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ANAEROBIC TREATMENT OF
WASTEWATER CONTAINING
FATTY ACIDS
IN UPFLOW REACTORS

BIBLIOTHEEK
DER
TECHNISCHE HOGESCHOOL
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ANAEROBIC TREATMENT OF WASTEWATER CONTAINING FATTY ACIDS IN UPFLOW REACTORS

PROEFSCHRIFT

TER VERKRIJGING VAN DE GRAAD VAN DOCTOR
IN DE TECHNISCHE WETENSCHAPPEN AAN DE
TECHNISCHE HOGESCHOOL DELFT, OP GEZAG
VAN DE RECTOR MAGNIFICUS PROF. DR. IR.
F. J. KIEVITS, VOOR EEN COMMISSIE AAN-
GEWEZEN DOOR HET COLLEGE VAN DEKANEN
TE VERDEDIGEN OP WOENSDAG 7 NOVEMBER
1979 TE 16.00 UUR.

DOOR

ROBERT RENE VAN DER MEER

scheikundig ingenieur
geboren te 's-Gravenhage

1602 4242

Dit proefschrift is goedgekeurd door de promotor
PROF. DR. IR. P. M. HEERTJES

Aan mijn vader.
Aan Lies.

DANKWOORD

Op deze plaats betuig ik graag mijn hartelijke dank aan allen die een bijdrage hebben geleverd aan het werk dat heeft geleid tot dit proefschrift.

Mijn dank gaat vooral uit naar :

- De afstudeerders : Ir. C. Buijs (Hoofdstukken IV en VI), Ir. J.W. Buijs (II en VI), Mej.Ir. A.D. Hulsmann (I), Ir. G.J. Keilman (I en II), Ir. L.J. Kuijvenhoven (III), Ir. J.H.G. Lok (III), en Ir. R.P. Plantenberg (V).
- Ir. R. de Vletter en de andere medewerkers van de Afdeling Milieuzaken van de Centrale Suiker Maatschappij.
- De leden van de Werkgroep Anaerobe Zuivering.
- Mej. Th.E. van Bruggen, de Heren L. Boshuizen, J.J.B. van Holst, W.J. Jongeleen, J.H. Kamps, M. den Os, G.L.H. van Schaik, J. Sikkens en de andere medewerkers van het analyselaboratorium, de werkplaats, de elektronische dienst en de huishoudelijke dienst van het Laboratorium voor Chemische Technologie.

De Minister van Volksgezondheid en Milieuhygiëne en de Direktie van de Centrale Suiker Maatschappij ben ik erkentelijk voor de financiering van een belangrijk gedeelte van het onderzoek, dat daardoor kon worden uitgevoerd met reaktoren van semi-technische en technische schaal.

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HISTORICAL BACKGROUND

The microbiological conversion of organic matter into methane has attracted the interest of chemists, (micro)biologists and sanitary engineers, already for two centuries. Alexandre Volta demonstrated in 1776 (1) that "combustible air" that had been liberated by stirring in marshy pools, could be collected and used for many experiments. He was the first who related the decay of organic matter to the formation of gas.

Systematic research started at the end of the last century with the work of Hoppe-Seyler, Omeliansky, Mazé, and others (2). At the beginning of this century, Söhngen conducted an interesting study at the Delft University of Technology (3), in which he described the digestion of the salts of fatty acids by enriched cultures in the absence of oxygen, with methane and carbon dioxide as the main products. He found that other organic compounds such as albumen, sugars and cellulose can also be digested, and discovered that hydrogen gas could be used by the anaerobic bacteria to convert carbon dioxide into methane. This reaction was more thoroughly investigated by Schnellen, also in Delft (4).

The biological conversion of waste into other products that were less water polluting and had a less offensive odour, had already been applied for a long time : the septic tank, for example, was introduced by Mouras in 1860 (5). In such installations the efficiency of the water purification is governed by anaerobic processes, which was in 1920 understood by Groenewege (6). This investigator applied the fundamental microbiological research of Söhngen to the anaerobic digestion process for wastewater treatment. He considered the conversion of more complicated organic molecules into organic acids and alcohols as the first stage of the anaerobic purification process and stated that in a second stage these products would be converted into methane (having shown in the laboratory that most of these intermediary products were easily converted into methane). In addition, he speculated that the slow growth of the methane bacteria, observed in the laboratory, could cause the slow start-up of the septic tanks.

These early studies of municipal installations gave technological data and recommendations of interest; for instance, it appeared possible to stop the feed of the installations for months without a significantly lower biological activity of the sludge at a new start-up, although it was better to supply the waste continuously, as an equilibrium could then be established between the growth of new biomass and the washout of sludge in the effluent.

Groenewege concluded that washout occurs because biomass is whirled up by

the rising gas bubbles and circulates through the whole tank, and therefore can easily be carried off with the effluent.

Anaerobic digestion has been applied for the treatment of sewage sludge since the last decade of the previous century. By 1920, the development of this process (that took place mainly in Germany, Great Britain and the U.S.A.), made this an established operation at sewage treatment plants, despite the fact that digesters were often unreliable.

In the 1930's, investigators such as Buswell et al. (7) looked for possibilities to apply the anaerobic digestion process to the purification of industrial wastewaters (dairies, distilleries, etc.). They were reasonably successful, because at liquid residence times of 20-30 days, purification efficiencies of up to 90-95 % were obtained. For example, it was calculated that in the case of dairy wastewater, treatment of 1 kg (dry solids) of milk waste products required a reactor volume of 0.372 m^3 . At a price of about \$ 17.50 per m^3 reactor volume (in 1932) the total investment for 95 % purification of 1 kg milk waste products was estimated at \$ 6.57 (which was a low price). Moreover, the yield of gas was calculated to be $0.54\text{-}0.80 \text{ m}^3/\text{kg}$ (7).

In spite of these encouraging results, the process did not find practical application. One of the reasons was that anaerobic treatment of wastewater was thought to be too delicate an operation in relation to process control; because if process control failed, fatty acids were produced, which caused environmental nuisance and could also result in the process going down. In the latter situation, new start-up was a tedious procedure. Other reasons had to do with the technological improvement of industrial processes to which anaerobic treatment could be applied; with the fact that the concentration of the waste and the temperature of the wastewater decreased by the increased use of cheap process water; and finally, with the low price of energy that made the profit of the produced biogas only marginal.

Around 1960, McCarty (8) recognized that the improvement of the technology gave new possibilities to the anaerobic treatment of wastewater. This resulted in the application of a number of full-scale installations which operated with varying success (9-13).

However, a real breakthrough in industrial wastewater treatment was not attained for the following reasons :

- the organic loadings that could be applied (to about $5 \text{ kg COD}/\text{m}^3\text{d}$), were of the same order of magnitude as the loadings applied in highly loaded active aerobic sludge installations, (the latter being thought of as a more reliable process);

- because of the slow growth of methane forming bacteria, a long sludge retention time is necessary (10-30 days), while for economic reasons a short liquid residence time is required.

Lately, because of the energy crisis and the introduction of environmental legislation (which in many countries is coupled with a tax system), there has been an alteration in the costs of disposal and aerobic treatment of wastewater, leading to more interest in the anaerobic process. The purpose of the study described in this thesis, was to investigate whether the technology of this process could be improved to such an extent that, besides from the economical point of view, it is also technologically attractive and preferable.

CHAPTER I

GENERAL INTRODUCTION OF THE STUDY OF THE UPFLOW REACTOR

I.1 Introduction

As one of the possibilities for reducing water pollution, anaerobic treatment can be useful in a concentration range of the waste of $0.4 - 10 \text{ kg TOC/m}^3$. In this process, wastewater is biologically purified by anaerobic bacteria that convert the organic pollutants into methane and carbon dioxide (14-19). The process shall be technologically attractive if it is simple and reliable, if short liquid residence times (2 - 24 h) can be applied, and if the reactor can be highly loaded (at least $4 \text{ kg organic carbon per m}^3$ reactor volume per day). From introductory studies (20) it was apparent that such favourable conditions may be obtained if the methane forming reactions of the process are performed in a separate reactor. In such a case, two reactors are used : in the first reactor, hydrolysis and acid formation take place; in the second reactor, the fatty acids are converted into methane and carbon dioxide (21).

For the methane forming step, an upflow reactor with an internal settler appeared to be most appropriate in comparison with the upflow filter and the stirred tank with separation and recirculation of the sludge (20). It was concluded that with the upflow reactor, high purification efficiencies at short liquid residence times can be attained if :

- a sludge bed - sludge blanket system is used;
- the influent is uniformly distributed over the bottom of the reactor;
- the contents of the reactor is mixed by the gas produced;
- an effective separation of effluent, gas and biomass is obtained, with the biomass recirculated to the reactor.

By the separation of the methane forming bacteria from the acid formers, optimal process conditions can be chosen and maintained for both steps separately, so that optimal enrichment of the bacterial cultures can be obtained (22). In mixed cultures as used, adaptation of the bacteria takes place by changes of their phenotype and by preferential growth (23, 24). Because the rate of growth of methane forming bacteria is very small, their adaptation is very slow; consequently it will take a long time before a stationary situation is obtained. Before it is possible to establish the optimum conditions for the conversion of fatty acids, more insight in the process is necessary (25-29). Therefore, the intention of the study described in this thesis, was to obtain more technological know-how of the process. To this end, experiments were performed in upflow reactors of different size.

Because, for the greater part the work has been carried out with pilot-plants and a full-scale reactor, which were just constructed, most of the experiments were performed under conditions that were not optimal and only steady over short periods of time. Furthermore, the long adaptation time and the changes of the influent composition caused the properties of the sludge to vary during the experimental period.

In the following sections of this introductory chapter will be described : the performance of the processes taking place in the upflow reactor; the upflow reactors of different size as used; the most important process conditions during the experimental period; and the sludge properties, including the origin and history of the sludge.

I.2 Considerations about the upflow reactor

All the reactors used were designed according to the same principle, but on different scales : a laboratory reactor of approximately 60 l, pilot-plants of

6 m³ and of 30 m³, and a full-scale reactor of 200 m³. The last three reactors had been placed at sugar factories of the C.S.M.* in Breda and Halfweg.

The reactors contain a suspension of anaerobic sludge. The influent in which the organic compounds to be converted are dissolved, enters the reactor at the bottom; the effluent leaves the reactor via the settler at the top. During the residence time of the upflowing liquid, the main part of the substrate is converted into gas. In the internal settler at the top of the reactor, most of the

Fig. I-1. Upflow reactor with an internal settler for anaerobic treatment of waste water; (o) : gas bubbles.

anaerobic sludge that would otherwise leave the reactor together with the effluent, is separated by settling, and is recirculated into the reactor (see Fig. I-1).

* Centrale Suiker Maatschappij.

Fig. I-2. Whirling up of sludge by rising gas bubbles.

In the reactor, a bed-blanket system is established from the originally concentrated suspension of sludge with which the reactor was filled. The density and good settling properties of well adapted sludge cause the formation of a sludge bed with an almost constant sludge concentration of $50\text{--}100\text{ kg/m}^3$ suspended solids (SS); while the production of gas causes the formation of the sludge blanket with a sludge concentration between 0 and 40 kg SS/m^3 on top of the bed (see the Figs. I-2 and I-3). An area of transition can sometimes be present between the bed and the blanket : in such an area the sludge concentration decreases with the height.

In the stationary situation, the transport of sludge upwards from the bed to the blanket is equal to the sum of the washout and the transport downwards, the latter being due to settling and recirculation with the backmix flow. This downward liquid stream is initiated by the rising gas bubbles and their wake : the backmix flow compensates for the loss in volume in the bed. The concentration of sludge in the blanket is a function of : the volume of gas produced in the sludge bed per unit of time; the concentration of sludge in the bed; the settling velocity; the linear velocity of the liquid in the reactor. The concentration of sludge in the bed is a function of : the density of the wet sludge particles; the particle size distribution; the linear velocity of the liquid in the reactor; the volume of gas produced in the sludge bed per unit of time. The influence of these parameters however, does not appear to be very strong, as the sludge concentration in the bed remained almost constant.

If a reactor is started-up with sludge from a municipal digester, the sludge bed may be absent because of the small settling velocity of such sludge. Then,

Fig. I-3. The "bed-blanket reactor",
schematically.

even at low gas productions, the sludge is distributed uniformly over the reactor. Such a situation has to be avoided because the settler will then easily be overloaded, which results in washout of large amounts of sludge. Anyhow, after some time, by adaptation and loss of sludge, the settling properties will improve, resulting in a bed-blanket system. The behaviour of the liquid and of the sludge are interdependent and influence the process performance (30). Furthermore, the liquid flow pattern is also influenced by the way the influent is distributed over the bottom of the reactor and by the behaviour of the gas bubbles. The most effective contact between the

sludge and the substrate would be obtained by homogeneous distribution of the influent over the bottom, e.g. by means of a sieve plate or of a rotating distribution system. But unfortunately, the small openings of a sieve plate will soon be blocked, resulting in inhomogeneous distribution of the influent; on the other hand, rotating systems, if used in the upflow reactor, are not accessible for repair. Therefore, a limited number of influent points with relatively large openings had to be used. In the 30 m³ pilot-plant two different systems could be studied :

- a system by which at each influent point the influent streams horizontally, in unidirectional flow over the bottom of the reactor;
- a system by which the influent is introduced vertically and thereafter is distributed radially over the bottom around the influent opening by means of a horizontal plate.

In the 200 m³ full-scale reactor ten influent points for horizontal unidirectional flow could be used. In this reactor, experiments were performed to study the number of influent points that were necessary to obtain an effective influent distribution.

The use of a limited number of influent points, together with the fact that the influent has a lower density than the sludge bed, results in preferential flow through the sludge bed, and this leads to short-circuiting channels and

Fig. I-4. Block diagram, giving a qualitative picture of the fluid flow pattern in the upflow reactor.

dead space regions. The rising gas bubbles, produced in the sludge bed, strengthen the bypassing effects, but reduce the dead spaces. In the main part of the sludge bed, the bacteria are sufficiently supplied with substrate, so that enough gas is produced to mix these areas. In the lower parts of the bed, the gas bubbles rise slowly because of the compactness of the bed. At higher places and in the sludge blanket, the gas bubbles will rise faster, giving better mixing. The backmix flow from the blanket to the bed, is generated to replenish the volumes of the gas bubbles (plus their wake) that have already left the sludge bed. This flow rate is dependent on the flow rate of gas. In Fig. I-4, an example has been presented of a combination of the above mentioned flow regimes.

The influent that bypasses the sludge bed contains substrate that has to be converted in the sludge blanket and in the area of transition between the bed and the blanket. The amount of substrate that enters these areas per unit of time, may not be larger than the local maximum conversion rates, as overloading could result in substrate entering the settler. The sludge present in the settler then starts to produce gas which results in turbulence, and therefore, strongly increases the washout of sludge (31, 32). Overloading of the sludge blanket therefore aggravates this situation as the decrease in the amount of sludge increases the overloading. In this way malfunctioning of the reactor may occur, which can only be kept under control by diminishing the organic loading in the reactor.

From the above it is evident that before the effluent leaves the reactor, an effective gas - liquid - biomass separation has to be achieved. If no measures are taken to separate the sludge from the effluent and to recirculate it, the liquid residence time has to be of the same order of magnitude as the sludge retention time. This would restrict the liquid residence time to large values (10 - 30 days), as otherwise, the slow growth of sludge would not be sufficient to compensate for the loss of sludge by washout (33 - 35). In the reactors used

Fig. I-5a. Settler with circulation stream. (For a top view of these settlers, see Fig. I-17).

Fig. I-5b. Settler with upward fluid flow.

this problem is solved by means of internal settlers in which the sludge can settle and recirculate by gravity. The construction of the settlers is such that at first, by means of plates, a separation between the gas bubbles and the sludge suspension is obtained. In the Figures I-5a and I-5b the fluid and gas flow as occurring in and below the settlers used in the 200 m³ reactor have been indicated. The plates that cause a separation between the influent of the settler and the gas, have been indicated with A. The separation between the sludge and the effluent can take place in the settling compartments B, in which the fluid is in viscous flow. Through a round opening (settler of the 6 m³ reactor) or a slit (settlers of the 30 m³ and 200 m³ reactor) in the bottom (C), the sludge can return into the reactor (30, 36).

The settlers in the 200 m³ reactor can be used in two different ways :

- if the inflow openings (the pipes indicated with D in Fig. I-5a) and the recirculation opening C are separated, a circulation stream is maintained by the difference between the densities of the gas-liquid-biomass mixture in the blanket of the reactor and the thickened suspension of sludge at the bottom of the settler (D→E→C, see Fig. I-5a);
- if, by blocking the inflow pipes, the recirculation opening C is also used as the inflow opening, the fluid flow in the settler is mainly upwards.

The settlers of the laboratory reactors were constructed according to the first principle.

1.3 The anaerobic sludge used

The properties of the sludge are dependent on the history of the sludge. Therefore, in this section a survey will be given of the history of the anaerobic

sludge used, and then the properties of well adapted sludge will be described. The sludge with which the process was started, originated from the settler of a stationary operated municipal digester. In mid-July 1974, the 6 m³ pilot-plant was started-up with about 60 kg SS (about 50 % VSS) of sludge, and loaded with wastewater of a sugar factory of C.S.M. During the sugar campaign (September - December) the substrate was a neutralised solution of lower fatty acids. During the intercampaign (January - August) the substrate was a partly (20 - 40 %) soured sugar solution : wastewater taken from a factory for by-products. It was stored in a pond in which acid formation took place until the pH had decreased to about the value 4. The 6 m³ reactor operated for two years continuously, apart from short interruption periods due to the necessity of repair work. After the period of adaptation that lasted about one year, the sludge formed consisted of well distinguishable particles with good settling properties. At the end of the series of experiments in the 6 m³ reactor (August 1976), the amount of sludge appeared to have increased to about 500 kg SS (~ 77 % VSS).

This sludge was used to start-up the 30 m³ pilot-plant in August 1976. During a period of one year this reactor was continuously in operation in the same manner as the 6 m³ pilot-plant. In March 1977 the reactor was stopped for some days to replace the settler for an improved version (see Figs. I-13 and I-14). At the end of the period, the amount of sludge had been increased to about 1800 kg SS (about 80 % VSS), as a result of both internal growth and the addition of about 400 kg SS (about 50 % VSS) sludge from a municipal digester. All this sludge was used to start-up the full-scale 200 m³ reactor in September 1977. This reactor was in operation during the next three months (campaign 1977), thereafter the reactor had to be stopped because no substrate was available during the intercampaign. Also in this reactor, growth of sludge had taken place. The increase of the volatile suspended solids in the reactor was estimated at about 1300 kg, but this was mainly due to the accumulation of about 6000 kg mud (about 20 % VSS) during two weeks at the end of October and the beginning of November. This mud was present in the influent as a result of a wrong operation of the wastewater circuit of the sugar factory (37). As the presence of mud influences the settling behaviour of the anaerobic sludge particles, these two kinds of solids in the reactor had to be distinguished. After an operational period of about one year, the properties of the sludge did not change very pronouncedly any more, the sludge was adapted to the wastewater. The sludge particles, formed during the campaign of 1975 and later, consisted of about 80 % volatile matter. About a quarter of the remaining 20 % of fixed suspended solids was ferrous sulphide.

Measurements with a CHN-analyser indicated that the volatile fraction of the suspended solids consisted of 45 - 50 % Carbon, about 7 % Hydrogen and about 10 % Nitrogen.

In the Figs. I-6a and 6 b cumulative settling velocity distribution curves for

Fig. I-6a. Cumulative settling velocity for unhindered settling; anaerobic sludge fed with a partially acidified sugar solution.

Fig. I-6b. Cumulative settling velocity for unhindered settling; anaerobic sludge fed with a completely acidified sugar solution.

Fig. I-7. Cumulative settling velocities, measured at depths of 0.7 m (x), 1.2 m ($\square$) and 1.7 m (o).

unhindered settling of the sludge particles present in the inter-campaign and the campaign of 1976, have been presented. During the inter-campaign the settling velocity was somewhat smaller due to the presence of acid forming sludge. The good settling properties of the sludge were due to the cluster forming properties of the sludge particles; the form of the particles; and their relatively high

density ($1.05 - 1.07 \cdot 10^3 \text{ kg/m}^3$) that has probably been caused by the presence of ferrous sulphide. In Fig. I-7 the influence of this formation of clusters on unhindered settling can be seen.

The mud present in the 200 m^3 reactor during the second part of campaign 1977,

Fig. I-8. The mean settling velocity of anaerobic sludge particles as a function of the concentration of mud present : (●) measured in the batch settling column; (○) measured in the settler of the 200 m^3 reactor.

hindered free settling of the sludge particles. In Fig. I-8 a relation between the settling velocity of the sludge particles and the concentration of mud is presented (as measured for the sludge in the 200 m^3 reactor). From the particle size distributions, measured by means of wet sifting in March and September 1977 (see Figs. I-9a and 9b, it appeared that the average size of the particles had increased slowly with time. Therefore, one may predict that the multiplication of the anaerobic bacteria in the particles will further improve the settling properties.

Fig. I-9a. Particle size distribution of the sludge from the bed of the upflow reactor.

Fig. I-9b. Particle size distribution of the same sludge, half a year later.

I.4 Process performance, reactors used, process conditions.

I.4a Experiments with the laboratory reactor

The "two-reactor process" that was studied, took place firstly in a reactor in which a sugar solution was converted into a fatty acids solution and secondly in a reactor in which the effluent of the first reactor was treated to produce methane and purified water. In Fig. I-10 a flow diagram of the "two-

Fig. I-10. Flow sheet of the two-reactor process in the laboratory.

reactor process" and in Table I-1 the compositions and flow rates of the solutions used, have been presented. The first reactor, almost identical in design with the second, served only to produce the solution of fatty acids and was therefore operated almost stationary during the total experimental period. The composition and the flow rate of the influent obtained after dilution of a concentrated sugar solution, as well as of the effluent of the first reactor have been presented in Table I-2. Tap water could be supplied between the two reactors, which made variations in the hydraulic and organic loadings and in the influent concentrations of the second reactor possible. The height of the sludge bed in the second reactor, and the specific organic loading of the sludge could be varied independently of the reactor loading by supplying or discharging of sludge (this sludge stemmed from the 30 m³ pilot-plant). The laboratory reactor, in which the conversion of organic acids into methane takes place, is schematically presented in Fig. I-11. The following is indicated : where the influent enters (I) and the effluent leaves (E) the reactor;

TABLE I - 1.

Composition of the concentrated food-, nutrients- and buffer solutions, used for the "two-reactor process" as performed in the laboratory.

food solution	buffer solution
first reactor	first reactor
flow rate : 0.27/0.4 l/h	flow rate : 0.54 l/h
[sugar] = 256 kg/m ³	[NaOH] = 34 kg/m ³
[Mg ²⁺] = 0.31 "	[NaHCO ₃] = 42 "
[SO ₄ ²⁻] = 1.25 "	[K ⁺] = 0.72 "
[Fe ²⁺] = 0.9 "	[PO ₄ ³⁻] = 0.87 "
[NH ₄ ⁺] = 0.87 "	
buffer solution	nutrients solution
second reactor	second reactor
flow rate : 0.54 l/h	flow rate : 0.12 l/h
[NaOH] = 10 kg/m ³	[Mg ²⁺] = 0.78 kg/m ³
[NaHCO ₃] = 60 "	[SO ₄ ²⁻] = 3.12 "
[K ⁺] = 0.72 "	[Fe ²⁺] = 2.25 "
[PO ₄ ³⁻] = 0.87 "	[NH ₄ ⁺] = 16.8 "

where sample points (SP) are located; the construction of the settler (S); where the gas (G) leaves the system. A stirring mechanism (SM) worked continuously during the experiments on the kinetics of the reactions. The dimensions of the laboratory reactors are presented in Table I-3. The experiments performed with the second reactor can be classified in four categories :

- 1a. measurements on the carbon-balance;
- 1b. investigations of the kinetics of the process;
2. determining the fluid flow pattern under various process conditions;
3. study of the behaviour of the sludge in relation to the volume of gas produced per unit of time;
4. test of the substrate model.

The most important conditions during the experiments are presented in Table I-4.

TABLE I - 2.

Performance of the first reactor of the "two-reactor process".

Period I : experiments 1-4 on the carbon-balance
 Period II : experiments 5, 6 on the carbon-balance
 Period III : experiments on the kinetics and residence time distributions
 Period IV : experiments to check the substrate model

	Period I	Period II	Period III	Period IV
τ (h)	7.5-8	7.5-8	7.5-8	7.5-8
ϕ''_S (kg C/m ³ d)	~ 11.5	~ 16	~ 11.5	~ 11.5
C_{SO} (kg C/m ³)	3.6	5.4	3.6	3.6
ϕ_g (l/h)	15-20	20-30	15-20	15-20
pH	4.5-5.5	4.5-5.5	4.5-5.5	4.5-5.5
% acidification	95 $\pm$ 6	95 $\pm$ 6	100 $\pm$ 3	95 $\pm$ 8
% acetate	42 $\pm$ 7	44 $\pm$ 5	50 $\pm$ 2	55 $\pm$ 4
% propionate	16 $\pm$ 2	6 $\pm$ 1	18 $\pm$ 2	23 $\pm$ 4
% butyrate	38 $\pm$ 6	28 $\pm$ 3	22 $\pm$ 2	15 $\pm$ 3
% valerate	2 $\pm$ 4	-	10 $\pm$ 1	7 $\pm$ 1
% caproate	2 $\pm$ 2	22 $\pm$ 3	-	-

TABLE I-3.

The dimensions of the laboratory reactors used.

During the kinetic experiments, the reactor for the methane forming step was continuously stirred : 60 rpm, $d(\text{stirrer}) = 0.26$ m.

first reactor :

second reactor :

acid formation

methane formation

	carbon-balance sludge distribution	kinetics/RTD/check of the substrate model
$V_R = 60.0$ l	$V_R = 63.5$ l	$V_R = 51.9$ l
$V_{st} = 11.5$ l	$V_{st} = 11.5$ l	$V_{st} = 11.5$ l
$d_R = 0.292$ m	$d_R = 0.292$ m	$d_R = 0.292$ m
$h_R = 0.98$ m	$h_R = 1.04$ m	$h_R = 0.87$ m
$h_{st} = 0.26$ m	$h_{st} = 0.26$ m	$h_{st} = 0.26$ m
$h_b = 0.3-0.6$ m

Fig. I-11. Laboratory reactor for the conversion of a totally acidified sugar substrate into methane and carbon dioxide.

TABLE I-4.

Process conditions for the experiments with the methane forming laboratory reactor.

general : pH = 6.8 - 7.4

temperature = 30 - 35°C

VSS = 75 % of SS

1 g VSS = 0.5 g C

gas : 77 % CH₄, 18 % CO₂, 4 % H₂O, 2 % N₂

experiments on the carbon-balance and sludge distribution :

X = 0.5 - 0.85 kg C

$C_{sm,o} = (0.105 \pm 0.027)C_{so}$

kinetic experiments :

X = 0.03 - 0.125 kg C

$C_{sm,o} = (0.097 \pm 0.027)C_{so}$

$C_{so} = 0.36 - 0.9 \text{ kg C/m}^3$

$\tau = 4.3 - 6.5 \text{ h}$

experiments on the fluid flow pattern : see Chapter III.

I.4b Experiments with the pilot-plants

Of the experiments with the 6 m³ pilot-plant (see Fig. I-12) most results have already been presented (38). It was found that the process as performed in the upflow reactor is a reliable method to purify wastewater containing fatty

acids, and that very high loadings (up to 32 kg COD/m³ d) could be converted effectively. To study the fluid flow pattern in this reactor, one residence time distribution experiment was performed that will be discussed in Chapter III.

The experiments with the 30 m³ reactor that will be described in the following chapters concern :

1. the fluid flow pattern under various process conditions;
2. the distribution of sludge over the reactor in relation to the hydraulic loading, the volume of gas produced per unit of time and the properties of the sludge;

Fig. I-12. 6 m³ pilot-plant.

TABLE I-5.

Dimensions of the 30 m³ pilot-plant, locations of the sample points.

	August 1976 - March 1977					March 1977 - September 1977							
V_R (m ³)	30.2					33.2							
V_{st} (m ³)	1.5					5.6							
d_R (m)	2.6					2.6							
h_R (m)	6.0					6.5							
h_{st} (m)	1.1					1.6							
sp	I-1	I-2	I-3	I-4	1	2	3	4	5	6	7	8	
h (m)	0.05	0.05	0.05	0.05	0.2	0.3	0.4	0.5	0.72	1.02	1.32	1.62	
d (sp-wall)	0.8	0.8	0.8	0.8	0.8	0.6	0.8	0.6	0.2	0.2	0.2	0.2	
sp	9	10	11	12	13	14	15	16	17	18	19	20	
h (m)	1.92	1.92	1.92	1.92	2.2	2.5	2.8	3.1	3.42	4.0	4.6	5.6	
d (sp-wall)	1.3	0	1.0	0.6	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2

Fig. I-13. 30 m³ pilot-plant as used from August 1976 until February 1977, top and bottom of the reactor.

Fig. I-14. 30 m³ pilot-plant as used from March 1977 until September 1977, top of the reactor.

3. the performance of the settler in relation to the properties of the sludge, the linear liquid velocity and the flow rate of the gas.

The differences of the two settlers which are used, can be seen in the Figs. I-13 and I-14. Fig. I-13 shows the reactor with the first settler as it was used from August 1976 until February 1977; and Fig. I-14 the reactor with the second settler, used from March 1977 until December 1977. In Table I-5 the most important dimensions of the reactor are presented, together with the locations of the sample points. Figure I-15 is the flow diagram of the installation used. The possibility of supplying nutrients and soda (as is indicated in this diagram) was only used during the intercampaign. During the campaign, the amount of nutrients present in the wastewater was usually sufficient and the pH was in general acceptable (6 - 6.5). In Table I-6 examples are given of the composition of the influent during the campaign in 1976 and the intercampaign in 1977.

Table I-7 gives a survey of the process conditions during the whole experimental period. These data have been extracted from the description of the process given by C.S.M. (39). Further details of the process conditions and the results of the measurements of the process operation can also be found in this reference.

Fig. I-15. Flow sheet of the pilot-plant system.

TABLE I-6

Composition of the influent of the 30 m³ pilot-plant.

The campaign wastewater contained 0.01 - 0.06 kg/m³ N(NH₃) and (5-10)10⁻³ kg/m³ P(PO₄³⁻).

The intercampaign wastewater contained 0.02 - 0.1 kg/m³ N(NH₃) and ~3 10⁻³ kg/m³ P(PO₄³⁻). During this period, part of the effluent was recirculated.

	Influent during campaign 1976	Influent during intercampaign 1977
total-COD (kg/m ³)	2.0-4.0	2.0-6.0
centrifugate-COD (kg/m ³)	1.3-2.6	2.0-6.0
fatty acids-COD (kg/m ³)	1.0-2.0	1.0-4.0
BOD ₅ (kg/m ³)	1.2-2.4	-
C _{mo} (kg SS/m ³)	0.2-0.85	0.1-0.6
Date	Oct. 15 1976	April 04 1977
C _s (kg C/m ³)	0.532	1.17
C _{ST} "	-	2.10
[Ac ⁻] "	0.20	0.30
[Pr ⁻] "	0.18	0.065
[But ⁻] "	0.03	0.12

TABLE I-7

Process conditions during the experiments in the 30 m³ pilot-plant.

Period 1 : Mid-August - September 1976 (intercampaign).
 Period 2 : October - Mid-November 1976 (campaign).
 Period 3 : Mid-November - December 1976 (campaign).
 Period 4 : March - April 1977 (intercampaign).

	Period 1	Period 2	Period 3	Period 4
X (kg SS)	490	420	630	1100
h _d (m)	1.2	1.0	1.2	2.2
τ (h)	2.3-7.5	3.3-8.1	8.0-30	5.0-20
Ø _s ^{'''} (kg C/m ³ d)	0.8-4.1	2.0-6.6	0.8-3.0	2.0-8.0
% conversion (TOC)	~75	~63	~75	~93
pH	6.8-7.7	7.1-7.8	~7.2	6.7-7.2
temperature (°C)	20-32	24-37	~33	28-31

I.4c Experiments with the full-scale reactor

The 200 m³ reactor is presented schematically in the Figs. I-16, 17 and 18, giving the cross-section and front, the top and the bottom of the reactor respectively. In these drawings the places of the sample points are also indicated. The most important reactor data are given in Table I-8.

From the flow diagram, presented in Fig. I-19 it can be seen that the reactor was part of the wastewater circuit of the sugar factory. About one quarter of the effluent of the factory could be treated. Shortly after the start-up of the reactor, hydrogen sulphide was produced by reduction of sulphate that was temporarily present in the waste-water circuit. The problem of the bad smell of this gas was solved by supplying ferric chloride that with hydrogen sulphide gives the almost insoluble ferrous sulphide. After some weeks, the concentration of sulphate in the wastewater circuit and in the influent, and the concentration of sulphide in the effluent were negligible again. To prevent formation of foam at the top of the reactor and in the gas outlet, about 1 kg/d of an antifoam agent had to be added during the whole experimental period with this reactor. In Table I-9 examples have been given of the composition of the influent during the starting-up period of the reactor (about three weeks), and during a relatively stationary period of one week at the end of November 1977. A survey of the process conditions and the influent composition is given in the Tables I-10 and I-11 (again taken from the work presented by C.S.M.) (39).

Fig. I-16. Outline of the cross-section (A-A) and front of the 200 m³ full-scale reactor.

Fig. I-17. Outline of the top of the 200 m³ full-scale reactor.

TABLE I-8

Dimensions of the 200 m³ reactor, locations of the sample points.

$V_R = 204 \text{ m}^3$ $d_R = 7.5 \text{ m}$ $h_{st} = 1.9 \text{ m}$ $l_{st} = 7.5 \text{ m}$ $V_{st} = 53.5 \text{ m}^3$ $h_R = 4.5 \text{ m}$ $b_{st} = 3.5 \text{ m}$													
sp	1	2	3	4	5	6	7	8	9	12	13	18	
h (m)	0.8	0.8	1.0	0.3	1.0	1.0	0.3	1.0	0.8	1.5	1.5	1.5	
sp		S1	S2	S3	S4	S5	S6	S7	S8	S9	S10	S11	S12
depth (m)		2.0	0.9	1.7	1.1	0.8	0.4	0.8	0.6	0.9	1.7	0.8	0.6

The experiments with this reactor, to be presented in the Chapters, III, IV and V, were performed to study :

The experiments to study the fluid flow pattern and the distribution of sludge, were performed simultaneously. Because this reactor was part of the wastewater

Fig. I-19. Flow sheet of the full-scale system.

circuit of the sugar factory, it was of prime importance that the purification efficiency was good : to attain optimum operation, the process conditions were always adapted to the situation in the wastewater circuit. As a consequence, the conditions could not be chosen freely for the experiments. Fortunately, the organic loading and the gas production changed during the campaign in such a way, that their influences could after all, still be examined. The properties of the sludge particles as such did not change significantly, but because mud had accumulated in the reactor, the properties of the mixture of sludge and mud changed sharply. This also influenced the efficiency of the settlers used. Until mid-November both settlers of the reactor were operated with circulation flow (see Fig. I-5a).

From mid-November until the end of the campaign, in one settler the influent pipes for the circulation flow were blocked, so that in this settler the fluid flow pattern was changed. Here, the liquid flowed mainly upwards; while in the other, the circulation flow was still maintained.

TABLE I-9

Examples of the composition of the influent of the 200 m³ reactor.

	October 24 1977	November 30 1977
Total-COD (kg/m ³)	3.40	5.60
Centrifugate-COD (kg/m ³)	2.92	5.0
C _{so} (kg C/m ³)	0.90	1.42
[Ac ⁻] (kg C/m ³)	0.65	0.63
[Pr ⁻] "	0.34	0.60
[But ⁻] "	0.025	0.06
[Val ⁻] "	-	0.005
N(NH ₃) (kg/m ³)	0.023	0.045
P(PO ₄ ³⁻) (kg/m ³)	0.006	-
pH	7.3	7.0

TABLE I-10.

Composition of the influent of the 200 m³ reactor during the experimental period.

The influent contained 0.011-0.051 kg/m³ N(NH₃), the effluent 0.018 - 0.063 kg/m³ N(NH₃) and about 5 · 10⁻³ kg/m³ P(PO₄³⁻).

Period 1 : Start-up period, October 1977.

Period 2 : "Stationary" process operation, from November 22 1977 till December 2, 1977.

	Period 1	Period 2
Total COD (kg/m ³)	0.73-3.45	4.23-5.6
Centrifugate-COD (kg/m ³)	0.57-2.89	3.41-4.98
C _{so} (kg C/m ³)	0.4 - 0.9	0.93-1.46
C _{mo} (kg SS/m ³)	0.2 - 1.4	0.65
C _{fa,o} (kg C/m ³)	0.06-1.0	0.84-1.39
% acetate	54-77	48+3
% propionate	22-42	46+6
% butyrate	0-4	5+2
% valerate	-	-

TABLE I-11.

Process conditions during the experiments in the 200 m³ reactor.

	Period 1	Period 2
X (kg SS)	1800-5690	6000-6500
X (kg VSS)	1320-2330	2090-2500
h_b (m)	0.4	0.4
τ (h)	7.7-10	5-10
ϕ_s'' (kg C/m ³ d)	1.3-2.4	3.3-4.2
% purification	65-88	90-95
pH	7.2-7.6	7.0
temperature (°C)	30-35	30-33

I.5 Discussion

The data presented in this chapter depict the development of the process for anaerobic treatment of wastewater in the upflow reactor during the time interval 1974 - 1977. It appears that as a result of the adaptation of the sludge, especially at more or less constant process conditions, the process performance strongly improved, so that high organic loadings could be applied and a good purification efficiency was obtained. At the end of this period it was even possible to operate the 200 m³ reactor with high loadings, at a time when this reactor still contained a small amount of sludge and too much mud. Because these good results were obtained with wastewater that was partly soured during one period (intercampaign) or that was totally soured during the other period (campaign), this process has many potential applications.

The purification efficiencies obtained and organic loadings applied, show that the maxima of the potentialities of the reactors had not yet been reached. With the scale-up of the 6 m³ pilot-plant to the 30 m³ pilot-plant, the ratio of the height to the diameter of the reactor was kept roughly constant (~ 2) (this influenced the bed height). At scale-up of the 30 m³ pilot-plant to the 200 m³ technical scale reactor, the ratio of the height to the diameter was changed to 0.6, which meant that the diameter of the reactor was disproportionately scaled-up. The technological consequences of scale-up will be discussed in Chapter III (the fluid flow pattern) and in Chapter IV (the sludge behaviour).

With respect to the work performed after the above mentioned period of development, it can be reported that, based on the work described here, the C.S.M. has constructed a 800 m³ full-scale reactor. This reactor has been tested during the campaign of 1978 at the sugar factory in Halfweg. In this reactor all the wastewater of the factory had to be treated. Because of the short period during which the reactor had operated and the small amount of sludge, of course incomplete results were obtained, nevertheless, these were promising.

CHAPTER II

KINETICS

II.1 Introduction

For a description of the methane forming process, data are necessary, concerning the conversion rate of the substrate and the formation rate of the products. If it is known how these conversion and formation rates are influenced by the process variables, and what the optimum values of the process conditions are, the process can be optimised. Variables that are of importance for this process are the concentration of the substrate in the influent and in the reactor, the liquid residence time, the concentration of sludge in the reactor, and the sludge retention time. Process conditions of importance are the composition of the substrate, the pH, and the temperature.

Because the sludge consists of many different species of micro-organisms, the model description of the kinetics should be developed for an unstructured, non-segregated biomass (40). The model should be deterministic, and would only have any exact predictive value if it were to be applied for mixed cultures and for process conditions that correspond fully with those used for the development of the model. Our purpose with the kinetic model was to use it as support for the designing and scale-up of reactors for the methane forming step, requiring it to be as simple as is compatible with the reliability. Moreover, it should be rather easy to find the values of the constants of the model, for any set of process conditions that would be applied. With regard to the derivation of the model, literature concerning the kinetics of continuous cultures was used as a starting point (53-56).

From the literature concerning the kinetics of the anaerobic process for wastewater treatment (31, 41-45, 57-65), most of the data are inconsistent and *fair* therefore, seem to be unreliable. For this reason it is necessary to perform experiments that provide more reliable data.

The lack of reliable data is due to the specific character of the anaerobic process, giving the following experimental difficulties :

- Introductory experiments in this laboratory demonstrated that data from batch experiments (31, 41, 44) cannot be used to predict the operation of continuous processes such as the one to be described here, so that experiments on the kinetics had to be performed in a continuously operated reactor.
- The growth rate of anaerobic sludge is very small; this requires a long sludge retention time and long adaptation times when the composition of the substrate has been changed.

- The use of mixed cultures and the fact that the conversion of substrate is a succession of processes often involving several kinds of bacteria (46-51), are reasons why the systems for which the parameters and constants have to be measured, often differ widely; moreover, they are in many cases inexactly described, so that the results have a poor reproducibility.
- The possibility that mass transport of substrate from the bulk of the solution to and into the bacteria can be a process limitation (52), has to be recognized.
- The difficulty of maintaining anaerobic conditions.

The kinetics can be based upon several types of mass-balances. Because of the automatic analytical techniques available for the measurement of TC, TOC, CHN, the composition of the gas, etc., the carbon-balance is preferable, as both the conversion of substrate and the formation of products can be included. Furthermore, with this balance, no confusion is possible as no special conversion factors relating COD, BOD, VSS, DNA, etc., have to be used.

II.2 Models for the kinetics

Since the purpose of a biological process for wastewater treatment is to convert water pollutants effectively into non-pollutants, the model of the kinetics must have a two-fold function : 1) it has to give a reliable description of the conversion of the substrate; 2) it has to be informative about the formation rates of the products. Of the unstructured models used for the description of microbiological processes, the model of Monod (53) has most often been applied. In this model, the specific growth rate of the micro-organisms (μ) is described as a function of the concentration of the substrate that limits the growth of the bacteria. It is assumed that this growth is proportional to the conversion of the substrate, and that the Michaelis-Menten equation can be used with the restriction that the so-called saturation constant K_s has no physical meaning in the model of Monod :

$$\mu = \mu_{\max} \frac{C_s}{K_s + C_s} \quad \text{II-1}$$

and

$$dX = -Y dS \quad \text{II-2}$$

Many biological processes showed large deviations from this model at low growth rates. Therefore, the model has been extended by assuming that the yield factor (Y) is a variable instead of a constant. This has been done in two ways :

- a) by including endogeneous respiration in the model (Herbert; 54); or
- b) by assuming that part of the substrate is converted for cell maintenance (Pirt; 55, 56).

a) The endogeneous respiration can be fitted in Monod's model by introducing a negative term in the equation for the growth of the sludge :

$$r_m = (\mu - k_d) C_m = (\mu_{\max} \frac{C_s}{K_s + C_s} - k_d) C_m \quad \text{II-3}$$

This approach has often been used for kinetic models of anaerobic treatment of wastewater (57 - 61).

Sometimes in this equation, instead of Monod's term for the growth of sludge, an analogous term for the conversion of substrate has been used; it is assumed that $\mu/q = a$ is constant :

$$r_m = (a q_{\max} \frac{C_s}{K_s + C_s} - k_d) C_m = a r_s - k_d C_m \quad \text{II-4}$$

In this equation r_s is the total substrate conversion. The equation can be re-written as :

$$r_s = \frac{r_m}{a} + \frac{k_d}{a} C_m \quad \text{II-5}$$

The constant k_d in these equations is the so-called micro-organism decay coefficient (62). The equations II-3, II-4 and II-5 fit the experiments well for small growth rates and have thus been used for the description of the kinetics of the anaerobic process, because anaerobic bacteria grow slowly as they have little energy for the formation of new cell material (63 - 68).

Although the above models usually fit the data well, it must be noted that in fact the equations do not give a sound description for anaerobic processes. In the mass-balance of the sludge material, a negative term has been introduced that is interpreted as a decay coefficient, while it is well known that anaerobic bacteria do not decay easily, unless poisonous substances are present. Moreover, one of the most important features of the process : the production of methane, does not show in the balances. The discrepancy between model and fact stems from the fact that the k_d factor was introduced for the kinetics of low loaded aerobic cultures. In such systems the decay coefficient may indeed be important. However, for anaerobic systems this introduction has no sense. It must be a coincidence that the model fits well with the data provided by experiments with anaerobic cultures (16).

b) In the second modification of Monod's model, Pirt separated the total substrate conversion into a term for growth and a term for energy supply (maintenance) :

$$r_s = r_s(\text{growth}) + r_s(\text{maintenance}) \quad \text{II-6}$$

In anaerobic treatment of wastewater, the production of methane will be one of the results of conversion of substrate for energy supply. For stationary situations there may be a constant ratio of the total conversion of substrate and the growth of sludge, which means that Eq. II-2 may be used. In addition, it may be assumed that the ratio between the amount of substrate used for growth, and the amount of newly formed sludge (the growth itself), is constant (Y^*), and that the amount of substrate converted for bacterial energy supply is proportional to the amount of sludge present. Then, Eq. II-6 can be written as:

$$r_s = -\frac{r_m}{Y} = -\frac{r_m}{Y^*} - m C_m \quad \text{II-7}$$

Because this equation is identical with Eq. II-5 (Y^* and m are equal to a and k_d/a in that equation), it can now be understood why the first modification of Monod's model could be used for the anaerobic process although its starting point is disputable : it appears that mathematically no difference exists with the more appropriate Pirt-equation (16). Because in this last case it holds that $r_m = \mu C_m$, it can be derived that

$$Y = \frac{\mu}{m + \mu/Y^*} \quad \text{II-8}$$

In the literature this approach has been worked out for the anaerobic process of wastewater treatment (69), and used to predict the concentrations of substrate and products. For this purpose, the kinetic constants, such as the maximum specific growth rate, the saturation constant, the minimum generation time, and the yield factors for biomass and other products were determined. If the carbon-balance is used, both modifications of the model of Monod can be simplified. The yield factors Y^* and a (the ratio between the amount of substrate used for growth, and the growth itself), are equal to unity. Eq. II-8 can be changed into :

$$Y = \frac{\mu}{m + \mu} = \frac{\mu}{q} \quad \text{II-9}$$

In the study described here, the second modification will be used as a starting point. In the next section, the carbon-balance, needed for the evaluation

of the kinetics will be developed in further detail.

II.3 The carbon-balance

II.3a Theory

The carbon-balance can be specified if the amounts of soluble organic and inorganic carbon in the influent and effluent are measured with TOC and TC analyses; the composition of the gas is determined; the amount of sludge in the influent, effluent, and the reactor are measured; and the percentage of carbon in the sludge has been ascertained by a CHN-analysis. The flow rates of the influent, effluent and the gas must be known accurately. Finally, the concentrations of fatty acids in the influent and effluent have to be known. At the experiments on the carbon-balance, performed in this laboratory, a difference was found between $C_{s,e}$ and $C_{fa,e}$ (the TOC and the concentration of fatty acids in the effluent (both in kg C/m³)). These unidentified organic products will have been synthesised by the anaerobic bacteria, and therefore it is appropriate to take them into account in the carbon-balance. It is unknown whether the formation of these (soluble) products is exothermic or endothermic; therefore, the term describing their presence in the carbon-balance (a) has to be distinguished from the term m that describes the formation of gas and bicarbonate (the reaction providing the maintenance energy of the bacteria), and also from μ that gives the amount of substrate used for new cells. The balance can be written in the form :

$$q = m + \mu + a \quad \text{II-10}$$

For a continuously, stationary operating methane forming reactor, where the influent is the effluent of the acid forming reactor, the various terms of the carbon-balance can be determined as follows :

Specific substrate conversion rate :

$$q = F(C_{so} + C_{sm,o} - C_{fa,e})/X \quad \text{II-11}$$

Specific production rate of gas and bicarbonate :

$$m = \phi_g C_g/X + F((C_{so} - C_{s,e}) - (C_{ST,o} - C_{ST,e}))/X \quad \text{II-12}$$

Specific growth rate of sludge :

$$\mu = F(C_{ST,o} + C_{sm,o} - C_{ST,e})/X - \phi_g C_g/X \quad \text{II-13}$$

Specific production rate of unidentified products :

$$a = F(C_{s,e} - C_{fa,e})/X \quad \text{II-14}$$

Because the unidentified products are introduced in the mass-balance, Eq.II-9 for the yield-factor for sludge has to be adapted, and gives :

$$Y = \frac{\mu}{m + \mu + a} = \frac{\mu}{q} \quad \text{II-15}$$

II.3b Experiments for the determination of the carbon-balance

The system used for the experiments (as presented in the Figs. I-10 and I-11) consists of a stationary working laboratory reactor for the acid formation; an intermediate, completely stirred tank with such provisions that the methane forming reactor could be operated independently from the first reactor; and the methane forming reactor itself. In the Tables I-1 and I-2 the compositions of the concentrated food, nutrients and buffer solutions, the effluent of the acid forming reactor, and the influent of the methane forming reactor have been presented.

The duration of the experiments for the determination of the carbon-balance was several hours, so that only one experiment could be performed per day. After finishing the experiment, new conditions were established, and during the night the system could adapt itself to these conditions. After this, the reactor was in a stationary situation, and ready for a next experiment. The experiments involved continuous sampling of the influent and the effluent, and continuous measurement of the volume of gas produced per unit of time. The fluid flow was determined before and after each experiment, the composition

Fig. II-1. Flow sheet of the analytical procedure.

TABLE II-1

Results of the measurements to establish the carbon-balance in the methane forming reactor.

	Period 1	Period 2	Period 3	Period 4	Period 5	Period 6
C_{so} (centrifugate) (kg C/m^3)	0.73 ± 0.1	0.61 ± 0.04	1.03 ± 0.11	1.48 ± 0.1	1.92 ± 0.2	2.94 ± 0.19
τ (h)	6.2	6.2	6.5	4.6	6.1	7.5
reactor loading, based on centrifuged samples ($\text{kg C/m}^3 \text{d}$)	2.8 ± 0.5	2.4 ± 0.2	3.8 ± 0.5	7.7 ± 0.6	7.6 ± 0.9	9.4 ± 0.7
% conversion (fa)	100	100	100	99	98	97
% conversion (TOC)	86 ± 5	91 ± 5	87 ± 3	86 ± 2	89 ± 3	91 ± 1
ϕ_g (l/d)	197 ± 4	177 ± 6	275 ± 11	525 ± 13	575 ± 17	684 ± 17

TABLE II-2

Model parameters, derived from the data given in Table II-1.

	$\Delta(\%)$	Period 1	Period 2	Period 3	Period 4	Period 5	Period 6
q (d^{-1})	23	0.32	0.24	0.30	0.56	0.51	0.70
m "	36	0.26	0.20	0.26	0.49	0.42	0.64
a "	37	0.04	0.02	0.04	0.08	0.06	0.05
μ "	100	0.012	0.016	0.002	-0.005	0.036	0.017
Y	77	0.038	0.067	0.005	-0.009	0.070	0.024

$$k_{sg} = (1.22 \pm 0.01 \text{ m}^3 \text{CH}_4 / \text{kg TOC}; \bar{Y} = 0.03 \pm 0.023; \bar{Y}_{(\text{VSS-balance})} = 0.077 \pm 0.023$$

of the gas was measured several times during the series of experiments. Before and after this series, the amount of sludge in the reactor was measured by sampling the reactor that was completely mixed at that moment. The following analyses were made of the influent, effluent and reactor-samples :

TC ($C_{ST,o}$, $C_{ST,e}$), TOC (C_{so} , $C_{s,e}$), fatty acids ($C_{fa,o}$, $C_{fa,e}$), VSS and CHN ($C_{sm,o}$, $C_{sm,e}$, X). A flow sheet of the analytical procedure is presented in Fig. II-1.

The most important experimental data are presented in Table II-1, and the values of the parameters of the model in Table II-2. For the determination of the values of the yield factor (which are also included in this table), Eq. II-15 was used. It is possible to calculate the mean value of this factor by means of the sludge-balance : during the total experimental period 8.675 kg TOC was converted and 0.67 kg of sludge (expressed in kg Carbon) had grown in the reactor (this latter value is the difference between the sludge contents before and after the experiments, plus the cumulative amount that washed out during the series of experiments). Therefore, the mean yield factor $\bar{Y} = 0.077 \pm 0.023$.

Fig. II-2. The production rate of methane as a function of the conversion rate of total organic carbon, measured for the laboratory reactor.

If the volume of methane, produced per day, is plotted against the amount of substrate converted per day (see Fig. II-2) a straight line is obtained. From the slope of the line it has been found that $k_{sg} = 1.22 \pm 0.01 \text{ m}^3 \text{ CH}_4/\text{kg TOC}$.

II.3c Remarks about the carbon-balance

The accuracy of the carbon-balance depends on the accuracy of each measurement. With the methods used, the error in the determination of the influent TC and TOC, and of the effluent TC were about $\pm 5\%$, for the effluent TOC this was $+15\%$ and for the effluent VSS even as large as $\pm 50\%$. The error in the fatty acids analysis was $\pm 12\%$ and in the CHN-analysis $\pm 2\%$. Because of these values each experiment was performed three times and sampled in triplicate; each of the nine samples was analysed in duplicate.

For each experimental period, the sum of the terms m , a and μ that are given in Table II-2, is equal to the value of q ; this means that the balance as formulated in the Eqns. II-10 to II-14 inclusive, is correct. The quantitative values of the terms of the carbon-balance show that the term indicating the presence of unidentified products is necessary in a complete balance. The differences between the percentages of conversion based on fatty acids and on TOC (see Table II-1), are caused by the presence of these products. The 90 % reliability intervals of the terms, given in the first column of Table II-2, show that this kind of experiment cannot be used to find reliable values for the growth of the bacteria; but for q , m and a , this method is appropriate. Fig. II-3 is presented as an example of the product streams in anaerobic treatment of wastewater. The quantitative values in this figure have been found at period 3.

Fig. II-3. The carbon-balance, determined for the methane forming reactor, period 3.

The two values of the yield factor found independently, indicate that the real value will be about 0.05 kg C(sludge)/kg C(fatty acids). This agrees with the average value of 0.044 kg biomass/kg COD given in the literature for the conversion of fatty acids (64).

From the conversions presented in Table II-1 it can be concluded that the two-reactor system operated properly. In the methane forming reactor (that is process limiting), a loading of 9.4 kg C/m³d is converted with a good efficiency (97 %, period 6). The production of gas during this experiment, expressed in m³ of gas per m³ reactor volume per day, was 10.6 d⁻¹ (about 80 % CH₄). As can be seen in Table I-2, the acid formation could easily be maintained stationary and it seems even possible to load this reactor more than the highest loading applied during this series of experiments.

II.4 Kinetic model used

II.4a Theory

Of the parameters appearing in the mass-balance (Eq. II-10), it is of importance to know the kinetics. The parameters q and a (the specific substrate conversion rate, and the specific formation rate of unidentified products), give information about the conversion capacities of the process with respect to fatty acids and total organic carbon. The parameter m consists of three terms viz. the specific production rates of methane, carbon dioxide and bicarbonate. With regard to the retention of sludge, the specific growth rate (μ) is of importance. With the apparatus and methods used, the kinetic constants for the growth of sludge could not be determined with some accuracy. The reason was that the washout of sludge, being small because of the use of a settler, could only be measured with a large error.

For stationary conditions the mass-balance for the conversion of substrate is as follows :

$$F(C_{so} + C_{sm,o} - C_{fa}) - qC_m V_R = 0 \quad \text{II-16}$$

For the specific substrate conversion rate, the Monod equation has been used, in the form :

$$q = q_{\max,fa} \frac{C_{fa}}{K_{fa} + C_{fa}} \quad \text{II-17}$$

By measuring C_{fa} and q at different organic loadings, the constants $q_{\max,fa}$ and K_{fa} can be determined.

Although the kinetics of the gas and bicarbonate can be measured, we were only

interested in the production of methane and carbon dioxide. The mass-balance for the gas is as follows :

$$\phi_g = \frac{r_g V}{C_c} \quad \text{II-18}$$

This balance can be rewritten for methane by the introduction of the fraction of methane in the gas :

$$\phi_{g,me} = f_{me} \phi_g = \frac{r_{me} V}{C_c} \quad \text{II-19}$$

For the methane production rate an equation analogous to that of Monod and to Eq. II-17 was chosen :

$$r_{me} = \frac{r_{me, \max, fa} C_m}{K_{me} + C_{fa}} \quad \text{II-20}$$

By measuring C_{fa} and r_{me} at different organic loadings, the constants $r_{me, \max, fa}$ and K_{me} can be determined.

For the formation of unidentified products, it is assumed that already formed products are not converted into gas and that the formation of the products is proportional to the difference of the fatty acids concentrations in the influent and the effluent. The mass-balance for the unidentified products is as follows :

$$FC_{ap} = aC_m V R \quad \text{II-21}$$

in which $C_{ap} = C_s - C_{fa}$ and $aC_m = r_{ap}$. According to the assumption mentioned, the rate of formation of unidentified products is :

$$r_{ap} = k_{ap} (C_{so} - C_{fa}) \quad \text{II-22}$$

By measuring C_{so} , C_s and C_{fa} at different loadings, the rate constant k_{ap} can be determined.

II.4b Experiments and results

The experiments were performed in the methane forming reactor of laboratory scale. The only difference with the carbon-balance experiments was that the reactor was continuously mixed to obtain a uniform substrate concentration. The experimental procedure was identical : one experiment per day. The duration of these experiments was restricted to several hours because then it could

be assumed that the amount of sludge was constant during the experiment. The results are presented in the Figures II-4, 5 and 6.

Fig. II-4. The inverse of the specific substrate conversion rate as a function of the inverse of the fatty acids concentration in the reactor. By extrapolation to $1/C_{fa} = 0$, the value of $q_{max,fa}$ is found, the slope of the line gives $K_{fa}/q_{max,fa}$.

Fig. II-5. The inverse of the methane production rate as a function of the inverse of the fatty acids concentration. Extrapolation to $1/C_{fa} = 0$ gives the value of me_{max} , the slope of the line gives K_{me}/me_{max} .

In Table II-3 the results of the regression analyses of the data in the figures are presented together with the values of the kinetic constants obtained by these analyses. The intervals in the table indicate 90 % reliability.

Fig. II-6. The formation rate of unidentified products as a function of the difference in influent and effluent fatty acids concentration in the laboratory reactor. The slope of the line gives the rate constant k_{ap} .

TABLE II-3.

Statistical constants and the results of the experiments on the kinetics.

substrate	methane	unidentified products
$n = 10$	$n = 11$	$n = 13$
$r = 0.64$	$r = 0.66$	$r = 0.87$
$m = 8.13 \cdot 10^{-2} \text{ kg m}^{-3} \text{ s}$	$m = 1.64 \cdot 10^3 \text{ kg m}^{-3} \text{ s}$	$m = 5.47 \cdot 10^{-6} \text{ s}^{-1}$
$b = 2.12 \cdot 10^4 \text{ s}$	$b = 4.05 \cdot 10^4 \text{ s}$	$b = 3 \cdot 10^{-7} \text{ kg m}^{-3} \text{ s}^{-1}$
$\sigma_x = 5.06 \text{ m}^3 \text{ kg}^{-1}$	$\sigma_x = 4.93 \text{ m}^3 \text{ kg}^{-1}$	$\sigma_x = 0.22 \text{ kg m}^{-3}$
$q_{\max, fa} = (4.7 \pm 0.8) \cdot 10^{-5} \text{ s}^{-1}$	$m_{e, \max, fa} = (2.5 \pm 0.3) \cdot 10^{-5} \text{ s}^{-1}$	$k_{ap} = (5.5 \pm 0.5) \cdot 10^{-6} \text{ s}^{-1}$
$K_{fa} = (38 \pm 3) \cdot 10^{-3} \text{ kg/m}^3$	$K_{me} = (41 \pm 3) \cdot 10^{-3} \text{ kg/m}^3$	

II.5 Discussion and conclusions

The assumption behind the choice of short experimental periods was that in three to five hours no measurable growth of sludge would take place in the reactor. A simple calculation shows that even at the highest organic loadings, the growth of new sludge remained under 1 % of the amount of sludge present. Therefore, the total amount of sludge could safely be supposed constant during each experiment. Another assumption that needs explanation was made for the model that describes the formation of unidentified products. It was assumed

that no conversion of these products takes place. The tests in the laboratory reactor showed no conversion after 24 hours, but in a longer run (about 20 days) conversion was observed.

It has to be noted that the data show only a moderate correlation with the lines obtained by regression analysis, because most of the data points lay in the same area. In fact, better results should be obtained if the concentration range was extended to values below 0.05 kg C/m^3 , but at these low values process limitation by mass transport processes in the sludge particles may occur.

If the quantitative values of the saturation constants K_{fa} and K_{me} are compared, no significant difference can be observed. This indicates that these processes are kinetically coupled (70). In such a case, a constant can be determined that associates the rate of methane production with the rate of substrate conversion:

$$\alpha_{s,me} = \frac{m_{e,max,fa}}{q_{max,fa}} = 0.53 \pm 0.15$$

This kinetic coupling is confirmed by the straight line obtained after plotting the methane production against the TOC-conversion (see Fig. II-2).

For the kinetics describing the formation of unidentified products, an analogy with the model of Monod could not be used. It appeared that a linear relation between the formation of these products and the difference of the fatty acids concentrations in the influent and the effluent agreed well with the measured data. This implies that there is no association between the rate of substrate conversion and the rate of formation of the unidentified products. This could be an indication that these products are waste products of the bacteria.

If analogous to the methane production is assumed that the growth of sludge is associated with the substrate conversion, the yield factor for the sludge can be interpreted as an association constant. Because this factor could be estimated from the results of the carbon-balance experiments, $Y \approx 0.05$, the model gives some information that can be used to predict the growth of an-aerobic sludge: $\mu_{max,fa} = q_{max,fa} \cdot Y \approx 2.3 \cdot 10^{-6} \text{ s}^{-1}$. ($\approx 0.2 \text{ d}^{-1}$)

Since the results obtained are very system dependent, i.e. dependent on the process conditions applied, on the composition of the substrate, and on the mixed culture used (for instance the degree of enrichment obtained by an adaptation period); a comparison with the values of kinetic constants given in the literature is difficult. For example, the values of the constants found in this chapter are only valid for a substrate contained by complete acidification of a sugar solution and for sludge obtained after an adaptation period of more than a year; while in the literature, mostly the kinetics of the conversion of single fatty acids have been studied (64, 69, 71) and much shorter adaptation periods have been applied, resulting in a

lower degree of enrichment of the cultures. Moreover, most models used in the literature have primarily been developed to describe the growth of anaerobic sludge and are mostly based on the disputable introduction of a decay coefficient in the equation for the growth of sludge. Therefore, a comparison, although very interesting, cannot be very fruitful and will not be made. Instead, some general conclusions that can be drawn from the results, will be presented.

- The carbon-balance is an adequate means to quantify the substrate and product streams in anaerobic treatment of wastewater.
- In this process it has to be taken into account that about 10 % of the waste will be converted into unidentified organic products, presumably formed as waste products of the bacteria.
- Based on the carbon-balance and Pirt's modification of Monod's model, a kinetic model was developed that appeared to be useful in describing the conversion of the substrate (in this case a completely acidified sugar solution) and the production of methane : $q_{\max,fa} = 4.7 \cdot 10^{-5} \text{ s}^{-1}$, $K_{fa} = 38 \cdot 10^{-3} \text{ kg C/m}^3$; $m_{\max,fa} = 2.5 \cdot 10^{-5} \text{ s}^{-1}$, $K_{me} = 41 \cdot 10^{-3} \text{ kg C/m}^3$.
- The constant, associating the rate of methane production with the rate of substrate conversion, appeared to have the value $\alpha_{s,me} = 0.53 \text{ kg C/kg C}$, and this is in agreement with the experimentally found relation between the methane production and the TOC-conversion : $k_{sg} = 1.22 \text{ m}^3 \text{ CH}_4/\text{kg TOC}$ (it has to be taken into account that the value of α is based on the conversion of fatty acids).

CHAPTER III

FLUID FLOW PATTERN

III.1 Residence time distribution

For a good conversion of the organic biodegradable compounds in wastewater, the water has to come into close contact with the anaerobic sludge. The idea was that such contact is established in an upflow reactor containing a sludge bed and a sludge blanket, in which the gas produced mixes the contents. However, because this contact is dependent on the fluid flow pattern, that depends on the design of the reactor and the process conditions, it is necessary to know more about the way the fluid flows through the upflow reactor. To measure this, residence time distribution experiments can be performed.

The residence time of a fluid element in a vessel is defined as the time interval between the moments of entering and leaving. The distribution of the residence times of the fluid elements can be studied by means of the stimulus-response technique (72); the stimulus can be an amount of tracer that is injected into the influent of the vessel, the response is the concentration of tracer on a certain point, measured as a function of the time interval since the start of the injection.

If from the moment $t = 0$, the tracer is injected continuously into the influent, an up-step stimulus is obtained. The response to such a stimulus, measured in the effluent of the vessel, is the so-called F -curve. If initially, no tracer was present, this curve is simply obtained by plotting C/C_0 against $\theta = t/\tau$. The quantitative value of F at a moment θ after the start of the injection, represents the fraction of traced liquid elements in the effluent at that moment; or in other words, represents all the fluid elements in the effluent with a residence time between 0 and θ .

If at the moment $t = 0$, an amount of tracer is injected as a pulse into the influent, the so-called C -curve can be measured in the effluent of the vessel. The quantitative value of C at a moment θ after the injection, represents the fraction of traced fluid elements in the effluent at that moment, that is, it represents all the fluid elements in the effluent with residence time θ . The curves obtained from such experiments have to be evaluated, which can be done by different methods. A method that is very well suited for large systems with relatively long residence times, is evaluation by means of a so-called "combined model", that is a model description of the fluid flow pattern, obtained by combining regions of different flow regimes. The following can be

distinguished as flow regimes : perfect mixing, plug flow, and dead space. The flows that connect the regions are : cross flow, bypassing flow, and recirculation flow (73). For instance, the fluid flow pattern in a chemical reactor can often be described as a combination of perfect mixing, dead space, and bypassing flow (74).

In a "combined model", an up-step stimulus-response experiment can be simulated by assuming that at a certain moment $t = 0$, the concentration of a tracer in the model's influent, increases at once from zero to the value C_0 . By means of the mass-balance for the tracer over the total model, the simulated response or the F -curve of the model can be calculated. Mass-balances for the tracer over sub-units of the combined model (to obtain $C/C_0 - \theta$ curves for these sub-units) can be used as well. The F -curve and the other $C/C_0 - \theta$ curves, calculated for the model, can be compared with the F -curve and the other dimensionless concentration-time curves, measured in the effluent of the reactor and also in the reactor itself.

Because regions and flows can be distinguished whose physical meanings are closely related to reality, and because these regions and flows can easily be combined in appropriate models, this method of evaluation of the results of residence time distribution experiments is very useful for the characterisation of the fluid flow pattern in the upflow reactor for anaerobic treatment of wastewater.

III.2 Model

On the basis of observations and considerations already presented (see Chapter I), it is assumed that : a) in the bed-blanket-reactors used, a distinction can be made between the fluid flow patterns in the sludge bed, the sludge blanket and the settler (30); b) as flow regimes in the sludge bed, perfect mixing and dead space may be present; c) the sludge blanket can be described as one ideal mixer; and d) the fluid flow in the settler can be described as plug flow. Cross flows will occur between the bed and the blanket, and bypassing of the sludge bed is possible. Combination of all these sub-units gives the qualitative model description of the hydrodynamical situation in the upflow reactor as is shown in Fig. I-4. For the sake of brevity this will further be called "model".

If in this model an up-step stimulus-response experiment is simulated, the following mass-balances can be derived for the tracer in the sludge bed and in the sludge blanket :

$$V_b \frac{dC_b}{dt} = F_o C_o + F_{db} C_d - F_{bd} C_b \quad \text{III-1}$$

$$V_d \frac{dC_d}{dt} = F_k C_o + F_{bd} C_b - F_{db} C_d - F C_d \quad \text{III-2}$$

To solve these differential equations, the values of the independent parameters - the volumes V_b and V_d , and the flow rates F , F_k and F_{db} - must be known.

Furthermore, the following relations hold :

$$F_o = F - F_k \quad \text{III-3}$$

$$F_{bd} = F_o + F_{db} \quad \text{III-4}$$

$$V_R = V_b + V_d + V_{pf} + V_{ds} \quad \text{III-5}$$

The responses of the sludge bed and the sludge blanket in the model are obtained by plotting the calculated values of C_b/C_o and C_d/C_o against $\theta = t/\tau$, in which $\tau = V_R/F$. The dimensionless concentration-time curve for the tracer in the effluent (the F -curve) is obtained by shifting the $C_d/C_o - \theta$ curve horizontally over the reduced residence time of the plug flow region ($\theta_{pf} = V_{pf}/F\tau$).

The choice of the quantitative values of the independent parameters was limited because we intended the model to be in close agreement with the physical reality. Therefore, the values of V_b , V_d , and V_{pf} naturally had to represent the reactor volumes occupied by the bed, blanket and the settler. The value of F_k must lie between zero and F ; the backmix flow F_{db} is generated by the production of gas in the sludge bed, and should be chosen in relation to this gas flow.

The volume of the region of dead space in the reactor can be estimated by means of the measured F -curve if V_R , F , C_o and C are accurately known. Because

$$\int_0^{\infty} (1 - F(\theta)) d\theta = 1 \quad \text{III-6}$$

and this integral is equal to the area between the F -curve and the line $C/C_o = 1$ in the $C/C_o - \theta$ diagram, the dead space can be estimated by measuring this area. If this turns out to be smaller than unity, the reactor volume occupied by the flowing liquid must have been smaller than the real volume of the reactor. The difference between these volumes is the volume of the region of dead space.

The responses of the model are compared with the locally measured, dimensionless concentration-time curves obtained from up-step experiments. To find the

best approximation between the model and the experiment, the values of the independent parameters are varied, until (as the result of this trial-and-error method) the best values of the parameters are found.

III.3 Experimental methods

III.3a Residence time distribution experiments

The stimulus was obtained by pumping (from the moment $t = 0$) a concentrated LiCl solution continuously into the influent (see the Figs. I-10, 15 and 19). The distance between the point of injection in the influent, and the place where the influent enters the reactor was chosen such that the LiCl was distributed homogeneously in the influent as this entered the reactor. In the time interval of 2-3 liquid residence times ($2-3 \tau$) after the start of the injection, samples were taken at several places, the sampling time intervals being 0.1τ or longer. The influent flow rate, the flow rate of the concentrated LiCl solution and the volume of gas produced per unit of time were measured with intervals of about 0.25τ . The samples were filtered over cotton wool and analysed for Li^+ with a spectrophotometer.

In Table III-1 the general conditions of the RTD-experiments are presented

TABLE III-1.

Experimental conditions for the study of the fluid flow pattern.

aspect studied	injection	reactor volume	LiCl kg/m^3	injection flow rate (l/h)	injection time (h)
RTD	up-step	51.9 l	2.5	0.54	7.5
RTD	up-step	30 m^3	14	35 - 50	10 - 14
RTD	up-step	200 m^3	60 - 100	15	30 - 35
influent	pulse	30 m^3	100-133	1200	0.025
influent	pulse	200 m^3	100	375	0.067

together with those of the other (pulse) tracer experiments for the measurement of the distribution of the influent in the lower parts of the reactor. In the full-scale 200 m^3 reactor, the duration of each experiment (RTD) was 25-35 hours; during the night, an automatic sampling apparatus was used.

One of the experiments performed in the laboratory reactor (experiment 4a), was conducted in a different way than the others : the tracer was injected continuously into the sludge blanket in an area of good mixing. C_0 for this experiment, was calculated by the multiplication of the tracer concentration in the injection, with the ratio of the flow rates of the injection and the influent. To prevent density problems at the injection, the LiCl was - for this experiment - dissolved in a 1 : 1 methanol-water mixture. After about 3 τ , the sludge bed was sampled in two places. This procedure was followed to study the flow rate of the backmix stream from the blanket to the bed. During the other experiments in the laboratory reactor, the sludge bed was not sampled because this would have influenced the local fluid flow pattern too strongly.

III.3b Experiments on the influent distribution system

Additional to the continuous injection method, used to obtain an overall picture of the fluid flow in the reactor in dependence on the influent distribution, for the characterisation of the fluid flow pattern around the influent points, pulse tracer experiments are particularly useful. By a pulse stimulus, quickly changing tracer concentrations in the area around the influent point are obtained, and these responses will be strongly influenced by the local fluid flow pattern. The differences appear with local sampling of very short time intervals.

This method has been used for the characterisation of the fluid flow pattern around the different influent systems used in the 30 m³ pilot-plant and for the estimation of the area covered with one unidirectional influent point in the 200 m³ reactor. The continuous injection has been used to compare the fluid flow pattern in the 30 m³ pilot-plant at radial influent distribution, with that of a unidirectional influent distribution. In the 200 m³ reactor the fluid flow pattern is compared, initially at ten, then at four influent points for unidirectional influent distribution.

III.4 Results

In this section the results of fourteen residence time distribution experiments will be presented in such a way that they can be compared with one another, to study the influence of the process conditions and the dimensions of the reactors. First the results will be evaluated to obtain model descriptions valid for the hydrodynamic situation belonging to the process conditions and the reactor dimensions applied during the experiments. As described in section III-2, this will be done by trial-and-error, in which the model presented in

Fig. I-4, and the estimated values of the independent parameters of the model are used as starting points. It is necessary to follow this evaluation procedure, because the rough data do not give enough information for direct use in quantitative comparison.

When the most appropriate "combined model" for each set of dimensionless concentration-time curves has been found, quantitative comparison of the results is possible by comparing these models. The differences between the models are caused by the different process conditions and reactor dimensions applied, so that then quantitative information about the influences of these conditions and dimensions concerning the fluid flow pattern is obtained.

The results of the experiments on one scale are presented together with the combined models after evaluation of these experimental data, so that by comparison of these models, the influence of the process conditions can be studied for each scale separately. In the discussion, the models obtained for the different reactors are compared with one another, providing conclusions as to the influence of the reactor dimensions.

III.4a Experiments in the laboratory reactor

In the Figs. III-1 to III-5 inclusive, the results of the experiments are

Fig. III-1. Results of stimulus-response experiment 1 (laboratory reactor) and the calculated responses of model 1. (●) reactor; (○) effluent; (—) response V_d ; (---) response V_{pf} .

Fig. III-2. Results of stimulus-response experiment 2 (laboratory reactor), and the calculated responses of model 2. (●) reactor; (○) effluent; (—) V_d ; (---) V_{pf} .

Fig. III-3. Results of stimulus-response experiment 3 (laboratory reactor) and the calculated responses of model 3. (●) reactor; (o) effluent; (—) V_d ; (---) V_{pf} .

Fig. III-4. Results of stimulus-response experiment 4 (laboratory reactor) and the calculated responses of model 4. (●) reactor; (o) effluent; (—) V_d ; (---) V_{pf} .

Fig. III-4a. Results of stimulus-response experiment 4a (laboratory reactor) and the calculated responses of model 4a. (●) reactor; (o) effluent; (—) V_d ; (---) V_{pf} . At $\theta = 3$, SP-2 : $C/C_0 = 0.87$; SP-3: $C/C_0 = 0.27$.

Fig. III-5. Results of stimulus-response experiment 5 (laboratory reactor) and the calculated responses of model 5. (●) reactor; (o) effluent; (—) V_d ; (---) V_{pf} .

presented as six sets of dimensionless concentration-time relations. The models obtained from evaluation of these results are presented in Table III-2. The

TABLE III-2.

The models representing the fluid flow patterns in the laboratory reactor.

For comparison, the starting values of the independent parameters of the trial- and-error procedure are also presented.

results	experiment no :					
	1	2	3	4	4a	5
V_b (% of V_R)	18.9	7.4	35.5	34.5	34.5	30.0
V_d "	58.8	70.3	43.2	43.2	43.2	47.7
V_{pf} "	16.7	9.4	22.3	15.7	15.7	20.5
V_{ds} "	5.6	12.9	0	6.6	6.6	1.8
F_k (% of F)	60 $\pm$ 15	80 $\pm$ 10	20 $\pm$ 5	20 $\pm$ 5	20 $\pm$ 5	5 $\pm$ 5
F_{db} "	65	25	70	40	40	35

starting values :

V_b (% of V_R)	18.9	7.7	44.3	44.3	44.3	40.4
V_d "	59.0	70.2	33.6	33.6	33.6	37.5
V_{st} "	22.1	22.1	22.1	22.1	22.1	22.1
F_{db} (% of F)	65	25	70	40	-	35

data points are the actual results of the measurements; the lines are the responses of the models that fit these data points.

Because the backmix stream is a function of the local production rate of gas in the sludge bed, a separate experiment (4a) was performed to examine this relationship. By injection of tracer into the blanket, a concentration profile for the tracer in the reactor developed, that was dependent on the backmix flow rate, and therefore on this production rate of gas. Sampling in the sludge bed after 3τ (SP.3, see Fig. I-11) showed a local tracer concentration of $0.27 C_0$. By means of adapted mass-balances for the tracer over the bed and the blanket, in which the concentrated injection stream of tracer has been taken up, model responses could also be calculated for this experiment. The best approximation between the responses and the measured data, was obtained if a backmix stream $F_{db} \approx 0.4 F$ was chosen.

From the process conditions during this experiment ($\phi_g = 21.0 \cdot 10^{-3} \text{ m}^3/\text{h}$, $F = 45.6 \cdot 10^{-3} \text{ m}^3/\text{h}$, $F_k = 0.2 F$, hydrostatic pressure $\sim 0.4 \text{ m}$ water column)

$\phi_{gb} \approx 0.4 F$ could be calculated; it can therefore be concluded that the back-mix stream is roughly proportional to the volume of gas produced per unit of time in the sludge bed. This conclusion has been used to estimate the backmix flow rates in the models for the other experiments in the laboratory reactor. In Fig. III-15* the block diagram is presented that gives a generalised picture of all the models obtained for the laboratory experiments. This was possible

Fig. III-15. * Generalised block diagram of the models 1-5 (see also Table III-2).

because all the models consisted of the same sub-units and differed only in the values of the parameters to be filled in. The general block diagram appears to conform well to the model that was assumed in Chapter I. The only difference is the occurrence of dead space in the settler, instead of in the sludge bed. This effect must have been caused by the amount of sludge accumulated in the settler, because no dead space was detected after removing this sludge ($V_{ds} = 0$ in Model 3, exp. 3, see Table III-2), and the largest volume of dead space ($V_{ds} = 12.9\%$) was found just before this removal (exp. 2, Model 2). Furthermore, from Table III-2 it appears that the distribution of the volumes of the flow regions in the models conforms well to the distribution of the actual volumes of the bed, blanket, and the settler in the reactor. In the following paragraphs, the models will be compared with one another in relation to the different process conditions presented in Table III-3. It can be expected that the process conditions which influence directly the fluid flow pattern in the reactor, and whose influence should therefore be traced in the models, are : the volume or the height of the sludge bed (V_b or h_b); the influent flow rate (F); and the volume of gas per unit of time (ϕ_g).

* This figure number deviates from the usual sequence, because in this chapter, the numbers of the experiments, models and graphs had to be unified.

TABLE III-3.

Process conditions of the residence time distribution experiments
in the laboratory reactor.

	experiment no.					
	1	2	3	4	4a	5
V_b (m^3)	0.01	0.004	0.023	0.023	0.023	0.021
h_b (m)	0.18	0.09	0.37	0.37	0.37	0.34
q (d^{-1})	2.24	3.30	1.17	1.10	1.16	-
F ($10^{-3} m^3/h$)	23.9	23.9	27.7	45.6	45.6	17.9
τ (h)	2.17	2.17	1.87	1.14	1.14	2.90
C_{so} (kg C/ m^3)	0.82	1.24	0.94	0.54	0.54	-
$C_{s,e}$ (kg C/ m^3)	0.28	0.74	0.15	0.08	0.06	-
% conversion (TOC, centri- fuged samples)	66	40	84	86	88	-
ϕ_s'' (kg C/ $m^3 d$)	9.1	13.7	12.0	11.4	11.4	-
ϕ_g ($10^{-3} m^3/h$)	18.3	11.0	26.0	21.0	21.0	7.7
ϕ_g'' (h^{-1})	0.35	0.21	0.50	0.40	0.40	0.15
$C_{Li,o}$ ($10^{-3} kg/m^3$, calc.)	8.12	-	6.88	-	-	8.05
$C_{Li,o}$ ($10^{-3} kg/m^3$, meas.)	8.0	-	6.80	-	-	8.10

Comparing the models 1, 2, 3, and 5, it follows that the differences between the volumes of the sludge bed have indeed influenced the fluid flow pattern, because the volumes of the first ideal mixers (V_b) in the models, showed similar differences. These differences (between the bed volumes or bed heights) had another effect : the bypassing streams in the models 1, 2, 3, and 5 were also different. It can be concluded that in a low sludge bed large bypassing streams are present, while in a higher sludge bed smaller bypassing effects occur.

From the comparison of the models 1 and 2 it follows that the bypassing stream is not increased by a larger gas production. This is confirmed by comparison of model 3 with model 4. This comparison leads furthermore to the conclusion that the influent flow rate does not significantly affect the bypassing stream. It should, however, be noted that this last conclusion is only true for a laboratory scale reactor in which the linear fluid velocity is very small (0.3 - 0.75 m/h).

III.4b Experiments in the pilot-plants

The residence time distribution measurements in the pilot-plants resulted in five sets of dimensionless concentration-time relations, presented as the data points in the Figs. III-6 to 10 inclusive. After evaluation, five models were

Fig. III-6. Results of stimulus-response experiment 6 (6 m^3 pilot-plant) and the calculated responses of model 6. (x) sludge bed; (o) effluent; (—) V_b ; (---) V_{pf} .

Fig. III-7. Results of stimulus-response experiment 7 (30 m^3 pilot-plant) and the calculated responses of model 7. (x) sludge bed; (□) transitional area; (V) sludge blanket; (o) effluent; (—) V_b ; (-.-) V_{ta} ; (---) V_d ; (...) V_{pf} .

Fig. III-8. Results of stimulus response experiment 8 (30 m^3 pilot-plant) and the calculated responses of model 8. ($\square$) transitional area; ($\times$) sludge blanket; ($\circ$) effluent; (—) V_b ; (-.-) V_{ta} ; (---) V_d ; (...) V_{pf} .

Fig. III-9. Results of stimulus-response experiment 9 (30 m^3 pilot-plant) and the calculated responses of model 9. ($\times, \square$) sludge bed; ($\blacktriangle, \nabla$) transitional area; ($\bullet, *$) sludge blanket; ($\circ$) effluent; (—) V_b ; (-.-) V_{ta} ; (---) V_d ; (...) V_{pf} .

Fig. III-10. Results of stimulus-response experiment 10 (30 m^3 pilot-plant) and the calculated responses of model 10. (x, $\square$) sludge bed; ($\blacktriangle$, ∇) transitional area; ($\bullet$, $*$) sludge blanket; (o) effluent; (—) V_b ; (---) V_{ta} ; (-.-) V_d ; (...) V_{pf} .

obtained, the parameters of which have been collected in Table III-4; the responses of these models are the lines in the five figures.

Analogous to the presentation of the laboratory experiments, a block diagram giving a generalised picture, has been presented in Fig. III-16. This diagram differs from the model as suggested in Chapter I, on two main points. Instead of two perfectly mixed regions, three have to be distinguished in the reactor, and the dead space is absent. The first and third ideal mixer are interpreted as the sludge bed, and the sludge blanket; the second perfectly mixed region can be seen as an area of transition between the bed and the blanket. Such an area is also found by measurements of the sludge concentration over the reactor height. The second difference, the absence of dead space, indicates that the total reactor volume contained flowing liquid. This absence can be attributed to the pilot-plants operating for some weeks prior to the experiments being performed. Dead space that may have been present in the beginning, could have been supplied with substrate by diffusion process, resulting in both gas production and mixing. In such a way, the dead spaces may have diminished or even disappeared. The absence of dead space perhaps also implies that the influent distribution systems were well chosen. Similar to the laboratory experiments (the preceding section), the influence

TABLE III-4

The models representing the fluid flow pattern in the pilot-plants.

For comparison, the starting values of the independent parameters are also presented.

results	experiment no. :				
	6	7	8	9	10
V_b (% of V_R)	25+5	5+2.5	5+2.5	25+5	25+5
V_{ta} "	30+5	20+5	20+5	20+5	20+5
V_d "	40+5	70+5	70+5	50+5	50+5
V_{pf} "	5+2.5	5+2.5	5+2.5	5+2.5	5+2.5
F_k (% of F)	40+5	33+5	33+5	60+5	60+5
F_{k1} "	-	-	-	7+5	13+5
F_{k2} "	-	-	-	53+5	47+5
$F_{ta,b}$ "	170+50	170+50	170+50	160+50	160+50
$F_{d,ta}$ "	170+50	170+50	170+50	160+50	160+50
starting values					
V_b (% of V_R)	50	20	20	32	32
V_d "	42	75	75	51	51
V_{st} "	8	5	5	17	17

Fig. III-16. Generalised block diagram of the models 6-10, see also Table III-4.

of several process conditions on the fluid flow pattern could be determined by a comparison of the models developed for these five pilot-plant experiments. In Table III-5, the process conditions for these experiments are collected.

TABLE III-5.

Process conditions of the residence time distribution experiments in the pilot-plants.

	experiment no. :				
	6	7	8	9	10
date	May 1976	Sept. 1976	Sept. 1976	March 1977	April 1977
reactor	6 m ³	30 m ³ (1)	30 m ³ (1)	30 m ³ (2)	30 m ³ (2)
influent system	radial	radial	radial	one-dir.	radial.
X (kgSS)	340	490	490	1150	1150
% VSS	70	77	77	84	84
V _b (m ³)	~ 3	6	6	12	12
h _b (m)	~ 1.5	1.2	1.2	2.2	2.2
C _{md} (kg SS/m ³)	16	3.5	7	18	17
q (d ⁻¹)	~ 0.3	~ 0.65	~ 0.8	~ 0.5	~ 0.45
F (m ³ /h)	1.4	10.5	10.1	7.6	7.7
τ (h)	4.4	2.9	3.0	4.4	4.3
C _{so} (kg C/m ³)	~ 0.9	~ 0.4	~ 0.6	~ 1.1	~ 1.0
C _{s,e} (kg C/m ³)	~ 0.1	~ 0.1	~ 0.1	~ 0.05	~ 0.05
% conversion (TOC, centrifuged samples)	~90	~75	~85	~ 95	~ 95
ϕ _s ^{'''} (kg C/m ³ d)	~ 4.5	~ 2.5	~ 4.0	~ 5.8	~ 5.3
ϕ _g (m ³ /h)	1.54	2.7	4.5	6.5	6.0
ϕ _g ^{'''} (h ⁻¹)	0.25	0.09	0.15	0.20	0.18
C _{Li,o} (10 ⁻³ kg/m ³ , calc.)	-	-	-	11.5	-
C _{Li,o} (10 ⁻³ kg/m ³ , meas.)	8.0	10.3	10.4	11.76	12.82

During the experiments 6 and 8, the bed heights in both reactors were about the same, although the height of the 30 m³ pilot-plant (exp. 8) was twice that of the 6 m³ pilot-plant (exp. 6). This happened because the 6 m³ pilot-plant contained a sufficient amount of sludge, and the other reactor did not.

Therefore, the distributions of the total reactor volumes in the volumes of the sub-units were different for models 6 and 8 : in model 8, the first perfectly mixed region was extremely small (5 %), while the third was extraordinarily large (70 %); in model 6 the values were respectively 25 % and 40 % (see Table III-4).

The comparison of the bypassing flow rates (F_k) in both models led to the

conclusion that no significant difference was observed, despite the different volumes of gas per m^3 per unit of time during both experiments.

For experiments 7 and 8, the models appear to be identical. This indicates that the different flow rates of gas, caused by the different organic loadings and purification efficiencies during both experiments (see Table III-5), did not result in significantly altered fluid flow patterns. Hence, productions of $0.09 - 0.15 \text{ m}^3$ of gas per m^3 reactor volume per hour (ϕ_g'') cause similar mixing effects in an anaerobic upflow reactor.

Experiments 6 and 10 were performed in the 6 m^3 and the 30 m^3 pilot-plant. Both reactors worked efficiently during these experiments, and the 30 m^3 reactor contained a sufficient amount of sludge compared with earlier experiments (due to the growth of sludge in the course of time). The models obtained were almost similar with regard to the distribution of the sub-volumes; the small differences are attributed to the different geometries of the reactors. A more pronounced difference was observed between the flow rates of the bypassing streams. Because the liquid residence times and the volumes of gas per m^3 reactor volume per unit of time during these experiments corresponded, the different bypassing flow rates must have been caused by the different bed heights : a higher bed giving larger bypassing streams. A general conclusion can be drawn from the comparison of models 6, 8, and 10 : in scaling-up the amount of sludge is of first importance; this amount fixes the height of the sludge bed, which appears to influence the magnitude of the bypassing stream.

Because the process conditions and the settlers used during experiments 8 and 10 were quite different, the comparison of these experiments is complicated. Therefore, a systematic analysis of the differences has to be made. These differences are found in the volumes of the first and the third ideal mixers, and in the bypassing flow rates (see Table III-4). Since the plug flow regions were the same in both models, it can be concluded that the different settlers used, did not seriously influence the fluid flow pattern in the reactor. The different liquid residence times that were applied, did not significantly influence the fluid flow pattern either : during experiment 10, the influent flow rate was smaller than during experiment 8, so this parameter cannot be the cause of the larger bypassing stream in model 10. Based on the conclusion obtained from the comparison of the models 7 and 8, it may be assumed that the significant difference between the bypassing flow rates in the models 8 and 10, was not caused by the different flow rates of gas. It can be concluded that the differences between models 8 and 10 were mainly caused by the different amounts of sludge for the respective experiments. This accounts for the differences of the first perfectly mixed regions in both models (respectively 5

and 25 %), and the increase of the bypassing flow rate from 33 to 60 %. These conclusions confirm earlier conclusions.

III.4c Experiments in the full-scale reactor

The results of these experiments are presented as the data points in the four sets of dimensionless concentration-time relations in the Figs. III-11 to 14 inclusive. In these figures, the lines are the responses of the models. The parameters of these models are collected in Table III-6, together with the starting values of the trial-and-error procedure to find the best conforming values.

The generalised block diagram that represents these models (see Fig. III-17), and the suggested model in Chapter I, differ only in two cases by the presence of dead space in the settler. This was found when a large amount of mud was

Fig. III-11. Results of stimulus-response experiment 11 (200 m^3 reactor) and the calculated responses of model 11. Figure at the left : (x) SP1; (o) SP2; (□) SP4; (Δ) SP7; (---) V_b . Figure at the right : (o) influent; (●) effluent; (Δ) SP3, 6; (□) SP12; (x) SP18; (---) V_b , V_d ; (-.-) V_{pf} .

Fig. III-12. Results of stimulus-response experiment 12 (200 m^3 reactor) and the calculated responses of model 12. Figure at the left : (o) SP4; (□) SP7; (●) SPI; (x) SPII; (★) SPIII; (Δ) SPIV; (■) SPV; (all SP's in the sludge bed); (---) V_b . Figure at the right : (o) influent; (●) effluent; (x) SP5; (★) SP1, 2; (□) SP12, 13; (---) V_b , V_d ; (-.-) V_{pf} .

Fig. III-13. Results of the stimulus-response experiment 13 (200 m^3 reactor) and the calculated responses of model 13. Figure at the left : (Δ) SP4; (□) SP7; (●) SPI; (o) SPII; (x) SPIII; (SP's in the sludge bed); (---) V_b . Figure at the right : (▽) influent; (★) effluent; (●) SP1; (o) SP2, 3; (□) SP6; (Δ) SPI2; (x) SPI3, 18; (---) V_b , V_d ; (-.-) V_{pf} .

Fig. III-14. Results of stimulus response experiment 14 (200 m³ reactor) and the calculated responses of model 14. Figure at the left : (Δ) SP4; (□) SP7; (●) SPI; (○) SPII; (x) SPIII; (SP's in the sludge bed); (---) V_b . Figure at the right : (▽) influent; (*) effluent; (●) SP1, 2; (○) SP 3, 6; (□) SP12, 18; (x) SP13; (Δ) SP14; (---) V_b , V_d ; (-.-) V_{pf} .

Fig. III-17. Generalised block diagram of the models 11-14, see also Table III-6.

present both in the reactor and in the settler. Apparently, the fluid flow pattern in the settler was considerably influenced by this mud. For the fluid flow pattern in the reactor, the presence of this mud had no significant influence, as there was an almost homogeneous distribution of it over the reactor by the turbulence that was caused by the production of gas. The mud did not contribute to the formation of a larger sludge bed.

The process conditions during the experiments were such that by means of the models, the effects of the flow rate of gas, the liquid residence time, the

TABLE III-6

The models representing the fluid flow pattern in the 200 m³ reactor.

For comparison, the starting values of the independent parameters are also presented.

Results	experiment no. :			
	11	12	13	14
V_b (% of V_R)	18+2.5	22+2.5	11+2.5	12+2.5
V_d "	59+5	66+5	66+5	67+5
V_{pf} "	23+2.5	12+2.5	8+2.5	10+2.5
V_{ds} "	0+2.5	0+2.5	16+2.5	12+2.5
F_k (% of F)	61+5	71+5	86+5	83+5
F_{db}	160+50	160+50	160+50	160+50
starting values :				
V_b (% of V_R)	8.5	7.4	10.3	8.6
V_d "	65.3	66.4	63.5	65.2
V_{st} "	26.2	26.2	26.2	26.2

influent velocity per influent point, and the concentration of sludge, could be studied (see Table III-7).

Models 11 and 12 differ only in the volumes of the plug flow regions in the settlers, and the bypassing flow rates in the reactor (see Table III-6).

During experiment 11, almost all the fluid in the settlers was in plug flow, which indicated that the flow rate of gas (see Table III-7) did not generate a circulation stream in the settlers. During experiment 12 the plug flow region was smaller, which indicated that part of the settler volume was either well mixed or dead space. Since dead space was absent in the models developed for experiments 11 and 12, the different plug flow regions must have been the result of mixing in the settlers. Apparently, the larger gas production during experiment 12 did generate a circulation stream in the settler, resulting in a decrease of the plug flow region, and an increase of the second perfectly mixed region in model 12. The larger bypassing stream for model 12 would also be caused by the larger gas production, as the other differing process condition - the liquid residence time - could not be the cause, since during experiment 12 the influent flow rate was smaller than during experiment 11.

In the models 12 and 13 the volumes of the second perfectly mixed regions are equal, and the plug flow regions have almost equal volumes; but the first

TABLE III-7

Process conditions of the residence time distribution experiments in the 200 m³ reactor.

	experiment no. :			
	11	12	13	14
	Oct. 1977	Oct. 1977	Dec. 1977	Dec. 1977
number of influent points	10	10	10	4
X (kg SS)	1795	3510	6465	5565
% VSS	73.5	53.5	38.5	38.5
V _b (m ³)	17.3	15.5	20.8	17.7
h _b (m)	0.39	0.35	0.47	0.40
C _{md} (kg SS/m ³)	5.74	17.3	33.6	29.9
q (d ⁻¹)	0.65	0.53	0.73	0.85
F (m ³ /h)	35.5	20.9	24.5	30.5
τ (h)	5.7	9.8	8.3	6.7
C _{so} (kg C/m ³)	0.50	0.94	1.43	1.18
C _{s,e} "	0.19	0.15	0.10	0.09
C _{fa,e} "	0.05	0.05	0.05	0.06
% conversion (TOC, centrifuged samples)	61.5	84.0	93.0	93.0
% conversion (fa)	91.0	95.0	97.0	95.0
ϕ _s " (kg C/m ³ d)	1.3	1.9	3.8	3.9
ϕ _g (m ³ /h)	11.8	18.8	46.0	40.5
ϕ _g " (h ⁻¹)	0.06	0.09	0.23	0.20
V _{L,o} (m/s)	0.9	0.5	0.6	1.9
C _{Li,o} (calc. 10 ⁻³ kg/m ³)	-	-	5.64	8.0
C _{Li,o} (meas. 10 ⁻³ kg/m ³)	5.78	9.87	5.61	7.92

perfectly mixed regions, and the bypassing flow rates, are different. Furthermore, in model 13 a dead space region is present. Since all other process conditions had the same order of magnitude during both experiments, these differences can only be caused by the different flow rates of gas and the different concentrations of sludge in the blanket (C_{md}). A larger gas production gives better mixing, but due to the extremely small height of the sludge bed, it may also influence the bypassing streams. The high value of C_{md} during experiment 13 (see Table III-7) caused accumulation of sludge in the settler which resulted in the occurrence of dead space regions there. Also at the bottom of

the reactor, dead spaces may have been present; this follows from model 13, as the volume of the first perfectly mixed region was halved. Moreover, these dead spaces can also be the result of the large C_{md} value, which caused extra thickening of sludge at the bottom of the reactor.

III.4d Influent distribution

The process conditions during the pulse tracer experiments in the 30 m³ pilot-plant are presented in Table III-8. The results obtained, i.e. the tracer curves measured at the sample points around the influent openings, fluctuated

TABLE III-8

Results of the pulse-tracer experiments to study the fluid flow pattern in the 30 m³ pilot plant as a function of the influent distribution system.

A <u>peak</u> signal rapidly rises and falls.									
An "extra strong signal" (<u>ess</u>) increases to higher values than the mean signal in the reactor.									
The <u>mean</u> signal is a relatively slowly increasing signal of the maximum value C_{Li} .									
	experiment no. :								
	I-1		I-2			I-3		I-4	
Influent system	radial		unidirectional			radial		unidirectional	
$\emptyset_g$ (m ³ /h)	4.98		4.89			3.25		3.07	
F (m ³ /h)	2.35		3.0			4.35		4.57	
$C_{Li,o}$ (kg/m ³)	0.65		0.86			0.62		0.48	
C_{Li} (10 ⁻³ kg/m ³)	4.0		4.0			2.0		1.5	
Peak in SP no :	1	9	-			3	9	2	3
t (min)	4.0	5.0	-			6.5	8.0	7.0	10.0
C_{Li} (max) (10 ⁻³ kg/m ³)	25	18	-			78	18	68	66
v_k (10 ⁻³ m/s)	2.1	4.3	-			1.6	2.7	2.4	1.8
<u>ess</u> in SP no :	I-1	I-4	2	5	9	-		I-2	
t (min)	9	8	8	12	6	-		9	
C_{Li} (max) (10 ⁻³ kg/m ³)	14	24	7	8	15	-		12.5	
<u>mean</u> in SP no :	4	I-3	3	4		2	4	5	9
t (min)	6.5	4	7.0	7.5		6	9	10	6.5

strongly and were therefore evaluated before they have been presented in Table III-8. Each curve was examined for the presence of a peak signal (indicating a bypassing stream); an extra strong signal, marked ess (indicating preference flow, not as strong as bypassing streams); and the mean signal (indicating that the volume around the sample point is well mixed). From these results (see Table III-8) it appeared that the fluid flow pattern around the radial influent opening varied both with the process conditions applied, and in the long run with time. When this system was used, both horizontally and vertically directed bypassing streams could be detected at the same time. When a horizontal, unidirectional influent system was applied, peaks were absent or came later than for the radial system. It was also found that the direction of a bypassing stream changed from horizontal to vertical, this appeared from the response at SP I-2, Exp. I-4, that is located between the SP's 2 and 3 (but closer to the bottom). At this sample point, a response was found that was smaller than a peak signal (characteristic for bypassing), but stronger than the mean signal that is usually found in well mixed regions.

The process conditions during the residence time distribution experiments, that were performed to study the effect of the influent distribution system on the fluid flow pattern in the 30 m^3 pilot-plant (experiments 9 and 10), have already been presented in Table III-5. The models, developed for these experiments, are presented in Table III-4. From the small differences found, it is concluded that the radial influent distribution system, and the unidirectional horizontal influent distribution system give about the same fluid flow pattern, and therefore about the same influent distribution. From the pulse tracer experiments it was found that the radial system gives an influent distribution of local bypassing streams that change with time.

On this basis, the unidirectional influent distribution system was chosen for application in the 200 m^3 reactor. Because the bottom surface was about 44 m^2 , several influent points had to be used. Since the 30 m^3 reactor had operated well in the absence of dead spaces, with only one influent point on a bottom surface of 5.39 m^2 , the influent for the 200 m^3 reactor was distributed over 10 influent points (one point on 4.4 m^2).

In the 200 m^3 reactor, some pulse tracer experiments were performed to study the dimensions of the area covered by one influent point. For this purpose, the concentrated tracer solution was brought into the influent at point D (see Figs. I-16 and I-18, and Table I-8), while the surrounding sample points were used for frequent sampling. The results of such an experiment with the reactor in a start-up situation, are presented in Table III-9 (1). In Table III-9 (2) the results of an experiment with the reactor in full operation

TABLE III-9

Results of pulse-tracer experiments to study the fluid flow pattern around influent point D in the 200 m³ reactor.

Experiment 1 : Reactor in start-up situation, 4 influent points in use, (B, D, G, J), December 1977, $\phi_g = 8 \text{ m}^3/\text{h}$, $F = 26 \text{ m}^3/\text{h}$.

time after injection θ	SP-1 $C_{\text{Li}}/C_{\text{Li},o}$	SP-2 $C_{\text{Li}}/C_{\text{Li},o}$	other SP's $C_{\text{Li}}/C_{\text{Li},o}$
0.004	0	10950	0
0.008	0	11600	0
0.013	0	9250	0
0.017	20	6300	0
0.021	10	5250	0
0.025	525	2950	0
0.030	700	175	0

Experiment 2 : Reactor in operation, 10 influent points in use, October 1977, $\phi_g = 29.6 \text{ m}^3/\text{h}$, $F = 26.8 \text{ m}^3/\text{h}$.

θ	SP-1 $C_{\text{Li}}/C_{\text{Li},o}$	SP-2 $C_{\text{Li}}/C_{\text{Li},o}$	SP-3 $C_{\text{Li}}/C_{\text{Li},o}$	SP-4 $C_{\text{Li}}/C_{\text{Li},o}$	SP-5 $C_{\text{Li}}/C_{\text{Li},o}$	SP-6 $C_{\text{Li}}/C_{\text{Li},o}$
0.007	0	91	34	330	216	5
0.011	114	91	239	352	136	11
0.015	68	136	125	193	57	34
0.020	68	46	80	102	68	57
0.024	46	80	91	102	170	68
0.033	57	57	68	114	68	80

are collected. From these tables it can be seen that in a properly operating reactor the sample points at a distance of 1.5 m from the influent point D (SP's 3, 4, 5 and 6) are sufficiently supplied with tracer and thus also with the substrate. If the reactor does not produce enough gas to mix the system, a very significant bypassing stream occurs (measured in SP 2, just above influent point D (see Table III-9 (1)), that brings almost all the influent directly into the sludge blanket. A similar effect was observed in the 30 m³

pilot-plant.

From the four residence time distribution experiments, experiments 13 and 14 were performed to compare the fluid flow patterns in the reactor if either 10 or 4 (B, D, G, J) influent points were used. The conditions, results and models of these experiments have already been presented (see Tables III-6 and 7, and Figs. III-13 and 14). The models derived from the experimental data are almost similar, so it was concluded that instead of 10 influent points, 4 influent points were sufficient for a good influent distribution on a bottom surface of 44 m^2 (the implication being that with this geometry, only one influent point is sufficient to cover 11 m^2 of bottom surface). This may be true in a stationary situation when no start-up problems are to be expected anymore. However, in experiment 11, seven days after the start-up of the reactor, the sample points 4 and 7 clearly showed that dead spaces were present in the reactor (see Fig. III-11); the pulse experiments in the reactor at start-up (described in the foregoing paragraph) showed a large bypassing stream. Therefore, one influent point on a bottom area of 5 m^2 is recommended.

III.5 Discussion and conclusions

In Fig. III-18 the reactors in which the residence time distribution experiments have been performed, are presented schematically, together with the generalised block diagrams obtained. Because all the models were developed from the same starting point, namely the model depicted in Chapter I, all the models were combinations of ideal mixers, plug flow, bypassing flow, and backmix streams. As quantitative starting points for the development of the models, the values for the volumes of the bed, blanket and settler, have been used that were derived from the dimensions of the reactors, and from the measurements of the sludge concentration over the reactor height. In all cases, this resulted in models whose responses fitted the experimental data well, while the volumes in these models were about equal to the actual volumes in the reactors.

Similarities between the models, that did not follow directly from the starting points used, but from the interpretation of the experimental data, are :

- the presence of dead space regions in the models for the laboratory reactor and the full-scale reactor;
- the presence of three perfectly mixed regions in the models for the pilot-plants.

Both in the settler of the laboratory reactor and in those of the full-scale reactor, dead spaces were present, if there was an accumulation of sludge. In the settlers this sludge produced almost no gas, because of the absence of

Fig. III-18. Collection of the reactors in which residence time distribution experiments were performed, and the fluid flow models derived.

substrate. This leads to the conclusion that non-active sludge blocks the fluid flow. In the full-scale reactor, this accumulation was the result of the increased mud concentration in the blanket; in the laboratory reactor, the recirculation opening was too small, and therefore easily blocked. Another similarity between the fluid flow patterns in both reactors, was that both models contained two ideal mixers, which resulted from the bed heights in both reactors having the same order of magnitude. For a laboratory reactor of about 60 l, a bed height of about 0.4 m was suitable; for a full-scale reactor with a total reactor height of 4.5 m, a bed height of 0.4 m was extremely small. The very large bypassing streams in this reactor were caused by this extraordinarily small bed height.

The fluid flow patterns in the pilot-plants corresponded fairly well. This is furthermore illustrated by the fact that, if by means of the volumes of

the second perfectly mixed regions, the heights of the areas of transition between the sludge bed and the sludge blanket are calculated, for the 6 m^3 pilot-plant a height of $0.9 \pm 0.15 \text{ m}$ is found, and for the 30 m^3 pilot-plant a height of $1.2 \pm 0.3 \text{ m}$; these values are about the same as experimentally observed.

The most important differences between the generalised block diagrams are :

- the models of the fluid flow patterns in the pilot-plants contain three perfectly mixed regions, while the other models contain only two;
- the models for the pilot-plants do not contain dead space regions, that are present in several other models;
- the bypassing flow rate in the models for the full-scale reactor is much larger than in the other models.

The first difference has to do with the various bed heights in the pilot-plants and the other reactors. In the laboratory reactor and the full-scale reactor, the bed heights of about 0.4 m were so small, that in fact only the area of transition and the sludge blanket were present. If the full-scale reactor had contained sufficient sludge, the bed height would have been $1.5 - 2.5 \text{ m}$ and, based on the results of the experiments in the pilot-plants, it would be expected that a "three mixer model" might have been found. The laboratory reactor contained a proper amount of sludge. It should be noted that the size of this reactor was very small compared with the other reactors, implying that the models for the laboratory reactor can hardly be compared with the other models. Therefore, the relation between the local production of gas in the sludge bed, and the backmix flow, determined in this reactor, was only used to derive models for the laboratory experiments.

The absence of dead space regions in the pilot-plants has already been explained in section III.4b. The presence of dead space in the settlers of the other reactors had invariably been the result of accumulation of sludge in these settlers. Such an effect did not occur in the 30 m^3 pilot-plant, as a large recirculation opening was present in the bottom of the settler, and the concentration of sludge in the blanket did not exceed 20 kg SS/m^3 .

The high values of the bypassing flow in the full-scale reactor (such as 86 % of the total influent), were due to the small height of the sludge bed. Moreover, the influent velocity at any influent point was such that large bypassing through the low sludge bed could be expected. In the laboratory reactor, the linear influent velocity at an influent point, did not exceed 0.12 m/s , so that here, only bypassing of minor importance occurred at bed heights of about 0.4 m .

On the basis of similarities and differences discussed here, the following conclusions can be drawn with respect to the scaling-up of an upflow reactor : if the reactor height is changed from 3 to 6 m, and the bed height is increased accordingly, quite similar fluid flow models are found. The only difference that must be mentioned, is the occurrence of larger bypassing streams the higher the sludge bed. Therefore, the height of the sludge bed is restricted to a maximum, since the bypassing streams can become so large that the effective use of the sludge in the lower parts of the bed would be reduced. A maximum value of the height of the sludge bed of 2 to 3 m is therefore recommended. The models obtained from the experiments in the 200 m^3 reactor, contain one ideal mixer for the sludge bed, although the scattering of the data measured in different places in the bed indicated that, rather than one ideal mixer, a combination of several parallel ideal mixers would have been appropriate. If the distribution of the influent over the bottom is uniform, no distinction will be found between the tracer curves measured in different places in the bed; and one ideal mixer, together with one bypassing stream might very well be used to describe the fluid flow pattern in the bed. In case of the 200 m^3 reactor, it can therefore be concluded that the influent was sufficiently but not uniformly distributed. This fact, and the other similarities between the models for the 30 m^3 pilot-plant and the 200 m^3 reactor, that have already been discussed, lead to the conclusion that scaling-up of a reactor in horizontal directions is quite feasible, and, provided that the influent is well distributed over the bottom, not restricted to any maximum. Finally, it follows that the pilot-plants gave valuable data for scale-up, as well as useful fluid flow models. The laboratory reactor, however, did not give many results that can be used for scale-up. In the 200 m^3 reactor, the amount of sludge was too small, giving too low a sludge bed. Therefore, the results of these experiments are somewhat extraordinary, and further experiments, when this reactor contains a normal amount of sludge, may give other models and results.

CHAPTER IV

DISTRIBUTION OF SLUDGE IN THE UPFLOW REACTOR

IV.1 Model

It follows from the mass-balance

$$\frac{dx_R}{dt} = \phi_{m,o} + \text{growth} - \phi_{m,e} - \phi_{m,o'} \quad \text{IV-1}$$

that the amount of solids in the upflow reactor can increase in two ways (see Fig. IV-1) : if the growth of sludge exceeds its washout; and if suspended solids in the influent accumulate in the reactor (if $\phi_{m,o} > \phi_{m,o'}$). The second possibility will seldom occur, because usually precautions are taken to prevent this. Therefore, Eq. IV-1 can often be simplified to :

$$\frac{dx_R}{dt} = \text{growth} - \phi_{m,e} \quad \text{IV-2}$$

Integration of this equation for the period of time that the reactor operates, gives a total for the increase or decrease of sludge. The slow growth of anaerobic sludge (see Chapter II), causes that the effluent may contain only small amounts of sludge in order that $\text{growth} > \phi_{m,e}$; and even then it takes a long time before an increase of the amount of sludge in the reactor can be detected (see for instance the increase of sludge as described in Chapter I).

It appears therefore, that for the establishment of a model that describes the distribution of sludge in the upflow reactor, it can be assumed that for short periods the amount of sludge in the reactor is constant. As has already been discussed in Chapter I, in an upflow reactor that contains anaerobic sludge, the flow

Fig. IV-1. Transport of solids in the anaerobic process of wastewater treatment.

rate of gas, and the fluid flow rate, cause a distribution of sludge over the height of the reactor. In the sludge bed, the concentration of sludge was

found to be high ($50-100 \text{ kg SS/m}^3$) and almost constant over the ranges of the process conditions that were mostly applied. In the sludge blanket, the sludge concentration is dependent on the process conditions, and lower than in the bed. To predict the performance of wastewater treatment in both the bed and the blanket, the quantitative relations between the concentration of sludge (in these reactor parts) and the process conditions, have to be known. Furthermore, it has to be known in what manner the washout of sludge via the settler depends on the concentration of sludge in the blanket. During all the experiments that were performed after the adaptation period of about one year, the sludge concentration in the bed was almost constant, so that for the model this concentration will be assumed to be constant.

Fig. IV-2. Transport of solids in the upflow reactor, containing a bed and a blanket.

Therefore, the behaviour of the sludge in an upflow reactor can be depicted schematically as in Fig. IV-2.

If no accumulation of influent solids occurs in the reactor ($\phi_{m,o} = \phi_{m,o'} = \phi_{m,o''}$), if the settler operates efficiently, and if the process is stationary (with the consequence that, at a given set of process conditions, C_{md} is constant), the following equation holds :

The model described in this chapter presents the relation between the sludge concentration in the blanket and the process conditions. In the next chapter the behaviour of the sludge in the settler will be discussed. For the model presented here, the settler can be seen as an apparatus by which the sludge that is carried upwards in the wake of the gas bubbles in the blanket, is separated from the gas, and - with a certain efficiency - is recirculated to the top of the blanket, from where it can fall back into the direction of the bed. In other words, the settler can be seen as an apparatus that reduces the concentration of sludge in the exit of the blanket.

$$\phi_{m,bd} = \phi_{m,db}$$

IV-3

The transport of sludge upwards is caused by rising gas bubbles that whirl up the sludge bed and carry a sludge suspension in their wake upwards. The linear fluid velocity in the reactor is mostly small compared with the settling velocity of the sludge (so that this stream does not contribute to the upward sludge transport). Therefore, the transport of sludge from the bed to the blanket ($\phi_{m,bd}$), is proportional to the sludge concentration in the bed (C_{mb}), and to the upward fluid flow that is the sum of the wakes of the gas bubbles that rise per unit of time from the bed to the blanket ($F_{bd,w}$) :

$$\phi_{m,bd} = k_{mb} F_{bd,w} C_{mb} \quad \text{IV-4}$$

in which

$$F_{bd,w} = k_w \phi_{gb}, \quad \text{IV-5}$$

therefore,

$$\phi_{m,bd} = k_w k_{mb} \phi_{gb} C_{mb} \quad \text{IV-6}$$

or

$$\phi_{m,bd} = k_{tb} \phi_{gb} C_{mb} \quad \text{IV-6}$$

The transport of sludge downwards, from the blanket to the bed ($\phi_{m,db}$), is caused on the one hand by the settling of sludge and on the other hand by the transport of the sludge by the backmix stream (F_{db}). Both transports are proportional to the sludge concentration in the blanket (C_{md}) :

$$\phi_{m,db} = (v_s - v_L) A_R C_{md} + k_{md} F_{db} C_{md} \quad \text{IV-7}$$

Because it can be assumed that $F_{db} = F_{bd,w}$, Eq. IV-7 can be rewritten as :

$$\phi_{m,db} = (v_s - v_L) A_R C_{md} + k_w k_{md} \phi_{gb} C_{md} \quad \text{IV-8}$$

or

$$\phi_{m,db} = (v_s - v_L) A_R C_{md} + k_{td} \phi_{gb} C_{md} \quad \text{IV-8}$$

According to Eq. IV-3 the upgoing sludge stream is equal to the sludge stream going downwards, so Eqns. IV-6 and IV-8 can be combined to obtain :

$$k_{tb} \phi_{gb} C_{mb} = (v_s - v_L) A_R C_{md} + k_{td} \phi_{gb} C_{md} \quad \text{IV-9}$$

from which C_{md} can be calculated :

$$C_{md} = \frac{\phi_{gb} k_{tb} C_{mb}}{\phi_{gb} k_{td} + (v_s - v_L) A_R} \quad \text{IV-9}$$

In this model two extreme situations can be distinguished :

- a. If free settling of the sludge particles is hindered by a high sludge concentration in the blanket, or if the reactor contains sludge with poor settling properties, the term $(v_s - v_L) A_R$ should be small compared with the term $\phi_{gb} k_{td}$. In that case, Eq. IV-9 can be simplified to :

$$\frac{C_{md}}{C_{mb}} = \frac{k_{tb}}{k_{td}} \quad \text{IV-10}$$

or, the ratio of the sludge concentrations in the blanket and in the bed becomes equal to the ratio of the sludge transport factors.

- b. If the volume of gas produced per unit of time is low, and the sludge has good settling properties, the term $(v_s - v_L) A_R$ will be large compared with the term $\phi_{gb} k_{td}$. Then, Eq. IV-9 can be changed into :

$$C_{md} = \frac{\phi_{gb} k_{tb} C_{mb}}{(v_s - v_L) A_R} \quad \text{IV-11}$$

which means that C_{md} is proportional to the volume of gas per unit of time at the top of the sludge bed (see also (77)).

To use the model presented in Eq. IV-9, the values of the following parameters have to be known : C_{md} , C_{mb} , ϕ_{gb} , v_s , v_L and A_R . To find the values of k_{tb} and k_{td} , experiments must be performed during which the distribution of sludge over the height in the reactor is measured for different volumes of gas produced per unit of time (ϕ_g).

By means of the measured values of ϕ_g , the values of ϕ_{gb} can be calculated if it is known how much of the substrate is converted into gas in the sludge bed. In other words, the distribution of substrate over the bed and the blanket has to be known. This distribution is in accordance with the models of the fluid flow pattern in the upflow reactor that had already been obtained (see Chapter III) : in these models the influent splits into a stream F_o , that enters the bed, and a stream F_k , that bypasses the bed and enters the blanket. The distribution of the substrate over the bed and the blanket is proportional to

the influent distribution over both reactor parts. If the conversion in the bed and in the blanket is complete, the ratio of the gas production in the bed and in the blanket is equal to the ratio F_o/F_k . Consequently, the following relation between the total volume of gas produced and the volume of gas produced in the bed, both at atmospheric pressure, can be formulated :

$$\phi_{gb} = \frac{F_o}{F} \phi_g \quad \text{IV-12}$$

The volume of gas per unit of time at the top of the sludge bed ($\phi_{gb'}$) is not equal to ϕ_{gb} because at this place, the pressure is higher than the atmospheric pressure due to the presence of a liquid column with a height of $h_d + h_{st}$ above the bed :

$$\phi_{gb'} = \phi_{gb} \left(\frac{10}{10 + h_d + h_{st}} \right) = \frac{F_o}{F} \phi_g \left(\frac{10}{10 + h_d + h_{st}} \right) \quad \text{IV-13}$$

The parameter h_{st} (the height of the settler) in this equation, is a reactor constant. The height of the sludge blanket follows from the measurements of the distribution of sludge in the reactor.

The settling velocity v_s must be measured separately. The linear liquid velocity in the reactor (v_L) follows from the value of the influent flow rate (F) (being a process condition that can easily be measured), and the value of A_R (a reactor constant). If Eq. IV-9 is changed into :

$$\frac{C_{mb}}{C_{md}} = \frac{(v_s - v_L) A_R}{k_{tb}} \cdot \frac{1}{\phi_{gb'}} + \frac{k_{td}}{k_{tb}} \quad \text{IV-14}$$

and v_s and v_L are constant, it can be seen that a straight line should be obtained if C_{mb}/C_{md} is plotted against $1/\phi_{gb'}$. From the slope of this line, the value of k_{tb} can be calculated and by extrapolation of the line to $1/\phi_{gb'} = 0$, the value of the ratio k_{td}/k_{tb} is obtained, and thus by means of k_{tb} also k_{td} is found.

IV.2 Experimental

The measurements have been performed in the laboratory reactor, the 30 m³ pilot-plant, and the 200 m³ full-scale reactor, that have already been described in Chapter I. To check the model, and to find the values of k_{tb} and k_{td} , the distribution of sludge over the height of the reactor, was measured for different gas productions. In the 200 m³ reactor, the distribution of sludge over the diameter was also determined at several heights. For the

measurements, samples were taken from sample points, mounted in the reactor wall, as well as by means of a sampler that could be used at several places at the top of the reactor and that could be moved in a vertical direction. The settling velocity of the sludge was determined by means of a batch settling column ($h = 2 \text{ m}$, $d = 0.14 \text{ m}$). This column was filled with the sludge suspension from the blanket in the reactor and then mixed carefully. After that, the concentration of sludge, 1 m below the surface, was determined as a function of time. Via this procedure, the cumulative settling velocity distribution is obtained (78). From this distribution, the median settling velocity v_s (50 % settles faster, 50 % slower) has been used. If the sludge concentration was higher than $15\text{--}20 \text{ kg SS/m}^3$, the sludge particles hindered settling in such a way that all the particles settled with about the same average velocity. This velocity could be determined by means of the batch column. In the 200 m^3 reactor, such a situation existed by the presence of mud. Here the average settling velocity could be determined by means of sludge concentration measurements in the settler with upward fluid flow. In the batch column, no concentration measurements had to be performed, because for the determination of the average settling velocity, only the boundary layer between the sludge suspension and the supernatant had to be followed on its way downwards. In the settler of the 200 m^3 reactor, this boundary layer would stay at the horizontal plane where the linear, upward fluid velocity is equal to the average settling velocity. To calculate this velocity, the surface area of the boundary layer in the settler has to be known, together with the flow rate of the effluent.

The determination of the sludge concentration was performed by centrifuging a sample for 20 minutes in a laboratory centrifuge (4000 rpm); then decanting; after which, the solids were dried at 105°C until a constant weight. At low sludge concentrations, after decanting the solids were washed with distilled water, again centrifuged and decanted, and then dried.

IV.3 Results

IV.3a Experiments in the laboratory reactor

In Table IV-1, the results of the measurements in the laboratory reactor have been collected, together with the most important process conditions applied during these experiments. After plotting C_{mb}/C_{md} against $1/\phi_{gb}$, a straight line is obtained by means of linear regression analysis of the data points (see Fig. IV-3). Therefore, it can be concluded that the model presented for the description of the behaviour of the sludge, can be used for this system.

TABLE IV-1

Distribution of sludge in the laboratory reactor.

Process conditions : $v_s = (0.56 \pm 0.09) 10^{-3} \text{ m/s}$
 $C_{mb} = (70.1 \pm 1.7) \text{ kg SS/m}^3$
 $F = 10.8 \cdot 10^{-3} \text{ m}^3/\text{h}$
 $h_b = 0.25 \text{ m}$
 $F_k = 0.33 F = 3.6 \cdot 10^{-3} \text{ m}^3/\text{h}$

Experiment no.	ϕ_g $10^{-3} \text{ m}^3/\text{h}$	$\phi_{gb'}$ $10^{-3} \text{ m}^3/\text{h}$	C_{md} kg SS/m^3
1	14.1	9.0	3.7 ± 0.12
2	14.1	9.0	3.9 ± 0.15
3	19.8	12.6	4.5 ± 0.13
4	18.2	11.6	5.6 ± 0.15
5	29.0	18.4	7.8 ± 0.5
6	29.6	18.8	7.1 ± 0.3

Fig. IV-3. The ratio of the sludge concentrations in the bed and the blanket of the laboratory reactor, as a function of the inverse of the volume of gas per unit of time at the top of the sludge bed.

From the slope of the line, the value of k_{tb} has been calculated to be 0.8 ± 0.25 . After extrapolation of the line to $1/\phi_{gb'} = 0$, the ratio of k_{td}/k_{tb} appeared to have the value +1, but if the 90 % reliability interval (that is also

presented in Fig. IV-3) is taken into account, it has to be concluded that this ratio can vary from -1 to +3.

IV.3b Experiments in the 30 m³ pilot-plant

From the beginning of October until the end of November, well adapted sludge consisting of quickly settling sludge particles, was present in the 30 m³ pilot-plant. In a second period, from the end of November until the end of the campaign, an extra amount of sludge from a municipal digester was present. This sludge had inferior settling properties compared with the sludge particles already present, and was therefore, even at low gas flow rates, distributed uniformly over the reactor.

TABLE IV-2

Distribution of sludge in the 30 m³ pilot-plant (first period).

experiment no.	ϕ_g m ³ /h	$\phi_{gb'}$ m ³ /h	$\phi_{g,ta'}$ m ³ /h	F m ³ /h	C _{md} kg/m ³ (SS)	C _{mb} kg/m ³ (SS)	C _{m,ta} kg/m ³ (SS)	v _s 10 ⁻³ m/s
1	6.3	2.8	3.0	8.0	9.8+1.2	-	-	-
2	4.2	1.8	1.9	8.5	5.2+1.0	-	-	9.4
3	3.6	1.6	1.7	8.0	5.7+0.6	77.9+2.1	13.4	13.8
4	3.6	1.6	1.7	8.4	5.2+0.7	88.3+4.7	21.1	-
5	5.3	2.4	2.6	8.0	8.2+1.3	94.5+6.8	45.3	13.8
6	5.3	2.4	2.6	8.0	8.3+1.1	84.3+2.5	40.9	10.7
7	7.0	3.0	3.2	8.6	11.6+2.1	84.7+4.3	49.6	8.6
8	6.8	3.0	3.2	8.0	7.8+0.9	88.7+6.4	44.3	-
9	6.0	2.6	2.8	4.3	9.9+1.1	82.8+5.3	36.5	11.5
10	7.6	3.4	3.6	4.2	10.4+1.1	83.4+4.2	9.2	14.0
11	4.4	2.0	2.1	7.0	11.0+1.1	83.2+3.8	52.2	-
12	4.4	2.0	2.1	7.9	8.3+0.9	80.3+3.3	29.2	-
13	6.3	2.8	3.0	6.8	9.4+0.4	79.3+5.0	35.9	10.4
14	6.4	2.8	3.0	6.7	9.7+1.5	82.0+5.9	27.5	-
15	2.1	1.0	1.0	4.2	3.6+0.4	77.4+5.8	5.0	11.1
16	1.5	0.6	0.6	2.5	2.7+0.3	83.3+4.3	5.3	12.6

For the measurements of the sludge concentration in the pilot-plant during these periods, from each sample point (indicated in Fig. I-13) at least duplicate samples were taken. It appeared that the sludge concentrations determined in the samples from both the sludge bed and the sludge blanket, were almost constant, and therefore, for the measurements over these two areas, the two mean values are calculated and presented. During the first period, the sludge concentrations obtained by sampling in the area between the bed and the blanket were significantly different from both the concentration in the bed and that in the blanket. Therefore, the mean concentration of the sludge in this area of transition has also been presented.

The results obtained during the first period, together with the values of the process conditions that are needed to use Eq. IV-14, are presented in Table IV-2. The results obtained and presented, give three possibilities to check the model, namely : by plotting $C_{mb}/C_{m,ta}$ and C_{mb}/C_{md} against $1/\phi_{gb}$, and $C_{m,ta}/C_{md}$ against $1/\phi_{g,ta}$. Of these, the first two plots are indeed straight lines (see Fig. IV-4 and Eq. IV-14), while the third plot did not conform to the model. Here a plot of $C_{m,ta}/C_{md}$ against $\phi_{g,ta}$, appeared to give a straight line (see Fig. IV-5).

Fig. IV-4. The ratio of the sludge concentrations in : (●) the bed and the blanket in the 30 m³ pilot-plant, or (x) the bed and the transitional area in this reactor, as functions of the inverse of the volume of gas per unit of time at the top of the sludge bed.

Fig. IV-5. The ratio of the sludge concentrations in the transitional area and the blanket in the 30 m³ pilot-plant, as a function of the volume of gas per unit of time at the top of the transition area.

As the settling velocity of the sludge during this period was almost constant, the mean value $v_s = 11.6 \cdot 10^{-3}$ m/s has been used for both plots in Fig. IV-4 to determine the transport factors; for $C_{mb}/C_{md} - 1/\phi_{gb}$, $k_{tb} = 12.8 \pm 1.8$; and for $C_{mb}/C_{m,ta} - 1/\phi_{gb}$, $k_{tb} = 20.2 \pm 2.1$. The difference between these values originates from the efficiency of the sludge transport upwards, that should be lower the larger the height above the bed. By extrapolation of the $C_{mb}/C_{md} = 1/\phi_{gb}$ -graph to the ordinate in Fig. IV-4, the ratio of the two k-factors is found: $k_{td}/k_{tb} = 3.0 \pm 0.9$.

Fig. IV-5 indicates that at small volumes of gas produced per unit of time no area of transition will be found, and that at higher gas productions (above 2 m³/h) the importance of such an area will increase. From the linear regression analysis of the data in this figure, it was found that $C_{m,ta}/C_{md} = (1.07 \pm 0.1)\phi_{g,ta} + (1.3 \pm 0.3)$.

The results of the measurements during the second period are presented in Table IV-3. The large scatter of the data made a linear regression analysis impossible. This was caused by the presence of very fine sludge particles with poor settling properties. The amount of this sludge was not constant, because part of it washed out during the experimental period. To find the separate values of the transport factors for the two kinds of sludge present, the sludge concentrations of both the sludge particles and the fine sludge had to be calculated from the measured values of the total sludge concentrations. For this purpose, the VSS-balance could be used, as it appeared that the fine

TABLE IV-3

Distribution of sludge in the 30 m³ pilot-plant, with extra sludge from a municipal digester (second period).

v_s (sludge particles) = $8.3 \cdot 10^{-3}$ m/s (about 7.5 kg/m³ "digester sludge" present).

v_s ("digester sludge") = $0.45 \cdot 10^{-3}$ m/s.

exp. no.	ϕ_g m ³ /h	$\phi_{gb'}$ m ³ /h	F m ³ /h	C_{md} kg/m ³ (SS)	$C_{md,p}$ kg/m ³ (SS)	C_{mb} kg/m ³ (SS)	$C_{mb,p}$ kg/m ³ (SS)	k_{tb}
1	0.6	0.26	0.6	9.3+0.7	1.5	88.6+2.7	76	12.5
2	4.3	2.0	4.2	15.9+0.7	6.5	82.5+5.0	75	9.0
3	8.1	3.6	4.1	18.2+1.3	11.5	81.9+2.1	77.5	11.5
4	7.2	3.2	4.2	15.4+1.2	8.7	80.3+3.0	76.5	8.4

sludge had a VSS-value of 50 %, which differed from the value of the well settling sludge particles (80 %). For the sludge particles, the results of these calculations are presented in Table IV-3 ($C_{mb,p}$ and $C_{md,p}$); for the fine sludge, this method was too inaccurate so that for this sludge no transport factor could be calculated.

For the calculation of the transport factor k_{tb} of the sludge particles, it was assumed that the ratio of the sludge transport factors for the downward and upgoing fluid streams, k_{td}/k_{tb} , was the same as found for the sludge particles in the first period. This calculation has been performed for the four experiments for which the data have been presented in Table IV-3. The mean of the four k_{tb} -values obtained is 10.4 ± 2.3 , which accords well with the value of this transport factor, found during the first period.

IV.3c Experiments in the full-scale reactor

The results of the measurements of the distribution of sludge in the 200 m³ reactor are presented in Table IV-4. The measured data indicate that - apart from the different sludge concentrations in both the bed and the blanket - no sludge gradients were present over the height and the diameter of the reactor.

Because a large amount of mud had accumulated in the reactor during the experimental period, again no linear regression analysis of the results was possible, and the transport factors had to be calculated. Before this

TABLE IV-4

Distribution of sludge in the 200 m³ reactor.

exp.	C_{md} kg/m ³	C_{mb} kg/m ³	$C_{md,p}$ d ≥ 0.35 mm kg SS/m ³	$C_{mb,p}$ d ≥ 0.35 mm kg SS/m ³	$C_{md,f}$ d < 0.35 mm kg SS/m ³	$C_{mb,f}$ d < 0.35 mm kg SS/m ³
no.	(SS)	(SS)				
1	34.3±0.7	76.4±9.8	7.5	41	-	-
2	48.3±1.2	94.1±11	8.0	33.5	-	-
3	44.3±0.8	88.2±5.1	6.4	32	-	-
4	33.8±1.3	88.2±7.8	5.6	46	28.2	42.2
5	33.6±1.4	97.4±6.4	6.2	49	27.3	27.3
6	32.3±2.1	84.7±6.2	7.3	50	25.0	34.8
7	29.9±1.6	76.9	5.8	47	24.1	43.3

could be done, the total solids concentrations measured, had to be separated into concentrations of mud, and of sludge particles with good settling properties. For this purpose, the VSS-balance can be used (the mud consisted for 20-30 % of VSS, and the sludge particles for 80 %), as well as a sieve analysis (the fraction with $d \leq 0.35$ mm (22.3 % VSS) was mud, the fraction with $d > 0.35$ mm (80 % VSS) was the anaerobic sludge, present from the start-up of the reactor). The latter method was preferable because of its accuracy, and was therefore used. The calculated concentrations of mud ($C_{md,f}$) and of sludge ($C_{md,p}$) are presented in Table IV-4.

In order to calculate with these data the transport factor of the sludge particles ($k_{tb,p}$), it was assumed that the ratio between the transport factors found in the 30 m³ pilot-plant ($k_{td}/k_{tb} = 3.0 \pm 0.9$) could be used. For the calculation of the transport factor of the mud, $k_{tb,f}$, this assumption could not be used because this material was distributed almost uniformly over the reactor. Here it was assumed that Eq. IV-10 could be used.

The settling velocities of the mud and the sludge were determined in a different manner : because of the influence of the mud on the settling velocities of the sludge particles, at first, the relation between the settling velocity of the sludge particles and the concentration of mud was determined (see Fig. I-8). This relation could then be used to find the settling velocity belonging to the distribution of sludge and mud as measured.

For the mud, the settling velocity was determined by means of the linear fluid velocity at the boundary layer between the sludge blanket and the supernatant

TABLE IV-5

Process conditions and the results of the evaluation of the data of Table IV-4.

exp. no.	ϕ_g m^3/h	ϕ_{gb} m^3/h	F m^3/h	$v_s(p)$ $10^{-3} m/s$	$k_{tb,p}$	$k_{tb,f}/k_{td,f}$
1	29.5	4.2	26.6	1.5	20.5	-
2	12.1	1.7	21.9	0.3	12.7	-
3	16.1	2.3	29.3	0.3	4.0	-
4	37.4	5.3	35.1	1.2	5.6	0.67
5	43.8	6.2	24.5	1.4	6.5	1.0
6	32.1	4.6	26.4	2.0	16.5	0.72
7	40.6	5.8	30.4	2.4	11.5	0.56

in the settler with upward fluid flow. The mean value of this settling velocity was : $v_s = (1.83 \pm 0.2) 10^{-4} m/s$. In Table IV-5, the process conditions and the resulting calculated values of $k_{tb,p}$ and $k_{tb,f}/k_{td,f}$ are presented.

IV.4 Discussion and conclusions

The values of the transport factors for sludge particles, found in the $30 m^3$ pilot-plant and the $200 m^3$ full-scale reactor, are almost equal (see Table IV-6).

TABLE IV-6

Sludge transport factors, determined in reactors of different size.

Reactor	v_s $10^{-3} m/s$	k_{tb}	Δk_{tb}	n
laboratory	0.56	0.8	0.25	6
$30 m^3$ pilot-plant (first period)	11.6	12.8	1.8	16
$30 m^3$ pilot-plant (second period)	8.3	10.4	2.3	4
$200 m^3$: sludge particles	0.3-2.4	11.1	6.5	7

This indicates that the reactor diameter (in this comparison the most important scale-up variable), has no important influence on the transport of sludge up-

wards. The other scale-up variable, the height of the reactor, may have a more important influence on the transport of sludge, because in the laboratory reactor with a height of about 1 m, a transport factor is found that is about 15 times smaller than the factor found with the pilot-plant. The factor describing the relative volume of the wake of a rising gas bubble, k_w , is unknown in this model. The value of this factor cannot be determined, due to the unreliability of the data given in the literature (75, 76), and because the experimental data given here are not suited for this purpose. This factor will be dependent on the velocity and the size of the gas bubbles; both conditions can differ locally, but will be constant with time. It is therefore probable that k_w will be larger for a higher reactor. It has to be noted that the reactor height is not the only possible cause of the small k_{tb} value found in the laboratory reactor, since the slow settling velocity of the sludge in the laboratory reactor (see Table IV-6) can also have influenced the transport factor.

In the 30 m³ pilot-plant a transitional area with sludge concentrations higher than the concentrations in the blanket, but lower than in the bed, existed over a reactor height of about 1 m. Although this area is not present at low gas productions, it may become more important at higher gas productions. In this area the sludge concentration decreased from the value found in the sludge bed to the value found in the blanket. The model derived and described in this chapter appeared to be suited to the description of the distribution of sludge over the bed, blanket and area of transition, expressed in mean sludge concentrations. With this model, the gradient in the sludge concentration - as found in the area of transition - could not be described, due to the use of constant transport factors. This gradient would be caused by the efficiency of sludge transport upwards decreasing with the height above the bed.

The presence of the sludge gradient confirms the presence of a separate fluid flow region between the sludge bed and the blanket (as was demonstrated in Chapter III). This supports the recommendation that a distance of about 1 m should be maintained between the bottom of the settler and the top of the sludge bed.

The principal conclusions from the experiments concerning sludge behaviour were :

- The model presented for the description of the behaviour of anaerobic sludge in the upflow reactor was checked in a laboratory reactor as well as in a 30 m³ pilot-plant, and appeared to be correct.
- The sludge transport factor with an upward-going fluid stream caused by rising gas bubbles, has been determined for well settling sludge particles :

$$k_{tb} = 12.8 \pm 1.8.$$

- The sludge transport factor with a downward fluid stream has been determined for sludge particles that settled well : $k_{td} = 38 \pm 12$. For fine sludge and mud this factor could not be measured.
- For mud the ratio of the transport factors with upward and downward fluid flow was determined : $k_{tb,f}/k_{td,f} = 0.74 \pm 0.27$.
- In a reactor that contains well settling sludge (30 m^3 pilot-plant, first period), at gas flow rates above $0.7 \cdot 10^{-4} \text{ m/s}$, an area of transition is present between the bed and the blanket, in which the sludge concentration depends on the flow rate of gas and decreases with the height in the reactor.
- At scale-up in horizontal directions (compare the dimensions of the 30 m^3 and 200 m^3 reactor) no problems have been found with respect to the behaviour of the sludge. The presence of an area of transition between the bed and the blanket, in which the concentration of sludge depends on the flow rate of the gas, leads to the recommendation that a distance of at least 1 m should be present between the top of the bed and the bottom of the settler.

THE SETTLER

V.1 Introduction

Due to the slow growth of the sludge, washout can easily result in a decrease of the amount of sludge in the reactor. In order to prevent the washout exceeding the growth, the efficiency of the separation and recirculation of the sludge must be very good. For example, if the yield factor $Y = 0.05$, the concentration of substrate in the influent $C_{so} = 2.0 \text{ kg/m}^3$, and the concentration of sludge in the blanket $C_{md} = 10.0 \text{ kg/m}^3$, it can be calculated that the efficiency must be at least 99 %.

Because methane forming sludge has good settling properties if it is well adapted to the substrate, a settler designed for a good efficiency of say 99 %, can have dimensions such that it can be placed on top of the upflow reactor. This has the advantage that no pumps are needed to recirculate the sludge from the settler back into the reactor, and that no extra place is required.

In the literature, two settler types are described (79, 86) :

- settlers with horizontal fluid flow (horizontal settlers);
- settlers with mainly vertical fluid flow (vertical settlers);

Vertical settlers are often used with municipal digesters (80, 81); however, because the liquid residence time in this type of settler is usually more than a factor of 100 larger than in the settler of the upflow reactor, the data presented referring to their operation, cannot be used for our study. Settlers of the horizontal type, with short liquid residence times, are in operation with the aerobic active sludge process (81-85); with tertiary treatment of municipal wastewater; and for other inorganic flocculated suspensions. Unfortunately, the settling properties of such suspensions cannot be compared with those of anaerobic sludge particles. Therefore, these results are also not very relevant for this study. Settlers, in particular used for the effluent of the anaerobic process for wastewater treatment, have not been described in the literature (87).

With a settler built on top of the upflow reactor, high separation efficiencies can be attained, if the following requirements are fulfilled :

- Turbulent fluid flow should not occur in the settling compartment. Therefore, the settler and the reactor part in which the purification process takes place must be separated, so that the gas production and the turbulence of the blanket do not disturb the fluid flow in the settler. Before the

suspension enters the settling compartment, the dissolved gas that liberates by expansion effects at the top of the reactor, has to be intercepted. Furthermore, to prevent the production of gas in the settler, the retention time of the sludge in the settler must be short, and the purification efficiency of the reactor must be high.

- Because anaerobic sludge particles have cluster forming properties (see Chapter I), the presence of a sludge blanket in the settler is beneficial. But the sludge of course may not accumulate in the settler, and to prevent this, the recirculation of sludge has to function properly.
- The plug flow in the settler compartment must be uniform; in other words, bypassing streams and dead spaces must be absent. To obtain this, the supply of settler influent, the discharge of effluent and the recirculation of sludge have to take place continuously and uniformly.

The settler of the laboratory reactor was too small to give relevant experimental data. The settlers, chosen for the pilot-plants and the full-scale reactor, were designed in such a way that most of the above formulated requirements could be realised (see Fig. V-1). Therefore, the settler's inflow, the effluent, and the recirculation stream, have been separated. Furthermore, the

Fig. V-1. Schematic generalised picture of the cross-section of settlers used in the pilot-plants and the full-scale reactor. At the left side of this picture, the flow patterns of the fluid and sludge have been shown. At the right side, it has been tried to indicate the presence and distribution of sludge (S), gas (G) and liquid (L) in the compartments in and below the settler.

settler construction was such that, at first, a separation of gas from the mixture of gas (G), liquid (L) and solids (S) was obtained; after that, the suspension of solids in the liquid entered a compartment where dissolved gas could escape by expansion and where the turbulence that was generated in the blanket, was damped. Finally, the sludge and the effluent could be separated in the settling compartment at the bottom

of which a sludge blanket could be present that continued into the compartments for expansion and recirculation. From this last compartment, the thickened sludge suspension (S) could fall into the reactor. Because it has been found that if the bottom plates each are placed under an angle of 45° or steeper, the sludge particles slide or roll slowly downwards, this angle was applied to the bottom plates of all the settlers that were studied. The procedure described here can be formulated as follows :

$$(G + L + S) \rightarrow G \uparrow + (L + S) \rightarrow S \downarrow + L$$

For a continuous supply of the sludge suspension from the blanket in the reactor, and recirculation of the thickened sludge suspension back to the blanket, a circulation stream is used that is generated by the difference of the densities of the thickened suspension in the settler and the three-phases suspension in the blanket. The flow rate of this stream depends mainly on the flow rate of gas in the reactor. The application of this circulation stream in a settler, has been patented by C.S.M. (36). For scaling-up purposes, the settlers of the 30 m^3 pilot-plant and the full-scale 200 m^3 reactor, were rectangular. In the experiments with the settlers of the 6 m^3 , 30 m^3 and 200 m^3 reactor, the fluid flow pattern, the distribution of sludge, and the efficiency of the separation and recirculation of sludge, were determined for various process conditions.

V.2 Fluid flow pattern in the settler

V.2a Methods

For an overall analysis of the fluid flow pattern in the settler, F -experiments can be used. The time interval between the dimensionless concentration-time curve measured in the blanket, and the F -curve (measured in the effluent), indicates what percentage of the reactor can be described as a plug flow region. This percentage can then be compared with the actual volume of the settling compartment in which laminar fluid flow would occur. Furthermore, such experiments give an estimation of the dead space regions present in both the reactor and the settler (via Eq. III-6).

A more detailed description of the flow pattern is obtained by frequent local sampling after a pulse injection of tracer into the influent of the reactor. With such an experiment, a quickly changing tracer signal is measured on several locations in the settler. With the use of automatic time controlled sampling pumps, the total duration of each experiment could be restricted

TABLE V-1

Conditions of the tracer experiments to study the fluid flow pattern in the settler.

injection	reactor	[LiCl] kg/m ³	injection time min.	injection flow rate m ³ /h
pulse	30 m ³	56.7	1.5	1.2
pulse	200 m ³	200.0	8	0.375

to about 0.5 h, as the most significant changes in the tracer concentration occur in that time interval. See Table V-1 for the experimental conditions.

V.2b Results and discussion

The efficiency of the first of the two settlers, applied in the 30 m³ reactor (see Fig. I-13), was insufficient. The circulation stream was present in this settler, but it was so turbulent that the settling of sludge was disturbed. It was concluded that the compartment of the settler which was occupied by the circulation stream, had to be separated from the actual settling compartment. In the second settler used in the 30 m³ reactor (see Fig. I-14), the construction of which was based on this conclusion, the separation of these two compartments appeared to prevent turbulence in the settling compartment sufficiently.

The residence time distribution experiments 9 and 10 (see Chapter III), performed in the 30 m³ pilot-plant (on top of which the second settler had been placed), indicated that the settling compartment in which plug flow was present, occupied about 5 % of the reactor volume. Dead spaces were not present during these experiments. Because the total settler volume was about 17 % of the reactor volume, a large part of the settler (~ 70 % or ~ 12 % of the total reactor volume) must have been well mixed. The same conclusion can be drawn from the results of experiment S1 (a pulse tracer experiment, see Table V-2). During this experiment, it appeared that the sample points in the inflow opening in the settler, and in the recirculation opening in the bottom, gave identical signals almost simultaneously (this moment has been defined as $\theta = 0$). In the effluent and 0.4 m below the water level in the settler, the tracer signal arrived at $\theta = 0.06$. From these and other data in the table, it can be calculated that a plug flow region of 5 % of V_R must have been present between the

TABLE V-2

Process conditions during, and results of the tracer experiments in the settlers of the 30 m³ pilot-plant and the 200 m³ reactor.

(c) : circulation flow; (u) upward fluid flow.

	experiment no.				
	S-1	S-2	S-3	S-4(c)	S-4(u)
reactor	30 m ³	200 m ³	200 m ³	200 m ³	
date	March 1977	Oct.1977	Nov.1977	Dec.1977	
ϕ_g (m ³ /h)	2.7	24.3	26.3	43.9	43.9
ϕ_g'' (m/h)	0.51	0.55	0.6	1.0	1.0
F (m ³ /h)	4.1	26.7	21.5	26.3	26.3
v_L (settler) (m/h)	0.72	0.61	0.50	0.60	0.60
v_s (m/h)	~20	5.4	0.5	0.65	0.65
C_{md} (kg SS/m ³)	~ 8	16.6	41.5	32.3	32.3
$C_{m,e}$ (kg SS/m ³)	~ 0.3	0.084	0.21	0.29	0.29
eff. settler (%)	~96	99.5	99.5	99.1	99.1
θ - infl.settler	0	0	0	0	0
θ - effluent	0.06	0.15	0.05	0.10	0.10
θ - recirc. opening	0	0.007	0.045	0.007	~
θ - on depth 0.8 m	0.01	0.085	0.02	0.013	0.02
θ - " " 0.6 m	0.01	-	0.02	0.035	~
θ - " " 0.4 m	0.06	0.15	0.02	-	0.04

effluent and the sample points on a depth of 0.6 - 0.8 m. From the differences measured at the sample points on different depths, it was concluded that the fluid velocity in the settling compartment was not completely uniform. Nevertheless, from the effluent sludge concentration ($C_{m,e}$) and the efficiency of the settler, it could be concluded that this settler operated properly. During the first and second month that the 200 m³ reactor operated (see Chapter I), a circulation stream was present in the two equal settlers that had been constructed on top of the reactor. In this period, the pulse tracer experiments S2 and S3 were performed. During the third month, experiment S4 was performed with one "circulation flow settler" S4(c), and one "upward fluid flow settler" S4 (u). This was achieved by blocking the inflow openings of one of the "circulation flow settlers", so that both the inflow and recirculation took place through the slit in the bottom. The results of the experiments S2,

S3, and S4, are presented in Table V-2, together with the most important process conditions. In Table V-4, the distribution of the sludge over the settlers during these experiments has been presented.

The results show the influence of the amount of sludge in the settler (compare the results of S2 and S3), and of the flow rate of gas (compare S2/S3 - S4 (c)), on the fluid flow pattern, especially on the size of the plug flow region and the region in which the fluid flow is not defined. From experiment S4(u) the flow pattern in the settler with upward fluid flow can be derived and this result can be compared with that of experiment S4(c).

The volume of the plug flow region is derived from the interval $\Delta\theta$ between the curves measured in the settler's inflow and the effluent: $V_{pf} = \Delta\theta V_R$. This interval is measured at the value $C/C_0 = 0.5$, which was chosen because at that value the curves are mostly the steepest, so that the measurement is there the most accurate.

It is possible to specify this plug flow region in more detail. To do this, it has to be assumed that all the liquid that enters the settling compartments, leaves them as effluent, and is therefore equal to the influent flow rate F . The time interval between the tracer signal at a depth of 0.8 m, where the settling compartments begin, and the signal in the effluent then gives the plug flow region present in the upper compartments of the settler. Subtraction of this value from the total plug flow region in the settler, gives the plug flow region in the compartments below a depth of 0.8 m. By means of the actual volumes of the upper and lower compartments (respectively 18.0 and 8.7 m³ in both settlers of the 200 m³ reactor), the volumes with non-defined fluid flow (at this point it is unknown as to whether or not it is well mixed or dead space) can then be calculated. The results of these analyses have been presented in Table V-3.

From the comparison of the results of the experiments S2 and S3, it appears that the increase of the amount of sludge results in a decrease in volume of the plug flow regions, especially in the plug flow regions in the lower settler compartments. Because the recirculation opening appeared to be partly blocked, this decrease would probably have been caused by the development of dead space regions. The most noticeable difference between the process conditions of the experiments S3 and S4(c), appears in the gas production. Apparently, an increase of the flow rate of gas has positive effects on the volumes of flowing liquid in the settlers, as the plug flow region has increased, and the blockage of the recirculation opening disappeared. This last effect is also a result of the lower concentration of sludge in the blanket. The main effect of the increased gas flow rate is that the circula-

TABLE V-3

Results of the evaluation of the data of the tracer experiments in the 200 m³ reactor-settler.

	experiment no.			
	S-2	S-3	S-4(c)	S-4(u)
V_{st} (m ³)	26.7	26.7	26.7	26.7
V_{pf} (total) (m ³)	15.3	5.1	10.2	10.2
V_{nd} (total) (m ³)	11.4	21.6	16.5	16.5
V_{pf} (upper comp.) (m ³)	6.6	3.0	8.9	8.2
V_{pf} (lower ") "	8.7	2.1	1.3	2.0
V_{nd} (upper ") "	11.4	15.0	9.1	9.8
V_{nd} (lower ") "	0	6.6	7.4	6.7

tion stream is also larger. This stream mixes the mayor part of the volume of non-defined fluid flow in the lower settler compartments, and this results in short local sludge retention times.

As follows from Table V-3, the settler with upward fluid flow (exp. S4(u)), gives similar results. However, there is a difference with the settler with circulation flow; that is, in the region of non-defined fluid flow in the "upward settler", circulation flow is absent, so that mixing is almost impossible. Most of this region must have been dead space.

Since the pulse tracer experiments and the residence time distribution experiments in the 200 m³ reactor were performed in the same period, and thus under the same process conditions (compare Tables III-7 and V-2), the results are complementary. If they are compared, it appears that the model descriptions obtained differently, conform well. The absence of dead space in model 12 (see Table III-6), indicates that the region of non-defined fluid flow during experiment S2, was most probably a well mixed region. The finding of dead space in model 13, and to a lesser degree also in model 14, confirms that during experiment S3, the small plug flow region was a result of dead space present in the settler. This effect is caused by the amount of sludge that accumulated in the settler.

V.3 Distribution of sludge in the settler

From the distribution of sludge in the settler, the efficiencies of the separation and the recirculation of the sludge can be derived. The most important parameters that influence the sludge distribution, are : the sludge concentration in the blanket (C_{md}); the flow rate of the circulation stream; the settling velocity of the sludge (v_s); and the local linear fluid velocity in the settling compartment (v_L). The parameters C_{md} , v_s , and v_L , can be measured quite easily. Quantitative determination of the flow rate of the circulation stream has not been successful up to now.

The measurements were performed by local sampling of the settler, the reactor, and the effluent, during periods with constant process conditions. The samples were analysed for the concentration of suspended solids. Some results of these measurements in the settlers in both pilot-plants, and in the 200 m³ reactor, are presented in Table V-4. In Fig. V-2 the effluent sludge concentration of

TABLE V-4

Distribution of sludge in the settlers of the pilot-plants and the full-scale reactor.

date	April 13 1976	March 29 1977	April 13 1977	Oct. 17 1977	Nov. 1 1977	Dec. 7 1977
reactor	6 m ³	30 m ³	30 m ³	200 m ³	200 m ³	200 m ³
ϕ_g (m/h)	0.6	0.4	0.86	0.55	0.6	1.0
v_L (settler) (m/h)	1.1	0.66	0.41	0.61	0.5	0.6
v_s (m/h)	7.2	21.6	36	5.4	0.5	0.65
C_{mo} (kg SS/m ³)	~0.2	0.33	0.30	0.31	0.86	0.12
C_{md} (kg SS/m ³)	15.9	7.9	15.3	16.6	41.5	33.0
C_m (depth 1.7 m)	-	-	-	19.5	49.0	~38
C_m " 1.3 m	-	8.6	2.7	-	-	-
C_m " 1.1 m	-	-	-	16.1	34.7	~35
C_m " 0.8 m	20.7	4.8	1.3	0.7	34.0	~35
C_m " 0.4 m	14.4	4.2	0.4	0.5	33.0	~30
C_m " 0.2 m	13.7	3.5	0.3	-	0.9	~1
C_m " 0.1 m	12.0	1.3	0.03	-	-	-
C_m " 0.05 m	11.8	-	-	-	-	-
$C_{m,e}$ (kg SS/m ³)	0.13	0.33	0.16	0.084	0.21	0.38
eff.settler (%)	99	96	99	99.5	99.5	98.8

Fig. V-2. The concentration of sludge in the effluent of the 200 m³ reactor as a function of the linear fluid velocity.

the 200 m³ reactor is plotted as a function of the mean linear fluid velocity at the top of the settler. This figure shows that washout of sludge is only to a small degree, dependent on the linear velocity. From the data in the table it follows that the settler efficiency is neither sensitive to changes of the settling velocity, nor to the concentration of sludge in the blanket. Both parameters varied strongly for the different experiments. From the high efficiencies obtained, it can again be concluded that the settlers used, functioned very well. However, sometimes process operation had to be adjusted to maintain this situation.

Overloading of the settler can occur if the linear velocity at the top of the settler exceeds the settling velocity of the sludge. This happened temporarily in the 200 m³ reactor (November 1977). Another cause of overloading can be that the small recirculation of sludge is exceeded by the supply of sludge in the settler. This happened in the 6 m³ pilot-plant when the recirculation opening in the bottom had a diameter of only 0.06 m. Washout of sludge was prevented in such a situation, by stopping the process and waiting until the sludge that had accumulated in the settler, had fallen back into the reactor. The experiments in the 30 m³ pilot-plant, showed no thickening of sludge in the lower settler parts, at gas flow rates of 0.86 m/h; while at 0.4 m/h, the sludge concentration at the bottom of the settler was slightly higher than that in the blanket. Although the settling properties may have had some influence, it is thought that the small flow rate of gas, giving a small recirculation flow rate, indirectly produces this thickening of the sludge. This is confirmed by the results of the October-measurements in the 200 m³ reactor, that correspond with those of the measurements in the 6 m³ pilot-plant. The November and December-measurements in the 200 m³ reactor gave different

results because of the accumulation of mud.

Comparison of the results obtained by the measurements in the two settlers with different fluid flow patterns (upwards, circulation) in the 200 m^3 reactor, led to the conclusion that both settler types functioned properly under what appeared to be difficult circumstances.

V.4 Conclusions and recommendations

In those situations that the sludge concentration in the blanket and accordingly also in the influent of the settler, was high, settler operation appeared to be difficult : in a considerable part of the settler no fluid flow occurred while in other parts preference flow occurred, causing washout of sludge. These problems do not show in the data presented, because as soon as increased wash-out was observed, process operation was adjusted. Because such blockage and preference flow occurred at $C_{\text{md}} \geq 40 \text{ kg SS/m}^3$, and it may be assumed that the higher C_{md} the stronger these effects, it has been concluded that the critical or maximum value of C_{md} to be applied is 40 kg SS/m^3 .

From the foregoing sections it can be concluded that usually with these settlers the high standards can be reached that are required to maintain a steady process in the upflow reactor. At the end of the experimental period it appeared that the average settler efficiency even had been better than was strictly necessary, since the amount of sludge in the reactor had increased (see Chapter I). The dimensions of the settlers used, were for the 6 m^3 pilot-plant : diameter = 1.34 m, depth = 0.7 m; 30 m^3 pilot-plant (second settler) : length = width = 2.6 m, depth = 1.6 m; 200 m^3 reactor (two equal settlers) : length = 7.5 m, width = 3.5 m, depth = 1.9 m. Because no scale-up limitations were met, it can be concluded that the length, width and depth of a settler as such are not the most important parameters in scaling-up.

One of the main causes of the successful operation of the settlers studied, is the application of a circulation stream for a steady supply of settler influent and for recirculation of sludge, for which last function, the angles of 45° of the bottom plates are of prime importance. Another important factor may be the separation of the settling compartment from the recirculation and expansion compartments, that, with the steady recirculation of sludge, results in a sludge free zone in the upper layers of the settling compartment. It can thus be recommended to use the type of settler that was studied, and to adjust the dimensions of such a settler to those of the reactor. In case that the depth of the settler would be too large so that too much reactor volume is occupied by the settler, more than one settler can be used. Recommended dimensions of such settlers are : width 2 - 4 m, depth 1 - 2 m, length to be

chosen in accordance with the dimensions of the reactor.

As follows from the conclusions of Chapter VI, in the upflow reactors studied, higher loadings than those used up to now can be applied. In that case the volume of gas produced will also be much higher. In such a situation, the measures to separate the gas from the suspension of solids, that have been applied in the settlers studied, may be insufficient. Moreover, at a high flow rate of gas below the settler, expansion effects, floating sludge and foam can become serious problems. Both for the settlers with either upward fluid flow or circulation flow, that have been described in this chapter, a solution for these problems has to be found. The measures proposed are : to separate the settling compartment better from the rest of the settler; to prevent foam, floating sludge and gas to escape and accumulate at the top of the settler; and to generate an extra circulation stream near the recirculation opening of the

Fig. V-3. Outline of : cross-section A-A (VI-3a); top-view (VI-3b); cross-section B-B (VI-3c); of a settler to be built-in on top of an upflow reactor for anaerobic treatment of wastewater.

settler, by which the sludge transport back into the reactor is stimulated. A proposal for a settler, set-up along these lines, is given in Fig. V-3. In this apparatus, the extra circulation stream flows around the gas shield below the settler. This shield, together with the roof-construction inside the settler and the small gas shields below the inflow openings, keeps the gas bubbles in the reactor away from the settling compartment. The construction inside the settler furthermore prevents foam, floating sludge, turbulent fluid flow and gas that liberates by expansion, to disturb the laminar fluid flow in the settling compartment. Via an opening in the side of the settler, this gas, foam and floating sludge can return to the reactor.

CHAPTER VI

MODEL FOR THE CONVERSION OF FATTY ACIDS IN THE UPFLOW REACTOR

VI.1 Model

The model for the conversion of fatty acids in the methane forming step of the anaerobic process for wastewater treatment in the upflow reactor, is obtained by the combination of the three parts already studied :

- the fluid flow pattern in the upflow reactor;
- the behaviour of the sludge;
- the kinetics of the conversion of fatty acids into several products.

Using this model, the conversion of substrate can be calculated; and because the independent parameters of the model can be varied, it can also be used to predict which process conditions, and what process performance, have to be applied for an optimal operation of the reactor, provided that the variations of the parameters are kept within the domain applied in the experiments. More specifically, the model can be used to predict how much sludge should be maintained in the reactor; how this sludge should be distributed over the height of the reactor; and what dimensions the reactor should have, so that optimum process operation is guaranteed for given compositions, concentrations, and flow rates of the influent.

The following assumptions are made :

- the substrate is a solution of neutralised fatty acids; it is not present in the settler, and is not converted in the bypassing channels from the influent openings to the sludge blanket;
- the amounts of bacterial waste products, produced by the conversion of the fatty acids in the exchange streams between the bed and the blanket, are negligible as compared with the amounts produced by direct conversion of the influent fatty acids;
- both the growth of sludge and the washout are neglected.

The model is based on the fluid flow model that is presented in Fig. I-4, and quantified in the fluid flow experiments in Chapter III. This model is extended with the kinetics and the sludge behaviour as described in the Chapters II and IV. The combination obtained is presented in the following equations for the conversion of substrate, the production of gas, and the production of bacterial waste products :

Conversion of substrate :

$$V_b \frac{dC_{fa,b}}{dt} = F_o C_{so} + F_{db} C_{fa,d} - F_{bd} C_{fa,b} - q_b C_{mb} V_b \quad \text{VI-1}$$

$$V_d \frac{dC_{fa,d}}{dt} = F_k C_{so} + F_{bd} C_{fa,b} - F_{db} C_{fa,d} - F C_{fa,d} - q_d C_{md} V_d \quad \text{VI-2}$$

Gas production :

$$\phi_{gb} = \frac{r_{me,b} V_b}{C_{me}} \quad \text{VI-3}$$

$$\phi_{gb'} = \phi_{gb} \frac{10}{10 + h_d + h_{st}} \quad \text{VI-4}$$

$$\phi_{gd} = \frac{r_{me,d} V_d}{C_{me}} \quad \text{VI-5}$$

$$\phi_g = \phi_{gb} + \phi_{gd} \quad \text{VI-6}$$

Bacterial waste products :

$$V_b \frac{dC_{ap,b}}{dt} = F_{db} C_{ap,d} - F_{bd} C_{ap,b} + r_{ap,b} V_b \quad \text{VI-7}$$

$$V_d \frac{dC_{ap,d}}{dt} = F_{bd} C_{ap,b} - F_{db} C_{ap,d} - F C_{ap,d} + r_{ap,d} V_d \quad \text{VI-8}$$

The values of the parameters, needed to solve the preceding equations, can be found in the Chapters II, III, and IV.

From the kinetics (Chapter II), the following relations are obtained :

$$q_b = q_{\max,fa} \frac{C_{fa,b}}{K_{fa} + C_{fa,b}} \quad \text{II-17}$$

$$q_d = q_{\max,fa} \frac{C_{fa,d}}{K_{fa} + C_{fa,d}} \quad \text{II-17}$$

$$r_{me,b} = me_{\max,fa} \frac{C_{fa,b} C_{mb}}{K_{me} + C_{fa,b}} \quad \text{II-20}$$

$$r_{me,d} = me_{\max,fa} \frac{C_{fa,d} C_{md}}{K_{me} + C_{fa,d}} \quad \text{II-20}$$

$$r_{ap,b} = k_{ap} (C_{so} - C_{fa,b}) \quad \text{II-22}$$

$$r_{ap,d} = k_{ap} (C_{so} - C_{fa,d}) \quad \text{II-22}$$

For the fluid flow pattern, the following equations should be used :

$$V_R = V_b + V_d + V_{pf} + V_{ds} \quad \text{III-5}$$

$$F = F_o + F_k \quad \text{III-3}$$

$$F_{bd} = F_o + F_{db} \quad \text{III-4}$$

From Chapter III it follows furthermore that the magnitude of the bypassing stream is related to the height of the sludge bed. This relation can be estimated by means of the results of the residence time distribution experiments in the 30 m³ pilot-plant, performed at different heights of the sludge bed.

Linear interpolation between the F_k -values gives :

$$F_k = (a_1 + a_2 h_b) F \quad \text{VI-9}$$

in which the constants are $a_1 = 0.126$ and $a_2 = 0.17 \text{ m}^{-1}$.

From the laboratory experiment with the tracer injection in the blanket (also described in Chapter III), the value of the constant a_3 in the relation of the backmix stream to the volume of gas per unit of time at the top of the bed, has been obtained. In this relation :

$$F_{db} = a_3 \phi_{gb} \quad \text{VI-10}$$

the constant $a_3 = 1.0$.

The results of the measurements of the distribution of sludge (Chapter IV) give the equation for the concentration of sludge in the blanket, that is based on a constant value of the sludge concentration in the bed :

$$C_{md} = \frac{\phi_{gb} k_{tb} C_{mb}}{\phi_{gb} k_{td} + (v_s - v_L) A_R} \quad \text{IV-9}$$

By means of the set of equations given, using the values of the parameters and constants that were already found (see Table VI-1), the quantitative values of the following parameters are calculated : the fatty acids concentration, the production rate of gas, and the concentration of waste products. All these values are calculated, both in the bed and in the blanket. With these data, the effluent concentrations of fatty acids and waste products, and the total volume of gas produced per unit of time can be calculated.

TABLE VI-1

a. List of the parameters of the substrate model and their origin.

Fluid flow model	$v_b, v_d, v_{pf}, v_{ds}; F_k, F_{db}$
Sludge model	C_{md}
Kinetics	$C_{fa,b}, C_{fa,d}, \phi_{gb}, \phi_{gd}, \phi_g, C_{ap,b}, C_{ap,d}$
Experimental conditions	v_R, F, C_{so}

b. List of the constants of the substrate model and their origin.

Fluid flow model	$a_1 = 0.126; a_2 = 0.17 \text{ m}^{-1}; a_3 = 1.0$
Sludge model	$k_{tb} = 12.8 \pm 1.8; k_{td} = (3.0 \pm 0.9) k_{tb}$
Kinetics	$q_{\max,fa} = (4.7 \pm 0.8) 10^{-5} \text{ s}^{-1};$ $K_{fa} = (38 \pm 3) 10^{-3} \text{ kg C/m}^3$ $m_{\max,fa} = (2.5 \pm 0.35) 10^{-5} \text{ s}^{-1}$ $K_{me} = (41 \pm 3) 10^{-3} \text{ kg C/m}^3$ $k_{ap} = (5.5 \pm 0.5) 10^{-6} \text{ s}^{-1}$
Experimental conditions	sludge : %C, %VSS, $v_s, C_{mb};$ $C_{me}; v_L; h_{st};$ % acidification influent.

As has already been stated, the model can also be used to optimise the process. For a good process performance, the effluent concentration of fatty acids and the linear fluid velocity have to remain below certain values. In the optimisation procedure these values have to be specified; and because the height of the internal settler is fixed, it is then possible to calculate the dimensions of the reactor as a function of the influent flow and concentration.

If for a certain wastewater, the two last mentioned variables are known, together with the effluent concentration chosen, it can be calculated how much substrate has to be converted per unit of time. By means of the substrate model, this amount of substrate is brought into relation with an amount of sludge that is minimally needed to obtain the desired conversion as well as the desired effluent concentration. Furthermore, the distribution of this sludge over the bed and the blanket is determined together with the distribution of the substrate over both reactor parts. As a result of these calculations, the minimally required volumes of the bed and the blanket are obtained. Because the bottom surface of the reactor is fixed by the flow rate of the wastewater stream and the chosen value of the linear fluid velocity, the minimally required heights

Fig. VI-1. Block diagram 1: the methane forming step of the anaerobic process for wastewater treatment in the upflow reactor.

of the bed and the blanket are found in combination with the required volumes. Together with the height of the settler, the total reactor height is obtained. The eight basic equations of the substrate model (VI-1 - VI-8) and all the other equations to be used, produce a complex network of relations. These relations have been collected in block diagram 1, presented in Fig. VI-1. It has been derived for a stationary situation; therefore, all concentrations are independent of time.

VI.2 Test of the model

To test the model, the results of the computations of C_{fa} , C_{ap} , and ϕ_g , have to be compared with the values measured in a reactor in which the process

TABLE VI-2

Values of the parameters and constants, used for the test of the substrate model in the laboratory reactor.

General : Reactor dimensions see Table I-3.

$C_{mb} = 20.0 \text{ kg C/m}^3$
 $q_{\max,fa} = 4.7 \cdot 10^{-5} \text{ s}^{-1}$
 $K_{fa} = 38 \cdot 10^{-3} \text{ kg C/m}^3$
 $k_{ap} = 5.5 \cdot 10^{-6} \text{ s}^{-1}$

$m_{e\max,fa} = 2.5 \cdot 10^{-5} \text{ s}^{-1}$
 $K_{me} = 41 \cdot 10^{-3} \text{ kg C/m}^3$

Substrate : 100 % fatty acids, composition fatty acids:
see Table I-2.

Exp. no.	V_b 10^{-3} m^3	V_d 10^{-3} m^3	C_{md} kg/m^3 (C)	C_{so} kg/m^3 (C)	F 10^{-6} m^3/s	F_k 10^{-6} m^3/s	q d^{-1}
1	30.0	20.6	0.6	0.61	2.86	0.14	0.25
2	22.5	17.8	0.6	0.54	12.7	2.58	1.16
3	22.5	17.8	0.6	0.54	12.9	2.58	1.10
4	22.5	17.8	0.55	0.81	7.67	1.53	1.03
5	22.5	17.8	0.75	0.94	7.67	1.53	1.17
6	11.1	29.2	0.5	1.4	1.92	1.15	0.82
7	12.5	27.8	0.5	1.57	2.0	1.2	0.85
8	11.1	29.2	0.4	0.82	6.5	3.9	2.24
9	3.8	36.5	0.55	1.24	6.67	5.34	3.30
10	3.8	36.5	0.55	3.25	2.08	1.66	2.60
11	3.8	36.5	0.55	3.28	2.13	1.7	2.75

conditions have been varied. In the first instance, the test has been performed with the laboratory reactor. To determine the sludge model in this reactor, only a few data points were available, giving a moderate correlation (see Chapter IV). Therefore, with respect to the distribution of sludge, independent concentration measurements were performed. The data of these and the other measurements necessary to calculate the substrate model, have been collected in Table VI-2. The computations have been performed by means of a digital computer (88).

Because of a spread in the values of the parameters and constants used for the computations (mostly indicated by means of a 90 % reliability interval), their deviations from the mean values were taken up in the calculations. In the Figs. VI-2a, b and c, the computed and measured results are graphically presented.

Fig. VI-2a. Laboratory test of the model presented in figure VI-1, for the effluent concentration of fatty acids, (XXXX) measured; ($\square$) calculated.

From these figures it can be concluded that the result of the test of the substrate model was positive. The measured fatty acids concentrations all lay within the calculated intervals, and mostly close to the mean value. The calculated concentrations of the bacterial waste products, agreed sufficiently with the measured values. In this case, the calculated intervals were smaller than the measured intervals, which was caused by the fact that the inaccuracy of the measurements was large, due to the results being obtained by two large values subtracted from each other ($C_{ap,e} = C_{s,e} - C_{fa,e}$); the calculations, on the other hand, were based on a kinetical constant determined via the mean values. The computed gas productions also agree well with the measurements.

Fig. VI-2b. Laboratory test for the effluent concentration of bacterial waste products.

Fig. VI-2c. Laboratory test for the total methane production.

TABLE VI-3

Laboratory test of the model presented in figure VI-1, for the conversion of substrate (TOC).

Experiment No.	Predicted Values		Measured Values	
	% conversion		% conversion	
	Interval	Mean	Mean	Interval
1	92.1-85.3	89.9	91.7	93.3-90.7
2	90.9-78.9	85.2	88.2	90.5-86.3
3	90.9-78.9	85.2	85.8	88.6-83.5
4	90.8-82.8	86.6	83.2	86.4-80.8
5	92.6-82.7	87.9	83.9	87.0-81.4
6	83.9-71.5	74.2	74.8	78.6-71.0
7	83.7-65.2	74.5	73.9	78.9-70.0
8	78.6-54.8	65.5	65.9	73.8-54.8
9	51.5-19.2	33.2	40.8	52.0-31.9
10	65.1-33.7	46.6	38.9	50.5-30.0
11	63.2-30.5	43.3	38.0	50.0-28.7

For the experiments 9, 10, and 11, this comparison could not be made, due to a failing gasmeter.

Finally, in Table VI-3 the calculated and measured values of the conversion, based on TOC, are presented. Because they are in very close agreement for a wide range of process conditions, it can be concluded that the model can be used to predict the process performance in the laboratory reactor.

The computed concentrations of fatty acids and waste products in the effluent, flow rate of gas, and conversion of substrate, have also been checked for the 30 m³ pilot-plant. The model was constructed with the assumption that the reactor could be described as a combination of only two ideal mixers, despite the fact that in the models of the fluid flow pattern in the pilot-plants three ideal mixers have been found : the bed, the blanket, and the area of transition (see models 6 - 10, Chapter III).

TABLE VI-4

Values of the parameters and constants, used for the test of the substrate model in the 30 m³ pilot-plant.

General : Reactor dimensions see Table I-5.

$$C_{mb} = 34.7 \text{ kg C/m}^3$$

$$k_{tb} = 12.8$$

$$k_{td} = 38.4$$

$$\text{Exps. 1 - 6 : } v_s = 11.6 \cdot 10^{-3} \text{ m/s; } f_{me} = 0.83$$

$$\text{Exps. 7, 8, 9 : } v_s = 7.8 \cdot 10^{-3} \text{ m/s; } f_{me} = 0.675.$$

$$q_{\max, fa} = 4.7 \cdot 10^{-5} \text{ s}^{-1}; \quad m_{e, \max, fa} = 2.5 \cdot 10^{-5} \text{ s}^{-1}$$

$$K_{fa} = 38 \cdot 10^{-3} \text{ kg C/m}^3 \quad K_{me} = 41 \cdot 10^{-3} \text{ kg C/m}^3$$

$$k_{ap} = 5.5 \cdot 10^{-6} \text{ s}^{-1}$$

composition influent : see Table I-6

Exp. no.	V_b (m ³)	V_d (m ³)	C_{so} (kg C/m ³)	F (10 ⁻³ m ³ /s)	F_{db} (10 ⁻³ m ³ /s)
1	6.5	21.0	0.33	3.6	6.1
2	6.5	21.0	0.74	2.1	3.6
3	6.5	21.0	0.38	2.8	4.8
4	6.5	21.0	0.48	2.2	3.7
5	6.5	21.0	0.53	2.3	3.9
6	6.5	21.0	0.73	2.5	4.3
7	6.5	21.0	0.55	2.2	3.8
8	6.5	21.0	0.74	1.7	2.8
9	11.9	15.7	1.12	0.6	1.0
10	11.9	15.1	0.54	1.0	1.6
11	11.9	15.1	1.17	0.7	1.1

Unfortunately it was impossible to test the model for the 200 m³ full-scale reactor as it contained an insufficient amount of sludge and very much mud during the campaign of 1977.

As has been described in Chapter I, the influent composition of the pilot-plant was not as accurately known as that of the laboratory reactor (see Table I-6). From this table it appears for instance, that the dissolved substrate consisted only partly of fatty acids, and that furthermore, also an amount of suspended not dissolved matter could be present in the influent. Whether all the products in the influent, that are no fatty acids, can be anaerobically converted, is still unknown.

In Table VI-4, the values of the parameters to be used in the model, have been presented. They lay in a narrow range, as it was of prime importance that this reactor operated efficiently, and was not overloaded (which was contrary to the situation with the laboratory reactor). Therefore, the range in which the model for the 30 m³ pilot-plant could be tested, was restricted to those circumstances under which the purification efficiency was reasonable to good. The first eight tests of the model were performed with data obtained after the period of start-up, in the campaign of 1976; the last three, with data collected during the intercampaign of 1977. The results obtained have been presented in the Figures VI-3a, b, and c, and in Table VI-5. They are based on the TOC-

Fig. VI-3a. Pilot-plant test of the model presented in figure VI-1, for the effluent concentration of fatty acids, (hatched) measured; (white) calculated.

Fig. VI-3b. Pilot plant test for the effluent concentration of bacterial waste products.

Fig. VI-3c. Pilot plant test for the total production of gas.

TABLE VI-5

Pilot-plant tests of the models presented in the Figures VI-1 and VI-5, for the conversion of substrate (TOC).

Exp. no.	Model (Fig. VI-1)		Measured Values		Simplified Model	
	Predicted Values		% conversion		Predicted Values	
	% conversion		% conversion		% conversion	
	Interval	Mean	Mean	Interval	Interval	Mean
1	95.0-88.6	93.3	71.2	80.6-60.0		
2	94.2-89.5	91.7	68.9	79.1-56.8		
3	95.8-86.3	92.0	76.1	87.5-60.0	96.2-83.0	92.3
4	95.5-83.9	90.6	78.2	88.3-62.5	95.7-86.2	90.6
5	95.5-85.7	91.8	79.9	89.4-65.9	95.7-87.6	92.0
6	95.0-89.1	92.8	78.1	85.0-69.7	95.8-89.1	93.1
7	94.6-87.6	91.8	80.0	84.5-72.0	95.4-87.5	90.7
8	92.6-86.6	90.0	78.4	85.2-75.3		
9	78.1-65.0	73.3	81.2	87.0-73.0		
10	87.7-74.5	82.3	81.5	87.5-73.5		
11	82.4-70.5	77.0	91.5	94.3-87.7		

values of the centrifuged influent as C_{SO} -values, as it was unknown what part of the suspended, undissolved matter in the influent was digestible. It appears that the measured and computed values sometimes differ somewhat, but because they always have the same order of magnitude, and the differences are due to the complicated composition of the influent (see later), it can be concluded that this test of the model was also successful, and that therefore the two-mixer model can also be used for the semi-technical reactor. The fact that for the behaviour of the liquid and that of the substrate different models can be used, can perhaps be explained as a result of the complex reaction between the substrate and the sludge.

Because of the use of C_{SO} -values, based on TOC-measurements of centrifuged influent, the organic loading of the reactor in the model calculations could have been smaller than the actual loading. That this was in fact the case, appears from Fig. VI-3c, in which the measured gas productions either agreed with the predicted values, or were higher. It can thus be concluded that the undissolved matter in the influent will have contained a fraction of anaerobically convertible material. During the campaign, another part of the substrate was not digestible, this follows from the first eight experiments as presented in Fig. VI-3b and Table VI-5. From the figure it appears that the amount of waste products measured, was larger than the predicted amount. This must have been a result of the fact that non-biodegradable products present in the influent, will be measured as bacterial waste products in the effluent. This effect depresses the conversion of substrate, as can be seen from the table.

During the experiments 9, 10, and 11, performed in the intercampaign, the influent was only partly acidified (20 - 40 %). This may have been the cause of the larger differences between predictions and measurements, as compared with the other experiments. From the measured values of the substrate conversion for these experiments (see Table VI-5), it follows that the substrate was almost completely biodegradable, resulting in high gas productions (see Fig. VI-3c). Because the kinetic constant for the formation of waste products had been determined for a completely acidified substrate, the predictions of the model with respect to the waste products for these three experiments, were not very reliable.

VI.3 *Simplified model*

The network of relations, presented in block diagram 1, is rather complicated for the reason that this model has to give a proper description of the process over a wide range of process conditions. For instance, it has to be valid, both

for an efficiently operating reactor, as well as for a strongly overloaded reactor. In practice, overloading will always be avoided as much as possible, and therefore a simplified model that only holds for those situations when the reactor operates effectively, has enough practical value. In such a situation, it has been found that the fatty acids concentrations in the bed and the blanket are almost equal, and are small compared with the influent concentration; so that the sludge in both reactor parts is loaded almost equally. Therefore, to simplify the model, it can be assumed that for a well working reactor, equality exists for the q-values of the bed and the blanket, so that Eqns. VI-1 and VI-2 can be simplified to :

$$V_b \frac{dC_{fa,b}}{dt} = F_o (C_{so} - C_{fa,b}) - q C_{mb} V_b \quad \text{VI-11}$$

$$V_d \frac{dC_{fa,d}}{dt} = F_k (C_{so} - C_{fa,d}) - q C_{md} V_d \quad \text{VI-12}$$

or, under stationary conditions :

$$F_o (C_{so} - C_{fa,b}) = q C_{mb} V_b \quad \text{VI-13}$$

$$F_k (C_{so} - C_{fa,d}) = q C_{md} V_d \quad \text{VI-14}$$

From these equations the following can be derived :

$$h_b = \frac{F_o (C_{so} - C_{fa,b})}{q A_R C_{mb}} \quad \text{VI-15}$$

$$h_d = \frac{F_k (C_{so} - C_{fa,d})}{q A_R C_{md}} \quad \text{VI-16}$$

The equations for the production of bacterial waste products can also be simplified. Under stationary conditions, Eqns. VI-7 and VI-8 change to :

$$Q = F_{db} C_{ap,d} - F_{bd} C_{ap,b} + r_{ap,b} V_b \quad \text{VI-17}$$

$$Q = F_{bd} C_{ap,b} - F_{db} C_{ap,d} - F C_{ap,d} + r_{ap,d} V_d \quad \text{VI-18}$$

From these two equations it follows that :

$$r_{ap,b} V_b + r_{ap,d} V_d = F C_{ap,d} \quad \text{VI-19}$$

and with Eq. II-22, this equation can be transformed :

$$C_{ap,d} = \frac{k_{ap}}{F} (C_{so} - C_{fa,d}) (V_b + V_d) \quad \text{VI-20}$$

or

$$C_{ap,d} = \frac{k_{ap}}{v_L} (C_{so} - C_{fa,d}) (h_b + h_d) \quad \text{VI-20}$$

The equations for the gas production can also be simplified, as an identical empirical relation between the methane production and the TOC-conversion was found in the measurements with the 200 m³ reactor and the 60 l reactor. This simple, empirical relation will be used here :

$$\phi_{g,me} = k_{sg} F (C_{so} - C_{s,e}) \quad \text{VI-21}$$

or

$$\phi_{g,me}'' = k_{sg} v_L (C_{so} - C_{s,e}) \quad \text{VI-21}$$

For the measurements with the 200 m³ reactor : $k_{sg} = 1.12 \pm 0.02 \text{ m}^3 \text{CH}_4 / \text{kg TOC}$
(see Fig. VI-4); for the laboratory reactor : $k_{sg} = 1.22 \pm 0.01 \text{ m}^3 \text{CH}_4 / \text{kg TOC}$

Fig. VI-4. The production rate of methane as a function of the conversion rate of total organic carbon, measured for the 200 m³ reactor.

(see Fig. II-2). The value of k_{sg} depends somewhat on the composition of the substrate (for these compositions see Chapter I).

The total gas production rate can be derived from Eq. VI-21, if the fraction

of methane in the gas is known :

$$\phi_g'' = \frac{\phi_{g,me}''}{f_{me}}$$

VI-22

The simplified model, resulting from the considerations given above, has been presented in block diagram 2, see Fig. VI-5.

Fig. VI-5. Block diagram 2. A simplified version of block diagram 1.

VI.4 Restrictions of the simplified model

The restricting values for the range of the influent concentrations, stem from the experiments on the kinetics and the carbon-balance, performed in the laboratory reactor. The kinetic constants found (and therefore also the simplified model presented in this chapter), may be used for the following range of influent concentrations : $0.4 \leq C_{SO} \leq 3.0 \text{ kg C/m}^3$.

The bed height is also subjected to restrictions; for instance the measurements in the 200 m^3 reactor showed that in a bed with a height as small as 0.4 m, large bypassing streams occur. Such an unfavourable effect becomes smaller and thus less important the higher the bed; but it increases again if the bed becomes too high. The measurements in the 30 m^3 pilot-plant indicated that at a bed height between 1.2 and 2.2 m, the bypassing effect is acceptable, and therefore, the model was developed for bed heights that lay between these boundary conditions.

Because the experiments on the fluid flow pattern and those on the sludge behaviour in the 30 m^3 pilot-plant were performed with sludge with good settling properties ($v_s = (6 - 14)10^{-3} \text{ m/s}$), and at a linear fluid velocity of 2 m/h maximum, the model is only valid for this type of sludge and below this maximum linear fluid velocity.

The last restriction concerns the height of the reactor ($h_R = h_b + h_d + h_{st}$). The highest reactor used was the 30 m^3 pilot-plant, with $h_R = 6.5 \text{ m}$ (of which the settler took 1 m). The model is therefore only guaranteed for reactors up to this height.

VI.5 Test of the simplified model

For this test, five experiments were selected from the list presented in Table VI-4, for which the 30 m^3 pilot-plant operated in such a way that the concentration of fatty acids in the effluent remained under 30 mg C/l. This meant that the purification efficiency (based on the fatty acids concentration) was very good. However, the efficiency based on TOC was much smaller (69-80 %), as the adaptation time needed for the total organic mixture appeared to be much longer than the month gone since the start-up of the reactor in campaign 1976.

The results of the test for the percentage conversion of the substrate (TOC), have been presented in Table VI-5. The data for the simplified model agree very well with those of the non-simplified model, and sufficiently with the measured values. The check for the concentrations of fatty acids and bacterial waste products, gave similar results. The calculated flow rates of gas agreed

even better with the measured values than the gas flow rates calculated with the non-simplified model. This was caused by the use of the relation between the substrate conversion and the gas production, that was empirically determined for the 200 m³ reactor. It can be concluded that the simplified model predicts the performance of an efficiently operating reactor well, and that it therefore may be used instead of the model presented earlier.

VI.6 Optimisations with the simplified model

VI.6a Procedure

The model contains five independent parameters, namely : the influent concentration C_{so} , the effluent concentration $C_{s,e}$, the linear fluid velocity v_L , the bottom surface of the reactor A_R , and the height of the settler h_{st} . In order to present the results of the model calculations as a function of only the influent concentration, the quantitative values of the last four parameters have to be chosen on the basis of practical relevancy.

Since the settler operation was not included in the model, a sufficient efficiency for the separation and recirculation of the sludge had to be guaranteed by suitable choices of process conditions based on experimental information of the settler operation. From the experiments with the 30 m³ pilot-plant and with the 200 m³ full-scale reactor, it appeared that if the linear fluid velocity at the top of the settler (that is the same as the linear fluid velocity at the top of the reactor under the settler), was equal to 1 m/h, the amount of sludge in the reactor increased; while if $v_L = 2$ m/h, the amount of sludge remained almost constant. Therefore, computations of the model were made for these two values of v_L : 1 m/h if the amount of sludge available is insufficient, so that an increase would be necessary ; 2 m/h for those situations when a sufficient amount of sludge is present. For the height of the settler, a value of $h_{st} = 1$ m, should be chosen, as applied. Finally, because it was found that by the production of gas in the settler its efficiency was decreased, no production of gas should be able to occur there, provided that the conditions prevail for which the model will be calculated. Measurements in reactors of different size, indicated that if $C_{fa,e} \leq 30$ mg C/l, almost no gas was produced in the settler, and almost no washout of sludge occurred. For the calculations, the maximum value of the effluent fatty acids concentration is therefore taken as 30 mg/l. This agrees well with the condition that the reactor has to operate efficiently, so that the effluent concentration of fatty acids is small. The maximum value taken, implies for the interval of influent concentrations already given, a purification efficiency of 92.5 - 99 %. From the kinetics it

follows that this can be obtained if $q = \sim 0.5 q_{\max}$.

These values, together with those of the constants and other experimental conditions, that are collectively presented in Table VI-6, can be introduced into the model. The calculated values of the following parameters are then obtained as functions of the influent concentration : the heights of the bed and the blanket (giving with h_{st} the height of the reactor), the organic loading, the concentration of sludge in the blanket, and the volume of gas produced per unit of time.

TABLE VI-6

List of the boundary conditions, the values of the constants, and the experimental conditions, needed for the computation of the substrate model.

Boundary conditions	Parameters/constants of the sub-models	Experimental conditions
$C_{fa} \leq 30 \text{ mg C/l}$	$a_1 = 0.126$	sludge : VSS = 80 %
$1.2 \leq h_b \leq 2.2 \text{ m}$	$a_2 = 0.17 \text{ m}^{-1}$	%C = 50 % of the VSS
$h_d \geq 1.7 \text{ m}$	$a_3 = 1.0$	$v_s = 11.6 \cdot 10^{-3} \text{ m/s}$
$0.4 \leq C_{so} \leq 3.0$	$k_{tb} = 12.8$	$C_{mb} = 34 \text{ kg C/m}^3$
kg C/m^3	$k_{td} = 38.4$	$f_{me} = 0.83$
$h_R \leq 6.5 \text{ m}$	$q_{\max, fa} = 4.7 \cdot 10^{-5} \text{ s}^{-1}$	$v_L = 1 \text{ m/h}$ and 2 m/h
	$K_{fa} = 38 \cdot 10^{-3} \text{ kg C/m}^3$	influent : 100 % fatty acids
	$k_{ap} = 5.5 \cdot 10^{-6} \text{ s}^{-1}$	$k_{sg} = 1.12 \text{ m}^3 \text{CH}_4/\text{kgTOC}$
		$h_{st} = 1.0 \text{ m}$
		$A_R = 1.0 \text{ m}^2$

As the model is only guaranteed in a restricted range of conditions, the results of the calculations naturally have to lie within these ranges. For the calculated height of the sludge bed, a consequence is, that if at the calculations of the bed height, it turns out to be smaller than 1.2 m, further calculations will be made with a fixed value of $h_b = 1.2 \text{ m}$, which results in a lower specific loading of the sludge in the bed than has hitherto been assumed. Another consequence is that the computer calculations will stop if the height of the sludge bed comes above 2.2 m, and if the reactor height turns out to be larger than 6.5 m.

A minimum value for the height of the blanket follows from the finding that with large gas productions, a transitional area of about 1 m in height is

Fig. VI-6. Block diagram 3, used for the calculations of the optimum reactor dimensions and process operation.

present between the bed and the blanket. In this area, sludge concentrations of more than 40 kg SS/m^3 often occur, while the settler efficiency will decrease if the sludge concentration in its inflow is larger than this value. Therefore, a minimum distance of about 1 m is required between the bottom of

the settler and the top of the sludge bed. Furthermore, because in the apparatus used, the bottom of the settler was about 0.7 m lower than the top of the blanket, it could be derived that the minimum height of the blanket, including an area of transition, had to be 1.7 m. Therefore, if by computer calculation the height of the blanket turns out to be smaller than this minimum, the computer proceeds with $h_d = 1.7$ m (which then results in a lower specific loading of the sludge in the blanket than has been assumed). In block diagram 3 the simplified model, in which the restrictions have been incorporated, is presented (see Fig. VI-6).

VI.6b Results

The results have been presented in the Figs. VI-7a, b and c for a linear fluid velocity $v_L = 1$ m/h, and in the Figs. VI-8a, b and c for $v_L = 2$ m/h. It has been assumed that the model can be used for less active sludge with a lower

Fig. VI-7. Results of the optimisations obtained from the simplified model, presented in block diagram 3, as a function of the influent substrate concentration; calculated for various specific substrate conversion rates and the linear fluid velocity $v_L = 1$ m/h.

Fig. VI-7a. The heights of the bed, blanket and the reactor, respectively h_b , h_d and h_R .

Fig. VI-7b. The concentration of sludge in the blanket.

Fig. VI-7c. The organic loading that can be converted (ϕ'''), and the volume of methane produced per unit of time per m^3 reactor volume ($\phi'''_{g, me}$).

q -value, and therefore, the calculations have been performed for several specific substrate conversion rates as is indicated in the figures. The largest q -value used ($q = 2.0 \text{ d}^{-1}$) fulfils the boundary condition $q = 0.5 q_{\max, fa} = 0.5 (4.7 \cdot 10^{-5}) \text{ s}^{-1}$.

By means of these results, the dimensions of the reactor, the amount of sludge necessary, the organic loading that can be applied, and the purification efficiency and gas production that can be expected, can be determined. These data (to be used for optimisation) are obtained by reading from the figures at given C_{SO} -value, q -value and v_L -value. Therefore, in order to make the correct choices in these figures, the flow rate and the concentration of the influent have to be known; the maximum specific substrate conversion rate of the sludge (because $q = \sim 0.5 q_{\max, fa}$), and the saturation constant (K_{fa}), also have to be known. As said before, the linear fluid velocity to be applied has to be chosen : 1 m/h if an increase of the amount of sludge is necessary (in this case, the Figs. VI-7a, b and c have to be used for the optimisation), 2 m/h if sufficient sludge is available (Figs. VI-8a, b and c).

Fig. VI-8a. The heights of the bed, blanket and the reactor, respectively h_b , h_d and h_R .

Fig. VI-8b. The concentration of sludge in the blanket.

Fig. VI-8. Results of the optimisations obtained from the simplified model, presented in block diagram 3, calculated for the linear fluid velocity $v_L = 2$ m/h.

Fig. VI-8c. The organic loading that can be converted (ϕ_S'''), and the volume of methane produced per unit of time per m^3 reactor volume ($\phi_{g,me}'''$).

VI.6c Some remarks

It has to be noted that not one of the reactors used for this study was designed optimally, as often choices had to be made for parts of the design that were based on other than technological considerations. Furthermore, process operation was rarely at an optimum because of starting-up problems of the reactors as well as difficulties with the wastewater circuit of the sugar factory.

The results obtained from the model-calculations, show for both the 30 m³ pilot-plant and the 200 m³ reactor, that, if they are operated optimally, much higher loadings than have been used until now can be applied. For instance, at an influent concentration of 3 kg C/m³ (fatty acids) and a linear fluid velocity of 2 m/h, in the 200 m³ reactor an amount of 25.2 kg C can be converted per m³ reactor volume per day, if the reactor contains a sufficient amount of sludge, and if this sludge is as active as the sludge used in the experiments on the kinetics (Fig. VI-8c). With such a loading the concentration of sludge in the blanket will be 8.8 kg C/m³ (Fig. VI-8b), and the volume of methane produced per cubic meter per day will amount to 28 d⁻¹ (or for the total reactor 5610 m³/d). From the heights of the bed and the blanket, respectively 1.2 and 2.3 m (Fig. VI-8a), and the local sludge concentrations, it follows that 2750 kg C, or 6350 kg SS (80 % VSS, of which 45 % C) of sludge has to be present. For the 30 m³ pilot-plant it follows from Fig. VI-8a that the height has been estimated a factor 1.5 too large, which means that in this reactor per m³ reactor volume a smaller amount of substrate can be converted (about 16.8 kg C/m³d).

As we have never worked with an optimally operating reactor of technical or semi-technical scale, the results of optimisations will always - to a certain extent - be predictions that cannot be fully guaranteed. The high gas productions that appear to be possible, as has been predicted by means of the model, have not yet been realised, although with the pilot-plants already some spectacular results were obtained. However, it is not known whether at such gas flow rates the fluid flow pattern can still be described as a combination of two ideal mixers (as has been done in the model), or whether only one ideal mixer for the fluid flow in both the bed and the blanket would be more appropriate. In the second case, of course another optimum would have been found. Because better mixing results in an improved contact between sludge and substrate, and in a better transport of substrate to the bacteria in the flocs, it is probable that this optimum will even be higher than the value predicted. So it appears that in our optimisation a conservative procedure was followed,

and consequently it can be stated that this aspect of the high flow rate of gas does not affect the predictive value of the model. Another aspect is the fact that at these high values, the efficiency of the settler can be influenced negatively. This has been discussed in Chapter V. The problem may be anticipated by application of a new type of settler based on the measures proposed in that chapter.

Because of the low sludge bed in the 200 m^3 reactor, the model could not be checked for this reactor. Therefore, the results of the optimisations for reactors of this size (or larger) have to be used with some reservation. In this connection it is very encouraging that in spite of the too small sludge contents of both the 200 m^3 reactor and the 800 m^3 reactor, already good results were obtained with these reactors.

The model was developed for the bed-blanket system with bypassing streams present through the bed. Although by a correct design of the influent distribution system, the magnitude of the bypassing stream can be reduced, it has to be noted that bypassing or channeling is an effect that will always be present in this kind of reactor. Furthermore, the presence of a sludge bed through which the fluid flows in an upward direction, and in which gas is produced, has as a consequence that the sludge blanket must also be present.

To enlarge the possibilities of the application of the model, more information is necessary about the influence that the height of the sludge bed has on the occurrence of bypassing streams, so that Eq. VI-9 can be determined more accurately for a larger range of bed heights. If more information were available about the influence of the settling velocity on the behaviour of the sludge, the accuracy of the model would be improved. This is in particular the case for settling velocities smaller than about $6 \cdot 10^{-3} \text{ m/s}$.

VI.7 Sensitivity analysis

To study which parameters have the largest influence on the quantitative results of the substrate model, and therefore should be most accurately measured, a sensitivity analysis of this model was made. The sensitivity of the organic carbon concentration in the effluent $S_{(C_{s,e})}$ to a change of a parameter P has been defined as follows :

$$S_{(C_{s,e})} = \frac{(\Delta C_{s,e} / C_{s,e})}{(\Delta P / P)} \quad \text{VI-23}$$

In view of the variations found in the constants, the sensitivity of the model to these variations has also been analysed :

$$S_{(C_{s,e})} = \frac{(\Delta C_{s,e}/C_{s,e})}{(\Delta K/K)}$$

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The results of these analyses are collected in Table VI-7 and presented graphically in the Figs. VI-9 and 10.

It appears from the results that of the parameters studied, the flows F and F_k and the concentrations C_{so} and C_{mb} show the largest influence. Therefore, in the operation, special attention has to be paid to keep the influent flow rate and concentration steady. Furthermore, it can be noted that changes of A_R (the bottom surface of the reactor), the settling velocity v_s and the heights of the bed and the blanket, respectively h_b and h_d , are of lesser importance, but are nevertheless still significant. The model appeared to be quite insensitive to changes of the backmix flow rate F_{db} .

The sensitivities to variations in the values of the constants, all show the same order of magnitude (see Table VI-7). In Fig. VI-10 the sensitivity of the substrate model has been presented for the 90 % reliability intervals for these constants. It indicates that if the inaccuracies of all the constants influence the model in the same way, then the model will bear little relation to reality. However, the effect of an inaccuracy of one constant will usually be counteracted by another inaccuracy of a second constant.

TABLE VI-7

Results of the sensitivity analysis of the simplified model.

Parameter (P)	$S_{(C_{s,e})}$	Constant (K)	$S_{(C_{s,e})}$
F	-0.6	$q_{\max,fa}$	-0.6
C_{so}	1.2	K_{fa}	0.4
A_R	0.2	f_{me}	0.2
F_k	0.8	k_{ap}	0.6
F_{db}	-0.02	k_{tb}	-0.5
C_{mb}	-0.7	k_{td}	0.5
h_b	0.2	k_{sg}	-0.2
h_d	-0.3		
v_s	0.2

Fig. VI-9. Parameter sensitivity of the simplified model presented in block diagram 2 (Fig. VI-5).

Fig. VI-10. Sensitivity of the simplified model presented in block diagram 2, to variations of its constants.

Given this analysis, and because uncertainties are present in the data, it has to be recommended that, if the present model is used to predict optimum process conditions and to design an optimum reactor, safety factors should be used in the choice of some parameters. To prevent overloading that could lead to the deterioration of the process, it is of prime importance that the sludge blanket should be oversized. A factor of 1.5 for the height seems to be

sufficient. Overdimensioning the sludge bed is not strictly necessary, as in this reactor part, overloading will not directly result in increased washout of sludge.

VI.8 Conclusions

- A model has been presented that describes the conversion of substrate for the methane forming step of the anaerobic process of wastewater treatment in the upflow reactor. This model has been developed for the bed-blanket system with bypassing streams present through the sludge bed. It can be used both for process description and for process optimisation.
- From the tests of the model it has been concluded, that, for a wide range of process conditions, it can be used to predict the process performance in a laboratory reactor (60 l) and in a reactor of a semi-technical scale (30 m³).
- A simplified model has been presented for a smaller range of process conditions. This model appeared to be valid for
 $C_{fa,e} \leq 30 \text{ mg/l}$ or $q \leq 0.5 q_{\max,fa}$; $0.4 \leq C_{so} \leq 3.0 \text{ kg C/m}^3$;
 $1.2 \leq h_b \leq 2.2 \text{ m}$; $h_d \geq 1.7 \text{ m}$; $h_R \leq 6.5 \text{ m}$; $v_s (6 - 14) 10^{-3} \text{ m/s}$ and
 $v_L \leq 2.0 \text{ m/h}$.
- From the sensitivity analysis of the simplified model it has been concluded that if the model is used to calculate optimum reactor dimensions, a safety factor of 1.5 for the height of the sludge blanket has to be recommended.
- If a sufficient amount of active, anaerobic sludge ($q_{\max,fa} = 4.7 10^{-5} \text{ s}^{-1}$) is available in a reactor that is dimensioned according to the results of the simplified model, it may be expected that an amount of 16.8 - 25.2 kg C can be converted per m³ reactor volume per day.

CHAPTER VII

SYNOPSIS

The objective of this work was to study the features of the methane forming step of the anaerobic process of wastewater treatment in upflow reactors, and to investigate whether from a technological point of view, such a process could be attractive for treatment of industrial wastewater. In co-operation with C.S.M.*, we were in the fortunate position to perform studies on reactors of a semi-technical and a technical scale. The results thus obtained, contributed directly to the development of a full-scale reactor. The Department for Public Health and Environmental Hygiene granted part of the finances for both the studies and reactors, because of its policy to stimulate the prevention of water pollution at the source of contamination.

A negative consequence of this type of co-operation was that the work had to be carried out over short periods of time, using reactors that neither operated stationary nor optimally. The studies of the reactors therefore mainly concerned the fluid flow pattern, the behaviour of the sludge, and the operation of the settler, as these measurements took only a few hours to a whole day, which were periods of time during which the process operation could be maintained stationary. However, for the study of the kinetics, the reactor had to operate stationary during longer periods, with the possibility to choose the process conditions freely. These experiments were therefore performed in a laboratory reactor.

In the diagram (see below), a survey has been given of the subjects studied, and their interrelations in the performance of the process. This information, together with experimental knowledge for solving practical problems, is necessary for the development of a full-scale upflow reactor. The work described in this thesis was organised and performed according to this diagram.

In Chapter I general information about the process, reactors, settlers, sludge, and the process performance, needed in the following chapters, has been collected. The recent literature about the technological features of this process, was used for the start of the study.

The study of the kinetics, described in Chapter II, was aimed at getting reliable data for the conversion of fatty acids and the formation of gas, biomass and other products, valid for the well adapted methane forming sludge used in this process.

* Centrale Suiker Maatschappij, Amsterdam, Nederland.

Survey of the subjects studied and their inter-relations in the process performance.

The sludge, that came from the 30 m³ pilot-plant, was studied in the 60 l laboratory reactor that was continuously loaded with a completely acidified sugar solution. By varying the amount of sludge and the organic loading applied, as well as by measuring the production rate and composition of the gas, and the concentrations of TOC and fatty acids in the influent and effluent; the kinetic constants that describe the conversion of fatty acids, the production of methane and the formation of bacterial waste products could be determined. Furthermore it could be concluded that the equation of Monod can be used for the kinetics of the conversion of substrate and that Pirt's modification of Monod's model, namely the introduction of a maintenance term in the substrate balance, gives an appropriate description of this process. The carbon-balance appeared to be an adequate means to quantify the substrate and product streams in anaerobic treatment of wastewater.

From the quantitative results obtained and presented in Chapter II, it follows that for fatty acids concentrations in the interval $0.4 \leq C_{so} \leq 3.0 \text{ kg C/m}^3$, even at the high specific substrate conversion rate $q = 2.0 \text{ d}^{-1} = 0.5 q_{\text{max,fa}}$, still very good purification efficiencies of the wastewater can be obtained, namely 92.5 - 99 %. For the prediction of the purification efficiency (based on TOC), the production rate of bacterial waste products has to be known. It was found that a linear relation may be assumed between this production

rate and the difference of the fatty acids concentrations in the influent and in the effluent.

The aim of the study on the dynamics of the liquid flow, described in Chapter III, was in the first place to develop a physical description of the fluid flow pattern in an upflow reactor with anaerobic sludge present as a sludge bed, sludge blanket and sometimes an area of transition, after which the influence of the gas production, bed height, distribution of sludge, linear fluid velocity, and the size of the reactor, on the fluid flow pattern had to be determined. These studies were performed by means of residence time distribution experiments with lithium chloride as the tracer. The data obtained from up-step stimulus-response experiments (by sampling in the effluent and on several places in the reactor), have been evaluated by means of a trial-and-error procedure, by which models were obtained that describe the fluid flow pattern in a reactor as a combination of regions with different flow regimes.

It appeared that the fluid flow in the upflow reactor can be described as a combination of ideal mixers, bypassing flow, backmix flow and (shortly after start-up), dead space regions. The regime of fluid flow in the settler, is mostly plug flow, sometimes in combination with an ideal mixer and dead space. Both the number and volumes of the ideal mixers as found in the reactor, were strongly influenced by the volume and height of the sludge bed : at a bed of 0.4 m, which is low as compared with a total reactor height of 4.5 m, two ideal mixers were found : a small one in the bed and a large one in the blanket (this is the area between the top of the bed and the settler at the top of the reactor); with a higher bed (1.2 - 2.2 m) three ideal mixers have been distinguished in the reactor : the first for the bed, the second for the area of transition between the bed and the blanket, and the third for the blanket.

Of the process conditions influencing the fluid flow pattern, the gas production must be mentioned first, as the rising gas bubbles produce the mixing in the reactors. However, it has also been found that gas bubbles rising through the sludge bed, strengthen the bypassing of influent through the bed. For gas flow rates increasing from $1.4 \cdot 10^{-4}$ m/s to $3 \cdot 10^{-4}$ m/s, the model of the fluid flow pattern did not change significantly.

The height of the sludge bed appeared to strongly influence the magnitude of the bypassing flow rates. At a bed height of 0.4 m, very large bypassing streams were found ($F_k/F = 0.7 - 0.86$); at a height of 1.2 m, only one third of the influent bypassed the bed; while this stream was again larger at a height of 2.2 m. It was concluded that this last effect (which may be a result of thickening in the lower parts of the bed), restricts the height of the bed.

In the reactors of different size that were studied (the 6 m^3 , 30 m^3 and 200 m^3 reactors, height : 3 - 6.5 m; diameter : 1.5 - 7.5 m; bottom surface : $1.5 - 45\text{ m}^2$) the fluid flow patterns were not significantly different. However, for the 200 m^3 reactor "two-mixer models" have been found, and for the other reactors "three-mixer models"; but this was due to the relatively small amount of sludge in the 200 m^3 reactor and not a result of the different reactor dimensions. On the basis of this study, it could be recommended that scaling-up should be effected by enlarging the horizontal dimensions of the reactor. In the studies as described in Chapter III, no indication was found that such enlarging of the bottom surface of the reactor is restricted to a maximum value. In this chapter, the studies concerning the influent distribution system have also been described. The work was performed by means of pulse tracer experiments and sampling close to the influent point, as well as by means of up-step stimulus-response experiments. From the comparison of the horizontal, unidirectional influent distribution system with the horizontal, radial distribution system (both used in the 30 m^3 pilot-plant), it appeared that the first system is to be preferred, since the influent flow pattern is constant. The measurements in the 30 m^3 pilot-plant and the 200 m^3 reactor showed furthermore, that one unidirectional influent point on a bottom surface of $5 - 11\text{ m}^2$ gave a good distribution of the influent, resulting in a good contact between the substrate and the sludge. Only during starting-up periods with a thick inactive sludge bed present in the reactor, dead space regions were found and strong bypassing streams existed. In such periods, temporary enlarged fluid flows have to be applied in each influent point, to disturb the compactness of the bed. Therefore, it has been recommended that the design of the influent system is such that each influent point can be opened and closed individually, and that the points are located and directed such that the effect of one point on the fluid flow pattern in the reactor is multiplied by the effects of other points. An example of such a system can be seen in the influent distribution system of the 200 m^3 reactor (as presented in Chapter I).

The measurements of the distribution of sludge in several reactors, showed that in the sludge bed the concentration of sludge did not depend on either place or time, and that in the sludge blanket the concentration was everywhere the same, but depended on the flow rate of gas and the settling velocity of the sludge. This work has been described in Chapter IV.

At the study of the 30 m^3 pilot-plant, that contained well adapted, well settling sludge particles, a linear relation was found between the ratio of the sludge concentrations in the bed and the blanket, and the inverse of the quantity of gas per unit of time leaving the top of the sludge bed. This is in agree-

ment with the model and the resulting equation developed for the behaviour of the sludge in the upflow reactor. Therefore, this equation and the values found for the factors that describe the transport of sludge from the bed to the blanket by the rising gas bubbles, and from the blanket back to the bed by the backmix stream, have been used in the substrate model as described in Chapter VI.

The 200 m³ reactor contained well settling sludge particles that came originally from the 30 m³ pilot-plant, as well as a fraction of mud that had poor settling properties. It appeared that for the well settling sludge particles the factor for the transport of sludge upwards from the bed to the blanket agreed well with the value found for the 30 m³ pilot-plant. It has therefore been concluded that enlarging of the bottom surface of the reactor from 5.4 m² to 44.2 m² does not have consequences for the behaviour of the sludge in the upflow reactor, provided that the recommended influent distribution is applied.

From the studies on the distribution of the sludge it followed furthermore, that at a gas flow rate of about $0.7 \cdot 10^{-4}$ m/s or higher, an area of transition can be present between the bed and the blanket, where the sludge concentration decreases with the height above the bed. The height of this area in both pilot-plants was about 1 m. As it was also found that the settler efficiency decreases at high sludge concentrations directly below the recirculation opening, it is therefore recommended that the distance between the top of the bed and the bottom of the settler should be more than 1 m. This was the case in the 30 m³ and 200 m³ reactors. From the studies described in Chapter IV, no indication was obtained as to whether the height of the sludge blanket should be restricted by any maximum. Even in a blanket with a height of 3.5 - 4.5 m (30 m³ pilot-plant, 200 m³ full-scale reactor) the sludge concentration appeared to be independent of the place in the blanket.

The most important conclusions drawn from the studies of the settlers (described in Chapter V) were that the settling compartment has to be separated effectively from the turbulent sludge blanket in the reactor; that the retention time of the sludge in the settler has to remain short; and that foam, floating sludge and gas that liberates by expansion have to be collected before they can disturb the laminar fluid flow in the settling compartment. If effective measures are taken to solve these problems, a sludge free zone can be maintained in the upper layers of the settling compartment. In such a situation, the good settling properties of anaerobic sludge guarantee that high separation efficiencies can be obtained.

In the apparatuses used, the thickened sludge was transported back to the reactor by means of a circulation stream that is generated by the difference of

the densities of the suspension at the bottom of the settler and the three-phases suspension in the reactor. For this transport, it was very important that the bottom plates of the settler each are placed under an angle of 45° since in that situation the sludge particles slide or roll downwards along the plates. Because of the good separation efficiencies (more than 99 %) that were reached in each of the settlers studied, this type of settler could be recommended for application on top of an upflow reactor. If the dimensions of the reactor would result in a settler with too large a depth, more than one settler could be applied.

The settlers as proposed are simple and cheap and have the advantage that a scraper system, that transports the sludge to the recirculation opening, is unnecessary. Advantages of building a settler on top of the reactor are that less room is necessary for the total system, and that the anaerobic bacteria do not come into an alien environment before they are recirculated.

Finally, in order to describe and to optimise the anaerobic process in the upflow reactor, in Chapter VI a model for the conversion of fatty acids, based on a combination of the dynamics of the liquid flow, the kinetics, and the behaviour of the sludge, has been developed. This model is based on a fluid flow pattern in the reactor consisting of two ideal mixers (one for the bed and the other for the blanket), a bypassing stream and backmixing. In the mass-balances for the substrate over the bed and the blanket, conversion terms are included; while the model description of the sludge behaviour has been introduced via the concentrations of sludge in both reactor parts. The operation of the settler was indirectly taken into account via the boundary conditions of the model.

The substrate model was tested empirically, and used to study which parameters have the largest influence on the process. From the tests it was concluded that the model adequately describes the process in the upflow reactor, especially if a completely acidified substrate is used. Furthermore, it was concluded that the model, if applied to efficiently operating reactors, can be simplified considerably, as in that case the concentrations of fatty acids in the bed and blanket are almost equal. By means of a sensitivity analysis, it was found that the influent concentration and flow rate, the bypassing flow rate, and the concentration of sludge in the bed are the most influential parameters in this process. It was concluded that the practical and fundamental information obtained and incorporated in the substrate model, gives a proper insight into the relations between the process performance and the operating conditions, with which it has become possible to develop an optimally operating upflow reactor.

The growth of sludge and the operation of the settler have not been incorporated in the model, so that the simplified model has only been calculated for those conditions in which efficient settler operation (resulting in more growth of sludge than washout), could be guaranteed : $v_s = 11.6 \cdot 10^{-3} \text{ m/s}$, $v_L = 1 \text{ m/h}$ or 2 m/h , $C_{\text{md}} \leq 40 \text{ kg SS/m}^3$, $h_{\text{d,min}} = 1.7 \text{ m}$, $C_{\text{fa,e}} \leq 30 \text{ mg C/l}$. The calculations indicated that if the 200 m^3 reactor were to be operated optimally, and would contain sufficient sludge of a good activity ($q_{\text{max,fa}} = 4.7 \cdot 10^{-5} \text{ s}^{-1}$), at an influent concentration of 0.4 kg C/m^3 of fatty acids, an amount of $8 \text{ kg C/m}^3\text{d}$ or about $20 \text{ kg COD/m}^3\text{d}$ can be converted. At an influent concentration of 3.0 kg C/m^3 , an amount of even $25.2 \text{ kg C/m}^3\text{d}$ or about $60 \text{ kg COD/m}^3\text{d}$ can be converted. The purification efficiency obtained, would be 92.5 - 99 %. From these results it can be concluded that anaerobic treatment of wastewater containing fatty acids has many technological points of recommendation for industrial applications.

SLOTBESCHOUWING EN SAMENVATTING

De doelstelling van het werk, dat is neergelegd in dit proefschrift, was het bestuderen van de technologie van de methaan vormende stap van het anaerobe waterzuiveringsproces in een opstroomreaktor. Hierbij werd onderzocht of dit proces technologisch geschikt is voor toepassing op industrieel afvalwater. Door een intensieve samenwerking met de C.S.M.* konden we dit onderzoek uitvoeren in reaktoren van semi-technische en technische schaal, en daardoor droegen de resultaten direkt bij tot de ontwikkeling van een reaktor van praktijkgrootte. Het Ministerie van Volksgezondheid en Milieuhygiëne heeft met een bedrag van ongeveer een miljoen gulden bijgedragen in de kosten van het onderzoek en van de reaktoren. Deze subsidie was een gevolg van het stimuleringsbeleid dat erop is gericht waterverontreiniging bij de bron te voorkomen. Een negatieve consequentie van deze samenwerking was, dat het werk moest worden uitgevoerd in perioden van beperkte tijdsduur, met reaktoren die geen van alle stationair of optimaal funktioneerden. De onderwerpen die in deze reaktoren werden bestudeerd, betroffen het vloeistofstromingsgedrag, het gedrag van het slib, de werking van de bezinker(s), aangezien de hiervoor noodzakelijke metingen meestal slechts enige uren tot een dag in beslag namen, en gedurende deze tijdsduur het proces doorgaans wel in een stationaire toestand kon worden gehouden. Bij de experimenten ter bestudering van de kinetiek moest het proces gedurende langere perioden stationair blijven en moesten de procesomstandigheden nauwkeurig instelbaar zijn. Daarom werden deze experimenten uitgevoerd in een laboratorium reaktor.

In het onderstaande diagram is een overzicht gegeven van de onderwerpen, die moeten worden bestudeerd om uiteindelijk te komen tot een goed funktionerende opstroomreaktor van praktijkgrootte. De combinatie van deze onderzoeken geeft inzicht, zowel in het proces zelf, als ook in de invloed van de schaalgrootte op het proces. Het werk, dat in dit proefschrift is beschreven, is opgezet en uitgevoerd aan de hand van dit schema.

In Hoofdstuk I is de algemene informatie verzameld, die nodig is voor de daarna volgende hoofdstukken. Deze informatie betreft het proces, de procesvoering, de reaktoren, de bezinkers en het slib. De recente literatuur over de technologie van dit proces, is gebruikt als uitgangspunt van het onderzoek.

De in Hoofdstuk II beschreven studie van de kinetiek van de omzetting van vetzuren, en de vorming van gas, biomassa en andere produkten, was gericht op het verkrijgen van betrouwbare gegevens die geldig zijn voor het goed geadapteerde methaanvormende slib dat is gebruikt. Het slib, afkomstig uit de 30 m³ semi-

* Centrale Suiker Maatschappij, Amsterdam.

Overzicht van de bestudeerde onderwerpen en hun onderlinge relaties.

technische reaktor, is voor deze studie gebracht in een 60 l lab-reaktor die kontinu werd belast met een volledig verzuurde suikeroplossing. De kinetiek konstanten konden worden bepaald door variatie van zowel de hoeveelheid slib als de organische belasting, en door meting van de produktiesnelheid en de samenstelling van het gas, en de concentraties aan TOC en vetzuren in het influent en effluent. Met deze konstanten was het mogelijk om de omzetting van vetzuren en de vorming van methaan en bakteriële afvalprodukten kinetisch te beschrijven. Er kon worden gekonkludeerd dat de vergelijking van Monod gebruikt kan worden voor de beschrijving van de substraatafbraak en dat de door Pirt voorgestelde modifikatie van het model van Monod een voor dit proces geschikte uitbreiding is. Deze modifikatie omvat de introductie van een term in de substraatbalans die de omzetting van substraat ten behoeve van celonderhoud beschrijft. Om de substraat- en produkt stromen bij het anaerobe waterzuiveringsproces kwantitatief weer te geven bleek de koolstofbalans een geschikt middel te zijn.

Uit de kwantitatieve resultaten van dit onderzoek is af te leiden, dat voor vetzuur concentraties in het interval $0,4 \leq C_{SO} \leq 3,0 \text{ kg C/m}^3$, zelfs bij de hoge specifieke substraatverbruiksnelheid $q = 2,0 \text{ d}^{-1}$, nog een zeer goede zuivering van het afvalwater kan worden bereikt, namelijk 92,5 - 99 %. Om de zuivering op basis van TOC te kunnen voorspellen moet de vormingssnelheid van

bakteriële afvalprodukten bekend zijn. Het bleek mogelijk om het verband tussen deze vormingssnelheid en het verschil in de vetzuur concentraties in het influent en effluent te beschrijven met een lineaire relatie.

Het onderzoek naar het dynamisch gedrag van de vloeistof is beschreven in Hoofdstuk III. De doelstelling was in de eerste plaats het ontwikkelen van een geschikte fysische beschrijving van het vloeistofstromingsgedrag in een opstroomreaktor met daarin anaeroob slib (aanwezig als een slibbed, slibdeken, en soms een overgangsgebied tussen het bed en de deken), en in tweede instantie het bepalen van de invloed van de gasproductie, bedhoogte, verdeling van het slib, lineaire vloeistofsnelheid, en de afmetingen van de reaktor, op dit dynamisch vloeistofgedrag. Deze studies zijn uitgevoerd met behulp van verblijftijdspreidingsexperimenten met lithiumchloride als tracer. De gegevens, die zijn verkregen door middel van stapvormige stimulus-respons experimenten (hierbij werden zowel in het effluent als op meerdere plaatsen in de reaktor monsters genomen), zijn geëvalueerd met een "trial-and-error" methode. Hiermee zijn modellen verkregen die de vloeistofstroming in een reaktor beschrijven als een combinatie van gebieden met verschillende stromingsregimes.

Als model van de vloeistofstroming in de opstroomreaktor bleek in het algemeen een combinatie van ideale mengers, kortsluitstroming, terugmengstroming en (kort na de start) dode ruimte, goed te voldoen. Het stromingsregime in de bezinker was meestal propstroming, soms in combinatie met een ideale menger en dode ruimte. Zowel het aantal als het volume van de ideale mengers in de reaktor werd sterk beïnvloed door het volume en de hoogte van het slibbed: bij een bedhoogte van 0,4 m, hetgeen laag is bij een totale reaktorhoogte van 4,5 m, bleken twee ideale mengers een geschikte beschrijving te geven: een kleine menger voor het slibbed en een grote voor de slibdeken (dit is het vloeistofvolume tussen de bovenkant van het slibbed en de bezinker bovenin de reaktor); bij een hoger bed (1,2 - 2,2 m) konden drie ideale mengers worden onderscheiden: de eerste voor het bed, de tweede voor het overgangsgebied tussen het bed en de deken, en de derde voor de deken.

Van de procesomstandigheden die het stromingspatroon van de vloeistof beïnvloeden, moet in de eerste plaats de gasproductie worden genoemd, aangezien de opstijgende gasbellen de menging veroorzaken. Er is ook gevonden, dat in het slibbed opstijgende gasbellen de kortsluitstroom van influent door het slibbed heen naar de deken, versterken. Het bleek dat bij een toename van de gassnelheid van $1,4 \cdot 10^{-4}$ m/s tot $3 \cdot 10^{-4}$ m/s, de modellen voor het vloeistofstromingspatroon niet significant veranderden.

De hoogte van het slibbed bleek ook een belangrijke invloed te hebben op de grootte van de kortsluitstroom: bij een bedhoogte van 0,4 m werden grote

kortsluitstromen gemeten ($F_k/F = 0,7 - 0,86$), bij een hoogte van 1,2 m passeerde slechts eenderde van het influent het bed. De kortsluitstroom werd weer groter bij een bedhoogte van 2,2 m. Dit laatstgenoemde effect (hoogstwaarschijnlijk een gevolg van indikking onderin het slibbed) levert uiteindelijk een beperking op voor de toe te passen bedhoogte.

De modellen van de vloeistofstroming voor de verschillende reaktoren, die zijn bestudeerd (6 m^3 , 30 m^3 , 200 m^3 ; hoogte : 3 - 6,5 m, diameter : 1,5 - 7,5 m; bodemoppervlak : 1,5 - 45 m^2), bleken onderling geen significante verschillen te vertonen. Weliswaar werden voor de 200 m^3 reaktor "twee-menger modellen" gevonden en voor de andere twee reaktoren "drie-menger modellen", maar dit verschil was een gevolg van de relatief kleine hoeveelheid slib in de eerstgenoemde reaktor, en niet van de verschillende reaktorafmetingen. Op basis van deze studie kan dus de aanbeveling worden gedaan, dat bij schaalvergroting de horizontale afmetingen van de reaktor vergroot moeten worden. Bij de studies, die zijn beschreven in Hoofdstuk III, is geen aanwijzing gevonden dat de vergroting van het bodemoppervlak van de reaktor aan een limietwaarde zou zijn gebonden.

In ditzelfde hoofdstuk is ook het onderzoek aan het influentverdeelsysteem beschreven. Dit werk is uitgevoerd door middel van puls-tracer experimenten met monsternamen vlak bij het influentpunt, en door middel van stap-tracer experimenten. Bij de vergelijking van het horizontale, eenduidig gerichte verdeelsysteem met het horizontale, radiale verdeelsysteem (beide gebruikt in de 30 m^3 semi-technische reaktor), bleek dat aan het eerste systeem de voorkeur moest worden gegeven, aangezien hier het stromingspatroon bij het influentpunt konstant was. De metingen in de 30 m^3 en 200 m^3 reaktor toonden verder aan dat met één eenduidig gericht influent punt op een bodemoppervlak van 5 - 11 m^2 een goede influentverdeling en dus een goed contact tussen het slib en het substraat wordt verkregen. Slechts in opstartperiodes, bij aanwezigheid van een ingedikt, inactief slibbed, zijn dode ruimten aangetoond en is sterke voorkeursstroming aanwezig.

Tijdens zulke periodes moet in elk influentpunt tijdelijk een grote uitstroomsnelheid kunnen worden toegepast, waardoor het bed als het ware wordt opgebroken. Om die reden is het aan te bevelen dat elk influentpunt afzonderlijk geopend en gesloten kan worden, en dat ze zodanig worden geplaatst en gericht dat ze elkaars uitstroomeffect versterken. Een voorbeeld van een dergelijk systeem is het influentverdeelsysteem van de 200 m^3 reaktor.

De metingen aan de verdeling van het slib in de verschillende reaktoren toonden aan, dat in het slibbed de slibconcentratie onafhankelijk is van de plaats en van de proces omstandigheden en dat in de slibdeken de slibconcentratie

overal gelijk is, maar afhankelijk is van de gasproduktiesnelheid en de bezinkingsnelheid van het slib. Dit onderzoek is beschreven in Hoofdstuk IV.

Bij het onderzoek aan de 30 m^3 reaktor, waarin zich goed geadapteerd, goed bezinkend, korrelvormig slib bevond, bleek de verhouding van de slibconcentraties in het bed en de deken omgekeerd evenredig te zijn met de hoeveelheid gas die per tijdseenheid werd geproduceerd in het slibbed. Dit was in overeenstemming met het model en de daaruit voortgekomen vergelijking voor het gedrag van anaeroob slib in de opstroomreaktor. Deze vergelijking is, tezamen met de waarden voor de factoren voor transport van slib van het bed naar de deken door opstijgende gasbellen en vice versa door meevoering met de terugmengstroom, gebruikt in het substraatmodel dat in Hoofdstuk VI is beschreven.

De 200 m^3 reaktor bevatte, behalve de goed bezinkende slibdeeltjes die afkomstig waren uit de 30 m^3 reaktor, ook een hoeveelheid modder (die via het influent de reaktor was binnengekomen) waarvan de bezinkeigenschappen slecht waren. Uit de metingen bleek dat voor de goed bezinkende slibdeeltjes, de slibtransportfaktor voor het opwaartse transport goed overeen kwam met de waarde van deze faktor die was gevonden voor het slib in de 30 m^3 reaktor. Hieruit kan worden gekonkludeerd, dat het vergroten van het oppervlak van de reaktorbodan geen gevolgen heeft voor het gedrag van het slib in de reaktor, mits het aanbevolen influentverdeelstelsel wordt toegepast.

Bij de bestudering van het gedrag van het slib bleek voorts dat bij een gasnelheid in het slibbed van ongeveer $0,7 \cdot 10^{-4} \text{ m/s}$ en hoger, een overgangsgebied aanwezig was tussen het bed en de deken, waarin de slibconcentratie afnam met de hoogte boven het slibbed. De hoogte van dit gebied was in beide semi-technische reaktoren ongeveer 1 m. Daar eveneens was vastgesteld dat de effectiviteit van de slibafscheiding in de bezinker afneemt bij hoge slibconcentraties direkt onder de bezinker, is het aan te bevelen om tussen de onderkant van de bezinker en de bovenkant van het slibbed een afstand van tenminste 1 m aan te houden. Dit was zowel in de 30 m^3 als de 200 m^3 reaktor het geval. Bij het onderzoek dat in Hoofdstuk IV is beschreven, is geen aanwijzing gevonden of de hoogte van de slibdeken aan een maximum is gebonden. Zelfs in een deken met een hoogte van 3,5 - 4,5 m bleek de slibconcentratie plaatsanafhangelijk.

De belangrijkste konklusies van het onderzoek aan de bezinker (dit is beschreven in Hoofdstuk V) waren dat de bezinker zodanig moet zijn ingericht, dat zowel het gas als de turbulentie in de reaktor niet kunnen doordringen in de bezinkruimte omdat daar anders de laminaire vloeistofstroming wordt verstoord; dat de slibretentietijd in de bezinker kort moet zijn; en dat schuim, drijfslib en het door expansie vrijkomende gas moeten worden afgevangen voor aankomst in de bezinkruimte. Als effectieve maatregelen zijn getroffen om aan deze voor-

waarden te voldoen, kan bovenin de bezinkruimte een slibvrije zone worden gehandhaafd. In dat geval zijn de goede bezinkeigenschappen van anaerobe slibdeeltjes een garantie dat hoge slibafscheidingsefficiëncies kunnen worden bereikt.

In de gebruikte slibafscheiders werd door middel van een circulatiestroming die wordt opgewekt door het dichtheidsverschil tussen de ingedikte suspensie onderin de bezinker en de drie-fasen suspensie in de reaktor, het ingedikte slib teruggevoerd naar de reaktorruimte. Bij dit transport was van groot belang dat de schuine bodemplaten elk onder een hoek van 45° stonden, omdat dan de slibdeeltjes langs deze platen langzaam omlaag glijden. Aangezien met elk van de bezinkers een goede slibafscheidingsefficiëncy (meer dan 99 %) werd bereikt, kan dit type worden aanbevolen voor toepassing bovenop een opstroomreaktor. Indien, door de afmetingen van de reaktor, de erop te plaatsen bezinker een te grote insteekdiepte zou krijgen, kunnen meer dan een bezinker worden toegepast.

Het aanbevolen type is eenvoudig van konstruktie en daarmee goedkoop, terwijl het niet nodig is om een schrapersysteem toe te passen voor het transport van slib terug naar de reaktorruimte. Andere voordelen van het toepassen van de bezinker(s) bovenop de reaktor zijn, dat het geheel minder plaatsruimte nodig heeft, en dat de anaerobe bakteriën niet aan andere (vaak ongunstige) omstandigheden worden blootgesteld alvorens ze worden gerecirculeerd.

Om tenslotte te komen tot een beschrijving van het gehele proces, is in Hoofdstuk VI een model voor de omzetting van vetzuren ontwikkeld. Dit model is gebaseerd op een combinatie van de modelbeschrijvingen van de kinetiek, het vloeistofstromingsgedrag en het slibgedrag. Hierbij is uitgegaan van een vloeistofstromingspatroon in de reaktor, dat bestaat uit twee ideale mengers (een voor het bed, de andere voor de deken), een kortsluitstroom en terugmengstroming. In de massabalansen voor het substraat over het bed en de deken zijn konversietermen opgenomen; de modelbeschrijving van het slibgedrag is opgenomen via de concentraties van het slib in de beide reaktorgedeelten. De werking van de bezinker is indirect meegenomen via de grenswaarden waarbinnen het model geldig is. Dit substraatmodel biedt ook de mogelijkheid om het anaerobe waterzuiveringsproces in een opstroomreaktor door te rekenen en te optimaliseren.

Het substraatmodel is empirisch getoetst en gebruikt om te bepalen welke parameters in dit proces van belang zijn. De konklusie die uit de toetsingen kon worden getrokken was, dat het model een goede beschrijving van dit proces oplevert en wel vooral wanneer een volledig verzuurd substraat wordt gebruikt. Voorts bleek het goed mogelijk om het model sterk te vereenvoudigen indien het

alleen wordt gebruikt voor efficiënt werkende reaktoren. In dat geval zijn de vetzuurconcentraties in het bed en de deken nagenoeg gelijk. Uit een gevoeligheidsanalyse bleek, dat de influentconcentratie, het influentdebiet, het debiet van de kortsluitstroom en de slibconcentratie in het bed de parameters zijn met de grootste invloed. Er kan gekonkludeerd worden dat de praktische en fundamentele informatie, die is verkregen met en is gebruikt in het substraatmodel, ook na vereenvoudiging een goed inzicht geeft in de relaties tussen de werkingssomstandigheden en het proces. Hierdoor is het mogelijk geworden om een optimaal funktionerende opstroomreaktor te ontwikkelen.

Zowel de groei van het slib als de werking van de bezinker zijn niet opgenomen in dit model en daarom is dit vereenvoudigde model slechts doorgerekend voor die omstandigheden waarbij een effectieve slibafscheiding in de bezinker kan worden gegarandeerd (hierbij is de aangroei van slib groter dan of gelijk aan de uitspoeling) : $v_s = 11,6 \cdot 10^{-3}$ m/s, $v_L = 1$ m/h of 2 m/h, $C_{md} \leq 40$ kg SS/m³, $h_{d,min} = 1,7$ m, $C_{fa,e} \leq 30$ mg C/l. Uit de berekeningen bleek, dat als de 200 m³ reaktor optimaal zou funktioneren en voldoende goed actief ($q_{max,fa} = 4,7 \cdot 10^{-5}$ s⁻¹) slib zou bevatten, bij een influentconcentratie $C_{so} = 0,4$ kg C/m³ (vetzuren) een hoeveelheid van 8 kg C/m³d of 20 kg COD/m³d kan worden omgezet. Bij een influentconcentratie van 3,0 kg C/m³ kan zelfs 25 kg C/m³d of 60 kg COD/m³d worden omgezet. De zuiveringsefficiency zal in het interval 92,5 - 99 % liggen. Uit deze resultaten kan worden gekonkludeerd, dat de anaerobe zuivering van vetzuren houdend afvalwater technologisch zeer aantrekkelijk is voor industriële toepassing.

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87. D.E. Simpson, *Water Res.*, 5 523 (1971).
88. Continuous System Modeling Program III, program no. 5734-XS9, IBM.

LIST OF SYMBOLS

a	specific production rate unidentified products	s^{-1}
a	constant; yield (Chapter II)	..
A	surface area	m^2
b	statistical constant	..
b	width	m
C	response on a pulse stimulus	-
C	concentration	$kg\ m^{-3}$
d	distance; diameter	m
F	response on an up-step stimulus	-
F	flow	$m^3\ h^{-1}$
f	fraction	-
h	height	m
K	constant	..
k	constant	..
l	length	m
m	specific substrate conversion rate for maintenance	s^{-1}
m	statistical constant	..
me	specific methane production rate	s^{-1}
n	number of measurements	-
P	parameter	..
q	specific substrate conversion rate	s^{-1}
r	correlation coefficient	-
r	production or conversion rate per m^3 reactor volume	$kg\ m^{-3}\ s^{-1}$
S	substrate	kg
S	sensitivity	-
t	time	s
V	volume	m^3
v	velocity	$m\ s^{-1}$
X	amount of sludge	kg
Y	sludge yield factor	-
y^*	sludge yield factor	-
α	association constant	-
Δ	inaccuracy; variation; interval	..
$\emptyset_g$	volume of gas per unit of time	$m^3\ h^{-1}$
$\emptyset''_g$	volume of gas per m^2 per unit of time	$m\ h^{-1}$
$\emptyset'''_g$	volume of gas per m^3 per unit of time	h^{-1}
$\emptyset_m$	mass flow	$kg\ h^{-1}$

ϕ_s''	specific conversion of substrate	$\text{kg m}^{-3} \text{d}^{-1}$
ξ	conversion	-
μ	specific growth rate of sludge	s^{-1}
θ	relative time	-
σ	standard deviation	-
τ	residence time	h

LIST OF INDICES AND ABBREVIATIONS

ap	unidentified products
b	sludge bed
bd	from bed to blanket
c	carbon in gas
d	blanket; decay (Chapter II)
db	from blanket to bed
ds	dead space
e	effluent
eff	effective
f	fine sludge or mud
fa	fatty acids
g	gas
gb	gas produced in the sludge bed (1 atm)
gb'	gas at the top of the sludge bed
gd	gas produced in the blanket (1 atm)
k	bypassing
L	linear
Li	lithium
m	solids; biomass; sludge
max	maximum
mb	sludge from the bed
md	sludge from the blanket
me	methane
min	minimum
mo	biomass in the influent
nd	non defined
o	influent
o'	originally coming from the influent
o''	originally coming from the influent
p	sludge particles
pf	plug flow
R	reactor

s substrate; settling
so substrate in the influent
sm biomass converted as substrate
sp sample point
ST total carbon
st settler
sg substrate-gas
ta area of transition
ta' at the top of the area of transition
tb transport of sludge from the bed
td transport of sludge from the blanket
w wake
x x-direction

BOD biological oxygen demand
C carbon
CHN Carbon-Hydrogen-Nitrogen
COD chemical oxygen demand
E effluent
FI flow indication
FIR flow indication, recording
FRC flow recording, controlling
G gas
I influent
L liquid
LC level controlling
RTD residence time distribution
S settler; solids
SM stirring mechanism
SP sample point
SS suspended solids
TC total carbon
TOC total organic carbon
VSS volatile suspended solids

Kleinschalige methaangisting is zelfs als lokale deeloplossing van het wereld-energievraagstuk ongeschikt.

R.B. Singh, *Compost Sci.*, 13 20 (1972).

P.J. Meynell, *Methane : planning a digester*, Prism Press Dorchester, Engeland (1976).

Methane generation from human, animal, and agricultural wastes, National Academy of Sciences, Washington D.C. (1977).

L.J. Fry, *Practical Building of Methane Power Plants for Rural Energy Independence*, Santa Barbara, California : Standard Printing (1974).

Het model van Monod en daarvan afgeleide kinetiekmodellen zijn niet geschikt om de procesdynamica van anaerobe waterzuivering te beschrijven.

J.F. Andrews, *Proc.Ind.Waste Conf.*, 23 285 (1968).

De kleine waarde van de tijdkonstante (0,25 - 1,0 h) voor een stapvormige versterking van de influentsubstraatconcentratie van opstroomreactoren voor anaerobe waterzuivering van vetzuren bevattend afvalwater, is een garantie voor de stabiliteit van dit proces.

J.W. Buijs, *Afstudeerverslag TH-Delft*, (1978).

De door Lettinga en medewerkers geconstateerde achteruitgang in de zuiverings-efficiency en de methaanproductie van een met een oplossing van suikerextract belaste continue reaktor voor anaerobe waterzuivering, die na verloop van tijd bij lage belastingen (ca. 5 kg COD/m³d) optrad, is het gevolg van niet op elkaar afgestemde hoeveelheden zuurvormende en methaanvormende bacteriën in de reaktor. Dit kan worden voorkomen door deze bacteriën in twee verschillende reaktoren onder te brengen.

G. Lettinga, A.G.N. Janssen en P. Terpstra, *H₂O*, 8 530 (1975).

P.M. Heertjes and R.R. van der Meer, *Proc.Ind.Waste Conf.*, 34 (1979).

De door Williamson en McCarty aanbevolen snelle methode om de verzadigings-konstante in het kinetiekmodel van Monod te bepalen, heeft als voordeel dat inderdaad snel resultaten worden verkregen, echter als nadeel, dat deze resultaten onbruikbaar zijn voor continue, gemengde kulturen.

K.J. Williamson and P.L. McCarty, *Biotechnol.Bioeng.*, 17 915 (1975).

STELLINGEN

1

Juist in een tijd van grote jeugdwerkloosheid is het aan te bevelen om de militaire dienstplicht uit te breiden tot een maatschappelijke dienstplicht en deze verplicht te stellen voor elke Nederlandse vrouw en man in de leeftijdsgroep van 18 - 22 jaar.

2

De rustpauzes in een concert dienen als een overgang tussen de onderdelen en hebben derhalve een muzikale functie. Het verbreken van de rust door gehoest en geschraap van kelen, waarvoor het publiek deze pauzes denkt te moeten gebruiken, is derhalve af te keuren.

3

Het uitoefenen van een leraarsfunctie direkt na het behalen van een universitaire graad, dient te worden vermeden.

4

Door het toepassen van slip tussen de banden van een persbandfilter, kan slib goed worden ontwaterd.

J. Ooms, afstudeerverslag TH-Delft, 1977.

5

Voor de ontwikkeling van doelmatige slibontwateringsapparatuur zou onderzoek moeten worden verricht naar de afhankelijkheid tussen de ontwaterbaarheid, de rheologische eigenschappen en de compressibiliteit van slurries.

P. Aarne-Vesilind, Treatment and Disposal of Wastewater Sludges, Ann Arbor Science (1974).

Derek B. Purchas (ed), Solid-liquid separation, equipment, scale-up. Uplands Press Ltd., Croydon Engeland (1977).

6

Bij het onderzoek naar niet verslavende pijnstillers dient meer aandacht te worden besteed aan de enzymatische afbraak van pijnstillers in het lichaam, en in het bijzonder aan een mogelijke restbinding tussen achtergebleven afbraakprodukten en de receptoroppervlakken op de voor het verminderen van pijn actieve plaatsen.

S.H. Snyder, Chem.Eng. News., Nov. (1977) 26.