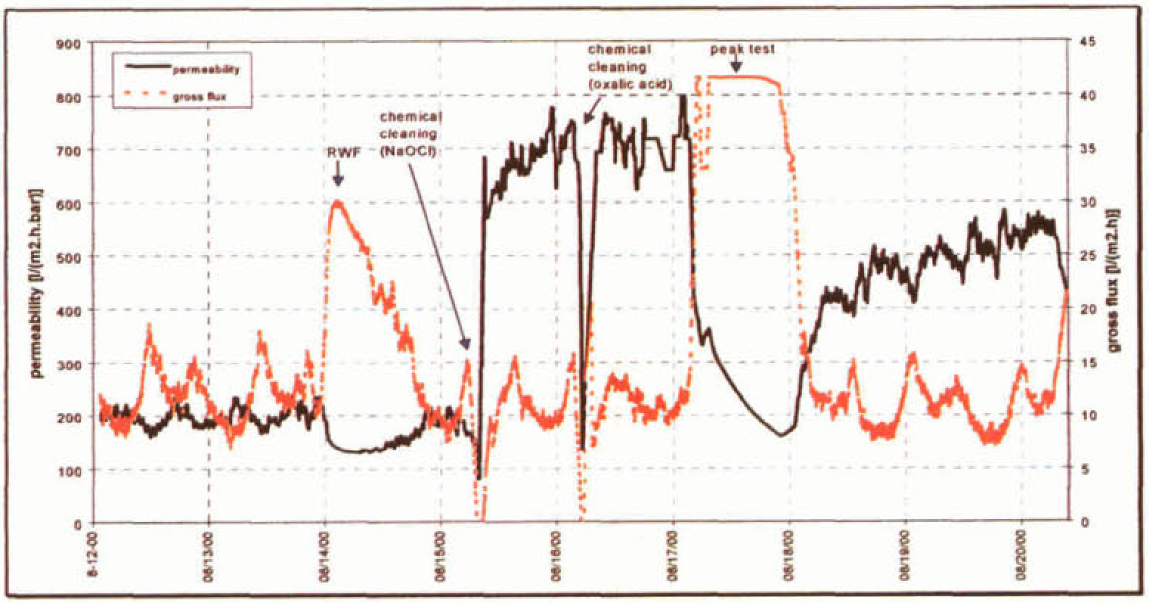


Figure 40 - Permeability and gross flux during and after cleaning and peak testing

As a result of the NaOCl cleaning the permeability increased from 200 to 700 $l/(m^2.h.bar)$. This was equal to the clean process permeability as measured at the start of the project. The oxalic acid cleaning had little additional effect on the permeability. The fouling was predominately biomass surface fouling or organic in nature.

After the membranes were satisfactorily cleaned, a peak test was executed at a peak flux of 41.7 $l/(m^2.h)$. The peak test lasted approximately 20 hours at a process temperature of 23-24°C. The permeability decreased catastrophically from 700 to 170 $l/(m^2/h.bar)$. The system was manually stopped and set to proportional flow. The permeability recovered within 12 hours to >400 $l/(m^2.h.bar)$ and in one month it had eventually recovered to 700 $l/(m^2.h.bar)$, the original permeability just after cleaning. The rapid decline in permeability at high fluxes suggested that the filtration process rapidly regressed to cake filtration, and the subsequent recovery process indicated that relaxation was necessary to re-establish a clean membrane surface.
From the peak tests it was concluded that the Kubota membranes could not handle a 38.5 or 41.7 $l/(m^2.h)$ net peak flux, however, a 32.5 $l/(m^2.h)$ was possible under the circumstances described (T>20°C). It was concluded that the Kubota plant must be optimised by introducing a relaxation period between process modes.

Peak test with relaxation
The introduction of relaxation resulted in an efficiency reduction to 80% and would have to be compensated in the peak flux (software). After the introduction of the relaxation mode, a 100 hour peak flux was executed. The permeability and the gross flux during and after the peak tests are presented in Figure 41. During the peak test the process temperature ranged from 21 to 23°C.

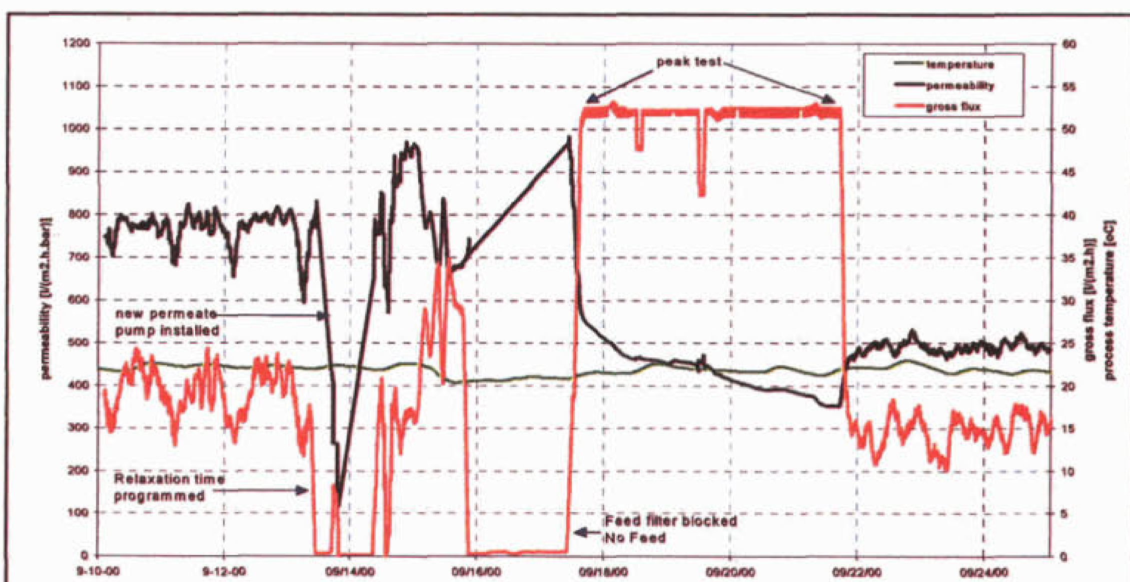


Figure 41 - Permeability, gross flux and process temperature during and after peak test with relaxation

Before the peak test started, the permeability varied from 650-950 $l/(m^2.h.bar)$, depending on the DWF variations, indicating a clean membrane. The peak test was started at 42 $l/(m^2.h)$ net or 52.5 $l/(m^2.h)$ gross, the permeability rapidly decreased to 550 $l/(m^2.h.bar)$ and subsequently slowly to 450 $l/(m^2.h.bar)$ in 48 hours. The permeability further decreases to a minimum level of 360 $l/(m^2.h.bar)$. On stopping the peak test, the permeability recovered steadily to around 500 $l/(m^2.h.bar)$ within 12-24 hours. The permeability remained stable at this level which was lower than the permeability before the peak test, but was still relatively high.
It was concluded that the Kubota membranes were able to handle the 42 $l/(m^2.h)$ peak flow under the circumstances described above. The relaxation seemed to have a most positive impact on the membrane performance at peak flow.

The above test defined the maximum operating net flux of the membrane and proved to be sustainable for several days. The argument however, was then directed to the operating temperature of the peak test, 22.5°C. Was the peak flux also sustainable at lower temperatures as could be expected under winter conditions.

Low temperature test

From 26/09/00 to 10/10/00 a low temperature test was executed. Biomass was circulated through a heat exchanger to achieve an overall biomass temperature of 10°C. During the cooling period the membranes ran continuously between 12 to 25 $l/(m^2.h)$ gross flux, but unfortunately due to the size of the pilot the feed flow had to be dropped further to reduce the heat sink. Eventually the 10°C was reached with a feed flow of 950 l/h, the cooling process lasted 14 days and finally achieved a temperature of 10°C which avoided excessive biomass deterioration. At the end of the cooling period a 24 hour peak test was executed. During the peak test the permeate flow was re-circulated and the process temperature increased slightly to 11.5°C due to additional feed heat sink and more energy input. The result of the low temperature test and the peak test is presented below in Figure 42.

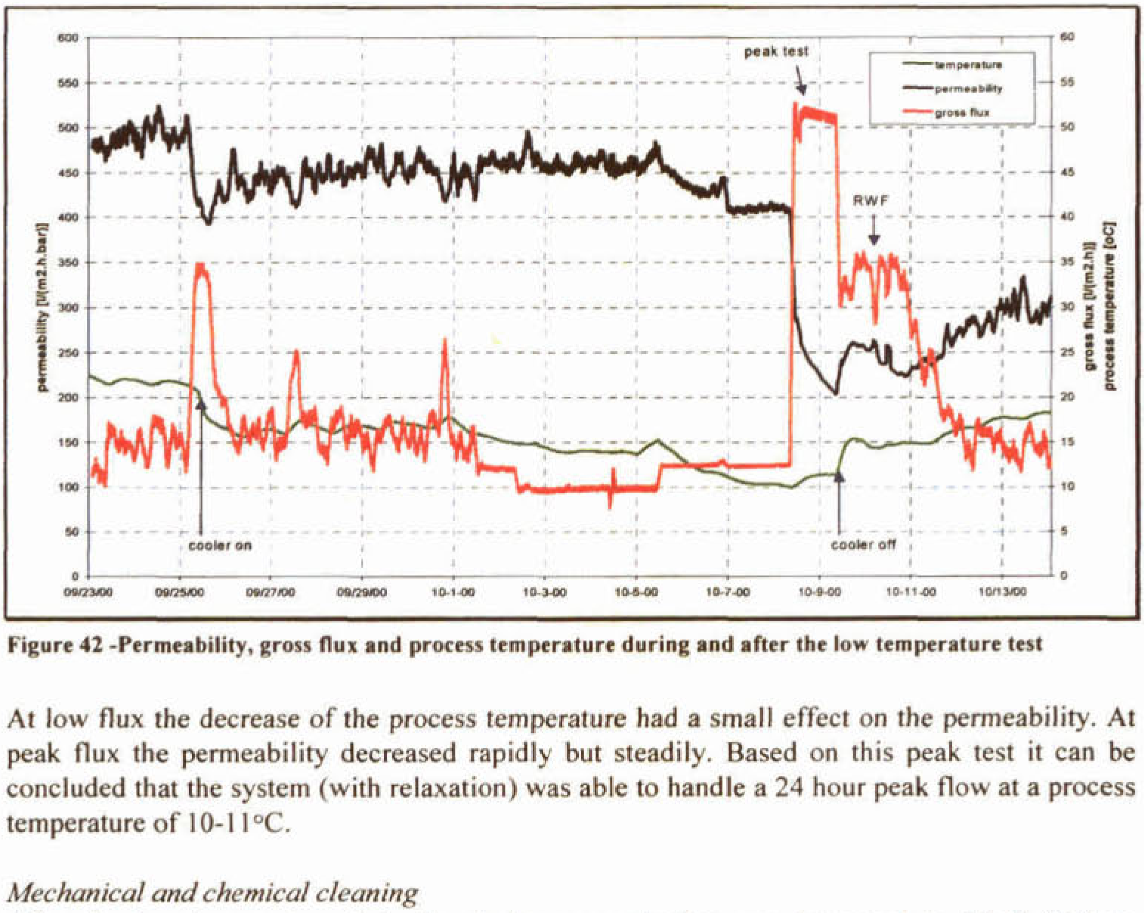


Figure 42 -Permeability, gross flux and process temperature during and after the low temperature test

At low flux the decrease of the process temperature had a small effect on the permeability. At peak flux the permeability decreased rapidly but steadily. Based on this peak test it can be concluded that the system (with relaxation) was able to handle a 24 hour peak flow at a process temperature of 10-11°C.

Mechanical and chemical cleaning

After the low temperature test, the sludge concentration was increased to 15g/l (Kubota requirement for better system performance), the permeability however, decreased further and the oxygen concentration in the system remained low. The latter, combined with visual observations of the membrane tank surface, suggested that the aeration system was partially blocked. On 30/10/2000 the membrane system was mechanically cleaned. Between the membrane plates was a thickened sludge layer with a concentration up to 15%. The aeration system under both modules was partially blocked and there was a sludge build-up under the modules and secondary aeration system. This may partly be caused by the introduction of influent particles that could not be removed by the 2 mm pre-screening.

The membrane plates required mechanical cleaning and a chemical disinfection with a 150 mg/l NaOCl solution was also carried out. The aeration system would be modified to incorporate a flushing / venturi system in the DD (Double Decker) configuration. After the mechanical cleaning the permeability recovered immediately to 600-800 l/(m^2.h.bar) and membrane integrity was maintained.

6.4.2 *Phase 2 - Primary settled wastewater and simultaneous precipitation*

The simultaneous precipitation in the pilot was started on 13/10/00 due to the prolonged cooling test. After the mechanical cleaning the system was set to proportional flow where it continued to operate until the rebuilding of the pilot to the so called DD configuration. During this time the permeability remained above 400 l/(m^2.hr.bar) suggesting a correctly working membrane. All hydraulic peaks in this period were processed without loss of overall permeability and subsequently were always followed by a recovery. The membrane system was stable as can be seen from Figure 43.

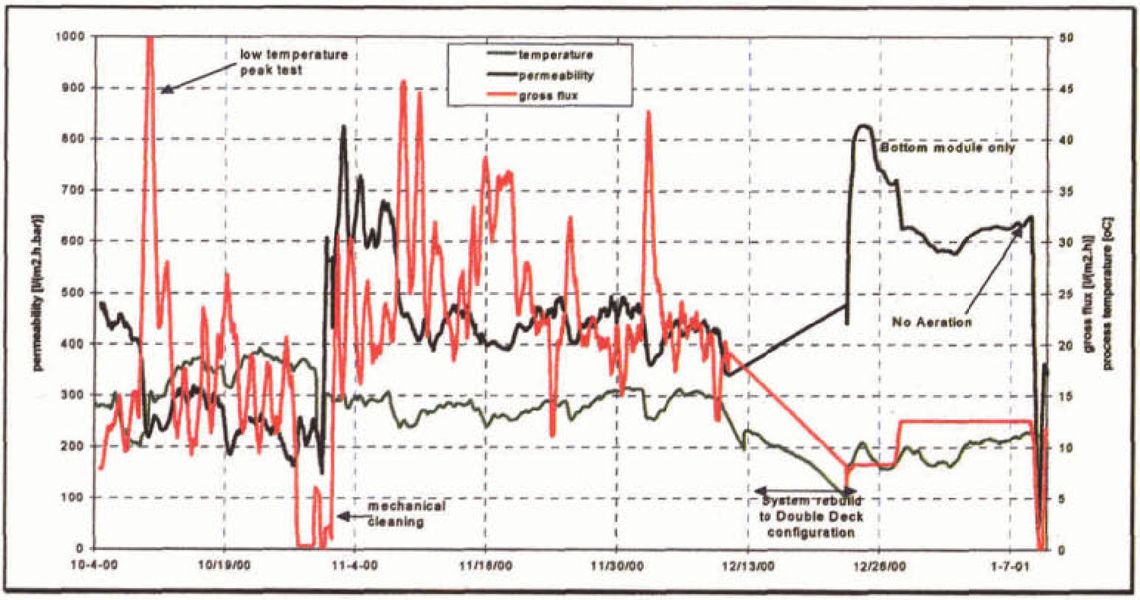


Figure 43 - Permeability and gross flux during and after mechanical cleaning procedure

6.4.3 *Phase 3 - Raw influent and simultaneous precipitation*

Between the 8th and 22nd December 2000 the pilot plant was rebuild to a so called DD configuration, in which the two existing modules were placed on top of each other. Each module had it's own permeate pump and control system. In this period the sludge was aerated continuously and fed at a very low rate to maintain the biomass as good as possible. The membranes were kept outside in the open air for one week and subsequently in a water filled tank with a low NaOCl concentration (disinfection) for a few days.

Initially only the bottom module was started up at low flux, continuous operation at approximately 10 l/(m^2.h). As a consequence the permeability decreased to around 600 l/(m^2.h.bar) where it became stable. Eventually the top module was put into operation. According to Kubota's experience, the top module could handle a higher flux than the bottom module, due to a better bubble distribution effect. Therefore, the flux from the two modules was not equal and depended on the feed flow. The operation of the modules is described below and the settings were optimised during the research period that followed.

The coarse bubble aeration in the membrane tank had a capacity of 90-180 Nm^3/h and could be adjusted. At the start of the DD operation, the aeration capacity was set at a constant level of 115 Nm^3/h. Unfortunately the bottom module was allowed to permeate without aeration for up to 5 hours, this caused a dramatic reduction in permeability. In the month that followed at reduced flux, the module recovered to 400 $l/(m^2.h.bar)$. The latter, as seen already in the peak testing without relaxation, was a rapid build up of sludge on the surface of the membrane, followed by cake filtration. The month of reduced flux and continuous aeration at 130 Nm^3/h loosened the cake form the surface until the entire surface was once again available for filtration - self recovery. The aeration was returned to normal at 115 Nm^3/h, and once the DD configuration was fully established and the biomass had recovered to normal running, the membranes were cleaned with the same intensive procedure as before. The cleaning procedure was again successful with a 100% recovery seen on the top module and 95% recovery on the bottom module (due to cake build up).

6.4.4 *Phase 4 – Raw influent with biological phosphorus removal*

Various ratios of flux were compared between the top and bottom modules, these ranged from 60:40 to 50:50. The latter optimisation was required due to the better scouring effect of the air on the top module as compared to the bottom module (bubbles expand as they gain height in the tank). It was concluded that 60:40 was the best ratio of flux operation, but only after the air locking in the top module was solved. The DD configuration, due to the lower static head on the top module, plus the increased air volume, caused excessive dissolved air extraction through the membrane, thus causing artificially low permeability (the air built up in the headers and plates and caused an extra pressure drop). The air locking was removed by venting the permeate piping during relaxation mode. Here, the static head of water above the module was sufficient to purge the air out of the headers. The performance of the DD configuration is shown below in Figure 44 up until 3/7/01.

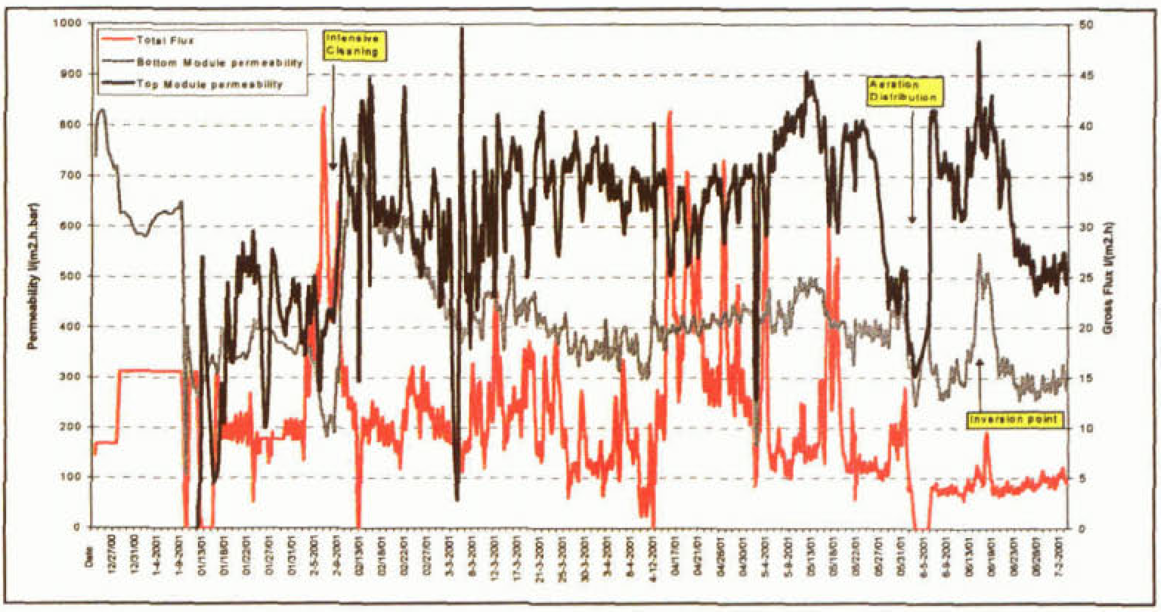


Figure 44 - Permeability and gross flux for top and bottom modules until 3/7/01

Several interesting points can be seen regarding the DD configuration. The permeability of the top module is more erratic than that of the bottom module. This was most probably caused by excessive air entrainment during permeate extraction (top sees more air). The permeability of the top module remains parallel to that of the bottom module despite the higher flux according to the 60:40 split, this suggests that the two modules are in balance.

The 50:50 ratio until 4/3/01 proved detrimental to the performance of the bottom module, thereafter, at 60:40 ratio the bottom module recovered and the top module remained relatively unaffected.

Partial blocking of the aeration system, around 25/5/01, promoting an inefficient air distribution over the two modules, caused a more dramatic impact on the top module than the bottom module. The aeration system was later fitted with a biomass-air flushing system designed in Japan. No further occurrences of preferential aeration occurred at a flushing frequency of once per day. The flushing system has also been tested at full scale on multiple cassette configurations.

The most interesting point is depicted by the inversion point. Here, the permeability increased as the flux increased. The latter was seen during a long period of extreme dry weather where the total flux barely reached 5 $l/(m^2.h)$ for the installation. At such low flux air in the plates may have caused air locking and thus restricted the passage of permeate to the headers, the latter suggested that the available filtration area was reduced or partially/temporally redundant. At the onset of a small flux increase on 17/6/01 the restrictions in the modules were purged out, thus increasing the available surface area for filtration and resulting in an increased permeability. In fact the permeability was always higher than suggested by the electrical data. The latter phenomena was included into the operators 'System Assessment', here the permeability was always carried out at a constant flux of 20 $l/(m^2.h)$, thus avoiding partial permeation. The data generated here suggested a steady decrease in permeability for both modules, see Figure 45.

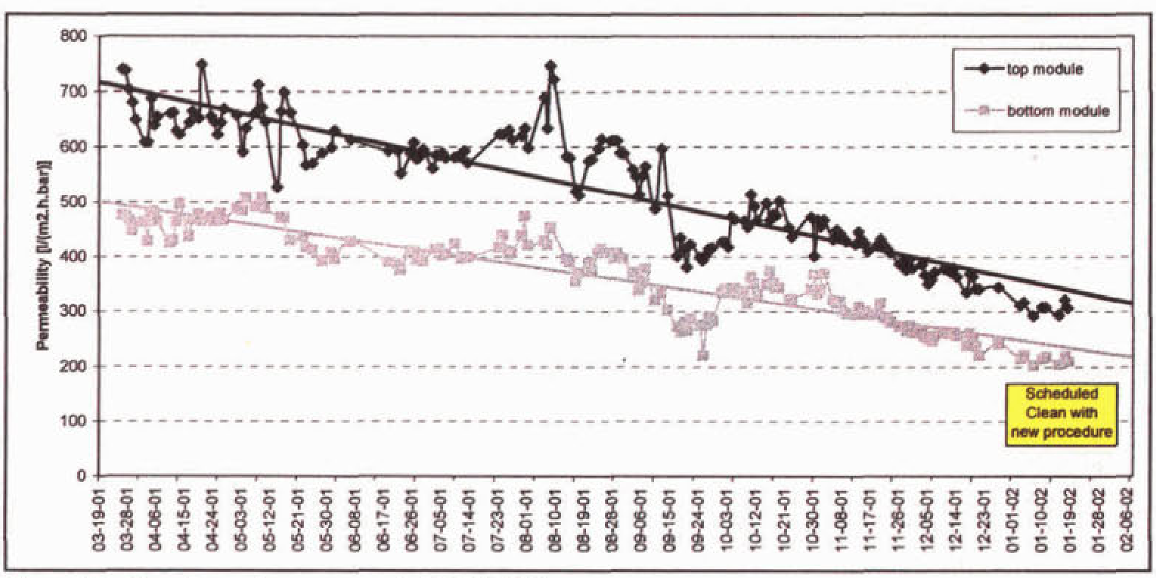


Figure 45 - Operator system assessment at 20 $l/(m^2.h)$

The above tool is a method to assess the onset of cleaning requirements as compared to the actual 'real' electrical data, however, it must be noted that at full scale with several compartments available for filtration, it would be highly unlikely that a single compartment would ran down to such low fluxes. Since the decrease in permeability is linear (Figure 45) for both modules, the cleaning frequency can be calculated. Considering a clean membrane at 700 $l/(m^2.h.bar)$ and a dirty membrane at 200-300 $l/(m^2.h.bar)$, this allows a permeability decline of 400-500 $l/(m^2.h.bar)$ in time. The gradient of the top and bottom modules were found to be 1.2 and 0.84 $l/(m^2.h.bar.d)$ respectively. The latter suggested an intensive clean of the top module every 400 days (1x per year) and every 500 days (1x per 1.4 years) for the bottom module. Overall, the intensive cleaning would be made for both modules at 1x year prior to the onset of winter.

An important aspect of membrane filtration is the effect of temperature on the permeability. The Kubota membrane proved to be relatively insensitive to temperature variations and decreases in permeability were solely associated with increased flux requirements and the onset of cake filtration processes.

The change from a simultaneous P removal process to a biological P removal process (BioP) had no detrimental effect on the membranes performance.

The DD configuration was allowed to foul further and additional cleaning optimisation in a separate research programme was scheduled for January/February 2002. This cleaning would attempt to reduce the high chemical concentrations used in the Intensive Cleaning procedure and reduce the associated biological toxic shock and chemical handling problems.

6.5 Conclusions

The biological performance of the Kubota pilot system was in agreement with theoretical design calculations. The COD removal was constant at high levels, nitrogen removal was as expected. Simultaneous phosphorus removal gave a better effluent quality than pre-precipitation, even at very low ferric dosing ratios. Biological phosphorus removal resulted in a low and relatively stable permeate concentration.

The sludge concentration in the aeration tank was kept at a level of 12 kg MLSS/m^3, to optimise the oxygen transfer in relation with the α-factor. The α-factor was according to literature values. The floc structure and the sludge characteristics were good.

A peak flux of 42 l/m^2.h net was achieved for the required operating period of more than 3 days continuous (100 hours proven) with relaxation. At reduced temperature, 10°C, the peak flux of 42 l/m^2.h net was achieved for the required operating period of 24 hours continuous.

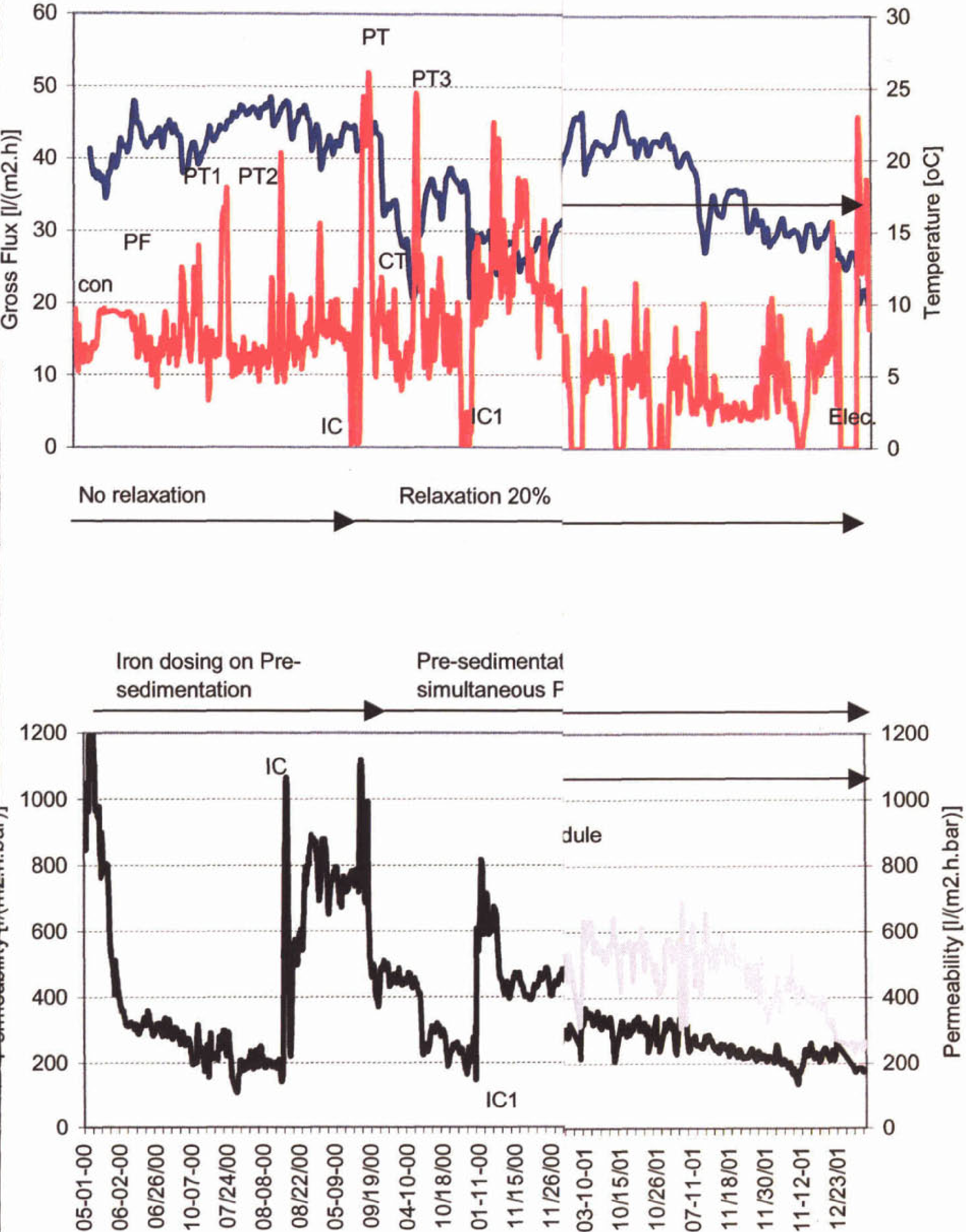
For the Dutch circumstances, with major flow variations and high peak fluxes, the membranes were found to be unreliable without aerated relaxation. Relaxation or variable relaxation was essential for full membrane availability and system performance.

The suggested cleaning of the modules at twice per year was proven to be pessimistic. The DD configuration suggested that one intensive cleaning per year would be sufficient. This cleaning should be planned in October/November prior to the onset of the winter period.

The DD configuration has proven to be reliable, with careful attention to the aeration distribution and the permeate extraction ratio. The focus of the research has shifted towards chemical cleaning and the optimisation of the cleaning procedures in regard to scale-up requirements.

KUBOTA Foldout Graphic

PF	Transition to Proportional flow (direct signal STP Beverwijk)
con	Transition to Constant flow
PT	Peak Flux test at suppliers maximum flux specification with relaxation
PT1	Failed Peak Flux test at suppliers maximum flux specification, without relaxation (dirty membranes)
PT2	Failed Peak Flux test at suppliers maximum flux specification, without relaxation (clean membranes)
PT3	Peak Flux test period at cold temperatures
CT	Artificially induced Cold Test using 22kW biomass cooler.
RB1	Rebuild 1, membrane configuration to Double Deck layout
FB	Feed pumps set to raw influent and feed filter set up modified
RB2	New feed filter placed with screenings collection.
AF	Membrane module Aeration system failure/blockage
IC	Intensive Cleaning Procedure according to KUBOTA specification
IC1	Mechanical Cleaning of module
60:40	Ratio Top module to Bottom module permeate extraction
50:50	Ratio Top module to Bottom module permeate extraction


Gross Flux [l/(m2.h)]
Temperature [oC]
PT
PT3
PT1
PT2
PF
con
CT
IC
IC1
Elec.
No relaxation
Relaxation 20%
Iron dosing on Pre-sedimentation
Pre-sedimentat
simultaneous F
dule
IC
IC1
Permeability [l/(m2.h.bar)]
05-01-00
06-02-00
06/26/00
10-07-00
07/24/00
08-08-00
08/22/00
05-09-00
09/19/00
04-10-00
10/18/00
01-11-00
11/15/00
11/26/00
03-10-01
10/15/01
10/26/01
07-11-01
11/18/01
11/30/01
11-12-01
12/23/01

Kubota

7 MITSUBISHI

7.1 Introduction

Figure 46 - Photograph Mitsubishi pilot plant

In this chapter the performance of the Mitsubishi pilot plant is described. In section 7.2 the system configuration and design figures are given. In section 7.3 and 7.4 respectively the biological and membrane performance is presented. Finally in section 7.5 the conclusions are summarised.

The Mitsubishi pilot plant was in operation since week 17 of the year 2000. In Table 12 the main activities for the Mitsubishi pilot plant research are summarised.

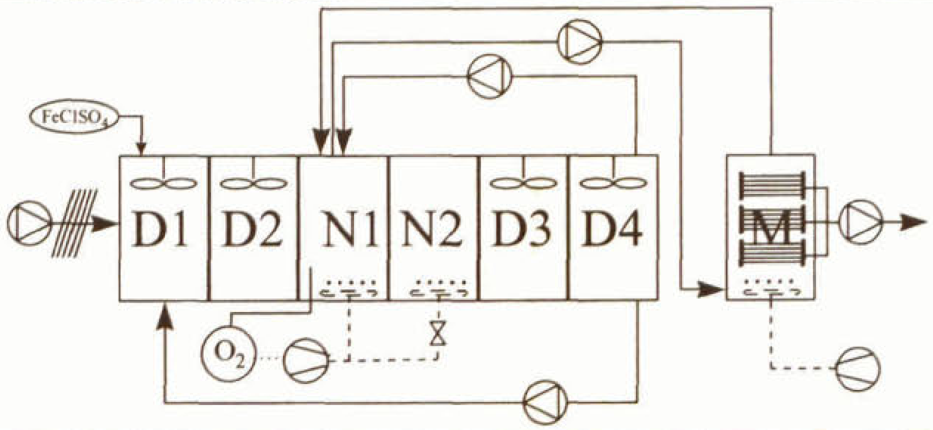
Table 12 - Research activities

phase 1	**primary clarification with pre-precipitation** 01/05/00 - 03/10/00 week 17/2000 - 40/2000
week 17	testing installation and seeding with sludge
week 18-23	constant flow
week 24-29	proportional flow
week 30-33	peak testing, recovery and cleaning
week 34-42	proportional flow, incl. chemical cleaning
phase 2	**primary clarification with simultaneous precipitation** 04/10/00 - 10/01/01 week 40/2000 - 02/2001
week 42-44	constant flow
week 44-46	system out of operation, adjustments made on hard- and soft-ware
week 47-52	limited proportional flow and constant flow introduction back-pulse procedure and extended back-pulse procedure
week 01-02	proportional flow
week 02	introduction of day / night simulation
phase 3	**raw influent with simultaneous precipitation** 11/01/01 - 11/05/01 week 02/2001 - 19/2001
week 02-08	proportional flow, optimisation operational settings
week 08	new module installed
week 09-19	constant flow, optimisation maintenance cleaning, peak tests
phase 4	**raw influent with biological phosphorus removal** 12/05/01 - 31/12/01 week 20/2001 - 52/2001
week 20-52	constant flow, optimisation (enhanced) maintenance cleaning, peak tests

7.2 System configuration

The configuration and design figures are presented in Table 13. The configuration and some of the process settings have been changed during the research period. The main changes are described in the table.

Table 13 - Design figures of the pilot plant



process part	parameter	unit	value
influent pump	capacity	m^3/h	10
	RWF design flow	m^3/h	6.4 → 9.6 [1]
	design flow	m^3/d	32 → 48 [1]
influent screen	type	-	parabolic sieve
	slot size	mm	0.75
aeration tank	total volume (and depth)	m^3	34.2
	- anoxic volume (D1/2/3/4)	m^3 (m)	4 x 3.9 (1.70)
	- oxic compartment (N2)	m^3 (m)	2 x 3.9 (1.70)
	- membrane tank (M)	m^3 (m)	10.8 (5.0)
aeration biology	compressor capacity total	Nm^3/h	160
	capacity N1 / N2	Nm^3/h	80 / 80
ferric dosing [2]	type	-	$FeClSO_4$
	ferric content	%	12.3
	capacity (at Me/P = 0.8)	ml/h	160
membrane filtration	number of modules	-	3
	total surface	m^2	315
	max. net flux (at RWF)	$l/(m^2.h)$	20.3 → 30.6 [1]
aeration membranes	compressor capacity	Nm^3/h	75 - 120
	specific capacity	$Nm^3/(m^2.h)$	0.24 - 0.38
re-circulation flows	sludge N1 → M	m^3/h	20-40
	internal D4 → N1	m^3/h	17-25
	nitrate D4 → D1	m^3/h	17-25

1: the hydraulic design load is increased during phase 1 of the pilot research

2: the ferric dosing is stopped at 11 May 2001

Aeration of sludge in compartments N1 and N2 was achieved with one compressor. The aeration in compartment N2 could be shut down manually if required. The aeration control was based on the oxygen measurement in the first aeration compartment (N1). In practise the aeration was in operation for both compartments almost continuously. The coarse bubble aeration for the membrane module was mostly fixed at a constant level, but could be linked to the permeate flow.

The original design flow of the Mitsubishi aeration tank was set at 34 m^3/d. From week 21 it was decided to increase the design flow to 48 m^3/d, to improve the sludge quality.

On 3rd October the chemical dosing for the pre-precipitation process was stopped. The simultaneous precipitation was started on 15th November due to technical reasons (rebuild). The dosing was stopped on 21st December after the membrane chemical cleaning. To summarise, phase 2 (simultaneous precipitation period) only lasted from 15th November 2000 until 21st December 2000. In phase 3 the simultaneous precipitation was in operation continuously, until the first week of May 2001 when the ferric dosing was stopped. In phase 4, phosphorus removal was based on biological uptake only. No additional measures were taken to improve the biological phosphorus uptake.

7.3 Biological performance

7.3.1 Process conditions

The main process conditions are presented graphically in Figure 47 to Figure 49. The average process conditions during the phases of the pilot research, including periods in which the installation was out of operation, are presented in Table 14.

During the first weeks after start-up the pilot was fed with a constant flow. Since week 25 the feed to the pilot was related proportionally to the feed of the full-scale Beverwijk treatment plant. The average influent flow during the research period varied strongly due to rain weather conditions. The average flow during the whole research period (incl. rain) amounted 40 m^3/d, which is lower than the design flow. This was caused by technical problems, rebuilding of the pilot, and software changes, as described in section 7.4. The Mitsubishi pilot plant was located outside. During the winter the feed filter froze or blocked under low flow loading this reduced the pilots feed supply and in turn also restricted the pilot's permeate production.

Table 14 - Process conditions of the pilot plant

parameter	unit		phase 1	phase 2	phase 3	phase 4
influent flow	m^3/d		38	44	40	55
process temperature	°C	average	20	14	12	18
		range.	12 - 31	5 - 27	5 - 21	7 - 31
pH	-		7.4	7.3	7.3	7.4
biological loading	kg COD/(kg MLSS.d)		0.043	0.045	0.059	0.084
sludge concentration	kg MLSS/m^3		8.9	9.9	10.8	11.6
organic part	%		62	62	63	65
sludge production	kg MLSS/d		3.5	5.4	11.8	15.5
sludge age	d		87	62	31	26
ferric dosing AT	l/d		0	0.9 [1]	2.1	0
ferric dosing ratio	mol Fe/mol P		0	0.30	0.58	0

1: simultaneous precipitation from 15/11/2000 till 21/12/2000

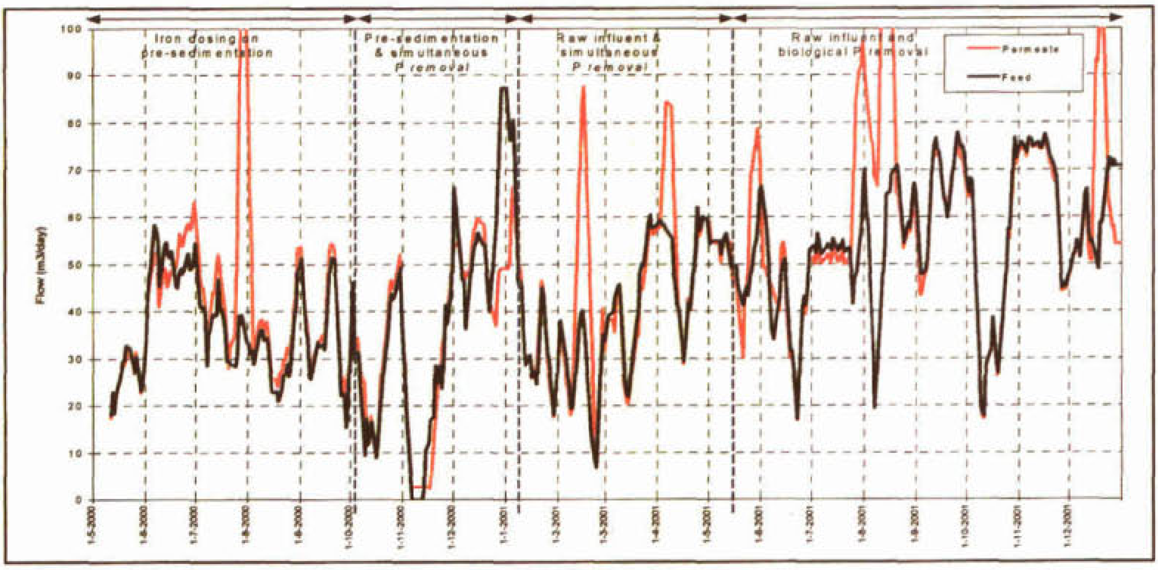


Figure 47 - Influent and permeate flow of the Mitsubishi pilot plant

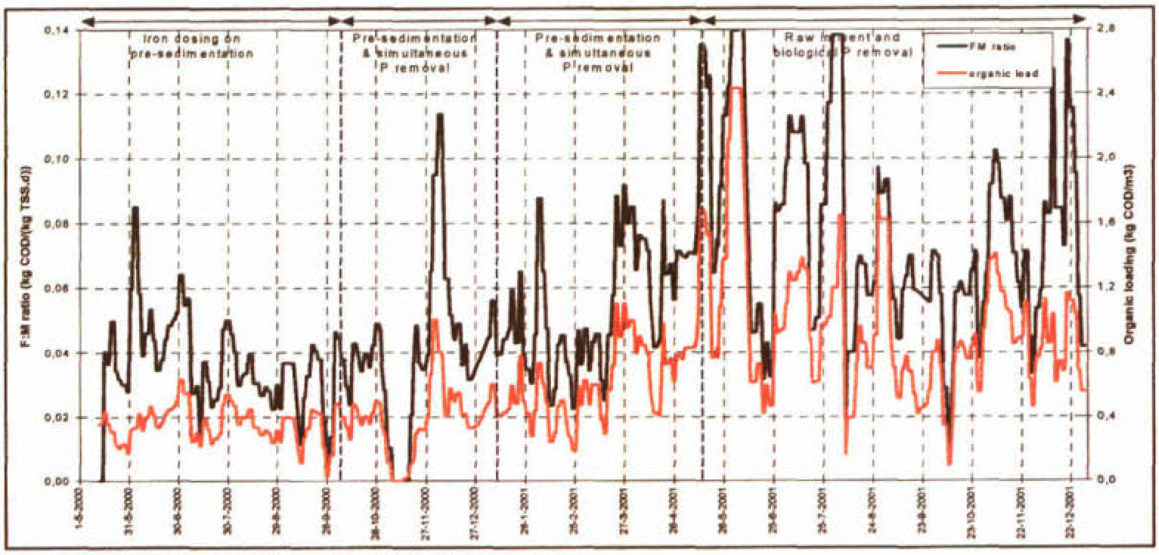


Figure 48 - Biological loading of the Mitsubishi pilot plant

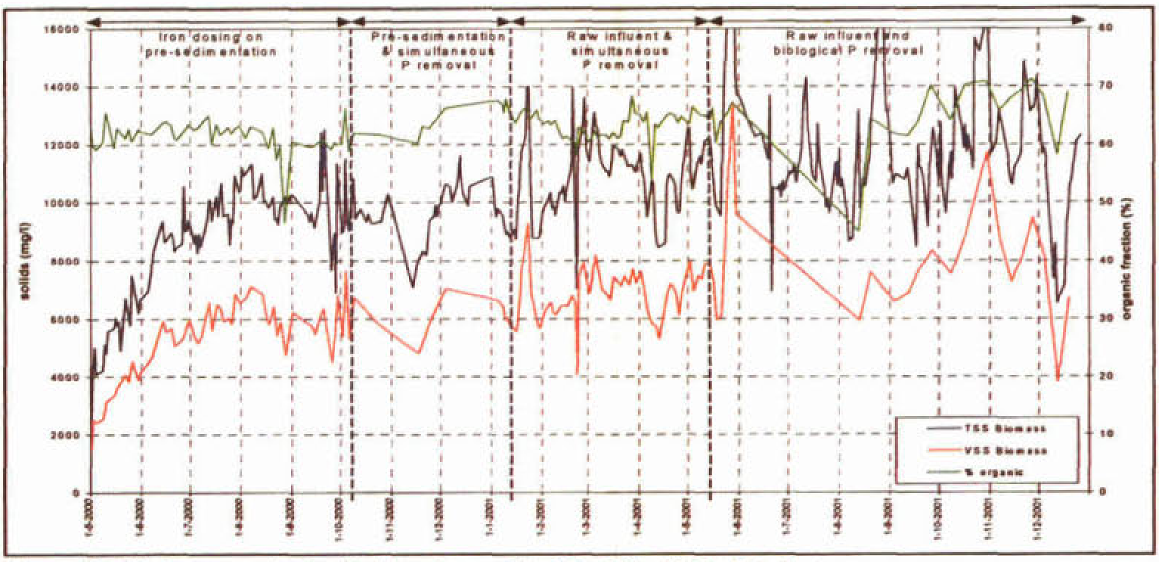


Figure 49 - Sludge concentration (total and organic) of the Mitsubishi pilot plant

The set-point for the sludge concentration was 10 kg MLSS/m^3. The process conditions during the research periods differed significantly as shown in Table 15. Besides the influent flow variations, the process temperature also changed. The ferric dosing ratio during phase 2 of the simultaneous precipitation period was relatively low. The average biological loading was stable.

7.3.2 *Results*

The main biological parameters are presented graphically in Figure 50 to Figure 52. In Table 15 the average feed and permeate concentrations are presented.

Table 15 - Feed and permeate concentrations

parameter		unit	phase 1	phase 2	phase 3	phase 4
COD	feed	mg/l	364	325	548	605
	permeate	mg/l	31	25	30	34
	efficiency	%	91	92	95	94
N_{kj}	feed	mg/l	65	44	57	59
	permeate	mg/l	2.6	1.1	2.1	4.2
NO_3-N	permeate	mg/l	9.2	8.0	9.0	4.4
N_{total}	permeate	mg/l	11.9	9.1	11.1	8.6
	efficiency	%	82	79	81	85
P_{total}	feed	mg/l	7.9	6.6	9.3	12.1
	permeate	mg/l	2.1	1.9	0.7	1.2
	efficiency	%	74	71	92	90

COD removal

The COD-permeate concentration was relatively low due to the fact that the non-soluble fraction was completely removed by the membrane filtration. The COD permeate concentration in phase 2 measured only 25 mg/l, which was lower than the concentration in phase 1, 3 and 4. This was due to the dilution effect caused by long periods of rain weather.

Nitrogen removal

In phase 1 the nitrate concentration in the permeate was relatively high with an average concentration above 10 mg/l due to the pre-precipitation process. The COD/N ratio of the feed was 6 which was relatively low. In phase 2 this ratio increased to 8 and consequently nitrogen removal improved. In this respect also the dilution effect must be taken into account.
The nitrogen removal in phase 3, without primary clarification, performed worse in the beginning of this period. This is mainly a consequence of the technical failures of the system in that period. Since the end of March 2001 the nitrogen removal performed well at a permeate concentration < 10 mg N_{total}/l.

Phosphorus removal

In phase 1 it was not possible to achieve a phosphorus permeate concentration < 1 mg P_{total}/l at a Me/P dosing ratio of 1.2. Both the removal efficiency of the primary clarifiers and the sludge production were too low. For phase 3 the same problems as with the nitrogen removal occurred at the beginning of this period. With a relatively high chemical dosing (compared to the other pilot systems), the permeate concentration remained > 1 mg P_{total}/l. Since March 2001 the phosphorus removal has improved significantly to a permeate concentration < 1 mg P_{total}/l. In phase 4, without any additional measures for phosphorus removal, the removal efficiency is 90%. This indicates that enhanced biological phosphorus removal is occurring.

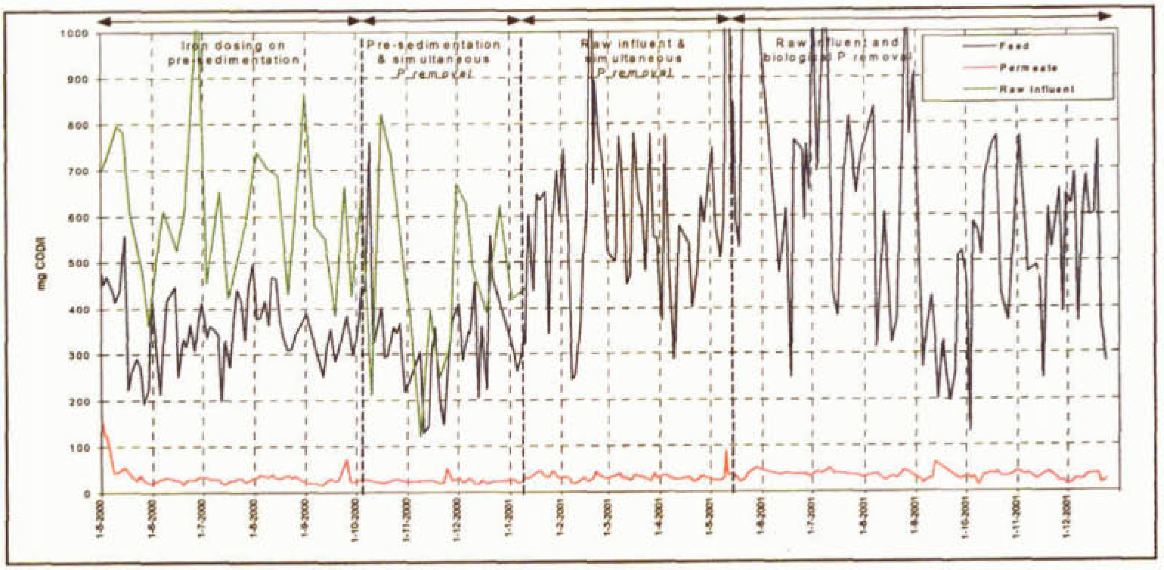


Figure 50 - COD feed and permeate concentration of the Mitsubishi pilot plant

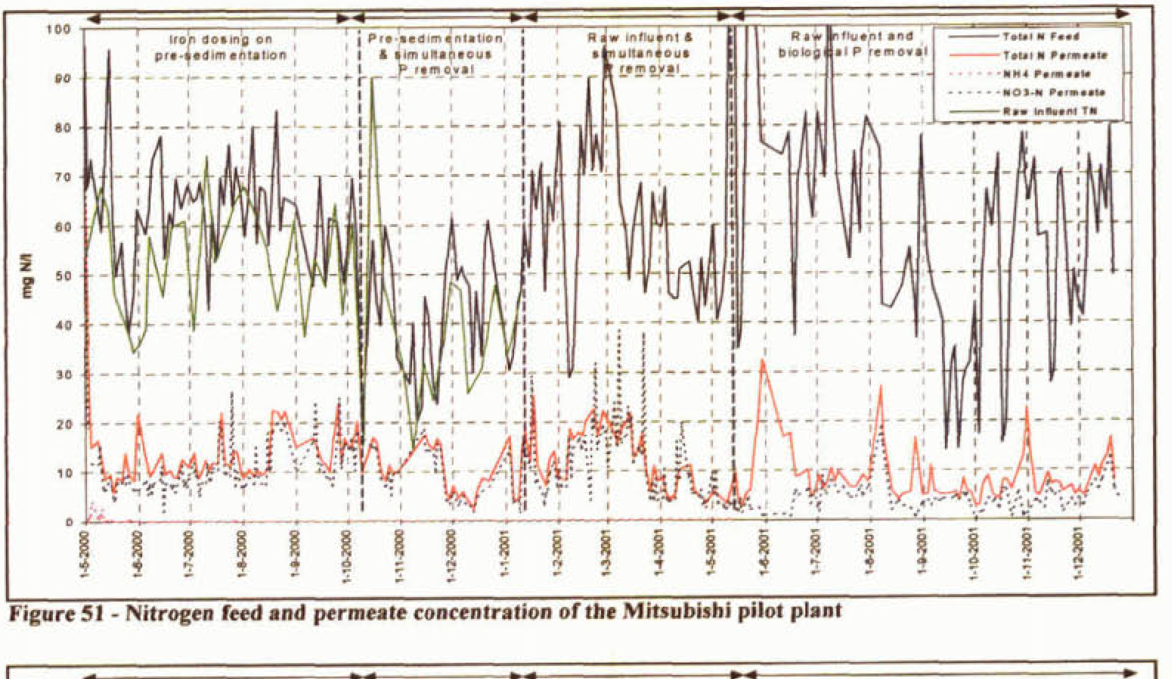


Figure 51 - Nitrogen feed and permeate concentration of the Mitsubishi pilot plant

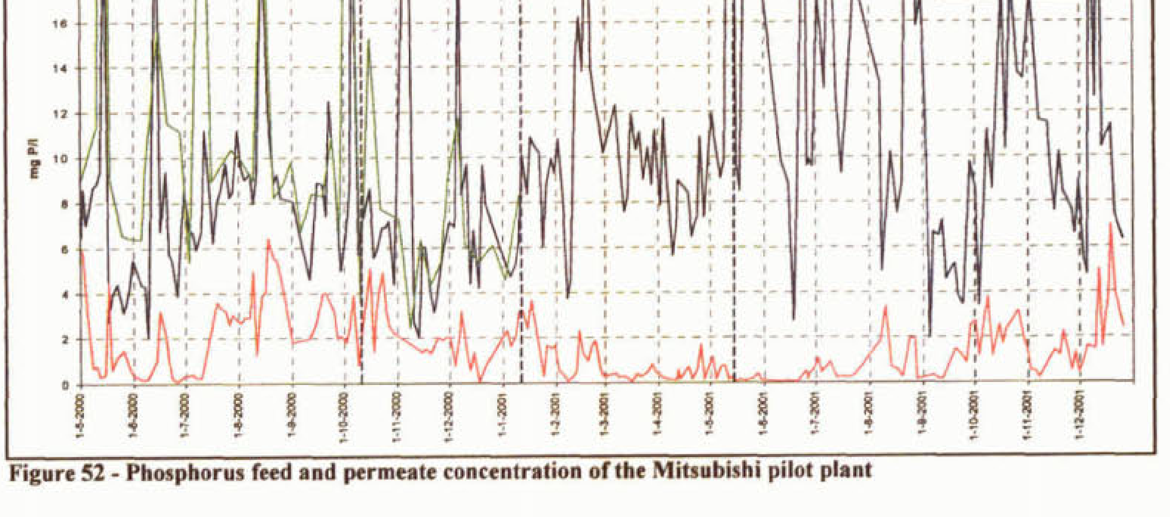


Figure 52 - Phosphorus feed and permeate concentration of the Mitsubishi pilot plant

7.3.3 *Sludge characteristics*

The main sludge characteristics are presented in Table 16. The DSVI was relatively stable at a level of 100-140 ml/g. Both the CST and the Y-flow can be related to the sludge concentration. The viscosity of the MBR sludge varied from 6-13 mPa.s, which was relatively low for the applied sludge concentrations. The sludge viscosity appeared to be related to the sludge concentration. The first week of November 2000 aeration tank mixers were changed from high turbulent pumps to low turbulent mixers. As a consequence the sludge characteristics improved.

Table 16 - Sludge characteristics

	unit	phase 1	phase 2	phase 3	phase 4
Sludge characteristics					
DSVI	ml/g	130	140	120	100
CST	s	130	130	120	80
Y-flow	s	320	690	n.a.	130
Viscosity					
viscosity value	mPa.s	6.9	10.3	9.4	7.5
shear rate	1/s	100	130	130	90
α-factor					
surface aeration	-	0.43	0.50	0.53	n.a.
bubble aeration	-	0.38	n.a.	0.48	n.a.
Gravity thickening					
settling velocity	cm/h	3.9	4.5	3.4	n.a.
maximum concentration	%	2.0	2.4	2.6	n.a.
Mechanical thickening					
MLSS at 3900 rpm / 10 min	%	8.0	9.3	6.7	n.a.
MLSS at 1000 rpm / 3 min.	%	2.4	2.9	2.7	n.a.

The α-factor with bubble aeration in phase 1 varied between 0.3 and 0.5 at a sludge concentration of 5 to 12 kg MLSS/m^3. The average α-factor at 10 kg MLSS/m^3 was approximately 0.4. With a surface aerator 15% higher values were achieved. In phase 3 the average α-factor with surface aeration increased to 0.5-0.6, probably due to system improvements resulting in a lower energy input. The average settling velocity at the beginning of the gravity thickening test amounted to 2 - 6 cm/h, dependent on the sludge concentration. The maximum concentration after 24 hours was lower than 3%, indicating that gravity thickening was not efficient. The maximum sludge concentration with mechanical thickening amounted to 6 to 9%.

Since the start-up of the MBR system the flocs decreased to a small floc size with an open structure. Since the start-up monocultures were present. The size of these monocultures increased a little, and sometimes the monocultures were larger than the flocs present. In the end of phase 2 their number decreased dramatically. Normally, some crawling ciliates (Aspidisca) and free-swimming ciliates (Euplotus) were observed. Their occurrence indicated a well-aerated system with minor disturbances. Especially during the start-up some filamentous organisms occurred, but the number steadily dropped below 1 (according to Eikelboom).

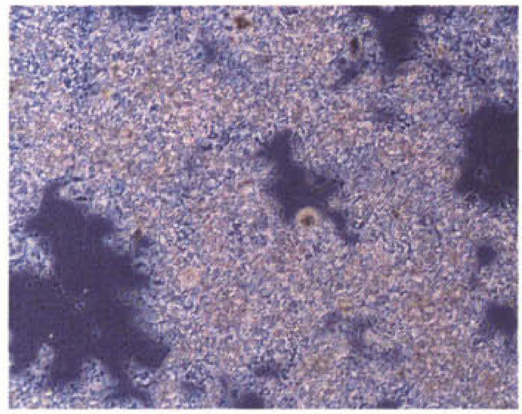


Figure 53 - Microscopic view of the activated sludge

7.4 Membrane performance

The results of the whole research period are presented graphically in the fold out page and further explanation is given in the text. The data presented is damped to aid readability and conclusions are made regarding the membrane performance and the relevant operating & design criteria.

It was expected that the maximum achievable flux would be 32.5 $l/m^2.h$ for a period of 3 days continuous (20.3 $l/m^2.h$ net originally specification). The cleaning was estimated at twice per year and the cleaning mode was set according to Mitsubishi's specifications of a maximum of 5,000 mg/l NaOCl followed by an acidic solution. Relaxation was used between process modes and no other recovery steps or modes were anticipated.

Periodically the Mitsubishi membranes have relaxation. During the research period the ratio between net and gross flux was 79 to 87%. All data for flux in the graphic is gross flux.

7.4.1 Phase 1 - Primary settled wastewater and pre-precipitation

Start up and Definitions

To achieve a stable biological start up the system was run at the 24 hour average daily flow rate of 32 m^3/day. The latter was later increased to 45 m^3/day to increase the biological loading and increase sludge growth. The membranes determined the flow through the system and the feed pump followed via a set level in the bioreactor.

In June 2000 the pilot showed stable enough operation with an acceptable sludge growth to proceed to a proportional flow control at a ratio of 833:1 to that of Beverwijk WWTP. The permeability of the Mitsubishi pilot plant steadily decreased from 500 $l/(m^2.h.bar)$ to 50 $l/(m^2.h.bar)$ in a 5 month duration. Limited recovery of permeability was seen during periods of DWF, and changes in the process mode conditions had little effect on the permeability decay.

In Figure 54 a graphical representation of the first phase is presented.

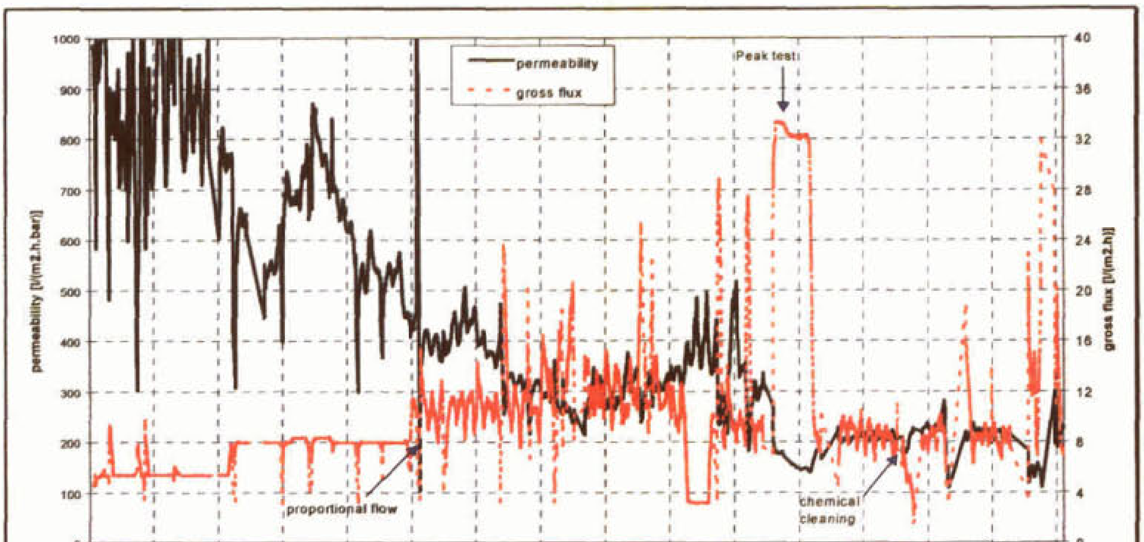


Figure 54 - Permeability and gross flux of the Mitsubishi pilot plant

Some recovery was seen during longer periods of system shutdown, and as a result this mode was later incorporated into the systems software (night relaxation) during phase 2 with some success. One concern for the capillary type fibre membrane used by Mitsubishi was the potential to sludge up with thick, non-removable sludge. This raised serious questions about the systems integrity. The latter was also a possible (theoretical) explanation of why the permeability only declined. On two separate occasions an electrical fault in the instrumentation caused the system to de-water, and as a result the sludge was thickened to 25 kg MLSS/m^3. Under these conditions the membranes were extremely vulnerable to sludging, but after simple re-aeration of the filtration tank and the establishment of a workable dissolved oxygen concentration, the system was set back to process with no apparent permeability loss. It must be stated that the Mitsubishi module displayed negligible sludging during the pilot trial, despite the original concerns. This was put down to the pre-screening used and the turbulence generated in the filtration zone.

Peak tests

A 95 hour peak test was executed. During the peak test the maximum hydraulic loading of the pilot was 10.2 m^3/h. This corresponded to a net flux of 28.1 l/(m^2.h) and a gross flux of 32.5 l/(m^2.h). The process temperature was 20-21°C. The permeability and the flux before, during and after the peak tests are presented in Figure 55.

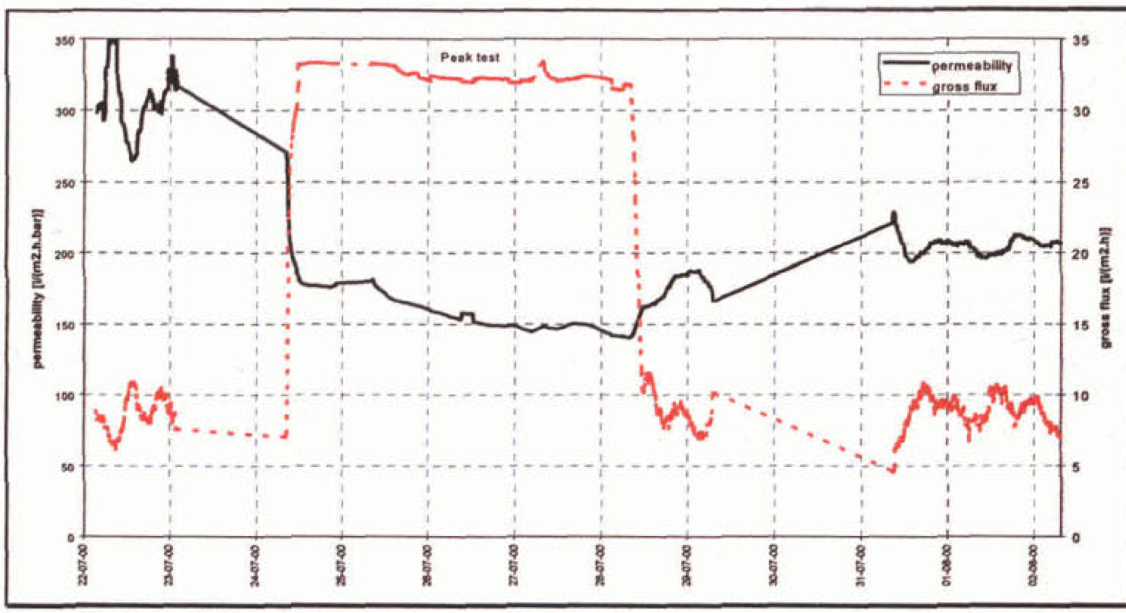


Figure 55 - Permeability and gross flux during peak testing

Before the peak test started, the permeability varied from 270-350 l/(m^2.h.bar), depending on DWF flow variations. On starting the 95 hour peak test, the permeability rapidly decreased and subsequently slowly stabilised at 140-150 l/(m^2.h.bar). Stopping the peak test caused the permeability to recover steadily to about 175-200 l/(m^2.h.bar) in a period of 24 hours. In the following days the permeability remained stable at that level, much lower than the permeability level before the peak test. This caused some concern as a membrane that can not recover adequately on it's own accord would require intensive cleaning on a regular basis. Despite the latter the membrane was put back to DWF and the permeability returned to it's original permeability decline rate.

From the peak test trial it was concluded that the Mitsubishi membranes could handle a peak flow of 28.1 l/(m^2.h) net at 20°C but with limited stability. The relative bad sludge characteristics in this phase (due to high turbulent mixing) probably had a negative influence on the membrane performance. Further peak tests were not carried out and more effort was spent on the membrane recovery.

Cleaning procedure

On 9/8/00 the first chemical CIP cleaning with NaOCl, NaOH and HCl was executed in-situ in biomass. This cleaning was not totally according to the official Mitsubishi CIP procedure, and the permeate pump was too small to facilitate an effective back pulse, as the pilot plant was not designed for it. The effect of the cleaning was limited as the permeability remained at 200 l/(m^2.h.bar). A second chemical cleaning on 22/08/00 was carried out in a period of electrical/software problems and this cleaning also had limited effect on the membrane performance, as the permeability increased only slightly. During the second cleaning a 5,000 ppm NaOCl solution was pumped through the membranes in a closed loop at 40°C, also in-situ in biomass. The second clean was according to Mitsubishi procedures for in-situ cleaning.

7.4.2 *Phase 2 - Primary settled wastewater and simultaneous precipitation*

The system was cleaned using the intensive cleaning procedure during 21 and 22/09/00, this involved tank drainage and internal and external membrane cleaning with NaOCl and citric acid. This cleaning returned permeability to 250 l/(m^2.h.bar), but within one month dropped to 50 l/(m^2.h.bar). At such low permeability the only course of action was to return the system to continuous permeate flow. At this stage it was decided to modify the system. Technical adjustments were made as well as changes in the software. This rebuilding of the pilot was carried out between 01/11/00 and 15/11/00, and in this period the system was fully tested and the membrane permeability returned to a respectable clean process starting point of 400 l/(m^2.h.bar). The intensive cleaning procedure used was in accordance with Mitsubishi but used a longer contact time than the cleanings procedures carried out up to that time. The intensive cleaning procedure was proven and the fouling was found to be predominantly organic, in the form of surface fouling, covering an inorganic iron based fouling on the membrane surface. The results of phase 2 are shown in Figure 56. The simultaneous precipitation in the bioreactor was started on 15/11/00.

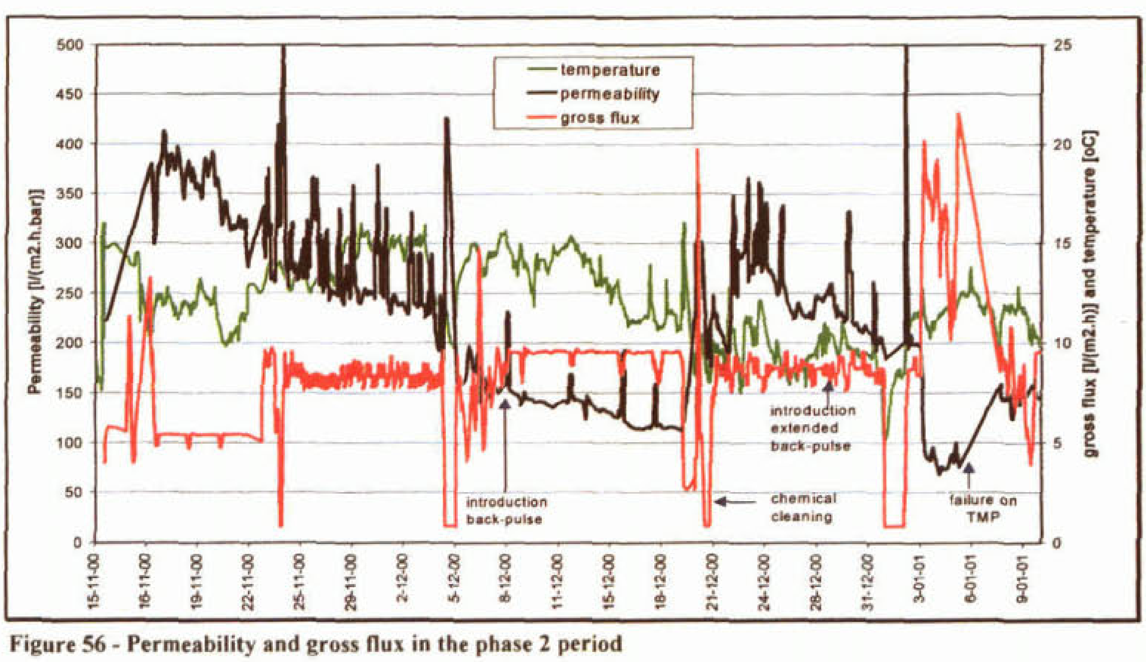


Figure 56 - Permeability and gross flux in the phase 2 period

After adjusting the pilot plant, it was ran at a constant gross flux of 5.4 $l/(m^2.h)$, which suggested a net flux of approximately 4.6 $l/(m^2.h)$. In this short period of constant low flow, the permeability decreased from around 380 to around 320 $l/(m^2.h.bar)$.

During the subsequent two weeks the constant gross and net flux were increased to 9.5 $l/(m^2.h)$ and 8.2 $l/(m^2.h)$. In this period the permeability decreased further from around 320 to around 220 $l/(m^2.h.bar)$. Despite the introduction of a back flush mode, the permeability decreased rapidly even at constant low flux. Many forms of optimisation were tested in the above phase 2 period. Unfortunately, the introduced back pulse procedures only had a limited effect. The same conclusion was drawn from the introduction of extended back-pulse procedures. Finally a relaxation period of 4 hours from midnight until 4.00 a.m. was introduced and tested in phase 3.

7.4.3 *Phase 3 - Raw influent and simultaneous precipitation*

The system feed was changed on 10/1/01 and fully operational on 11/1/01. Two additional cleaning procedures were carried out. On inspection of the module it was found that the membrane fibres were weakened. Mitsubishi was also present to witness the damage to the fibres and samples were taken back to Japan for investigation. The fibres had been chemically weakened, caused by over contact with NaOCl, possibly from the procedures carried out in September 2000. It was decided to replace the old module and to test the membranes at a higher flux in the last two weeks. Prior to installation of the new module the old one would be cleaned according to a standard Mitsubishi intensive cleaning procedure but at a lower NaOCl concentration of 1,000 mg/l.

The results were surprising. At higher fluxes of 18 and eventually 21 $l/(m^2.h)$ gross (15 and 18 $l/(m^2.h)$ net respectively) the membrane showed a steady decline in permeability - the same trend from on the beginning of the research. It was clear that this membrane could be operated in a continuous permeation mode at higher fluxes than previously tested. The decline could be calculated as it was almost linear, and the time interval between in-situ or intensive cleaning could be easily determined. The latter was a new philosophy for the operation of the Mitsubishi membrane in accordance with the Dutch situation. After two weeks of high continuous flux the membranes were cleaned using a standard Intensive Cleaning procedure but at 1,000 mg/l NaOCl - the cleaning was successful and the process permeability returned to 350 $l/(m^2.h.bar)$.

On 22/2/01 a new module was installed and run in at constant low flux. The back-flush and night relaxation procedures were also incorporated. Preliminary results appeared to be positive. and the goal was to push for higher fluxes with the incorporation of maintenance cleaning (MC) rather than intensive cleaning (IC). Figure 57 displays the progress of the MC procedures and the membrane performance until the beginning of July 2001.

The new module was not directly subjected to routine MC procedures but ran according to the Mitsubishi standard criteria of low, but continuous flux. The first 1.5 months yielded the same result as at the beginning of the pilot study - a continuous decline in permeability. MC in air procedures were started on 27/3/01 but not as a routine but more as a semi-intensive clean and were actually started on an already fouling membrane.

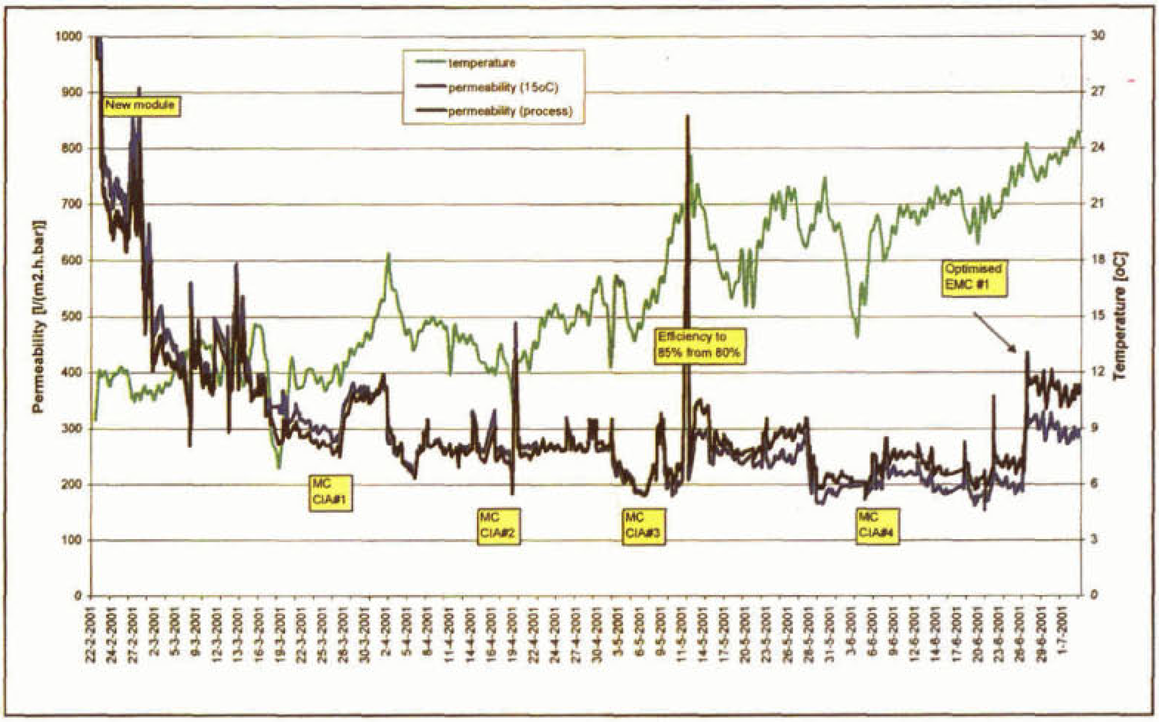


Figure 57 - Permeability and temperature and MC progression for new module

7.4.4 *Phase 4 – Raw influent with biological phosphorus removal*

After 4 MC procedures were carried out in a period of 3 months it was clear that the procedure had an advantage but only minor, as the permeability decline remained prominent. It was established that the MC back flush flux was of critical importance and this gave rise to the EMC procedure - enhanced maintenance cleaning in air. The general MC in air procedure remained intact but the back flush flux was forced to 20 l/(m^2.h). The latter guaranteed that all fibres received an adequate pressure differential and back flush volume.

The results of the first EMC were most positive, and subsequent EMC procedures on a routine basis have helped maintain a high and stable permeability of around 400 l/(m^2.h.bar). It is believed that at the lower back flushing fluxes the capillary fibres were not equally flushed, and as a result preferential cleaning took place (usually on the cleanest fibres). The latter would facilitate the linear permeability decline already seen. The use of high back flushing fluxes eliminated the preferential cleaning and thus returned the total membrane surface to full availability. The latter procedure is not uncommon on capillary drinking water membranes.

The period after 1 July 2001 saw a further development of the EMC. What became evident was the necessity to maintain a routine EMC rather than ad-hock cleaning. The latter was emphasised prior to the onset of colder weather where a number of weekly cleans were dropped from the schedule, the subsequent EMC procedures thereafter lacked in efficiency and the system became more temperature / fouling sensitive. This was seen in a reduction of the base line permeability of 200-300 l/(m^2.h.bar), see Figure 58.

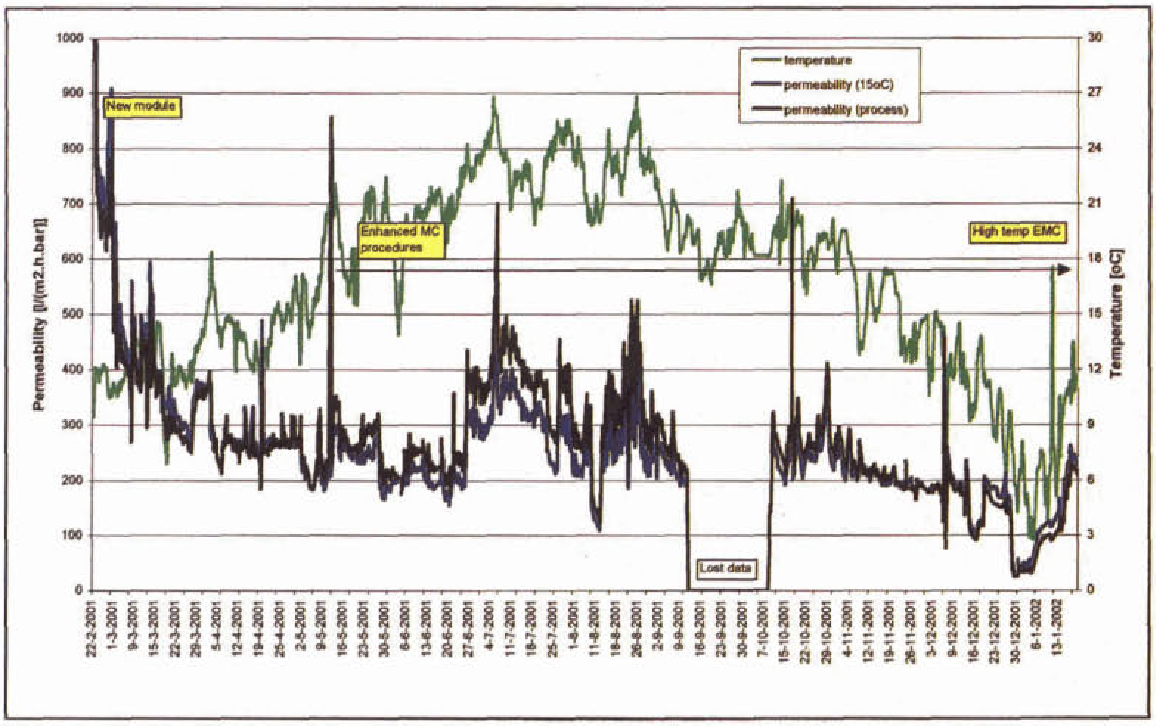


Figure 58 - Permeability and temperature for progression of EMC procedures

As seen in the foldout graphic net fluxes of 20.8 $l/(m^2.h)$ have been successfully achieved for up to 5 days. A net peak flux of 31.25 $l/(m^2.h)$ was tested for 5 days and displayed a rapid drop in permeability, this however was directly accompanied by a full recovery promoted by the EMC procedures. As already discussed the Mitsubishi membrane displays strong temperature dependability, above 15°C the membrane performance was reliable, under 10°C the permeability was strongly dependent on the process temperature and the EMC procedure was unable to maintain a stable operating condition. It should be noted that the pilot (located outside) saw an extremely low temperature of 2.7°C during December/January, and under the 5°C for 1 week, here the permeability reached its lowest base line of only 30 $l/(m^2.h.bar)$ and the membrane flux had to be reduced. A small modification to the EMC amounting to a simple cleaning temperature increase from 15°C to 30°C facilitated a full recovery to the cold weather permeability base line of around 250 $l/(m^2.h.bar)$. The potential to return to proportional flow is now under investigation.

Figure 57 and Figure 58 show a distinct relationship between permeability and process temperature. This was partly related to the viscosity, but mainly to the physical characteristic of the membrane itself. The predominate fouling characteristic remains cake filtration but the permeability trend follows the temperature variation.

7.5 Conclusions

The biological performance was acceptable but less than expected according to theoretical design calculations. This was due to a combination of technical malfunctioning of the software programme, low and irregular feeding conditions and the sub-optimal biological process control.

The sludge concentration in the aeration tank was kept at a level of around 10 kg MLSS/m^3, to avoid problems in the membrane zone and to optimise the oxygen transfer in relation to the α-factor. The α-factor had increased during the research, mainly due to system improvements resulting in a lower shear stress for the sludge. In phase 3 the α-factor was as expected from literature.

A peak flux of 28.1 l/(m^2.h) net was achieved for the required operating period of more than 3 days continuous (95 hours proven) at a temperature above 20°C. The membrane was unable to satisfactorily process successive large peaks during successive RWF conditions. Continuous flow mode seemed to be the best mode of operation for the module.

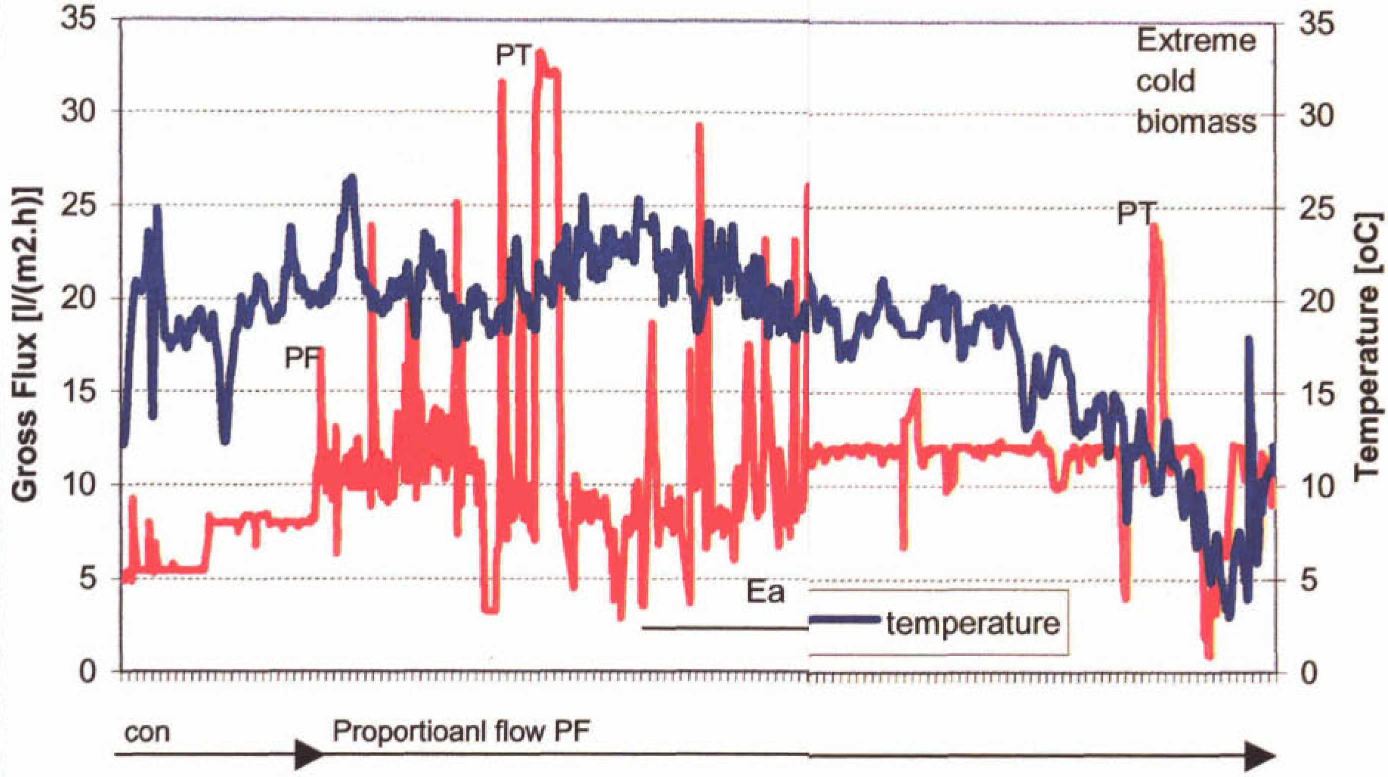
The maximum continuous design flux above temperatures of 10°C was established at 15 to 18 l/(m^2.h) net. At these fluxes a cleaning in the form of in-situ or intensive could be planned every 2-4 months of operation. The required cleaning frequency of the module is 4 to 6 times per year, at lower chemical concentrations than expected.

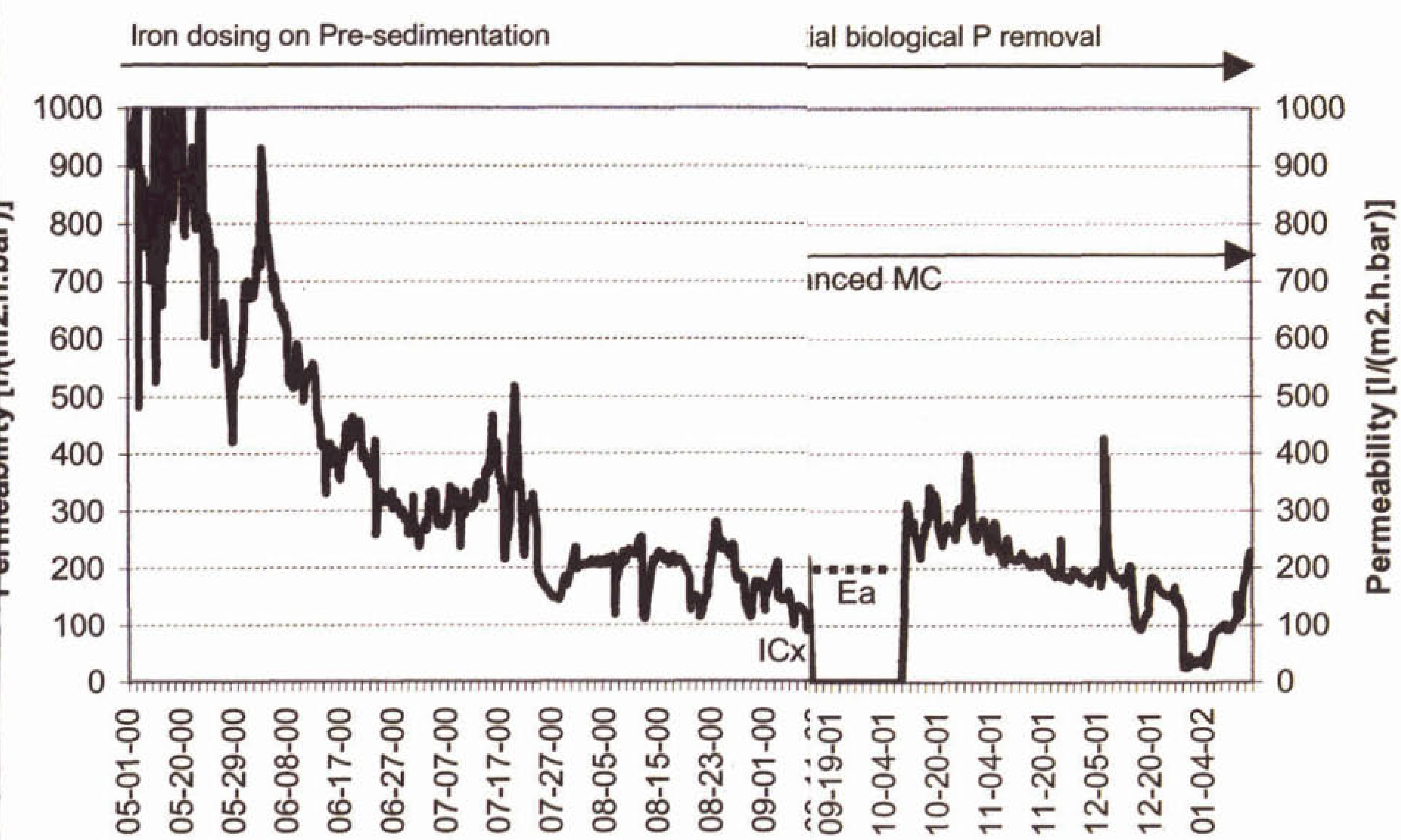
A new module was installed and run in at constant low flux. The new module has been pushed for higher fluxes with the incorporation of maintenance cleaning (MC) rather than intensive cleaning (IC) and later by enhanced maintenance cleaning (EMC). Results appeared positive and the EMC proved to be a successful routine procedure. Back flushing at 20 l/(m2.h) also proved to be advantageous to the membrane performance in combination with the EMC. The expected cleaning frequency based on the preliminary results with the new module is 1 to 2 times per year IC in combination with a weekly or 2-weekly EMC.

The incorporation of the EMC has lead to repeated peak tests of 20.8 l/(m^2.h) for up to 5 days. A net maximum peak flux of 31.25 l/(m^2.h) was tested for 5 days and displayed a rapid drop in permeability, this however was directly accompanied by a full recovery promoted by the EMC procedures. The EMC procedure is being further developed and will be optimised in the near future for proportional flow, the procedure is flexible and can be adapted to difficult process conditions, such as extreme low process temperatures.

Mitsubishi Fold-out Graphic

PF	Transition to Proportional flow (direct signal STP Beverwijk)
con	Transition to Constant flow
PT	Peak Flux test period at suppliers maximum flux specification
RB1	Rebuild 1, software, electrical and mechanical systems renewed
RB2	System configuration optimised, MC software, relaxation etc.
RB3	New module Installed
Ea	Electrical/software Alarms, system tripped automatically due to circumstances
MC	Standard Maintenance Cleaning Procedure according to Mitsubishi specification
IC	Intensive Cleaning Procedure according to Mitsubishi specification
ICx	Intensive Cleaning Procedure incorrectly executed
OptC	Optimisation of in-situ cleaning (Chemicals, Back-flush, extended BF, relaxation)
Test	Test period during RB1


Gross Flux [l/(m2.h)]
Temperature [oC]
Extreme
cold
biomass
PT
PF
PT
Ea
temperature
con
Proportioanl flow PF


Iron dosing on Pre-sedimentation
ial biological P removal
Permeability [l/(m2.h.bar)]
Permeability [l/(m2.h.bar)]
inced MC
Ea
ICx
05-01-00
05-20-00
05-29-00
06-08-00
06-17-00
06-27-00
07-07-00
07-17-00
07-27-00
08-05-00
08-15-00
08-23-00
09-01-00
09-19-01
10-04-01
10-20-01
11-04-01
11-20-01
12-05-01
12-20-01
01-04-02

Mitsubishi

8 X-FLOW

8.1 Introduction

Figure 59 - Photograph X-Flow pilot plant

In this chapter the performance of the X-Flow pilot plant is described. In section 8.2 the system configuration and design data are given. In section 8.3 and 8.4 respectively the biological and membrane performance are presented. Finally in section 8.5 the conclusions are summarised.

The pilot was in operation since week 18 of the year 2000 and shown on the photograph in Figure 59. In week 26 of the year 2001, a new pilot plant has been taken in operation, which is shown in Figure 60.

In Table 17 the main activities are summarised and divided for the four phases.

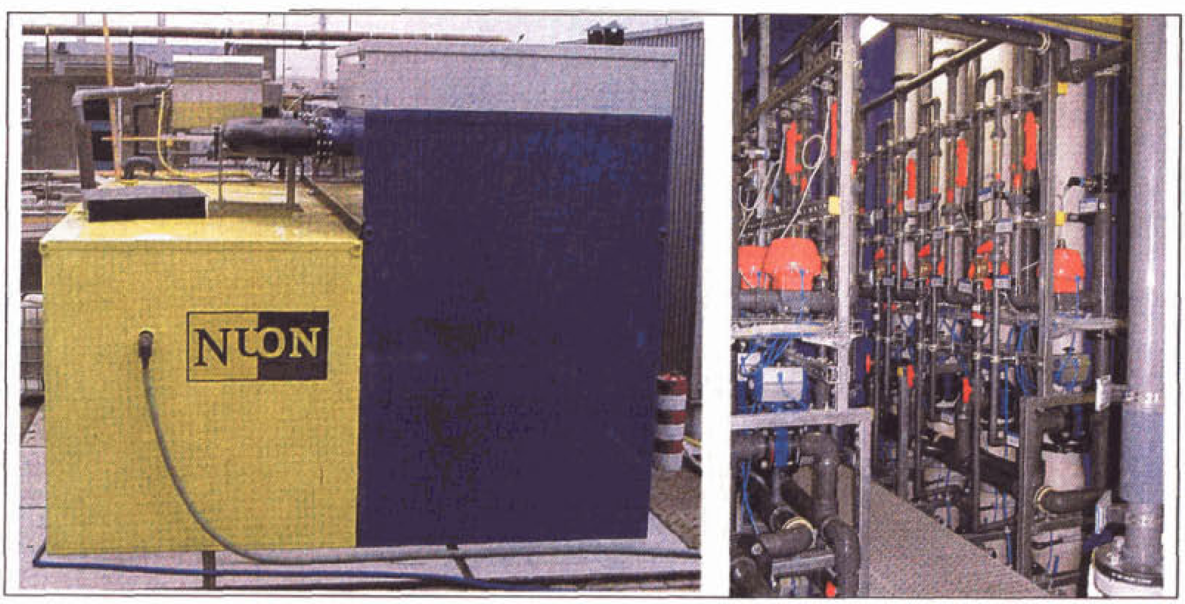


Figure 60 – Photographs of the new X-Flow pilot plant

In the period between the weeks 20 and 22 of the year 2001, the old pilot plant was still in operation but at a very low level. The sludge was thickened in order to produce seeding sludge for the new X-Flow pilot installation. Therefore, the results for the weeks 20-22 are not taken into account. The new pilot installation, with 8 X-Flow modules and a capacity of 10 m^3/h, is in operation since June 2001.

Table 17 - Research activities

phase 1	**primary clarification with pre-precipitation** 01/05/00 - 03/10/00 week 18/2000 - 40/2000
week 18-19	testing installation and seeding with sludge
week 20-29	constant flow
week 27	adjustments on the pilot system
week 30-34	proportional flow, including peak test
week 35-38	optimisation of pilot settings
week 39-41	proportional flow
phase 2	**primary clarification with simultaneous precipitation** 04/10/00 - 10/01/01 week 40/2000 - 02/2001
week 41-42	chemical cleaning (in place) and peak flow
week 42-44	constant flow
week 44-50	proportional flow with limited peak flow, including chemical cleaning 06/11/00 - adjustments operational settings membrane system 11/12/00 - introduction weekly maintenance cleaning procedure
week 51-52	constant flow
week 01	introduction daily 4 hour relaxation period
phase 3	**raw influent with simultaneous precipitation** 11/01/01 - 11/05/01 week 02/2001 - 19/2001
week 02-11	optimisation of process settings, higher continuous flux
week 12	new module installed
week 13-19	constant flow with peak test, optimisation maintenance cleaning
phase 4 (new pilot plant)	**raw influent with biological phosphorus removal** 12/05/01 - 31/12/01 week 20/2001 - 52/2001
week 20-22	low flux, preparations for new pilot-installation
week 23-26	no pilot plant in operation
week 27-34	new pilot plant, start up
week 35-52	multiple module sequencing, maintenance cleaning and peak testing

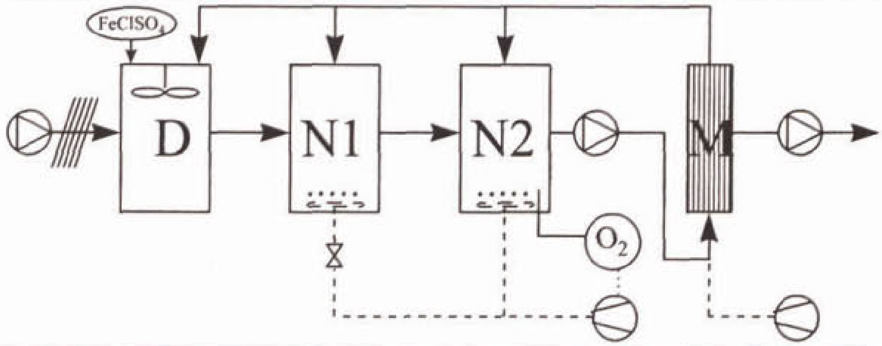
8.2 System configuration

The configuration and design figures of the two X-Flow pilot plants is schematically presented in Table 18 and Table 19.

The configuration and some of the process settings were changed during the research period. On 6 November 2000 (week 45/2000) the settings of the membrane installation were changed, based on tests performed by X-Flow. From these tests it was concluded that the cross-flow should be increased at peak flow. The details from the new settings are described in section 8.4.

For the aeration of the sludge in the compartments N1 and N2 there is one compressor. The aeration in compartment N1 can be shut down by the aeration control. The aeration control was based on the oxygen measurement in the second aeration compartment (N2).

Table 18 - Design figures X-Flow pilot plant



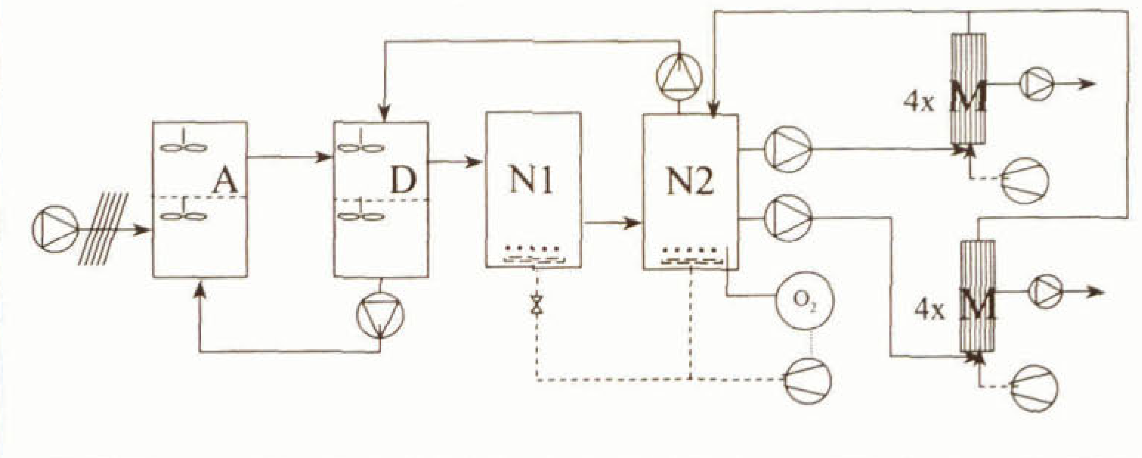
process part	parameter	unit	value
influent pump	capacity	m^3/h	0 - 2.5
	RWF design flow	m^3/h	1.8 $\rightarrow$ 1.25 [1]
	Design flow	m^3/d	9.0
influent screen	type	-	rotating drum
	slot size	mm	0.50
aeration tank	total volume	m^3	5.70
	- anoxic volume (D)	m^3 (l x w x wd)	1.40
	- anoxic/oxic comp. (N1)	m^3 (l x w x wd)	2.10
	- oxic compartment (N2)	m^3 (l x w x wd)	2.10
	- membrane module (M)	m^3	$\approx$ 0.10
	depth aeration tank	m	1.75
aeration biology	compressor capacity	Nm^3/h	100
	number of domes	-	12 (2 x 6)
ferric dosing [2]	type	-	$FeClSO_4$
	ferric content	%	12.3
	capacity (at Me/P = 0.8)	ml/h	40
membrane filtration	number of modules	-	1
	surface	m^2	30
	max. net flux (at RWF)	$l/(m^2.h)$	60 $\rightarrow$ 41.7 [1]

technical adjustments in week 27 and 45 of 2001			*< week 27*	*wk 27-45*	*> week 45*
aeration membranes	air flush	Nm^3/h	17	17	
	air flush ON/OFF	s	5 / 300	5 / 480	
	continuous air lift	Nm^3/h	8-10	8-12	
re-circulation flows	pump N2 $\rightarrow$ M	m^3/h	40	16	24-34
	sludge M $\rightarrow$ N2	m^3/h	30	14	22-32
	internal M $\rightarrow$ N1	m^3/h	8	0	0
	nitrate M $\rightarrow$ D	m^3/h	2	1.8	2

1: the hydraulic design load is increased during phase 1 of the pilot research

2: the ferric dosing is stopped at 11 May 2001

Table 19 - Design figures new X-Flow pilot plant



process part	parameter	unit	value
influent pump	capacity	m^3/h	0 - 10
	RWF design flow	m^3/h	10
	Design flow	m^3/d	50
influent screen	type	-	rotating drum
	slot size	mm	0.50
aeration tank	total volume	m^3	40.8
	- anaerobic volume (A)	m^3	12.0
	- anoxic comp. (D)	m^3	7.0
	- anoxic/oxic comp. (N1)	m^3	10.5
	- oxic compartment (N2)	m^3	10.5
	- membrane modules (M)	m^3	≈ 0.80
	depth aeration tank	m	3.0
aeration biology	compressor capacity	Nm^3/h	20-140 (FO)
membrane filtration	number of modules	-	8 #
	surface each	m^2	30
	surface total	m^2	240
	max. net flux (at RWF)	$l/(m^2.h)$	41.7
aeration membranes	air flush	Nm^3/h.module	15-20
	air flush ON/OFF	s	7 / 200
	continuous air lift	Nm^3/h	10-15
re-circulation flows	pumps N2 → M	m^3/h	2 x 80
	internal D → A	m^3/h	15
	internal N2 → D	m^3/h	15

\# 2 parallel streets 4 modules with 5.2 mm tubes and 4 modules with 8.0 mm tubes

8.3 Biological performance

8.3.1 Process conditions

The main process conditions are presented graphically in Figure 61 to Figure 63. The main average process conditions during the phases of the pilot research are presented in Table 20.

Table 20 - Process conditions

parameter	unit		phase 1	phase 2	phase 3	phase 4
influent flow	m^3/d		8.0	9.6	5.9	33
process temperature	°C	average	23	13	14	n.a.
		range	15 - 35	3 - 20	5 - 25	
pH	-		7.4	7.4	7.3	7.4
biological loading	kg COD/(kg MLSS.d)		0.070	0.063	0.091	0.054 #
sludge concentration	kg MLSS/m^3		7.5	9.9	7.2	10.6
organic part	%		61	59	54	63
sludge production	kg MLSS/d		0.64	0.96	0.98	8.8
sludge age	d		66	59	42	34 #
ferric dosing AT	l/d		0	0.75	0.71	0
ferric dosing ratio	mol Fe/mol P		0	1.0	1.2	0

\# calculation based on the activated sludge volume excl. anaerobic tank (28.8 m^3)

The design flow at dry weather flow conditions amounted to 9.0 m^3/d for the old pilot plant and 50 m^3/d for the new pilot plant. The average influent flow during phase 1, 2 and 3 varied strongly due to long rain weather conditions. The average flow was lower than the design flow, mainly due to technical reasons: software and electrical problems and frozen feed pipes and feed pump. In phase 4 the average daily flow of the new pilot plant was lower than the design flow, due to failures in the membrane section.

The set-point for the sludge concentration was 10 kg MLSS/m^3. The process conditions during the four research periods differed significantly as shown in Table 20. Besides the large influent flow variations also the process temperature and the biological loading changed.

8.3.2 *Results*

The main biological parameters are presented graphically in Figure 64 to Figure 66. In Table 21 the average feed and permeate concentrations are presented.

Table 21 - Feed and permeate concentrations

parameter		unit	phase 1	phase 2	phase 3	phase 4
COD	feed	mg/l	370	345	632	569
	permeate	mg/l	32	33	43	36
	efficiency	%	91	90	93	94
N_{kj}	feed	mg/l	65	43	64	56
	permeate	mg/l	5.0	7.1	7.1	3.6
NO_3-N	permeate	mg/l	11.2	6.3	10.6	4.2
N_{total}	permeate	mg/l	16.2	13.4	17.7	7.8
	efficiency	%	75	69	72	86
P_{total}	feed	mg/l	8,3	8.4	10.7	11.3
	permeate	mg/l	2,5	1.5	0.6	1.4
	efficiency	%	69	83	94	88

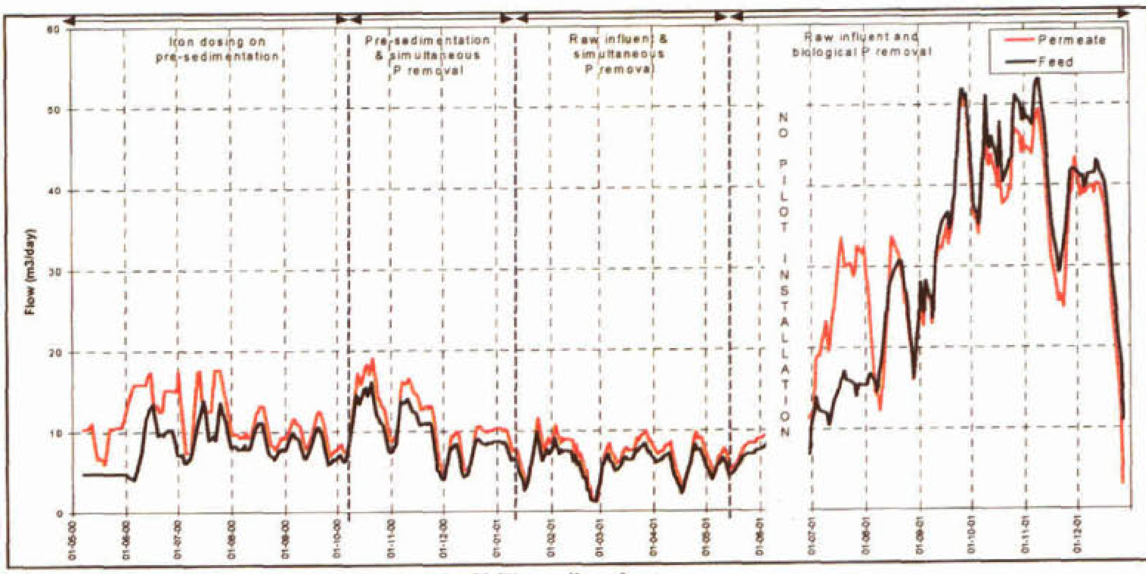


Figure 61 - Influent and permeate flow of the X-Flow pilot plant

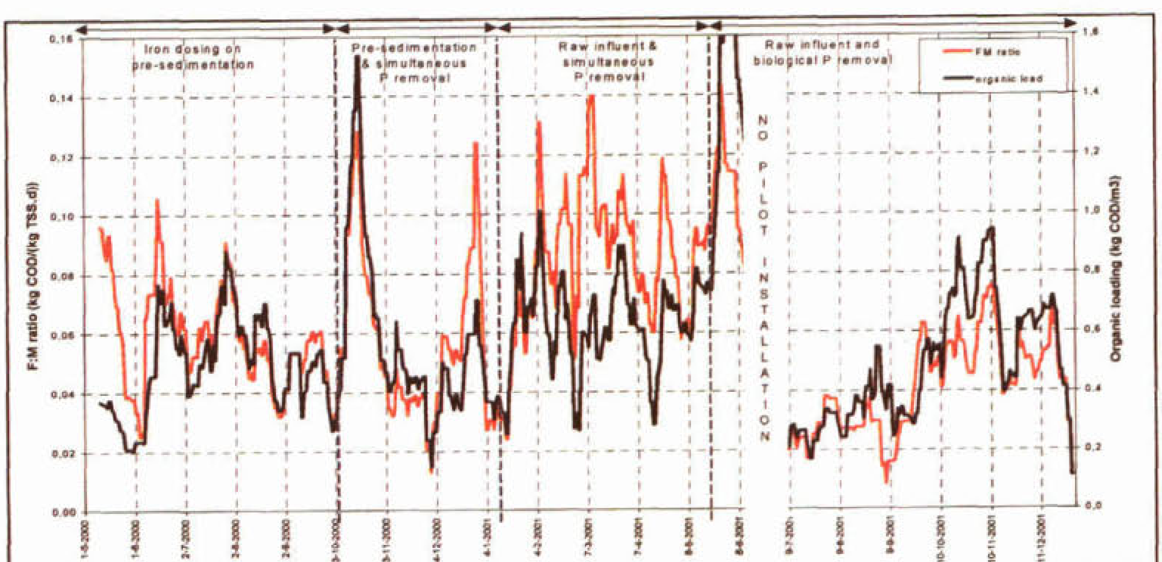


Figure 62 - Biological loading of the X-Flow pilot plant

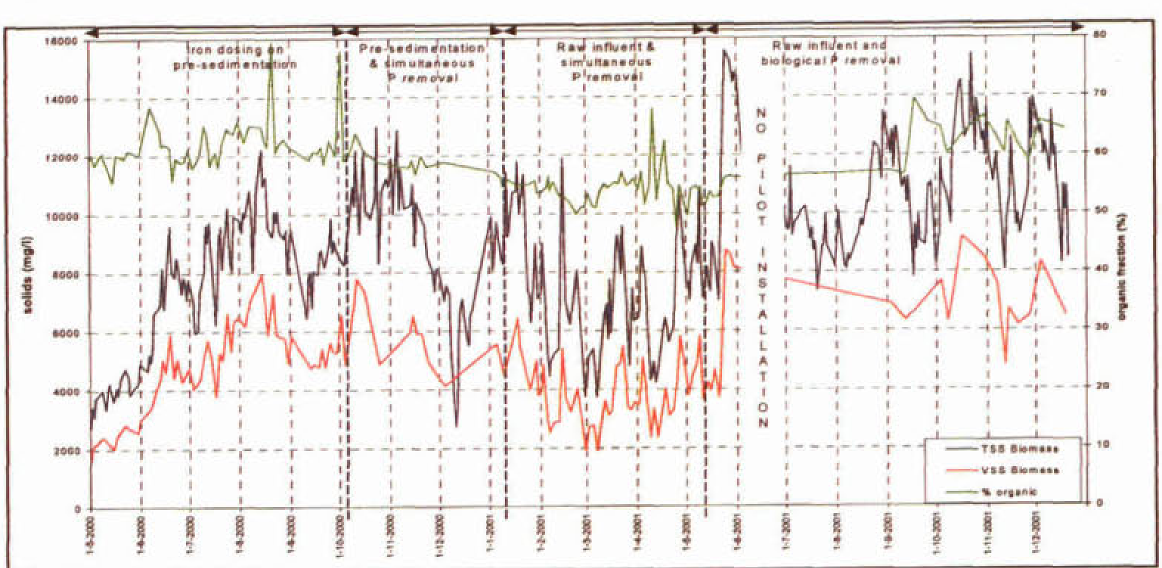


Figure 63 - Sludge concentration (total and organic) of the X-Flow pilot plant

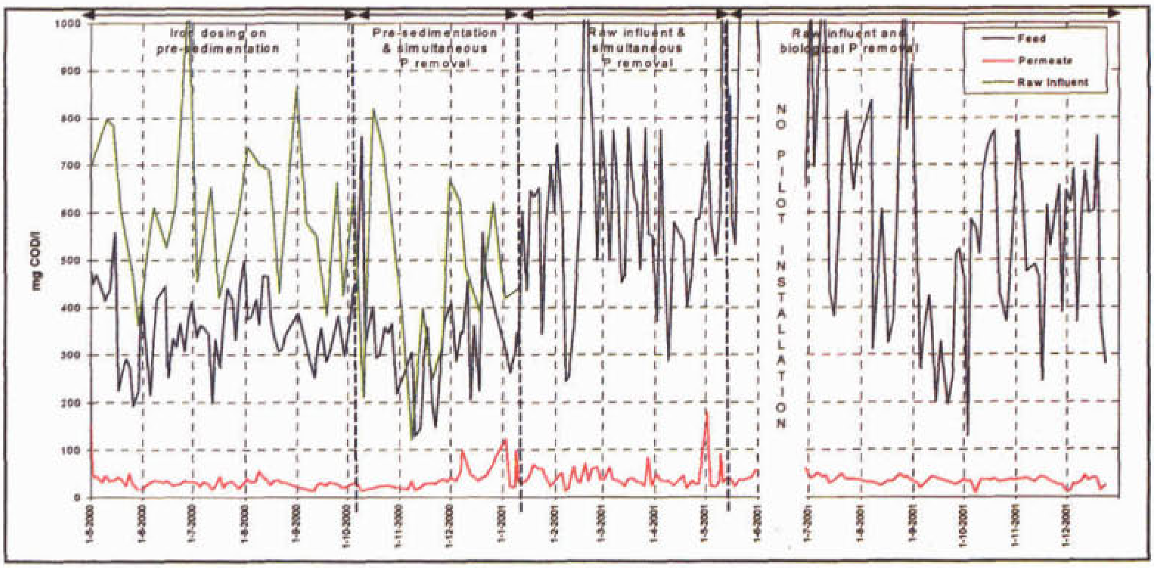


Figure 64 - COD feed and permeate concentration of the X-Flow pilot plant

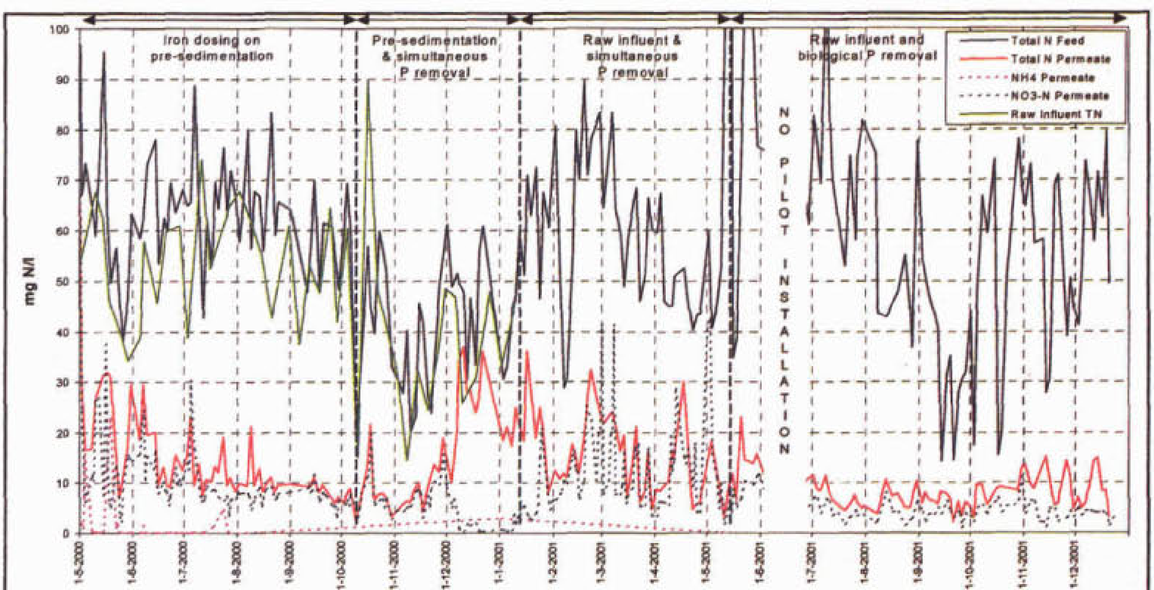


Figure 65 - Nitrogen feed and permeate concentration of the X-Flow pilot plant

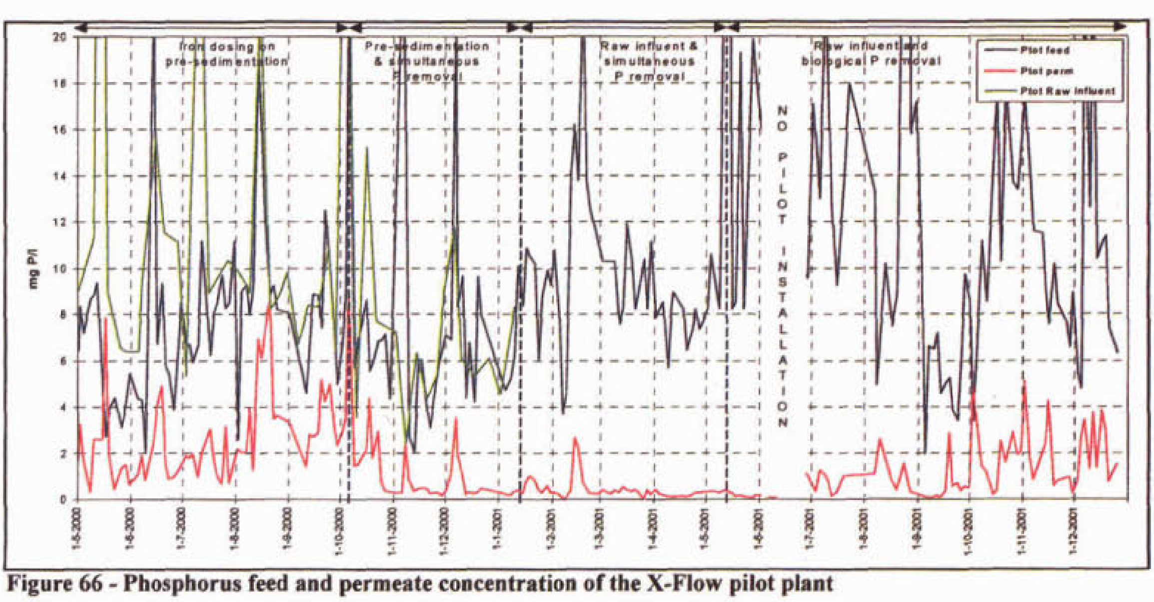


Figure 66 - Phosphorus feed and permeate concentration of the X-Flow pilot plant

COD removal
The COD-effluent concentration was relatively low due to the fact that non-soluble fraction was completely removed by the membrane filtration, COD removal efficiency remained at 90-94%.

Nitrogen removal
At the beginning of December 2000 the nitrogen total permeate concentration increased to a level of over 30 mg N_{total}/l. The nitrate concentration was almost zero, which suggested that the ammonium concentration had increased. The inhibition of the nitrification process was probably due to a NaOCl overdosing during a number of cleaning cycles. This was caused by a failure of the process control.
In phase 4 the nitrogen removal improved significantly to a nitrogen effluent concentration lower than 8 mg N_{total}/l. This is partly due to the low loading of the pilot plant in this period.

Phosphorus removal
In phase 1 it was not possible to achieve a phosphorus effluent concentration below 1 mg P_{total}/l at a Me/P dosing ratio of 1.2. Both the removal efficiency of the primary clarifiers and the sludge production were too low. In phase 2 an average permeate concentration 1.5 mg P_{total}/l was achieved at a relatively high dosing ratio. The permeate concentration decreases in phase 3 to < 1 P_{total}/l. In phase 4 the phosphorus influent concentration was relatively high and the sludge loading was relatively low. As a consequence the phosphorus effluent concentration was > 1 mg P_{total}/l.

8.3.3 Sludge characteristics

The main sludge characteristics are presented in Table 22.

Table 22 - Sludge characteristics

	unit	phase 1	phase 2	phase 3	phase 4
Sludge characteristics					
DSVI	ml/g	90	90	90	100
CST	s	60	110	70	70
Y-flow	s	160	300	n.a.	120
Viscosity					
viscosity value	mPa.s	6.0	6.9	6.6	7.0
shear rate	1/s	95	100	95	100
α-factor					
surface aeration	-	0.58	0.62	0.72	n.a.
bubble aeration	-	0.52	0.43	0.82	n.a.
Gravity thickening					
settling velocity	cm/h	4.3	12.1 [1]	9.6	n.a.
maximum concentration	%	2.3	3.3 [1]	2.3	n.a.
Mechanical thickening					
MLSS at 3900 rpm / 10 min	%	6.4	8.2	n.a.	n.a.
MLSS at 1000 rpm / 3 min.	%	2.3	4.7	6.4	n.a.

1: based on 1 measurement at a sludge concentration of 7 kg MLSS/m^3

The DSVI was relatively stable at a level of 80 to 90 ml/g. Both the CST and the Y-flow can be related to the sludge concentration. The viscosity of the MBR sludge varied from 5-10 mPa.s, which was relatively low for the applied sludge concentrations. The sludge viscosity also appeared to be related to the sludge concentration.

The α-factor with bubble aeration varied between 0.4 and 0.8 at a sludge concentration of 7 to 11 kg MLSS/m^3. The average α-factor at 10 kg MLSS/m^3 was approximately 0.6. With a surface aerator values 15% higher were achieved. In phase 3 the average α-factor with surface aeration increased to around 0.7-0.8, which was related to the relatively low sludge concentration (7-9 kg MLSS/m^3) in this period.

The average settling velocity at the beginning of the gravity thickening test amounted to 3-10 cm/h, strongly dependent on the sludge concentration. The maximum concentration after 24 hours was lower than 3% in general, indicating that gravity thickening was not really efficient. The maximum sludge concentration with mechanical thickening was 6 to 8%.

The floc structure of the MBR sludge has changed dramatically from flocs with a medium size and a compact structure to a very small one, like pin flocs, with a somewhat open structure. Of course the number of flocs also increased due to the increased dry solid content of the system.

Since the summer of 2000 the floc has become denser and solid, the size was very small. The number of mono cultures has more or less stabilised to a small number per sample, but by the end of phase 2 the number had decreased dramatically.

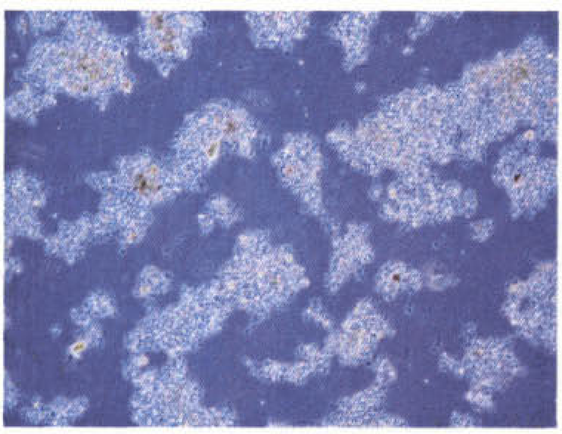


Figure 67 - Microscopic view of the activated sludge

During the test the number of filamentous organisms increased to 1 (Eikelboom), but during the last phase the number varied between 1 and 2 according Eikelboom. The type was not identified, but appeared floc bound. During phase 3 the flocs were more compact, and the number of crawling ciliates were increasing steadily.

8.4 Membrane performance

The results of the whole research period are presented graphically in the fold out page and further explanation is given in the text. The data presented is damped to aid readability, and conclusions are made regarding the membrane performance and the relevant operating & design criteria.

It was expected that the maximum achievable flux would be 60 l/(m^2.h) net for a period of 3 days continuous. The cleaning was estimated at routine intervals dependent on the permeability, 2-4 times per year was suggested and the cleaning mode was set according to X-flow's specifications of NaOCl solution followed by an acidic solution. Back flush was used between process modes and no other recovery steps/modes were anticipated.
The X-flow installation has been under constant development and optimisation. During the research period the ratio between net and gross flux varied between 60 and 90%, depending on the research period and the feed flow.

8.4.1 *Phase 1 - Primary settled wastewater and pre-precipitation*

Start up and definitions

The membranes were set up in a semi-cross flow system (~1 m/s) and set to run at a constant flux of 15 $l/(m^2.h)$ gross. After the biological system was established this was increased to 25 $l/(m^2.h)$ gross. It should be noted that the flow through the system was controlled via the permeation rate coupled to the bioreactor level. In both circumstances the average feed flow suggested a membrane efficiency of 60-65%.

After a short period of time it was clear that the semi-cross flow system was not performing, with high energy usage and unstable operability. The system was modified to a low pressure cross flow (LPCF) system based on experience with vertical air-flushing. The pressure drop over the module was reduced and the speed of sludge and air in the tubes reduced to < 1m/s. A permeate extraction pump was used to generate a trans-membrane pressure (TMP) across the membrane due to the absence of the cross flow driving force. Increased stability and operability was brought to the installation as can be seen from the comparison between Figure 68 and Figure 69.

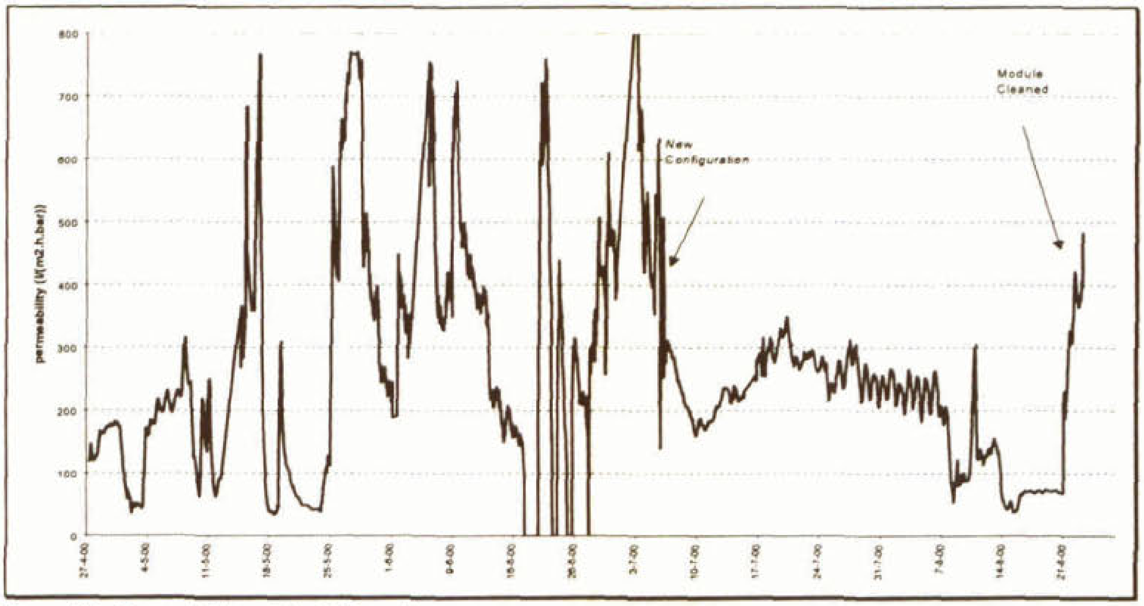


Figure 68 - Permeability of the X-flow pilot

After the modifications, the pilot plant was continuously fed with a relatively high influent flow of 525 l/h, which related to a gross flux of 25 $l/(m^2.h)$ on the membrane with a 72% efficiency. The new configuration allowed the membrane to recover and work under less strenuous processing conditions. Since recovery was seen, with only the low pressure cross flow (LPCF) and aeration on, after an electrical problem. The system was considered ready for proportional flow.

The proportional flow was started at 24/07/00 at the design flow of the pilot plant (9.5 m^3/d) with a scaling to the Beverwijk WWTP of 4500:1. As a consequence the gross flux decreased to an average of 15 $l/(m^2.h)$. The gross flux remained relatively low until the end of the proportional flow period as there was no real RWF situation. During the proportional flow period the permeability slowly decreases from 270 $l/(m^2.h.bar)$ to 250 $l/(m^2.h.bar)$. This was probably due to slime formation on the permeate side of the module as a result of after growth.

This phenomenon was visually registered and a regular chemical back-flush prevented this subsequently.

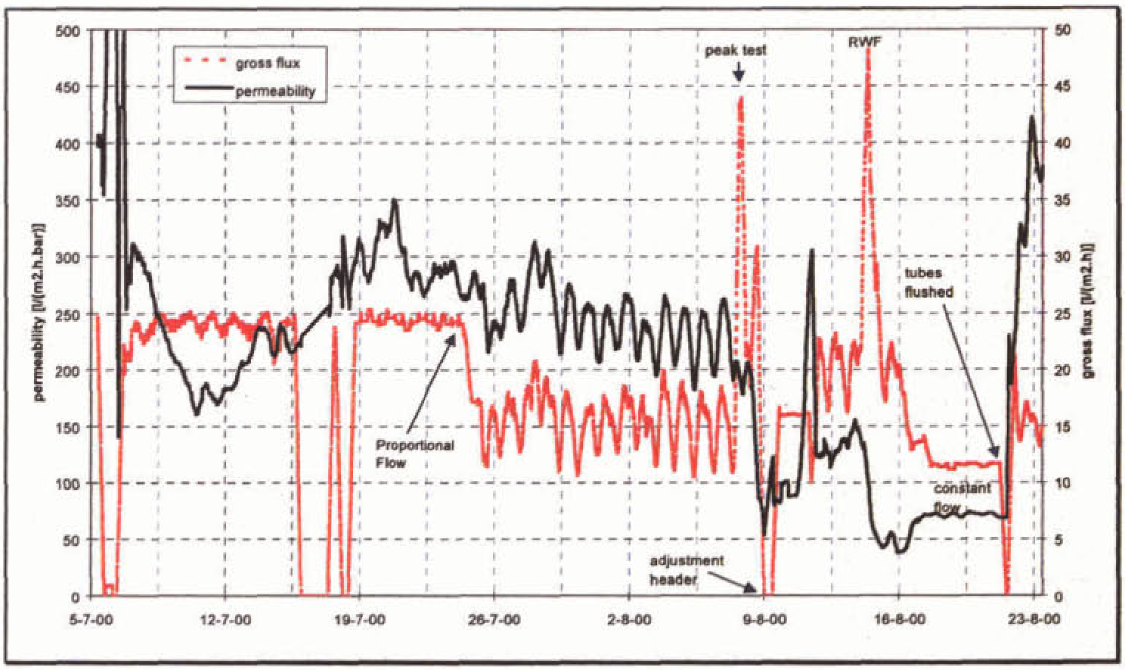


Figure 69 - Permeability and flux of the adapted X-flow pilot plant

Peak tests

The first peak test was executed on 7/8/00. The gross flux was 60 $l/(m^2.h)$ with a net yield of 50 $l/(m^2.h)$. The peak test lasted 4 hours. As a result of the peak flow, the permeability decreased catastrophically from 250 $l/(m^2.h.bar)$ to around 75 $l/(m^2.h.bar)$ suggesting that the X-flow membrane was not able to handle a peak flux of 50 $l/(m^2.h)$ net let alone the design peak flux of net 60 $l/(m^2.h)$.

The conclusion of the peak test was that the header configuration did not function well and generated too much pressure loss. During back-flush, not only permeate but also air was brought into the system and as a result the efficiency of the back-flush was probably too low to keep the membrane clean. The headers were enlarged and thereafter the system was set back to proportional flow. However, during a RWF situation on 14/08/00 the permeability dropped rapidly to around 50 $l/(m^2.h.bar)$. On visual inspection it became clear that about 60% of the tubes were blocked with thickened sludge. The tubes were manually cleaned with a lance and as a result the permeability increased to more than 300 $l/(m^2.h.bar)$.
On 28/8/00 a second peak test was executed. The gross flux was 45 $l/(m^2.h)$ or 33 $l/(m^2.h)$ net. Again, the tubes blocked and the permeability had decreased significantly. The adjustments to the headers did not seem to have the desired effect, or the flux required was just too high.

There followed a period of considerable flux optimisation, using variation cross-flow, aeration, relaxation and chemical injection modes, the peaks are shown in Figure 70. The permeability and gross flux were made more stable and some permeability recovery was seen.

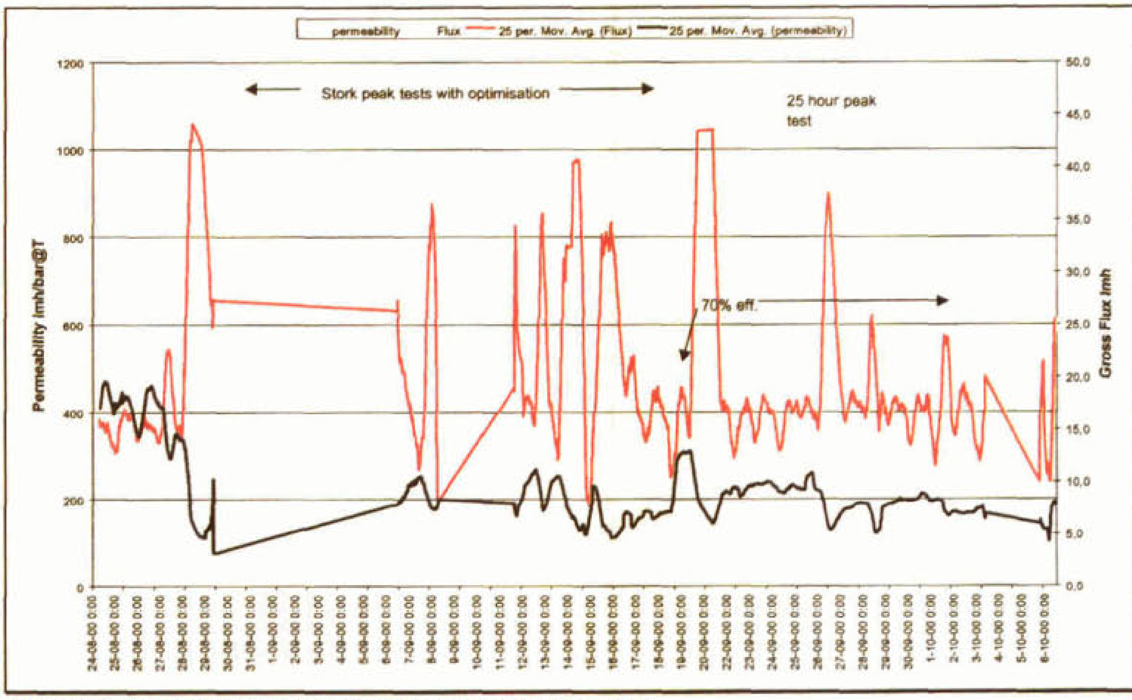


Figure 70 - Permeability and flux during optimised peak testing

8.4.2 *Phase 2 - Primary settled wastewater and simultaneous precipitation*

In phase 2 of the research program the system was further optimised. In this section the main events are described chronologically.

On 4/10/00 the back-flush settings were changed and at the same time the switch between pre-precipitation and simultaneous precipitation took place. In the following period, between 8 and 11/10/00 a RWF period occurred and the membrane was not able to handle this long-term peak flow. The permeability decreased rapidly from 190 l/(m^2.h.bar) to below 100 l/(m^2.h.bar). Based on the results it was decided to remove the module and clean it at a X-flow facility in Friesland, The Netherlands. On arrival 50-80 tubes were clogged and had to be unplugged. After this a chemical cleaning took place in different steps using Ultrasil 75 (combination of sulphuric acid and phosphoric acid) and NaOCl (500 ppm). The cleaning re-established the permeability to > 500 l/(m^2.h.bar).

After re-installing the module it was fluxed with a peak gross flux of 40-43 l/(m^2.h) for 4 days. As a consequence the permeability decreased to around 200 l/(m^2.h.bar). After switching to a constant gross flux of 22.5 l/(m^2.h), the permeability recovered to 225-300 l/(m^2.h.bar). The peak test was proven at 43 l/(m^2.h) gross or 32 l/(m^2.h) net and showed recovery (see Figure 71).

Further optimisation of the process and membrane running was continued to late December, including limited proportional (peak flux limits), night time relaxation, and lower TSS concentrations in the bioreactor. In all cases the peaks eventually caused a deterioration in the permeability to levels where the module had to be cleaned. It was eventually concluded that the membrane was unsuitable to handle flux variations of >1.5x and the system was set to continuous permeation mode (normal set up for industrial MBR systems). The results can be seen from Figure 72.

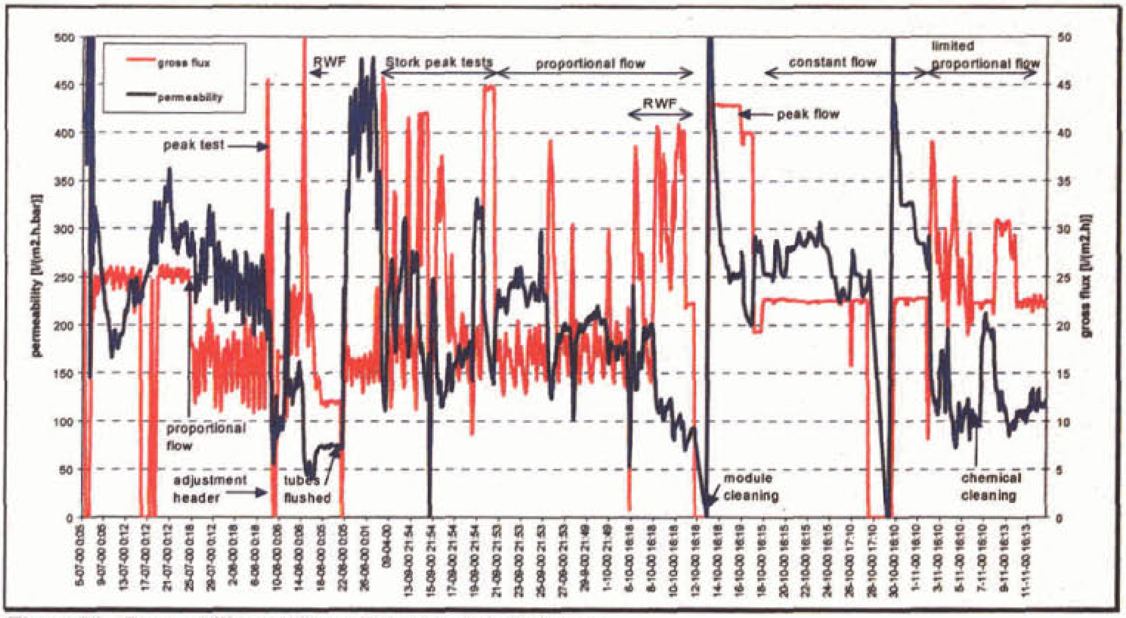


Figure 71 - Permeability and flux of the adapted pilot plant

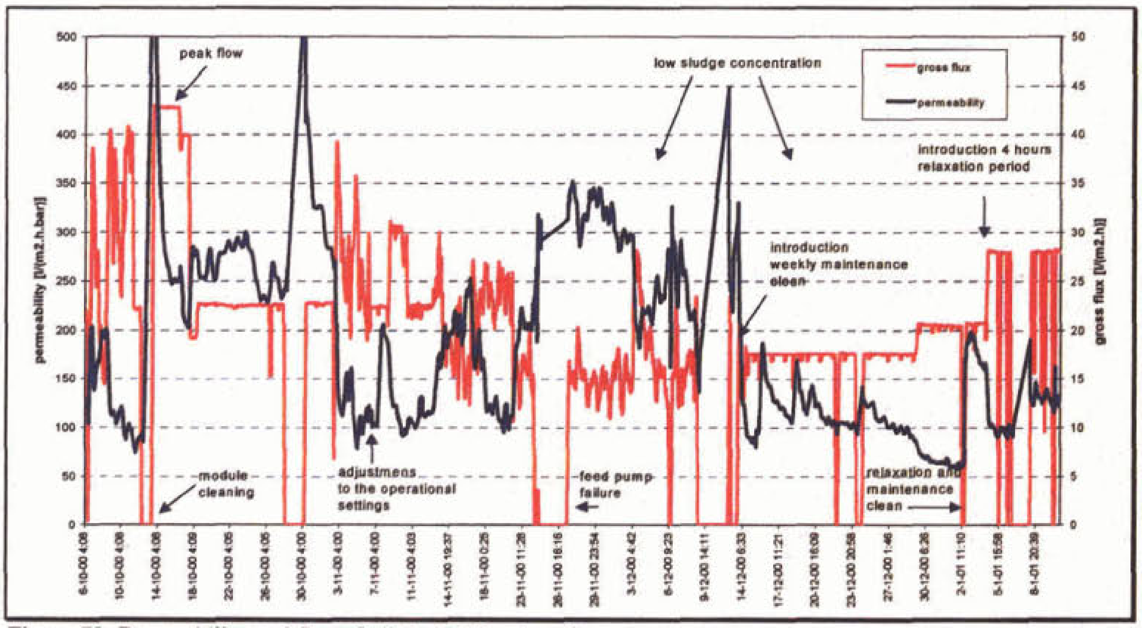


Figure 72 -Permeability and flux of pilot plant in phase 2

The system was set to continuous mode of permeate extraction from on 11/12/00. The optimisation remained and was further developed. In full-scale a part of the MBR system would be shut off at low flows during the night (except for RWF conditions), thus modules would always be allowed to relax. To simulate this procedure, the operational settings were changed on 02/01/01. Since that time a relaxation period of 4 hours was introduced.

8.4.3 *Phase 3 - Raw influent and simultaneous precipitation*

The system was yet further optimised under the constant flow conditions using a maintenance cleaning (MC) of once per week. The gross flux was eventually set at a constant 28 $l/(m^2.h)$ gross, but the permeability remained low. Unfortunately a major problem occurred with the control system of the pilot, on 18-2-01. The permeation remained on whilst the bioreactor level decreased. During inspection, it seemed that the module was not recoverable and consequently a new module was installed.

As expected the new, clean module performed well under constant flow conditions, it was found that this new module could be run at a constant 42 $l/(m^2.h)$ gross 37 $l/(m^2.h)$ net and the system remained in an optimisation period with the incorporated MC procedures. The module remained in good workable condition until a new pilot was commissioned. The single module system was used to thicken up a seed sludge for the new pilot to 16 g/l before the old pilot was decommissioned on 6/6/01.

8.4.4 *Phase 4 - Raw influent and Biological P removal*

A new 10-12 m^3/h X-flow pilot was installed in June 2001, to extend and optimise the LPCF technique further. The system control was modified to avoid the onset of tube sludging and excessive permeation. The system was installed with 4 parallel 5.2mm compact modules and 4 parallel 8mm compact modules. The 5.2mm modules were put into operation, initially to ascertain comparable data to that generated by the single module pilot. The system was run at a constant 2.5m3/h feed flow on two parallel modules and later set to 3m3/h for 3 modules. In each case the reserve module was taken out of operation and rotated in the module sequence, so all modules received the same number of operating hours. Results were encouraging as continuous fluxes were achieved of 50 l/(m2.h) with an efficiency of above 90%. A peak test of 70 l/(m2.h) was executed but lead to rapid cake-filtration and permeability loss. The module however, recovered without additional cleaning steps. Figure 73 below displays data for 1 of the 4 5.2mm compact modules, UF1 m120.

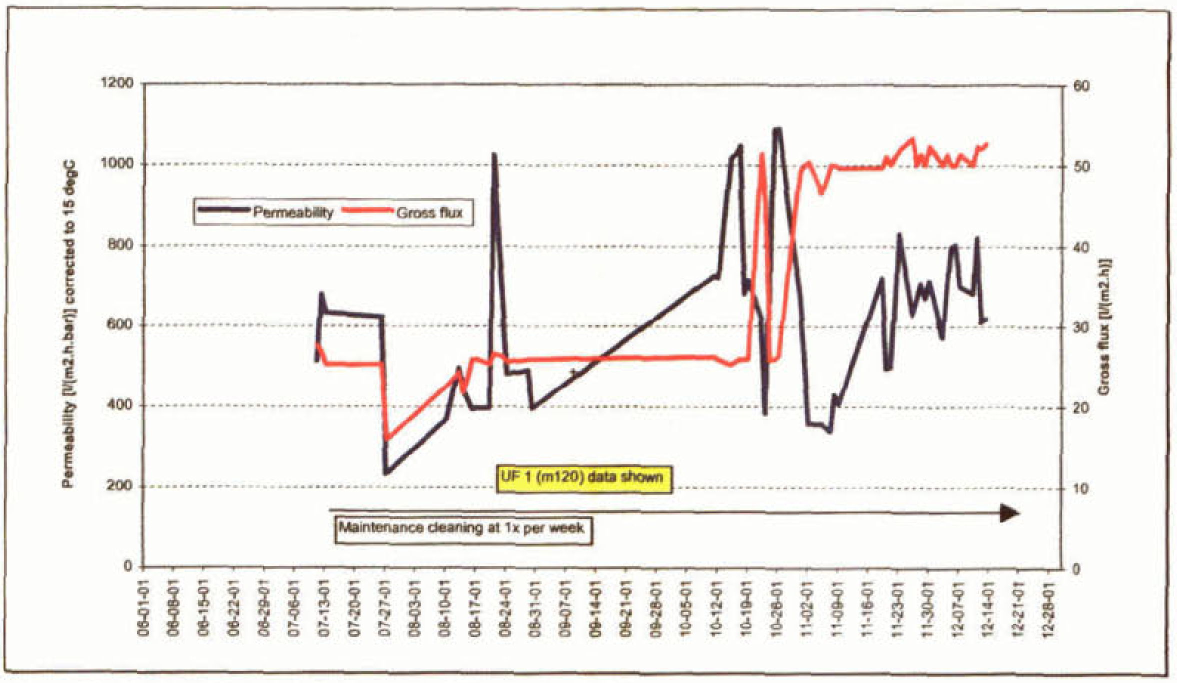


Figure 73 - UF1 m120 module data, permeability and gross flux

The system ran with weekly maintenance cleaning, this was further optimised. The efficiency, air scouring and turbulence were also optimised under steady state process conditions. Due to time constraints the effect of 'real' cross-flow filtration during rain weather periods was not assessed, or the effect of daily cross-flow increases on the modules cleaning requirements. It should be noted that the modules under the LPCF configuration were ran down to well under the 1m/s and good results were achieved at 0.5 m/s, the air flow supplied an additional 0.25-0.5 m/s to the biomass' turbulence but remained constant under normal processing conditions. With further development with the known increased fluxes associated with higher cross-flow speeds, it is believed that the LPCF system could be optimised to achieve the goals set out in the projects objectives.

During the six months of intensive testing of the new pilot the modules were not subjected to an intensive clean, maintenance cleaning on a routine basis was adequate to maintain process operability. The latter procedure was optimised during the trial and eventually required less than 45 minutes to execute on a weekly basis.

General:
The module appeared not to be affected by temperature variation, but the data for permeability was so unstable that a true correlation could not be made. The rapid decrease in permeability coupled with the fact that the module was always easier to back flush than to extract permeate, suggested the fouling on the module was predominately cake filtration.

8.5 Conclusions

First pilot :
In general, the biological performance of the X-flow pilot system was not as good as could be expected from theoretical design calculations. This was mainly due to technical failures and as a result low and irregular feeding conditions. In periods where the system was functioning technically well, the biological performance was also seen to improve.

The sludge concentration in the aeration tank was kept at a level of around 10 kg MLSS/m^3, to avoid problems with the membrane tubes and to optimise the energy consumption in relation to the α-factor. The α-factor was as expected from literature.

A peak flux of 50 l/m^2.h net was achieved for an operating period of 4 hours continuous at a temperature above 20°C. This was unsatisfactory.
The membrane was unable to satisfactorily process successive large peaks during RWF.
Continuous flow mode was found to be the best mode of operation for the module.
The maximum continuous design flux above temperatures of 5°C was established at 22.5 l/(m^2.h) net and further optimised to 37 l/(m^2.h) net. At these fluxes a cleaning in the form of a weekly maintenance clean (MC) was required with an intensive clean estimated at 4-8 times per year.
The cleaning of the module at 2-4 times per year was not proven to be correct, 4 to 8 times per year was more realistic, but at lower chemical concentrations. MC proved to be a most successful procedure as was night-time relaxation. The standard Back flushing procedure proved to be a critical aspect on the membrane performance and efficiency.
The membrane was difficult to stabilise. At fluxes of less than 20 l/(m^2.h) the permeability remained under control and the membrane achieved stability. Above 20 l/(m^2.h) the system required extensive optimisation to maintain performance between the weekly MC procedure.

New pilot :

In the biological performance of the X-flow system was as good as the theoretical design calculations. The biological phosphorus removal process was proven to work and the nitrogen removal process was according to the design criteria. Biological phosphorus removal had no detrimental effect on the membrane performance.

The sludge concentration in the aeration tank was kept at a level of around 10 kg MLSS/m^3, to avoid problems with the membrane tubes and to optimise the energy consumption in relation to the α-factor. The sludging problem that occurred with the first pilot was not seen to the same extent in the large pilot, some rapid cake formation was evident but the modules were able to recover without further assistance.

A continuous flux of 50 l/(m^2.h) net was achieved for an operating period of several weeks, even at low process temperatures of >10°C. This was aided by the rotation of modules in the processing sequence (full scale simulation).

The membrane was unable to satisfactorily process successive large peaks beyond 70 l/(m^2.h). Continuous flow mode was found to be the best mode of operation for the module as seen in the first pilot.

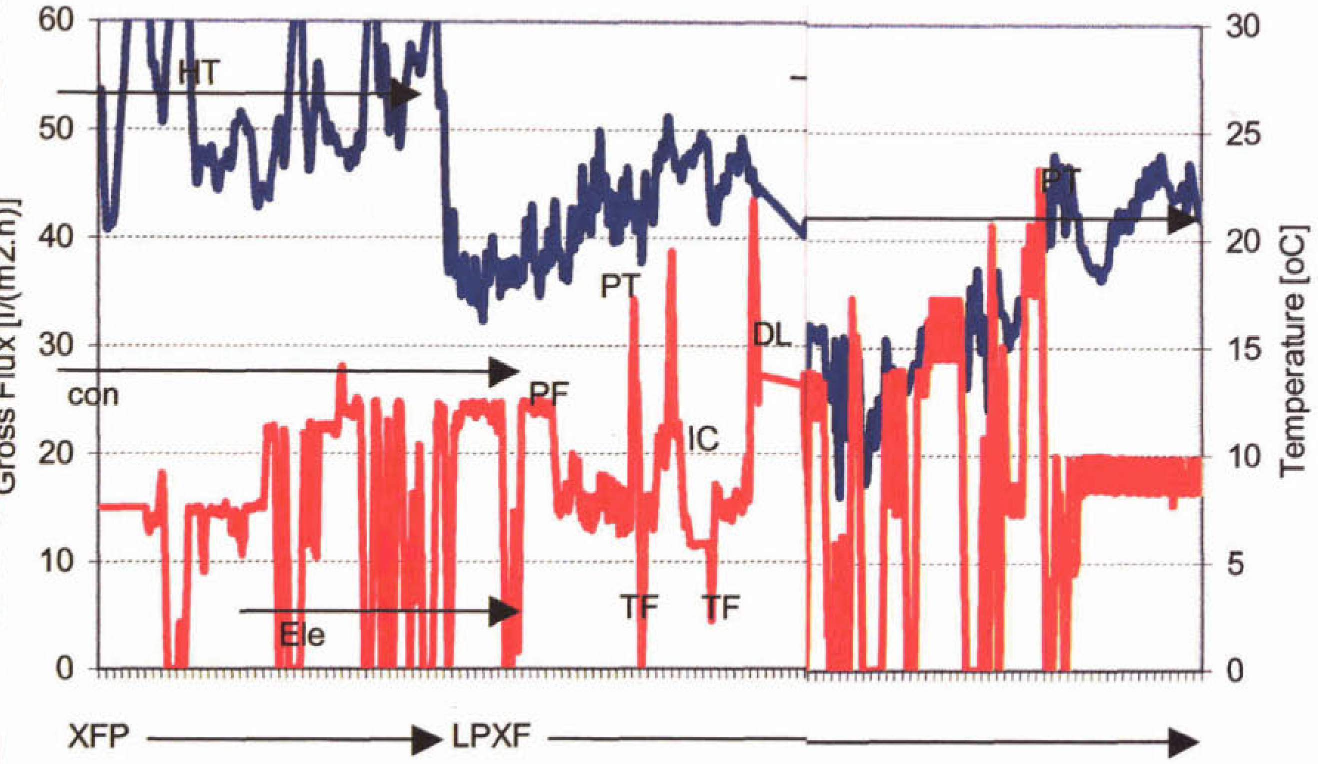
The maintenance cleaning procedure was found to perform adequately in order to maintain the process permeability under all process and temperature conditions. As confirmed by the first pilot the normal processing mode consisted of a suction step followed by a short and intensive back wash. The MC procedure was set at 1x per week and during the 6 month trial the modules were not subjected to an intensive clean.

The large pilot was fully automated and operated via the telephone/internet – this proved successful and as a result reduced the man hours required for daily supervision. The software was also optimised via the Internet and the system optimised efficiently.

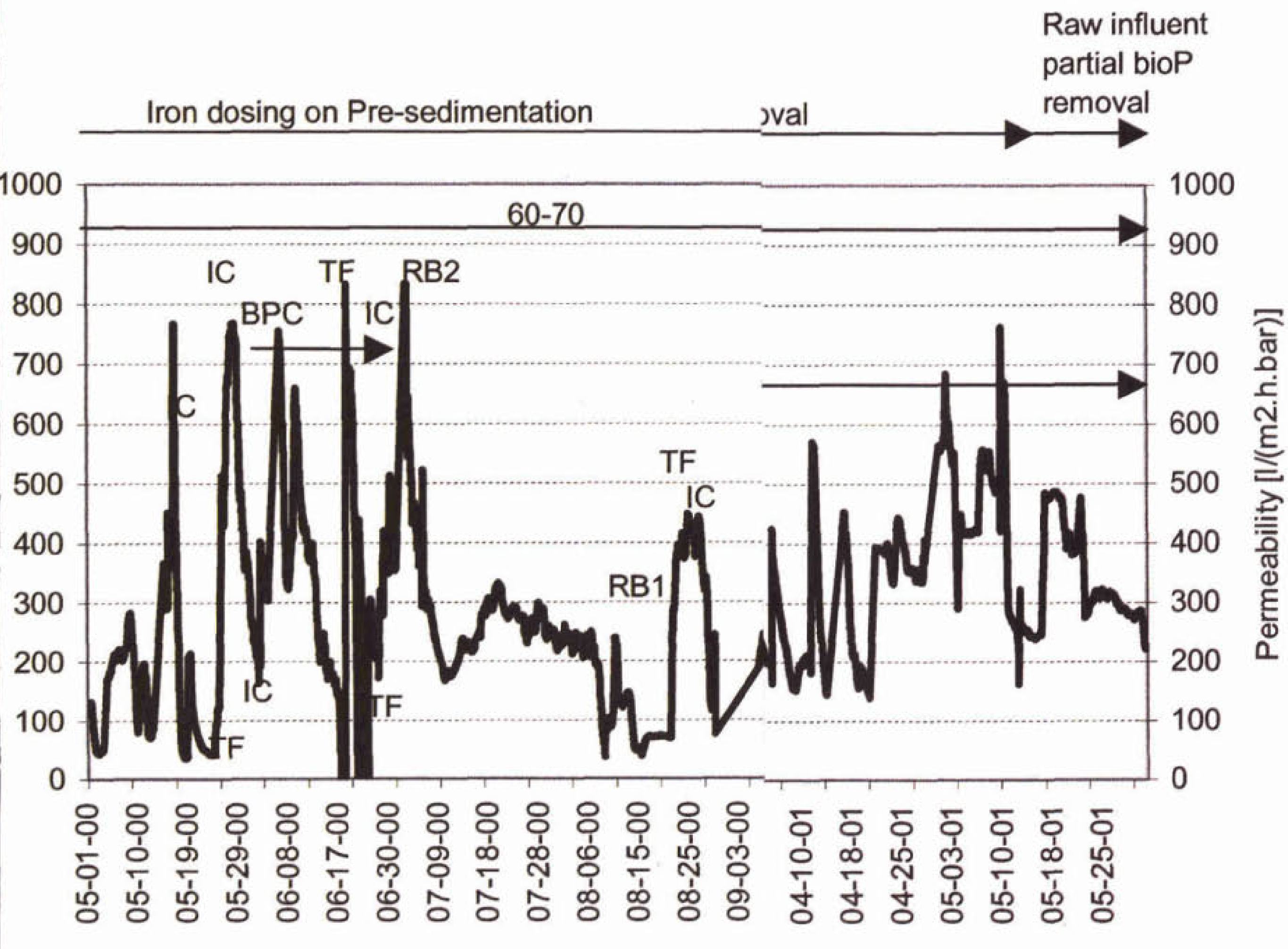
Due to time limitations, the effect of increased cross-flow on the cleaning, and the addition flux generated for the RWF situation were not investigated. Here the goal was to ascertain the best ratio for LPCF and full cross-flow under the DWF and RWF circumstances. The latter will be investigated further.

X-FLOW *Fold-out Graphic*

PF	Transition to Proportional flow (direct signal STP Beverwijk)
con	Transition to Constant flow
PT	Peak Flux test period at suppliers maximum flux specification
OPT	Optimised flux tests, aeration flow, relaxation and cross-flow.
RB1	Rebuild 1, reduction in piping resistance - lower Cross-flow pressure
RB2	System converted from Cross flow configuration to Low Pressure Cross-Flow
Ele	Electrical Problems
DL	Data logger not on line
TF	Membrane module tube flush (mechanical cleaning)
MC	Standard Maintenance Cleaning Procedure
HT	High temperature caused by Cross-flow operation
XFP	Cross flow mode of operation
LPXF	Low Pressure Cross flow mode of operation
IC	Intensive Cleaning Procedure correctly executed
BPC	Extended back-pulse with chemicals
LPF	Limited Proportional flow
Rx	Night time relaxation period incorporated for 2-4 hrs
NM	New Module installed
CB	Complete non-recoverable blockage of tubes
XFT	Cross flow testing
60-70	System efficiency
80-90	System efficiency



Gross Flux [l/(m2.h)]
Temperature [oC]
HT
con
PF
PT
DL
IC
TF
TF
Ele
PT
XFP
LPXF



Iron dosing on Pre-sedimentation
Raw influent
partial bioP
removal
60-70
IC
TF
RB2
BPC
IC
IC
TF
TF
IC
RB1
Permeability [l/(m2.h.bar)]
05-01-00
05-10-00
05-19-00
05-29-00
06-08-00
06-17-00
06-30-00
07-09-00
07-18-00
07-28-00
08-06-00
08-15-00
08-25-00
09-03-00
04-10-01
04-18-01
04-25-01
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05-18-01
05-25-01

X-Flow

9 ZENON

9.1 Introduction

Figure 74 - Photograph Zenon pilot plant

In this chapter the performance of the Zenon pilot plant is described. In section 9.2 the system configuration and design data are given. In section 9.3 and 9.4 respectively the biological and membrane performance are presented. Finally in section 9.5 the conclusions are summarised.

The pilot was in operation since week 12 of the year 2000 and shown on the photograph in Figure 74. In Table 23 the main activities are summarised and divided for the three phases.

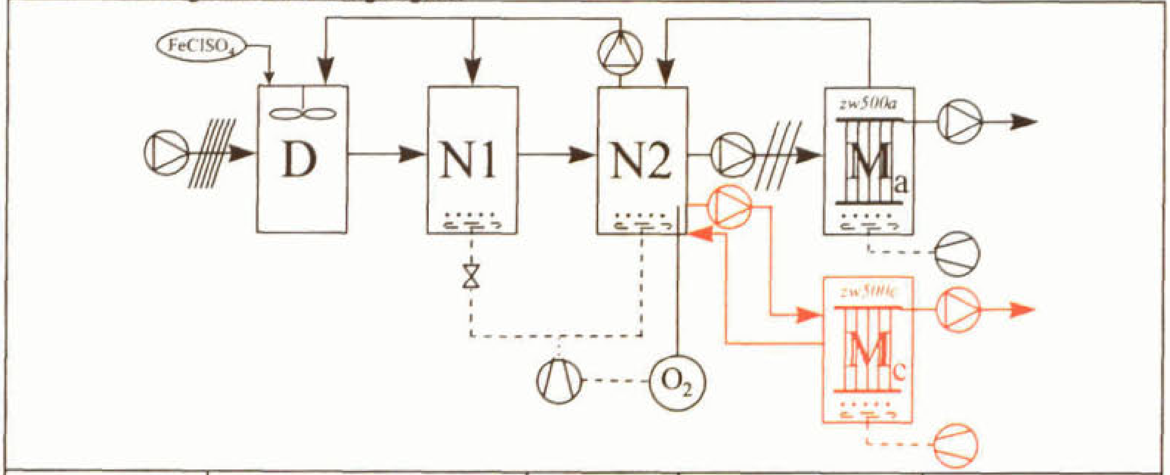
Table 23 - Research activities

phase 1		**primary clarification with pre-precipitation** 27/03/00 - 03/10/00 week 12/2000 - 40/2000
week 12-13		testing installation and seeding with sludge
week 13-18		constant flow
week 19-40		proportional flow
week 26-27		peak flow test
week 31-34		low temperature test, followed by chemical cleaning
phase 2		**primary clarification with simultaneous precipitation** 04/10/00 - 10/01/01 week 40/2000 - 02/2001
week 41		manual and chemical recovery cleaning
week 48-49		optimisation pilot plant configuration and operation, chemical recovery cleaning
week 50-02		optimisation of the maintenance cleaning procedure
phase 3		**raw influent with simultaneous precipitation** 11/01/01 - 11/05/01 week 02/2001 - 19/2001
week 02-19	zw500a	proportional flow, optimisation maintenance cleaning
week 11		new membrane section (ZW500c) in operation
week 12-19	zw500c	constant and peak flow, optimisation maintenance cleaning
phase 4		**raw influent with biological phosphorus removal** 12/05/01 - 31/12/01 week 20/2001 - 52/2001
week 20-52	zw500a	proportional flow, optimisation cleaning procedure
week 20-52	zw500c	constant flow, optimisation cleaning procedure week 46: one of the 3 modules has been replaced

9.2 System configuration

The configuration and design of the Zenon pilot are presented in Table 24. The configuration and some process settings were changed during the research period. The main changes concerned the re-circulation flow from the membrane section M to the aeration tank, as shown in Table 24. This re-circulation was switched from N1 to N2 in week 28 to reduce the oxygen re-circulation to N1 to favour de-nitrification in this compartment. For the aeration of the activated sludge in N1 and N2 only one compressor was installed. The aeration in N1 could be shut down by the aeration control, which was based on the oxygen measurement in N2. The coarse bubble aeration in the membrane section was at a fixed constant level, but cycled over two of the four installed modules at all time.

Table 24 - Configuration and design figures



process part	parameter	unit	zw500a < *week 11/2001*	zw500a / zw500c > *week 11/2001*
influent pump	capacity	m^3/h	≈ 15[1]	
	RWF design flow	m^3/h	7.6	
	design flow	m^3/d	38.0	
influent screen [1]	type	-	static half drum with brush	
	slot size	mm	0.75	
aeration tank	total volume (and depth)	m^3	23.6	26.6
	- anoxic volume (D)	m^3 (m)	4.38 (1.75)	4.38 (1.75)
	- anoxic/oxic comp. (N1)	m^3 (m)	7.66 (1.75)	7.66 (1.75)
	- oxic compartment (N2)	m^3 (m)	7.66 (1.75)	7.66 (1.75)
	- membrane tank (M_a)	m^3 (m)	3.92 (2.80)	3.92 (2.8)
	- membrane tank (M_c)	m^3 (m)	-	3.0 (2.0)
aeration biology	compressor capacity	Nm^3/h	100	
	number of domes	-	24 (2 x 12)	
ferric dosing [2]	type	-	$FeClSO_4$	
	ferric content	%	12.3	
	capacity (at Me/P = 0.8)	ml/h	80-100	
membrane filtration	number of modules	-	4	4 / 3
	surface each module	m^2	46	46 / 20
	total surface	m^2	184	184 / 60
	max. net flux (at RWF)	$l/(m^2.h)$	41.3	35 / -
aeration membranes	compressor capacity	Nm^3/h	100 (cycled)	100 / 60 (cycled)
	specific capacity	$Nm^3/(m^2.h)$	0.54	0.54 / 0.50
re-circulation flows			*< 07/12/00*	*> 07/12/00*
	sludge N2 → M_a	m^3/h	24	18-30 (5:1) [3]
	internal N2 → N1	m^3/h	8	8
	nitrate N2 → D	m^3/h	14.5	14.5
	sludge N2 → M_c	m^3/h	-	5 - 15 (x:1) [4]
sludge screen	type	-	basket filter	-
	slot size	mm	2.0 (later removed)	

1: *the influent micro screen is in operation since 8 September 2000*

2: the ferric dosing is in operation since 4 October 2000, and until 11 May 2001

3: the re-circulation flow from N2 to M_a is related to the influent flow in a ratio 5:1

4: the re-circulation flow from N2 to M_c is related to the influent flow in a ratio x:1 (adjusted during research)

In week 11 of 2001 a second membrane pilot system was installed. This section contained a new Zenon module (ZW500c), which displayed an optimised configuration compared with the "old" Zenon ZW500a module. The second installation has a surface area of 60 m^2 and a capacity of 1.7 - 3.0 m^3/h. The new installation is fed with a constant flow from the N2 tank, the overflow is returned to the N2 tank.

9.3 Biological performance

9.3.1 Process conditions

The main process conditions are presented graphically in Figure 75 to Figure 77. The overall process conditions during the phases of the pilot research are presented in Table 25.

Table 25 - Process conditions

parameter	unit	phase 1	phase 2	phase 3	phase 4
influent flow	m^3/d	46	74	45	44
process temperature	°C average	20	15	14	20
	range	11 - 25	10 - 21	8 - 20	10 - 28
pH	-	7.5	7.6	7.5	7.5
biological loading	kg COD/(kg MLSS.d)	0.075	0.093	0.110	0.086
sludge concentration	kg MLSS/m^3	10.4	10.9	10.0	11.2
organic part	%	63	64	63	64
sludge production	kg MLSS/d	4.8	9.3	10.2	10.1
sludge age	d	51	28	26	29
ferric dosing AT	l/d	0	2.1	2.2	0
ferric dosing ratio	mol Fe/mol P	0	0.38	0.51	0

The design flow at dry weather flow conditions amounted to 38 m^3/d. The average influent flow during the research period varied strongly due to excessive rain weather conditions. The average flow amounted to 140% of the design flow. The design sludge concentration was 10 kg MLSS/m^3. Due to technical issues and sludge de-watering tests at semi-full scale, especially during phase 1 variations between 8 and 13 kg MLSS/m^3 occurred.

The process conditions during the research periods differed significantly as shown in Table 25. Besides the large influent flow variations also the process temperature and the biological loading changed. The ferric dosing ratio during the simultaneous precipitation period was relatively low.

9.3.2 Results

The main biological parameters are presented graphically in Figure 78 to Figure 80. In Table 26 the average feed and permeate concentrations are summarised.

COD removal

The COD-effluent concentration was relatively low due to the fact that the non-soluble fraction was completely removed by the membrane filtration. The low COD effluent concentration in phase 2 was due to the dilution effect under rain weather conditions. The average COD removal of the MBR system was > 93% for all phases.

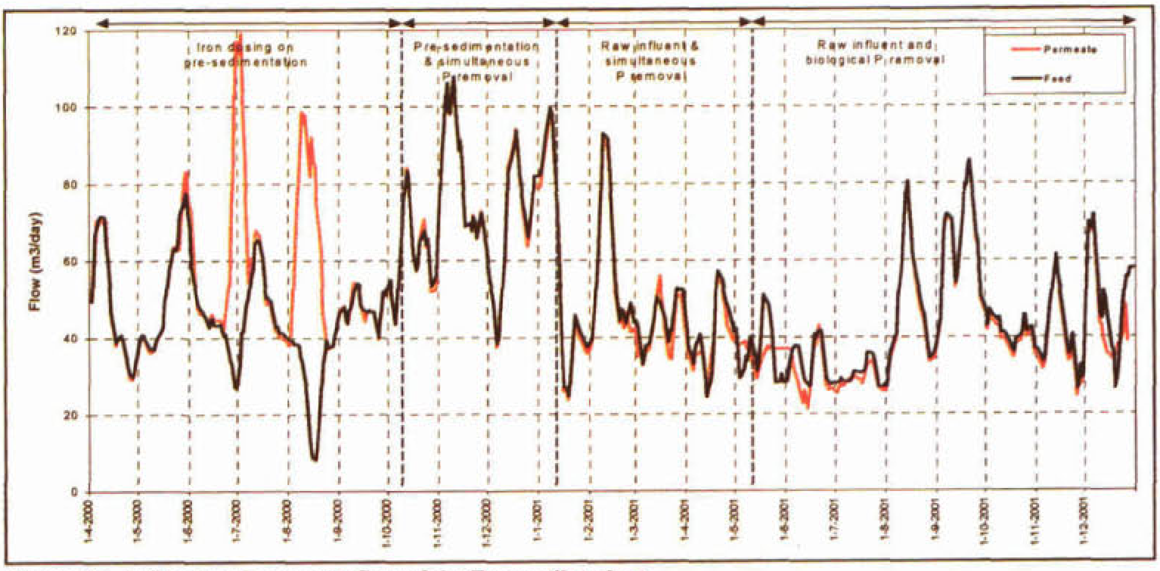


Figure 75 - Influent and permeate flow of the Zenon pilot plant

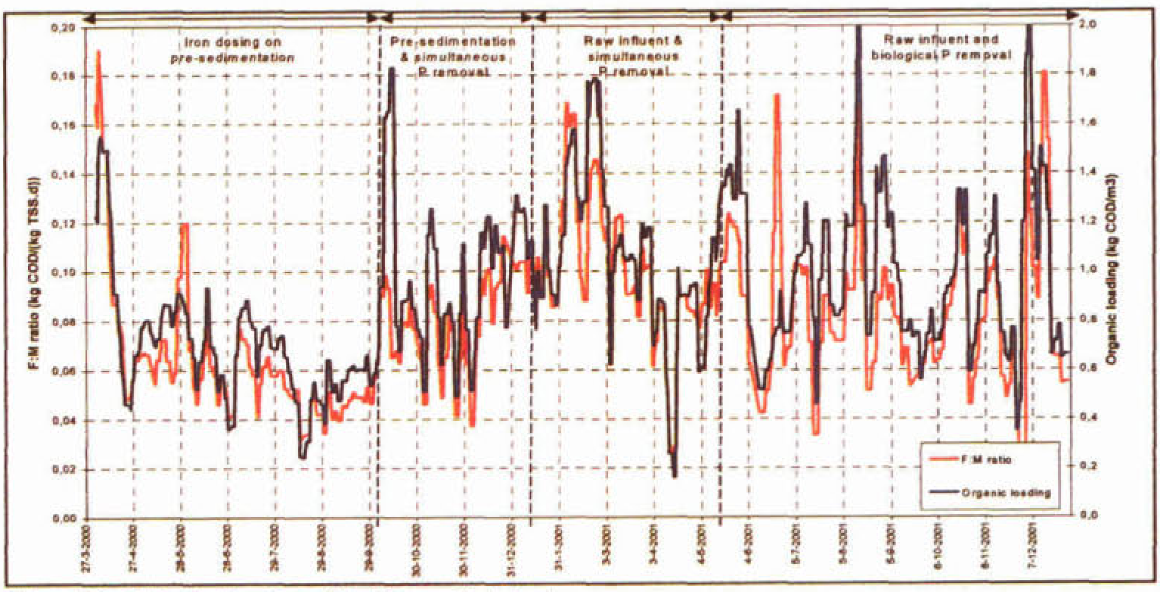


Figure 76 - Biological loading of the Zenon pilot plant

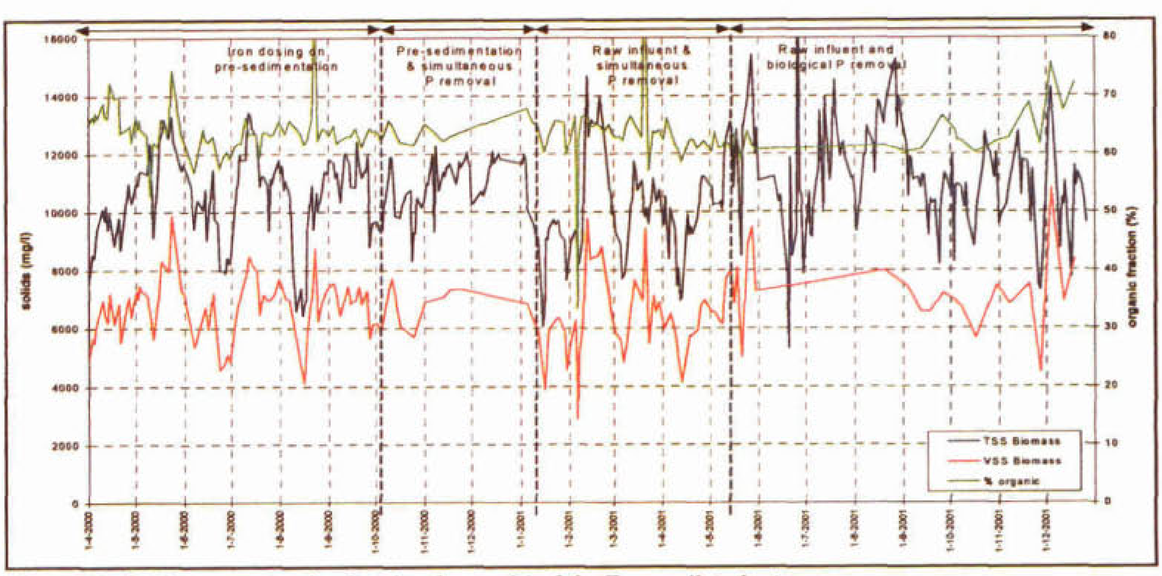


Figure 77 - Sludge concentration (total and organic) of the Zenon pilot plant

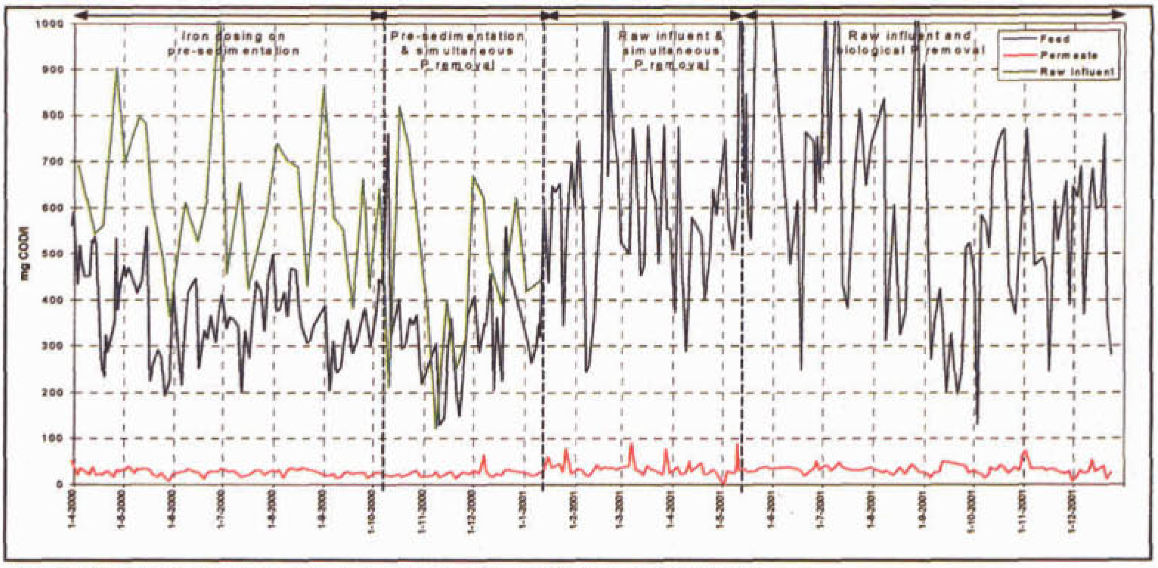


Figure 78 - COD feed and permeate concentration of the Zenon pilot plant

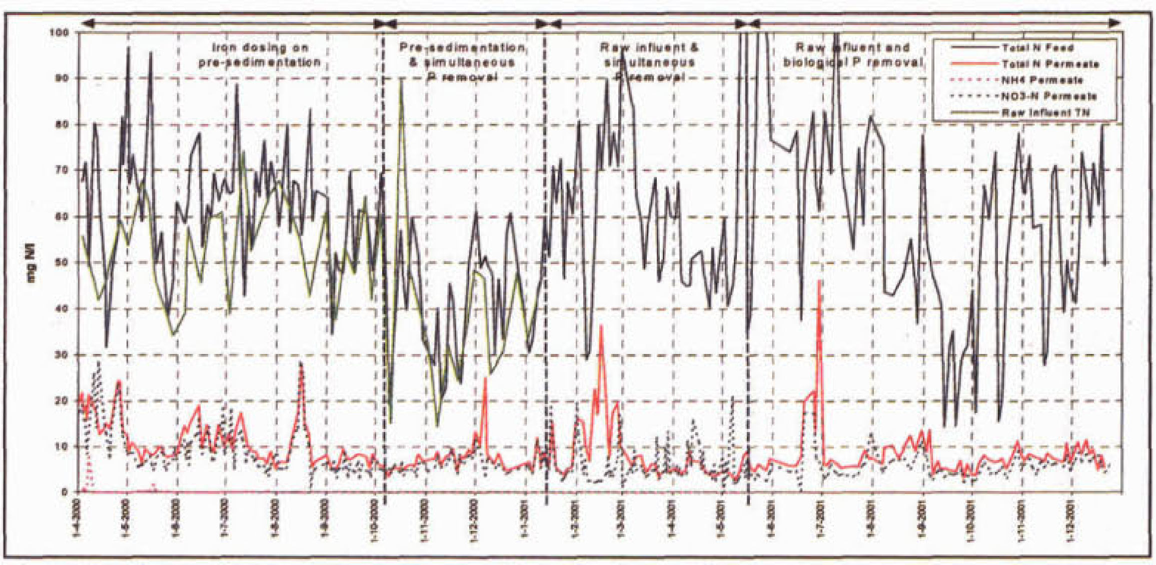


Figure 79 - Nitrogen feed and permeate concentration of the Zenon pilot plant

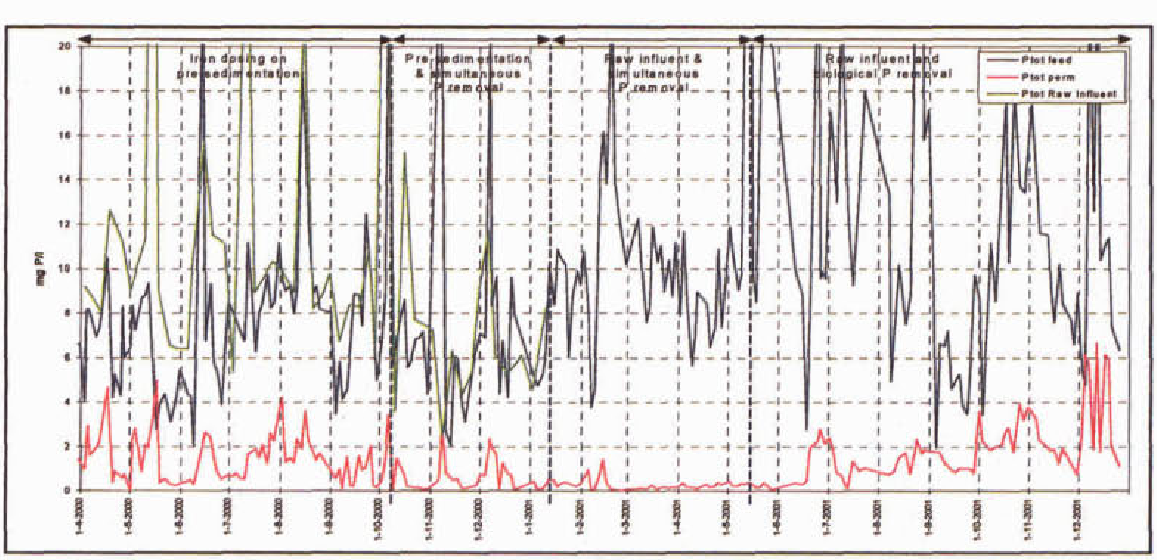


Figure 80 - Phosphorus feed and permeate concentration of the Zenon pilot plant

Table 26 - Feed and permeate concentrations

parameter		unit	phase 1	phase 2	phase 3	phase 4
COD	feed	mg/l	422	314	576	605
	permeate	mg/l	31	23	35	33
	efficiency	%	93	93	94	95
N_{kj}	feed	mg/l	67	39	58	59
	permeate	mg/l	2.8	1.2	4.2	2.7
NO_3-N	permeate	mg/l	10.2	6.1	5.3	5.8
N_{totaal}	permeate	mg/l	13.1	7.3	9.5	8.5
	efficiency	%	80	81	84	86
P_{total}	feed	mg/l	7.1	7.6	9.9	12.0
	permeate	mg/l	1.5	0.7	0.3	1.9
	efficiency	%	80	90	97	84

Nitrogen removal

In phase 1 the nitrate concentration in the permeate was relatively high with an average concentration above 10 mg/l due to the pre-precipitation process. The COD/N ratio of the feed was 6 which was relatively low. In phase 2 this ratio increased to 8 and consequently nitrogen removal improved. In this respect also the RWF dilution effect must be taken into account.

In phase 3 (February 2001) the sludge concentration increased to around 13 kg MLSS/m^3. Visual inspection showed a white air layer on the top of the aeration tank, indicating a decreasing solubility of air. This led to a longer period of erroneous oxygen measurement and a strong reduction of aeration input in the aeration tank. As a consequence the nitrification process was temporarily disturbed and the N_{kj} effluent concentration increased to 9 mg/l. In phase 4 the permeate nitrogen concentration was relatively stable.

Phosphorus removal

In phase 1 it was not possible to achieve a phosphorus effluent concentration below 1 mg P_{total}/l at a Me/P dosing ratio of 1.2. Both the removal efficiency of the primary clarifiers and the sludge production were too low. In phase 2 and 3 the phosphorus effluent concentration was considerably below 1 mg P_{total}/l and coupled to a relatively low dosing ratio. The biological phosphorus uptake of the sludge was relatively high, which indicates the possibility of enhanced biological phosphorus removal. This was demonstrated in phase 4, in which period no chemical dosing occurred and the phosphorus permeate concentration increased to 1,9 mg P_{total}/l. Nevertheless, the biological phosphorus uptake in this period was >4% (weight) on average.

9.3.3 *Sludge characteristics*

The main sludge characteristics are presented in Table 27.

The DSVI was relatively stable at a level of 100 to 120 ml/g. Both the CST and the Y-flow can be related to the sludge concentration. The viscosity of the MBR sludge varied from 5-12 mPa.s, which was relatively low for the applied sludge concentrations. The sludge viscosity seemed to be related to the sludge concentration. At lower sludge concentrations viscosity decreased even further to 2 mPa.s.

The α-factor with bubble aeration was seen to vary between 0.24 and 0.77 at a sludge concentration of 5 to 15 kg MLSS/m^3. The lowest value was measured during the low temperature tests and was presumably due to the reduced feed loading at that time. The average α-factor range at 10.5 kg MLSS/m^3 was between 0.4 and 0.6. With a surface aerator higher values were achieved in phase 1 and 2. In phase 3 the values measured with a surface aerator were 10% lower than measured with bubble aeration.

Table 27 - Sludge characteristics of the Zenon pilot plant

	unit	phase 1	phase 2	phase 3	phase 4
Sludge characteristics					
DSVI	ml/g	110	120	110	100
CST	s	90	100	70	50
Y-flow	s	370	690	n.a.	120
Viscosity					
viscosity value	mPa.s	8.2	9.1	8.7	7.6
shear rate	1/s	120	140	130	110
α-factor					
surface aeration	-	0.51	0.48	0.52	n.a.
bubble aeration	-	0.41	0.34	0.64	n.a.
Gravity thickening					
settling velocity	cm/h	4.3	3.6	4.3	n.a.
maximum concentration	%	2.3	2.7	2.4	n.a.
Mechanical thickening					
MLSS at 3900 rpm / 10 min	%	6.9	7.5	7.9	n.a.
MLSS at 1000 rpm / 3 min.	%	2.8	2.1	3.4	n.a.

The average settling velocity at the beginning of the gravity thickening test amounted to 3 - 6 cm/h, dependent on the sludge concentration. The maximum concentration after 24 hours was lower than 3%, indicating that gravity thickening was not efficient. The maximum sludge concentration with mechanical thickening amounted to 6 to 8%.

During the first part of the research the floc structure of the Zenon pilot sludge changed dramatically in time, from medium sized flocs with a compact structure to a very small floc with an open structure. Later in the research the floc became less open and more compact again. Figure 81 shows a microscopic view of the Zenon MBR sludge.

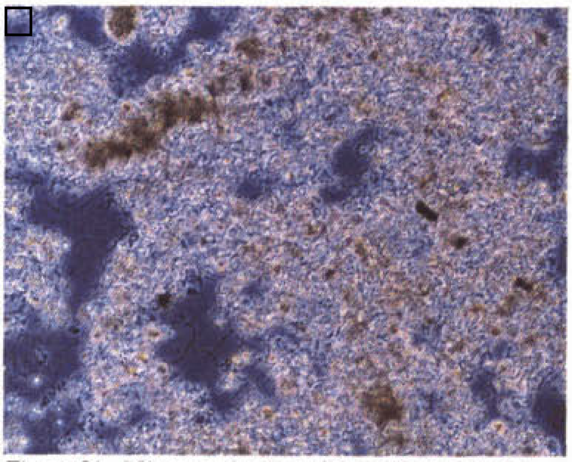


Figure 81 - Microscopic view of the activated sludge

During the start-up of the pilot plant a few filamentous organisms were present, but their number dropped near the end of phase 2. During phase 3 they displayed a slight increase again.

Since the summer of 2000 mono cultures were found, often larger than the sludge flocs. Mostly some crawling ciliates (Aspidisca) and free swimming ciliates (Euplotus) were observed. Their occurrence indicated a well aerated system with minor disturbances. Since the end of phase 2 no scale amoebas were observed.

9.4 Membrane performance ZW500a

The results of the whole research period are presented graphically in the fold out page and further explanation is given in the text. The data presented is damped to aid readability and conclusions are made regarding the membrane performance and the relevant operating and design criteria.

It was expected that the maximum achievable flux would be 41.3 l/(m^2.h) net for a period of 3 days continuous. The cleaning was estimated at twice per year and the cleaning mode was set according to Zenon's specifications of a NaOCl solution followed by a citric acid solution in an intensive cleaning procedure. The potential of maintenance cleaning was also suggested, but not incorporated in the beginning.

Back flush was used between process modes and no other recovery steps/modes were anticipated.

Periodically the Zenon membranes were back-flushed with permeate as a form of in-situ cleaning. During the research period the ratio between net and gross flux ranged between 83 to 85%.

9.4.1 Phase 1 - Primary settled wastewater and pre-precipitation

Start up and definitions

To achieve a stable biological start up the system was run at the 24 hour average daily flow rate of 40 m^3/day. The membranes however, could not run at a flow of less than 3,250 l/h due to the installed size of the permeate pump. The latter situation afforded an intermittent membrane permeation pattern based on the bioreactor level variation with only 61% operability of the membranes and consequently a regular relaxation. Related to the average daily flow rate an effective working net flux of 9 l/(m^2.h) was calculated.

The graphical fold out displays the gross flux, but during these continuous flow regimes the red line of gross flux actually represents the constant running of the permeate pump at 3,250 l/h with the relaxation time incorporated, hence net flux. All gross flux data presented up to a flux of 17.7 l/(m^2.h) gross, i.e. 3,250 l/h: 184 m^2 is actually compensated for the automatic relaxation periods. All gross flux data presented above 17.7 l/(m^2.h) is the actual gross flux as the permeate pump thereafter runs continuously without the sporadic interruptions of the relaxation modes. The latter procedure occurred automatically.

Failure air distribution system

The air distribution system of the membrane aeration failed 3 times during phase 1. In normal operation the blower aerated the two sets of modules every 10 seconds by an intermittent aeration control. During these technical failures (see Graph - AF) two of the modules were aerated continuously and the other two modules were not aerated at all, permeation however, continued.

The first failure occurred on 16/04/00 for less than one day. The hydraulic capacity of the membrane filtration system was constant (1,667 l/h) and therefore a relatively low net permeation rate. As a result of this failure the permeability dropped. After remedying the problem the system recovered quickly, without chemical cleaning.

On 26/05/00 the second failure occurred, this time for the whole weekend. During this period it also rained continuously, therefore the hydraulic loading of the membranes was high. As a result, the permeability of the membranes decreased rapidly to 100 l/(m^2.h.bar). After repairing the aeration device the permeability recovered quickly under normal operating conditions to about 260 l/(m^2.h.bar). The permeability of each set of two modules was separately measured, the continuously aerated modules were at 324 l/(m^2.h.bar) (clean) and the non aerated modules were at 134 l/(m^2.h.bar) (dirty). The non aerated modules were therefore considerably fouled.

On 15/06/00 the two non-aerated modules were cleaned in three steps. Firstly some residue sludge was washed off, then the modules were chemical flushed with chlorine (1,000 mg/l, by gravity only) and finally the modules were chemical washed with hydrochloric-acid. After this cleaning procedure the permeability of the membranes was increased to about 300 l/(m^2.h.bar). From 16/06/00 to 19/06/00, the same aeration device failure occurred for the third time. The effect of the failure was less extreme due to the relative low hydraulic loading during this weekend. The permeability of the membranes decreased from 300 to about 200-220 l/(m^2.h.bar) but recovered to 260 l/(m^2.h.bar) after the repair was made.

On 21/06/00 all the modules were cleaned in-situ (in biomass) in two steps, an extended back-pulse with chlorine (250 mg/l, 30 minutes) and an extended back-pulse with hydrochloric-acid (10% pH2.5 for 30 seconds). Due to a power break-down during cleaning, the sludge in the membrane filtration system remained at a pH of 5.5 for about 4 hours. After this manual extended back-pulse the permeability of the total membrane system increased to about 300 l/(m^2.h.bar).
To avoid the possibility of further aeration device failures the air distribution was replaced with a more heavy duty device. The latter air failures pointed out a serious risk involved with submerged membranes. The aeration of the modules is essential and must be treated with utmost respect in designing a full scale installation. The modules must never be permeated or operated in the total absence of aeration.

Peak Flux Testing
Peak flux testing was carried out to establish the maximum operating flux under the Dutch Operating Regime (up to 5:1 peaks). The objective was to run the membranes at a high and continuous flux for a period of 3 to 4 days, then observe the recovery of the permeability. During the peak test the influent pump was set to a constant flow of 1,667 l/h, resulting in the same biological loading as during the start up of the research program. The permeate was re-circulated back to the bioreactor to artificially increase the hydraulic loading of the membranes but as not to overload the biological system. To simulate the peak hydraulic loads, a permeate re-circulation pump with a capacity of 6,000 l/h was installed in the CIP tank of the pilot, this flow was controlled to balance the total hydraulic loading fed to the pilot. The total hydraulic loading was 7,667 l/h, which corresponded with a net flux of 41.7 l/(m^2.h) and a gross flux of 50 l/(m^2.h). In Figure 82 the results are graphically presented.

On Thursday 23/06/00, the first trial peak test was started. It was planned to do an 8 hour peak test. After 8 hours the pressure and permeability were relatively stable, and it was decided to increase the duration of the peak test. After 24 hours the peak test was automatically interrupted due to a low level indication in the CIP tank. The peak of 41.7 l/(m^2.h) net was not continuously maintained due to the CIP tank trip, but it could be seen that the permeability recovered to its original level of 300 l/(m^2.h.bar) after relaxing for a period of 12 hours.

For a further period of 104 hours the peak test was allowed to continue with the result that the permeability dropped from 350 to 240 l/(m^2.h.bar). Thereafter the system was set directly to proportional flow (DWF), the absence of flow peaks in this period allowed the permeability to recover back to 300 l/(m^2.h.bar) and remained stable. It should be noted that the peak test was carried out at an average process temperature of 22.5°C. The oxygen concentration in the re-circulated permeate was about 5 mg O_2/l during the peak test trial resulting in a deterioration of the nitrogen removal process.

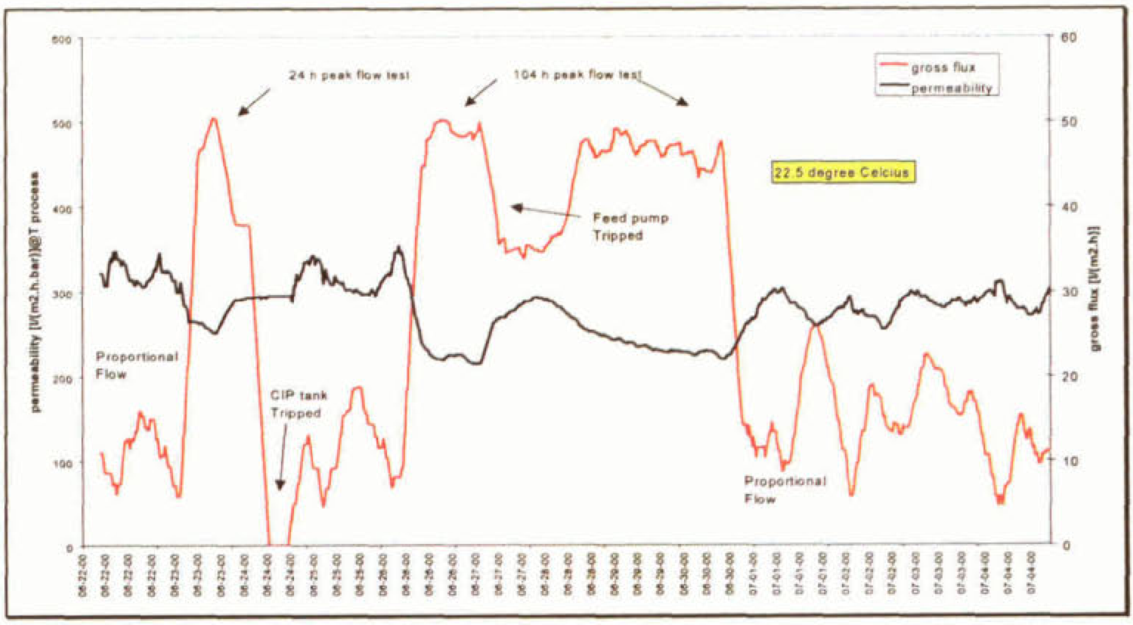


Figure 82 - Permeability and flux during peak test

The above test defined the maximum operating net flux of the membrane and proved to be sustainable for several days. The argument however, was then directed to the high operating temperature of the peak test. Was the peak flux also sustainable at lower temperatures as expected in winter conditions.

Cold Temperature Simulation and Peak Test at reduced temperature

To achieve a reliable understanding of the peak flux at lower temperatures the operating process temperature would have to be dropped to 10°C. The latter was made possible with the use of a 22 kW cooler and heat exchanger. Permeate from the CIP tank was circulated through the heat exchanger to achieve a water temperature of 10°C, however the system was very slowly lowered in temperature to avoid biological problems. The feed was set to the 24 hour daily average flow of 1,667 l/h to assist the stable temperature transition. During the cooling period the membranes ran continuously at 25 to 35 $l/(m^2.h)$ to generate enough permeate circulation for the cooler, the latter continued for 11 days to cool the bulk, thereafter the feed was stopped for two days to remove the feed heat sink and finally achieved a temperature approaching 10°C, ready for the peak test.

The peak test carried out on 14/08/00 lasted 17 hours before the system tripped out at a too high trans-membrane pressure. It was clear that the membranes could quite easily run continuously at fluxes averaging 30 $l/(m^2.h)$ gross flux at temperatures ranging from 11 to 24°C, but the peak flux of 41.7 $l/(m^2.h)$ was in some question. During the cooling test the permeability was seen to decrease from 250 $l/(m^2.h.bar)$ to as low as 60 $l/(m^2.h.bar)$. Once the temperature and the normal proportional DWF was re-established the permeability recovered to 175 $l/(m^2.h.bar)$ at 23°C. The failure to fully recover suggested the membranes were fouled.

The membranes were inspected on 18/08/00 and intensively cleaned. The membranes showed signs of extensive sludging thus a reduction in the available area for permeation. The membranes were cleaned according to standard procedures for intensive cleaning with 1,000 mg/l NaOCl at pH 10.5 followed by 3 g/l citric acid at pH 2.5. The contact time was in effect to short and the recovery achieved was only to 280 $l/(m^2.h.bar)$, whereas 350 $l/(m^2.h.bar)$ had been expected.

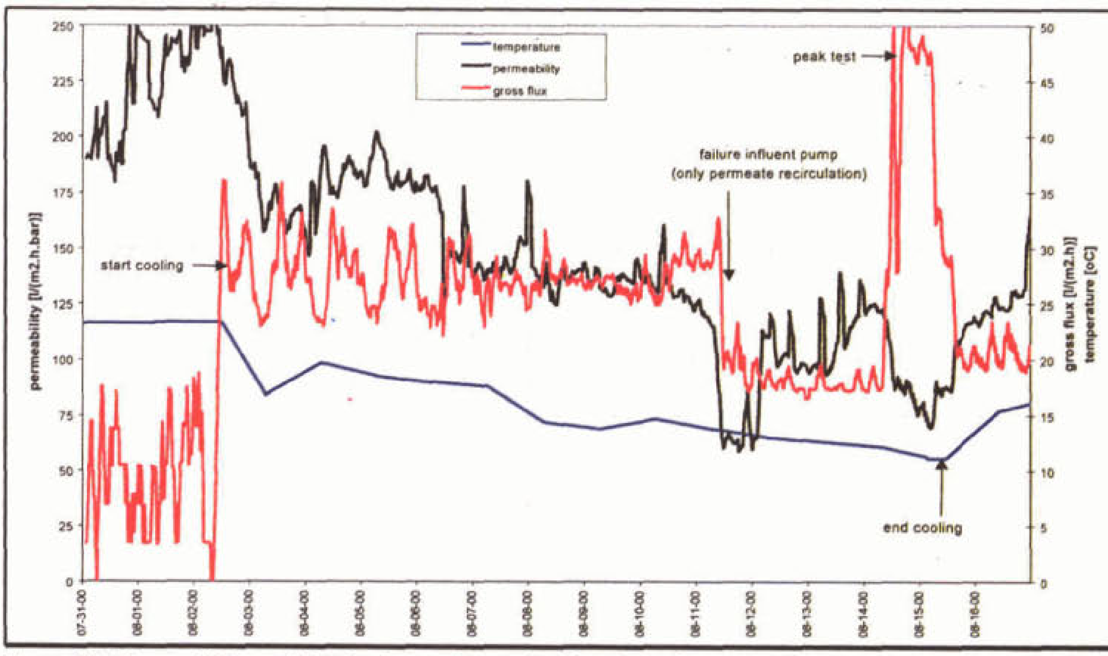


Figure 83 - Permeability and flux during cooling and cold peak testing

The last period of phase 1 leading into phase 2 saw huge amounts of rainfall, a decrease in average process temperature and a stepwise rapid decrease in membrane permeability. No recovery was possible before the next peak RWF attacked. The membranes suggested limitations to successive peak fluxes.

9.4.2 *Phase 2 - Primary settled wastewater and simultaneous precipitation*

The chemical dosing to the pilot plant was changed from pre-precipitation to simultaneous precipitation on 4 October 2000. The performance of the membrane system for that period is presented in Figure 84.

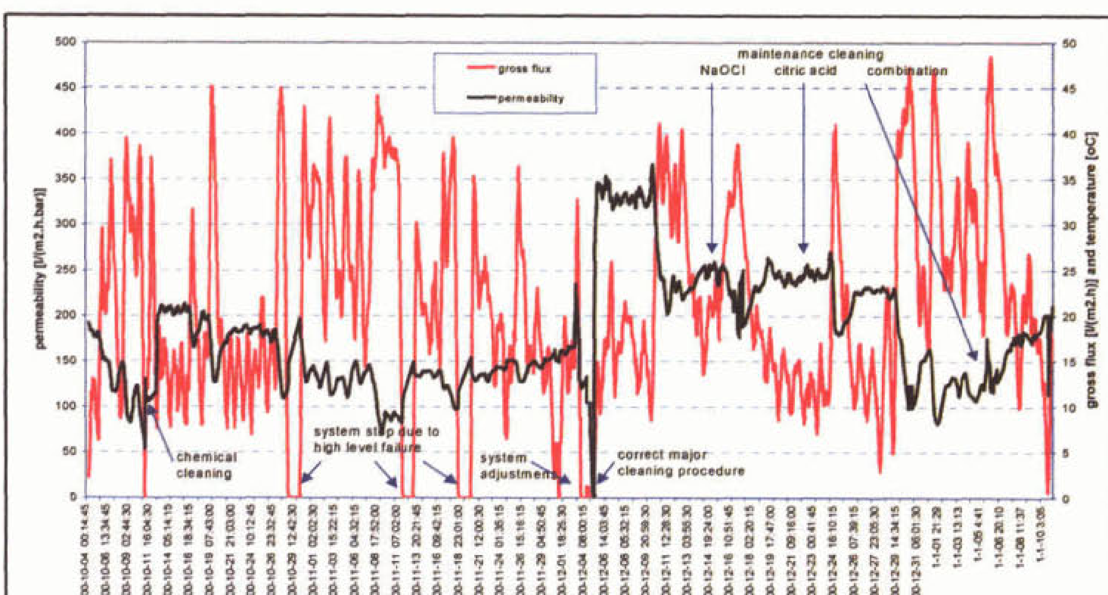


Figure 84 - Permeability and flux in phase 2

During the period 18/08/00 to 03/12/00 the system received numerous RWF peaks where the membrane system was forced to run at its maximum design flux, the average temperature for this period was also low at 12°C, but ranged from 20 to 10°C. It was seen that the permeability reacted in a step wise manner which, if not given enough time to recover, continued in a steady downwards trend, eventually tripping the system out on the maximum allowable TMP.

The system was also cleaned for the second time using the intensive cleaning procedure, again the module was sludged and the cleaning procedure appeared once again to be relatively ineffective at restoring the original permeability. This time a permeability of only 210 $l/(m^2.h.bar)$ was achieved.

It was concluded that the Zenon membrane system had difficulties in handling peak fluxes if not allowed to sufficiently recover. After a peak flux, followed by yet more peak fluxes, the permeability decreased without recovery, as a result the permeability decrease in time was relatively rapid. This was probably due to the system configuration, which caused sludging of the modules. The configuration of the membrane section was therefore optimised and modified in the first week of December 2000. The adjustments to the system referred to the introduction of a maintenance cleaning and technical changes to improve hydraulic patterns and reducing sludge concentration in the membrane tank. In this period both recovery (Intensive Clean) and maintenance cleanings were executed. The results of these cleaning procedures are described in this section.

Intensive and maintenance cleaning

On 5 December 2000 a regular major chemical cleaning was executed with NaOCl and citric acid. As a result the permeability increased from around 150 to 320-350 $l/(m^2.h.bar)$, a level which was expected for Zenon membranes. The maintenance clean was required at an interval of 1x per week. The procedure was carried out in air, based on previous non-effective in-situ results. This air cleaning was easy to perform and less chemical was wasted into the biomass. The first MC was carried out with only NaOCl and the second MC only with citric acid. The independent use of the chemicals in the MC procedure had little to no effect. The 3rd and all subsequent MC procedures were carried out with a combination of both chemicals and seem to be effective. The effect of the maintenance cleaning is presented below in Figure 85.

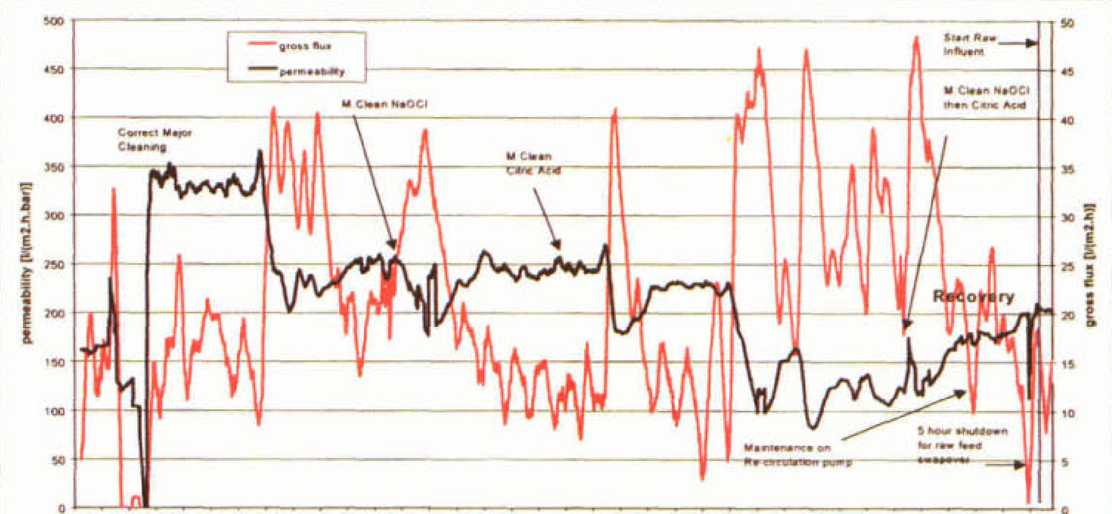


Figure 85 - Permeability and flux during the maintenance cleaning procedures

During the Christmas period of 2000, snow melted causing huge RWF peaks for a 12 day period at 12 to 10°C, the membranes in combination with the MC procedure treated the peaks with no problem with a full recovery in permeability in the subsequent 15 days of cold DWF. The maximum peak flux seen was 46 l/(m^2.h) gross.

9.4.3 *Phase 3 - Raw Influent and simultaneous precipitation*

The system feed was changed as of 10/01/2001 and fully operational on 11/01/2001. The design maximum flux was adjusted to 35 l/(m^2.h) net as suggested by Zenon. After a problematic swap over to raw feed caused by the huge amounts of screenings being produced and feed filter screenings collection capacity. The pilot was set back to proportional flow after a 7 day period at 1,667 l/h.

During the beginning of February 2001 The Netherlands was blanketed in snow which rapidly melted. This melt water caused a very low waste water temperature of 7°C and associated gross fluxes of 35 l/(m^2.h). The membrane, which was receiving maintenance cleaning every week, was absolutely clean at a permeability of 280 l/(m^2.h.bar) at 12°C. The combination of high flux and low temperature caused an extreme decline in process permeability, eventually causing the membrane system to almost trip out on high TMP. The peak was caught in time and the maximum flux adjusted to 30 l/(m^2.h) net which caused the permeability to stabilise around 100 l/(m^2.h.bar). The maximum peak flux was further optimised to 32 l/(m^2.h) net at 7°C. Once the peak subsided the MC procedures recovered the permeability back to the original 280 l/(m^2.h.bar) at 12°C.

9.4.4 *Phase 4 - Raw Influent with biological phosphorus removal*

The system was allowed to run under normal proportional flow conditions for the remaining period of the study, this was in order to ascertain the most optimum MC procedure and get an insight into the life expectancy of the module. In May 2001 the iron dosing was stopped and the biological system allowed to perform without chemical phosphorus removal. The latter yielded a surprisingly high level of P removal under biological conditions, where the DN zone was probably operating as a partially anaerobic zone (nitrate and DO concentrations were low in the circulation stream). The absence of iron salts had no affect on the performance of the membranes.

In October 2001 the MC for the ZW500a modules was changed, reflecting the general trend in chemical use and the EOX discharge requirements set in other European countries. The performance of the membrane suggested no detrimental effect regarding alternative chemical usage. In February 2002 the ZW500a module was cleaned with an Intensive procedure according to the standard supplier procedure for cleaning. The permeability was seen to return to the original process permeability as at the beginning of the pilot trial. The life expectancy of the module could not be assessed within the duration of the research.

From on November 2001 the DS in the circulation to the membrane zone was measured using on-line instrumentation – Modified Turbidity measurement from E&H. The return sludge from the membrane zone was also measured. The data proved to be reliable to the extent that the surplus sludge pump could be directly coupled to the bioreactor sludge DS measurement. This was positive regarding the rapid build up of solids in the relatively small MBR volume.

9.4.5 *General Points*

Mode frequency and efficiency

Despite the changing of the Back-flush frequency during phase 2 no positive effects can be concluded from the data collected. This area of potential optimisation must be addressed further. Currently the system is optimised for a recovery of 84% and there are no arguments to suggest that this could not be further increased to 90% with careful process optimisation.

Temperature Effect and Permeability

As mentioned elsewhere the permeability has not been corrected for temperature back to a standard reference temperature. Figure 86 depicts the difference between actual measured permeability and temperature corrected permeability for a reference temperature of 15°C.

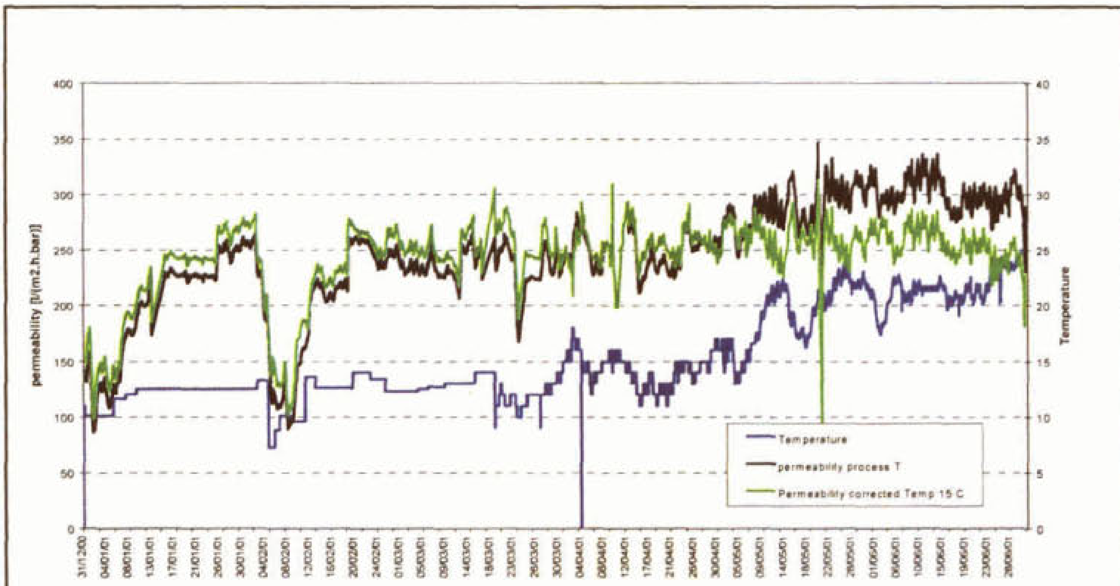


Figure 86 - Permeability and temperature correction

The corrected permeability to 15°C (the average wastewater temperature at Beverwijk) shows no major deviation from the actual measured value based permeability at the process temperature. The above, in combination with the relatively low measured viscosity of the biomass, suggests that the viscosity effect of water has little to do with the decreases in permeability at lower temperatures. This leads to a different methodology for permeability decrease in a Zenon MBR. The curves suggest a rapid transition over to cake filtration at increasing flux rates. The back-flushing of the module goes some way in re-establishing the water layer on the membrane but the MC procedure recovers the membrane surface for a better more stable water layer, ready for filtration.

9.5 Conclusions (Biology and ZW500a)

The biological performance of the Zenon pilot system was in accordance with the theoretical design calculations. The COD removal was constant at high levels, nitrogen removal was as expected, except for phase 3, which was explicable to oxygen input problems. Simultaneous phosphorus removal gave better effluent quality than pre-precipitation, even at very low ferric dosing ratios. It was a highly probable that biological P-removal occurred.

The sludge concentration in the aeration tank was kept at a concentration of 10-11 kg MLSS/m^3, to avoid problems in the membrane zone and optimisation of the energy consumption due to an optimal α-factor. This factor was as expected from literature.

A peak flux of 41.3 l/(m^2.h) net was achieved for the required operating period of more than 3 days continuous (104 hours proven).
At reduced temperature, 10°C, the peak flux of 41.3 l/(m^2.h) net was achieved for 17 hours before the maximum specified TMP was reached. At lower temperatures of 5 to 10°C the maximum design flux should be lowered to a more operable 30 to 32 l/(m^2.h) net.
Use of the routine MC procedure in combination with better filtration tank hydro-dynamics has virtually eliminated the known sludging problem of the ZW500a module.

The expected intensive cleaning of the module twice per year was proven to be pessimistic. The use of the routine maintenance clean in air suggested that the intensive cleaning procedure was reduced to 1x per year at worst.

9.6 Membrane Performance (ZW500c - "octopus")

During the course of the MBR research programme at Beverwijk WWTP it came to light that Zenon would discontinue the production of the ZW500a module in favour of the new ZW500c module or "octopus". All full scale municipal systems after January 2001 would be fitted with the ZW500c module - this module however, was not tested for the full scale MBR installation anticipated for Beverwijk.

Zenon claimed:
- The new module displayed negligible sludging as compared to the ZW500a module
- The module was manufactured using machines able to supply a better quality, the membrane remained unchanged (supported fibre).
- Air penetration through the fibres was increased, thus reducing concentration pockets whilst also decreasing the total air requirement
- The fibre packing density was increased, reducing footprint of installed modules for a given membrane area.

The construction and operation of the module was discussed and a new pilot was placed alongside the existing ZW500a pilot. The pilot took sludge from the biology and returned this sludge via an overflow back to the biology, all permeate produced was returned to the biology, thus creating an off-line testing facility irrespective of the proportion flow of the ZW500a pilot. The following objectives were defined:
1. The new module would be tested at various sludge re-circulation rates to increase the possibility of sludging, combined with peak flux testing.
2. The module would be ran with all optimisations from the ZW500a module, including: air-cycling, back-flushing and routine maintenance cleaning. (NB the ZW500c module has only one permeate header as compared to two headers on the ZW500a module). The ZW500a module would also be optimised as not to sludge.
3. If item 1 was successful the module would be tested for peak flux.
4. If item 3 was satisfactory, the chemical consumption of the module during MC procedures would be optimised.

Overall the objectives were to ascertain the utmost limits of the new module in a short time frame. The objectives were tested in a period of 4 months, until the end of the main stream research programme. The results are displayed in Figure 87.

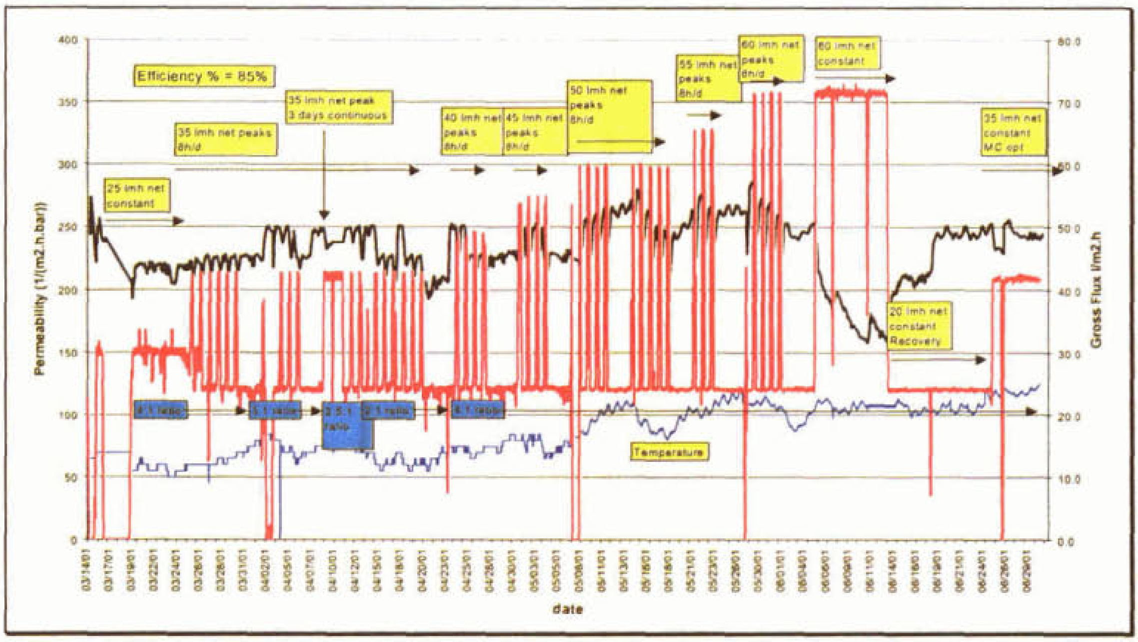


Figure 87 - Research programme ZW500c

During study items 1 through 3 the MC procedure remained unchanged routine. No peripheral parameters were adjusted on the module, e.g. efficiency, process time, back-flush time, air flow, etc. One major difference between the ZW500c and ZW500a is the base line clean process permeability. Due to the single permeate header of the ZW500c module the permeate transport system generates a small additional pressure drop in comparison to the two headed ZW500a module. The difference amounts to a lower clean process permeability in the ZW500c module of 50 l/(m^2.h.bar) and ranges from 225 to 270 l/(m^2.h.bar), whereas the ZW500a module ranges from 280 to 325 l/(m^2.h.bar).

Study Item 1 - sludging

After a bedding-in period of two weeks at a constant flux of 25 l/(m^2.h) net and a sludge circulation ratio of 4:1 (based on net permeate flow) the sludging test commenced. An 8 hour 35 l/(m^2.h) net peak was set each day followed by a constant flux night-time operation at 20 l/(m^2.h) net to facilitate recovery. At all times the circulation ratio was fixed according to the required weekly ratio. Figure 87 displays 5 distinct sludge/permeate ratios. The ratio of 4:1 was tested first as this was the design basis for Zenon municipal MBR's. This was followed by 3:1, 2.5:1 and 2:1, before being returned to 4:1.

The ratios, 4:1, 3:1, and 2.5:1 yielded no sludging deviation in the permeability curve and the trend followed the temperature profile. At these ratios the peaks and recovery periods were easily processed. The older ZW500a module expressed an onset of sludging at ratios of less than 3.5:1. The final ratio of 2:1 suggested that the module was beginning to foul with a minor loss of process permeability that did not follow the temperature profile, but displayed characteristics of cake filtration. The return to a ratio of 4:1, accompanied by the routine MC procedure returned the membrane to its original process permeability.

The module was subsequently removed from the filtration tank and inspected - negligible sludge build up was seen. It appeared that sludge had been building up but was dislodged by the air and moveable bottom header. Considerable amounts of sludge had fallen out of the module in the 5 week test and had settled on the bottom of the tank. This sludge most definitely originated from the membrane due to the fibre impressions in most of the sludge debris.

It was concluded that the ZW500c module was a significant improvement on the ZW500a module regarding sludging. The ZW500c module, under the operating environment tested did not sludge up, but transported the sludging generated to the bottom of the tank. The latter implied that at full scale a sludge removal system should be built under the membranes. The phenomena of sludging and sludge deposition from the module gave rise to how much and how fast. These items would be addressed in the following research programme, but until such time, The ZW500c module should not be operated at ratios of less than 2.5:1.

Study item 2 - optimisation

The operability of the ZW500c module proved to be exactly the same as its predecessor. The single permeate header yielded no apparent disadvantage in the beginning despite the offset in base-line process permeability. Due to the inability to sludge-up more membrane surface area was available for filtration, thus reducing the sharp decrease in permeability at the onset of peak fluxes.

Since the incorporation of the routine MC procedure on the ZW500a pilot the membrane had not sludged and will be left running to ascertain the long term effect of MC operation. The ZW500c module was also run with weekly MC procedures.

Study item 3 - peak testing

The objective here was to keep the sludge circulation ratio constant at the design basis of 4:1 and steadily increase the net flux extracted. Based on the success of the consecutive peak tests of 8 hours per day over one week, the flux would be increased for the following week by 5 $l/(m^2.h)$. As in study item 1 the 8 hour daily peak was followed by a recovery step at 20 $l/(m^2.h)$. The test was started with a net flux of 40 $l/(m^2.h)$ and increased through 45, 50, 55 and 60 $l/(m^2.h)$ net. At this stage the peak flux could no longer be increased due the permeate pump capacity. In general the 8 hour peak fluxes saw the typical step decline in permeability but always recovered to the original permeability after the recovery period. As in study item 1 the permeability trend followed the temperature profile, as shown in figure 84.

The 8 hour peak flux testing was followed by a week of recovery at 20 $l/(m^2/h)$ and recovery was seen. This was immediately followed by a continuous peak test at 60 $l/(m^2.h)$ net for as long as possible. The system continued for 9 days, including two MC procedures (routine) clearly seen in the permeability profile. The membrane never reached its maximum trans-membrane pressure but the back-flush pressure reached its maximum safety pressure, noting that the back-flush flow was 125% of the permeate flow. The membrane was inspected after the continuous peak flux and the same phenomena of sludge deposition was seen under the module the module itself had less sludge entrainment as compared to study item 1, suggesting that sludge circulation was more critical regarding sludging than higher fluxes.

The module was allowed to recover for two weeks and the recovery was substantial, each successive MC procedure returned more permeability.

Study Item 4 - MC chemical optimisation

The membrane was set at 35 $l/(m^2.h)$ net continuous permeate extraction and a sludge circulation rate of 4:1. The idea was to slowly reduce the NaOCl MC step from 6 to 1 thus reducing the chemical usage in steps of 17%. The citric acid step was also reduced in time from 6 to 2 steps, thus reducing the citric acid consumption by 66%. The optimised cleaning was set at 4 NaOCl steps followed by 2 citric acid steps. The latter was maintained for the rest of the study. Figure 88 displays the remainder of the ZW500c membrane performance.

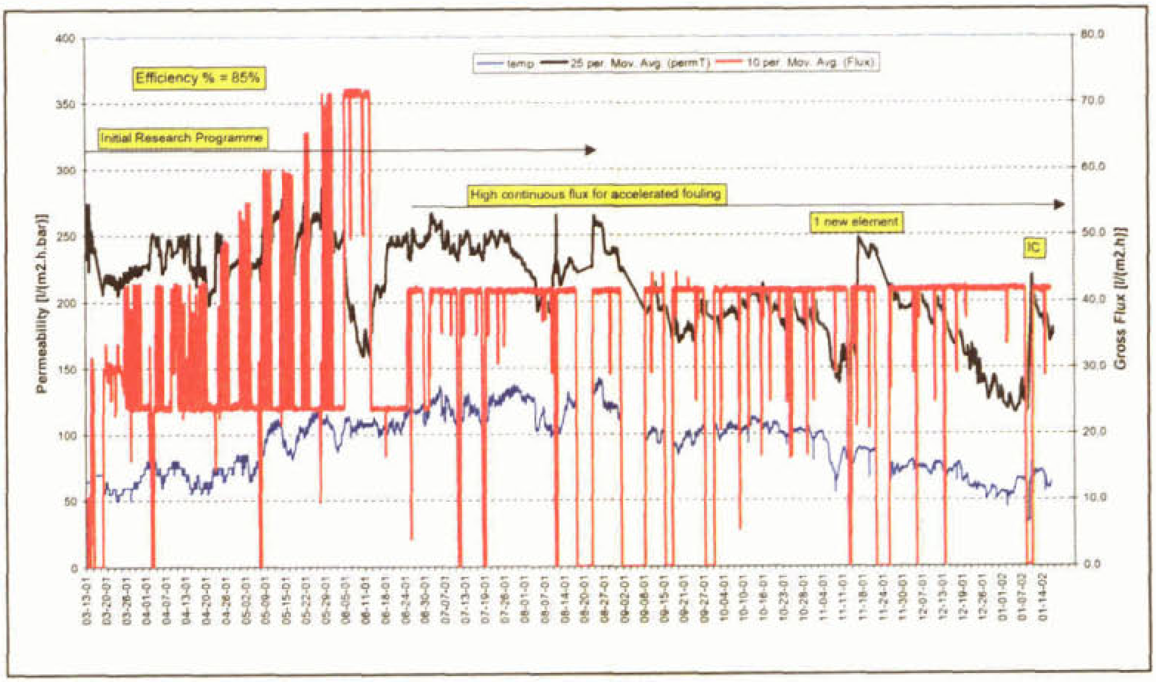


Figure 88 - ZW500c membrane performance until January 2002

Bearing in mind that the net flux was continuously high, it was expected that the permeability would decline at an increased rate. At operating temperatures of 20°C and above the high flux had little effect on the permeability decline. As the process temperature decreased the permeability became difficult to re-establish with the weekly MC procedure. Eventually it was concluded that the membrane was fouling according to pore fouling, bio-fouling or scaling, rather than the typical cake-formation, and the MC procedure had little to no effect on the recovery. One of the three ZW500c elements was removed and returned to Zenon Canada where it would be investigated thoroughly. A new element was set in its place. The single clean element and the remaining two 'dirty' elements continued to foul rapidly (most work being done by the new element), and accelerated due to the lower operating temperatures associated with the winter period. Within 6 weeks the ZW500c was fouling at such a rate that an Intensive Cleaning procedure was necessary. Zenon's research on the removed element yielded an intensive cleaning procedure that should recover the membrane completely. This was carried out in the pilot in week 2 of 2002. The cleaning procedure was of limited success and the membranes were recovered to only 82% of the original process permeability. The period after the IC procedure saw cold process temperatures where the MC procedure did little to help the permeability decline, the decline remained rapid.

The above phenomena could not be related to the membrane material itself or the process conditions around the module, as the ZW500a module experienced none of the permeability declines of the ZW500c. Under cold processing conditions the ZW500a modules were recovering with a simple MC procedure and at no stage put the processing of water at risk. The ZW500c module performance in comparison to the ZW500a module was less reliable, but the water processing was also never at risk.

Despite the claims of the supplier regarding the single header permeate extraction on the ZW500c module, the results express a clear detrimental performance of the ZW500c module in direct relation to the double headed ZW500a module (more so at lower processing temperatures). The latter is not expressed in flux performance but more in terms of recoverability and process reliability. It must be stressed that the ZW500c module was pushed extremely hard regarding flux rate, but this would only act to increase the onset of fouling that the module would have experienced over a much longer period of time at a lower flux. The ZW500c module requires further investigation and optimisation before it achieves an equivalent status of the ZW500a module.

The ZW500c module as a concept is a good idea. Regarding sludging and peak flux it out performs it predecessor. However, the two header configuration for permeate extraction may have been better suited for reliable process operation on the longer term.

9.7 Conclusions ZW500c

The ZW500c showed negligible sludging within the module itself but tended to deposit the sludge blocks under the module. The sludging effect is still present but is shifted to a simple mechanical removal of the sludge at the bottom of the tank. The sludge build up and deposition rate must still be defined. Sludge circulation ratios of greater than 2.5:1 (based on net permeate flow) would be advisable.

The maximum short term peak flux was not established due to limitations of the pilot, however an 8 hour peak flux of 60 $l/(m^2.h)$ net was successfully achieved. The latter raised questions regarding the maximum design flux for the Dutch situation, this will be established.

A maximum continuous flux of 60 $l/(m^2.h)$ net was achieved for 9 days before the pilot limited further progression. The latter raised questions regarding design fluxes for continuous flow installations or fixed flow compartmentalised municipal systems.

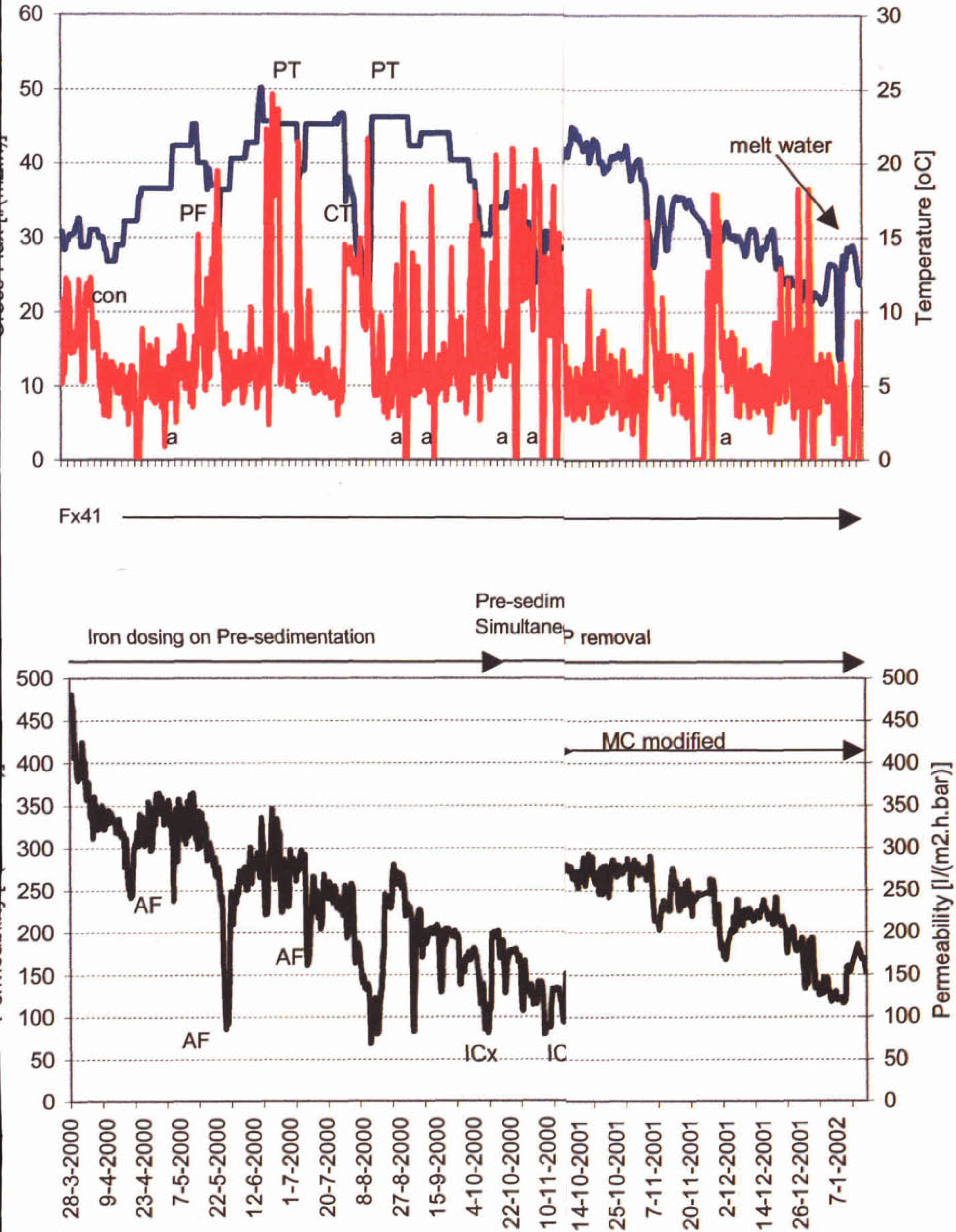
Chemical reduction in the MC in air procedure regarding NaOCl consumption yielded possible reductions of up to 50%.

The ZW500c module appeared more temperature sensitive than the ZW500a module. The ZW500c as a result of lower operating temperatures was more difficult to recover with MC procedures and only achieved 82% of it's original permeability after an intensive cleaning procedure. The permeability decreased rapidly under cold processing conditions whereas the ZW500a module could be recovered.

The ZW500c module displayed many improvements over the ZW500a module regarding peak fluxes and the lack of sludging. However, the reduction in permeate headers from two to one may have adversely effected the permeation characteristic of this module configuration. Serious thought must be put into the concept of a ZW500c module with two permeate headers in order to gain the advantages of both module types.

ZENON Fold-out Graphic ZW500a

PF	Transition to Proportional flow (direct signal STP Beverwijk)
con	Transition to Constant flow
PT	Peak Flux test period at suppliers maximum flux specification
CT	Artificially induced Cold Test using 22 kW cooler.
RB1	Rebuild 1, filtration zone modifications for improved mixing
RB2	Feed pumps set to raw influent and feed filter set up modified
RB3	Screenings collection and filter position improved
a	Alarm, system tripped automatically due to circumstances
AF	Membrane module Aeration system failure
MC	Standard Maintenance Cleaning Procedure according to ZENON specification
MC1	Maintenance Cleaning Procedure with NaOCl only
MC2	Maintenance Cleaning Procedure with citric acid only
IC	Intensive Cleaning Procedure according to ZENON specification
ICx	Intensive Cleaning Procedure incorrectly executed
Fx41	Standard maximum net Flux phase 1 and 2
Fx35	Standard maximum net Flux phase 3 and 4
Fx30	Maximum net Flux set point during extreme temperature conditions



PT
PT
melt water
PF
CT
con
a
a
a
a
a
a
Temperature [oC]
Fx41
Pre-sedim
Iron dosing on Pre-sedimentation
Simultane
P removal
MC modified
AF
AF
AF
AF
ICx
IC
Permeability [l/(m2.h.bar)]
28-3-2000
9-4-2000
23-4-2000
7-5-2000
22-5-2000
12-6-2000
1-7-2000
20-7-2000
8-8-2000
27-8-2000
15-9-2000
4-10-2000
22-10-2000
10-11-2000
14-10-2001
25-10-2001
7-11-2001
20-11-2001
2-12-2001
14-12-2001
26-12-2001
7-1-2002

Zenon

10 SIDE STUDIES

10.1 Side study 1: Pre-treatment

From experiences on several municipal MBR plants it is known that membrane systems can be very sensitive for macro-fouling by debris. The occurrence of debris has led to problems regarding throughput and eventual membrane damage. In some cases this has lead to a complete replacement of the membrane modules.

On some small scale MBR plants the effect of debris build-up in the membrane modules is accepted as a fact of life and as a result is routinely removed. For larger scale municipal MBR plants this is not an option as this would be very labour intensive. Consequently, a thorough and reliable pre-treatment of the wastewater is of vital importance for the well functioning of a MBR system. The main purpose of the pre-treatment is to remove solids (screenings) which are harmful for membranes, such as coarse solids (plastics, leaves, seeds, sand particles), oils, fats and hairs.

To prevent fouling as much as possible the MBR pilot research at Beverwijk WWTP was started in phase 1 by feeding préprecipitated wastewater. Depending on the MBR system, an influent micro- or fine-screen or an in-line sludge screen was used in that period. As the research period was extended and the pre-treatment was changed to only pre-sedimentation and even raw wastewater, it was decided to implement influent micro-screens on all MBR systems.

During four months four different micro screens were tested which were used as a pre-treatment for the MBR installations (Figure 89).

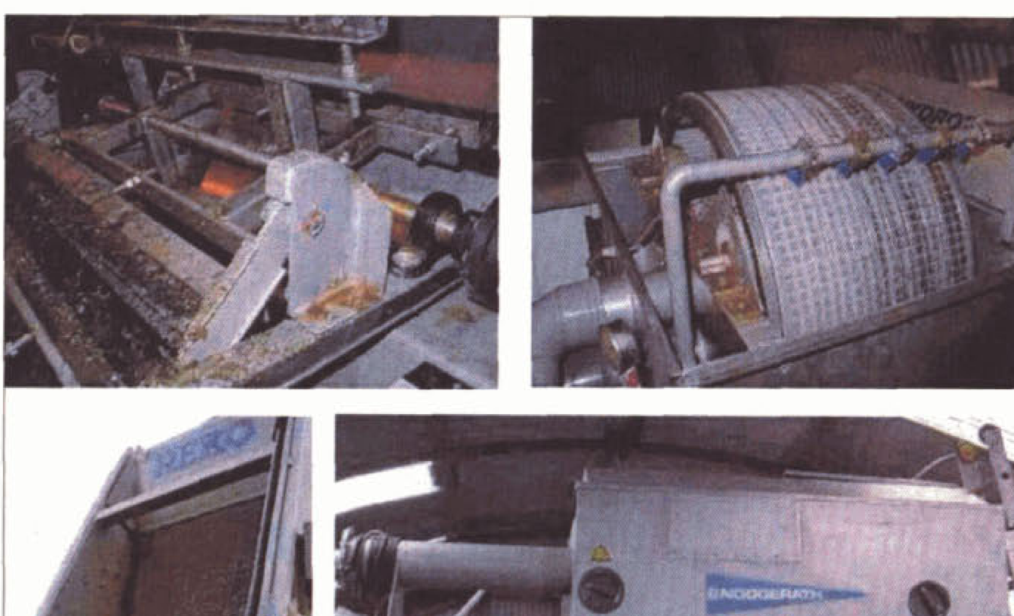


Figure 89 - Micro-screens tested in Beverwijk

The micro-screens differ in working (filtration) principle and filter mesh; a vibrating screen (0.75 mm), a cylindrical screen with rotating brush (0.75 mm), an internally fed drumfilter (0.50 mm) and an internally fed rotary wedgewire screen (0.25 and 1.0 mm). The choise of these micro-screens was based on a short literature survey and limited by the practical applicability and costs.

The aim of the side study was to investigate the performance of different types of micro-screens (including maintenance and cleaning requirements) and to quantify the removal efficiency on the relevant parameters.

The micro-screens have been tested in phase 2 and 3 of the research.

The systems show significant differences concerning the solids production, the SS and COD removal and the maintenance requirements. The most interesting results for phase 3 (raw wastewater) are presented in Table 28.

Table 28 – Micro screen performance pilot installations

parameter	unit	vibrating static screen	rotating brush raked screen	drumfilter	rotary wedgewire screen	
		(0.75 mm)	*(0.75 mm)*	*(0.5 mm)*	*(0.25 mm)*	*(1.0 mm)*
COD removal	%	12	9	29	13	-
SS removal	%	44	20	63	28	-
retained screenings	g ds/m^3	14	23	94	34	8
Fats removal	%	42	24	58	28	-
Mineral oils removal	%	68	56	49	76	-

The drum filter retained the highest amount of fats and solids, and produced the highest amount of screenings. The drum filter was the only one which was discontinuously cleaned (rinsed), resulting in a sludge layer on the filter surface and consequently an artificially obtained finer filter medium (pre-coat). Striking is that in contrast to the other screens no hair was found in the drum filter screenings.

The SS removal efficiency of the continuously cleaned systems is 30-60% in general, which is relatively high. The application of a micro screen will influence the biological performance of the system as the surplus sludge production will decrease and the sludge composition will change. The COD removal efficiency for a continuously cleaned micro-screen amounts 10-15%. The application of a micro-screen will influence the denitrification potential of the MBR system. In case of strict effluent requirements the choice of the micro screen type and mesh size should be thoroughly considered.

Some general observations were:

- the specific screenings retention for raw wastewater (phase 3) was 30-80 times bigger than for pre-sedimentated wastewater (phase 2);
- the use of a 1.0 mm filter mesh for the rotary wedgewire screen resulted in a much lower screenings production;
- from visual observations it was concluded that the screenings of the continuously cleaned screens (all except for the drum filter) had a similar composition and contained coarse organic material (leaves, seeds) and hairs.

All screens operated automatically and did not require special attention during normal operation. The maintenance requirements were different for the four systems. The rotating brush raked screen did not require any maintenance at all.

From the observations during the side study it has become clear that in case of raw wastewater the use of a micro-screen is definitely required.

10.2 Side study 2: Membrane fouling and cleaning

MBR technology has several advantages compared to traditional activated sludge processes. However, one of the main disadvantages is the necessity to routinely clean the membranes. Cleaning of the membranes, irrespective of the procedure used has a knock on effect to the membrane life-expectancy and there after operational replacement costs. Additionally the chemical usage required for cleaning purposes impacts a toxic shock on the biological system, as well as a concentrated corrosive shock for the membrane material of construction.

It was of tantamount importance to undertake an investigation into the mechanics behind the membrane operation, the types of fouling that can promote membrane performance decline, and methods to clean the membranes in the most optimal manner, both from an environmental and economical standpoint. Once investigated an optimised operating window could be defined for all relevant circumstances involved with MBR membrane operation.

Fouling within MBR systems applicable to the treatment of municipal wastewater is a complex phenomenon, mainly due to the problematic and variable composition of the raw wastewater and the suspension requiring filtration. The biological suspension (activated sludge) not only contains biological flocs, formed by the conglomeration of micro-organisms within a floc matrix, but also a whole range of soluble, insoluble and colloidal compounds, either bought in by the influent to be treated or resulting from bacterial metabolism.

In simple terms the objective of this side study was to bring the fouling and cleaning methodology under control within defined safe operational parameters. The pilots were already protected against the real MBR 'membrane killer' - hair, debris and coarse material. As the latter had been documented on numerous full scale installation. This in mind, the pre-treatment to the installations was also studied under side study 1. It was important to this study item that the most common cause of membrane performance deterioration be categorically eliminated. The only fouling permitted within the MBR configurations was that related to the biological process and the permeate extraction. This basis was the foundation to developing optimised operating criteria.

Fouling Expectations
Six fouling mechanisms were expected to occur in MBR micro- and ultra-filtration applications, and were split into two sub-categories of micro-fouling and macro-fouling:

Micro-fouling	Macro-fouling
1. Scaling 2. Bio-fouling 3. Organic fouling/adsorption 4. Pore blocking	5. Cake formation on the membrane surface 6. Feed debris, grease balls, plastic, hair, paper, etc

Each mechanism was investigated, practically and historically to understand the basis of fouling and to envisage potential solutions.

Factors affecting fouling
The nature and extent of membrane fouling was expected to be influenced by the following factors:

1. The physical-chemical nature of the membrane
    - Hydrophilic characteristics of membrane material
    - Surface topography
    - Charge of the membrane
    - Pore size
    - Surface modification
2. The physical-chemical nature of the solute
    - Sludge and effect of EPS (side study 3)
    - Water phase
3. The configuration of the system and dynamics
    - Temperature
    - Hydrodynamics
    - Pressure

The latter were tested analytically as well as practically on the pilots.

Fouling in practice
To assess the membranes for the above expected fouling phenomena the membranes were subjected to various scanning electron microscope tests, including SEM (EDX), Cryro-SEM, and CLSM. The samples included, new membranes, dirty membranes and membranes after a cleaning procedure and each sample was scanned on the surface and as a cross section. Comparisons were made of the systems and conclusions drawn.

Controlling influences and tools
In practice the most important fouling mechanism seen was cake formation. Other fouling mechanisms were of importance but remained under control via maintenance cleaning or process optimisation. To control fouling in MBR systems the focus must be on the control of cake formation. Certain factors of influence were defined with a high priority and tools were found to bring these factors under control. The most important factors associated with cake formation were:

- Debris
- Sludge
- Hydrodynamics/aeration
- MC / IC (Maintenance and Intensive Cleaning)
- Process optimisation

Each of these factors was investigated and theorised, and numerous links made in order to gain control over the onset of cake formation. This was achieved in the form of process operating windows which were clearly defined for each system.

Cleaning
Commonly a membrane is considered to be clean when a permeability is obtained of a comparable new membrane. Actually, it may not be possible to obtain the initial clean water permeability, as this permeability usually drops to a stable value after just a few runs. The importance of restoring the clean water permeability can be over emphasised, leading to an excessive (non-

economical) cleaning frenzy. The most important criteria regarding whether a cleaning is successful or not should be that the previous (clean) process permeability is restored (first few runs after installation or previous cleaning).

Based on the fouling factors identified and isolated in the previous section the membranes were subjected to a number of cleaning procedures to ascertain the type and quantity of the fouling. It should be noted that the effectiveness of a particular cleaning procedure directly relates to the fouling occurring during normal processing.

As a rule the type of fouling determines the type of cleaning required.

Types of cleaning
Two main types of cleaning can be distinguished: chemical cleaning and mechanical cleaning.

1. Chemical cleaning
Chemical cleaning is essentially a physical-chemical reaction between the cleaning chemical and the foulant. During a chemical clean the fluid mechanics, temperature and contact should be considered.

2. Mechanical cleaning
Mechanical cleaning is a term used for the physical removal of suspended solids from the membrane material. Mechanical cleaning is usually based on turbulence and fluid mechanics and in extreme cases manual participation.

<u>Turbulence and fluid mechanical cleaning</u>
- Circulation, based on a shear rate of sludge over membrane
- Air-flush/aeration, also based on a shear rate of an air-water mix
- Back-flush, a reversal of the filtrate flow
- Relaxation, shear rate at the outside of the membrane (by aeration) and no permeate production.

<u>Extreme Mechanical cleaning</u>
These procedures are only carried out when the process has failed in a major way, the chance that the membranes would be damaged is high. The procedure involves the removal of the membranes from their normal environment, and with the use of hose pipes, brushes, sponges etc the membrane surface is cleaned.

Main relations in cleaning
It can be concluded that every type of fouling that occurs on/in the membrane can be treated with simple chemicals and procedures under simple hydraulic parameters. Each chemical used can be, if necessary, substituted for other more specific cleaning chemicals compatible to the membrane requiring cleaning, e.g. detergents for removing oil, fat and grease.

In municipal MBR systems mainly bio-fouling, and cake formation occur. Bio-fouling can be treated with an NaOCl cleaning eventually followed by an acid cleaning. Cake formation can be treated hydraulically. Optimising the hydraulic cleaning was an important aspect to prevent fouling of membranes, and was directly related to the membrane configuration and process control. Considering the simplicity of the chemical and hydraulic procedures all forms of fouling for most types of wastewater can be tackled with the same degree of success as seen in the pilots.

The frequency of the IC (Intensive clean) or MC (Maintenance clean) chemical cleanings are much lower than the frequency of the hydraulic cleanings. The most important aspect is to optimise the hydraulic cleaning and use the IC or MC as the back up procedure for process integrity.

Overall cleaning and fouling evaluation of MBR Systems

The most important aspect is the operation of the MBR to prevent the forming of a cake on the membranes, followed by the removal of the cake layer and the cleaning of the membranes with chemicals.

Operation window of MBR system

Some basic rules were defined to prevent cake formation on the membranes. These rules were:

- use a membrane with a high maintainable permeability; lower velocities to the membrane surface prevent cake formation.
- reduce dead zones in the membrane module/tank, and avoid a too high packing density of the membranes.
- select responsible operating fluxes, however, during peaks temporary higher fluxes can be maintained.
- generate larger sludge flocs to allow a more permeable cake layer on the membrane.
- install and maintain good biological conditions thus preventing the formation of EPS and filamentous organisms.

If a cake layer is formed some actions can be undertaken to remove the cake layer from the membrane. The most important tools to remove the cake layer are:

- high turbulence/re-circulation of the fluid in the membrane tank, but the sludge flocs must not be macerated or stressed (This will cause a less permeable cake layer on the membrane).
- temporally more air bubbling or intermittent air can break down the cake layer on the membrane. A high cross-flow along/through the membranes will also help.
- where possible, a back flush is effective, but will only help the membrane if the frequency is high. In the first 3 - 4 minutes of a process mode the main part of the cake layer is formed.
- relaxation of the membranes is very effective to remove the cake layer. Disadvantage is that the membrane system reduces productivity.

10.3 Side study 3: Energy consumption / α-factor

The MBR can be build very compact because high sludge concentrations can be maintained. Recent research showed that oxygen transfer rate at high sludge concentrations was very low. This implies that the energy consumption of aeration will be higher in a MBR than a conventional system.

This side-study was performed to examine the mechanisms which influence the α-factor in a aeration system and to formulate the ideal setting for a MBR referring to oxygen input.

The study was split into two parts:

1. A theoretical part containing a literature study and a brainstorm;
2. A practical part containing a microscopic study, specific measurements on the pilot-installations and small experiments.

The transfer of oxygen from air into water depends on the characteristics of the transfer layer, air bubble size and mean residence time of the bubble in the water. These last two parameters depend strongly on the viscosity of the solution to be aerated. Viscosity it self can be influenced by reactor configuration and mixing/aeration device and by the concentration and the characteristics of the activated sludge.

Bacteria are able to influence the α-factor by the way they grow. Under some conditions they produce excessive amounts of extra-cellular polymeric substances (EPS). This EPS makes the sludge voluminous thereby increasing the viscosity of the sludge (see Figure 90). EPS can also have a high content of proteins thereby making the sludge hydrophobic. Some EPS is loosely attached to the floc and therefore easily excreted in the water-phase. Due to latter diversification and numerous causes it is difficult to find a common factor responsible for stimulating the production of EPS.

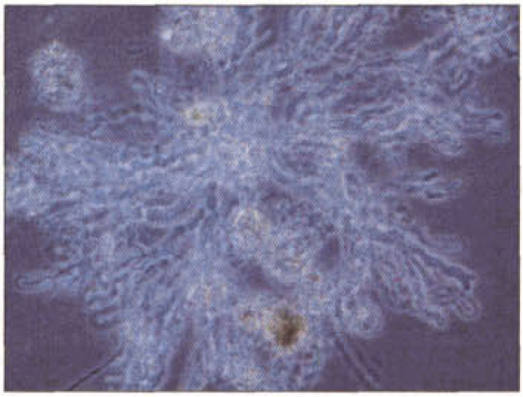


Figure 90 - EPS sludge

Growth of filamentous bacteria in the sludge will increase the sludge volume and thus increasing the viscosity. Some filamentous bacteria are known to be hydrophobic and to excrete hydrophobic compounds in the water-phase thus creating their own environment.
Large microbial products excreted into the water, soluble microbial products (SMP), will be retained by the membrane of a MBR and not washed out like in a conventional system. These compounds are then able to influence the aeration characteristics or foul the membranes.

Experiments were performed to determine if the water-phase contained compounds responsible for lowering the α-factor. They showed that there was no accumulation of compounds in the water-phase responsible for lowering the α-factor. This meant that the cause of low α-factors measured lays in the sludge concentration and sludge characteristics.
That these characteristics were closely linked to the viscosity was proven by plotting the gathered viscosity and α-factor data over a period of three months. A nearly straight relationship showed up, where the α-factor decreased at a increasing viscosity. This implies that the viscosity of the sludge in a MBR should be kept as low as possible at higher sludge concentration.
Since there were no significant differences between the mean α-factor of the 4 pilot-installations no conclusions could be drawn about the optimal sludge composition. However, since excessive growth of filamentous bacteria and EPS productions were not observed during the test period, it can not be excluded that these growth forms have a negative influence on the α-factor.
The viscosity assay shows that viscosity is strongly influenced by the mixing energy. This implies that the choice of aeration/mixing device and reactor configuration will be a very important aspect in engineering a MBR at higher sludge concentrations.

10.4 Side study 4: Effluent quality

One of the main advantages of the MBR system is the "superior" effluent quality that can be achieved. This item will be thoroughly investigated in the STOWA project on the MBR pilot Maasbommel, which will start at the beginning of 2002. To illustrate: in Figure 91 the effluent of the Beverwijk WWTP secondary clarifier and the MBR pilot plants permeate are shown.

The goal of the side study was to measure the most important effluent parameters and to get a global overview on the expected effluent quality.

Literature review
In general only COD, BOD, SS and nutrient removal in MBR systems are described in literature. The results on nutrient removal were often very good due to the very low loading of the systems. The SS permeate concentration was usually below detection level.

For a number of (full scale) installations the microbial parameters were measured and published. In general, the removal of total and faecal coliforms was around log 6, the removal of viruses was varying from log 2-4. An advantage of this disinfection method is that the micro-organisms are totally removed and not just inactivated with a risk of re-activation.

Only few publications were found in which attention was paid to the removal of micro-pollutants. AOX's, medical substances, pesticides and endocrine substances (phthalates) were measured in both the MBR's permeate and the Immenstaad WWTP effluent. The AOX level in the MBR permeates was equal to the conventional effluent. For all the other substances significant lower concentrations were measured in the MBR permeate.

Measurements
The measurement programme was defined in association with the USHN project group and the STOWA steering committee. The sampling and analyses have been executed by different laboratories. The measured parameters are:

- microbial parameters: colony number (22°C), thermotolerant coli-bacteria, bacteriophages;
- heavy metals: As, Cd, Cr, Cu, Hg, Pb, Ni, Zn;
- polycyclic aromatic hydrocarbons (PAH's);
- extractable organic halogens (EOX's);
- phthalates.

Pesticides have not been measured as the measurement period was mainly in winter time, a period in which pesticides are not expected to be traceable in wastewater.

The main results from the measurements are:

- thermotolerant coli-bacteria are removed completely;
- the results on bacteriophages showed relatively major variations, in general it can be concluded that the bacteriophage concentration in a MBR permeate was 100 to 1,000 times lower than in the WWTP effluent;
- the removal efficiency of the heavy metal components depends to what extent the metal is attached to the sludge; especially Cu, Hg, Pb and Zn are removed significantly;
- the PAH's are almost completely attached to sludge, which implies that by decreasing the effluent SS concentration, the PAH's level can be reduced to nearly complete removal;
- as NaOCl is used for membrane cleaning, chlorine compounds are introduced in the MBR system on a regular basis and, as a consequence, high peak EOX concentrations were measured.

In a full scale design attention should be paid to the cleaning procedures, to limit the EOX increase in the permeate.

10.5 Side study 5: Sludge treatment

MBR technology has several advantages compared to traditional activated sludge processes, e.g. high effluent quality, limited space requirements and modular set up. However, one area of concern where very little work has been carried out is that of MBR sludge treatment. Due to higher stress conditions the MBR sludge composition (floc size and structure) differs from conventional WWTP's. The aim of this study was to investigate the treatment characteristics and possibilities of the MBR sludge in relation to that of a conventional WWTP. A research programme was defined and split into three sub-categories: digestion, thickening and de-watering.

1. Digestion

A batch and semi-continuous test was carried out on the various MBR sludges available from the MBR pilots, and compared to the conventional Beverwijk WWTP.

The batch test yielded the maximum gas production of each MBR sludge and appeared similar to that produced by the conventional activated sludge. In light of these positive results a semi-continuous digestion test was executed on one MBR sludge (Zenon) and compared to a laboratory reference plant receiving a mixed sludge feed of primary and secondary sludge.

The acclimatisation of the digester to the MBR sludge took 2 months and as a result the findings were limited. After a stable period was achieved the gas production was at 75% of the batch test for a similar 23 day retention. In comparison to the mixed sludge reactor the results were poor with only 35% of the gas production achieved.

2. Thickening

The MBR sludge was compared for thickening on two devices, a belt-thickener and a centrifuge.

In phase 2 the belt thickener yielded positive results but at higher, non economical polymer dosing rates. The polymer dosing was reduced significantly once the MBR feed waters were set to raw/screened influent in phase 3. Here the results were comparable to that of conventional activated sludge thickening.

Sludge thickening with a centrifuge proved to be possible without polymer on the conventional activated sludge but yielded only 80% recovery on the MBR sludges.

Figure 92 - Thickening centrifuge

3. De-watering

A laboratory scale filter press and a centrifuge were used to dewater the various sludge samples. Direct dewatering was not possible with the filter press as the flocs produced were small and relatively unstable under the conditions of high shear and pressure (6.5 bar).

The centrifuge yielded better results and were comparable to the conventional activated sludge on the same equipment. It is believed that the higher specific gravity of the MBR sludge aided the de-watering process. A full scale centrifuge was used to dewater the conventional sludge and the DS% reached was higher than that of the test equipment. The latter however could not be extrapolated to the MBR sludge.

11 EVALUATION

Biological performance

The biological performance of the four systems is summarised in Table 29 and split into the four different phases. During these phases not only the pre-treatment and the phosphorus removal process was different, but also process conditions like sludge loading, temperature and hydraulics. Judging the data one has to realise that the pilot plants have regularly been optimised to achieve better performance. These technical adjustments and specific tests to examine the membrane performance (cold water and peak tests) gave foreseen (major) disturbances in the biological performance. Furthermore the pilot plants were designed to operate under the conditions of phase 1, i.e. primary settled wastewater and pre-precipitation. Without major adaptations the pilots were also deployed for phases 2, 3 and 4.

From Table 29 the following conclusions can be drawn:

- The operational conditions of phase 1 were comparable to the design parameters. The differences between the design of Kubota and Mitsubishi on one hand, and X-Flow and Zenon on the other hand are related to the different designers. As can be seen from Table 29, during the different phases the operational load on the Kubota pilot plant has increased. Related to the maximum hydraulic load applicable on the Mitsubishi pilot installation, the biological load was significantly lower compared to the other pilots. This gave certain problems concerning sludge quality and consequently membrane performance. In all phases the N-load on the pilot systems remained more or less the same, due to the relatively small effect of pre-clarification and precipitation on N-influent concentrations.
- As a consequence of a COD/N ratio of lower than 6 during phase 1, the average TN concentration amounts to a level higher than 10 mg/l. In phase 2 this ratio climbed up to 7.7 and influent concentrations dropped down due to extreme rainfall. Despite the lower process temperatures the TN permeate concentrations were below 10 mg/l, except for the X-Flow pilot. This was due to technical disturbances and adaptations. Although the COD/N ratio in phase 3 increased further to 9.1, the TN concentrations increased to a higher level. Also here the circumstances mentioned before were the main causes. Looking at the technical and technological limitations (a.o. the limited recirculation flow capacity) of the pilot installations without doubt it will be possible to achieve lower TN concentrations at full scale. This item is further discussed at the end of this paragraph.
- P-removal during phase 1 with pre-precipitation was not optimal with average permeate concentrations above 1 mg/l. This was mainly due to the extreme varying influent concentrations and the process control of the chemical dosing installation at the Beverwijk WWTP. In phase 1 this process was not optimised as the membrane performance was of much more importance. During the phases 2 and 3 with the application of simultaneous precipitation it has proven to be possible to achieve very low permeate concentration. Starting at Me/P ratios similar to the previous pre-precipitation phase (1.0 - 1.2), TP concentrations below 0,1 mg/l were achieved, even without an optimal chemical dosing control. The chemical dosing set points were changed to get the required permeate concentration of 1 mg P_{total}/l. Based on the results achieved it can be stated that it is possible to reach very low P-concentrations, giving the impression that also biological uptake plays a significant role.

  Since March 2001 the possibilities of 100 % biological P-removal are investigated in the Kubota pilot plant. Since then, the phosphorus permeate concentration decreased rapidly to <0.2 mg P_{total}/l and remained stable for a few months. Since May 2001 all pilot plants are in operation without additional chemical phosphorus removal, the new X-Flow pilot is also configured with an anaerobic tank. All pilot plants show enhanced biological phosphorus removal and relatively low phosphorus permeate concentration, taking into account the high influent phosphorus concentrations and large fluctuations.

Table 29 - Biological performance pilot installations

		Kubota	Mitsubishi	X-Flow	Zenon
Design					
MLSS concentration	g/l	12	10	10	10
Total volume	m^3	30.8	34.2	5.7	23.6
Sludge load	kg COD/(kg MLSS.d)	0.049	0.043	0.073	0.074
	kg N/(kg MLSS.d)	0.008	0.007	0.012	0.012
Phase 1 - Primary settled wastewater and pre-precipitation					
Sludge load	kg COD/(kg MLSS.d)	0.054	0.043	0.070	0.075
	kg N/(kg MLSS.d)	0.009	0.008	0.012	0.012
MLSS concentration	g/l	10.5	8.9	7.5	10.4
Process temperature	°C	21	20	23	20
COD concentration / efficiency	mg/l / %	32 / 91	31 / 91	32 / 91	31 / 93
TN concentration / efficiency	mg/l / %	11 / 81	12 / 82	16 / 75	13 / 80
TP concentration / efficiency	mg/l / %	2.8 / 62	2.1 / 74	2.5 / 69	1.5 / 80
Sludge production / sludge age	kg/kg $COD_{removed}$ / days	0.31 / 70	0.27 / 87	0.24 / 66	0.27 / 51
DSVI / CST	ml/g / seconds	80 / 50	130 / 130	90 / 60	110 / 90
Phase 2 - Primary settled wastewater and simultaneous precipitation					
Sludge load	kg COD/(kg MLSS.d)	0.060	0.045	0.063	0.093
	kg N/(kg MLSS.d)	0.008	0.006	0.008	0.012
MLSS concentration	g/l	12.0	9.9	9.9	10.9
Process temperature	°C	13	14	13	15
COD concentration / efficiency	mg/l / %	21 / 93	25 / 92	33 / 90	23 / 93
TN concentration / efficiency	mg/l / %	9 / 79	9 / 79	13 / 69	7 / 81
TP concentration / efficiency	mg/l / %	1.3 / 84	1.9 / 71	1.5 / 83	0.7 / 90
Sludge production / sludge age	kg/kg $COD_{removed}$ / days	0.45 / 41	0.41 / 62	0.32 / 59	0.43 / 28
DSVI / CST	ml/g / seconds	90 / 70	140 / 130	90 / 110	120 / 100
Phase 3 - Raw influent and simultaneous precipitation					
Sludge load	kg COD/(kg MLSS.d)	0.084	0.059	0.091	0.110
	kg N/(kg MLSS.d)	0.009	0.006	0.009	0.011
MLSS concentration	g/l	10.6	10.6	7.2	10.0
Process temperature	°C	12	12	14	14
COD concentration / efficiency	mg/l / %	31 / 95	30 / 95	43 / 93	35 / 94
TN concentration / efficiency	mg/l / %	10 / 84	11 / 81	18 / 72	10 / 84
TP concentration / efficiency	mg/l / %	1.3 / 88	0.7 / 92	0.6 / 94	0.3 / 97
Sludge production / sludge age	kg/kg $COD_{removed}$ / days	0.49 / 27	0.57 / 31	0.28 / 42	0.42 / 26
DSVI / CST	ml/g / seconds	90 / 60	120 / 120	90 / 70	110 / 70
Phase 4 - Raw influent and biological phosphorus removal					
Sludge load	kg COD/(kg MLSS.d)	0.100 #	0.084	0.063 #	0.089
	kg N/(kg MLSS.d)	0.009 #	0.008	0.006 #	0.009
MLSS concentration	g/l	10.8	11.6	10.6	11.2
Process temperature	°C	19	18	n.a.	20
COD concentration / efficiency	mg/l / %	32 / 95	34 / 94	36 / 94	33 / 95
TN concentration / efficiency	mg/l / %	11 / 81	9 / 85	8 / 86	8 / 86
TP concentration / efficiency	mg/l / %	0.8 / 93	1.2 / 90	1.4 / 88	1.9 / 84
Sludge production / sludge age	kg/kg $COD_{removed}$ / days	0.34 / 30 #	0.49 / 26	0.50 / 34 #	0.40 / 29
DSVI / CST	ml/g / seconds	90 / 50	100 / 80	100 / 70	100 / 50

\#   calculation based on the activated sludge volume excl. anaerobic tank

- The sludge production of the pilot plants was similar compared to traditional activated sludge systems. The differences between the pilot installations mainly occurred due to the technical disturbances mentioned before and the scale of the tests, giving certain inaccuracies. As expected, the specific sludge production for phase 1 was significantly lower than for the following phases. The significantly lower sludge production in the X-Flow pilot plant is mainly caused by the high efficiency micro-screen.
- The sludge characteristics expressed as sludge volume index were excellent during all phases, with an exception for the Mitsubishi pilot in phase 1. This was mainly due to the mechanical stress caused by high velocity pumps and the relatively low sludge load. For design purposes it is of great importance to make the right choices for technical equipment, since the floc structure influences the membrane performance.
- The energy consumption and input of all pilot installations was (much) higher than the expected energy requirement in a full-scale plant, which is ≤ 1 kWh/m^3.

Nitrogen removal

The effect of the pilot plant configuration and design capacity on the nitrogen removal potential of the pilot plant, is demonstrated with Figure 93. In this figure the relative size and the capacity of the internal recirculation flows from the four pilot plants is schematically presented.

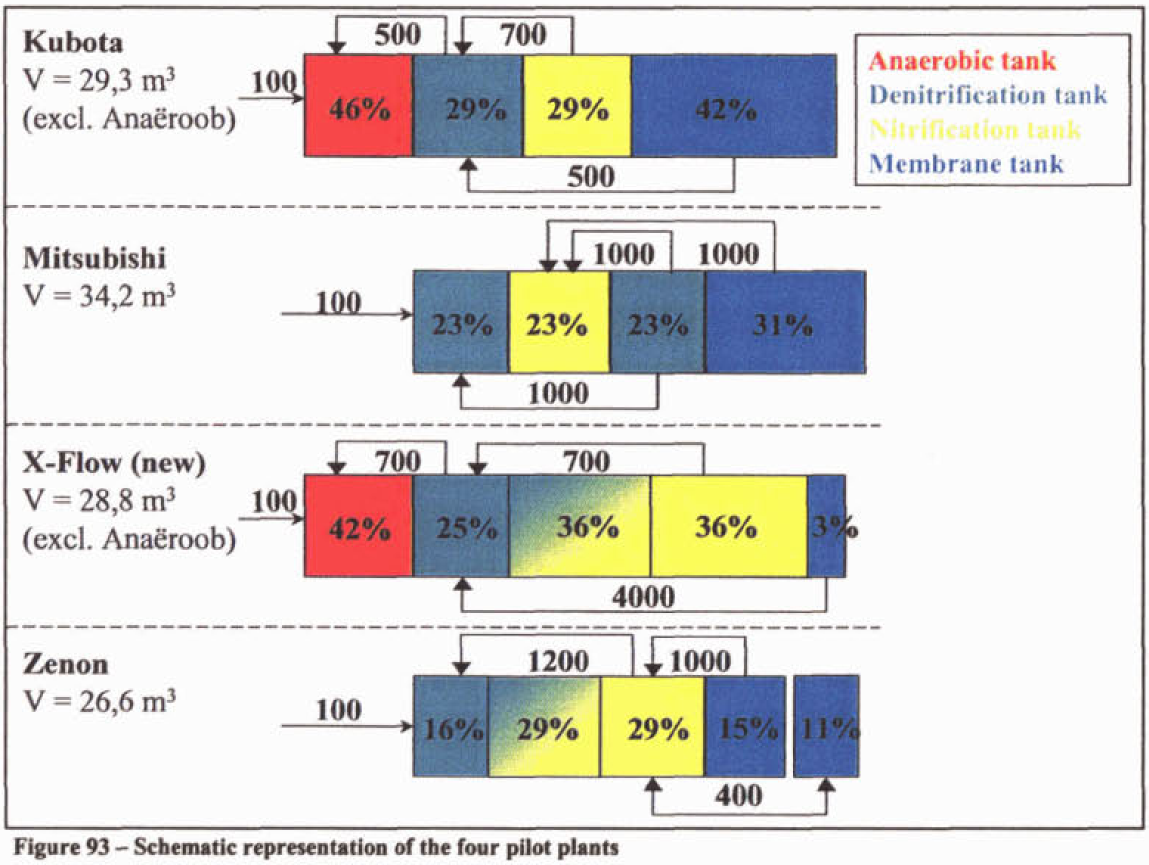


Figure 93 – Schematic representation of the four pilot plants

In phase 4, when the COD availability was not limiting, the recirculation flows were the main factor influencing the nitrogen removal capacity of the pilot plants. As shown, the membrane compartment of the Kubota pilot plant is relatively large and the recirculation flows are relatively small. As a consequence the expected and realised nitrogen removal capacity in this pilot plant is the lowest. Changing the recirculation flow from the membrane tank to the denitrification tank didn't have a significant effect on the results. The other three pilot plants have larger recirculation flows and therefore show a better performance on nitrogen removal.

Membrane performance

The membrane performance of the four systems is summarised in Table 30 and split into the different research phases. During these phases the membrane systems were all optimised and adapted many times. Therefore it is not possible to compare the results from the phases without the background described in the previous chapters. It can be stated that all systems performed better as the research period progressed.

From Table 30 and more in-depth results from previous chapters the following conclusions can be drawn:

- The maximum design flux for the pilot plants was mainly based on manufacturers experience and design philosophy. During the different phases it has been proven that this experience was not gained under specific Dutch circumstances. Especially the strong varying influent flows influenced the performance of the membrane systems, much more than expected, also compared to the different pre-treatment steps and process temperatures.

Table 30 - Membrane performance pilot installations

		Kubota	Mitsubishi	X-Flow	Zenon	
Design						
Capacity	m^3/h	7.8	6.4	1.8	7.6	
Daily flow	m^3/d	39	32	9	38	
Membrane surface	m^2	240	315	30	184	
Maximum flux net	$l/(m^2.h)$	32.5	20.3	60.0	41.3	
Phase 1 - Primary settled wastewater and pre-precipitation						
Average daily flow	m^3/d	48	38	8	46	
Maximum flux net	$l/(m^2.h)$	41.7	28.1	50.0	41.3	
Duration	h	100	95	4	104	
Operating mode [1]	CF / PF	CF => PF	CF => PF	CF => PF	CF => PF	
Cleaning procedure [2]	IC / MC	IC	IC	IC	IC	
Phase 2 - Primary settled wastewater and simultaneous precipitation						
Average daily flow	m^3/d	72	44	10	74	
Maximum flux net	$l/(m^2.h)$	41.7	8.2	22.5	41.3	
Duration	h	acc. to PF	continuous	continuous	acc. to PF	
Operating mode [1]	CF / PF	PF	PF => CF	PF => CF	PF	
Cleaning procedure [2]	IC / MC	-	IC	IC	IC + MC	
Phase 3 - Raw influent and simultaneous precipitation					*zw500a*	*zw500c*
Average daily flow	m^3/d	52	40	6	45	50
Maximum flux net	$l/(m^2.h)$	41.7	8.2 => 20.8	37	30-35 [3]	60
Duration	h	acc. to PF	continuous	continuous	acc. to PF	> 200
Operating mode [1]	CF / PF	PF	CF	CF	PF	CF
Cleaning procedure [2]	IC / MC	-	IC => (E)MC	IC + MC	MC	MC
Phase 4 - Raw influent and biological phosphorus removal					*zw500a*	*zw500c*
Average daily flow	m^3/d	51	55	33	44	50
Maximum flux net	$l/(m^2.h)$	41.7	31	25 => 50	35	35
Duration	h	acc. to PF	120	continuous	acc. to PF	continuous
Operating mode [1]	CF / PF	PF	CF	CF	PF	CF
Cleaning procedure [2]	IC / MC	- [4]	EMC	MC	MC	MC

1) CF = constant flow / PF = proportional flow, acc. = according
2) IC = intensive clean / MC = maintenance clean / EMC = enhanced maintenance clean
3) During temperatures below 10 °C maximum flux was restricted to 30 and 35 $l/(m^2.h)$
4) Intensive cleaning procedure planned for week 10 in 2002

- As a consequence of the introduction of a relaxation mode in the process control, Kubota performed better than expected. During all four phases the maximum net flux was maintained at 41.7 l/(m^2.h), which was presumably not even its maximum. This flux was combined with high permeability's and consequently enough reserve capacity to be able to cope with long term rain fall conditions.
- The Mitsubishi system successfully performed a peak test at the design flux, at a relatively high temperature. During the whole research period it has been proven that the system is not capable in dealing with large hydraulic fluctuations. Since phase 3 the pilot ran with an optimised process control at higher continuous fluxes. Under these circumstances including (enhanced) maintenance cleaning, the system performed well, accepting a low and relatively stable decrease of permeability. Recently a peak continuous flux of 31 l/(m^2.h) has been achieved for a 5 days period, at a temperature of >20°C.
- The X-Flow system was designed too optimistically. Comparable to the Mitsubishi system, the X-Flow module performed optimal in a constant permeation mode. With an optimal process control including MC and relaxation, both the old and the new pilot plant system could be run with constant high fluxes.
- The Zenon ZW500a system has shown to be reliable under the original design fluxes, in which a lower peak flux in phases 3 and 4 (without primary clarification) was foreseen. With the transition from IC to MC at the end of phase 2, the system was capable of handling peak flows, without loosing excessive permeability.
- The ZW500c module displayed many improvements over the ZW500a module regarding peak fluxes, chemical consumption and the lack of sludging. However, the ZW500c module appeared more temperature sensitive than the ZW500a module, as a result of lower operating temperatures it was more difficult to recover the performance with MC procedures. The reduction in permeate headers from two to one may have adversely effected the permeation characteristic of this module configuration.
- Under full-scale conditions membrane tanks will be designed with a specific hydraulic capacity. Under DWF conditions and during night membrane tanks will be out of operation, having the opportunity to be left in a relaxation mode. It has to be realised that these circumstances are much more optimal compared to those under which the pilot plants were operated.
- In determining the applicability of a MBR system, the maximum achievable flux and consequently the required membrane surface, is one of the parameters. Besides this, also the membrane compactness and consequently the required membrane tank volume and foot print, play an important role. From this point of view the Mitsubishi, Zenon and Kubota double deck system system have an advantage.
- Based on the experience gained at the Beverwijk WWTP it was necessary to further optimise the cleaning procedures and establish further savings on chemical consumption. Understanding of cleaning procedures and its influence was necessary to achieve results and satisfying membrane life time expectations. Technical and technological sub-optimal operations related to cleaning procedures and process control lead to the replacement of the Mitsubishi and X-Flow module.
- Under normal process conditions the following cleaning frequency is predicted:
  - ⇒ for the Kubota system 1 time per year IC;
  - ⇒ for the Mitsubishi system at high constant fluxes and dependent on process temperatures a regular (every week) enhanced MC procedure in combination with 1 IC per year;
  - ⇒ for the X-Flow system at high constant fluxes and dependent on process temperatures a regular (every week) MC procedure in combination with 1 IC per year;
  - ⇒ for the Zenon system a regular MC (every week) and one IC per year.

- The chemical dosing required for the four systems (as expected at the start of the research project and as used at the pilot plants after the optimisations described above), is summarised in table 30. Based on these figures it is concluded that the chemical consumption of the Kubota and Zenon (ZW500c) pilot plant has been significantly reduced and has reached a low level. The chemical consumption of the Mitsubishi and X-Flow pilot plant is (still) relatively high.

Table 31 - Chemical consumption pilot installations (in $g_{product}/m^3$ influent)

chemical		Kubota	Mitsubishi	X-Flow	Zenon a / c
NaOCl	- expected	0.59	4.3	0.11	0.64
	- current optimisation	0.25	0.8	1.4	0.46 / 0.23
Acid	- expected	1.2	2.2	0.22	1.6
	- current optimisation	0.5	1.6	2.9	1.2 / 0.6

- During the research period all four systems have been optimised and developed considerably. At this moment both the Kubota and Zenon system are well established and within a well defined operating window reliable for use in the municipal wastewater market for Dutch circumstances. Also the Mitsubishi and X-Flow system have significantly progressed. From results obtained, it is expected that with constant flow operations, optimised process control and cleaning procedures (MC + IC) reliable systems will be realised, although further test work has to be carried out to prove the expectations.

Design

The design of MBR installations for municipal waste water treatment up until now was based on experience gained in industrial applications and normal design procedures for traditional municipal wastewater treatment. The results achieved on the Beverwijk WWTP have made clear that regarding the design philosophy for MBR installations for full scale municipal WWTPs, the following items have to be taken into account:

- Pre-treatment of the influent is of utmost importance. Abrasive components, fat and hair have to be removed by fine screens to prevent problems with the membranes. Depending on the specific situation and further technical and technological requirements, a choice for an optimal screen has to be made.
- For a MBR installation fulfilling N and P requirements, the direction of the re-circulation flows especially from the membrane tank have to be considered seriously. In this respect it is important to realise and calculate the differences between an internal and external placement of membranes. The control of the nitrogen removal process on one hand and sludge concentration in the membrane zone on the other hand must be separated, if stringent nutrient removal is required and risks for sludging membrane modules have to be minimised. An important characteristic of a MBR in this respect is the intensive aeration in the last compartment of the installation. As a consequence the permeate is extracted at the location with the highest nitrate concentration. To be able to reach a lower TN permeate concentration, high recirculation flows should be introduced in the system configuration design. Furthermore it is important to consider the influence of influent components and microbial products related to the placement of membranes and the overall design of a MBR installation.
- Considering a full scale design of a MBR installation it is impossible to execute an intensive clean manually as is done at the MBR installations world wide in operation. For application the technology at bigger capacities, an automatic cleaning system has to be developed. This item is very much related to the conception of a standard membrane tank, to be developed for the demonstration installation.

- Concerning chemical membrane cleaning, attention must be focused on the decrease of the chemical use, the re-use of chemicals and the use of alternative (less harmful to the environment) chemicals.
- Related to a standard membrane tank it is important for the end customer to have the availability of membrane modules of different types, all fitting in one standard civil construction.
- Sludge treatment of MBR installations seems to be not significantly different from sludge treatment from traditional WWTP's. Based on the results so far the choice between gravitational and mechanical sludge thickening has to be reflected.
- As the membrane performance is strongly related to the sludge characteristics, the use of a selector tank may, also for a MBR system, be an option.
- In general the MBR technology is rather new and like all other technologies, consists of certain risks which have to be evaluated. A conscientiously executed risk analyses is necessary, giving back ground and motivation for process parts and equipment to be implemented or designed redundant. An example worthwhile to mention here is the air supply to the membrane modules. Without any doubt it can be said that this process is of utmost importance to be 100 % reliable.

Costs

Although the pilot study at the Beverwijk WWTP was not aiming at a direct relationship with costs of a MBR installation, information is available from different feasibility studies in which the costs for a MBR- and traditional installation under different circumstances are compared. In Figure 94 the investment costs for a new WWTP are compared.

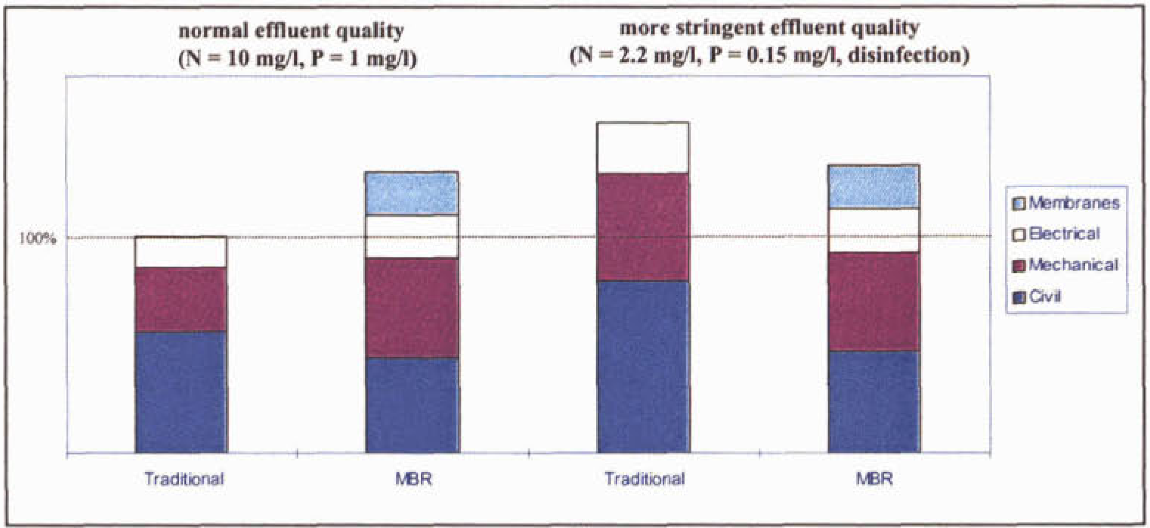


Figure 94 - Distribution of the capital costs for a 2,500 m³/h treatment plant

The investment costs for a completely new MBR installation aimed at normal effluent requirements is currently more expensive than a traditional WWTP. The decrease for the civil construction does not counterbalance the costs for the extra mechanical equipment and membrane modules. This conclusion can change dramatically if it is a matter of extension for a specific WWTP which would otherwise require too much space. Also if a more stringent effluent quality is required, the cost comparison can currently equal or be bettered for the MBR alternative.

For the above presented MBR system with the more stringent effluent requirements for nitrogen, phosphorus and disinfection, the annual costs are compared in Figure 95.

The annual costs of a MBR installation are strongly influenced by the depreciation for the mechanical equipment and the replacement of the membranes. Currently the latter is very difficult to predict, while practical experience from 3 to 8 years is not enough for a realistic estimation. With manufacturers, the discussion around lifetime leads to capacity guarantees.

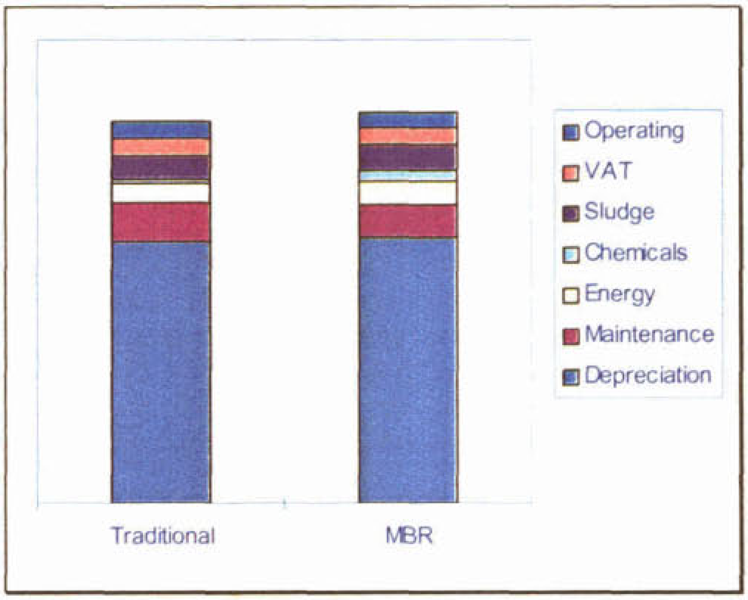


Figure 95 - Distribution of the annual costs for a traditional and a MBR system, with more stringent effluent requirements

Regarding variable costs, especially the energy consumption is an interesting item, which is mainly determined by the α-factor and consequently the energy for aeration. In the feasibility studies mentioned before, α-factors of 0.40 - 0.45 have been used for calculation purposes. Based on the results of pilot scale measurements, side study 3 and full scale measurements at WWTP's in Germany it is expected that these values are too low. Cost differences between a MBR and traditional WWTP concerning manpower, chemical consumption and sludge treatment is supposed to be minimal.

12 CURRENT STATUS AND PERSPECTIVE

In the course of the pilot trial, many of the uncertainties involved with the MBR technology were considered and tested. All the risks addressed at the onset of the pilot trial were pinpointed and eliminated where possible or significantly reduced to acceptable levels. Despite the defined operating windows, all four pilot systems can be further optimised and or developed in the following directions:

Kubota

- A reduction of chemical consumption for the cleaning procedure. The procedure is fixed and works, but the lack of flexibility in the procedure causes biological stress.
- An increase of efficiency and net flux. This would further reduce the installed membrane surface required and therefore investment cost. The efficiency is conservative at present and the limits are loosely defined.
- Proof of the long term operation of the Double Deck concept. This must be reliably established for the Dutch circumstances. There is one reference pilot which is being operated for one year in Japan under almost constant flow conditions.
- Investigation of gravity flow.

Mitsubishi

- An increase of the net flux efficiency (ratio between net and gross flux) may be possible.
- Optimisation of cleaning procedure and the reduction of chemicals must also be considered.

X-flow

- An optimisation of the process control and the operation philosophy. Regarding maximum achievable operating fluxes and cleaning criteria on the long term.
- An increase of the efficiency under all operating criteria is required.
- A reduction of the chemical consumption for the cleaning procedure must be considered.
- A comparison of 5.2 and 8.0 mm diameter tubes must be carried out.

Zenon

- Further development of the ZW500c module, with two permeate headers in order to gain the advantages of both module types.
- A further optimisation of the chemical consumption is required.

It is the intention to continue the pilot plant research at the Beverwijk WWTP and combine the further development on the long term effects of the membranes and the education of operating personnel.

At this moment both the Kubota and Zenon system are well established and within a defined operating window reliable for use in the Dutch municipal wastewater market. Also the Mitsubishi and X-flow system have significantly progressed. From results obtained, it is expected that reliable systems will be realised, although further test work has to be carried out to prove the expectations.

The objectives originally set for the pilot plant study at the Beverwijk WWTP have been achieved to a suitable level of satisfaction and most critical items have been addressed and quantified. After one year of intensive research and further development the technological feasibility of the MBR concept for the Netherlands has been proven and suitable to be extended to demonstration scale. The step from fundamental development on pilot scale to application on demonstration scale can now be made.

13 REFERENCE LIST

[1] Houten, R. van, TNO/STOWA: 1e Nederlandse symposium Membraan Bioreactor technologie, Apeldoorn, November 2000.

[2] Brindle, K. et al., Membrane Bioreactors for Wastewater Treatment, London, 2000.

[3] Houten, R. van, Brouwer, H., Magazine Afvalwater: Vakblad voor communaal, industrieel en agrarisch afvalwater, Nummer 4. 2001.

[4] Muller, E.B, Stouthamer, A.H., Verseveld, H.W. van, Eikelboom, D.H., Aerobic domestic waste water treatment in a pilot plant with complete sludge retention by cross-flow filtration, Water Research Vol. 29, No. 4, pp. 1179-1189, 1995.

[5] Churchouse, S., Operating experiences with the Kubota submerged membrane activated sludge process, March, 1997.

[6] Erftverband, A new option in waste water treatment. History of the Rödingen Sewage Treatment plant, 2000

[7] Zenon Environmental Systems Inc.- Municipal Division, Reference list, 2000

[8] Mathieu Praderie, Une Rapport intermédiaire sur les essais pilote de traitement des eaux résiduaires urbaines par le procédé BIOSEP, Maisson-Lafitte, July, 1995.

[9] Günder, Berthold, Das Membranbelebungsverfahren in de kommunalen Abwasserreinigung, München, 1999.

[10] ATO Wageningen UR, Reductie organische afvalstromen E.E.T. k98 021 / 39410-1210, oktober, 2000.

[11] Brindle, K., Stephenson, T., MBR- A compact wastewater treatment process. Reader Enquiry No. 285, December, 1996.

[12] Ghyoot, W., Vandaele., S., Reduced sludge production in a two-stage membrane-assisted bioreactor, Water Research. 34(1): 205-215, January, 2000.

[13] Zhang, B., Yamamoto, K., Ohgaki, S., Kamiko, N., Floc size distribution and bacterial activities in membrane separation activated sludge processes for small scale wastewater treatment/reclamation, Water Science Technology, 35(6), 37-44. 1997

[14] Jurgens K., Arndt H., Zimmermann H., Impact of metazoan and protozoan grazers on bacterial biomass distribution in microcosm experiments, Aquatic Microbial Ecology, 12,131-138, 1997

[15] B.S. Luxmy, Nakajima, F., Yamamoto, K., Predator grazing effect on bacterial size distribution and floc size variation in membrane-separation activated sludge, Water Science and Technology, Vol. 42 Nos. 3-4 pp. 211-217, 2000.

[16] Zhang, B., Yamamoto, K., The role of predators on sludge mineralisation in membrane separation bioreactor, 1997.

[17] Delanghe, B. , Nakamura, F. et al 1994. Biological denitrification with ethanol in a membrane bioreactor, Environmental Technology, 1994.

[18] Low, E.W., H.A. Chase, H.A., Reducing production of excess biomass during wastewater treatment, Water Research 33(5): 1119-1132, April, 1999.

[19] Ghyoot, W., Vandaele, S., et al, Nitrogen removal from sludge reject water with a membrane-assisted bioreactor, Water Research 33(10): 23-32, January, 1999

[20] Zenon Environmental Inc., ZeeWeed Engineering Manual (6.4.4, 6.4.5, 6,4,7), cleaning procedure, 2001.

[21] Brindle, K., Stephenson, T., 1996. The application of MBRs for the treatment of wastewater's, Biotechnology Bioeng 49, 601-610, 1996

[22] Ahn, K.H., Song, K.G., Treatment of domestic wastewater using MF for reuse of wastewater, Desalination: 126(1-3): 7-14, November, 1999.

[23] Defrance, L., Jaffrin, M.Y., Reversible of fouling formed in activated sludge filtration, Journal Membrane Science, 157(1):73-84, 1999.

[24] Behmann, H., Tonelli, F., Aerated hot membrane bioreactor process for treating recalcitrant compounds, Patent US 5558774, 1995.

[25] Jacobsen, B.N., Wastewater Quality Standards in European Countries - Effluent from municipal wastewater treatment plants, EWPCA Task Group, 1990.

[26] Ahn, K.H., Cha, H.Y., et al, Retrofitting municipal sewage treatment plants using an innovative MBR system, Desalination 124(1-3): 279-286, November, 1999.

[27] DHV Water BV, MBR Development for municipal waste water. Visit report Japan, January, 2001.

[28] Stephenson, Tom, Simon Judd, Bruce Jefferson and Keith Brindle (2000), Membrane bioreactors for wastewater treatment, IWA publishing, London, 2000

[29] Merlo, Rion P., Samer Adham, Paul Gagliardo, R. Shane Trussel, Rhodes Trussel and Montgomery Watson, Application of membrane bioreactor (MBR) technology for water reclamination

[30] Centraal Bureau voor de Statistiek, Waterkwaliteitsbeheer. deel B. Zuivering van afvalwater. 1997, Voorburg / Heerlen. 1999

[31] Ministerie van Verkeer en Waterstaat. Directoraat-Generaal Rijkswaterstaat. Rijksinstituut voor Integraal Zoetwaterbeheer en Afvalwaterbehandeling / RIZA. Rijksinstituut voor Kust en Zee / RIKZ. 1999, Hormoonontregelaars in water - Oriënterende studie naar de aanwezigheid van oestrogeen-actieve stoffen in watersystemen en afvalwater in Nederland

14 GLOSSARY

Activated sludge	Biological mass (flocs) produced in the treatment of wastewater by the growth of suspended bacteria and other micro-organisms under aerobic or anoxic conditions.
Aerobic condition	Descriptive of a condition in which dissolved oxygen is present.
Alpha factor	Ratio of the oxygen transfer coefficient in mixed liquor to the oxygen transfer coefficient in clean water.
Anaerobic condition	Descriptive of a condition in which dissolved oxygen, nitrate and nitrite are absent.
Anaerobic digestion	Anaerobic process which reduces the organic content of sludge.
Anoxic condition	Descriptive of a condition in which dissolved oxygen is absent and nitrite or nitrate are present.
Back pulse / flush	Reversed flow through the membrane to remove fouling on the membrane surface.
Cake filtration	Filtration across an artificial layer on the surface of the membrane.
Chemical precipitation	Conversion of components dissolved in water into undissolved form by chemical reaction with a precipitant.
Cleaning in place / CIP	Storage tank for permeate to enable flushing of the membrane
COD	Chemical oxygen demand; Mass concentration of oxygen equivalent to the amount of dichromate consumed when a water sample is treated with that oxidant under defined conditions.
Cross flow	A term used for the flow of liquid across the membrane surface.
CST	Capillary suction time; A method to measure the filterability of activated sludge.
Denitrification	Reduction of nitrate or nitrite to liberate mainly nitrogen gas by the action of bacteria.
DSVI	Diluted sludge volume index; Volume in millilitres occupied by 1 g of activated after settlement under specified conditions for 30 minutes.
Disinfection	Treatment of wastewater or sludge to reduce pathogenic activity below a specified level.
DWF	Dry weather flow; Flow at the wastewater treatment plant unaffected by rainfall or snowmelt.
Flux	The flow of liquid through a specific membrane surface area.

Intensive cleaning (IC)	A cleaning operation to return the membrane back to the original permeability.
Low pressure cross flow (LPCF)	Cross flow system with low speed, low pressure (generated by a permeate pump) and vertical air-flushing.
Maintenance cleaning (MC)	A (periodically executed) preventative cleaning of the membrane.
Membrane	A device for the separation of sludge from water.
Micro filtration	Filtration of particles to greater than 0,05 μm.
MLSS	Mixed liquor suspended solids; Concentration of suspended solids in the mixed liquor.
MLVSS	Mixed liquor volatile suspended solids; Concentration of organic suspended solids in the mixed liquor.
MTR (*Dutch*)	Abbreviation of Maximum Allowable Risk.
Nitrification	Oxidation of ammonium salts by bacteria. Usually, the end product of such an oxidation is nitrate.
N_{kj}	Kjeldahl Nitrogen (also TKN); Mass concentration of the sum of organic and ammonia-cal nitrogen.
N_{total}	Total nitrogen; Sum of the mass concentrations of Kjeldahl, nitrite and nitrate nitrogen.
Permeability	A measure for membrane performance (Flux / TMP).
Permeate	Effluent from a MBR installation.
Pre-precipitation	Chemical precipitation in the primary clarifier.
Proportional flow	A feed flow following the WWTP influent fluctuations.
P_{total}	Total phosphorus; Mass concentration of the sum of organic and inorganic phosphorus.
Relaxation	Mode during which no permeation of the membrane occurs.
Respiration rate	Rate of oxygen consumption due to respiration.
RWF	Rain weather flow; Maximum flow of wastewater a plant is designed to treat.
Screen	Device for removing coarse particles and objects from a flow of wastewater.

Simultaneous precipitation	Chemical precipitation in the activated sludge system.
Sludge	Mixture of water and solids separated from various types of wastewater as a result of natural or artificial processes.
Sludge age	Calculated time required to waste the total inventory of sludge being in the process tanks.
Sludge dewatering	Further process of reducing the water content of sludge, usually by mechanical means.
Sludge loading	Load of pollutants entering the biological treatment per unit mass of mixed liquor suspended solids or mixed liquor volatile suspended solids. It should be indicated whether the basis is total or volatile suspended solids.
Sludge treatment	Processing of sludge for its utilisation or disposal, eg sludge thickening, sludge stabilisation, sludge conditioning, dewatering, drying, disinfection, incineration.
Specific surplus sludge production	Ratio of mass suspended solids of surplus sludge to unit mass of COD removed.
SS	Suspended solids; Mass concentrations of solids in a liquid normally determined by filtration or centrifuging and then drying all under specified conditions.
Surplus sludge	Sludge that is removed from an activated sludge process.
TMP	Trans-membrane pressure; Driving force required to filter solids from liquid.
Ultra filtration (UF)	Filtration of particles larger than 0.005 μm.
Y-flow	A method to measure the dynamic viscosity of activated sludge.

APPENDIX 1 - PARTICIPANTS

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stowa

Execution

DHV

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Solis

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