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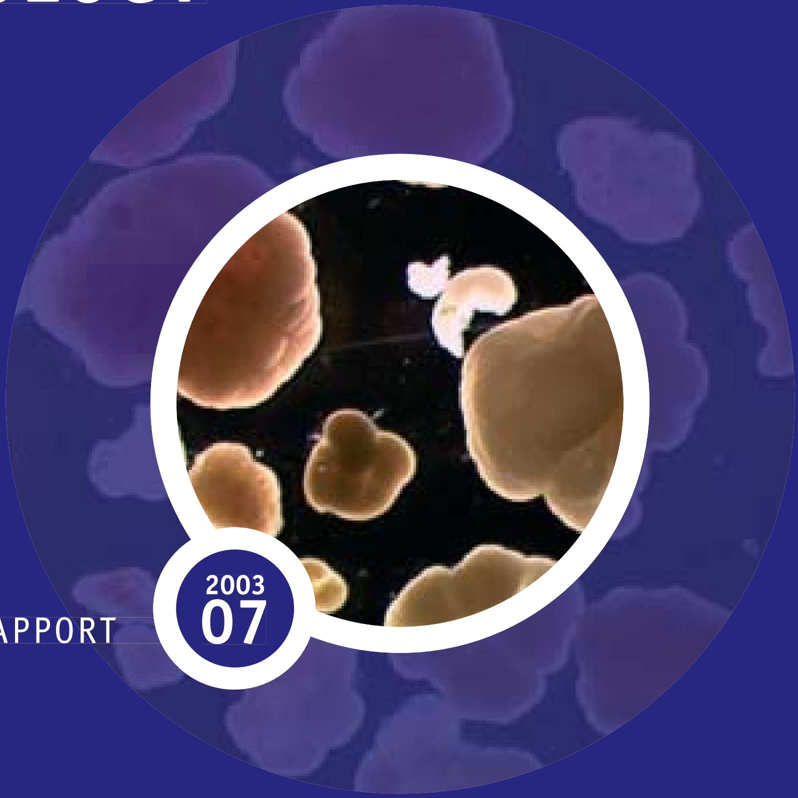
# DEVELOPMENT OF AEROBIC GRANULE REACTOR TECHNOLOGY



RAPPORT

2003
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# Final Report

Development of aerobic granule reactor technology

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# COLOFON

Utrecht, juni 2003

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**DRUK**
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# TEN GELEIDE

Een nieuwe ontwikkeling voor de zuivering van huishoudelijk afvalwater is de technologie gebaseerd op aëroob korrelslib. Onder specifieke procescondities kan korrelslib worden gekweekt. Door de unieke eigenschappen van dit korrelslib kunnen hoge volumetrische belastingen van beluchtingstanks worden bewerkstelligd en kan door de uitstekende bezinkingseigenschappen de scheiding van gezuiverd afvalwater en korrelslib bij hoge hydraulische belastingen plaatsvinden. Afhankelijk van de te kiezen procesconfiguratie is een goede effluentkwaliteit haalbaar, zodat naar verwachting zowel aan de vigerende als aan de te verwachten verscherpte effluenteisen voor rwzi's ten aanzien van stikstof en fosfaat zal kunnen worden voldaan. Hierdoor heeft de aëroob korrelslib technologie de potentie om een belangrijke bijdrage te kunnen leveren aan de zuivering van huishoudelijk afvalwater in Nederland.
Gedurende dit onderzoek zijn in het laboratorium de mogelijkheden onderzocht voor vergaande biologische stikstof- en fosfaatverwijdering. Verder is de technische en economische haalbaarheid van de technologie getoetst. Het is mogelijk gebleken om gedurende langere tijd stabiel korrelslib te handhaven en ten aanzien van CZV, totaal stikstof en fosfaat zijn hoge verwijderingsrendementen behaald. Verder is gebleken dat in vergelijking met conventionele zuiveringstechnieken een rwzi op basis van de aëroob korrelslib goedkoper is en een aanzienlijk kleiner ruimtebeslag heeft.
Gezien deze resultaten, alsmede het perspectief voor toepassing in de praktijk, heeft de STOWA besloten de ontwikkeling van deze technologie in de vorm van een pilot-onderzoek voort te zetten.

Voorliggend eindrapport beschrijft de resultaten van het onderzoek en de haalbaarheidsstudie. Het onderzoek is uitgevoerd in 3-liter reactoren op het laboratorium van de TU Delft, hoofdzakelijk op basis van synthetisch afvalwater.

Utrecht, juni 2003

De directeur van de STOWA

ir. J.M.J. Leenen

# SAMENVATTING

De meeste conventionele RWZI's, zoals de op actiefslib gebaseerde systemen, hebben een groot ruimtebeslag. Dit wordt veroorzaakt door de relatief slechte bezinkingseigenschappen van het actieve slib, hetgeen resulteert in lage hydraulische ontwerpbelastingen van nabezinktanks en lage toelaatbare drogestofgehaltes in beluchtingstanks. De laatste jaren zijn compactere systemen ontwikkeld, zoals slib-op-drager systemen (gepakt bed, gefluïdiseerd bed, airliftreactor), waarbij biomassa wordt geïmmobiliseerd op (gesuspen-deerde) dragerdeeltjes. Door het grote biofilmoppervlak en de hoge bezinksnelheid van de slib-op-dragerdeeltjes is een hoge volumetrische belasting mogelijk en dit resulteert in een beperkt ruimtebeslag. Recent laboratoriumonderzoek heeft uitgewezen dat in een discontinu gevoed systeem stabiele en gladde slibkorrels met een hoge dichtheid kunnen ontstaan, zonder dat hierbij dragermateriaal nodig is. Het discontinue voedingspatroon leidt tevens tot een efficiënte nutriëntenverwijdering.

In recent uitgevoerd laboratoriumonderzoek werd een stabiele korrelvorming verkregen met een synthetisch influent (acetaat ammonium oplossing), een pulsgewijze voeding van drie minuten ten opzichte van een totale cyclustijd van 180 minuten en een hoge concentratie opgelost zuurstof (DO). Bij een lage DO (40% van de verzadigingswaarde) en bij een langere beluchte voedingsperiode werden de korrels instabiel en vielen uit elkaar. Ook het gebruik van een bellenreactor in plaats van een airlift reactor was niet mogelijk onder deze omstandigheden. Voor toepassing van deze technologie op praktijkschaal is het echter essentieel om met langere voedingsperioden en lagere zuurstofspanningen te werken.
Tevens is uit economisch oogpunt het gebruik van een bellenkolom wenselijker dan een airlift reactor. Dit project had daarom als doel de knelpunten voor opschaling te benoemen en mogelijk te elimineren alsmede om de economische haalbaarheid van de aërobe korrelslib reactor (of Granule Sequencing Batch Reactor: GSBR) te evalueren.
Bij het laboratoriumonderzoek is met name ingegaan op de mogelijkheden voor vergaande simultane nutriëntenverwijdering. Daartoe werd de voedingswijze van het influent gevarieerd en is de korrelvorming en nutriëntenverwijdering bij lage zuurstof concentraties onderzocht. Verder is korrelvorming in een bellenkolom in plaats van een airliftreactor onderzocht. Ten slotte is in een kort vooronderzoek aangetoond dat korrelslib kan worden gevormd met voorbezonken huishoudelijk afvalwater als substraat.

Om het effect van een pulsvoeding te bewerkstelligen, is voorgesteld om gedurende de eerste 33% van de cyclustijd anaëroob te voeden. Wanneer in deze periode geen substraatopname plaatsvindt, zou de substraatconcentratie bij het starten van de beluchting in dat geval overeenkomstig zijn met die van een pulsdosering. Wanneer de voeding van het influent in een opwaartse stroom door het bezonken korrelbed onder anaërobe omstandigheden plaatsvindt, kan dit zelfs biologische fosfaatverwijdering stimuleren. In dat geval wordt het substraat wel tijdens de anaërobe voedingsperiode in de cellen opgenomen en opgeslagen als celintern polymeer. Bij een anaërobe voedingstijd van 60 minuten (totale cyclus was 180 minuten) en een DO concentratie van 20% is met een synthetisch influent gedemonstreerd dat een simultane CZV- (acetaat), fosfaat- en stikstofverwijdering mogelijk was (100% CZV (acetaat) en fosfaat verwijdering, 100% ammonia verwijdering door nitrificatie en 90% totaal-stikstof verwijdering). De biomassaconcentratie in dit type korrelreactor was ongeveer vijf maal zo hoog dan in een actiefslibsysteem met slibvlokken. Deze resultaten geven een indruk van de potentie van de technologie.

In een apart experiment is de vorming van aëroob korrelslib getest op basis van voorbehandeld (vergaand voorbezonken) huishoudelijk afvalwater als substraat. Aërobe korrels werden gevormd en alle biodegradeerbare CZV werd uit het influent verwijderd. De mogelijkheden van nutriëntenverwijdering zal in vervolgonderzoek moeten worden aangetoond.

De haalbaarheidstudie laat zien dat de aërobe korreltechnologie veelbelovend is. De totale jaarlijkse kosten van twee in beschouwing genomen GSBR-varianten (GSBR met voorbehandeling en GSBR met nabehandeling) zijn 6-16% lager dan die van de referenties (actiefslibsystemen). Uit de gevoeligheidsanalyse volgt dat de GSBR-technologie minder gevoelig is ten opzichte van de grondprijs en gevoeliger ten opzichte van een hogere RWA dan de referentiesystemen. Dit betekent dat de GSBR-technologie aantrekkelijker wordt bij lagere RWA/

DWA verhoudingen en bij hogere grondprijzen. Door de hoge volumebelasting is het benodigd oppervlak van een RWZI gebaseerd op aëroob korrelslib 25% van het benodigd oppervlak van de referentiesystemen. Met een GSBR in combinatie met alleen een voorbehandeling kan niet aan de huidige effluenteisen voor huishoudelijk afvalwater in Nederland worden voldaan. Dit wordt voornamelijk veroorzaakt door de verwachte hoge concentratie aan zwevend stof in het effluent.
Een groeiend aantal RWZI's in Nederland moet in de nabije toekomst gaan voldoen aan strengere effluenteisen. Dit betekent dat veel actiefslibinstallaties moeten worden uitgerust met een nabehandelingstap (b.v. zandfiltratie) of zullen worden omgebouwd naar MembraanBioReactoren. In dat geval kan een GSBR met zowel voor- als nabehandeling een aantrekkelijk alternatief vormen.

Op basis van de resultaten van zowel het onderzoek als de haalbaarheidsstudie wordt aanbevolen de aëroob korrelslibtechnologie verder te ontwikkelen, bijvoorbeeld door een pilotonderzoek bij een rwzi.

# DE STOWA IN HET KORT

De Stichting Toegepast Onderzoek Waterbeheer, kortweg STOWA, is het onderzoeksplatform van Nederlandse waterbeheerders. Deelnemers zijn alle beheerders van grondwater en oppervlaktewater in landelijk en stedelijk gebied, beheerders van installaties voor de zuivering van huishoudelijk afvalwater en beheerders van waterkeringen. In 2002 waren dat alle waterschappen, hoogheemraadschappen en zuiveringsschappen, de provincies en het Rijk (i.c. het Rijksinstituut voor Zoetwaterbeheer en de Dienst Weg- en Waterbouw).

De waterbeheerders gebruiken de STOWA voor het realiseren van toegepast technisch, natuurwetenschappelijk, bestuurlijk juridisch en sociaal-wetenschappelijk onderzoek dat voor hen van gemeenschappelijk belang is. Onderzoeksprogramma's komen tot stand op basis van behoefteinventarisaties bij de deelnemers. Onderzoekssuggesties van derden, zoals kennisinstituten en adviesbureaus, zijn van harte welkom. Deze suggesties toetst de STOWA aan de behoeften van de deelnemers.

De STOWA verricht zelf geen onderzoek, maar laat dit uitvoeren door gespecialiseerde instanties. De onderzoeken worden begeleid door begeleidingscommissies. Deze zijn samengesteld uit medewerkers van de deelnemers, zonodig aangevuld met andere deskundigen.

Het geld voor onderzoek, ontwikkeling, informatie en diensten brengen de deelnemers samen bijeen. Momenteel bedraagt het jaarlijkse budget zo'n vijf miljoen euro.

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# SUMMARY


Conventional sewage treatment plants (STPs) based on activated sludge technology require a large footprint. This is caused by the relatively poor settling characteristics of activated sludge, resulting in low permissible dry solids (DS) concentrations in aeration tanks and in a low maximum hydraulic loads of secondary sedimentation tanks. In the nineties of last century, compact attached growth technologies in several configurations were developed (immobilised bed, fluidised bed and airlift reactors). Main feature of these continuously operated reactors is a high volumetric load, occasionally without a separate sludge/water separation step resulting in a small footprint. Recent laboratory research showed that in a discontinuously fed system, dense, more stable and smooth granular sludge can be grown without a carrier. The discontinuous feeding pattern also allows an efficient nutrient removal.

During recent laboratory research stable granulation was obtained with a three-minute pulse feeding period (total cycle time 3 hours), a synthetic influent (acetate ammonium mixture) and a high dissolved oxygen (DO) concentration. At a low DO (40% of saturation concentration) and a longer feeding period the granules became unstable and fell apart. It also was not possible to maintain granules under these conditions in a bubble column reactor. However, for a full-scale application of the aerobic granular sludge technology it is essential to allow longer feeding periods and operation at low DO. For economic reasons a bubble column is preferred to an airlift reactor. Therefore, this project aimed at elimination of these bottlenecks for scale-up and at evaluation of the economic and technical feasibility of the granular sludge sequencing batch reactor (GSBR). Laboratory research focussed mainly on the possibilities for simultaneous nutrient removal. Therefore, the period and way of feeding was varied and granulation at low oxygen concentrations was studied. The influence of the use of a bubble column on granulation was experimentally determined as well. Finally, formation of aerobic granular sludge with pre-treated sewage was investigated.

In order to provoke a “pulse feeding” it was proposed to feed during 33 % of the cycle time under anaerobic conditions. If substrate is not converted during this period, substrate concentrations would be similar to a pulse feeding operation when aeration starts. More-over, feeding influent in a upflow mode through the granular sludge blanket will allow biological phosphate removal bacteria to proliferate. In this case, substrate is taken-up in the cells during the anaerobic feeding period and is cell-internally stored as a polymer. It was shown that a 60 minutes feeding period of synthetic influent (total cycle time 3 hours) and a DO of 20%, caused a simultaneous COD (Acetate), phosphate and total nitrogen removal, (100% COD (acetate) and P removal, 100% ammonia removal by nitrification and 90% total N-removal). Biomass concentrations that can be maintained in this type of reactors was around 5 times higher than in an activated sludge system. This shows the potential of the technology.

In a separate experiment granulation on pre-treated (pre-settled) sewage as carbon source was tested. This experiment was successful in showing granulation. All biodegradable dissolved COD was removed from the influent. The possibilities of extensive nutrient removal from municipal wastewater will have to be investigated in follow-up research.

The feasibility study showed that the aerobic granular sludge technology seems very promising. Based on total annual costs both GSBR variants (GSBR with pre-treatment and GSBR with post-treatment) prove to be more attractive than the reference activated sludge alternatives (6-16%). From a sensitivity analysis it appears that the GSBR technology is less sensitive to the land price and more sensitive to a higher RWF. This means that the GSBR technology becomes more attractive at lower RWF/DWF ratios and higher land prices.
Because of the high permissible volumetric load the footprint of the GSBR variants is only 25% compared to the references. However, the GSBR with only primary treatment cannot meet the present effluent standards for municipal wastewater in the Netherlands, mainly because of an expected too high suspended solids concentration in the effluent.

A growing number of sewage treatment plants in the Netherlands is going to be faced with more stringent effluent standards. In general, activated sludge plants will have to be extended with a post treatment step (e.g. sand filtration) or be transformed into a Membrane Bio Reactor. In this case a GSBR variant with primary treatment as well as post treatment can be an attractive alternative.

Based on both the good research results and the positive technical and financial evaluation for full-scale application for sewage treatment, it is recommended to develop the aerobic sludge technology further by means of pilot research at a sewage treatment plant.

# STOWA IN BRIEF

The Institute of Applied Water Research (in short, STOWA) is a research platform for Dutch water controllers. STOWA participants are ground and surface water managers in rural and urban areas, managers of domestic wastewater purification installations and dam inspectors. In 2002 that includes all the country's water boards, polder and dike districts and water treatment plants, the provinces and the State.

These water controllers avail themselves of STOWA's facilities for the realisation of all kinds of applied technological, scientific, administrative-legal and social-scientific research activities that may be of communal importance. Research programmes are developed on the basis of requirement reports generated by the institutes participants. Research suggestions proposed by third parties such as centres of learning and consultancy bureaux, are more than welcome. After having received such suggestions STOWA then consults its participants in order to verify the need for such proposed research.

STOWA does not conduct any research itself, instead it commissions specialised bodies to do the required research. All the studies are supervised by supervisory boards composed of staff from the various participating organisations and, where necessary, experts are brought in.

All the money required for research, development, information and other services is raised by the various participating parties. At the moment, this amounts to an annual budget of some five million euro.

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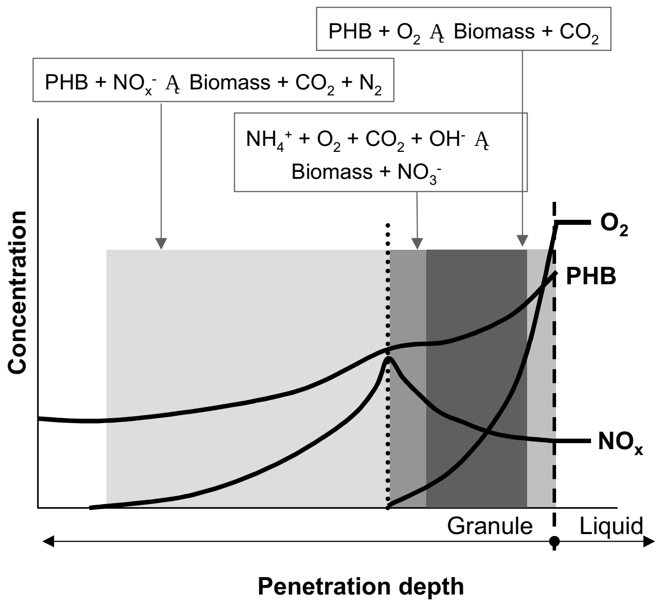
# 1 INTRODUCTION

Most wastewater treatment systems have the disadvantage of large area requirements and high excess biomass production. To build compact reactors, sludge can be grown in biofilms on carrier particles with high settling rates. The process conditions of aerated fluidised bed reactors are simple and the area requirement is small, because of the large specific biofilm area. Most of these compact biofilm reactors are continuously operated systems. An integrated settler on top of the reactor keeps the biofilm-covered particles in the reactor. Recent research showed the advantages of a discontinuously fed system and the possibility to grow stable granular sludge under aerobic circumstances (Beun, 1999). Full biological N- and P-removal can be easily integrated in a discontinuous system. Because of the high settling capacity of granules, the use of a traditional settler is not necessary and therefore, the installation can be built very compact.

**HISTORY OF THE PROJECT**

In previous research, a sequenced batch airlift reactor was used in order to obtain aerobic granular sludge (Beun, 2001). The used cycle configuration consisted of a pulse feed of the nutrient mixture, a long aeration period, a short settling period and an effluent withdrawal period. The total cycle time was three hours.

**FIGURE 1** SCHEMATIC CONCENTRATION PROFILES OF THE DIFFERENT COMPOUNDS IN THE GRANULE AND THEIR CONVERSION PROCESSES (PHB: POLY-β-HYDROXYBUTYRATE)




The first advantage of a discontinuous fed reactor is the possibility to select for particles with high settling velocities. This selection can be controlled through the duration of the settling phase. The particles which settle fast enough will be kept in the reactor, the slow settling floc structures will be washed out with the effluent. The second advantage is the high pulse COD concentration in the beginning of the cycle. Part of the COD will be used for aerobic growth, and part will diffuse into the granule and will be converted into poly-β-hydroxybutyrate (PHB) (feast period; period in which cell external substrate is available). When the COD in the solution is consumed, the organisms will use the slowly degradable storage product PHB as a carbon source for growth (famine period; starvation period in which cell external substrate is not available). The slowly degradable PHB leads to a low growth rate and therefore to a dense and smooth biofilm (Picioreanu, 1999). In the aerobic outer layer of a granule, PHB is converted with oxygen and nitrification occurs. Because of these processes, oxygen does not reach the inner part of the granule, in which nitrate and PHB will be present, leading to denitrification (figure 1). Preliminary modelling of this system showed that a dissolved oxygen concentration of 40% of air saturation would lead to a maximum nitrogen removal. Higher oxygen concentrations will lead to nitrate in the effluent, lower concentrations to ammonia (Beun et al., 2001).

**OBJECTIVES**

The main objective of this project was the development of the granule reactor technology, referring to optimal process conditions as well as the final technical design. The scale-up from laboratory to full-scale conditions played a central role during this research. The formation and the stability of granules under different circumstances were investigated. The circumstances were changed according to the scale-up criteria that followed from technical and economical design. Furthermore, the study of conversion processes in granular sludge was continued.

The research focused on the following subjects:

- Repetition of the granulation experiments and increasing the understanding of granulation;
- Effect of different feeding patterns and cycle configuration;
- Effect of DO concentration on granulation and conversion processes;
- Effect of the shear stress in the reactor (airlift or bubble column);
- Design and cost-calculations of a full-scale wastewater treatment plant based on aerobic granules.

In chapter 2 of this report, the used methods are described. Chapter 3 describes the different feeding patterns with a synthetic influent and their influence on the stability of the granules and the conversion processes in the granules. In chapter 4 the effect of varying oxygen concentrations will be presented with two different feeding patterns. In chapter 5 granulation in airlift reactors and bubble columns are compared applying a long feeding period with synthetic influent. In chapter 6 granulation on municipal sewage is described and in chapter 7 the economic and design part of the study is specified. Chapter 8 contains overall conclusions and recommendations.

# 2 METHODS

## REACTOR SET-UP

The experimental set-up is given in Figure 2. The reactors have a 3-litres working volume (diameter 6.25 cm; filled reactor height ±102 cm, total reactor height 130 cm), are double walled, contain an inner tube (in case of the sequencing batch airlift reactors (SBAR)) and an aerator to blow in the air at 4 l/s (= 87.5 m/hr). This airflow causes enough pressure difference between outer and inner tube to move the liquid and the granules. In case of the sequencing batch bubble column (SBBC) a larger sparger is used to distribute the gas bubbles more evenly and to obtain suspension of the particles. Effluent is extracted at a height of 50 cm, which means that after every cycle 1.4 litres remains in the reactor (an exchange ratio ($V_{effluent}/V_{filled\ reactor}$) of approximately 0.53).

## OPERATIONAL PROCEDURES

At start-up, the reactor was inoculated with activated sludge from sewage treatment plant (STP) Gouda. When a quick development of Bio-P activity was desired, sludge from STP Hardenberg was added as well, because of its high activity of phosphate accumulating organisms (PAO). During the start-up with pre-treated sewage (excessive pre-settling in the lab), the reactor was also inoculated with a small amount (<25 mg VSS) of crushed granules from the reactors operated with synthetic medium, in order to get a desired biomass composition, but to avoid the influence of granules on the overall granulation process.

The arrangement of the cycle is given in figure 3. The total cycle time was 3 hours and 2 hours in the experiment with pre-treated wastewater. A cycle consisted of a fill, aeration, settling and effluent withdrawal phase. The idle phase is not applied in the experiments. The reactors had the possibility to apply different filling regimes (figure 3). The influent was filled from the top (set-up 1) or from the bottom through the settled particle bed (set-up 2 and 3). During the aeration periods with controlled oxygen concentrations, air could be re-used via an off-gas circulation and a membrane pump (4 l/s). The system could also be operated without off-gas circulation and by applying pressurised air for high uncontrolled oxygen levels. The settling time (3 minutes) was chosen such that only particles with a settling velocity larger than 10 m/h were effectively retained in the reactor. In all systems, feeding was anaerobic and influent was either quickly fed from the top (first feeding pattern) or during a longer period from the bottom of the reactor and flowed in a plug-flow pattern through the bed (second feeding pattern).

The composition of the synthetic influent is given in table 1. The composition and selection of the pre-treated sewage water, which is used during the experiments, is given in chapter 6.

**TABLE 1 COMPOSITION OF THE DOSED MEDIA**

| Medium | component | final concentration in influent | remarks |
|---|---|---|---|
| Medium A: | NaAc | 371.5 mg COD/l | COD-load = 1.6 kg/m$^3$/day |
| | $MgSO_4 \cdot 7H_2O$ | 82.6 mg/l | |
| | KCl | 32.5 mg/l | |
| Medium B: | $NH_4Cl$ | 46.9 mg $NH_4$-N/l | COD/N ratio = 7.9 |
| | $K_2HPO_4$ | 12.5 mg $PO_4$-P/l | P-concentration = 20 mg P/l |
| | $KH_2PO_4$ | 6.3 mg $PO_4$-P/l | |
| | Trace element solution | 0.93 ml/l | |
| Water: | Tap-water | | total influent 1600 ml per cycle |

An overview of the experiments that are carried out is given in table 2. The operation time of the reactor under different circumstances was enough to reach a steady state if possible and operate the reactor for at least 3-4 sludge ages under steady state conditions.

TABLE 2 OVERVIEW OF THE LABORATORY EXPERIMENTS

| | Experiment | Reactor | Operation time | Remarks |
|---|---|---|---|---|
| 1a | Pulse feed, 100% DO (Stage I) | SBAR 1 | 143 days | Same experimental set-up as Beun |
| 1b | Pulse feed, 40% DO (Stage II) | SBAR 1 | 33 days | Continued from 1a |
| 1c | Pulse feed, 20% DO (Stage III) | SBAR 1 | 30 days | New start-up with low DO |
| 2a | Long feed, 100% DO | SBAR 2 | 233 days | First 103 days with nitrification inhibition by dosing ATU |
| 2b | Long feed, 40% DO | SBAR 2 | 184 days | Continued from 2a |
| 2c | Long feed, 20% DO | SBAR 2 | 250 days | Continued from 2b |
| 3 | Long feed, 40% DO | SBBC 1 | 385 days | |
| 4 | Pretreated sewage | SBAR 3 | 105ays | Details in table 6, chapter 6 |

**CONTROL**

The pH and oxygen concentration were measured continuously. The pH was controlled between pH 6.8 and 7.2 by the pH electrode and dosage of 1M NaOH or 1M HCl. The temperature was kept at 20 °C.

The DO concentration in the liquid was measured and controlled by dosing extra air or nitrogen gas in the off gas recycle flow via mass flow controllers (mfc, figure 2). This led to a stable DO in the reactor.

**ANALYTICAL PROCEDURES**

The $CO_2$ concentration in the off-gas was measured online with an infrared $CO_2$ analyser. Dry weight (TSS) in the reactor was determined by filtering and drying a sample for at least 24 hours at 105°C. Ash content was determined according to the Dutch standard method NEN6621 (NNI, 1982). Effluent biomass concentration was determined by the measured VSS (dry weight - ash content).

Solid retention time (SRT) is determined from the effluent biomass, which consists of granules and eroded material from the granule surface. Applying this method can lead to an underestimation of the sludge age in the inner part of the granule.

Bed volume of the settled bed was determined by reading the height of the biomass bed directly from the scale on the reactor at the end of a settling and effluent withdrawal period (in total 8 minutes settling time). Sludge volume index (SVI) was determined with the volume of the settled bed and the TSS concentration in the reactor. This unconventional method of SVI measurement was chosen to minimise the disturbance of the granules by leaving them in the reactor contrary to a conventional SVI measurement. Because of the compact structure of the granules, the SVI after 8 minutes of settling is equal to the SVI after an hour. Therefore, the SVI values obtained in this research can easily be compared with the ones obtained in activated sludge systems.

Concentrations of $PO_4^{3-}$, $NO_3^-$, $NO_2^-$, $NH_4^+$ and Acetate were measured regularly in filtered effluent samples and in filtered contents of the reactor after a filling period and occasionally every 10 to 20 minutes during one cycle (cycle measurement). The PHB fraction of the biomass was measured during the cycle measurements too. The influent concentration was calculated from the known medium concentration and additions. $PO_4^{3-}$ and $NH_4^+$ concentrations were measured with Dr. Lange test kits. $NO_3^-$ and $NO_2^-$ concentrations were regular-

ly checked with Merck indicator papers and measured spectrophotometrically. Acetate concentration of the filtered samples and PHB of freeze-dried granules were measured by gas chromatography.
The elemental composition of the granules (C, H, N, S) was measured by flash combustion in a partial oxygen atmosphere at 1,020°C using a Carlo Erba EA1108 elemental analyser.

Changes in morphology of the granules were followed by image analyses (IA), with a Lexmark Optra image analysis system. The average equivalent diameter (d) was measured, as well as the capriciousness of the particle surface (shape factor, 0=line, 1=circle) and the roundness of the particle (aspect ratio, 0=line, 1=circle). Also pictures were taken with this system to follow the development of the granules.

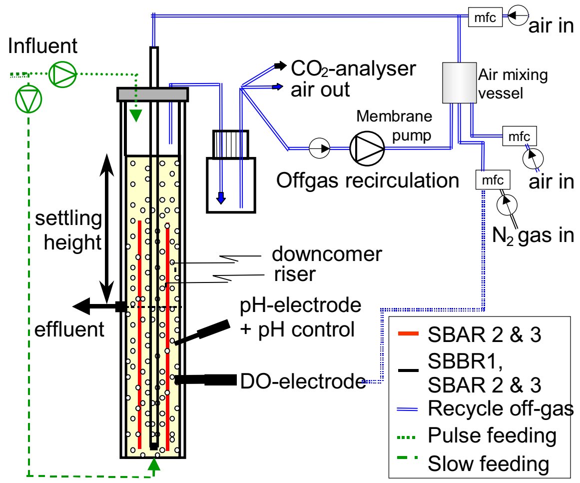



FIGURE 2 SET UP OF THE LABORATORY SCALE REACTORS

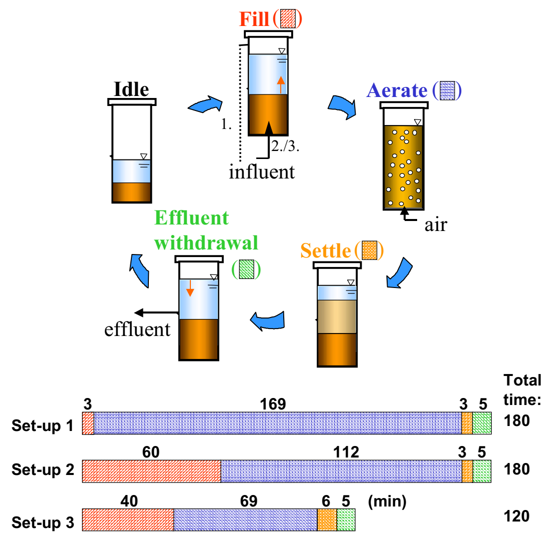


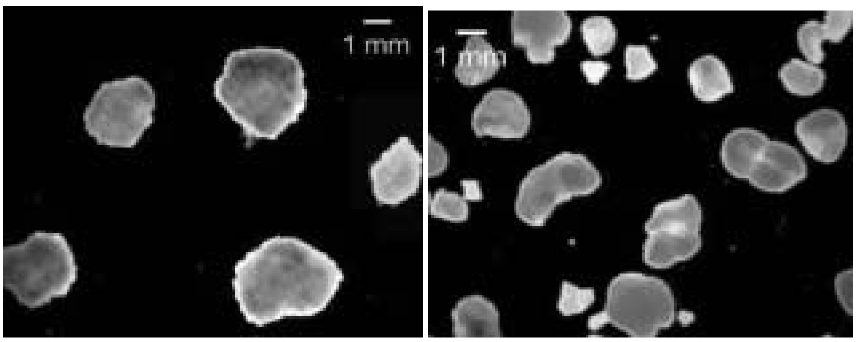

FIGURE 3 THE CYCLE CONFIGURATION AND DIFFERENT CYCLE TIMES FOR THE VARIOUS EXPERIMENTS

# 3 INFLUENCE OF FEEDING TIME

## 3.1 INTRODUCTION

In previous research (Beun, 2001, Etterer and Wilderer, 2001), aerobic granules were successfully grown in an airlift reactor with an anaerobic filling time of 3 minutes. This 3 minutes pulse feed was supposed to be a necessity for the development of smooth and dense particles. When aeration starts after the pulse feed and at the beginning of the aerobic period, the acetate concentration in the liquid is high enough to penetrate throughout the whole granule and to be partially converted in the slowly biodegradable storage polymer polyhydroxy-β-butyrate, or PHB (feast period). When all acetate is consumed, partly directly for growth (10-40%), partly for PHB production (60-90%), the organisms will use the stored PHB for growth and maintenance (famine period) (Beun et al., 2000). The conversion rate of PHB by heterotrophic bacteria is much lower than the conversion rate of easy biodegradable acetate. From earlier biofilm research it is known that a low conversion rate and thus low actual growth rates lead to denser and smoother biofilms than at a high conversion rate (Picioreanu, 1999, Van Loosdrecht, 1995). Formation of slowly degradable PHB out of rapidly degradable acetate is one of the most essential mechanisms for the production of dense, smooth and stable granules.

FIGURE 4 GRANULES GROWN WITH A SHORT FEEDING TIME (A) AND WITH A LONG FEEDING TIME (B).




During the aeration period of a cycle, oxygen will only partly penetrate into the granule, since it is consumed by the heterotrophic and autotrophic organisms in the outer layer of the granule. Inside the granule, oxygen is absent, but nitrite and nitrate, produced by autotrophic organisms in the outer layer, can diffuse into the core of the granules. In this zone PHB is available for the denitrifying growth of the heterotrophic organisms present in this part of the granule. This mechanism of the oxygen limited granule, instead of the substrate limitation as exists in continuous fed (biofilm) systems, plays a crucial role in simultaneous COD and nitrogen removal as exists in the aerobic granule system. This oxygen-limited system is only achieved when the acetate is fed as a pulse and the concentration of acetate in the reactor is very high at the beginning of the aeration period.

The experiment of Beun, with a 3-minute pulse feed of acetate, was successfully repeated and the granules were stable for 270 days, after which the experiment was stopped. As in the experiments of Beun (Beun et al., 2002), COD was totally removed and nitrogen removal was 62%, with oxygen saturation during the famine period. Also the start-up of the reactor was repeated several times, leading to stable granules after 60 days. After inoculation with nitrifying organisms, the ammonia concentration in the effluent started to decrease.

## 3.2 BOTTLENECK OF SCALE-UP

The pulse feed is seen as an important prerequisite for granulation. However, at full-scale this 3 minutes feeding period would imply very large pumps. As an example, for a 120,000 p.e. (Population equivalent) STP (see Chapter 7), this would mean pumps with a maximum working capacity of 5.7 $m^3/s$, which only work for three minutes per hour in a three reactor system or to work with much more and smaller reactors. Also large storage tanks would be needed and since one of the main goals of this system is surface area reduction, the use of storage tanks has to be avoided. All these approaches would increase the investment and production costs as well as the complexity of the system.

To be able to use smaller pumps and to avoid the need for storage tanks, a system with three reactors and a three-hour cycle was designed. In this configuration, the feeding time should be one hour. If the first reactor is filled, the second and the third are in the other phases and at the end of the filling phase of reactor 1, the influent flow can be switched to reactor two, and so on. This one hour filling time would decrease the maximum pump capacity drastically, but could disturb the pulse feeding character.

To maintain the concept of pulse feeding, an anaerobic one hour filling period was proposed, followed by an aeration period of 112 minutes and a short settling and effluent withdrawal time (respectively 3 and 5 minutes). In this design, no conversions would take place during this first hour filling. When the aeration is switched on, the granules and influent would be mixed and the reactor would experience a pulse feed. In practice, phosphate-accumulating organisms (PAO) will become dominant in the granules, which convert acetate during the anaerobic period to cell internally stored polymers, also leading to a low specific growth rate and thus to stable granulation (chapter 3.3.2).

To investigate simultaneously the possibility of no effluent extraction point, but pressing the effluent out of the reactor with the incoming influent, the reactor was fed from the bottom. With the low flow rate, a plug-flow of influent was created through the bed, which pushed the effluent from the pores to the upper part of the reactor. A colour tracer in the influent showed that no mixing occurred between the treated wastewater and the new influent. In the lab-scale set-up, an effluent withdrawal from the reactor at 50 cm, was maintained, because of practical reasons (maintaining a closed system, off-gas circulation).

## 3.3 RESULTS

The SBAR was operated for over 233 days, with a 60 minutes feeding period, in which the influent entered the reactor in a plug flow through the packed bed of granules. In this period oxygen was not controlled and reached full saturation during aeration. The first 65 days of the experiment were used for start-up of the reactor and to reach a steady state. The second period (day 87-day 117) was used for cycle measurements with inhibited nitrification, which was done by dosing allylthiourea (ATU). The last period was used to obtain full nitrification in the reactor and to observe whether the granules maintained stable.

### 3.3.1 GRANULATION

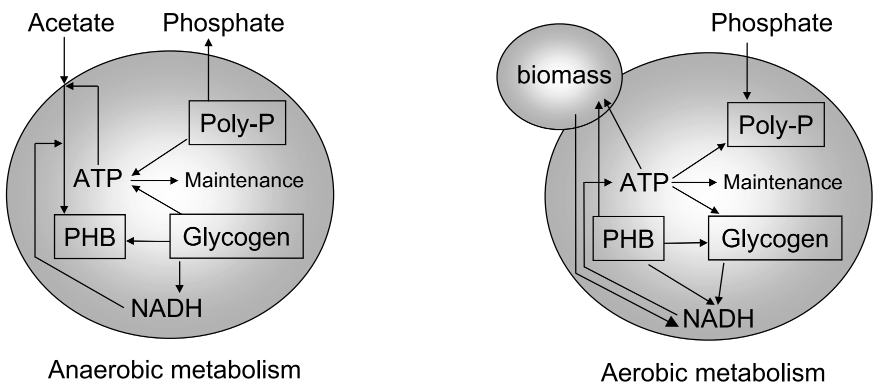
The granulation process during start-up of the reactor was comparable to the process with a 3 minutes pulse feeding. After 4 days, the first granules appeared. These first granules had many filaments at the surface, but after 9 days, smooth granules started to appear. After 40 days, all granules were smooth, regular granules, although a few very big granules were observed (diameter 6 mm). From day 68, the granule size started to decrease again to a minimum average size of 1.1 mm (figure 4). Small particles, which have good settleability characteristics, are very advantageous for the system, because this means a large biofilm surface area and therefore a higher substrate uptake rate per litre reactor volume. Another advantage is the small biofilm depth and therefore a larger active part in the granule. However, the particle size cannot be controlled in the reactor yet and more research is needed to find out which aspects influence this granule size and more important how to control these. The average sludge volume index was 27 ml/g, which is lower than the SVI as was found in the experiments with the pulse feed of Beun (39 ml/g).

The density of the granules increased from 38 gram TSS per litre of biomass to 121 g/l at the end of the experiment. The total biomass concentration increased to 8.5 g VSS per litre, with an ash content of 41% (TSS = 14.3 g/l). The density and VSS concentration were similar to the experiments of Beun, but the ash content was much higher (17% in Beun's experiments). Due to the anaerobic feeding phase, phosphate-accumulating organisms become dominant in the system. The stored poly-phosphates result in these high ash contents in the granule. The SRT was similar to the previous experiments, fluctuating around 40 days in the steady states of both experiments.

### 3.3.2 CONVERSION PROCESSES

The anaerobic feeding through the bed, combined with aerobic periods for growth and ammonium removal, are circumstances in which phosphate accumulating organisms (PAO) are favoured. PAO are capable of taking up easy degradable substrates as acetate, under anaerobic conditions and store them as polyhydroxyalkanoates (PHA), like PHB and to lesser extent PHV (polyhydroxyvalarate). The energy of this transport and storage is supplied by the hydrolysis of intracellular stored poly-P to orthophosphate, which is released from the cell into the liquid. The reduction equivalents required for the conversion of acetate to PHB are supplied by conversion of glycogen, which is stored during the aerobic period from PHB (Mino et al., 1995, Smolders, 1995, Brdjanovic, 1998). In the aerobic or anoxic phase, without the presence of an external substrate, PHA is used as substrate for cell growth, polyphosphate synthesis and glycogen formation. The anaerobic and aerobic metabolisms are schematically shown in figure 5. Since the acetate is fed through the packed bed, the PAO are capable of anaerobically consuming the acetate and using their storage products during the aerobic period. Besides PAO, glycogen-accumulating organisms (GAO) can be present in an anaerobic/aerobic system, because of their capability of converting acetate during the anaerobic period without releasing phosphate. When the aeration is started, no acetate is left for the heterotrophic organisms. Therefore, only organisms capable of anaerobic acetate storage can grow in this system.

**FIGURE 5** **METABOLISM OF THE PHOSPHATE ACCUMULATING ORGANISMS (SMOLDERS ET AL., 1994A).**




After 53 days of operation, the first decrease in effluent phosphate concentration was observed and after 68 days, the phosphate in the effluent was around zero (Figure 6). Phosphate stored as poly-P was removed with biomass present in the effluent. The biomass growth and thus the P-removal should be sufficient to take up the extra influent phosphate. Also all dosed acetate was consumed after the anaerobic period. The phosphate release per acetate taken up was 0.86 g P/g C. Smolders reports a ratio of 0.67 to 1.96 g P/g C, depending on the pH (pH ranging from 5.5 to 8.5) (Smolders et al., 1994b). These ratios were measured with batch tests with sludge from a SBR in which only PAO were present. The value measured in the aerobic granular sludge was within this range, but concerning the pH range the ratio is low. It is most likely that other organisms were present, like GAO. The average growth yield of the biomass was 0.16 g VSS/g COD (or 0.27 g SS/g COD), which was calculated from the biomass increase divided by the acetate consumption (the effluent VSS concentration, plus the increase of the biomass concentration in the reactor). This value is much lower than the observed growth yield during operation with a short feeding time (0.3 g VSS/g COD or 0.39 g SS/g COD). In literature the same yield values for PAO are reported, for example 0.23-0.27 g TSS/g COD for a denitrifying Bio-P biofilm system (Falkentoft, 2000) and 0.25-0.3 g SS/g COD for a denitrifying bio-P system and 0.35-0.4 for an aerobic bio-P system (Kuba et al., 1993). The low growth yield at the SRT of 40 days was sufficient for the complete removal of 20 mg P/l from the influent. Biomass wash-out form the reactor and thus phosphate removal partly occurs via fines, eroded and sloughed of the granules by shear forces and particle-particle collisions. It is important that the phosphate content of these fines is high enough to remove the surplus phosphate. The effluent concentrations of phosphate in solution showed that these fines contained enough phosphate to remain sufficient phosphate removal.

Nitrification was inhibited with ATU during the first 100 days in order to allow a stable bio-P build-up. After the ATU dosing was stopped it took 39 days to reach full nitrification (no nitrite in the effluent). The high nitrate concentrations in the effluent were due to insufficient total nitrogen removal (average 24%) at these high oxygen concentrations (chapter 4). Also no denitrification occurred during the feeding time, since the plug-flow feeding pattern, so hardly any mixing between COD from the influent and nitrate still present in the reactor occurred.

**FIGURE 6 THE PHOSPHATE CONCENTRATION IN THE EFFLUENT (◆) AND AFTER THE FEEDING PHASE, T=60 ( r )**

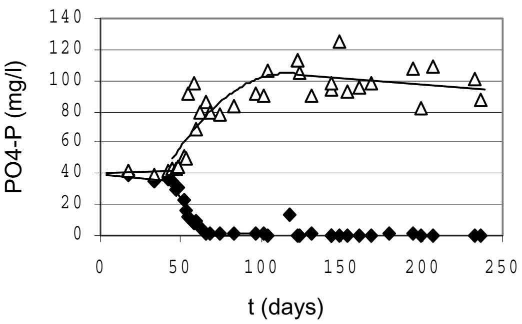
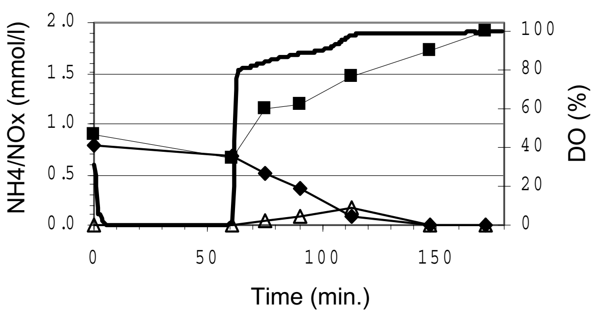



**FIGURE 7 A CYCLE MEASUREMENT OF THE AIRLIFT REACTOR AT DAY 199 WITH 60 MINUTES FEEDING. DO (-); $NH_4$ (◆), $NO_2^-$ ( r ), $NO_x$ (■),**




## 3.4 CONCLUSIONS

Operation of the reactor with a long feeding time through the settled bed of granules appeared feasible and is favourable over a short feeding time. The two most important reasons are:

The application of a long feeding phase at large scale treatment plant, will simplify the batch scheduling, the influent dosing and will make a storage tank not necessary.

Due to the alternate aerobic and anaerobic phases, PAO will outcompete fast growing heterotrophic organisms, leading to stable granules and full phosphate removal.

It can be said that the long anaerobic feeding period, most favourable with a plug-flow regime, is very promising for the future application of the system at large(r) scale.

# 4 INFLUENCE OF THE DISSOLVED OXYGEN CONCENTRATION

## 4.1 INTRODUCTION

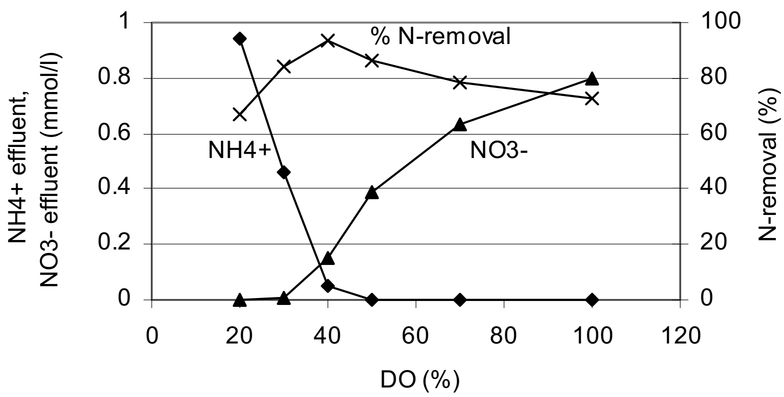
The distribution of biomass inside the biofilm or granule affects the performance of the processes taking place in the reactors. In aerobic biofilm systems, where COD removal and nitrification take place simultaneously, heterotrophic and autotrophic micro-organisms compete inside the biofilms for space and oxygen. When cell-external substrate (acetate) is present during the aerobic period, fast growing micro-organisms (heterotrophs) will be located in the outer layers while slow growing micro-organisms (autotrophs) will be confined to deeper zones of the biofilm with lower oxygen availability (Van Loosdrecht et al., 1995; Okabe et al., 1996). In the core of the granule, the oxygen concentration will be zero, but the nitrate produced by the autotrophic organisms will be available as electron acceptor for the heterotrophic organisms located in the deeper layers (figure 1, chapter 1).
In a system with an anaerobic feeding period and an aerobic reaction period phosphate accumulating organisms (PAO) become dominant (chapter 3). Contrary to the first mentioned granule system, in such system the aerobic outer layer can consist of a mixed system of PAO and autotrophs, because of their similar growth rates (respectively 1 – 1.5 $d^{-1}$ and 0.8 - 1 $d^{-1}$ (Brdjanovic et al., 1998, Henze et al., 1999). PAO can be divided into two different categories (Meinhold et al., 1998):

- Strict aerobic, unable to utilise nitrogen (nitrate or nitrite) as electron acceptor;
- Facultative aerobic (able to use oxygen or nitrogen as electron acceptor);

The first group can only be present in the outer layer of these phosphate-accumulating granules, the last group can also be present in the deeper layers.

Granular sludge growing at a high oxygen concentration shows a good COD and ammonium removal, but the nitrogen (nitrate) removal is insufficient (average of 24%). The type of feeding (long or short anaerobic feeding) has no influence on this removal capacity. In Dutch practice, larger STPs need to comply with an effluent standard of 10 mg total N/l (as an annual average). These standards might become stricter in the future, requesting even higher N-removal efficiency. With an average influent total Kjeldahl Nitrogen (TKN) of 55 mg/l, a total N-removal efficiency of 81% is needed. This can be partly done by a pre- or post treatment, but a large part should be removed by biological processes (>95% of the 45.7 mg/l dissolved N-kj) (chapter 7).

FIGURE 8 $NO_3^-$ AND $NH_4^+$ CONCENTRATION IN THE EFFLUENT AND PERCENTAGE OF N-REMOVAL AS A FUNCTION OF THE DO LEVEL IN ONE CYCLE AS CALCULATED BY BEUN



An increase in nitrogen removal in a granule system can be achieved by lowering the dissolved oxygen concentration in the reactor. Preliminary modelling of the system with a pulse feed (no PAO present), showed that a dissolved oxygen concentration of 40% of air saturation would lead to a maximum nitrogen removal. Higher oxygen concentrations might lead to nitrate in the effluent, lower concentrations to ammonia (figure 8, Beun et al., 2001).

For practical and economical reasons, a lower oxygen concentration is needed to reduce energy costs. It is however unclear whether a low DO concentration has negative effects on granule formation. Therefore, the long-term DO effect on granule formation and stability was investigated. Both modes of feeding patterns were included in the study.

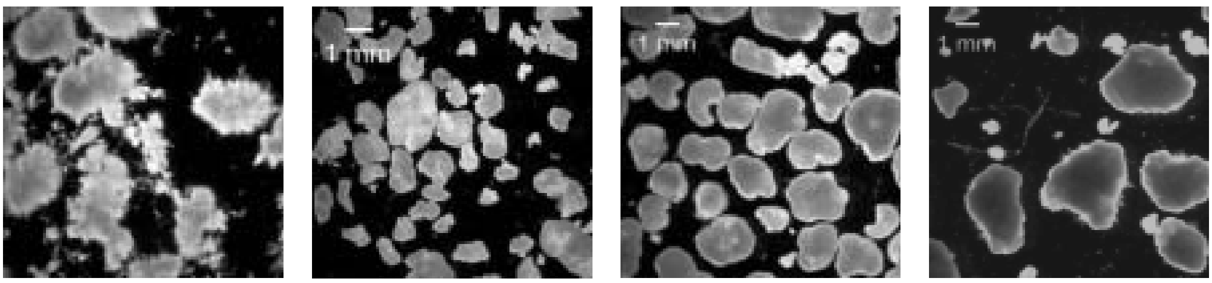
## 4.2 LOW DO AND PULSE FEED

### 4.2.1 GRANULATION

A reactor was operated during 143 days (Stage I) under the same operational conditions as described by Beun *et al*., 2000a, except for the feed composition (lower COD/N ratio; 7.9 instead of 14.3 as used in the experiments of Beun). The results (conversions and granulation) obtained during this first period were similar as the results Beun obtained and are described in chapter 3.

After 143 days the DO concentration was reduced to 40% of oxygen saturation (Stage II) in order to obtain higher nitrogen removal. After ten days, growth of filamentous bacteria was observed on the surface of the granules. From day 174 the granules started to disintegrate, granule size and shape factor decreased and biomass loss was observed (Figure 9a). The SVI was around 48-75 ml/g TSS (Stage I) reaching a maximum value of 100 ml/g TSS at the end of Stage II. Although for flocculent sludge this might be good, under the conditions in the granular sludge reactor the sludge is slowly washed out. This change in the granule structure can be attributed to limitation of oxygen in the inner parts of the granule, where it was previously available operating with DO 100%. These observations are in line with results reported by Picioreanu (1999). He reported that stronger substrate gradients (like when the DO concentration is reduced) lead to less compact biofilms. The biomass washout caused a decrease of biomass concentration in the reactor from 5 to 3 g VSS/L. At the end of the experiment, the sludge retention time (SRT) decreased sharply from 25 days to 8 days. The experiment was stopped at this point.

FIGURE 9 GRANULES AT DIFFERENT DO CONCENTRATIONS; SHORT FEEDING PERIOD AND DO 40% (A); LONG FEEDING PERIOD WITH DO 100% (B), DO 40% (C) AND DO 20% (D)



In a second experiment, the reactor was started up again with a 3-minute pulse feed, but with a DO concentration of 40% of the saturation value (stage III). The maximum biomass concentration reached was 0.9 g VSS/l, so the COD load per gram of biomass was much higher compared to former experiments. According to previous studies increasing the loading rate leads to weaker and less dense biofilms (Picioreanu, 1999, Kwok et al., 1998). Combined with a stronger oxygen gradient, this apparently prevents the formation of granular sludge. The SVI was very high (200 ml/g TSS) and the density of the particles was only 13 g $TSS/L_{granules}$. After 1 month the reactor was stopped due to the apparent impossibility of formation of stable granules under these conditions.

### 4.2.2 CONVERSION PROCESSES

At high DO, ammonium removal was 100%. However, the total N-removal was only 8%, presumably due to biomass formation. After the reduction of the DO to 40% (Stage II), first the denitrification efficiency increased and $NO_x^-$ was almost completely depleted, while the nitrification efficiency decreased and ammonia was present in the effluent. At the end of this Stage II, the ammonia was again completely nitrified, but nitrate and nitrite were not fully denitrified, although the N-removal capacity was higher (37%) than at DO 100%.

## 4.3 LOW DO AND LONG ANAEROBIC FEEDING

The experiment with a long feeding period, showed a population shift towards phosphate accumulating organisms (PAO) (chapter 3). Since these organisms store substrate during the anaerobic period, there is no influence of oxygen on the conversion from easily biodegradable substrate to slowly biodegradable PHB, as in the system with a pulse feed under aerobic conditions. Therefore, the influence of DO on granule stability is expected to be of less importance. The experiments as described in chapter 4.2, were repeated with a long feeding period.

The airlift reactor with a long feeding time has been operated as a stable system for 234 days at a DO concentration of 100% (chapter 3), for 184 days at 40% and for 207 days at 20%.

### 4.3.1 GRANULATION

A decreased oxygen concentration in a system dominated by PAO's had only small effects on the granule characteristics. These characteristics are given in table 3 together with the results of the pulse feed experiment.

The average dry weight in the reactor increased with lower DO concentrations; 12 g VSS/l at DO 40% and 16 g VSS/l at DO 20%. The ash content of the particles was stable and around 34% (a sludge content of 25 g TSS/l). The increase in dry weight in the reactor showed a dependency with the SVI and the granule diameter. In time, the settled bed filled up to the effluent withdrawal. In the present experimental set-up, the excess sludge produced was washed out with the effluent. The lower the particle diameter, the more particles fit in the reactor in the settled bed (packing of the particles). This causes also a decreased SVI and the lower the SVI (18 ml/g at DO 20%), the less biomass washed out. This led to an increase in total dry weight in the reactor.

The granule structure (size and shape) was changing in time and could not be controlled as such. However, the granules were stable without fluffy and filamentous structures, as observed with the short feeding time (figure 9). In general, it can be concluded that to be able to apply lower oxygen concentrations in this aerobic granule system, selection for slow growing organisms has to take place, which can be controlled by the feeding regime. Only then the granule structure will be dense and smooth. The reasons for the change in particle size distribution and particle shape of dense and smooth particles have to be investigated in later research.

### 4.3.2 CONVERSION PROCESSES

Table 3 illustrates the measured nitrogen removal at different DO concentrations. At lower DO concentrations, the nitrate concentration in the effluent decreases and the ammonia concentration in the effluent increases slightly, but remains low in all cases. After each change in DO concentration, it took approximately one sludge-age to react on the changed situation. Fist ammonia started to appear in the effluent, which was replaced by nitrate when the oxygen concentration remained low and population in the granule started to shift. The lowest concentration (D=20%) showed the same pattern, with the main difference that the nitrate concentration in the effluent remained low and the total N-removal in the system was 90% on average over the last 100 days of operation (Figure 10). The model of Beun predicted an increased amount of ammonia in the effluent at DO 20% and an optimal N-removal capacity at DO=40%. This model was based on a 3 minutes pulse feed and thus on a clear heterotrophic outer-layer of the granule with high growth rates. Since the heterotrophic PAO and the autotrophic nitrifyers have a comparable growth rate, these organisms are likely to be mixed in the outer layers and thus oxygen is more readily available for nitrification at DO 20%. The inner part of the particle uses enough nitrate for the conversion of PHB to maintain low total-N concentrations in the effluent.

The SRT of the granules was 37 days at DO 100%, 64 days at DO 40% and at DO 20% this value increased to 145 days, after which it decreased to 87 days. Because of this very high SRT, not enough phosphate was washed out with the effluent biomass and a 96% P-removal could not be maintained. With an increasing sludge age, the P-removal decreased to 40%. As soon as the SRT decreased again, the P-removal increased to 94%. This shows that in order to obtain a long-term sufficient phosphate removal, the sludge age has to be controlled by removing appropriate quantities of granules from the system.

**FIGURE 10** THE EFFLUENT $NO_x$ ( r ) AND $NH_4^+$ ( Ï ) AT DIFFERENT DISSOLVED OXYGEN CONCENTRATIONS AND THE TOTAL NITROGEN REMOVAL IN PERCENTAGE OF THE INFLUENT AMMONIA ( - )

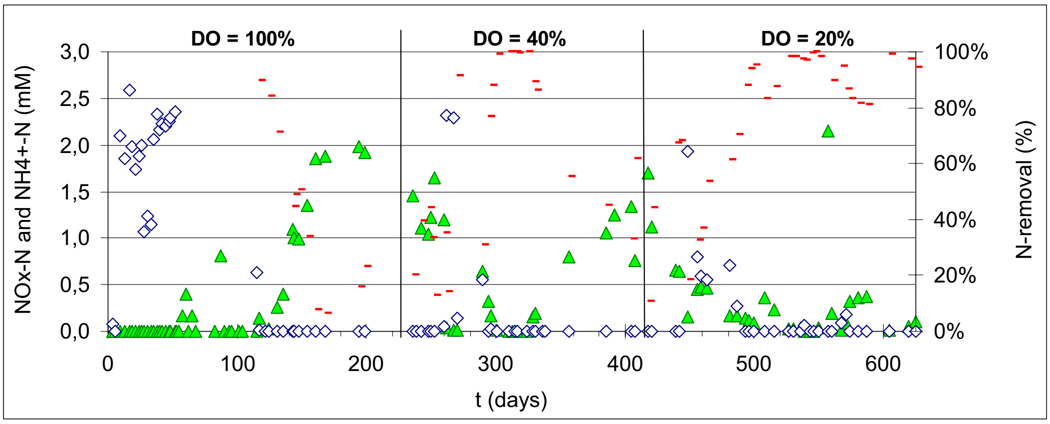



**TABLE 3** RESULTS OF THE EXPERIMENTS WITH DIFFERENT OXYGEN CONCENTRATIONS AND FEEDING PHASES (3 AND 60 MINUTES)

| Averaged values | Short feeding period | | | Long feeding period | | |
|---|---|---|---|---|---|---|
| | DO 100% (I) | DO 40% (II) | DO 40% (III)[2] | DO 100% | DO 40% | DO 20% |
| Feast phase (min) | 15.5 | 20 | 160 | -- | -- | -- |
| Bed volume (l) | 0.8 | 0.9 | 0.6 | 1.1 | 1.1 | 1.3 |
| SRT (days) | 25 | 8 | ND | 37 | 60 | 114 |
| Density granules (gTSS/lgranule) | 53 | 60 | 12.6 | 95 | 71 | 78 |
| Equivalent diameter (mm) | 1.56 | ND | 5.0 | 1.28 | 1.12 | 1.05 |
| gVSS/l (reactor) | 5.1 | 3 | 0.9 | 8.5 | 12 | 16 |
| GTSS/l (effluent) | 0.07 | 0.10 | 0.14 | 0.07 | 0.05 | 0.05 |
| SVI (ml/gTSS)[1] | 50 | 100 | 200 | 24 | 20 | 18 |
| mg TOC/l (influent) | 105 | 105 | 105 | 105 | 105 | 105 |
| mg TOC/l (effluent)[1] | 20 | 20 | ND | 27 | 12 | 20 |
| mg $NH_4^+$-N/l (influent) | 48 | 48 | 48 | 48 | 48 | 48 |
| mg $NH_4^+$-N/l (effluent) | 0 | 1 | ND | 0 | 0 | 0.1 |
| mg $NO_3^-$-N/l (effluent) | 35 | 21 | ND | 24.8 | 17 | 3 |
| mg $NO_2^-$-N/l (effluent) | 0 | 1.5 | ND | 0 | 0 | 0 |
| mg $PO_4$-P/l (influent) | -- | -- | -- | 20 | 20 | 20 |
| mg $PO_4$-P/l (effluent) | -- | -- | -- | 0.56 | 0.55 | 3.9 |

ND Not detected.

[1] after 5 min of settling.

[2] Values taken from the period when only granules where present but very variable due to wall growth.

## 4.4 CONCLUSIONS

Operation of the reactor at a low dissolved oxygen concentration is only feasible when a selection of slow growing organisms takes place. This means that a long anaerobic feeding time should be applied, in order to favour the growth of slow growing PAO over the growth of fast growing heterotrophic organisms. With a long anaerobic feeding period, stable granules will be formed.

Nitrogen removal increases with decreasing oxygen concentrations. At the lowest tested DO (20%), N-removal was sufficient to reach the effluent standard (total N<10 mg/l). In the effluent, only nitrate was present and ammonia could not be detected. This means in practice, even lower oxygen concentrations might be feasible. The possibility of applying low oxygen concentrations will reduce aeration and power costs.

Phosphate removal should be maintained by controlling the sludge age or solid retention time. If the SRT becomes too high, phosphate will not be properly removed.

It can be said that a low oxygen concentration combined with a long anaerobic feeding period is very promising for the future application of aerobic granular sludge at large(r) scale.

# 5 INFLUENCE OF THE REACTOR CONFIGURATION

## 5.1 INTRODUCTION

The experiments with aerobic granulation were performed so far in sequencing batch airlift reactors (SBAR). The disadvantage of this reactor type is the complicated and more expensive construction on large scale, when it is compared to a bubble column or aeration tank. Beun (1999) showed that in a sequencing batch bubble column (SBBC) fed with a three-minute pulse feed, less dense biofilm structures (ρ = 12 g TSS per litre of granules) were obtained. In an SBAR biofilm densities of 60 to 120 g X/l biomass were obtained. Stratification of granules in the bubble column caused a lower shear, originating from particle-particle interactions, in the upper part of the column. The fast growing organisms in the outer layer of the granule have the chance to form filaments and, therefore, less dense biofilms. This effect was intensified, because these less dense particles remained in this upper part of the reactor and were exposed to decreasing shear stress (Beun et al., 1999). In an airlift reactor, this effect does not take place, because of the excellent mixing of the particles in these type of reactors and the higher shear forces on the granules. Since the long feeding phase the selection for slow growing phosphate accumulating bacteria, the need for shear becomes less evident.

In order to investigate the necessity of an airlift reactor (the need of a riser) to generate stable and dense granules, a SBBC was started up. All process variables were kept the same as in the former experiments (chapter 3 and 4), only the riser was removed from one of the reactors.

## 5.2 RESULTS

### 5.2.1 GRANULATION

A SBBC was started up with a dissolved oxygen concentration of 40%. The granulation process in the SBBC was slower than in the SBAR. A week after start up in a SBAR, the first granules are observed and after 2 weeks they have an average diameter of 1 mm. The granules stabilise after 40 days at 1.3 mm (average diameter fluctuating between 1.1 mm and 1.6 mm at DO 100% and a maximum diameter of 1.2 mm at DO 40%).

In the SBBC, the first month the granules tend to form flocs during settling. After 27 days it was observed for the first time that the biomass only consisted of very small separate granules (average equivalent diameter was 0.433 mm). After 2 months, the particles reached an average diameter of 1 mm. The granules increased in diameter to an average of 1.3 mm after 3 months (fluctuating between 1 and 1.3 mm). The shape factor[1] of the granules varied between 0.67 and 0.74 and the aspect ratio[2] between 0.70 and 0.76, which is comparable to the values obtained with the SBAR (respectively 0.59 to 0.73 and 0.67 to 0.73).

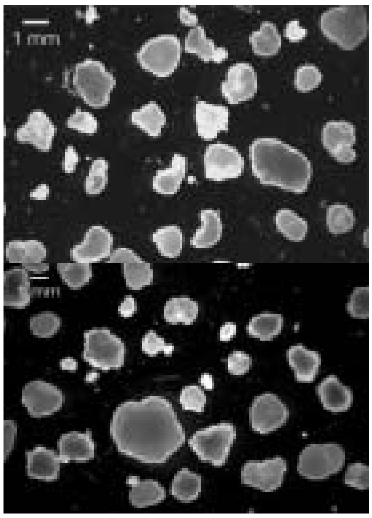



FIGURE 11 STABLE GRANULES IN A SBAR (A) AND IN A SBBC (B).

[1] 0 = very irregular structure; 1= smooth and round structure
[2] 0 = line; 1 = round

Table 4 shows the summary of results of the granules described above. All results show great similarity to the results of other researchers in their biofilm studies and what is stated in theory (Picioreanu, 1999, Tijhuis et al., 1995). By applying a long anaerobic feeding period to the granules, selection for slow growing phosphate accumulating bacteria takes place, resulting in dense, smooth and thus well settling granular structures. Therefore, the strong shear stress of the airlift is not necessary when the long feeding period is applied.

**TABLE 4** **DIFFERENCES IN GRANULE CHARACTERISTICS BETWEEN THE TWO REACTOR TYPES, FEEDING PATTERNS AND DISSOLVED OXYGEN CONCENTRATIONS**

| **Particle characteristics** | **pulse feed** | | | **long feeding period** | | |
|---|---|---|---|---|---|---|
| | **SBAR** | | **SBBC**[1] | **SBAR** | | **SBBC** |
| | DO 40% | DO 100% | DO 100% | DO 40% | DO 100% | DO 40% |
| Stability of granules | not stable | stable | n.a. | stable | stable | stable |
| Shear in reactor | high | high | low | high | high | low |
| Average growth rate | high | high | high | low | low | low |
| Average diameter (mm) | 5.0 | 1.56 | 2.0 | 1.12 | 1.28 | 1.14 |
| Biomass density in the particle (g TSS/ l biomass) | 13 | 53 | 12 | 71 | 95 | 101 |
| Dry weight in reactor (g VSS/l reactor) | 0.9 | 5.1 | 3 | 12 | 8.5 | 13 |
| SVI (ml/g) | 200 | 50 | n.a. | 20 | 24 | 19 |

[1] (Beun, 1999b)

### 5.2.2 CONVERSION PROCESSES

The application of a SBBC instead of a SBAR has no influence on the conversion processes of the granules. The acetate is totally converted after the anaerobic feeding period. The average phosphate removal is 93%. The N-removal was not very stable because of problems with the oxygen control, but reached values of 95% and was on average 83% during the stable periods. Because of the DO concentration of 40%, the nitrogen was present as $NO_3^-$ in the effluent.

## 5.3 CONCLUSION

The only difference in granulation between a bubble column and an airlift reactor was found in the duration of the start-up period, granule characteristics were highly comparable. As long as a slower start up of the reactor is not a problem, the use of a SBBC, combined with a long feeding period is feasible and has no negative influence on the granulation. Extra attention has to be given to the mixing of the granules and to the stability of the conversion processes.

# 6
# EXPERIMENT WITH PRE-TREATED SEWAGE

## 6.1 INTRODUCTION

The past results of aerobic granular SBAR and SBBC, as described in the former chapters, have been very promising for the use of this type of reactor in practice. Simultaneous full COD removal, high phosphate removal and high nitrogen removal have been reported. Also size and shape of the granules changed in time, but these characteristics were stable at different oxygen concentrations, when a long anaerobic feeding period was applied. The tests have been carried out with the use of synthetic influent, consisting of an acetate/ammonia mixture. No complex substrate and no suspended solids were present in the influent.

The next step of the design of a full-scale reactor is a pilot-phase, but before the investments of a pilot scale can be made, a test with presettled wastewater has to be performed to investigate the formation of granules on more complex influent. The behaviour in the reactor of suspended solids from the sewage is not investigated at this scale, since wall effects and shear stress will be too different from a full-scale installation.

## 6.2 SELECTION OF THE INFLUENT

### 6.2.1 PRE-TREATMENT

For the experiment with municipal wastewater, an influent was used that is very close to an actual pilot or full-scale situation, since this will give the best insight in the feasibility of the system. Wastewater was collected from the STP, right after the screen (5*25 litre per batch). The tanks were stored in a 4°C room, where the settlable suspended solids settled in the tanks. Once per two days, 25 litres of this influent was restored in a cooled storage tank (10°C) near the reactor. Per cycle 1.5 litres were pumped into a 2-litre storage tank to warm up to room temperature and subsequently dosed to the reactor. In this way a good pre-settled influent is prepared, without the use of flocculating agents. Pre-treatment of the raw wastewater was chosen, to avoid clogging of the tubes and other reactor parts. The use of flocculating agents was avoided, since their influence on granulation is not known and their use might lead to different results. In this experiment this was the step from relatively simple to more complex influent.

### 6.2.2 SELECTION OF THE RAW SEWAGE

During the first period (day 0 – 28) of the experiment, wastewater from STP Gouda (pipeline of Stolwijk) was used. The influent concentrations analysed during the fist period of this experiment showed a much more diluted wastewater than was expected.
A new influent with higher COD values was taken from the treatment plant of Berkel. This influent is applied from day 29 of the second start up. The total biodegradable COD load is still 31% lower than with the synthetic influent.
Table 5 shows the different used influent concentrations, compared with the synthetic influent, used in former experiments and the wastewater characteristics, used for the design of the large-scale reactor. The influent concentrations of the pre-treated wastewater, differ a lot from the synthetic influent. Especially the biodegradable COD concentrations are much lower (Table 5), resulting in a slow increase in biomass concentration. Potentially, always some intrinsic biomass washout occurs. When the COD load is low, a relatively high washout occurs in terms of washed-out biomass COD per influent COD. Therefore a longer time is needed before a clear increase in biomass will be observed, which results in a longer period needed for starting up the system.

**TABLE 5 INFLUENT CHARACTERISTICS**

| | Synthetic influent, first lab tests | Standard raw influent (STOWA) | Average presettled influent concentrations of | |
|---|---|---|---|---|
| | | | treatment plant Gouda [2)] | Treatment plant Berkel[3)] |
| Suspended solids (mg/l) | 0 | 250 | 84 | 75 |
| BOD ∞ (mg/l) | - | 220 | - | 70 |
| soluble COD (mg/l) | | 216 | 186 | 253 |
| COD total (mg/l)[1)] | 381 | 600 | 223 | 308 |
| VFA's (mg COD/l) | | - | 44 | 85 |
| $PO_4^-$-P (mg/l) | 19.6 | 9 | 5.35 | 8.5 |
| $NH_4^+$-N (mg/l) | | - | 31.6 | 57.1 |
| Total N, filtered (mg/l) | 50 | 45 | 42.2 | 67.2 |
| Total N, unfiltered (mg/l) | | 55 | - | 71.1 |
| Total COD/Total N | 7.6 | 10.9 | 5.3 | 4.3 |
| Total COD/Total P | 19.4 | 66.6 | 41.7 | 36.2 |

[1)] In case of the lab experiments, this is the value after settling in the 4°C room.

[2)] The average values of Gouda (Stolwijk pipeline) are based on 2 batches.

[3)] The average values of Berkel are based on 8 batches.

## 6.3 RESULTS

### 6.3.1 GRANULATION

The reactor was started up with wastewater from the treatment plant Gouda (table 5, table 6). The applied settling time of 5 minutes was too short, since the flocs (with granules) coagulated when aeration was stopped, and the settling velocity of the biomass was therefore too low to maintain the biomass in the system. Therefore, the reactor was started up again after 13 days of operation (=day 0) with new activated sludge. The reactor was also frequently inoculated with the dispersed cells and the little flocs from the effluent from the stable granular SBAR 2, operated with synthetic wastewater in order to enrich the biomass with PAO and autotrophic organisms.

The settling time was varied between 15 and 6 minutes, in order to keep granules with settling rate between 2 m/h and 5 m/h in the reactor. This was done in order to prevent too much biomass loss by wash out with the effluent. After 10 days, a riser was added to the SBBC, in order to speed up the granulation process (chapter 5).

**TABLE 6 OPERATION OF REACTOR 3, CHANGES IN SETTLING TIME, INFLUENT AND OTHER CHANGES IN OPERATION**

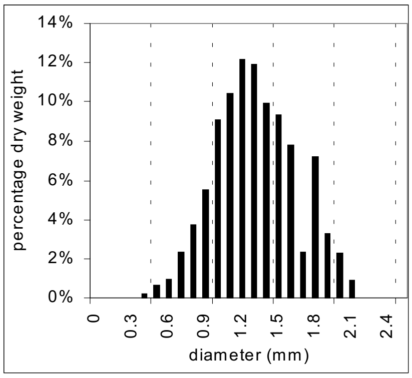
| Day | Settling time | Wastewater | Other Changes |
|---|---|---|---|
| -13 | 5 minutes | Gouda | start up first time |
| 0 | 15 minutes | Gouda | start up second time |
| 10 | 7 minutes | Gouda | from bubble column to airlift |
| 6-30 | varied between 6 and 7 minutes | Gouda | |
| 30 | 15 minutes | Berkel | |
| 50 | 10 minutes | Berkel | Total cycle time shortened to two hours to increase COD load |
| 69 | 6 minutes | Berkel | |

The increase in biomass content in the reactor with a low COD load is very difficult, since an intrinsic biomass wash-out will always occur (chapter 6.2.2). In practice at a wastewater treatment plant, the total cycle time would be shortened in order to increase the COD-load on the system. This method was not feasible at lab-scale, since the amount of available wastewater was limited. Therefore, the influent was changed to a more concentrated variant (Berkel) after 1 month of operation. Short after the change to the more concentrated wastewater, the granules did not flocculate anymore during settling and the granules started to increase in diameter to 0.5 to more than 1 mm after 40 days (figure 12). This is the first proof that granulation can occur with pre-treated municipal wastewater.

The development of the granules was slower than in the reactors operated with synthetic wastewater, mainly due to the lower availability of (easily biodegradable) COD. This improved when the cycle-time was decreased to 2 hours with the same amount of influent per cycle as in the 3 hour cycle (1600 ml/cycle), which caused an increased COD load from 1.16 to 1.74 kg/m$^3$/day (after storage the COD concentration in the wastewater decreases, leading to a COD load of 1.28 kg/m$^3$/day after 5 days of storage). The size of the granules reached an average equivalent diameter of 1.1 mm after 70 days, the aspect ratio 0.69 was and the shape factor 0.44, which is lower than on synthetic influent (respectively 0.70 and 0.73). The average SVI was 38 ml/g. The experiment was stopped while the biomass concentration in the reactor was increasing rapidly, but the last measured biomass concentration was 2.2 g TSS/l at day 100. Figure 13 shows a particle size distribution curve (day 63). Contrary to the smooth particles grown on synthetic wastewater, the surface of many granules grown on pre-treated wastewater were covered with small filamentous structures (figure 12).

FIGURE 12 SETTLING GRANULES IN THE REACTOR AT DAY 42 (A). GRANULES AFTER 63 DAYS (B) OF OPERATION


FIGURE 13 PARTICLE SIZE DISTRIBUTION OF THE GRANULES AT DAY 63




An explanation for the growth of filaments on the surface of the granules can be the presence of COD during the aerobic period. The fast degradable substrates (as acetate) are taken up during the anaerobic period, but much of the complex substrate and the low amounts of particulate COD are still available during the aerobic period. This favours fast growing heterotrophic organisms and among these filamentous organisms, that can use this substrate to grow during the aerobic period.

Shear can play an important role in the appearance of an irregular surface as well. Since only very small amounts of biomass are present in the reactor, there are very few particle/particle collisions. These particle/particle collisions play an important role in the amount of shear that the granules experience (Gjaltema et al., 1997). In previous research it was observed that filamentous growth on the surface of the granules increase with decreasing shear stress (Beun et al., 1999). In that case, the problem of filaments at the surface of the granules will be minimised when more granules are present in the system and thus more particle/particle collisions occur.

The amount of wall-growth was a problem during the experiment. Since an airlift reactor was used, a surface area of 0.5 m$^2$ was available for wall-growth (167 m$^2$/m$^3$). The biofilm quantity grown on the surface area of the reactor (measured once) plus the biomass in the effluent and the increasing dry weight concentration in the reactor, approximates the theoretical growth yield of 0.3 g SS/g COD for PAO (Kuba et al., 1993). Because of the extensive wall growth and the wash-out of organisms, the settled bed only increases with 0.35 cm per day, which is lower than as observed in the experiments with a synthetic wastewater. This extensive wall-growth is a typical small-scale problem. Such slow increase in biomass concentration in the reactor is not to be expected in a large- or pilot-scale system.

### 6.3.2 CONVERSION PROCESSES

The operation time of this experiment was too short to get stable conversion processes. The conversion processes derived from the reactor during operation at a cycle time of 2 hours and the wastewater from RWZI Berkel, will be given in this paragraph.

The VFA are totally removed during the anaerobic feeding period. The COD concentrations in the effluent were on average 0.17 g/l. This means that 60% of the biodegradable COD was converted. This biodegraded COD consists partly of particulate and colloidal material. At pilot-scale, this has to be improved and further investigated.

The P uptake per gram of biomass fluctuated largely, but during the last days of operation an average value 0.8 mg P/g d.w. was measured. In phosphate removing Bio-P systems values in the range of the 5-7 mg P/g d.w. reported by STOWA (STOWA, 2001). The sludge wash-out and wall-growth can have negative effects on the development of the slow growing PAO, as well as the growth of heterotrophs during the aeration phase (chapter 6.3.1).

The total N removal fluctuated between 5% and 24%. Balances, made from the measurements, showed that nitrification and denitrification occur simultaneously. Also the average N-conversion per gram dry weight was 11 mg N/g TSS and even 26 mg N/g TSS at day 54. Compared to the airlift reactor on synthetic wastewater this is very high, since this reactor removes up to 3.8 mg N/ g TSS. It has to be mentioned that this ratio cannot be higher in this last reactor, because it is limited by the ammonia loading rate.

At the end of the experiment, the reactor was fed during one cycle with the synthetic wastewater as is used in the other experiments. This leaded to a full COD (acetate) removal during the anaerobic period, 51% N-removal (as ammonia in the effluent) and 40% P-removal.
This result shows that part of the organisms are PAO and that the nitrate that is formed is immediately denitrified, since the amount of ammonia removed is too much to be only used for growth.

## 6.4 CONCLUSIONS

It can be said that the granulation process does occur with this complex influent. The granule formation was slower than on synthetic wastewater. This was due to the lower COD load of easily degradable substrate. Also, the granules were less regular and smooth. The formed granules are relatively small (1.1 mm) and have good settling characteristics (settling velocity > 5.5 m/h, SVI < 40 ml/g).

Biomass wash-out and wall-growth had a negative effect on the selection of the slow growing organisms and thus the processes of N-removal and P-removal. In a larger scale reactor, these effects will be less, since the ratio wall/volume ratio is much lower.
The COD, N and P loads have to be increased during the start-up, because this most certainly will have a positive effect on the growth of biomass.
From the results in the laboratory, we are not able to conclude that the conversion processes are sufficient for the treatment of the wastewater in practise. Results are promising though and the formation of granular aerobic sludge in this system is the main first step.

# 7 DESIGN AND ECONOMICS OF A FULL-SCALE GSBR TREATMENT PLANT

## 7.1 INTRODUCTION

Parallel to the research work a feasibility study was carried out for the full scale application of a GSBR treating municipal wastewater. For this purpose, two alternatives of the aerobic granular sludge technology are compared to three variants of conventional STPs based on activated sludge technology. Furthermore, costs determining process parameters are identified.

## 7.2 GENERAL BASIC CONDITIONS

### 7.2.1 CAPACITY AND WASTEWATER COMPOSITION

For the capacity a range for the pollution load is taken between 60,000 and 190,000 population equivalents (136 g TOD). For the wastewater the standard composition defined in STOWA research report 98-29 is assumed.
The ratio of dissolved and suspended particles is based on the results of the STOWA research "Physical/chemical pre-treatment of wastewater" in which the colloidal parts (with a diameter between 0.1 and 5 µm) are added up to the suspended particles. A summary of the wastewater composition and flow rates is given in table 7.

**TABLE 7 WASTEWATER COMPOSITION AND FLOW**

| Composition | dissolved (mg/l) | suspended (mg/l) | Total (mg/l) |
|---|---|---|---|
| COD | 216 | 384 | 600 |
| BOD | 106 | 114 | 220 |
| Suspended solids | n/a | 250 | 250 |
| TKN | 45.7 | 9.3 | 55 |
| $P_{total}$ | | | 9 |
| **Flow rate** | **Unit** | **Value** | |
| Average flow | l/(pe.d) | 160 | |
| RWF | l/(pe.h) | 34 | |
| RWF/DWF ratio | - | 2.5 | |

In table 8 the removal efficiencies of primary sedimentation with and without chemical dosing are summarised.

**TABLE 8** **REMOVAL EFFICIENCIES OF PRIMARY SEDIMENTATION WITH AND WITHOUT CHEMICAL DOSING**

| Composition | Primary sedimentation | Primary sedimentation with pre-precipitation (Fe and polymer) | Primary sedimentation with polymer dosing |
|---|---|---|---|
| COD | 31% | 53% | 53% |
| Suspended solids | 51% | 80% | 80% |
| TKN*) | 7% | 19% | 19% |
| $P_{total}$ | 7% | 70% | 38% |
| *) excluding the N-load in the rejection water of the sludge treatment | | | |

### 7.2.2 EFFLUENT REQUIREMENTS

The effluent requirements are based on the Discharge Decree for municipal wastewater in The Netherlands (see table 9). The requirements with regard to nitrogen and phosphate are determinative to a large extent for the technological design of STPs.

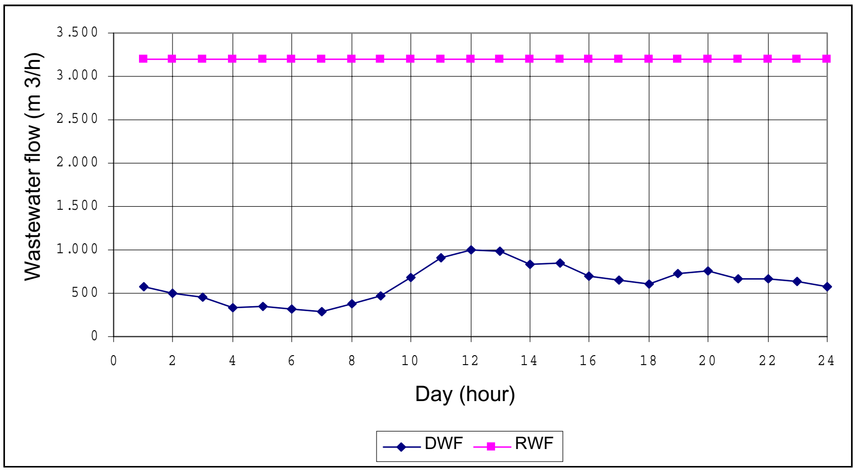
**TABLE 9** **EFFLUENT REQUIREMENTS**

| Parameter | Value (mg/l) | Type of requirement |
|---|---|---|
| COD | 125 | daily average |
| BOD | 20 | daily average |
| Suspended solids | 30 | daily average |
| $N_{total}$ | 10 | yearly average |
| $P_{total}$ | 1 | moving average of ten successive samples |

### 7.2.3 FLOW PATTERN

Figure 14 shows the daily flow pattern of a specific Dutch STP with a capacity of 400,000 p.e. at dry weather conditions as well as at rain weather conditions. As appears from Figure 14 the maximum flow is based on a day with a 24-hour continuous maximum rain weather flow (RWF).

**FIGURE 14** **FLOW PATTERN**



### 7.2.4 COST ESTIMATES

Table 10 summarises the starting points for the calculation of the capital costs. Capital costs are calculated based on annuities.

**TABLE 10 STARTING POINTS CALCULATION CAPITAL COSTS**

| Cost factor | Unit | Value |
|---|---|---|
| Depreciation period civil parts | year | 30 |
| Depreciation period mechanical parts | year | 15 |
| Depreciation period electro-technical parts | year | 15 |
| Interest rate | % | 6 |

In table 11 the starting points for the calculation of the operational costs are given.

**TABLE 11 STARTING POINTS CALCULATION OPERATIONAL COSTS**

| Cost factor | Unit | Value |
|---|---|---|
| Electricity | €/kWh | 0.054 |
| Sludge disposal | €/tonDrySolids | 320 |
| Iron chloride (41%) | €/ton | 115 |
| Poly electrolyte (liquid, 50%) | €/kg active | 3.6 |
| Maintenance civil parts | % of investments | 0.5 |
| Maintenance mechanical parts | % of investments | 2 |
| Maintenance electro technical parts | % of investments | 2 |

### 7.2.5 FOOTPRINT AND LAND PRICE

The footprint is calculated by the sum of the net surfaces of all process units and buildings, multiplied by a factor 1.3.
The land price is assumed to be 22.7 €/$m^2$. For the calculation of the annual costs associated with the purchase of land, only the interest costs are taken into account. Depreciation costs are not calculated, assuming that land keeps at least its original purchase value.

## 7.3 REFERENCE SYSTEMS

### 7.3.1 INTRODUCTION

The following three alternatives of the conventional activated sludge system (AS) are taken as a reference:

- a conventional AS-system with primary sedimentation, biological phosphate removal, additional simultaneous phosphate removal and sludge digestion;
- a conventional AS-system with primary sedimentation, pre-precipitation and sludge digestion;
- a conventional AS-sludge system with biological phosphate removal, additional simultaneous phosphate removal and simultaneous aerobic sludge stabilisation.

Global process flow diagrams of the aforementioned reference alternatives are presented in Appendix 1 (figure 1 to 3).

### 7.3.2 DESIGN ASSUMPTIONS

The following design assumptions are made:

- design temperature : 10 °C;
- maximum hydraulic load primary sedimentation : 4 m/h
- Me/P ratio based on the influent P-load (mol/mol):
- AS with primary sedimentation, Bio-P and sludge digestion : 0.2
- AS with primary sedimentation, pre-precipitation and sludge digestion : 1.5
- AS with Bio-P : 0.2
- poly electrolyte dosage with pre-precipitation : 0.8 $gPE_{active}/m^3$;
- pre-denitrification volume : 20%
- Sludge Volume Index (SVI) : 150 ml/g
- hydraulic retention time sludge digestion : 20 d
- average degradation MLVSS : 47 %
- poly electrolyte dosage sludge dewatering : 12 $gPE_{active}/kgDS$
- Dry Solids content dewatered sludge : 25 %
- N-total removal calculations according to Kayser (1983) and STOWA (1997);
- sludge production calculation according to Chudoba (1985);
- aeration by means of surface aerators with an efficiency of 2.0 $kgO_2/kWh$ in clean water;
- calculation of the oxygen demand according to Beuthe (1970);
- STORA-directives for the design of sedimentation tanks;
- gravitational thickening of primary sludge;
- mechanical thickening of surplus sludge;
- dewatering by centrifuges.

In table 12 the design sludge load of the aeration tanks are given for the reference alternatives. These design loads are based on the biological part of the sludge.

**TABLE 12 DESIGN SLUDGE LOADS OF AERATION TANKS IN REFERENCE ALTERNATIVES**

| Alternative | COD (kgCOD/(kgDS.d)) | Nkj (kgNkj/(kgDS.d)) |
|---|---|---|
| AS with primary sedimentation, Bio-P and sludge digestion | 0.14 | 0.018 |
| AS with primary sedimentation, pre-precipitation and sludge digestion | 0.11 | 0.019 |
| AS with Bio-P | 0.16 | 0.014 |

### 7.3.3 DIMENSIONS AND COSTS

In appendices 3 to 6 the dimensions and cost estimates of the three reference alternatives for a capacity of 120,000 p.e are summarised. Operational costs are based on main cost factors such as sludge disposal costs, power use aerators, chemical use and maintenance costs. Power use of the aerators for the reference alternatives with sludge digestion is corrected for power production from biogas.

Figures 15 and 16 show the investment costs and the total specific annual costs (capital costs and operational costs). As expected, the investment costs of the reference alternative “AS with primary sedimentation, Bio-P and sludge digestion” are the highest (see figure 15). The reference alternative “AS with Bio-P” results in the lowest investment costs.

Based on the total specific annual costs the reference alternative “AS with primary sedimentation, pre-precipitation and sludge digestion” is the most costly alternative throughout the range (see figure 16). In comparison with the alternative “AS with primary sedimentation, Bio-P and sludge digestion” the operational costs are substantially higher. This is caused by a higher sludge production and a higher chemical use required for phosphate removal. Compared to the reference alternative “AS with Bio-P” the higher capital costs and higher chemical use cause the cost difference.

The reference alternatives “AS with primary sedimentation, Bio-P and sludge digestion” and “AS with Bio-P” are economically the most attractive and comparable. The reference alternative "AS with Bio-P" is the most economic alternative in the range between 60,000 and 150,000 p.e. In this range the low capital costs are decisive rather than the relatively high sludge production and high electricity use. The reference alternative "AS with primary sedimentation, Bio-P and sludge digestion" is the preferred alternative from 150,000 p.e. on. As from this capacity the lower sludge production and lower electricity consumption weigh heavier in the total costs.

FIGURE 15 INVESTMENTS REFERENCE ALTERNATIVES

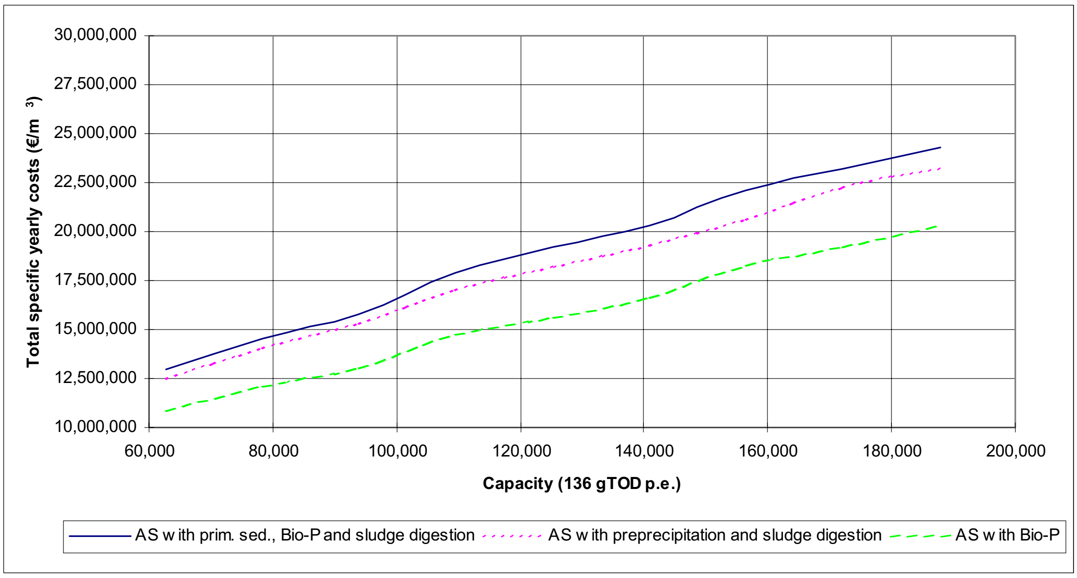
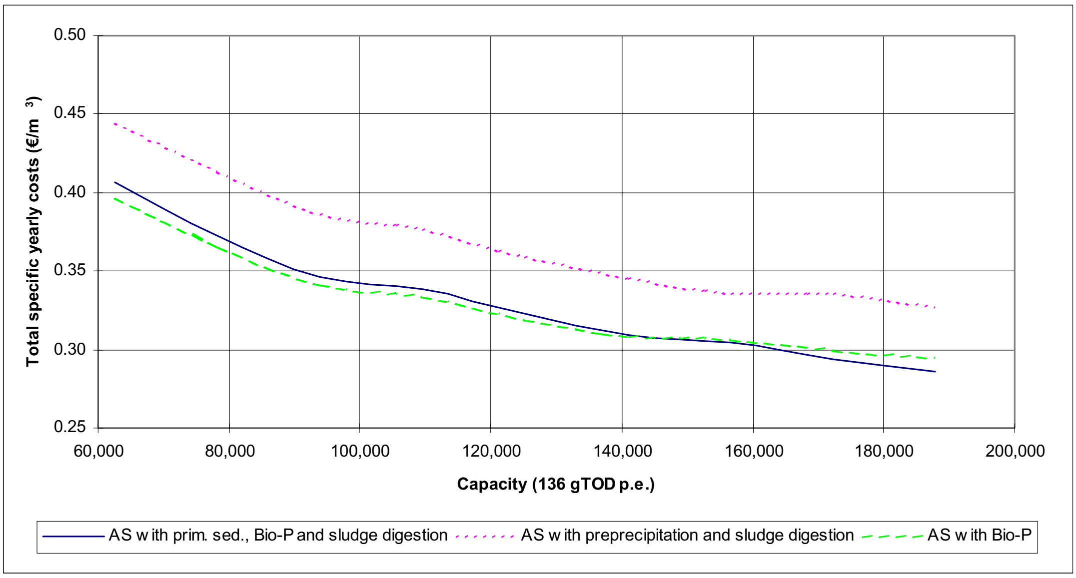



FIGURE 16 TOTAL SPECIFIC ANNUAL COSTS REFERENCE



## 7.4 INTERACTION BETWEEN FEASIBILITY STUDY AND RESEARCH

Because of the results from the feasibility study the research program was partly adjusted. Changes in reactor operation and their influences on the granulation are extensively discussed in the previous chapters. In this paragraph, the interactions between the laboratory research and the feasibility study are summarised.

### CYCLE CONFIGURATION:

- *Filling time*: A pulse feed in three minutes as performed during the first experiments is not feasible for a full-scale application (the flow during filling of the GSBR would be a factor twelve higher compared to the RWF). Therefore, experiments with a longer feeding period were performed (chapter 3);
- *Sedimentation:* For a full scale GSBR with a construction height of 5-6 metres and a particle sedimentation velocity of 15 m/h, a sedimentation time of 3 minutes is too short and should be extended to 15-30 minutes. Short sedimentation times result in low and undesirable Height/Diameter-ratios of the reactor. No further investigations on this topic were performed;
- *Effluent discharge:* Short decantation times lead to high discharge flows and expensive discharge constructions, despite the fact that effluent is discharged by gravity. The dimensions of a post treatment step depend also strongly on the maximum discharge flow, although this can be compensated with buffer tanks. Research was carried out to test the feasibility of simultaneous feeding and effluent withdrawal: incoming feed would push out the effluent in the reactor, assuming a plug flow character through the settled bed of granules. The main conclusion of these tests is that further research has to be carried out (e.g. on pilot scale) before conclusions can be drawn.

### PHOSPHATE REMOVAL

Initially chemical phosphate removal was considered. This seemed logical in combination with extensive primary treatment with pre-precipitation. Design calculations showed higher costs for chemical phosphate removal in comparison with biological phosphate removal. This was due to higher chemical use and higher sludge production. Based on these calculations biological phosphate removal is preferred to chemical removal. It was demonstrated that a longer filling time leads to a microbiological population shift towards PAOs (see chapter 3) enabling biological phosphate removal.

### NITROGEN REMOVAL AND OXYGEN SATURATION

Experiments proved that DO concentrations of 20% of the saturation value result in high Ntotal conversions. In the presence of PAOs stable granule formation is possible under these conditions.

### SLUDGE CHARACTERISTICS

Besides the granule characteristics (sedimentation velocities, size, density, etc.) the characteristics of the biomass which is washed out can also be important in case these 'fines' determine a large part of the sludge production. Important aspects in terms of sludge treatment are the thickening, digestion and dewatering characteristics of these 'fines'.
To evaluate the potential of biological phosphate removal a potential difference between the phosphate content of granules and fines is important. Research on this matter executed by TUD showed that this was not the case.

## 7.5 DESIGN GSBR

### 7.5.1 ASSUMPTIONS

**PRIMARY AND POST TREATMENT**

Considering the primary treatment two extreme alternatives are taken into account:

- GSBR with primary treatment including chemical dosing with extensive removal of suspended solids. Post treatment is assumed not to be required;
- GSBR with only post treatment. The post treatment consists out of removal of suspended solids from the effluent from the GSBR. For suspended solids removal the following conventional techniques are considered:
- conventional sedimentation;
- lamella settlers;
- rapid sand filtration.

**CONFIGURATION GSBR**

From the research appeared that granules can be grown in both an airlift reactor as in a bubble column. Because of the lower building costs bubble columns are taken as a basis for the calculations.

**SLUDGE CONCENTRATION**

The sludge concentration in the GSBR after effluent discharge is assumed to be 10 kg/m$^3$.
A free space of 1 m between sludge level and discharge location is assumed.

**SLUDGE PRODUCTION AND CHARACTERISTICS**

The design sludge load and sludge production in a GSBR are based on the results of the STOWA research with airlift reactors (97-25). In this research the net sludge production is zero, meaning equal sludge content in influent and effluent. A sludge production of 0.08 kgDS/kgCOD resulting from dissolved COD degradation is assumed. Total sludge production excluding chemical sludge before digestion amounts to 0.4 kgDS/kgCOD.
With regard to the sludge characteristics fines washed out from the GSBR are assumed to be removed by sedimentation. This means that biological processes in the GSBR convert colloidal particles in the influent into settlable solids. This is based on the results of the aforementioned STOWA research with airlift reactors (97-25). In this research SVI-values of fines washed out of the airlift reactors between 60 and 200 ml/g were reported. The removal and conversion of colloidal particles from the wastewater is an important point of interest in follow-up research.

**SLUDGE TREATMENT**

The sludge treatment for the GSBR-alternatives consists of thickening, digestion and dewatering, which corresponds to the sludge treatment process of the reference alternatives.

**COD- AND NITROGEN REMOVAL**

Extensive COD- and nitrogen is assumed to be possible. The design sludge load is based on the STOWA research 97-25. High removal rates for COD and N-total were reported at sludge loads of 0.3 kgCOD/(kgDS.d) respectively 0.05 kgNkj/(kgDS.d).
The assumed removal efficiencies for COD and suspended solids are given in table 13.

TABLE 13 REMOVAL EFFICIENCIES GSBR

| Parameter | Suspended | Dissolved |
|---|---|---|
| **GSBR with primary sedimentation and chemical dosing** | | |
| COD | 15% | 85% |
| Suspended solids | 15% | n.a. |
| **GSBR with post treatment** | | |
| COD | 20% | 85% |
| Suspended solids | 25% | n.a. |

**PHOSPHATE REMOVAL**
Only biological phosphate removal is considered.

**CONSTRUCTION HEIGHT**
One of the GSBR-alternatives is based on biological phosphate removal which is assumed to be possible in a full-scale GSBR applying alternating anaerobic feeding periods and aerobic reaction periods. In order to be assured of plug flow conditions while feeding from the bottom of the GSBR through the settled granules bed, the maximum hydraulic surface load during the feeding period is chosen at 7.5 m/h. This leads to a construction height of the GSBRs of 5-6 m.

**NUMBER OF TREATMENT LINES**
The number of parallel treatment lines is determined by the length of filling time compared to total cycle time. Assumed is that at the most one reactor can be fed with wastewater. For instance, if the filling time is a third part of the total cycle time, this would lead to three parallel treatment lines.

**EFFLUENT DISCHARGE**
The effluent is assumed to be discharged by gravity from a fixed height. Furthermore a maximum flow velocity in the effluent pipe of 2.8 m/s is assumed. This is a compromise between maximum discharge flow at one hand and hydraulic losses in the effluent pipes at the other hand. A low maximum flow rate leads to large diameters, a low hydraulic decline and high discharge flows (short decantation times). Disadvantages of high discharge flows are expensive discharge constructions as well as high costs for post treatment (if required). High design discharge flow velocities result in high hydraulic losses and therefore in long decantation times.

**CYCLE TIMES**
For the filling time and total cycle time respectively 20 and 60 minutes are assumed. The sedimentation time is calculated based on the sedimentation velocity and the reactor height. The decantation time is calculated based on the maximum flow velocity in the effluent pipe and the reactor height.

**POST TREATMENT**

| | | | |
|---|---|---|---|
| Surface load conventional sedimentation | : | 1 | m/h |
| Surface load lamella settler (based on footprint) | : | 8 | m/h |
| Surface load sand filters | : | 20 | m/h |

**ALTERNATIVES**
In appendix 2, figures 1 and 2, the global process flow diagrams for the GSBR alternatives are given. The alternative with post treatment is depicted with lamella settlers.

### 7.5.2 BUFFER CAPACITY

**RAW WASTEWATER**

The total wastewater flow including rainwater is treated in the GSBR. Considering the flat RWF flow pattern (see figure 14) buffering of raw wastewater is of no use. Because buffering of wastewater is pointless at RWF-conditions, buffering raw wastewater at DWF-conditions is not considered. This means that most of the time the volume of batches to be treated will vary over the day. In principle this is not different from conventional continuous systems, especially not when the number of parallel treatment lines of the GSBR increases.

**EFFLUENT GSBR**

In case of post treatment, the overflow from the GSBR is buffered. Because effluent from the GSBR is discharged by gravity, the initial flow is high. In order to reduce the dimensions of the post treatment step, the overflow of the GSBR has to be buffered.

### 7.5.3 DESIGN OUTLINE

In this paragraph the design of the GSBR with post treatment (lamella settler) and biological phosphate removal is specified for a capacity of 120,000 p.e. The cycle times as well as global dimensions are given in table 14.

**TABLE 14** CYCLE TIMES

| Fase | max. DWF | RWF |
|---|---|---|
| Filling | 20 | 20 |
| Aeration | 28.5 | 20 |
| Sedimentation | 4 | 10 |
| Decantation | 7.5 | 10 |
| Total | 60 | 60 |

| | | | |
|---|---|---|---|
| Number of treatment lines | : | 3 | |
| Batch size RWF | : | 1,335 | $m^3$ |
| Batch size max. DWF | : | 528 | $m^3$ |
| Design capacity per GSBR | : | 3,800 | $m^3$ |
| Volume per GSBR at RWF | : | 3,435 | $m^3$ |
| Sludge content GSBR at RWF | : | 4.6 | $kg/m^3$ |
| Volume GSBR at max. DWF | : | 2,630 | $m^3$ |
| Sludge content GSBR with max. DWF | : | 6.0 | $kg/m^3$ |
| Volume GSBR after decantation (incl. free space) | : | 1,570 | $m^3$ |
| Surface lamella settler | : | 500 | $m^2$ |

From the preceding it appears that for an important part the capacity of the GSBR at RWF is determined by the size of the batches.

### 7.5.4 EFFLUENT QUALITY

In table 15 effluent qualities for both alternatives are given. The alternative with primary treatment does not meet the required effluent quality with respect to suspended solids and P-total (compare with table 9) which is caused by a too high suspended solids concentration in the pre-treated wastewater (50 mg/l). Given the starting point that suspended solids in a GSBR are not removed, the suspended solids concentration in the effluent is too high. This also leads to increased levels of COD, N-kj and P-total. The GSBR with post treatment meets the effluent requirements.

The effluent requirements for the alternative with pre-treatment can only be met if the suspended solids concentration in the wastewater fed to the GSBR is less than 10-30 mg/l.

**TABLE 15** **EFFLUENT QUALITY**

| Parameter | GSBR with primary sedimentation and chemical dosing | GSBR with post treatment |
|---|---|---|
| COD (mg/l) | 80 | 40 |
| Suspended solids (mg/l) | 50 | < 10 |
| Nkj (mgN/l) | 4.1 | 2.0 |
| $NO_3^-$ (mgN/l) | 5.9 | 8.0 |
| Ptotal (mg/l) | 1.5 | 1.0 |

### 7.5.5 COSTS

In the appendices 6 and 7 the dimensions and cost estimates for the two GSBR alternatives with a capacity of 120,000 p.e. are given.

Figure 17 and 18 show the investment costs and the total specific annual costs (sum of capital costs and operational costs) in relation to the capacity. For the GSBR with post treatment lamella settlers are chosen. The costs for reference alternative 1 (AS with primary sediment ation, Bio-P and sludge digestion) are given. Also the costs of the GSBR alternative with primary treatment are depicted.

As can be seen both the investment costs and the total specific annual costs of the reference alternative are the highest for the complete range. Furthermore the costs of the GSBR with post treatment are, as expected, higher than the costs of the GSBR with primary treatment.

The investments costs of the GSBRs with primary treatment respectively post treatment are on the average 29% and 12% lower than the reference alternative. The total annual costs are 16% and 6% lower.

**FIGURE 17** **INVESTMENT COSTS**

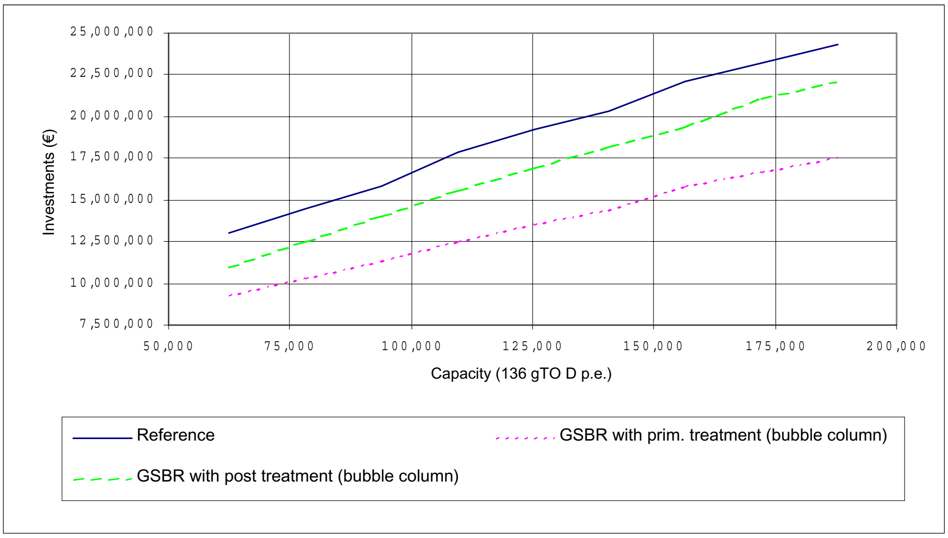
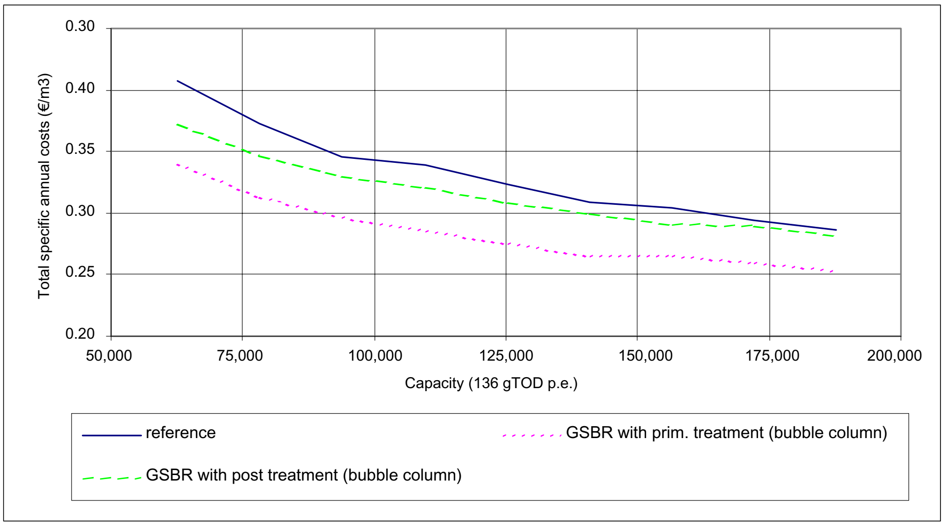


FIGURE 18 TOTAL SPECIFIC ANNUAL COSTS



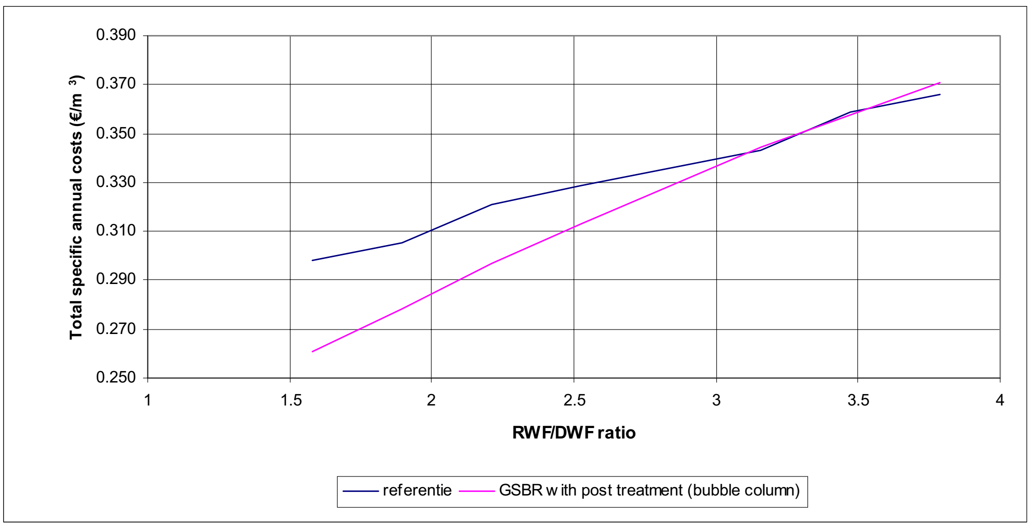

## 7.6 SENSITIVITY ANALYSIS

In this paragraph a sensitivity analysis for a number of important parameters is given. For this purpose a GSBR with post treatment, biological phosphate removal and a capacity of 120,000 p.e. is used. If not mentioned otherwise, a RWF/DWF ratio of 2.5 is applied (see also paragraph 7.2.1).

### 7.6.1 RWF/DWF RATIO

The RWF has a large impact on the design of the STP, especially for a STP based on a batch wise technology. At lower RWFs a GSBR system becomes relatively more attractive than a conventional activated sludge system. In the Netherlands a lower RWF is not imaginary given the increasing amount of improved separated sewage systems, the increasing level of disconnection of clean water and solid surface, optimisations of the wastewater chain and active control of pumping-stations (RTC). In figure 19 the influence of the RWF on the costs for a GSBR and reference alternative 1 are given. Below a RWF/DWF ratio of 3.5 a GSBR becomes more attractive.

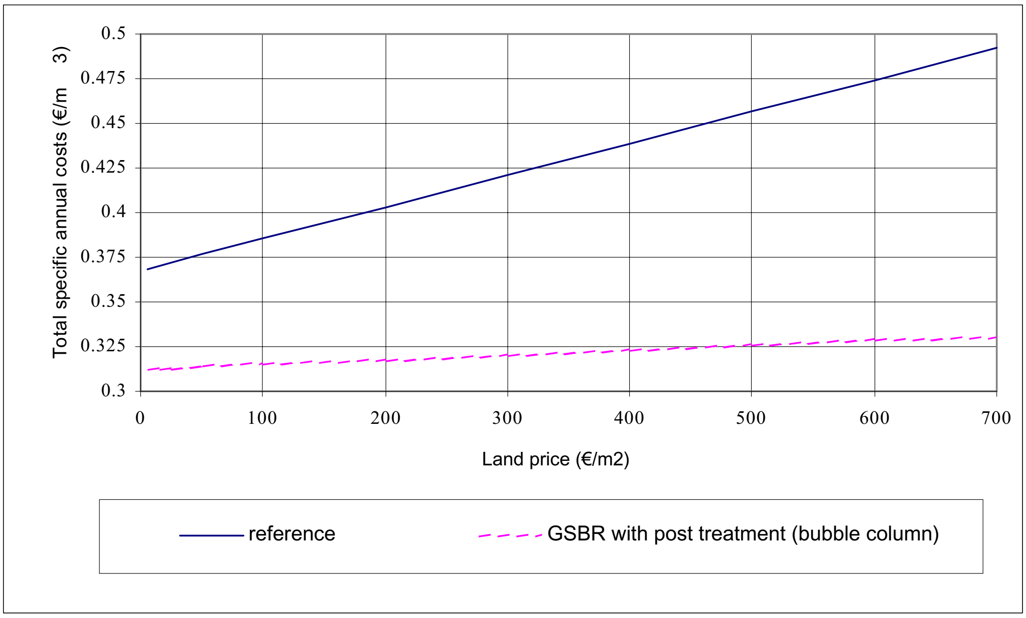
FIGURE 19 INFLUENCE RWF/DWF RATIO



### 7.6.2 LAND PRICE

The sensitivity of the land price on reference alternative 1 and the GSBR alternative with post treatment is shown in figure 20. As can be seen the GSBR is more attractive on the complete range.

**FIGURE 20** **INFLUENCE LAND PRICE**




### 7.6.3 POST TREATMENT ALTERNATIVES

In table 16 construction costs as well as total specific annual costs of the alternatives for post treatment are given (capacity 120,000 p.e.). Based on construction costs lamella settlers are the most attractive. The total costs of a GSBR alternative with post treatment based on conventional sedimentation are the least expensive. This is because the contribution of the mechanical parts of a lamella settler to the construction costs is substantial higher compared to conventional sedimentation. Sand filtration is more expensive compared to the other two alternatives.

**TABLE 16** **COSTS ALTERNATIVES POST TREATMENT**

| Building costs | Construction costs (€) | Total costs GSBR with post treatment and Bio-P (€/m$^3$) |
|---|---|---|
| Lamella settler | 1,810,000 | 0.313 |
| Conventional sedimentation | 1,960,000 | 0.310 |
| Sand filtration | 2,340,000 | 0.319 |

### 7.6.4 ROUND OR RECTANGULAR?

In table 18 the basic construction costs (no constructions for the incoming wastewater and effluent disposal, no division walls, etc.) are given for round and rectangular concrete tanks with a height of 10 m. It can be concluded that round tanks are substantially cheaper in comparison with rectangular tanks.

TABLE 18 BASIC CONSTRUCTION COSTS

| Capacity ($m^3$) | Construction costs (€/$m^3$) | |
|---|---|---|
| | Round tanks | Rectangular tanks |
| 3.000 | 107 | 151 |
| 6.000 | 78 | 128 |
| 12.000 | 57 | 104 |

7.6.5 CYCLE TIME

The influence of the cycle time is given in table 19. The cycle time appears to have a substantial influence on the batch size at RWF conditions and consequently on the volume of the GSBR. The difference in annual costs between a cycle time of 60 and 240 minutes is € 210,000,-.

TABLE 19 INFLUENCE CYCLE TIME

| Cycle time (minutes) | Batch size RWF ($m^3$) | Volume GSBR ($m^3$) | Volume buffer tank effluent GSBR ($m^3$) | Total specific annual costs (€/$m^3$) |
|---|---|---|---|---|
| 60 | 1,335 | 3,800 | 1,470 | 0,313 |
| 120 | 2,670 | 5,250 | 2,940 | 0,326 |
| 180 | 4,000 | 6,750 | 4,400 | 0,335 |
| 240 | 5,335 | 8,350 | 5,875 | 0,343 |

## 7.7 CONCLUSIONS AND POTENTIAL

7.7.1 REFERENCE ALTERNATIVES

The reference alternatives with biological phosphate removal are financially more attractive compared to the reference alternative with chemical phosphate removal. Chemical phosphate removal leads to a higher sludge production and a high chemical use. The differences between the reference alternatives 1 and 3 ("AS with primary sedimentation, Bio-P and sludge digestion" and "AS with Bio-P") are small. At lower capacities (< 150,000 p.e.) reference alternative 3 is cheaper, above 150,000 p.e. alternative 1 is more attractive.

7.7.2 GSBR

**TYPE OF PRIMARY TREATMENT**

A GSBR with only primary treatment can only meet the required effluent standards if the removal efficiency for suspended A GSBR with only primary sedimentation and chemical dosing can not meet the required effluent standards. This is caused by an insufficient suspended solids removal efficiency in the primary treatment combined with the assumption that suspended solids are not removed in a GSBR.
solids of the primary treatment is higher than 95%. This high value can be accomplished with new technologies such as membrane filtration.
During the pilot research the suspended solids removal efficiency in a GSBR will be determined. If higher suspended removal efficiencies can be established than assumed in this study, a GSBR with only primary treatment could be possible.

Under the assumptions presented in this report a GSBR with post treatment can meet the required effluent standards.

**POTENTIAL**

The aerobic granular sludge technology seems very promising!
Based on both the investments and the total annual costs a STP based on GSBR in the capacity range between 60,000-190,000 p.e. is economically more attractive compared to the reference alternatives based on activated sludge technology. The total annual costs for a GSBR with primary respectively post treatment are on the average 16% and 6% lower.
The footprint of a STP based on aerobic granular sludge technology amounts to only 25% compared to the references.

# 8 CONCLUSIONS AND RECOMMENDATIONS

The lab-scale experiments led to stable granular sludge and more insight in the important factors for simultaneous conversion of COD, P and N components from the influent. Experiments with different feeding periods, different reactor configurations and different oxygen concentrations with a synthetic influent were carried out as well as an experiment with a pre-treated sewage influent. Also the feasibility of a full-scale wastewater treatment plant with a granular sludge reactor was investigated. From this research the following conclusions can be drawn.

Long feeding times were feasible when an anaerobic feeding period is applied. A longer feeding phase is favourable because of the simplification of the scale-up and batch scheduling. Furthermore, the sequencing anaerobic feeding period and aerobic reaction period causes a domination of slow growing phosphate accumulating organisms. This last phenomenon is the main reason for stability of the granules at low dissolved oxygen concentrations. At a DO of 20% of the saturation concentration, simultaneous COD, phosphate and total nitrogen can be accomplished and effluent standards can be complied with. The use of a bubble column and thus applying lower shear was also achievable because of selection for slow growing PAOs.

Operation of a reactor with pre-treated municipal wastewater proved that granulation also occurs with complex substrate.

The feasibility study showed that the aerobic granular sludge technology seems very promising. Based on total annual costs both GSBR variants prove to be more attractive than the reference alternative (6-16%). From a sensitivity analysis it appears that the GSBR technology is less sensitive to the land price and more sensitive to a higher RWF. This means that the GSBR technology becomes more attractive at lower RWF/DWF ratios and at higher land prices. Because of the high permissible volumetric load the footprint of the GSBR variants is only 25% compared to the reference. However, the GSBR with only primary treatment can not meet the present effluent standards for municipal wastewater in the Netherlands, mainly because of a too high suspended solids concentration in the effluent.
A growing number of sewage treatment plants in the Netherlands is going to be faced with more stringent effluent standards. In general, activated sludge plants will have to be extended with a post treatment step (e.g. sand filtration) or transformed into a Membrane Bio Reactor. In this case a GSBR variant with primary treatment as well as post treatment can be an attractive alternative.

Based on both the good research results and the positive technical and financial evaluation for full scale application for sewage treatment, it is recommended to develop the aerobic sludge technology further by means of a pilot research at a sewage treatment plant.
Important aspects that need to be investigated are:

- the stability of granulation with domestic wastewater as a substrate and the influence of:
    - dynamic processes such as the wastewater flow pattern (at DWF and RWF) and process temperature;
    - the influence of suspended solids in the influent;
    - configuration of the GSBR (airlift, bubble column);
    - simultaneous fill and draw of the GSBR.
- volumetric loading rate;
- the effluent quality with respect to current and future effluent standards in the Netherlands;
- characteristics of granules and suspended solids washed out from the GSBR (settling characteristics, possibilities for digestion, influence of coagulants, granule size distribution)

Important aspects to investigate at laboratory scale are:

- granulation through slowly growing organisms others than PAOs;
- effects of COD/N-ratio;
- further research to the stability and composition of the granules;
- activity measurements, kinetics of conversion processes;
- degradation of slowly biodegradable components.

# 9 SYMBOLS AND ABBREVIATIONS

| | |
|---|---|
| Bio-P | Biological phosphate removal |
| COD | Chemical Oxygen Demand |
| DO | Oxygen concentration expressed as percentage from the maximum soluble oxygen |
| DWF | Dry weather flow |
| EBPR | Enhanced Biological Phosphorus Removal |
| EPS | Extra cellular Polymer Structures |
| GAO | Glycogen accumulating organisms |
| GSBR | Granule Sequencing Batch Reactor – full scale design of the granule system |
| PAO | Phosphate accumulating organisms |
| PHA | Polyhydroxyalkanoaat |
| PHB | Poly-β-Hydroxybutyrate |
| PHV | Polyhydroxyvalaraat |
| RWF | Rain weather flow (including DWA) |
| SBAR | Sequenced Batch Airlift Reactor |
| SBBC | Sequenced Batch Bubble column |
| SBR | Sequenced Batch reactors |
| SRT | Solid Retention Time |
| $SVI_5$ | Sludge Volume Index, measured after 5 minutes of settling |
| TOC | Total Organic Carbon |
| TSS | Total Suspended Solids |
| VSS | Volatile Suspended Solids |

Chapter 7:

| | |
|---|---|
| $a$ | specific surface ($m^2/m^3$) |
| $c^*_1$ | oxygen saturation concentration at half the height of the reactor and adjusted for oxygen depletion (ca. 14 mg/l) |
| $c_l$ | average oxygen concentration in GSBR (2 mg/l) |
| $k_l$ | mass transfer coefficient (m/h) |
| OC | aeration capacity (kg/h) |
| V | capacity GSBR ($m^3$) |
| $v_{gs}$ | superficial gas velocity (m/s) |

# 10
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# APPENDIX 1 PFDS REFERENCE SYSTEMS

**FIGURE 1** **PROCESS FLOW DIAGRAM AS WITH PRIMARY SEDIMENTATION, BIO-P AND SLUDGE DIGESTION**

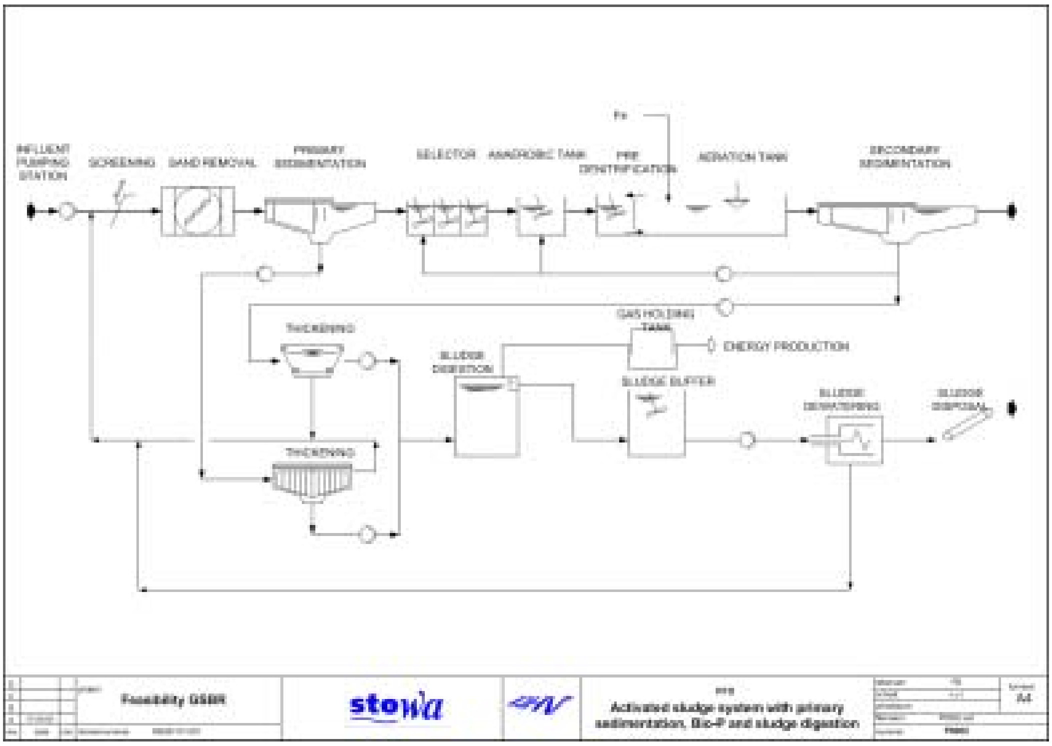



**FIGURE 2** **PROCESS FLOW DIAGRAM AS WITH PRIMARY SEDIMENTATION, PRE-PRECIPITATION AND SLUDGE DIGESTION**

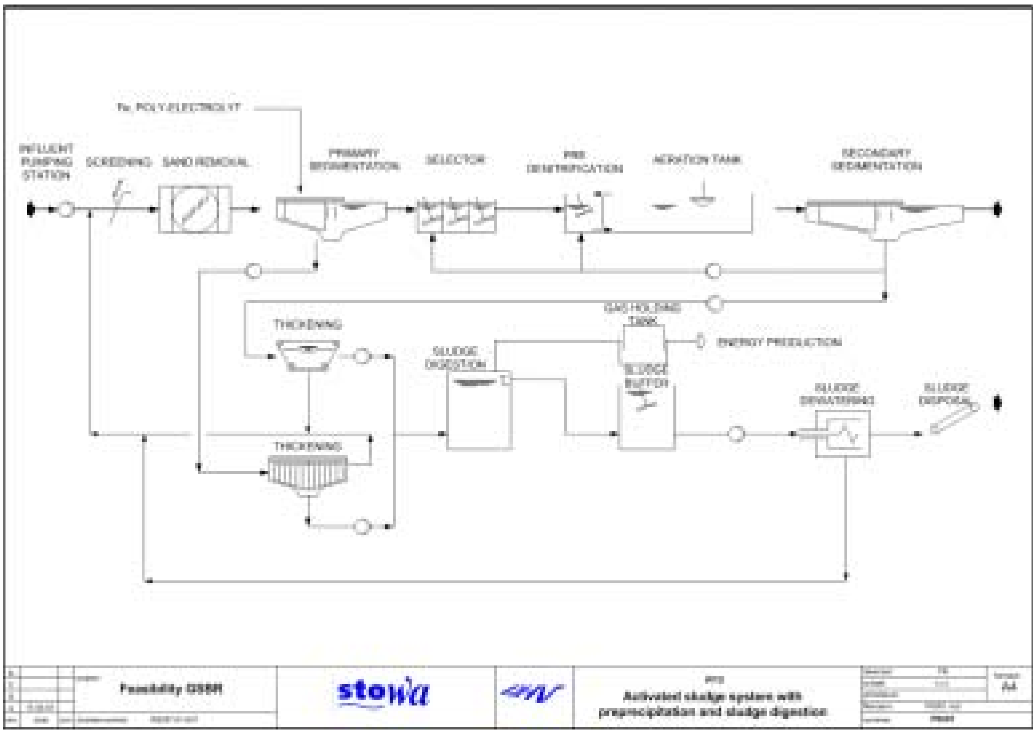
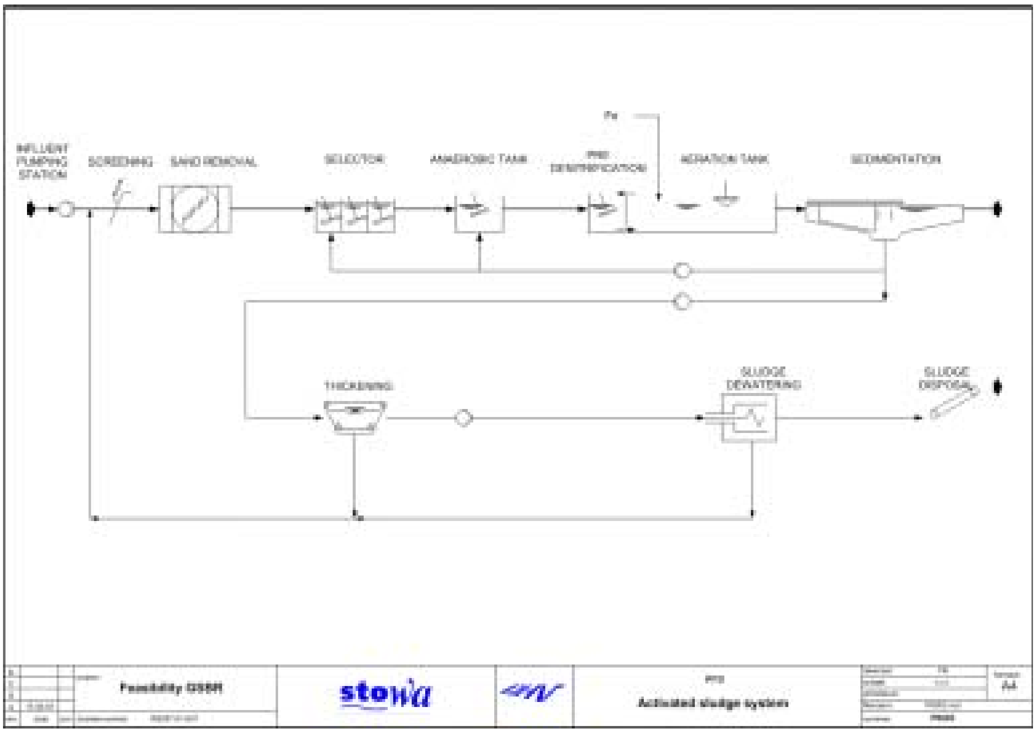


**FIGURE 3** **PROCESS FLOW DIAGRAM AS WITH BIO-P**



# APPENDIX 2 PFDS GSBR ALTERNATIVES

**FIGURE 1** **PROCESS FLOW DIAGRAM GSBR-ALTERNATIVE WITH PRIMARY TREATMENT**

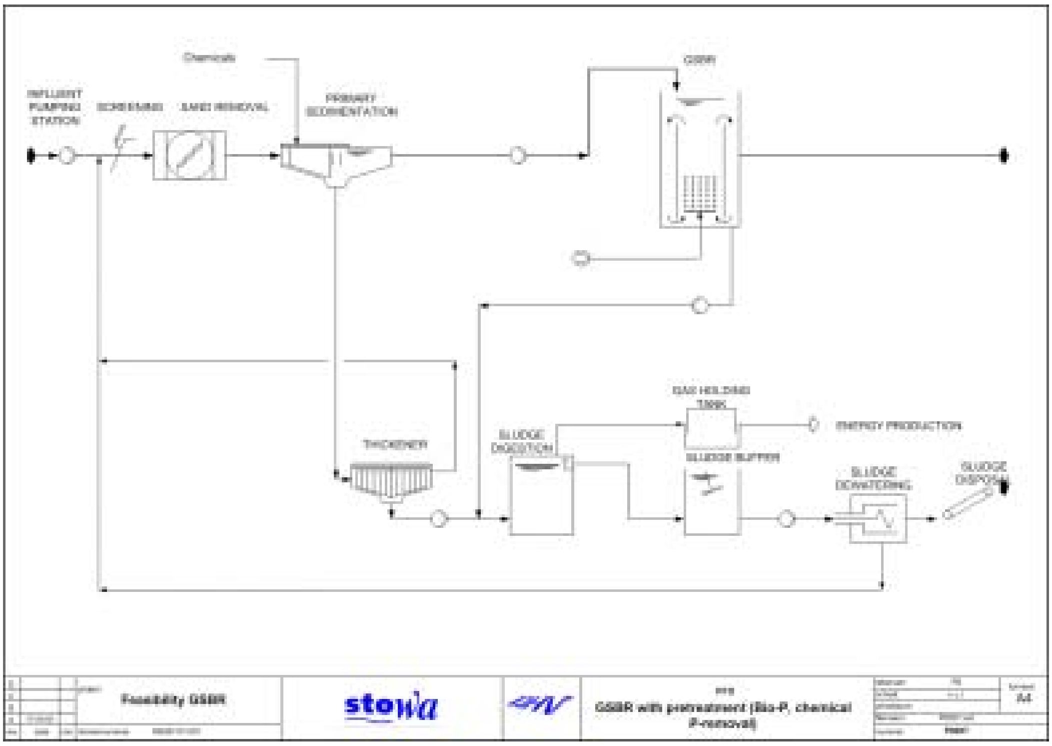
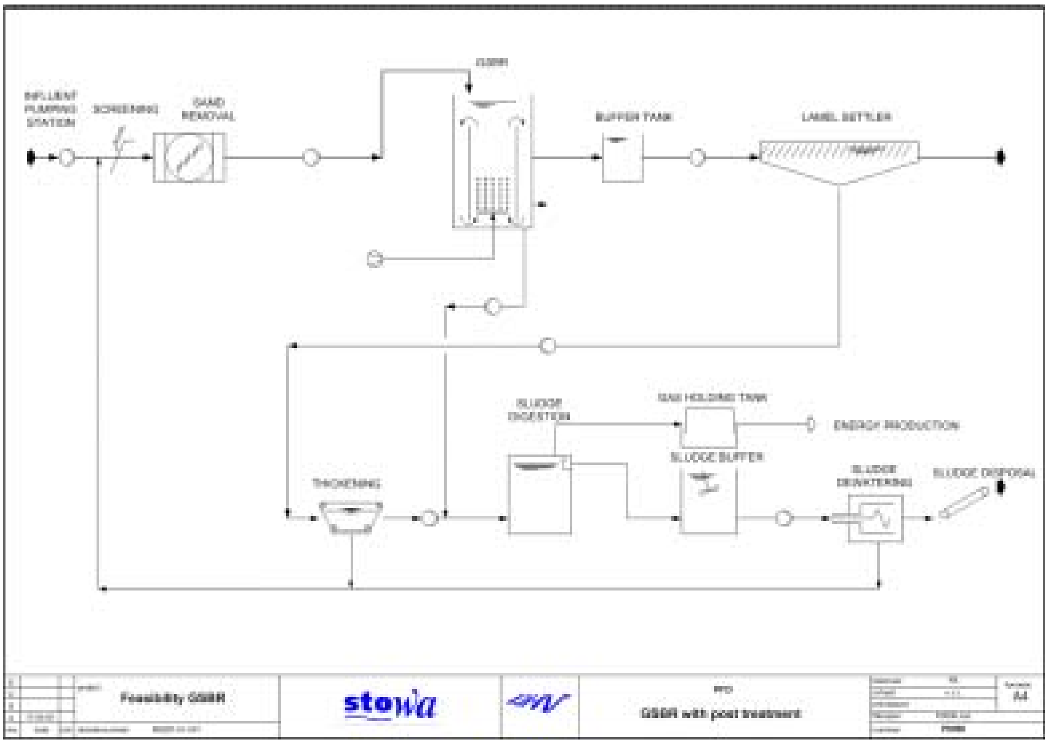



**FIGURE 2** **PROCESS FLOW DIAGRAM GSBR-ALTERNATIVE WITH POST TREATMENT**



# APPENDIX 3 DESIGN AND COSTS OF REFERENCE 1

| **REFERENCE ALTERNATIVE** | **Activated sludge with prim. Sed., Bio-P and sludge digestion** | | |
|---|---|---|---|
| **INFLUENT** | | **VALUE** | **UNIT** |
| COD | | 600 | mg/l |
| BOD | | 220 | mg/l |
| Nkj | | 55 | mg/l |
| SS | | 250 | mg/l |
| Average daily flow (ADF) | | **19,000** | m3/d |
| Rain weather flow (RWF) | | **4,000** | m3/h |
| Pollution load | | **118,939** | 136 TOD p.e. |
| **EFFLUENT** | | | |
| N-total | | **5** | mg/l |

| **Operational Costs** | | **Consumption/Production** | | **Costs [€/year]** | |
|---|---|---|---|---|---|
| Sludge production | | 1,179 | TonDS/y | 377,303 | |
| Energy consumption aerators minus energy production | | 1,214,109 | kWh/y | 65,562 | |
| $FeCl_3$ | | 327 | Ton FeCl3 (40%) | 37,594 | |
| Poly electrolyte | | 32 | Ton/year | 114,121 | |
| Maintenance | | | | 136,306 | |
| TOTAL | | | | 730,886 | |
| **Capital costs** | | | | **Costs [€/year]** | |
| Civil | | | | 946,861 | |
| Mechanical | | | | 416,751 | |
| Electrical | | | | 166,700 | |
| Total | | | | 1,530,312 | |
| Interest land purchase | | | | 20,992 | |
| **TOTAL YEARLY COSTS** | | | | **2,282,190** | |
| | | | | **0.33** | €/m3 |

| **PROCESS UNIT** | **TYPE** | | **NUMBER** | **CAPACITY** | **UNIT** | **CONSTRUCTION COSTS** Civil [€] | Mechanical [€] |
|---|---|---|---|---|---|---|---|
| **DESIGN WATER TREATMENT** | | | | | | | |
| INFLUENT PUMPING STATION | PUMPS | | 3 | 1330 | m3/h | 204,201 | 183,781 |
| SCREENINGS REMOVAL | OPENING SIZE (mm) | 3 | 2 | 2000 | m3/h | 181,512 | 108,907 |
| SAND REMOVAL (WATER LINE) | FLAT | | 1 | 100 | m2 | 164,442 | 81,171 |
| PRIMARY CLARIFIER | CIRCULAR | | 1 | 36 | m | 457,095 | 81,050 |
| PREPRECIPITATION | | | NO | | | | |
| | FeCl3-tank | | | | m3 | | |
| SELECTOR | ANOXIC | | 1 | 494 | m3 | | |
| MIXERS | | | | | kW | | |
| ANAEROBIC TANK | | | 1 | 3125 | m3 | 539,545 | |
| MIXERS | kW | | 4 | 3.9 | kW | | 53,318 |
| AERATION TANK | PREDENITRIFICATION | | 2 | 7092 | m3 | 1,899,824 | |
| AERATOR | SURFACE | | 4 | 90 | kW | | 149,529 |
| SIMULTANEOUS PRECIPITATION | FeCl3-DOSING | | 2 | 25 | l/h | | 2,503 |
| | FeCl3-tank | | 1 | 7.0 | m3 | | 15,126 |
| SECONDARY CLARIFIER | CIRCULAR | | 3 | 45 | m | 1,766,701 | 398,597 |
| RETURN SLUDGE PUMP | FO | | 3 | 1011 | m3/h | 54,454 | 98,078 |
| REJECTION WATER | RECIRCULATION | | | 78 | m3/d | | |
| **DESIGN SLUDGE TREATMENT** | | | | | | | |
| PRIMARY SLUDGE PUMP | | | 2 | 9 | m3/h | 1 | 7,427 |
| PRIMARY SLUDGE THICKENER | GRAVITARY | | 1 | 6.2 | m | 55,779 | 28,468 |
| SURPLUS SLUDGE PUMP | | | 2 | 16 | m3/h | 2 | 8,814 |
| SURPLUS SLUDGE THICKENER | MECHANICAL | | 2 | 14 | m3/h | 0 | 152,498 |
| DIGESTION TANK | | | 1 | 1808 | m3 | 409,629 | 151,736 |
| GAS STORAGE TANK | | | 1 | 367 | m3 | 0 | 88,902 |
| GAS MOTOR | | | 1 | 128 | kW | | 144,805 |
| SLUDGE BUFFER | | | 1 | 271 | m3 | 86,773 | |
| Mixers | | 7 W/m3 | 1 | 1.5 | kW | | 7,584 |
| SLUDGE DEWATERING | CENTRIFUGE | | 2 | 10 | m3/h | | 254,117 |
| PE DOSING | LIQUID | | 2 | 70 | l/h | | 119,279 |
| DEWATERED SLUDGE PUMP | | | 2 | 2 | m3/h | | 34,363 |
| CENTRATE PUMP | | | 2 | 16 | m3/h | | 29,044 |
| SLUDGE STORAGE | CONTAINER | | 2 | 20 | m3 | | 9,076 |
| | | | | | | | |
| PIPING | Deel civiel | 6% | | | | 449,596 | |
| OTHER | Deel civiel | 6% | | | | 337,197 | |
| | | | | | | | |
| BUILDINGS | | | | | | | |
| Operation building | | | | | | 273,000 | |
| Sludge treatment | | | | | | 409,000 | |
| | | | | | | | |
| Subtotal | | | | | | 7,088,750 | 2,201,452 |

| **Construction costs** | | | **Contingency** | **Costs [€]** |
|---|---|---|---|---|
| Civil | | | 20% | 7,088,750 |
| Mechanical | | | 20% | 2,201,452 |
| Electrical | | 40% | 20% | 880,581 |
| Total construction costs | | | | 12,204,939 |
| Lay-out costs | | | 2% | 244,099 |
| Additional costs | | | 5% | 610,247 |
| Unexpected | | | 10% | 1,220,494 |
| Total additional costs | | | | 2,074,840 |
| Construction + additional costs | | | | 14,279,779 |
| Design & Engineering | | | 10% | 1,427,978 |
| Subtotal | | | | 15,707,756 |
| VAT | | | 19% | 2,984,474 |
| **Investments** | | | | 18,700,000 |
| | | | | |
| **Land purchase** | | | | 349,865 |

| **OTHER** | | | |
|---|---|---|---|
| Average oxygen consumption | | 11,137 | $kgO_2/d$ |
| Methane production | | 880 | $m^3/d$ |
| Footprint | | 15,413 | $m^2$ |

# APPENDIX 4 DESIGN AND COSTS OF REFERENCE 2

**REFERENCE ALTERNATIVE** **Activated sludge with preprecipitation and sludge digestion**

| **INFLUENT** | **VALUE** | **UNIT** |
|---|---|---|
| COD | 600 | mg/l |
| BOD | 220 | mg/l |
| Nkj | 55 | mg/l |
| SS | 250 | mg/l |
| Average daily flow (ADF) | **19,000** | m3/d |
| Rain weather flow (RWF) | **4,000** | m3/h |
| Pollution load | **118,939** | 136 TOD p.e. |

| **EFFLUENT** | | |
|---|---|---|
| N-total | **10** | mg/l |

| **Operational Costs** | **Consumption/Production** | | **Costs [€/year]** | |
|---|---|---|---|---|
| Sludge production | 1,683 | TonDS/y | 534,081 | |
| Energy consumption aerators minus energy production | 848,033 | kWh/y | 45,794 | |
| $FeCl_3$ | 1,551 | Ton $FeCl_3$ (40%) | 179,390 | |
| Poly electrolyte | 47 | Ton/year | 168,035 | |
| Maintenance | | | 132,313 | |
| TOTAL | | | 1,059,332 | |
| **Capital costs** | | | **Costs [€/year]** | |
| Civil | | | 874,786 | |
| Mechanical | | | 416,170 | |
| Electrical | | | 166,468 | |
| Total | | | 1,457,423 | |
| Interest land purchase | | | 18,399 | |
| **TOTAL YEARLY COSTS** | | | **2,535,154** | |
| | | | **0.366** | €/m3 |

| **PROCESS UNIT** | **TYPE** | | **NUMBER** | **CAPACITY** | **UNIT** | **CONSTRUCTION COSTS** Civil [€] | **Mechanical** [€] |
|---|---|---|---|---|---|---|---|
| **DESIGN WATER TREATMENT** | | | | | | | |
| INFLUENT PUMPING STATION | PUMPS | | 3 | 1330 | m3/h | 204,201 | 183,781 |
| SCREENINGS REMOVAL | OPENING SIZE (mm) | 3 | 2 | 2080 | m3/h | 101,512 | 108,587 |
| SAND REMOVAL (WATER LINE) | FLAT | | 1 | 180 | m2 | 164,442 | 81,171 |
| PRIMARY CLARIFIER | CIRCULAR | | 1 | 36 | m | 457,035 | 81,050 |
| PREPRECIPITATION | $FeCl_3$-DOSING | | 2 | 140 | l/h | | 1,815 |
| | $FeCl_3$-tank | | | 25 | m3 | | 27,227 |
| SELECTOR | ANOXIC | | 1 | 466 | m3 | 177,465 | |
| MIXERS | | | 3 | 0.8 | kW | | 16,584 |
| ANAEROBIC TANK | | | NO | | | | |
| MIXERS | kW | | | | kW | | |
| AERATION TANK | PREDENITRIFICATION | | 2 | 5946 | m3 | 1,533,836 | |
| AERATOR | SURFACE | | 4 | 75 | kW | | 145,339 |
| SIMULTANEOUS PRECIPITATION | | | NO | | | | |
| | $FeCl_3$-tank | | 1 | | m3 | | |
| SECONDARY CLARIFIER | CIRCULAR | | 3 | 47 | m | 1,707,332 | 383,059 |
| RETURN SLUDGE PUMP | FO | | 3 | 879 | m3/h | 54,454 | 94,593 |
| REJECTION WATER | RECIRCULATION | | | 186 | m3/d | | |
| **DESIGN SLUDGE TREATMENT** | | | | | | | |
| PRIMARY SLUDGE PUMP | | | 2 | 21 | m3/h | 2 | 11,436 |
| PRIMARY SLUDGE THICKENER | GRAVITARY | | 1 | 9.2 | m | 83,816 | 32,047 |
| SURPLUS SLUDGE PUMP | | | 2 | 10 | m3/h | 1 | 7,460 |
| SURPLUS SLUDGE THICKENER | MECHANICAL | | 2 | 9 | m3/h | 8 | 127,270 |
| DIGESTION TANK | | | 1 | 2485 | m3 | 474,825 | 194,129 |
| GAS STORAGE TANK | | | 1 | 391 | m3 | 8 | 30,856 |
| GAS MOTOR | | | 1 | 136 | kW | | 154,512 |
| SLUDGE BUFFER | | | 1 | 373 | m3 | 104,106 | |
| Mixers | | 7 W/m3 | 1 | 2.6 | kW | | 9,487 |
| SLUDGE DEWATERING | CENTRIFUGE | | 2 | 10 | m3/h | | 254,117 |
| PE DOSING | LIQUID | | 2 | 70 | l/h | | 115,279 |
| DEWATERED SLUDGE PUMP | | | 2 | 2 | m3/h | | 34,046 |
| CENTRATE PUMP | | | 2 | 16 | m3/h | | 28,283 |
| SLUDGE STORAGE | CONTAINER | | 2 | 20 | m3 | | 9,076 |
| | | | | | | | |
| PIPING | Deel civiel | 6% | | | | 411,436 | |
| OTHER | Deel civiel | 6% | | | | 308,577 | |
| | | | | | | | |
| BUILDINGS | | | | | | | |
| Operation building | | | | | | 273,000 | |
| Sludge treatment | | | | | | 409,000 | |
| | | | | | | | |
| Subtotal | | | | | | 6,544,969 | 2,196,974 |

| **Construction costs** | | **Contingency** | **Costs [€]** |
|---|---|---|---|
| Civil | | 20% | 6,544,969 |
| Mechanical | | 20% | 2,196,974 |
| Electrical | 40% | 20% | 878,790 |
| Total construction costs | | | 11,544,867 |
| Lay-out costs | | 2% | 230,897 |
| Additional costs | | 5% | 577,243 |
| Unexpected | | 10% | 1,154,487 |
| Total additional costs | | | 1,962,627 |
| Construction + additional costs | | | 13,507,495 |
| Design & Engineering | | 10% | 1,350,749 |
| Subtotal | | | 14,858,244 |
| VAT | | 19% | 2,823,066 |
| **Investments** | | | 17,730,000 |
| | | | |
| **Land purchase** | | | 306,645 |

| **OTHER** | | |
|---|---|---|
| Average oxgen consumption | 9,694 | $kgO_2/d$ |
| Methane production | 939 | $m^3/d$ |
| Footprint | 13,509 | $m^2$ |

# APPENDIX 5 DESIGN AND COSTS OF REFERENCE 3

**REFERENCE ALTERNATIVE** **Activated sludge with Bio-P**

| **INFLUENT** | **VALUE** | **UNIT** |
|---|---|---|
| COD | 600 | mg/l |
| BOD | 220 | mg/l |
| Nkj | 55 | mg/l |
| SS | 250 | mg/l |
| Average daily flow (ADF) | **19,000** | m3/d |
| Rain weather flow (RWF) | **4,000** | m3/h |
| Pollution load | **118,939** | 136.100 p.e. |

| **EFFLUENT** | | |
|---|---|---|
| N-total | **10** | mg/l |

| **Operational Costs** | **Consumption/Production** | | **Costs [€/year]** |
|---|---|---|---|
| Sludge production | 1,658 | TonDS/y | 530,492 |
| Energy consumption aerators minus energy production | 2,595,418 | kWh/y | 140,153 |
| $FeCl_3$ | 411 | Ton FeCl3 (40%) | 47,310 |
| Poly electrolyte | 46 | Ton/year | 164,547 |
| Maintenance | | | 107,249 |
| TOTAL | | | 989,751 |
| **Capital costs** | | | **Costs [€/year]** |
| Civil | | | 790,754 |
| Mechanical | | | 317,376 |
| Electrical | | | 126,950 |
| Total | | | 1,235,080 |
| Interest land purchase | | | 19,781 |
| **TOTAL YEARLY COSTS** | | | **2,244,612** |
| | | | **0.32** €/m3 |

| **PROCESS UNIT** | **TYPE** | | **NUMBER** | **CAPACITY** | **UNIT** | **CONSTRUCTION COSTS** Civil [€] | Mechanical [€] |
|---|---|---|---|---|---|---|---|
| **DESIGN WATER TREATMENT** | | | | | | | |
| INFLUENT PUMPING STATION | PUMPS | | 3 | 1330 | m3/h | 204,281 | 183,781 |
| SCREENINGS REMOVAL | OPENING SIZE (mm) | 3 | 2 | 2000 | m3/h | 181,512 | 108,907 |
| SAND REMOVAL (WATER LINE) | FLAT | | 1 | 100 | m2 | 164,442 | 81,171 |
| PRIMARY CLARIFIER | | | NO | | | | |
| PREPRECIPITATION | | | NO | | | | |
| | FeCl3-tank | | | | m3 | | |
| SELECTOR | ANOXIC | | 1 | 534 | m3 | | |
| MIXERS | | | | | Kw | | |
| ANAEROBIC TANK | | | 1 | 2632 | m3 | 476,385 | |
| MIXERS | kW | | 3 | 4.4 | Kw | | 44,353 |
| AERATION TANK | PREDENITRIFICATION | | 2 | 6834 | m3 | 1,871,714 | |
| AERATOR | SURFACE | | 4 | 90 | Kw | | 149,929 |
| SIMULTANEOUS PRECIPITATION | FeCl3-DOSING | | 2 | 30 | l/h | | 2,666 |
| | FeCl3-tank | | 1 | 9.0 | m3 | | 16,639 |
| SECONDARY CLARIFIER | CIRCULAR | | 3 | 48 | m | 1,714,186 | 383,965 |
| RETURN SLUDGE PUMP | FQ | | 3 | 883 | m3/h | 54,454 | 95,059 |
| REJECTION WATER | RECIRCULATION | | | 50 | m3/d | | |
| **DESIGN SLUDGE TREATMENT** | | | | | | | |
| PRIMARY SLUDGE PUMP | | | NO | | | | |
| PRIMARY SLUDGE THICKENER | | | NO | | | | |
| SURPLUS SLUDGE PUMP | | | 2 | 27 | m3/h | 2 | 14,280 |
| SURPLUS SLUDGE THICKENER | MECHANICAL | | 2 | 22 | m3/h | 0 | 180,521 |
| DIGESTION TANK | | | NO | | | | |
| GAS STORAGE TANK | | | NO | | | | |
| GAS MOTOR | | | NO | | | | |
| SLUDGE BUFFER | | | 1 | 227 | m3 | 77,119 | |
| Mixers | | 7 W/m3 | 1 | 1.6 | Kw | | 7,423 |
| SLUDGE DEWATERING | CENTRIFUGE | | 2 | 5 | m3/h | | 226,890 |
| PE DOSING | LIQUID | | 2 | 60 | l/h | | 114,093 |
| DEWATERED SLUDGE PUMP | | | 2 | 2 | m3/h | | 31,795 |
| CENTRATE PUMP | | | 2 | 10 | m3/h | | 22,607 |
| SLUDGE STORAGE | CONTAINER | | 2 | 20 | m3 | | 9,076 |
| | | | | | | | |
| PIPING | Deel civiel | 8% | | | | 379,515 | |
| OTHER | Deel civiel | 6% | | | | 284,636 | |
| | | | | | | | |
| BUILDINGS | | | | | | | |
| Operation building | | | | | | 227,000 | |
| Sludge treatment | | | | | | 273,000 | |
| | | | | | | | |
| Subtotal | | | | | | 5,908,085 | 1,673,125 |

| **Construction costs** | | **Contingency** | **Costs [€]** |
|---|---|---|---|
| Civiel | | 20% | 5,908,085 |
| Mechanical | | 20% | 1,673,125 |
| Electrical | 40% | 20% | 669,250 |
| Total construction costs | | | 9,900,551 |
| Lay-out costs | | 2% | 198,011 |
| Additional costs | | 5% | 495,028 |
| Unexpected | | 10% | 990,055 |
| Total additional costs | | | 1,683,094 |
| Construction + additional costs | | | 11,583,645 |
| Design & Engineering | | 10% | 1,158,364 |
| Subtotal | | | 12,742,009 |
| VAT | | 19% | 2,420,382 |
| **Investments** | | | 15,200,000 |

| **Land purchase** | 329,681 |
|---|---|

| **OTHER** | | |
|---|---|---|
| Average oxygen consumption | 12,698 | $kgO_2/d$ |
| Methane production | 0 | $m^3/d$ |
| Footprint | 14,523 | $m^2$ |

# APPENDIX 6 DESIGN AND COSTS GSBR WITH POST TREATMENT AND BIO-P

| CONFIGURATION OVERVIEW | | |
|---|---|---|
| P removal | Biological | |
| Pretreatment | No pretreatment | |
| Cycle time | 60 | min |
| Reactor type | Bubble column | |
| Buffer effluent GSB | yes | |
| Post treatment | Lamella settlers | |
| Discharge method | Separate fill and draw GSBR | |

| INFLUENT | Value | Unit |
|---|---|---|
| COD | 600 | mg/l |
| BOD | 220 | mg/l |
| Nkj | 55 | mg/l |
| SS | 250 | mg/l |
| Ptotal | 9.0 | mg/l |
| Average daily flow (ADF) | 19,000 | m3/d |
| Rain weather flow (RWF) | 4,006 | m3/h |
| Dry weather flow (DWF) | 1,583 | m3/h |
| Pollution load | 119,000 | 136 g TOD p.e. |

| EFFLUENT | Concentration | |
|---|---|---|
| COD | 45 | mg/l |
| SS | 10 | mg/l |
| N sludge | 0.5 | mg/l |
| NH4 | 1.5 | mg/l |
| NO3 | 8.0 | mg/l |
| Ntotal | 10.0 | mg/l |
| Ptotal | 0.5 | mg/l |

| OPERATIONAL COSTS | Consumption | Production | Costs | |
|---|---|---|---|---|
| Sludge production | 1,095 | ton DS/j | 351,360 | €/year |
| Energy use airlift and other main users minus | 1,984,722 | kWh/j | 107,175 | €/year |
| $FeCl_3$ | 0 | ton/j | 0 | €/year |
| Poly electrolyte | 22 | ton/j | 78,083 | €/year |
| Maintenance | | | 168,395 | €/year |
| TOTAL | | | 704,013 | €/year |
| **Capital costs** | | | **Costs** | |
| Civil | | | 336,368 | €/year |
| Mechanical | | | 863,155 | €/year |
| Electrical | | | 265,262 | €/year |
| Total | | | 1,464,785 | €/year |
| Interest land purchase | | | 4,133 | €/year |
| TOTAL YEARLY COSTS | | | **2,170,000** | €/year |
| | | | **0.313** | €/m3 |

| PROCESS UNIT | | NUMBER | CAPACITY | UNIT | CONSTRUCTION COSTS | |
|---|---|---|---|---|---|---|
| | | | | | Civil | Mechanical |
| Influent pumping station | | 3 | 1,335 | m3/h | 204,808 | 184,148 |
| Screens | | 2 | 2,003 | m3/h | 181,888 | 108,880 |
| Sand removal | | 1 | 100 | m2 | 164,473 | 81,224 |
| Primary clarifier | no | FALSE | FALSE | m | FALSE | FALSE |
| FeCl3-dosing | | FALSE | FALSE | l/h | | FALSE |
| FeCl3-tank | | FALSE | FALSE | m3 | | FALSE |
| Pumping station GSBRs | | 3 | 1,335 | m3/h | 204,808 | 184,148 |
| GSBRs | Bubble column | | | | | |
| | round | 3 | 3,779 | m3 | 1,128,101 | |
| | FALSE | | FALSE | m2 | FALSE | |
| Compressors | | 3 | 6,121 | m3/h | | 242,816 |
| Diffusors | | 3 | 6,121 | m3/h | | 91,863 |
| Discharge construction | | 3 | 5,847 | m3/h | 0 | 446,209 |
| Effluent buffer tank | yes | 1 | 1,469 | m3 | 162,838 | |
| Effluent pumping station | | 3 | 1,335 | m3/h | 204,808 | 184,148 |
| Lamella settlers | yes | 1 | 4,006 | m3/h | 274,478 | 878,268 |
| Conventional sedimentatio | no | FALSE | FALSE | m | FALSE | FALSE |
| Hydrocyclones | no | FALSE | FALSE | m3/h | | FALSE |
| Sand filtration | no | FALSE | FALSE | m | FALSE | FALSE |
| Flush pump + blower | | FALSE | FALSE | m3/h | | FALSE |
| Clean water tank | | FALSE | 136 | m3 | | FALSE |
| Primary sludge thickener | no | FALSE | FALSE | m | FALSE | FALSE |
| Primary sludge pump | | FALSE | FALSE | m3/h | | FALSE |
| Surplus sludge pump | yes | 2 | 35 | m3/h | | 22,743 |
| Mechanical thickening | yes | 2 | 35 | m3/h | | 197,148 |
| Digestion tank | | 1 | 1,016 | m3 | 410,364 | 152,124 |
| Gas holding tank | | 1 | 326 | m3 | | 65,858 |
| Gas engine | | 1 | 113 | kW | | 128,590 |
| Sludge holding tank | | 1 | 272 | m3 | | 67,808 |
| Mixers sludge holding tank | | 1 | 2 | | | 8,808 |
| Sludge dewatering | | 2 | 10 | m3/h | | 284,117 |
| PE dosing sludge dewatering | | 1 | | | | 113,446 |
| Centrate pump | | 2 | 11 | m3/h | | 23,816 |
| Dewatered sludge pump | | 2 | 2 | m3/h | | 31,800 |
| Sludge containers | | 2 | | | | 9,076 |
| | | | | | | |
| Piping | | 8% | | | 234,858 | |
| Other | | 6% | | | 176,144 | |
| | | | | | | |
| BUILDINGS | | | | | | |
| Operation building | | | | | 272,368 | |
| Sludge treatment | | | | | 408,402 | |
| | | | | | | |
| Subtotal | | | | | 4,027,404 | 3,513,405 |

| Construction costs | | Contingency | Costs (€) |
|---|---|---|---|
| Civil | | 20% | 4,027,404 |
| Mechanical | | 20% | 3,513,405 |
| Electrical | 40% | 20% | 1,405,362 |
| Total construction costs | | | 10,736,405 |
| Lay-out | | 2% | 214,768 |
| Additional costs | | 5% | 536,770 |
| Unexpected | | 10% | 1,073,641 |
| Subtotal | | | 1,825,019 |
| Construction + additional costs | | | 12,560,424 |
| Design & Engineering | | 10% | 1,256,042 |
| Subtotal | | | 13,816,466 |
| VAT | | 19% | 2,625,129 |
| **Investments** | | | 16,400,000 |
| | | | |
| **Land purchase** | | | 68,889 |

| OTHER | | | |
|---|---|---|---|
| Average oxygen consumption | | | kgO2/d |
| Methane production | | 782 | m³/d |
| Footprint | | 3,036 | m² |

# APPENDIX 7 DESIGN AND COSTS GSBR WITH PRIMARY SEDIMENTATION AND BIO-P

| CONFIGURATION OVERVIEW | |
|---|---|
| P removal | Biological |
| Pretreatment | Primary sedimentation + polymer dosing |
| Cycle time | 60 min |
| Reactor type | Bubble column |
| Buffer effluent GSB | yes |
| Post treatment | No post treatment |
| Discharge method | Separate fill and draw GSBR |

| INFLUENT | Value | Unit |
|---|---|---|
| COD | 600 | mg/l |
| BOD | 220 | mg/l |
| Nkj | 55 | mg/l |
| SS | 250 | mg/l |
| Ptotal | 9.0 | mg/l |
| Average daily flow (ADF) | 19,000 | m3/d |
| Rain weather flow (RWF) | 4,086 | m3/h |
| Dry weather flow (DWF) | 1,583 | m3/h |
| Pollution load | 119,080 | 136 gTOD p.e. |

| EFFLUENT | Concentration | |
|---|---|---|
| COD | 65 | mg/l |
| SS | 57 | mg/l |
| N-sludge | 4.0 | mg/l |
| NH4 | 1.5 | mg/l |
| NO3 | 4.5 | mg/l |
| Ntotal | 10.0 | mg/l |
| Ptotal | 1.0 | mg/l |

| OPERATIONAL COSTS | Consumption/Production | | Costs | |
|---|---|---|---|---|
| Sludge production | 1,182 | ton DS/j | 378,311 | €/year |
| Energy use aeration and other main users minus | 1,757,886 | kWh/j | 56,931 | €/year |
| $FeCl_3$ | 0 | ton/j | 0 | €/year |
| Poly electrolyte | 51 | ton/j | 164,084 | €/year |
| Maintenance | | | 122,331 | €/year |
| TOTAL | | | 700,558 | €/year |
| **Capital costs** | | | **Costs** | |
| Civil | | | 492,786 | €/year |
| Mechanical | | | 464,584 | €/year |
| Electrical | | | 185,826 | €/year |
| Total | | | 1,143,196 | €/year |
| Interest land purchase | | | 3,646 | €/year |
| TOTAL YEARLY COSTS | | | **1,839,000** | €/year |
| | | | **0,278** | €/m3 |

| PROCESS UNIT | | NUMBER | CAPACITY | UNIT | CONSTRUCTION COSTS Civil | CONSTRUCTION COSTS Mechanical |
|---|---|---|---|---|---|---|
| Influent pumping station | | 3 | 1,395 | m3/h | 284,800 | 184,140 |
| Screens | | 2 | 2,003 | m3/h | 181,886 | 108,981 |
| Sand removal | | 1 | 100 | m2 | 184,473 | 61,224 |
| Primary clarifier | yes | 1 | 38 | m | 484,802 | 81,380 |
| FeCl3-dosing | | FALSE | FALSE | l/h | | FALSE |
| FeCl3-tank | | FALSE | FALSE | m3 | | FALSE |
| Pumping station GSBRs | | 3 | 1,395 | m3/h | 284,800 | 184,140 |
| GSBRs | Bubble column | | | | | |
| | round | 3 | 2,868 | m3 | 924,459 | |
| | FALSE | | FALSE | m2 | FALSE | |
| Compressors | | 3 | 7,030 | m3/h | | 378,259 |
| Diffusers | | 3 | 7,030 | m3/h | | 106,274 |
| Discharge construction | | 3 | 5,647 | m3/h | 0 | 446,289 |
| Effluent buffer tank | no | FALSE | FALSE | m3 | FALSE | |
| Effluent pumping station | | FALSE | FALSE | m3/h | FALSE | FALSE |
| Lamella settlers | no | FALSE | FALSE | m3/h | FALSE | FALSE |
| Conventional sedimentation | no | FALSE | FALSE | m | FALSE | FALSE |
| Hydrocyclones | no | FALSE | FALSE | m3/h | | FALSE |
| Sand filtration | no | FALSE | FALSE | m | FALSE | FALSE |
| Flush pump + blower | | FALSE | FALSE | m3/h | | FALSE |
| Clean water tank | | FALSE | 136 | m3 | | FALSE |
| Primary sludge thickener | yes | 1 | 8 | m | 72,896 | 31,110 |
| Primary sludge pump | | 2 | 24 | m3/h | | 20,885 |
| Surplus sludge pump | no | FALSE | FALSE | m3/h | | FALSE |
| Mechanical thickening | no | FALSE | FALSE | m3/h | | FALSE |
| Digestion tank | | 1 | 1,988 | m3 | 421,718 | 158,923 |
| Gas holding tank | | 1 | 351 | m3 | | 67,754 |
| Gas engine | | 1 | 122 | kW | | 138,446 |
| Sludge holding tank | | 1 | 65 | m3 | | 23,479 |
| Mixers sludge holding tank | | 1 | 0 | | | 4,915 |
| Sludge dewatering | | 2 | 5 | m3/h | | 226,080 |
| PE dosing sludge dewatering | | 2 | | | | 226,080 |
| Centrate pump | | 2 | 12 | m3/h | | 24,795 |
| Dewatered sludge pump | | 2 | 2 | m3/h | | 32,849 |
| Sludge containers | | 2 | | | | 9,076 |
| | | | | | | |
| Piping | | 8% | | | 211,881 | |
| Other | | 6% | | | 150,208 | |
| | | | | | | |
| BUILDINGS | | | | | | |
| Operation building | | | | | 272,268 | |
| Sludge treatment | | | | | 485,402 | |
| | | | | | | |
| Subtotal | | | | | 3,686,144 | 2,453,212 |

| Construction costs | | Contingency | Costs (€) |
|---|---|---|---|
| Civil | | 20% | 3,686,144 |
| Mechanical | | 20% | 2,453,212 |
| Electrical | 40% | 20% | 981,285 |
| Total construction costs | | | 8,547,188 |
| Lay-out | | 2% | 170,940 |
| Additional costs | | 5% | 427,358 |
| Unexpected | | 10% | 854,717 |
| Subtotal | | | 1,453,019 |
| Construction + additional costs | | | 10,000,188 |
| Design & Engineering | | 10% | 1,000,019 |
| Subtotal | | | 11,000,207 |
| VAT | | 19% | 2,090,039 |
| **Investments** | | | 13,100,000 |
| | | | |
| **Land purchase** | | | 60,808 |

| OTHER | | | |
|---|---|---|---|
| Average oxygen consumption | | | $kgO_2/d$ |
| Methane production | | 842 | $m^3/d$ |
| Footprint | | 2,879 | $m^2$ |