INTEGRATED APPLICATION OF THE UASB REACTOR AND PONDS FOR DOMESTIC SEWAGE TREATMENT IN TROPICAL REGIONS

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Up this earth, a man cannot possibly make
one step in a straight and a direct line.
John Donne, 1572-1631

To Adrianus, Tomaz, Yuri, Arturo,
Giovana and Úrsula.

Abstract

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Waste stabilization ponds are widely applied for domestic sewage treatment in Brazil. The main objective of conventional waste stabilisation ponds (WSP's) is, *nomen est omen*, to remove the organic material from wastewater. To achieve this objective, a quite long liquid retention time (R_h) is required: even under the favourable conditions of the tropics (high temperature, sunshine) a minimum liquid retention time of 20 to 30 days is necessary. It is well established that the optimal ponds configuration for domestic sewage treatment consists of a serial system composed of an anaerobic pond, followed by a partially aerobic (facultative) and one or more predominantly aerobic (maturation) ponds.

By substituting the anaerobic pond by an efficient high rate anaerobic reactor like the Up Flow Anaerobic Sludge Bed (UASB) reactor, the pond receiving the effluent of this reactor – an polishing pond – will receive a relatively very low organic load and therefore the design criteria for this pond are very similar to those for maturation ponds, that is the polishing pond is designed for pathogen and/or nutrient removal.

In semi-arid regions, like the North-East of Brazil, the use of treated sewage for irrigation water is an ecologically sound solution if precautions are taken to avoid contamination of the workers and of the consumers of the produced crops. According to the very sticky World Health Organization (WHO) standards for unrestricted irrigation, the helminth concentration must be less than 1 per litre and the faecal coliform (FC) concentration less than 10^3 colony forming unit (CFU) per 100 ml. On the other hand, other important fact in design of pond systems is the evaporation from the pond needs to be minimal, and consequently the pond area must be as small as possible. When UASB effluent (with low organic material and suspended solids concentrations) is post-treated in polishing ponds, these systems can be designed specifically for *FC* removal. For that reason, their surface area can be kept smaller than that of conventional ponds, designed for organic material removal. For reducing the liquid retention time (and surface area) of a polishing pond it is necessary to minimize the degree of mixing as much as possible, so that the hydraulic regime approaches a plug flow regime.

This thesis deals with aspects of design, performance and operation of sewage treatment systems consisting of an UASB reactor followed by a polishing pond. The thesis is directed to an experimental pilot plant investigation for assessing design and operation parameters with respect to the performance of the UASB reactor followed by post-treatment in different configurations of polishing ponds. With such a treatment system it was possible to produce an effluent that can be well used for unrestricted irrigation. The advantage of the system is that it allows a substantial reduction in the required surface area compared to the area required in conventional waste stabilization pond systems.

As far as UASB reactor design concerned, it was shown that the phase separator design is of paramount importance to the performance of the system. The addition of parallel plates to the conventional phase separator could double the treatment capacity of the UASB reactor. The experimental investigation also showed that the sludge age and not the liquid retention time is the fundamental operational parameter of the UASB reactor: UASB reactors with different liquid retention times but the same sludge age will tend to have the same efficiency.

An important aspect of UASB operation is the discharge of excess sludge. In this thesis a method was developed to estimate the frequency of excess sludge discharge as well as the optimal fraction of the total that can be discharged. It was shown that discharges of less than 50% of the sludge mass had a small and transitory effect on the UASB performance.

Polishing ponds were operated both with a continuous and sequential batch feeding. Based on the experimental results of the investigations, relevant operational and conceptual aspects were elucidated and elements for dimensioning and designing polishing ponds could be provided.

It was shown that the required liquid retention time for reduction of the BOD_5 and TSS concentrations to a virtually constant value only amounts to about 3 days, which is very short i.e. a factor 7 to 10 times shorter than the liquid retention time (25 - 30 days) in conventional WSP's. In this investigation the FC concentration in the digested sewage was of the order of 10^7 per 100 ml, so that the required removal efficiency for unrestricted irrigation ($FC < 10^3$ CFU/100 ml) was 99.99 % (4 log units). By using the value of the experimentally determined decay constant (2.0 to 2.2 day^{-1} at a depth of 0.65 m) the minimum time for this removal efficiency was calculated at 4 to 5 days for exponential decay. However the experimentally observed removal efficiency in a flow-through pond was much lower than expected on the basis of first order kinetics (exponential decay). This was attributed to the fact that partial mixing could not be avoided, even though the polishing pond was specifically designed to approach the plug flow regime as closely as possible. For this reason a minimum liquid retention time of about 10 days was required for a removal efficiency of 99.99 %.

Mixing in continuous reactors like polishing ponds was quantified by determining the residence time distribution of the liquid by applying a slug of a tracer in the influent and observing its concentration as a function of time in the effluent. The experimental data obtained in these investigations showed that even in the pond that was carefully designed and operated to avoid mixing as much as possible, the dispersion number was of the order of 0.1 to 0.2, which means that mixing intensity was moderate. The tracer studies also revealed a surprisingly high dead volume fraction in the pond. When the values of the dispersion numbers and the dead volume fractions were used to calculate the theoretical faecal coliform removal efficiency in the pond from the Wehner and Wilhelm equation, it was established that there was a good correlation between theory and experimental data.

Exponential decay for a first order process like FC removal can be guaranteed in a batch reactor. In this thesis the behaviour of ponds fed with sequential batches (SB) of anaerobically treated sewage was also evaluated. It was shown that the reduction in the FC concentration, was a good approximation, to the exponential decay of a first order process until a value of 1000 CFU/100 ml is attained, so that a very high removal efficiency (> 99.99%) of FC was obtained in a very short period of 4 to 5 days, as foreseen by theory for the experimental decay constant of 2.0 to 2.2 day^{-1}. During this period complete elimination of helminth eggs also occurred. In addition, due to the efficient pre-treatment, an aerobic environment prevailed in the sequential batch (SB) ponds and a substantial further reduction of the BOD_5 and TSS concentrations took place. Thus a final effluent was produced in ponds with a liquid retention time of only 4 to 5 days with a quality that complies with WHO standards for unrestricted irrigation. The liquid retention time required in a SB pond was only about half of the value for a flow-through polishing pond, so that its application leads to a reduction of the required pond area by a factor of 2. In conventional waste stabilisation pond systems the R_h time under tropical climate conditions is of the order of 25 days, which is a factor 5 to 6 times greater than in polishing ponds operated in sequential batch mode for the production of an effluent for unrestricted irrigation.

Other important finding of the investigations was that the rate of bottom sludge accumulation in polishing ponds remained much smaller than the values reported for conventional waste stabilization ponds, even though no sludge was discharged from the UASB reactor. The reduced bottom sludge accumulation has an important practical advantage: the frequency of desludging operation will be much lower than in anaerobic ponds and this operation may be unnecessary during the useful life span of the polishing pond

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Chapter 1

Introduction

Two important points were considered when defining the scope of this thesis: to contribute to a low cost solution for municipal wastewater treatment and at the same time to produce an effluent in accordance with local legislation for unrestricted irrigation. In Brazil there has been a lack of political will to implement basic sanitation for the population and this has been reflected in a lack of resources for investments in the construction of sewerage systems and treatment plants. Thus low cost solutions are necessary to minimize problems of pollution and contamination by municipal wastewater. Anaerobic treatment systems like septic tanks, Imhoff tanks and anaerobic filters as well as waste stabilization ponds are now widely applied. The most comprehensive report on low cost wastewater treatment plants in operation in Brazil [Andrade Neto,1997] shows that waste stabilization ponds and notably the so-called Australian configuration (an anaerobic pond followed by a facultative unit and one or more maturation units) is the most widely applied system for municipal sewage treatment. At the time of the report (1997) it was estimated that at least 700 waste stabilization ponds had been constructed in Brazil, with the highest number in the State of São Paulo. Most of these ponds were constructed in the 1970's and in many cases now operate under organic and hydraulic overload. As a result they have a poor performance and produce an effluent with unsuitable quality for final disposal in surface waters and/or utilization in agriculture or other productive activities.

The State of Paraíba, in North-East Brazil has 223 municipalities, of which only 11 have sewage treatment services, all managed by the State Sanitation Company, CAGEPA. A recent publication [König *et al.*, 2002] showed the result of an evaluation of the performance of 8 of the 11 systems in operation in the municipalities of Guarabira, Sapé, Campina Grande, Patos, Cajazeiras, Sousa, Monteiro and Itaporanga, indicated in color on the map of Figure 1.1. All these systems were waste stabilization ponds. Table 1.1, presents the configuration and the average values of the main parameters: BOD_5, *SS*, faecal coliforms (*FC*) and helminth eggs (*HE*) determined in the influents and effluents of the 8 evaluated systems. The report does not give information about the organic loads that was apply, so that it is not possible to state if the ponds are overloaded or not. However it can be concluded that the evaluated systems had poor performances with BOD_5 removal efficiencies ranging from 65 to 90%, *SS* removal between 22 and 81% and *FC* removal less that 3 logs (99.9%). Only helminth eggs were completely removed.

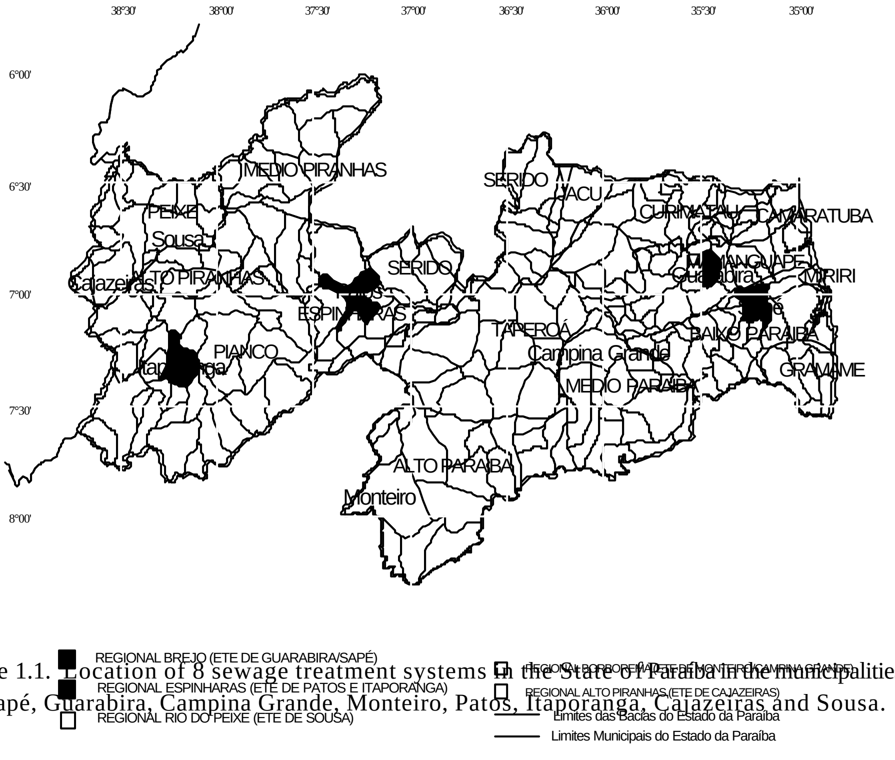

Figure 1.1. Location of 8 sewage treatment systems in the State of Paraíba in the municipalities of Sapé, Guarabira, Campina Grande, Monteiro, Patos, Itaporanga, Cajazeiras and Sousa.

Table 1.1. Configuration and average quality of raw sewage and effluents at 8 waste stabilization pond systems in operation in the State of Paraíba in North-East Brazil.

Municipality	Configuration	Parameter BOD_5 (mg/*l*)	 *SS* (mg/*l*)	 *FC* (log *CFU*)	 HE (eggs/*l*)
Campina Grande	2 AnP in series	334-54	224-65	7.8-5.5	45-0
Cajazeiras	Primary FP	118-41	88-66	6.9-4.3	11-0
Guarabira	2 series of AnP+FP	589-71	386-125	7.7-6.4	36-0
Itaporanga	Primary FP	428-43	328-61	7.7-5.9	16-0
Monteiro	2 series of AnP+FP	548-129	427-333	7.5-6.3	21-0
Patos	AnP	227-52	203-58	7.8-6.6	7-0
Sapé	Primary FP	434-68	318-183	7.2-5.4	47-2
Sousa	Primary FP	187-52	129-180	7.3-5.6	14-0

The state of Paraíba is located in North-East Brazil, a region with limited water resources. The scarcity is due to low and irregular rainfall. In several regions the storage of water is not even sufficient to reliably satisfy public demand. It goes without saying that under these circumstances there is also a lack of water for other uses such as food production. For this reason it is common practice to use effluents from sewage treatment plants for irrigation, ranging from animal fodder to lettuce. In regions with a lack of natural water resources, treated sewage may in part substitute irrigation water, supplying at the same time valuable nutrients. Before usage can be considered, wastewaters must first be treated to reduce the concentrations of suspended solids, organic material and pathogenic organisms. The latter requirement proves to be the most restrictive in practice. The World Health Organization guidelines for unrestricted irrigations (WHO, 1989) specifies the maximum concentrations of 1 helminth egg per litre and of 10^3 colony forming units (*CFU*) of *FC* per 100 ml.

Upflow anaerobic sludge blanket (UASB) reactors are very efficient in removing suspended solids (*SS*) and organic material (BOD_5 and *COD*) from wastewaters, when they are operated in regions with a tropical or subtropical climate. However, due to the short liquid retention time in the reactor, the removal of pathogens is only partial. Hence a post treatment unit is required, if the effluent is to be used for irrigation. In this thesis the application of polishing ponds proved to be an attractive option because it combines low investment and operational costs with simple operation and maintenance and high reliability. On the other hand, for waste stabilization ponds already in operation and demonstrating poor performance, substitution of the anaerobic pond by a UASB reactor may result in the production of an effluent quality compatible with the standards for environmental protection without the need for extension of the existing structures.

In this thesis it will be shown that the combination of the UASB reactor with a polishing pond (PP) has important advantages when compared with conventional waste stabilization ponds:

- PPs treating effluent from an efficient anaerobic pre-treatment unit receive such a low organic load that they are comparable with maturation ponds in conventional pond systems and for that reason the governing design criterion will be the removal of pathogens to be meet the standard recommended for final disposal:
- when polishing ponds are operated as sequencing batch ponds or flow through ponds with low dispersion, the required area is much smaller than that for a conventional pond systems, so that not only are costs drastically reduced but water losses due to evaporation are also much smaller. On the other hand the performance of currently overloaded conventional pond systems may be improved by substituting the anaerobic pond by a UASB reactor;
- the substitution of the anaerobic pond by UASB reactor can improve the performance of organically overloaded WSP's and also eliminate or greatly reduce the problem of odours.

Thus the aim of the present thesis is to evaluate the possibility of improving the performance of waste stabilization ponds and at the same time reducing costs by including an efficient anaerobic pre-treatment unit like the UASB reactor. To attain this general aim the following specific objectives were identified:

Objective 1 – to define design criteria and operational conditions to effect efficient suspended solids (*SS)* and organic matter (OM) removal in the UASB reactor, so that the main design criterion of the PP becomes pathogen instead of OM removal.

Although the feasibility of using the UASB reactor for removal of *SS* and OM has been amply demonstrated in full scale plants, the performance of these reactors can be further improved by innovations in design and operation. In Chapter 2 it will be shown that more efficient *SS* and OM removal can be obtained if sludge retention is maximized by having a more efficient phase separator and applying an adequate sludge wastage strategy. It will be shown that the presence of parallel plates above the conventional phase separator device leads to an appreciable improvement in effluent quality. The parallel plates enable the UASB reactor to retain sludge particles that could not be captured in the conventional phase separator and the contact between the small particles settled on the parallel plates apparently promotes flocculation and the return of bigger, better settling, sludge particles. The reduction in sludge loss due to “wash-out” promotes a longer sludge age, defined as the ratio between the sludge mass and the daily sludge loss. In Chapter 2 it will be shown that, in fact, the sludge age is the fundamental operational variable, the hydraulic retention time has importance only in that it must be sufficient to provide the necessary volume to accommodate the sludge mass.

In Chapter 3 a procedure is described for the strategy of sludge wastage. In this case an attempt has been made to develop a simple operational procedure that minimizes the frequency and time required for sludge wastage, making sure that at the same time a good effluent quality is assured at all times. It will be shown that large sludge discharges (50 % of the total mass or more) can be applied without impacting negatively on the reactor performance. The maximum sludge discharge and the discharge frequency depend on several factors such as: the applied organic load, the hydraulic retention time, the maximum sludge mass that can be retained in the reactor, the minimum sludge mass required for proper performance of the reactor and the specific sludge production.

Objective 2 – to obtain a final effluent quality compatible with the WHO guidelines by operating continuous flow polishing ponds for the post treatment of UASB effluent, seeking to reduce as much as possible the required *per capita* area.

In Chapter 4 the operation of continuous flow and sequencing batch polishing ponds is described. The liquid retention time and pond depth were identified as the main operational variables that affect the effluent quality.

In the case of continuous flow ponds the flow regime was evaluated by tracer studies and the dispersion number was determined for different retention times. The experimental data showed that the flow regime deviated substantially from the desired plug flow regime that would have been ideal for *FC* removal, even when the ponds were specifically designed for minimum mixing by having a very high length/width ratio. The mixing would appear to be due to unavoidable factors such as winds, density gradients, movements of microorganisms and rising of biogas bubbles. The experimental results obtained with a polishing pond with five compartment, having a total length/width ratio of 50/1 and dispersion numbers of 0.14 and 0.18, showed that after a liquid retention time of 3 days there was a complete removal of helminth eggs, whereas the *FC* concentration became less than 10^3 /100 ml with a minimum liquid retention time of 10 days. At this retention time the filtered BOD_5 was equal to 25 mg/*l*, which is in conformity with the standards of the *Council European Communities* (CEC, 1991) for discharge into surface waters.

The experimental data indicated that for a large range of depths (0.65 to 3 m) the decay constant of the *FC* was inversely proportional to the pond depth. This has an important consequence in that in principle the *FC* removal efficiency in a particular pond does not depend on the depth: the longer retention time in deep ponds (larger volume) compensates for the slower rate of *FC* decay. This is valid for all ponds independent of the flow regime.

Objective 3 – to obtain a final effluent compatible with WHO guidelines by operating sequential batch polishing ponds (SBPP) as a post treatment unit.

Reactor engineering theory shows that for the first order process of *FC* decay the highest efficiency is obtained in plug flow or batch reactors. In Chapter 3 it was shown that plug flow was not feasible, and as a consequence the required retention time was longer than the theoretical minimum. By operating ponds in sequential batch regime, it was possible to effect the *FC* reduction during the theoretical minimum retention time, which was considerable shorter than the necessary retention time in the continuous flow ponds.

Chapter 5 describes the configuration of two SBPP systems for the post treatment of UASB effluent. In the first configuration 5 parallel ponds were operated each being filled semi instantaneously every 5 days, with intervals of 1 day between consecutive ponds, so that this was the retention time in the ponds. In the second configuration, 4 parallel ponds with the same volume as in the first configuration were operated receiving the effluent from an accumulation or transfer pond, which in turn received the effluent from the UASB reactor. The transfer pond was operated with liquid retention times varying from 1 to 3.5 days. In both configurations the *FC* decay was exponential as is expected for the first order process and the required liquid retention time in the sequencing batch reactors was much shorter that in the flow-through ponds. The first configuration was marginally more efficient than the second one, but the latter could have some practical advantages.

Objective 4 – to evaluate the quantity and quality of bottom sludge formed in polishing ponds fed with effluent from a UASB reactor operated without sludge discharge.

Although the sludge produced in a UASB reactor is well stabilized for the sludge age normally applied in these units (> 50 days), a better effluent quality can be obtained by applying periodic sludge discharges as shown in Chapter 3. However in the UASB+PP system, the excess sludge from the UASB reactor can be discharged together with the effluent into the polishing pond, without substantially reducing the pond volume or measurably increasing the DO demand.

Chapter 6 deals with the qualitative and quantitative characterization of the bottom sludge accumulated in a pond that was operated over a period of 1 year, receiving raw UASB effluent together with the excess sludge. Over this period the accumulated sludge mass was 70 g/m^3 of treated sewage, which is about four times less than values found in ponds treating raw sewage. This means that in practice desludging of the pond may not be necessary during the useful life time of the pond system. This leads to an important reduction of the operational costs of the pond system. Due to the photosynthetic production of algae a high fraction of volatile solids was found in the bottom sludge (58 %). The high concentration of helminth eggs (651 per g *TSS*) hampers direct use of the dried sludge as fertilizer.

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Chapter 2

Influence of the Phase Separator Design on the Performance of the UASB Reactor and Excess Sludge Production

2.1 Introduction

The two main conditions for any well performing biological wastewater treatment system are: (1) to ensure good contact between the incoming substrate and the sludge mass in the system and (2) to maintain a large sludge mass in the system. In the UASB reactor the influent is distributed uniformly over the bottom of the reactor and then, following an up flow pathway, rises through a thick layer of anaerobic sludge, whereafter it is withdrawn at the top of the reactor. Thus the contact between the influent organic material and the sludge mass, in the reactor, is automatically guaranteed. In order to maintain a large sludge mass, the UASB reactor has a built-in phase separator, where the dispersed solids are retained by settling, so that an effluent virtually free from settleable solids can be discharged. The retained sludge particles will end up sliding back from the settler compartment into the digester compartment and accumulate there, thus contributing to the maintenance of a large sludge mass in the reactor.

Due to synthesis of biomass and to flocculation of particulate influent material, there is a continuous growth of the sludge mass in the reactor. However the reactor has a certain maximum sludge hold-up and once the reactor is "full", any new sludge production will lead to the loss of an equal sludge mass from the system. In that case, an increasing amount of sludge particles will be present in the effluent. From then onwards, the rate of (unintentional) sludge wastage will become equal to the rate of net sludge production and can therefore be determined experimentally as such. An alternative operational procedure is to discharge periodically part of the sludge mass (excess sludge), which should lead to a significant (though never complete) decrease of the sludge concentration in the effluent (Chapter 3, Section 3.2). In either case, it is possible to calculate the sludge age or the mean solids retention time. This is the ratio between the sludge mass present in the reactor and the rate of intentional + unintentional sludge wastage. It is important to emphasise here that, as a result of the solids retention mechanism, the solids retention time or sludge age (R_s) will always exceed the liquid retention time (R_h). The difference is more pronounced as the phase separator becomes more efficient. In a UASB reactor treating sewage in tropical regions, usually the R_h is of the order of 4 to 6 hours and the sludge age in the range of 20 to 50 days (van Haandel and Lettinga, 1994), so that $R_s/R_h \approx 120$ to 300.

The performance of the UASB reactor, as a unit for sewage treatment, under appropriate conditions is quite remarkable. In regions with a hot climate (i.e. with a sewage temperature above 18 ^oC), a very high removal efficiency of the organic material (65 to 80% of the influent COD) can be obtained in conventional UASB reactors with a short retention time (4 to 6 hours) (van Haandel and Lettinga, 1994 and Campos, 1999). This can be attributed mainly to the fact that through the application of a phase separator the sludge age becomes, at least in principle, independent of the liquid retention time. By maintaining a long sludge age, the sludge mass present in the reactor becomes high and this enhances efficient removal of biodegradable organic material.

The deterioration of UASB reactor performance at decreasing liquid retention times can be attributed to: a) the increasing inability of the phase separator to retain the sludge in the reactor for a sufficiently long period and b) the short time left for the retained sludge to convert the biodegradable and soluble COD. The increase of the COD fraction in the effluent is partly due to the presence of biodegradable influent material, which increases with shorter R_h. Also the sludge production increases when the R_h decreases because part of the influent particulate and biodegradable matter is discharged before hydrolysis can take place. Hence organic matter mixed with bacterial sludge will be present in the excess sludge. More efficient sludge retention could lead to a decrease of escaping particulate organic material, thus reducing COD fraction discharged together with the effluent. Moreover, efficient sludge retention would cause an increase in the sludge age in the reactor, which by itself would increase the efficiency of hydrolysis and subsequent digestion of the soluble matter. Therefore, a decrease in the liquid retention time can be accomplished without reducing the performance of the UASB reactor, if the phase separator design is improved.

One possible way of achieving effective sludge retention is by applying parallel plates in the zone above the conventional phase separator, thus creating a high rate settler, a system often used in water treatment plants. A UASB reactor equipped with such a high rate settler will have a better performance than a reactor equipped with only a conventional phase separator, operating under comparable conditions. Equally, the reactor with a more efficient separator can accommodate higher organic loads and yet have a similar performance to the conventional reactor. Hence the R_h may be reduced when the phase separator performance is improved.

This Chapter deals with results of an experimental investigation on the influence of the phase separator design on UASB reactor performance. For this purpose the *COD* removal efficiency and net sludge production were observed as functions of the liquid retention time in two reactors with identical dimensions and receiving the same wastewater load, but equipped with different phase separator designs. The first reactor had a conventional UASB separator (triangular prism with an open base, (Figure 2 1(***a***)) and the second had an improved design (Figure 2 1(***b***)) by having parallel plates above the conventional design.

2.2 Sludge Retention in UASB Reactors with Different Phase Separator Designs

The conventional phase separator is composed of prismatic elements placed in the UASB reactor, dividing it in a lower digestion zone and an upper settling section (Figure 2 1(***a***)). Gas-liquid, gas-solid and solid-liquid separation occurs below the prismatic units at the interface of the liquid phase in the gas chamber. Additional solid-liquid separation occurs in the settling zone above the separator elements: particles with sufficiently high settling rates will overcome the drag force of the upward liquid flow and eventually will settle on a separator element and from there, after having built up a layer with enough flocculent mass, slide back into the digestion zone.

Due to the devices in UASB reactors (phase separators, deflectors, gas and effluent collection), the liquid velocity varies during the upward path until the effluent discharge.

Van Haandel and Lettinga (1994), considered three key values for the liquid velocity for a particular influent flow, Q: (a) the velocity in the digestion zone, v_1, which is defined by the area A_1 of this zone (b) the velocity v_2 at the level of the apertures of the phase separator with an area A_2, normally the smallest area in the reactor and (c) the velocity v_3 at the level of effluent discharge, defined by the transversal area, A_3, at the effluent discharge level, available for settling. The main design criterion is the definition of the liquid velocity in the digestion zone, v_1. In practice, this value usually does not exceed 1 m/h in UASB reactors (van Haandel and Lettinga, 1994). The other velocities are related to v_1 by the respective areas: $Q = v_1A_1 = v_2A_2 = v_3A_3$. Thus the liquid velocity at the effluent discharge level is expressed as:

$$v_3 = \frac{Q}{A_3} \tag{2.1}$$

If sludge flocculation did not occur in the conventional reactor, only particles with a settling velocity exceeding the minimum rising velocity of the liquid could be retained. Flocs with a smaller settling rate will be washed out of the system by the liquid flow and will be discharged together with the effluent. Hence the condition for retention of a particular floc is given by:

$$s_c \geq v_3 = \frac{Q}{A_3} \tag{2.2}$$

where s_c is the critical settling velocity for floc retention by the conventional phase separator.

In the alternative design depicted in Figure 2 1(***b***), the phase separator is composed of two parts. The first part is like that of the conventional separator and effects the separation of the biogas and part of the sludge from the liquid. The additional second part consists of parallel plates, which are placed to settle and thus retain flocs escaping from the conventional separator. Now the retention efficiency of the solids is given by the critical settling velocity in the zone with the parallel plates, which is significantly smaller than the minimum settling velocity for retention in the conventional separator.

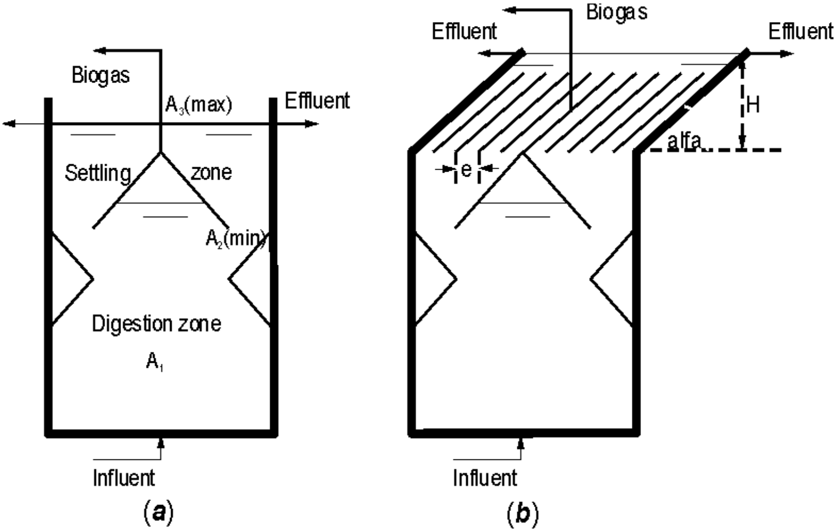

Figure 2.1. Different designs of phase separators for UASB reactors: (***a***) conventional, (***b***) with additional parallel plates.

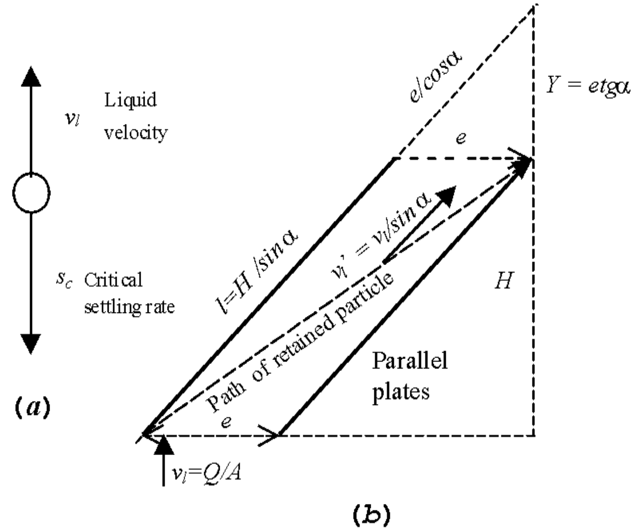

Figure 2.2. Representation of the settling mechanism for a conventional separator (***a***) and for a unit with parallel plates (***b***).

Figure 2.2 shows the path of a sludge particle moving between two plates: the particle enters the plate zone, next to the first plate and, as the liquid flows through the space between the plates, the particle settles and touches the second plate before the effluent leaves the plate zone. Such a particle would be retained and eventually be returned to the digestion zone. If the thickness of the plates is neglected the liquid velocity in the space between plates and the critical settling rate follows from:

$$v_l' = \frac{v_l}{\sin \alpha} \tag{2.3}$$

$$s_c' = \frac{v_l' . e . tg\, \alpha}{\dfrac{H}{\sin \alpha} + \dfrac{e}{\cos \alpha}} \tag{2.4}$$

where:

v'_l	flow velocity between plates ($v'_l = (Q/A_3).\sin \alpha = v_l / \sin \alpha$);
s'_c	critical settling rate;
α	angle of the parallel plates;
H	depth of the zone with plates;
e	spacing of the plates.

Hence the ratio of the critical settling velocity in a separator with parallel plates (Equation (2.4)) and in a conventional separator (Equation (2.2)) is given by:

$$s_c' / s_c = \frac{e . tg\, \alpha}{l \sin \alpha + e . tg\, \alpha} = \frac{1}{L \cos \alpha + 1} \tag{2.5}$$

where L is l/e, the ratio between the length "l" and the spacing "e" of the plates.

Equation (2.5) shows that the minimum settling velocity of the flocs that can be retained in the separator with parallel plates is always smaller than in the conventional separator. The difference becomes more pronounced as α become smaller and/or L larger. Hence for efficient floc retention by the parallel plates one must have: (1) a small distance between the plates (e), (2) a small angle of the plates with the horizon (α) and/or (3) long plates, i.e. a large depth of the zone with the plates (H). All three factors above are limited by practical considerations: (1) the distance between the plates cannot be very small to avoid problems with the removal of the sludge settled on the plates (blockages), (2) the angle of the plates must have a minimum value to make sure the settled sludge flocs will readily slide back into the digestion zone (in practice 45 to 60^o, (Valencia, 2000)) and (3) for economic reasons the depth of the zone with parallel plates cannot be very large.

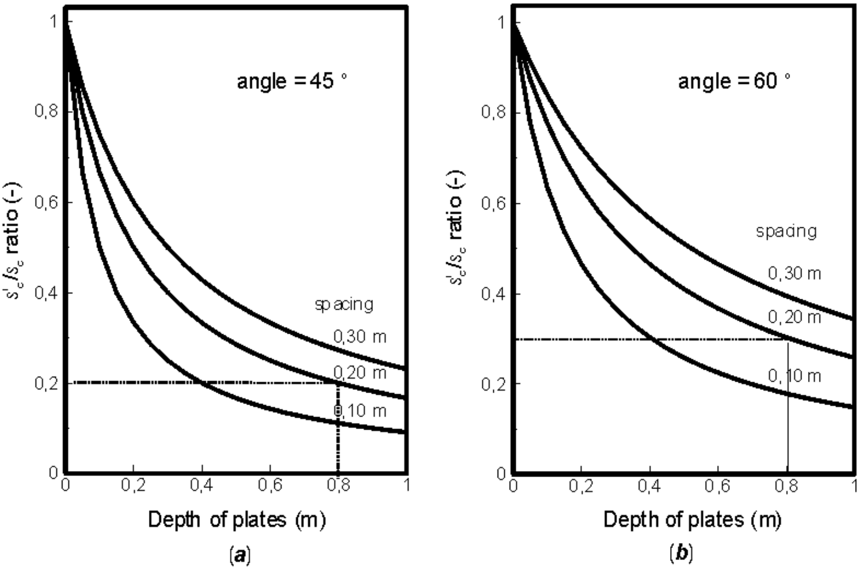

Figure 2.3. Ratio between the critical settling velocities in UASB reactors having both conventional separator and separator with parallel plates, as a function of the depth of the parallel plate zone for different plate spacings (*e*) and for (***a***) $\alpha = 45^o$ and (***b***) $\alpha = 60^o$.

Figure 2.3(***a*** and ***b***) shows the s'_c/s_c ratio as a function of the depth of the zone with plates for angles of 45 and 60^o. The ratios are calculated for three spacing between the plates: $e = 0.1$; 0.2 and 0.3 m. The results show that for practical values of *e*, α and *H*, theory predicts a large difference between the settling velocities of particles that can be retained in the reactor with parallel plates and in the conventional UASB reactor. For example, with plates at a distance of 0.20 m placed at an angle of 45^o and with a depth of 0.80 m, the settling velocity of flocs that can be retained in the reactor is a factor of only 0.20 or 1/5 of the minimum settling velocity of flocs that can be retained in the UASB reactor with a conventional separator.

Equally, it is possible to increase the hydraulic load of a reactor equipped with parallel plates and yet have the same efficiency of floc retention as, in the case, of the reactor with only a conventional separator. It is important to note that it is not possible to increase the sewage flow inversely proportional to the s'_c/s_c ratio, because, then, also the organic load (and consequently the sludge production) would increase. In Figure 2.3(***b***) it can be seen that for the same spacing and depth, considered in the example above but at an angle of 60^o, the ratio s'_c/s_c is 0.30, which is higher than in the case of 45^o, so that less particles are retained, but on the other hand the settled sludge will slide more easily back into the digestion zone.

An additional reason for the superior performance of the separator with parallel plates is that part of the floc with a settling velocity smaller than s'_c will still be retained, depending on their position between plates, when they enter the plate zone. By contrast, in the case of the conventional separator, if no flocculation occurs, all flocs with a settling velocity smaller than the rising velocity of the liquid will be washed out with the effluent.

2.3. Experimental Investigation and Results

The experimental investigation was carried out using two pilot scale UASB reactors of the same size with the objective of evaluating the effect of the phase separator design on the performance of the UASB reactor. The first reactor (A) had a conventional phase separator and the second (B) was equipped with additional parallel plates. A sketch of reactor (B) is presented in Figure 2.4. The UASB reactors were fed with a constant flow of raw municipal sewage from the main outlet of the sewer system of Campina Grande city in Brazil.

The digestion zone of the reactors was formed by two connected concrete rings with height of 1m each and a diameter of 0.80 m. The conventional phase separator, made of plexy glass, was placed in the brickwork section above the concrete elements and, in Reactor (B), on top of the conventional phase separator, parallel plates (also made of plexy glass) were placed. The plates had a length of 0.5m (depth of 0.35m) and they were placed at an angle of 45^o with a spacing of 0.07m. The reactors were operated under identical operational conditions. The liquid retention times were varied from 12 to 1.5 hours. After imposing a particular liquid retention time, the reactors were operated for a period of not less than 1 month before collecting the experimental data. The assessed parameters were related to: (1) operational stability (2) organic material removal efficiency and (3) net sludge production and composition. With respect to operational stability the effluent pH, total alkalinity ($TAlk$) and volatile fat acids (VFA) concentration were determined. For the sewage characteristics in Campina Grande ($TAlk$ = 350 mgCaCO$_3$/l, COD < 1000 mg/l and T = 25 ± 2^oC), the buffer index of the sewage was always sufficient to maintain the pH in the neutral range of 6.8 to 7.1. On the other hand, methanogenesis was always efficient: the VFA concentration in the effluent never exceeded 1 mmol/l (60 mg HAc/l), usually it was less than 0.5 mmol/l. This was true for both reactors and for the entire range of investigated retention times. As a consequence the operational stability was excellent throughout the investigation and there was no risk of souring.

Both reactors were operated without out intentional discharge of excess sludge, so that the maximum sludge mass was built up, whereafter sludge washed out at the same rate it was produced in the reactor. The reactors were operated at constant flow rates and the experimental data were collected only after the maximum sludge hold-up had been established for each of the investigated liquid retention times.

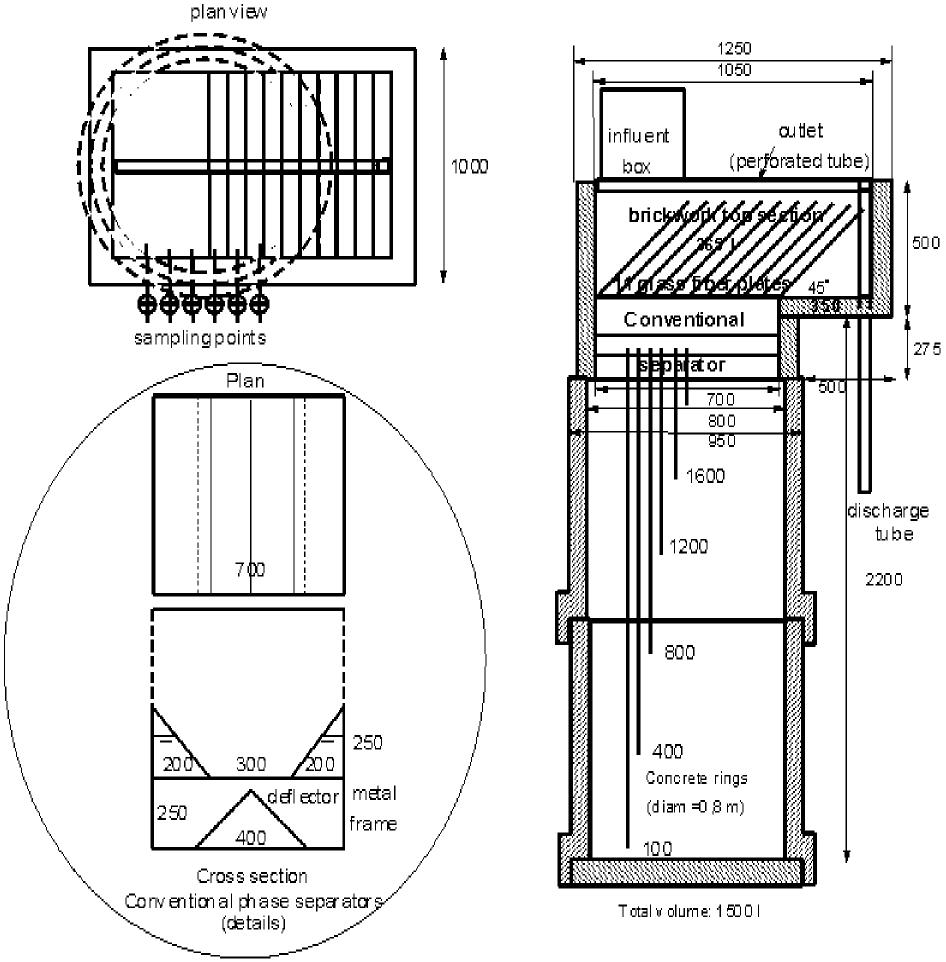

Figure 2.4. A representation of the UASB Reactor B used in the experimental investigation (values in mm).

In relation to the settleable fraction of the total suspended solids (*TSS*), it was considered that any particles settling in an Imhoff cone during 30 minutes were sludge particles and that the *COD* of the supernatant liquid was the true effluent *COD*. For this reason, both the raw and settled effluent *COD* were determined, the difference between the two effluent *COD*s was used to estimate the volatile sludge concentration in the effluent and hence the net sludge production in the reactor. Knowing that the *COD* of volatile sludge is, approximately, f_{cv} = 1.5 *gCOD/gVSS* (Marais and Ekama, 1976), the effluent volatile sludge concentration was calculated as:

$$X_{ve} = \frac{(S_{re} - S_{se})}{1{,}5} \tag{2.6}$$

where:

X_{ve} concentration of volatile sludge in the effluent;

S_{re} raw effluent *COD* concentration;

S_{se} settled effluent *COD* concentration.

Table 2.1. *COD* concentrations of influent (S_{ti}), raw (S_{re}) and settled effluent (S_{se}) and *VFA* for different R_h, in the conventional UASB reactor (A) and in the unit with an improved phase separator (B).

	COD (mg/*l*) and *VFA* (mgHAc/*l*) concentrations						
		Reactor A			Reactor B		
R_h (h)	Influent (S_{ti})	Raw Effluent (S_{re})	Settled Effluent (S_{se})	VFA	Raw Effluent (S_{re})	Settled Effluent (S_{se})	VFA
12	587	157	88	18	147	72	26
10	492	143	78	22	128	59	20
8	554	189	108	18	158	94	23
6	480	186	102	24	133	67	28
4	526	252	133	38	185	93	45
3	619	360	195	33	210	106	87
2	561	454	236	57	264	154	78
1.5	613	-	-	72	407	214	69

Table 2.1 shows the influent *COD* concentrations as well as the raw (S_{re}) and settled (S_{se}) *COD* effluent values and VFA effluent concentrations, for the reactors A and B, for the different retention times (R_h). In Table 2.2 the sludge mass and composition, in terms of g*TSS*/*l* and its volatile fraction are also presented. The sludge mass (total and organic) was calculated from linearized concentration profiles, using the sludge concentrations at the sample points (Figure 2.4). The volatile sludge production was estimated from the difference between the raw and settled effluent *COD* concentrations. The sludge age was calculated as the ratio between the volatile sludge mass in the reactor and the daily

production found in the effluent (Equation (2.6)), during the steady state operational conditions.

Table 2.2. Sludge concentration data (*TSS, g/l*), its volatile fraction (*VF*) and sludge age (R_s, days), as functions of liquid retention time R_h (hours) of reactors A and B.

	Sludge concentration, composition and sludge age					
	Reactor A			Reactor B		
R_h	R_s	*TSS*	*VF*	R_s	*TSS*	*VF*
12	122	20.6	0.54	205	36.6	0.56
10	98	18.0	0.57	155	29.7	0.58
8	58	16.1	0.58	120	27.0	0.57
6	44	16.0	0.61	64	19.7	0.57
4	21	17.5	0.65	47	28.2	0.61
3	13	16.9	0.67	39	35.6	0.61
2	6	14.6	0.68	17	23.7	0.63
1.5	-	-	-	11	28.8	0.68

2.4 Discussion

In the UASB reactor the influent *COD* (S_{ti}) is divided into three fractions: a fraction which leaves the system with the effluent, mS_e; a fraction which is converted into sludge, mS_x and a fraction digested to methane, mS_d. The objective of the anaerobic treatment is to minimize the first two fractions and to maximize the third. The first two fractions can be determined from data in Table 2.1, and the third can be calculated by subtracting these two fraction from unity:

$$mS_e = \frac{S_{se}}{S_{ti}} \tag{2.7}$$

$$mS_x = \frac{(S_{re} - S_{se})}{S_{ti}} \tag{2.8}$$

$$mS_d = 1 - mS_e - mS_x \tag{2.9}$$

where:

S_{ti}	influent *COD* concentration;
mS_e	fraction of *COD* influent present as organic material in the settled effluent;
mS_x	fraction of *COD* influent transformed into volatile sludge;
mS_d	fraction of *COD* influent digested in the reactor.

In Figure 2.5(***a***) the values of mS_e, mS_x and mS_d, calculated from the data in Tables 2.1 and 2.2 using Equations (2.7) to (2.9), are shown for both reactors A and B, as functions of the applied R_h. On the basis of these experimental data, empirical curves were drawn for the organic material fractions as a function of R_h for reactor A (interrupted curve) and B (drawn curve). Clearly, both fractions mS_e and mS_x increase, as the R_h. decreases. This is contrary to the objective of the anaerobic treatment systems, where these fractions must be the minimum. The curves also indicate a very strong influence of the phase separator on the performance of UASB reactor in terms of treatment efficiency. For the same digested *COD* fraction, the required retention time in reactor B is about half the value required in reactor A. Hence, the parallel plates in reactor B led to a doubling of its treatment capacity.

The data in Tables 2.1 and 2.2 can also be used to plot the three *COD* fractions as functions of the sludge age (log scale). This is shown in Figure 2.5(***b***). If the sludge age is used as the independent variable, the *COD* fractions mS_e and mS_x in reactors A and B, in good approximation, can be described with a single curve. This means that, for any particular sludge age, the fractions of the influent *COD* ending up in the effluent or converted into sludge is always the same, independent of the phase separator design or the liquid retention time that is applied. It can be concluded that the sludge age and not the liquid retention time is the relevant parameter to describe the performance of the UASB reactor for sewage treatment under specific conditions, e.g. temperature, sewage characteristics, etc. Figure 2.5(***a***) and 2.5(***b***) also reveal that, for the two reactors, there is a minimum R_h below which no methanogenesis will occur and consequently all the organic material will leave the reactor either in the effluent or as flocculated material in the excess sludge. This minimum R_h depends on the phase separator efficiency.

Unfortunately knowledge about UASB reactor design is still insufficient to make an *a priori* estimate of the UASB reactor sludge age, with a particular phase separator design and under specified operational conditions. This parameter can only be determined *a posteriori*, when the reactor has been built and is operating. The sludge hold-up does not only depend on the phase separator design, but also on the mechanical properties of the sludge, particularly its settling velocity and flocculation tendency and its mineral fraction. These properties depend on the operational conditions in the reactor as well as the influent characteristics. It appears that theory is not yet sufficiently developed to give an accurate estimate of the settling velocity of UASB sludge for sewage treatment.

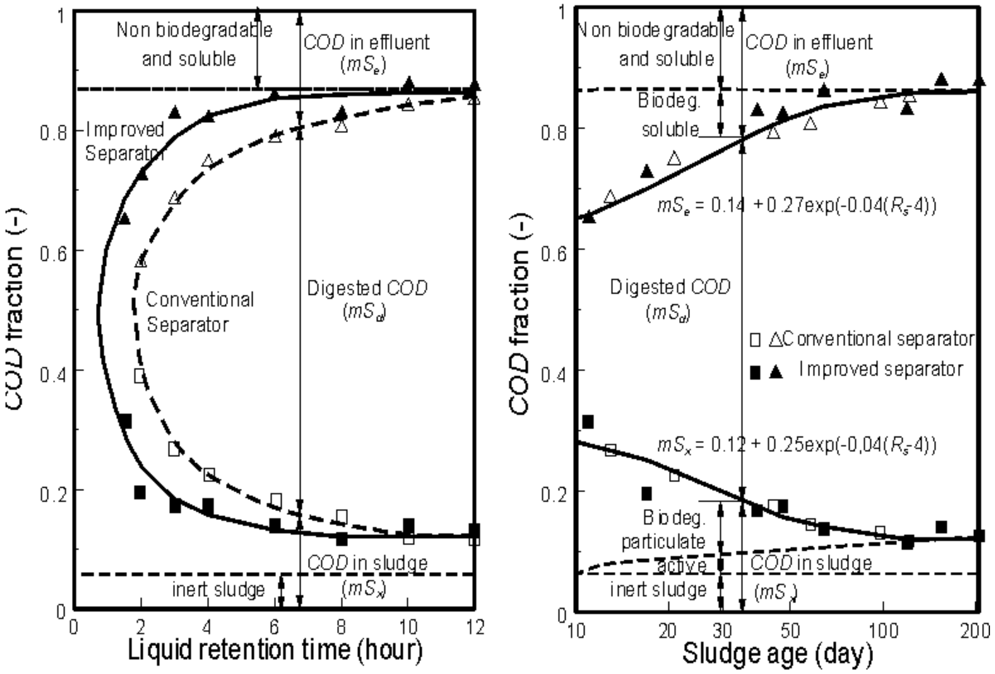

Figure 2.5. *COD* fraction in the influent, in the excess sludge and digested as a function of (***a***) R_h in reactor A (interrupted curve) and B (drawn curve) and (***b***) R_s reactor A and B.

Once the sludge age has been identified as the fundamental operational variable, it is interesting to express the *COD* fractions mS_e and mS_x as a function of this parameter. The following empirical expressions were found by trial and error from the data in Table 2.1 and 2.2 or Figure 2.5(***b***):

$$mS_e = 0.14 + 0.27 \exp\left[-0.04\left(R_s - 4\right)\right] \tag{2.10}$$

$$mS_x = 0.12 + 0.25 \exp\left[-0.04\left(R_s - 4\right)\right] \tag{2.11}$$

So that:

$$mS_d = 1 - mS_e - mS_x = 0.74 - 0.52 \exp\left[-0.04\left(R_s - 4\right)\right] \tag{2.12}$$

As the expressions of Equations (2.10), (2.11) and (2.12) were obtained from experimental data (Tables 2.1 and 2.2), their validity is restricted to the operational conditions during the experimental investigation.

Figure 2.5(***b***) is also very useful for evaluating of the composition of the *COD* fractions present in the effluent and converted into the sludge. Marais and Ekama (1976) divided the influent *COD* material in a biodegradable and a non biodegradable fraction, each with a soluble and a particulate component. At very long sludge ages, it may be assumed that the biodegradable material is completely used by the bacteria, so that only non biodegradable material persists (top section indicated in Figure 2.5(***b***)). At shorter sludge ages, there is a progressive increase in the presence of biodegradable material in the effluent as well as in the produced sludge. It is interesting to note that methanogenesis remained efficient even at sludge ages as short as 6 days (see Table 2.2): the average *VFA* concentration always remained below 60 mg/*l* (Table 2.1). Hence the increased biodegradable *COD* concentration at shorter R_s must be attributed to inefficiency of the processes preceding methanogenesis, *viz* hydrolyses and/or acidogenesis.

In Figure 2.5(***b***) the *COD* fraction in the effluent does not fall below a value of 0.14, even at very long sludge ages (R_s > 200 days). Apparently this value represents the fraction of non biodegradable and soluble material in the influent (f_{us}). Based on this value the fraction of biodegradable material in the effluent *COD* can be calculated as a function of the sludge age.

$$mS_{be} = mS_e - f_{us} = 0.27 \exp\left[-0.04\left(R_s - 4\right)\right] \tag{2.13}$$

Thus the fraction of biodegradable material in the effluent can be estimated as follows:

$$f_{be} = \frac{\left(mS_e - f_{us}\right)}{mS_e} \tag{2.14}$$

$$f_{be} = \frac{0.27 \exp\left[-0.04\left(R_s - 4\right)\right]}{0.27 \exp\left[-0.04\left(R_s - 4\right)\right] + f_{us}} \tag{2.15}$$

where:

mS_{be}	fraction of the biodegradable *COD* influent in the effluent;
f_{us}	non biodegradable soluble *COD* fraction in the effluent;
f_{be}	*ratio* biodegradable/ total *COD* in the effluent;
$(mS_e - f_{us})$	biodegradable *COD* fraction in the effluent.

Figure 2.6(***a***) shows the value of f_{be} as a function of the sludge age (logarithmic scale) for the data in Tables 2.1 and 2.2.

Similarly, the composition of the *COD* in the sludge may be evaluated. From the results in

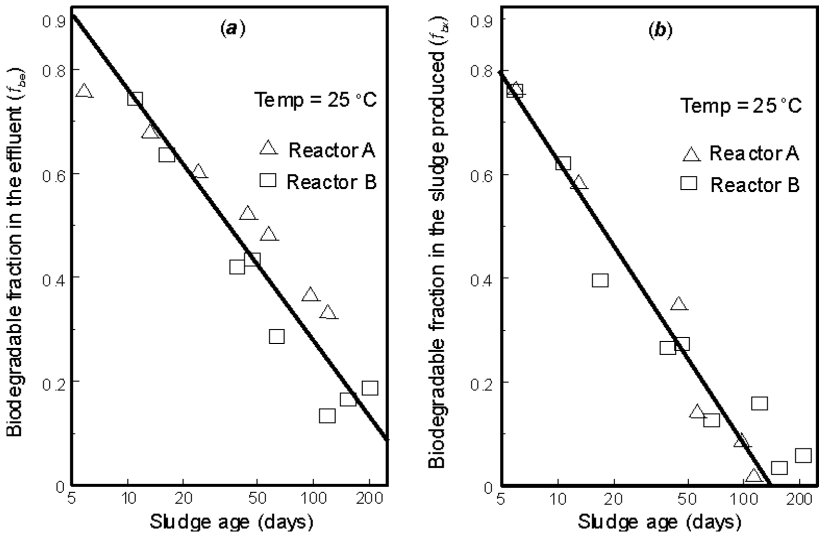

Figure 2.6. Fractions of biodegradable *COD* in (***a***) the effluent and, f_{be} (***b***) in the produced sludge, f_{bx}, as functions of the sludge age (logarithmic scale).

Figure 2.5(***b***) it is apparent that at very long sludge ages, the fraction of the influent *COD* converted into sludge becomes constant. Under these circumstances, sludge production may be attributed to two factors: (1) flocculation of non biodegradable particulate matter, forming the inert sludge mass fraction and (2) the biological sludge mass that is generated as a result of anabolic conversion in the reactor. It is considered that decay of the anaerobic sludge proceeds very slowly and that therefore the build-up of endogenous residue (the non biodegradable rest of decayed biological sludge) is negligible. Figure 2.5(***b***) and Equation (2.11) show that the minimum *COD* fraction converted into sludge is 0.12.

In order to estimate the proportion of the inert and bacterial material in the sludge, values must be attributed to the non biodegradable fraction of *COD* in the influent (f_{up}), that generates the inert sludge and to the yield coefficient, *Y*, which determines the bacterial sludge production. The inert material in the excess sludge is equal to non biodegradable and particulate *COD* in the influent, f_{up}:

$$mS_{xi} = f_{up} \tag{2.16}$$

where mS_{xi} is the influent COD fraction that is discharged as inert material in the excess sludge;

The fraction of *COD* converted into bacterial mass is much more difficult to estimate. If a unit mass of COD is metabolised, the anabolised and catabolised fractions can be estimated as:

$$f_{ana} = f_{cv}.Y \tag{2.17}$$

and

$$f_{cat} = 1 - f_{cv}.Y \tag{2.18}$$

where:

f_{ana}	COD influent fraction converted into bacterial mass;
f_{cat}	COD influent fraction catabolised;
f_{cv}	conversion factor between COD and VSS; 1.5 gCOD/gVSS (Marais and Ekama, 1976);
Y	yield factor or bacterial mass production per unit mass metabolised organic material (g VSS/gCOD).

It is important to note that the yield coefficient in Equations (2.17) and (2.18) is the "true" value as opposed to the apparent yield factor, Y_{ap}, which is calculated as the ratio between the observed volatile sludge production and the applied organic load. The latter value also includes particulate biodegradable and non biodegradable influent material that is incorporated in the sludge. For that reason Y_{ap} is always higher than Y and its value tends to increase as the sludge age decreases. Hence at very long sludge ages, when the biodegradable organic material, $(1 - f_{us} - f_{up}).S_{ti}$, is metabolised a fraction $f_{cv}Y$ is anabolised. Hence the fraction of influent COD converted into bacterial mass is:

$$mS_{xa} = f_{cv}.Y(1 - f_{us} - f_{up}) \tag{2.19}$$

where mS_{xa} is the fraction of influent COD converted into bacterial mass.

So, the minimum sludge production, which occurs at very long sludge ages, can be expressed as:

$$mS_{x,\min} = mS_{xi} + mS_{xa} = f_{up} + f_{cv}.Y(1 - f_{us} - f_{up}) \tag{2.20}$$

The expression above does not have an explicit solution because there are two unknowns (Y and f_{up}) and only one equation. The characteristic value of the non biodegradable and particulate fraction, determined in an aerobic activated sludge plant treating sewage in Campina Grande city was f_{up} = 0.06 (van Haandel and Marais, 1999). Although the non biodegradable fractions are not necessarily the same in aerobic and in anaerobic environments, it was adopted as the best estimate. Now the Y value can be calculated as follows. At very long sludge ages $mS_{v,min}$ = 0.12 with f_{us} = 0.14 and $f_{up} = mS_{xi}$ = 0.06 (Figure 2.5(***b***)), the value of Y can be calculated with the aid of Equation (2.20): Y = 0.05 mg *VSV*/mg *COD*. These constants can now be used to calculate the sludge composition at shorter sludge ages.

At any sludge age the fraction of influent *COD* that is catabolised in the treatment process can be expressed as Equation (2.9). Hence the influent *COD* fraction converted into anabolised material corresponds to Equations (2.12) and (2.17):

$$mS_{xa} = \frac{f_{cv}.Y}{(1 - f_{cv}.Y).mS_d} \tag{2.21}$$

Therefore the concentration of biodegradable influent material in the sludge can now be expressed as:

$$f_{bx} = (mS_x - mS_{xa} - S_{xi})/mS_x$$

$$f_{bx} = (mS_x - f_{cv}.Y/(1 - f_{cv}.Y).mS_d - f_{up})/mS_x \tag{2.22}$$

Figure 2.6(***b***) shows the composition of the excess sludge as a function of the sludge age in terms of the fraction of biodegradable influent *COD*. It can be noted that this fraction is very high when the sludge age is short. Figure 2.6(***b***) shows that at a sludge age of 50 days the biodegradable fraction of the sludge still amounts to about 24% i.e. 5 to 10 g*VSS*/*l* or 7.5 to 15 g*COD*/*l* for an assumed volatile solids concentration of 20 to 40 g/*l*. These values are much higher than the influent *COD*, normally in the range of 0.5 to 0.8 g/*l*.

A criterion that could be used to establish the minimum sludge age for a satisfactorily low biodegradable material concentration in the excess sludge of a UASB reactor would be the EPA recommendation (EPA, 1992) that defines stable sludge (for use in agriculture) “as sludge with the mass of volatile solids in the sewage sludge is reduced by at least 38% during sludge treatment” or “if it losses less than 17% additional volatile solids when it is anaerobically batch-digested in the laboratory in a bench-scale unit at 30 ^{o}C to 37 ^{o}C for an additional 17 days” (stability test). According to the results in Figure 2.6(***b***), for the case of this experimental investigation, the minimum sludge age would be 42 days.

Once the convenient sludge age of an anaerobic treatment unit has been established (for example, R_s = 42 days), the required R_h can also be estimated on the basis of the known influent *COD* concentration (and composition) and the prevailing average sludge concentration in the UASB reactor. This can be demonstrated as follows: the sludge mass in a UASB reactor can be expressed either as the product of the sludge age and the daily sludge production, which itself is the product of the apparent yield coefficient (Y_{ap}) and the organic load ($Q.S_{ti}$) or as the product of sludge concentration and reactor volume:

$$MX_t = R_s.ME_t = R_s.Y_{ap}.Q_i.S_{ti} = V_r.X_t$$

or,

$$R_h = \frac{V_r}{Q_i} = \frac{R_s.Y_{ap}.S_{ti}}{X_t} \tag{2.23}$$

where:

X_t	total sludge concentration (average);
MX_t	total sludge mass in the reactor;
ME_t,	daily total excess sludge discharge or daily total sludge production;
V_r	UASB reactor volume.

Equation (2.23) shows that the R_h is inversely proportional to the average sludge concentration in the treatment system. In practice this sludge concentration will depend very strongly on the efficiency of the phase separator: the more efficient it is, the higher a concentration can be maintained and therefore the higher is the treatment capacity of the reactor. In practice of municipal sewage treatment (where flocculent rather than granular sludge develops), the average sludge concentration usually is in the range of X_t =15 to 35 g/*l*; the lower value being typical for conventional systems and the higher for systems with efficient sludge retention. Considering a *COD* influent concentration of 800 mg/*l* and assuming a value of Y_{ap} = 0.15 *TSS*/g*COD* for R_s = 42 days (in accordance with experimentally found data (van Haandel and Lettinga, 1994)), the range of retention times for the minimum sludge concentration of 15g *VSS*/*l* to the maximum of 35 g *VSS*/*l*, R_h would be from 8 to 4 hours. It can be noted that the calculated R_h values for conventional reactors are compatible with the typical range found in real scale units.

2.5 Conclusions

The sludge age is the fundamental parameter to describe the performance of the UASB reactor for sewage treatment and, particularly, the division of the influent *COD* into three fractions: (1) discharged in the effluent, (2) converted into sludge and (3) digested to methane.

UASB reactors treating the same sewage at different liquid retention times will tend to have the same effluent quality and sludge production if the sludge age is the same. The shorter the sludge age, the higher are the fractions of the influent *COD* ending up in the effluent or in the excess sludge. The experimental data indicate that stable sludge production can be expected for a sludge age of about 40 to 50 days at 25 ^{o}C. At lower temperatures a longer sludge age is required.

The sludge age is strongly dependent on the efficiency of the sludge retention device of a UASB reactor. The application of parallel plates, in addition to a conventional phase separator, to form a high rate settler is extremely efficient. In this investigation, their application in a pilot UASB reactor (plates at 45^{o} with a depth of 0.35 m and with a spacing of 0.07 m) effectively doubled the treatment capacity of the reactor.

The required liquid retention time (and hence the reactor volume) depends on the average sludge concentration in the reactor which, in turn, is a function of the sludge retention efficiency of the phase separator.

Reduced efficiencies of the preparatory processes of hydrolysis, acidogenesis and acetogenesis, rather than methanogenesis itself are the cause of poor performance at short sludge ages.

2.6 References

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Chapter 3

Excess Sludge Discharge Frequency for UASB Reactor

3.1 Introduction

Biological treatment of wastewaters leads to the growth of sludge, so that there is a gradual increase of the sludge mass stored in the treatment system. Once the storage capacity is exhausted, one needs to discharge excess sludge or unintentional wash-out of sludge together with the effluent will occur. Without intentional sludge discharge, the rate of wash-out in the effluent is equal to the net sludge production rate. On the one hand it is desired to maintain a maximum sludge mass in the system, in order to achieve the maximum treatment capacity. On the other hand sludge wash-out must be avoided, because it decreases the effluent quality. In principle, two strategies can be adopted to minimize the presence of sludge in the effluent: (1) removal of the sludge in a post treatment unit or (2) avoidance of unintentional wash-out through planned discharges of excess sludge, before the sludge bed fills up the reactor completely. The sludge discharge may be continuous or semi continuous, if mechanical dewatering equipment is used like a centrifuge or belt filter. If a natural dewatering process is used, like a sludge drying bed, relatively large discharges of sludge are made at long intervals.

When applying a batchwise excess sludge discharge, it is important to know the fraction of the sludge mass in reactor that can be wasted. To simplify the operation of the treatment system, it is advantageous to apply large discharges with long intervals between these. However these discharges should be restricted, the remaining sludge mass always needs to exceed a certain minimum to prevent that the treatment efficiency and the operational stability are strongly affected.

In the present investigation the influence of the magnitude of the excess sludge discharge from UASB reactors on their performance was evaluated, in terms of organic matter removal efficiency and operational stability. It was observed that, for a liquid retention time (R_h) in the range of 4 to 8 hours, the performance of the pilot scale UASB reactors was hardly affected by discharges up to 50 to 60% of the maximum sludge mass.

3.2. Experimental Procedure and Results

The experimental investigation was carried out over a 10 months period, using three identical pilot scale UASB reactors (R1, R2 and R3) each with a volume of 126 litres and a height of 4 meters. The reactors were equipped with a phase separator in the form of a side arm placed at an angle of 45^o, through which the liquid flow passed before discharge. Figure 3.1 shows a schematic representation of one of the reactors. The reactors were fed with raw sewage from a trunk sewer in Campina Grande, at a constant flow maintained by dosing pumps. The concentration of pollutants varied with the sewage strength. The behaviour of the reactors was studied at R_h values of 4 and 8 hours, representing respectively a high and low loading for tropical countries. For each of the imposed liquid retention times, it was intended to evaluate: (1) the sludge production under steady state conditions and (2) the influence of different sludge discharges on the performance of the reactors. Table 3.1 summarizes the operational conditions in the reactors during steady state periods of the experimental investigation.

Sludge discharges were made from each reactor, operating at 8 and 4 hours of liquid retention time and under steady state conditions: at R_h = 8 hours 3 discharges were made, while 2 discharges were made at R_h = 4 hours. The percentages of the sludge mass released from the reactor as excess sludge for 8 and 4 hours of R_h are presented in Table 3.2.

Sludge production was determined first under steady state conditions on the basis of the sludge mass expelled from the reactor together with the effluent. The specific sludge production or yield coefficient (volatile and total sludge, Y_v and Y_t respectively) was determined for both liquid retention times. The sludge age under steady state conditions was determined as the ratio of the sludge mass in the reactors and the sludge production rate. Table 3.1 shows the experimental values of the specific sludge production and the sludge age for $R_h = 4$ and $R_h = 8$ hours.

Excess sludge was discharged, once the steady state sludge production had been established in the reactors, whereafter the recuperation of the sludge bed was observed until the reactor was "full" with sludge again and a new discharge could be made. The parameters used to characterize the performance of the UASB reactors after discharges of excess sludge were: (1) settled *COD* removal efficiency (2) operational stability reflected by *pH*, alkalinity and volatile fatty acids concentration and (3) sludge production and accumulation in the reactor. Table 3.3 shows the minimum and average settled *COD* removal efficiencies and *VFA* concentration, after discharges, when the reactors were operated under 4 and 8 hours of liquid retention time.

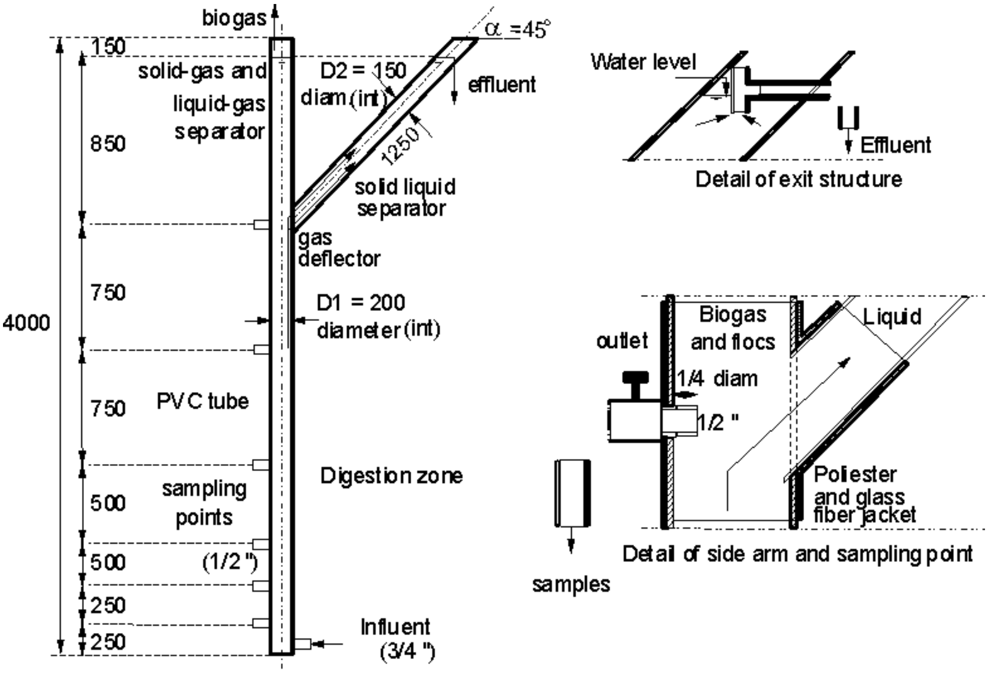


Figure 3.1. Schematic representation of the UASB reactor used in the investigation and detail of some special devices (sampling points, side arm, outlet).

Table 3.1. Average operational conditions and averages of performance parameters of the two phases of the investigation (R_h of 4 and 8 hours) during periods of steady state.

Parameter	Unit	R_h = 4 hours	R_h = 8 hours
Volume	*l*	126	126
Height	m	4	4
Flow rate	*l*.day^{-1}	756	375
Influent COD	mg.*l*$^{-1}$	589	622
Organic load	g*COD*.day^{-1}	442	233
Specific organic load	kg*COD*.m^{-3}.day^{-1}	3.5	1.8
Upflow velocity	m.h^{-1}	1.0	0.5
Temperature	^{o}C	28	25
Total sludge mass	kg *TSS*	2.95	3.35
Volatile sludge mass	kg *VSS*	1.62	1.76
Total sludge production	g*TSS*.day^{-1}	66	33
Organic sludge production	g*VSS*.day^{-1}	36	17
Specific total sludge production (Y_t)	kg*TSS*.kg^{-1}*COD*	0.15	0.13
Specific volatile sludge production (Y_v)	kg*VSS*.kg^{-1}*COD*	0.08	0.07
Sludge age	day	45	101

Table 3.2. Percentages of the sludge mass discharged in the experiments conducted at R_h = 4 hours (2 discharges in the reactors R_1, R_2 and R_3) and R_h = 8 hours (3 discharges in all reactors).

	R_h = 4 hours			R_h = 8 hours		
Discharges	R_1	R_2	R_3	R_1	R_2	R_3
1	52	66	77	22	48	54
2	54	63	81	24	46	56
3				60	78	81

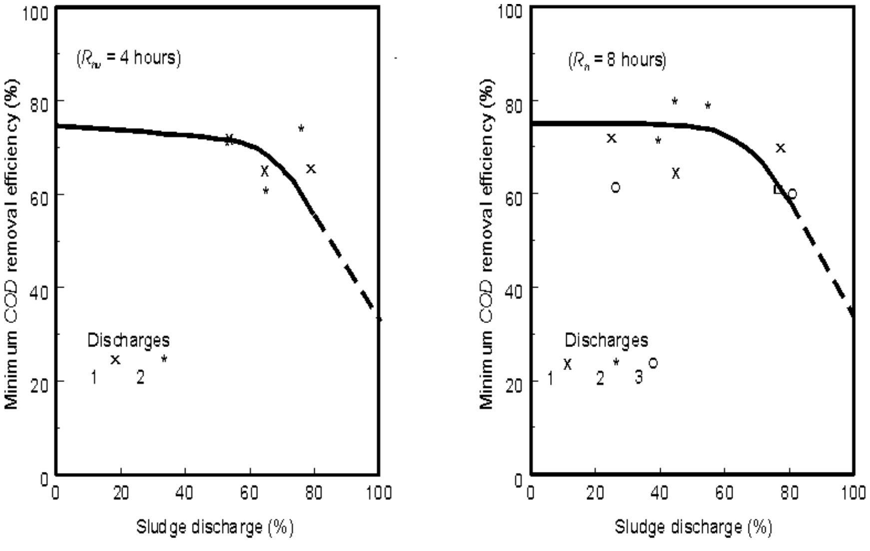
Table 3.3. UASB performance in terms of minimum and average *COD* efficiency removal and *VFA* concentration after discharges of sludge in the 3 reactors.

	R_h = 8 hours									R_h = 4 hours					
	COD removal (%)						*VFA*			*COD* removal (%)				*VFA*	
Reactor	minimum			average			mg*HAc/l*			minimum		average		mg*HAc/l*	
	1	2	3	1	2	3	1	2	3	1	2	1	2	1	2
R_1	63	72	62	80	80	69	32	25	29	72	71	81	74	28	32
R_2	71	80	60	79	83	68	29	26	32	64	59	79	65	32	41
R_3	70	79	63	79	80	73	44	46	40	64	75	76	62	55	53

Numbers 1, 2 and 3 refer to the first, second and third discharges.

3.2.1. Efficiency of Organic Material Removal

To establish the influence of sludge discharge on removal efficiency of organic material, the *COD* concentration in the effluent was observed as a function of time. For each discharge the following parameters were determined: (1) minimum *COD* removal efficiency on any particular day (normally the first day after sludge discharge) and (2) average *COD* efficiency in the interval between two discharges. Since, there were always sludge particles in the effluent, both the raw and the settled effluent *COD* were determined. Table 3.3 and Figures 3.2.and 3.3 show the minimum and average settled *COD* removal efficiencies, after discharges, when the reactors were operated under 4 and 8 hours of liquid retention time. The tests showed that after sludge discharges the *COD* removal efficiency tended to decrease, but this decrease was only significant in case of large discharges (higher than 50% of sludge mass).



(*a*) (*b*)

Figure 3.2. Values of the *minimum* settled *COD* removal efficiency after sludge discharges as a function of the percentage of wasted excess sludge for *Rh* = 4 (***a***) and 8 hours (***b***).

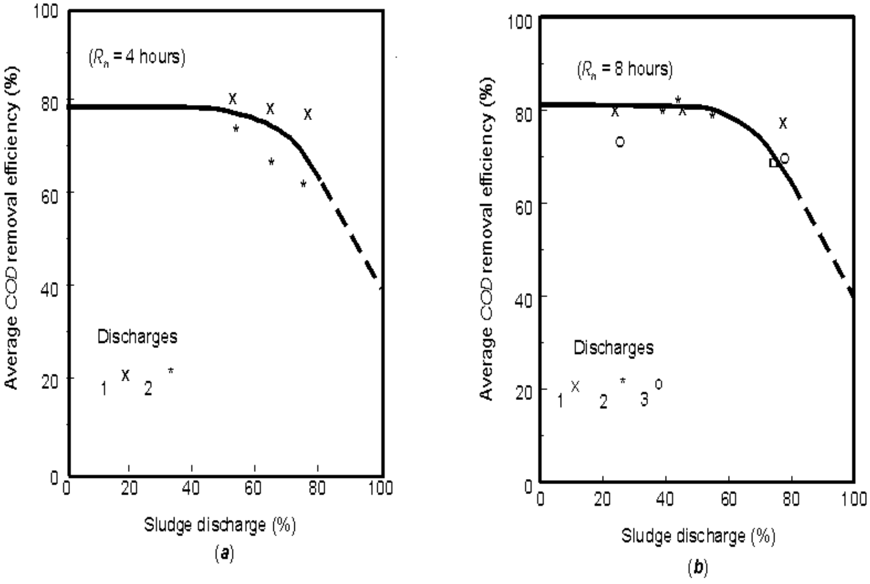


Figure 3.3. Values of the average settled *COD* removal efficiency between two sludge discharges as a function of the percentage of wasted excess sludge for (***a***) R_h = 4 and (***b***)8 hours.

3.2.2. Stability of the Anaerobic Digestion Process

With regard to operational stability of the reactor, the alkalinity (total and bicarbonate), *pH* and the concentrations of volatile fatty acids were determined. Table 3.3 shows the *VFA* concentrations after discharges when the reactors were operated under 4 and 8 hours of liquid retention time. Figure 3.4(***a***) shows the maximum daily concentrations of *VFA* for steady state on the day after different discharges of excess sludge for an R_h of 4 hours. Figure 3.4(***b***) shows corresponding data for R_h = 8 hours. It can be noted that the *VFA* concentration (usually less than 0.5 mmol HAc/*l*) always remained low and did not exceed 1 mmol.l^{-1} for any of the discharges. This means that methanogenesis was always efficient and that operational stability was always assured, even when very large discharges of 80% of the sludge were applied, although the *VFA* concentration did increase slightly with increasing sludge discharge. In Figures 3.2 to 3.4 the drawn curves are indicative for the experimentally observed reactor behaviour and the interrupted curves show the behaviour tendency for excessive sludge discharges (higher than 81%).

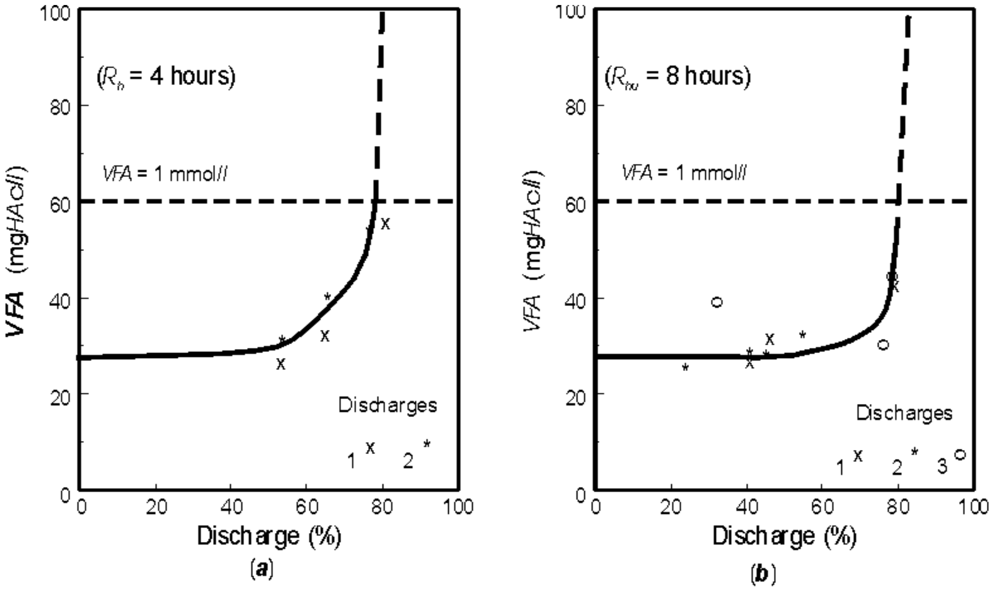


Figure 3.4. Values of the maximum *VFA* concentrations as a function of the discharged sludge mass percentage for (***a***) R_h = 4 and (***b***) 8 hours.

With regard to operational stability of the reactor, the alkalinity (total and bicarbonate), *pH* and the concentrations of volatile fatty acids were determined. Table 3.3 shows the *VFA* concentrations after discharges when the reactors were operated under 4 and 8 hours of liquid retention time. Figure 3.4(***a***) shows the maximum daily concentrations of *VFA* for steady state on the day after different discharges of excess sludge for an R_h of 4 hours. Figure 3.4(***b***) shows corresponding data for R_h = 8 hours. It can be noted that the *VFA* concentration (usually less than 0.5 mmol HAc/*l*) always remained low and did not exceed 1 mmol.l^{-1} for any of the discharges. This means that methanogenesis was always efficient and that operational stability was always assured, even when very large discharges of 80% of the sludge were applied, although the *VFA* concentration did increase slightly with increasing sludge discharge. In Figures 3.2 to 3.4 the drawn curves are indicative for the experimentally observed reactor behaviour and the interrupted curves show the behaviour tendency for excessive sludge discharges (higher than 81%).

3.2.3. Sludge production

In order to calculate sludge production following the discharges, it was considered that the sludge partially accumulated in the reactor, whereas the remainder washed out from the system with the effluent. The amount of sludge accumulated in the reactor was determined from sludge concentration profiles over the reactor height, by applying the following equation:

$$MX_t = \sum_{i=1}^{i=N} AL_i X_{ti} \tag{3.24}$$

where:

MX_t total sludge mass in the reactor;
AL_i volume of section "i" of the reactor with a concentration of X_{ti};
X_{ti} concentration of section number "i";
N number of sampling points ($N = 6$, see Figure 3.1).

The sludge profiles were determined weekly and, on the basis of the assessed profiles and calculated (MX_t) values, the rate of sludge accumulation in the reactor was determined. The daily sludge mass in the effluent was determined from the settleable solids concentration in the effluent. The data showed that the amount of settleable solids in the effluent remained substantially constant until the reactor became full of sludge, whereupon there was an increase of the settleable solids concentration in the effluent. From the data, the sludge mass leaving the reactor together with the effluent since the last sludge discharge was calculated as a function of time.

Finally the sludge production rate was calculated as the sum of the sludge accumulation rate in the reactor and the sludge loss rate in the effluent. The relative sludge growth rate was determined as the ratio between the sludge mass in the reactor and the sludge production rate. Figure 3.5 is an example of an R_h of 4 hours after a discharge of 54 % of the sludge mass. Figure 3.5(***a***) shows the accumulation of sludge (both *VSS* and *TSS*) in the reactor. After the discharge, the reactor initially had a sludge mass of 740 g*VSS* and attained 1952 g after 7 weeks, decreasing marginally to 1935 g in the eighth week. It was concluded therefore that after 7 weeks the reactor was full of sludge. The sludge accumulation data are somewhat irregular, because it was difficult to determine the sludge mass accurately; there were instances when biogas bubbles caused flotation of the sludge mass in the reactor. However, a tendency for linear sludge mass increase with time could be observed. The linearized sludge accumulation is also indicated in Figure 3.5.

Figure 3.5(*b*) shows an example of the unintentional sludge loss (*VSS* and *TSS*) in the effluent for the same conditions as in Figure 3.5(***a***). Again there is a tendency for a linear increase with time of the sludge mass discharged in the effluent. The conclusion that after 7 weeks the reactor was full is corroborated by the fact that in the eighth week an atypically high sludge loss rate was observed. Presumably, because no more sludge could be accommodated in the reactor, a sludge mass equivalent to entire production was expelled together with the effluent. Figure 3.5(***c***) shows the sludge production (*VSS* and *TSS*), expressed as the sum of the sludge accumulation (Figure 3.5(***a***)) and the sludge loss in the effluent (Figure 3.5(***b***)).

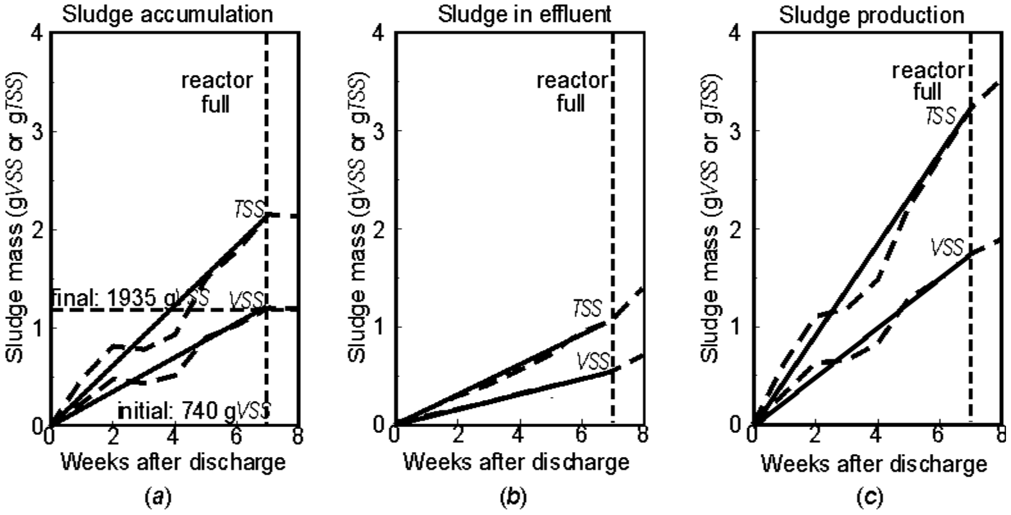


Figure 3.5.Values for the sludge mass accumulation in the reactor (***a***); expelled with the effluent (***b***) and total sludge production(***c***) as a function of the time after a sludge discharge of 54 % (R_h = 4 hours). Dashed and drawn lines are experimental data and linearized trends respectively.

The experimental data suggested the following behaviour pattern for the UASB reactors:

- before intentional sludge discharge, the sludge mass in the full reactor is constant and a sludge mass equivalent to the net production is discharged together with the effluent;
- after wastage of excess sludge, the sludge (settleable solids) concentration in the effluent decreases and maintains a constant value, while the sludge mass in the reactor increases; and
- when the sludge mass in the reactor attains its maximum value, sludge production again becomes equivalent to the discharged sludge in the effluent.

The sludge accumulation and sludge loss rates were determined from experimental values for the experiments with different sludge discharges at R_h = 4 and R_h = 8 hours. Table 3.4 shows experimental data for sludge loss in the effluent (g*VSS* and g*TSS* per week), whereas Tables 3.5 and 3.6 provide data on sludge accumulation and sludge production (g*VSS* and g*TSS* per week). The data show that an appreciable part of the produced sludge leaves the UASB reactor together with the effluent, even when the reactor is not full of sludge. For R_h = 8 hours there is an escape of 19 g of sludge per week, while the production is 112 g per week, so that a fraction of 19/112 = 17% is discharged involuntarily with the effluent. For an R_h of 4 hours this fraction increases to 79/240 = 32%. It is concluded that, even with the efficient phase separator in the UASB reactor, it was not possible to retain the entire sludge production in the reactors: about 1/6 of the produced sludge escaped with the effluent for R_h = 8 hours and approximately 1/3 for R_h = 4 hours.

Table 3.4. Sludge discharge rate in the effluent, in g*VSS* per week, at R_h = 8 and 4 hours, assessed during the operational steady state condition. The average influent *COD* is also indicated.

Reactor	R_h = 8 hours (*COD* = 622 mg/*l*)			R_h = 4 hours (*COD* = 589 mg/*l*)	
	Discharge 1	Discharge 2	Discharge 3	Discharge 1	Discharge 2
1	16.9	15.8	19.6	77	106
2	26.4	12.0	19.6	90	74
3	19.0	17.3	24.5	74	55
Average	20.8	15.0	21.2	80	78
Mean for R_h value	19.0			79	

Table 3.5. Sludge accumulation rate in the UASB reactor, in g*VSS* per week, at R_h = 8 and 4 hours.

Reactor	R_h = 8 hours			R_h = 4 hours	
	Discharge 1	Discharge 2	Discharge 3	Discharge 1	Discharge 2
1	137	86	115	152	173
2	81	109	64	138	127
3	79	39	125	124	252
Average	99	78	101	137	185
Mean for R_h value	93			161	

Table 3.6. Total sludge production rate (accumulation plus discharge in the effluent) in g *VSS* per week at R_h = 8 and 4 hours.

Reactor	R_h = 8 hours			R_h = 4 hours	
	Discharge 1	Discharge 2	Discharge 3	Discharge 1	Discharge 2
1	154	102	135	229	279
2	107	121	84	228	201
3	98	56	149	198	252
Average	120	93	122	217	263
Mean for R_h value	112			240	

Table 3.7. Specific sludge production for R_h = 4 and 8 hours.

R_h	g*VSS*/g*COD*			g*TSS*/g*COD*		
	Applied	Removed	Digested.	Applied	Removed	Digested
4	0.08	0.11	0.14	0.15	0.21	0.25
8	0.07	0.08	0.10	0.13	0.15	0.18

In order to evaluate the specific sludge production first the basis for calculations must be defined. There can be three interpretations: (1) sludge production per unit mass of applied *COD*, (2) per unit mass of removed *COD* and (3) per unit mass of digested *COD*. To calculate the sludge production per unit mass of applied *COD*, the daily sludge production should be divided by the influent organic load. For example: for R_h = 4 hours (flow rate of 126/4 = 31.5 *l*/h and a raw sewage influent *COD* of 589 mg/*l* the applied organic load is 31.5x24x0.589 = 442 g*COD*/day. Hence the specific sludge production or the apparent yield coefficient is calculated as 0.08 g*VSS*.g/g*COD* applied. Considering that the *COD* removal efficiency for the operational conditions was about 75% (see Figure 3.3) the corresponding value for the specific sludge production per unit mass of removed *COD* is calculated as 0.08/0.75 = 0.106g*VSS* per gram of removed *COD*. Knowing that the 0.106g of *VSS* produced per gram of removed *COD* represent a *COD* of 1.5x0.106 = 0.16g *COD*, the fraction of influent *COD* digested in the reactor is calculated as 0.75 – 0.16 = 0.59. Hence the specific sludge production per unit mass of digested *COD* is given by 0.08/0.59 = 0.135 g *VSS* per g of digested *COD*. Table 3.7 shows the values of specific sludge production per unit mass of applied, removed and digested *COD* for R_h = 4 and 8 hours.

3.3. Discussion

The experimental data showed that sludge mass discharges of 50 to 60% from UASB reactors treating domestic sewage at an R_h of 4 to 8 hours did not affect the effluent quality or the operational stability to any significant extent. At first sight it may appear surprising that such a large sludge discharge does not affect the performance of the reactor, but a more detailed analysis shows that indeed little effect is to be expected. For an R_h = 8 hours the maximum volatile sludge mass in the reactors was about 1760 g*VSS* (Table 3.1) so that a 60% discharge leaves 0.4*x*1760 = 680 g*VSS*. Under the circumstances the sewage flow is 126/8 = 16 *l*/h or 384 *l*/day. For an average *COD* concentration of 600 mg/*l* this means an organic load of 225 g*COD*/day. For a *COD* removal efficiency of 70% (raw effluent) the daily removed *COD* mass is 0.7x225 = 157 g*COD*/day. Hence the mass of *COD* removed per unit mass of sludge and per day (the required “in situ” sludge activity) is calculated as 157/680 = 0.23 g*COD*.g*VSS*$^{-1}$.day^{-1}. This value is small compared to the methanogenic activity determined for sludge from UASB reactors treating sewage at temperatures above 25 ^{o}C, which is in the range of 0.3 a 0.4 g*COD*.g^{-1}*VSS*.day^{-1} (van Haandel et al., 1992 and Haskoning, 1989).

On the other hand, similar calculations for a 60 % sludge discharge at an R_h of 4 hours show that the required “in situ” sludge activity for a 70% *COD* removal efficiency (raw effluent) would be 0.46 g*COD*.g^{-1}*VSS*.day^{-1}, which is a high value for sludge generated from raw sewage. In effect, the *COD* removal efficiency just after the discharge of 60% of the sludge mass at R_h = 4 hours was distinctly lower than 70% (see Figure 3.2) although the reactor rapidly recovered its usual performance. For discharges of 80% at R_h = 4 hours, the remaining sludge mass is heavily overloaded and the initial *COD* removal efficiency is low (Figure 3.2). The reactor was able to maintain the same average performance as shown for discharges of less than 50% (Figure 3.3). It is concluded that in order to maintain efficient UASB reactor performance after excess discharge, the treatment capacity of the remaining sludge mass must be compatible with the organic load of the influent.

In this context it is important to stress the importance of the phase separator: an efficient phase separator facilitates the retention of a large sludge mass in the reactor. For example, in the pilot scale reactors (126 litres) used in this experimental investigation, the maximum sludge mass was about 3kg*TSS* to 3.35 kg*TSS* (for R_h = 4 and 8 hours, see Table 3.1), which means an average sludge concentration of 24 to 26 g*TSS*/*l*, more than the mean concentration of sludge in UASB reactors with a conventional phase separator, which is of the order of 15 to 25 g*TSS*/*l* (Haskonig, 1989; van Haandel and Lettinga, 1994 and Vieira and Garcia, 1991). This is also borne out by the data in Section 2.1.

If a satisfactory performance of the UASB reactor is to be guaranteed it is necessary to maintain a minimum sludge mass in the reactor after discharge of a batch of excess sludge. The magnitude of the sludge discharge depends on the maximum hold-up of sludge in the reactor and the minimum necessary for proper performance. Neither of these two parameters are known and cannot be calculated theoretically. Hence they must be determined experimentally and for this experimental determination the following procedure can be suggested:

- establish experimentally the maximum sludge hold-up in the UASB reactor under normal operational conditions and without intentional sludge wastage;
- establish the minimum sludge mass that is required to maintain a well performing reactor in terms of *COD* removal efficiency and operational stability (this minimum may be estimated before the experimental determination, by assuming that the probable methanogenic activity of the sludge will be in the range of 0.3 to 0.4 $\mathrm{g}COD.\mathrm{g}^{-1}VSS.\mathrm{day}^{-1}$).

On the basis of the results for the maximum sludge hold up, two strategies may be followed (1) excess sludge is withdrawn when UASB reactor is full, i.e. when there is a sudden increase in the settleable solids in the effluent as can be determined easily by the Imhoff test or (2) establish a practical limit for the sludge hold up before the maximum is reached (for example 90% of the mass in the full reactor) and apply the sludge wastage when the calculation of the sludge mass from the concentration profile indicates that the practical maximum has been reached:

- determine experimentally the sludge production rate (this value may be estimated when the organic influent load and the sludge production per unit mass of applied *COD* are known) and the rates of sludge accumulation in the reactor and of sludge loss in the effluent, when the reactor is not yet full;
- calculate the period between the discharges as the ratio between the sludge mass to be accumulated in the reactor (=maximum less minimum sludge masses) and the sludge accumulation rate.

As an example the data for R_h = 4 hours may be used. It was experimentally established that the maximum sludge mass in the 126 *l* reactor was about 1600 g*VSS* (Table 3.1). From the tests, it is apparent that the reactor performance is hardly affected by sludge discharges of up to 50 % (Figures 3.2, 3.3 and 3.4). Thus an operational range is established for the sludge mass in the reactor with a minimum of 50% of 1600 i.e. 800 g*VSS* and a maximum of 90 % of 1600 g i.e. 1440 g*VSS*. Hence the sludge mass is varied between 800 and 1440 g*VSS*, so that the wasted sludge mass is 1440-800 = 640 g*VSS*.

In Table 3.7, for R_h = 4 hours the specific sludge production was determined 0.08 g*VSS*.g^{-1}*COD* applied, of which a fraction of 2/3 accumulates in the reactor and 1/3 is discharged together with the effluent (Tables 3.4 and 3.5). For the influent *COD* of 0.589 g/*l* and the flow rate of 24*x*126/4 = 756 *l*/day the organic load was calculated as 445g *COD*/day so that the sludge production rate is 0.08*x*445 = 36 g/day, of which 2/3*x*36 = 24g *VSS*/day accumulate in the reactor, while 1/3*x*36 = 12 g*VSS*/day escapes in the effluent. The period between sludge discharges is calculated as 640 g*VSS*/24g*VSS*.day^{-1} = 27 days.

Similarly for an R_h of 8 hours, accepting a maximum sludge mass of 1800 g *VSS* for the full reactor and an operational range between 40 to 90% (that established the minimum and maximum sludge masses of 720 and 1620 g*VSS*, respectively) and for a production rate of 0.07x225 = 15.75 g*VSS*/day, knowing that 5/6 of the produced sludge (13.1 g*VSS*/day) accumulates, whereas 1/6 (2.6 g/day) is discharged with the effluent, the period between excess sludge discharges is calculated as (1620-720)/13.1 = 69 days. Figure 3.6 (***a*** and ***b***) is a graphical interpretation of the examples. In practice, one would probably apply monthly discharges for R_h = 4 hours and bimonthly discharges for R_h = 8 hours.

In the above examples the importance of an efficient separator to reduce sludge losses due to wash out is apparent. In the examples, the sludge losses (12 and 2.6 g*VSS*/day for R_h = 4 and 8 hours, respectively) result in effluent sludge concentrations of 12/756 = 16 mg/*l* and 2.7/375 = 7mg/*l*, respectively. If the phase separator is inefficient, the concentration of settleable solids in the effluent will be higher than these values and the increase of unintentional sludge discharge, due to wash-out, will decrease the need for intentional excess sludge discharge. From the data of the examples, a sludge production of about 48 mg*VSS*/*l* or 90 mg*TSS*/*l* is estimated for an influent *COD* of 600 mg/*l* and an R_h of 4 hours. Hence, if the UASB reactor is unable to produce an effluent with a *TSS* concentration of less than 90 mg/*l*, the accumulation of sludge in the reactor would be insufficient to fill it with sludge and, consequently, there would never be a need for excess sludge discharges. An efficient phase separator will produce an effluent with a low settleable solids concentration and consequently the sludge accumulation rate will be high, which will result in a reduction of the start-up period.

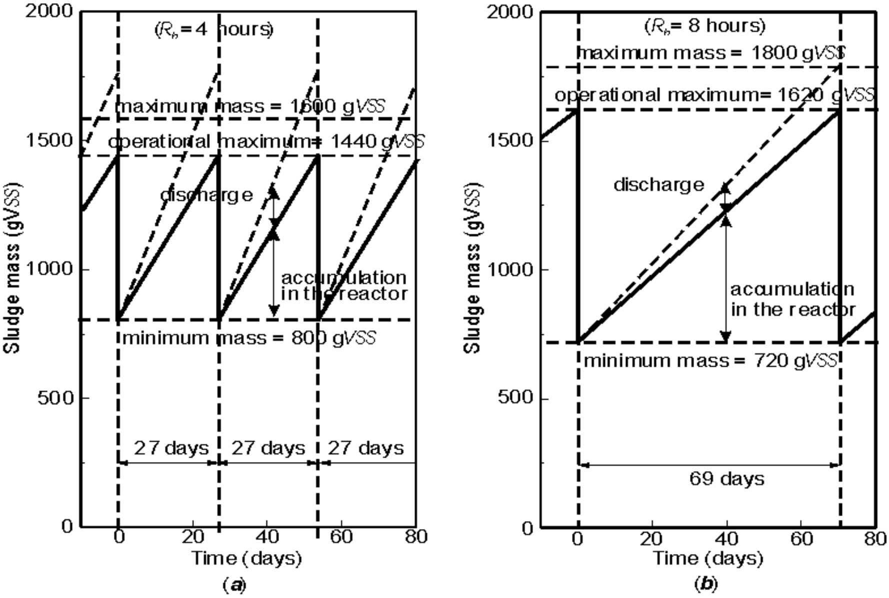


Figure 3.6. Schematic representation of the sludge mass variation in the UASB reactor with periodic discharges of excess sludge, as a function of time, that allowed the defining of the discharge frequency for R_h = 4 and 8 hours.

3.4. Conclusions

Normally the digestion capacity of the sludge accumulated in a UASB reactor treating municipal sewage is much larger than the applied organic load, so that it is feasible to waste a large fraction of the sludge in a full reactor without affecting the performance of the reactor in terms of the *COD* removal efficiency or of the operational stability.

For the investigated hydraulic retention times of 4 and 8 hours in pilot scale UASB reactors, with efficient phase separators, it was possible to discharge 50 to 60 per cent of the sludge without affecting significantly the performance of the UASB reactors. Discharges of up to 80 per cent led to a temporary reduction of the *COD* removal efficiency and an increase of the *VFA* concentration.

As long as the reactor is not full of sludge, a constant fraction of the produced sludge is discharged together with the effluent, while the remainder accumulates. The proportion between expelled and accumulated sludge depends on the hydraulic retention time and the phase separator efficiency.

The important operational parameters that determine the frequency of excess sludge discharges from UASB reactors are: (a) the applied organic load (b) the hydraulic retention time, (c) the maximum sludge mass that can be retained in the reactor (d) the minimum sludge mass required for a proper performance of the reactor and (e) the specific sludge production.

The sludge retention efficiency by the phase separator of UASB reactors is an important factor in the determination of the frequency and magnitude of excess sludge discharge.

3.5. References

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Chapter 4

Post Treatment of UASB Reactor Effluent in Flow-Through Polishing Ponds

4.1 Introduction

In Brazil, as well as in other regions with a hot climate, waste stabilization ponds systems (WSP´s) have been amply used for municipal sewage treatment. These WSP´s have, as their main objective, the reduction of the influent organic material (OM) concentration. To achieve this objective a long liquid retention time (R_h) is required even under the favourable conditions in tropical regions (high temperature and solar radiation intensity) R_h is in the range of 20 to 30 days (Mara, 1975, Mara and Pearson, 1987 and Yanez, 1993).

The long retention period of the liquid for the removal of organic material has an important indirect advantage: the liquid remains for a sufficiently long period to achieve complete removal of helminth eggs and a high removal efficiency of faecal coliform (*FC*). Thus a final effluent with a high hygienic quality is automatically assured. The large area required for the construction of a WSP´s constitutes the principal limitation of the use of these systems. For example, for R_h = 25 days and an average pond depth of 1 m, the required *per capita* area would be 3 m^2.

In order to obtain a good performance and, at the same time, to minimise R_h the WSP´s must be divided into several ponds, which are operated in series. In the first pond, which receives the raw sewage the applied high BOD_5 load will cause a predominantly anaerobic environment and, for this reason, it is called an anaerobic pond (AnP). Settling is the main mechanism of organic material removal in AnP's and the settled organic matter will be digested on the pond bottom. The AnP is followed by a partially aerobic pond, named a facultative pond (FP), and one or more predominantly aerobic ponds, known as maturation ponds (MP). The latter are used for complementary aerobic treatment and improvement of the hygienic effluent quality.

The anaerobic pond is very efficient compared to the other ponds, because at a relatively short liquid retention time (2 to 5 days) a large fraction of the influent organic material is removed (60 to 70 % of the BOD_5, Mara, 1975), especially at higher temperatures (higher than 20 ^oC). In the subsequent facultative pond the liquid retention time is of 7 to 10 days and in the maturation ponds a total of 10 to 15 days in total. It is important to note that the use of the AnP is not considered compulsory and it is sometimes excluded because of the bad odours that often emanate. In that case the facultative pond receives a much higher load and, as a result, the R_h must be longer.

Over the last decades, a large number of high rate anaerobic treatment systems have been implemented, particularly the so called up flow anaerobic sludge blanket (UASB) reactor or its variants (Campos, 1999). In these systems a high removal efficiency of organic material and suspended solids (*SS*) can be obtained (van Haandel and Lettinga, 1994), although the residual concentrations of these two pollutants in the effluent usually still exceed the maximum allowed by local environmental legislation.

When efficient anaerobic pre-treatment is applied, for example in the UASB reactor, the concentrations of OM and *SS* are drastically reduced and the removal of the residual concentration in ponds becomes much easier and, for this reason, it can be accomplished at a significantly shorter R_h. Under these conditions, the factor that limits the R_h of the pond and hence its volume is, no longer the stabilization of the OM, but rather the removal of pathogenic organisms and/or nutrients. In this thesis, the investigated ponds that receive effluent from an efficient anaerobic pre-treatment system are called polishing ponds (PP), to distinguish them from conventional waste stabilization ponds systems that treat raw sewage.

This chapter describes the operation and the performance of pilot scale ponds (PP´s), operated as flow-through systems and fed with UASB effluent. Design criteria for the optimization of the removal efficiencies of suspended solids, organic material (BOD_5 and

COD) and pathogens (*FC* and helminth eggs) will be presented. It will be shown that, under the experimental conditions, it was feasible to reduce BOD_5, *TSS* and helminth eggs concentrations to acceptable levels in only 3 days, while for an efficient removal of *FC* the required retention time was 10 days. The relatively long liquid retention time required for *FC* removal was due to the imperfections of the flow regime of the pond. Even though the ponds had been designed for a plug flow hydraulic regime, there was considerable mixing in reality, which was demonstrated by the high dispersion number (0.14 to 0.18) found in tracer studies. It was also shown that there was a considerable dead volume fraction (0.12 to 0.18), so that the flow regime in the ponds was far from plug flow. As a consequence, the exponential decay of *FC* (expected on the basis of reactor design theory) was not observed in the flow-through ponds.

An attempt was also made to assess the influence of depth on the *FC* removal by operating ponds with depth varying between 0.65 and 3.0 m. The decay constant (k_d) proved to be inversely proportional to the depth, so that the removal efficiency in the ponds is independent of depth: the longer retention time in deep ponds compensates the lower decay rate, due to the lower value of the decay constant.

4.2 Processes Developing in Ponds

The main objective of post treatment of anaerobically pre-treated sewage in polishing ponds is to improve the final effluent quality, so that it conforms conformity with the standards set by environmental control authorities or organizations for public health protection. This effluent can then be reused or discharged into surface waters without causing an adverse environmental impact or a public health hazard.

In Brazil the National Council for the Environment – CONAMA, through Resolution n$^{o.}$ 20, establishes the standards for water quality in water bodies in Brazil and for the discharge of effluents of pollution sources in surface waters. The resolution also establishes criteria for the hygienic quality in cases of reuse for unrestricted irrigation. Internationally important standards have been set by the Council of European Communities (CEC) and the World Health Organization (WHO). Table 4.1 shows the limits recommended by CONAMA, CEC and WHO for unrestricted irrigation and surface water discharge. In the same table the typical composition of raw and anaerobically pre-treated municipal sewage in Brazil is also shown. It can be noted that compared to the CEC standards, the CONAMA resolution is more restrictive for the limits on BOD_5 and *SS*, in cases of surface water discharge. Also the CONAMA resolution tends to be more restrictive than WHO recommendations for irrigation with treated sewage.

Table 4.1. Raw and anaerobically treated sewage characteristics and standards or recommendations for discharge of treated sewage in surface waters and unrestricted irrigation.

Parameter	Unit	Sewage		Desirable quality	Standard
		Raw	Digested		
COD	mg/*l*	600	120	125 (3)	CEC(1)
BOD_5	mg/*l*	250	50	5 25 (4)	CONAMA (1) CEC (1)
TSS	mg/*l*	400	100	≤ 150	CEC (1)
$N\text{-}NH_3$	mg N/*l*	50	45	≤ 5	CONAMA (1)
N_{total}	mg N/*l*			10 – 15 (5)	CEC (1)
P_{total}	mg P/*l*	10	9	1 – 2 (5)	CEC (1)
FC	CFU/100 m*l*	10^8	10^7	≤ 10^3 absence	WHO (2) CONAMA (2)
HE	No./*l*	100	50	<1	WHO (2)
pH		7.5	7.0	5-9	CONAMA (1)
Settle solid.	m*l*/*l*	20	2	1	CONAMA (1)

(1) Surface water discharge; (2) unrestricted irrigation; (3) sample without filtration or settling; (4) filtered sample, (5) population above 100,000 and from 10,000 to 100,000, respectively. WHO – WHO, 1989; CONAMA – CONAMA, 1986; CEC – CEC, 1991.

Although Table 4.1 shows that the anaerobic treatment is quite efficient at removing of organic material and suspended solids, the BOD_5 and the *TSS* concentrations in anaerobically treated sewage always exceed the recommended limits and therefore need to be corrected. The anaerobic process does not affect considerably other undesirable constituents of sewage such as pathogenic organisms, quantified by the *FC* and helminth egg (*HE*) concentrations and nutrients (notably the nitrogen and phosphorus concentrations). In addition post treatment may be needed for correction of other parameters such as the settleable solids concentration. The order of importance of the different parameters depends on the use of the final effluent, but generally the sequence can be listed as follows: (1) reduction of BOD_5 and *TSS*, (2) pathogen removal and, in case of discharge to lakes, dams or estuaries (3) nutrients removal.

4.2.1. Organic Material and Suspended Solids Removal

When wastewater is exposed to the conditions in polishing ponds, the concentrations of the different pollutants vary with time because of biological, physical and chemical processes that take place. The most important biological processes are: (1) photosynthesis, (2) oxidation of OM by aerobic bacteria and (3) fermentation of OM during anaerobic digestion. These three biological processes influence directly the removal of organic material in ponds. Figure 4.1 shows a schematic representation of the mechanisms that develop in ponds. It can be noted that there is an interaction between the biological processes of photosynthesis and oxidation: while the former produces organic material and oxygen from CO_2 and H_2O, the latter uses oxygen to oxidise organic material producing again CO_2 and H_2O.

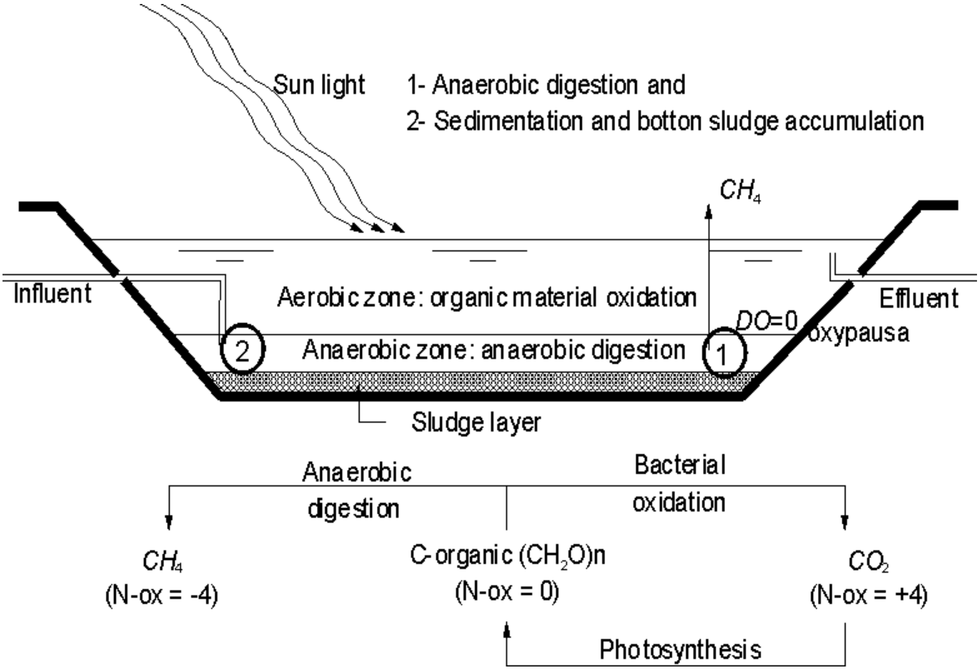

Figure 4.1. Schematic representation of a pond for the treatment of wastewater and the biological processes that take place in it.

In the process of **photosynthesis,** algae use solar energy to reduce carbon dioxide and transform it into cellular organic material. The process can be represented as follows:

$$xCO_2 + (y/2)H_2O \rightarrow C_xH_yO_z + 1/4(4x + y - 2z)O_2 \tag{4.1a}$$

It is important to note that the generation of organic material (in terms of COD) is accompanied by the generation of oxygen: stoichiometrically the mass of produced oxygen is exactly equal to the mass of oxygen necessary to oxidise the organic material produced in the process. However, the available oxygen in the pond is used by the bacteria not for the oxidation of algae but, primarily, for the oxidation of the faecal organic matter present in the sewage, as shown by the following equation of **aerobic oxidation by bacteria:**

$$C_xH_yO_z + 1/4(4x + y + 2z)O_2 \rightarrow xCO_2 + (y/2)H_2O \qquad (4.1b)$$

From Figure 4.1 and Equations (4.1a and b) it is clear that the processes of photosynthesis and oxidation are complementary in the sense that the products of one process are the reactants of the other. This shows that, in principle, there can be no removal of organic material by oxidation, if there is no oxygen source apart from photosynthesis, such as for example absorption of atmospheric oxygen. However the combined action of photosynthesis and oxidation can result in the removal of organic material from the liquid phase by other mechanisms: the net result of the two processes is the transformation of faecal organic material into algal mass, which can form flocs by spontaneous flocculation. Upon settling, they will become part of the bottom sludge in the pond, where the biodegradable part will be anaerobically digested, whereas the non biodegradable part will accumulate. This is indicated in Figure 4.1 by the numbers 1 and 2 respectively.

The relative rates of the processes of bacterial oxidation and photosynthesis affect directly important parameters like the dissolved oxygen (*DO*) and carbon dioxide (CO_2) concentrations, as well as, the BOD_5, *COD*, and *TSS* concentration and *pH*. Other parameters are indirectly affected such as the alkalinity and the concentrations of nutrients nitrogen (*N*) and phosphorus (*P*).

The rate of photosynthesis depends mainly on the transparency of the liquid phase, solar radiation, temperature and pond depth. The rate of bacterial oxidation of organic pollutants is influenced principally by temperature and the concentration of biodegradable organic material. If the concentration of organic material is high, the concentration of oxygen will be low and the environment of the pond will become anaerobic. At a lower organic loading rate in the pond an aerobic top layer develops, whereas the bottom layer (with less light for photosynthetic oxygen production) remains anaerobic. This is the typical situation for a facultative pond (FP). The oxypause refers to the transition level of the aerobic to the anaerobic environment. Naturally due to the cyclic nature of the process of photosynthesis (which requires sunlight) the position of the oxypause oscillates: lowering at daytime and rising when the sun sets. The average thickness of the aerobic layer depends on the relative rates of oxidation and photosynthesis. When the organic load is low, a large part of the pond will become aerobic and the anaerobic environment is restricted to the bottom sludge zone.

Several researchers (Mara and Pearson, 1987) have shown that in the case of raw sewage treatment the nature of the pond can be related to the applied organic loading rate. In regions with a tropical climate the recommended BOD_5 loading rate is presented in Table 4.2.

Table 4.2. Recommended BOD_5 loading rate for pond in tropical regions.

anaerobic ponds	> 400 gBOD_5.m^{-3}.day^{-1}
facultative ponds	< 300 kgBOD_5.ha^{-1}.day^{-1}
maturation ponds	< 150 kgBOD_5.ha^{-1}.day^{-1}

The third important biological process is **anaerobic digestion**, which develops at the bottom of the pond, where settled solids form a sludge layer and no dissolved oxygen is present. The process can be represented in a simplified manner as:

$$C_xH_yO_z + (4x - y - 2z)/4H_2O \rightarrow (4x + y - 2z)/8CH_4 + (4x - y + 2z)8CO_2 \qquad (4.1c)$$

Strictly speaking, the anaerobic digestion process does not destroy organic material: it transforms into methane gas, which is released to the atmosphere, resulting in physical removal of the organic material from the liquid phase.

4.2.2. Performance of Waste Stabilization Ponds

In Campina Grande, North East Brazil, an extensive experimental investigation has been carried out, at pilot scale, dealing with the application of waste stabilization ponds for the treatment of raw sewage. Table 4.3 shows the treatment performance average values obtained with a system comprising of 11 ponds. Figure 4.2 shows the lay-out of the pond system. The system consisted of an anaerobic pond (AnP) followed by four facultative ponds (FP), all operated in parallel and with different depths. The effluents of the FP were fed to a primary maturation pond (MP1). Three secondary maturation ponds MP2, MP3 and MP4, with depths of 0.90, 0.64 and 0.39 m respectively, also operated in parallel were fed with the effluent of pond MP1. The united effluents of the secondary maturation ponds were fed to two tertiary maturation ponds, MP5 and MP6, both with a depth of 0.60 m. The length/width ratio of pond MP5 was high (64/0.44), in an attempt to induce a plug flow regime. This pond system was operated at a low organic load (177 gBOD_5.m^{-3}.day^{-1}), during a period of 8 months.

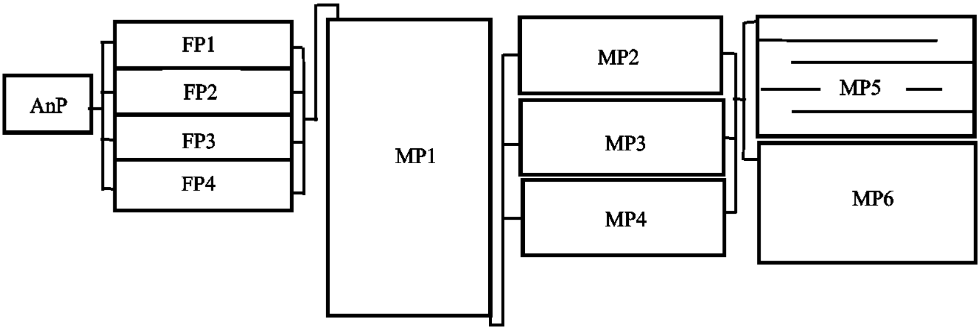

Figure 4.2. Schematic representation of a series of WSP´s, at pilot scale, operated with raw sewage at Campina Grande – Paraíba - Brazil.

The influent BOD_5 load of each pond, the main design parameters and the performance data are summarised in Table 4.3, where the filtered $BOD_{5,f}$ in the effluent of the MP are practically constant and always below 25 mg/*l*, so that they meet the standards of the Council of the European Communities. However, by contrast to the filtered $BOD_{5,f}$, the suspended solids (*SS*) concentration and the *COD* increased along the MP series. The average liquid retention time needed to obtain a concentration of $FC < 10^3$ was approximately 19 days.

Table 4.3. Some design parameters and average values (1) of the effluent quality parameters of a WSP system at pilot scale operated in Campina Grande with municipal sewage as feed from May to December 1992.

Parameters	RS	AnP	FP				Primary MP	Secondary MP			Tertiary MP	
			FP1	FP2	FP3	FP4	MP1	MP2	MP3	MP4	MP5	MP6
Dimensions and Operation:												
Length (m)		4.90	12.90	12.90	12.90	12.90	17.35	10.40	10.40	10.40	8.45	64
Width (m)		1.65	2.00	2.00	2.00	2.00	8.80	3.75	3.75	3.75	3.70	0.44
Depth (m)		2.50	1.00	1.33	1.67	2.00	1.00	0.90	0.64	0.39	0.60	0.60
Flow (m3/dia)		20	8	8	8	8	40	5	5	5	3.75	3.75
R_h (day)		1	3	4	5	6	3.8	7	5	3	5	5
Surface BOD_5 load (kgBOD5.ha^{-1}.day^{-1})		177$^{(2)}$	230	230	230	230	73	24	24	24	28	72
Effluent Quality:												
pH	7.3	7.3	7.6	7.7	7.7	7.7	8.1	8.9	8.9	8.9	9.2	9.3
$BOD_{5.f}$(mg/l)	179$^{(3)}$	53	22	19	17	17	15	17	16	20	20	17
COD(mg/l)	528	274	205	119	183	171	156	188	191	228	228	196
SS (mg/l)	285	121	81	77	67	63	53	74	73	115	101	107
$N\text{-}NH_3$	33.3	41.3	36.9	36.1	35.9	35.4	23.5	6.8	7.4	8.4	4.3	2.7
Chlorophyll a (µg/l)	Max/min		2334/398	2138/661	1321/379	1545/369	2990/359	3520/527	4058/622	4487/814	3993/621	4076/716
FC (CFU/100 ml)	3.2E+7	7.1E+6	1.1E+6	9.2E+5	9.2E+5	7.8E+5	2.3E+4	5.5E+2	6.8E+2	7.6E+2	3.5E+1	1.9E+1

(1) (Pearson *et al.*. 1996). (2) (g BOD_5.m^{-3}.day^{-1}) and (3) unfiltered sample.

4.2.3 Removal of Pathogenic Organisms

Municipal sewage contains a large variety of pathogenic organisms, including viruses, bacteria, protozoa and helminths (the latter mainly in the form of eggs). The most important indicator organisms used for hygienic quality are the concentrations of faecal coliform (*FC*) and helminth eggs (*HE*). The concentration of *FC* is expressed as the number of colony forming units per 100 ml of sample (CFU/100 ml) and the *HE* values are expressed as the count per litre of sample. The *FC* and *HE* values are important because they have a higher survival capacity than other organisms (*FC* in the treatment system and *HE* also in other environments). Thus, if the concentration of the indicator organisms remains well below the value set as standard, it can be inferred that the concentration of pathogenic microorganisms will also remain within the maximum limits for the intended use of the effluent.

Helminth Eggs Removal in Ponds

The reduction of *HE* and *FC* is obtained through completely different mechanisms. The removal of *HE* proceeds via physical mechanism of settling as a result of their density. By contrast, *FC* decay with time and cease to exist as live organisms. The removal of *HE* can be achieved by simple sedimentation or by absorption onto biological flocs in treatment systems. Silva *et al.*, 1996 showed that in a conventional WSP system operated by the team of research at Extrabes (Biological sewage treatment research facility) the effluent was free from *HE* after a retention time of 12.3 days (of with 6.8 days in an anaerobic pond and 5.5 in a facultative unit) and a complete removal of *Ancilostoma spp.* and *Ascaris lumbricoides* as well as larvae of the latter was attained, whereas any *Ancilostoma spp.* larvae were not found anymore after 17.8 days. Lakshminarayana and Abdulappa, 1969, reported that in a system with three ponds, each with a liquid retention time of only 2 days, all *Ancilostoma spp.* eggs settled in the first two ponds (4 days), but still larvae were present in the final effluent. These observations lead to questioning the validity of the criterion that water is or is not infective based on the fact that it has less than 1 *HE* per litre (Ancylostomatidae can hatch under favourable conditions 1 day after having been expelled with the faeces and after 3 to 4 days it can become infective larvae or filariae).

Experimental data, obtained by different researchers, on the time required to remove *HE* for wastewaters in ponds vary considerably. However, according to several researchers (Feachem *et al.*, 1983, Bartone, 1985, Mara and Cairncross, 1989 and Stott *et al.*, 1994), when efficient *FC* removal is obtained in a WSP system (99.99 %), it can be expected that *HE* removal will be complete. Table 4.4 shows the observed minimum liquid retention times for complete *HE* removal, found by different research workers. The very large spread can, in part, be attributed to the lack of a standardised method to count *HE* concentrations (Grimason *et al.*, 1996).

Stott *et al.* (1994) proposed an empiric equation to estimate the *HE* removal efficiency (Re*m*) as a function of the liquid retention time (R_h) in WSP (Equation (4.2)). For the estimated concentration of *HE* in raw sewage of Campina Grande (average of 500/*l*) the required R_h to reduce the number to <1 per litre would be 20 days.

$$\mathrm{Re}\,m = 100\left\{1 - 0{,}41 \exp\left(-0{,}49\,R_h + 0{,}0085\,R_h^2\right)\right\} \tag{4.2}$$

Table 4.4. Required R_h for the complete removal of *HE* in ponds (single or series units) according to several researchers.

Time for HE removal (day)	Scale and type	Count method	Place	Reference
2	AnP+FP (pilot scale)	Bailenger, modified by (Ayres and Mara, 1996)	Campina Grande-Pb, Brazil	Pearson *et al.*, 1996
11	AnP+FP+MP (pilot scale)	not described	Campina Grande-Pb, Brazil	Mara and Silva, 1986
12.3	AnP+FP	Extrabes$^{(1)}$	Campina Grande – Pb, Brazil	Silva *et al.*, 1996
12.4	FP+MP (real scale)	“A” $^{(2)}$	Dandora, Kenya	Grimason *et al.*, 1996
26.2	AnP+FP+MP+MP (real scale)	“A”	Eldoret	Grimason *et al.*, 1996
17.8	AnP+FP+MP (real scale)	“A”	Nakura, Kenya	Grimason *et al.*, 1996
34	AnP+FP+MP+MP (real scale)	NE$^{(3)}$	Jordan	Saqqar and Pescod, 1991
8	UASB+PP (pilot scale)	Bailenger, modified by (Ayres and Mara, 1996)	Itabira-MG, Brazil	América *et al.*, 2000

(1) Method developed at Extrabes, (2) described by Grimason et al., 1996 and (3) not specified.

Faecal Coliform Removal in Ponds

The removal of *FC* results from the decay of these bacteria (Marais, 1974), a biological process that has been described as a first order reaction (Marais and Shaw, 1961). In first order processes the rate is proportional to the concentration (Chick's law) and can be expressed as (Equation (4.3))

$$r_d = \frac{dN}{dt} = -k_d N \tag{4.3}$$

where:

r_d	decay (death) rate of the *FC*;
N	*FC* concentration (CFU/100 ml);
t	time (day);
k_d	decay constant for *FC* (day^{-1}).

When a pond of a treatment system is considered as a reactor operating with constant flow and volume (i.e., a particular liquid retention time), its removal efficiency of *FC* can be determined by solving the general differential equation (Equation (4.3)). Chemical reactor design theory shows that the solution of Equation (4.3) depends on the hydrodynamic behaviour in the reactor. For a first order process it is convenient to distinguish three different cases: (1) complete mixing, (2) no mixing and (3) partial mixing.

In a completely mixed pond the solution of Equation (4.3) is given by (Marais, 1974):

$$\frac{N_e}{N_i} = \frac{1}{1 + k_d R_h} \tag{4.4}$$

where:

N_e	effluent *FC* concentration (CFU/100ml);
N_i	influent *FC* concentration (CFU/100ml);

Equation (4.4) can be applied to a series of M completely mixed ponds, resulting in:

$$\frac{N_e}{N_i} = \frac{1}{(1 + k_d R_{h1})(1 + k_d R_{h2})\ldots(1 + k_d R_{hM})} \tag{4.5}$$

where R_{hM} is the liquid retention time in pond M

If in a series all ponds have the same liquid retention time ($R_{h1} = R_{h2} = \ldots R_{hM} = R_h/M$), Equation (4.5) can be simplified to:

$$\frac{N_e}{N_i} = \frac{1}{\left(1 + \frac{k_d R_h}{M}\right)^M} \tag{4.6}$$

In Figure 4.3(***a***), Equations (4.5) and (4.6) are presented graphically. The ordinate on the left of the Figure 4.3(***a***) presents the residual *FC* concentration, whereas the ordinate on the right expresses the percentage of *FC* removal (both log scale). The independent variable on the abscissa is the adimensional product k_dR_h. Figure 4.3(***a***) is very useful for the design with completely mixed ponds. For example, if a required removal efficiency of 99.9 % of the *FC* concentration is required in a system consisting of 5 completely mixed ponds of the same volume, it can be seen in Figure 4.3(***a***) that the product k_dR_h must have a value of 14.6. If the value of the decay constant is known (for example k_d = 2.0 day^{-1}), the required time to effect the removal can be calculated: 14.6/2 = 7.3 days. Equation (4.6) as well as Figure 4.3(***a***) show that in a series of ponds with any particular R_h, the removal efficiency of *FC* will be higher as the number of ponds in the series increases.

If there is no mixing in the pond, the hydrodynamic regime is a plug flow type and the solution of Equation (4.3) is an exponential expression:

$$\frac{N_e}{N_i} = \exp\left(-k_d R_h\right) \tag{4.7}$$

Equation (4.7) has also been presented in Figure 4.3(***a***). It can be noted that for any value of the total liquid retention time, the removal efficiency for *FC* in a pond without mixing is larger than in a series of ponds with completely mixed units. For example, for a required removal efficiency of 99.9 % of the *FC* in a pond without mixing the value of k_dR_h = 6.9 and for k_d = 2.0/day, the liquid retention time is R_h = 6.9/2 = 3.5 days, less than half the required R_h in a series with 5 ponds (see example above). It therefore it can be concluded that, in principle, it is advantageous to have no mixing in ponds, since, for the same R_h, the pond with a plug flow regime is more efficient than any other.

According to Agunwamba *et al*., 1992, Yanez, 1993 and von Sperling, 1999, mixing intensity in ponds decreases when shallow units are constructed with a large length/width ratio. This can be done by constructing a very long pond, or, more practically, by dividing the pond in a number of sections by walls, thus forming channels or lanes so that the large length/width ratio is artificially obtained.

Independent of the pond configuration, in practice, mixing can never be totally avoided.

However, the mixing intensity depends on several factors, some of which can be controlled, whereas others cannot. An important source of mixing, usually are the inlet and outlet structures of the pond. Also other mechanisms of mixing must be considered such as the mechanical action of wind and of biogas bubbles in the ponds. The wind tends to drag the superficial layer of liquid, causing a reflux in the bottom layer. The biogas bubbles tend to cause a mixing action of the liquid column along its vertical path to the water surface. Additional mechanical mixing will be caused by the locomotion of micro and macro organisms that live in the ponds. As far as algae are concerned, for example, König, 1984 observed that, in a pond operated at Campina Grande, the algae moved from the top layer to lower sections, during periods of intense sun shine. On the other hand, sunshine may cause thermal stratification over the depth of the pond: during daytime the top layer section will tend to be warmer and hence less dense than the bottom layer, whereas, at night, the pattern inverts with a colder and denser layer on the top. This daily thermal inversion tends to lead to vertical mixing of the liquid in the pond.

Wehner and Wilhelm, 1956 developed an expression (Equation (4.8)) for the determination of the efficiency of the first order process in a partially mixed reactor. In this expression the mixing intensity is quantified by the dispersion number, a dimensionless factor that, in principle, must be determined experimentally. The residual fraction of FC in a partially mixed pond is given by Wehner and Wilhelm, 1956:

$$\frac{N_e}{N_i} = \frac{4a.\exp^{\left(\frac{1}{2D}\right)}}{(1+a)^2 \exp^{\frac{a}{2D}} - (1-a)^2 \exp^{\frac{-a}{2D}}} \tag{4.8}$$

where $a = (1 + 4k_d R_h D)^{1/2}$ and D is dispersion number.

Figure 4.3 (***b***) shows the N_e/N_i ratio as a function of the adimensional parameter $k_{d..}R_h$ for different values of the dispersion number: D = 0 (plug flow), D = 0.01 (little mixing), D = 0.1 (moderate mixing), D = 1 (intense mixing) and $D = \infty$ (complete mixing). It is clear that the increase in the mixing intensity reduces the removal efficiency of the FC, especially if the required removal efficiency is high. For example, for 99.99 % (4 log cases) a value of $k_d R_h$ = 9.2 is required in a pond with plug flow (D = 0) but for moderate mixing, D = 0.1, the value of the product must be 17 (Figure 4.3(***b***)). Hence the required R_h in the pond with D = 0.1 will be 17/9.2 = 1.9 times larger than in the pond without mixing (D = 0). It is concluded that even moderate mixing in the pond may lead to a doubling of the required liquid retention time and hence the volume and the area of a pond.

The dispersion number D in Equation (4.8) is a quantitative measure of the mixing intensity in a pond and is calculated from experimental determination of the liquid retention time distribution. This distribution, in turn, is determined by adding a slug of a tracer to the influent and observing the effluent concentration as a function of time. The average liquid retention time is the time required to recover half of the added tracer mass. The dispersion number is calculated from the experimental data by a standardised calculation procedure (Levenspiel, 1972), considering the distribution of the actual liquid retention time.

Another important element that can be determined by tracer studies is the stagnant fraction of the pond volume, called dead volume fraction. It is determined from the difference of the actual and the nominal liquid retention time. The actual liquid retention time ($R_{h,a}$) is equal to the mean retention time, determined from the retention time distribution curve. The nominal liquid retention time is the ratio between the nominal volume of the pond and the applied flow. From the actual liquid retention time the actual reactor (pond) volume (V_a) can be calculated as:

$$V_a = Q_a R_{h,a}. \tag{4.9}$$

Since the nominal volume by definition is given by the product of the flow rate and the nominal liquid retention time, ($V_r = Q_a R_h$), the dead volume fraction can be determined experimentally as:

$$f_d = \frac{V_r - V_a}{V_r} = 1 - \frac{R_{h,a}}{R_h} \tag{4.10}$$

where f_d is the dead volume fraction.

Several researchers evaluated the value of the dispersion number in ponds. The data of Yanez, 1993, obtained in South America and of von Sperling, 1999, who analyzed specifically the situation in Brazil, show that the dispersion number in full scale ponds is considerable, even when they were specifically designed for reduced mixing by having a large length/width ratio. The dispersion number almost always exceeds 0.2 and is very often larger than 0.5. Agunwamba *et al.*, 1992 and Yanez, 1993 developed expressions that allow an estimate of the dispersion number D as a function of the dimensions of the pond (Equations (4.11) and (4.12)).

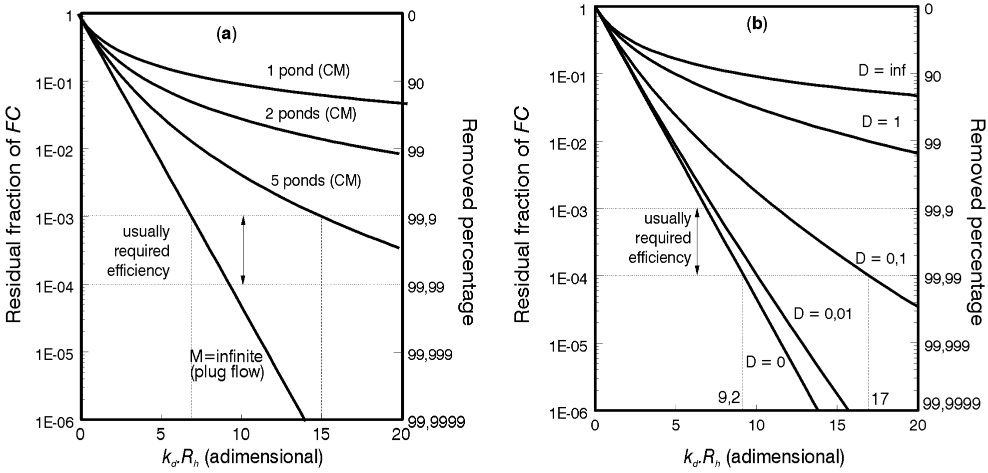

Figure 4.3. (a) Relationship between the residual fraction of *FC* (the removed percentage is also indicated) and the adimensional product $k_d.R_h$, for "M" ponds, "M" equal to 1, 2, 5 and ∞ and (*b*) for a single pond and different values of the dispersion number *D* equal to 0, 0.01, 0.1, 1 and ∞.

Agunwamba equation (simplified by von Sperling, 1999):

$$D = 0{,}102\left(3.\left(\frac{(B+2H).R_h.n}{4L.B.H}\right)\right)^{-0{,}410}.\left(\frac{H}{L}\right)\left(\frac{H}{L}\right)^{-(0{,}981+1{,}385H/B)} \qquad (4.11)$$

Yanez equation

$$D = \frac{L/B}{-0.261 + 0.254\left(\frac{L}{B}\right) + 1.014\left(\frac{L}{B}\right)^2} \qquad (4.12)$$

where:

í kinematical viscosity of the water (m^2/day) and

L and *B* are, respectively, the length and the width of the pond (m).

A factor not yet discussed concerns the value of the decay constant k_d. Several researchers showed that the value of the constant depends on a number of factors, among which the temperature, the pond depth and the *pH* are the most important. In the case of domestic sewage, the temperature of the pond is a variable that is determined by the air temperature. The pond depth is an important variable, because for any particular liquid retention time, the area is inversely proportional to the depth. The depth also has an indirect effect on *pH*: in deep ponds the mass of algae that receive light for photosynthesis (top layer) is relatively small, so that the CO_2 consumption remains low and hence also the *pH* increases little. By contrast, in shallow ponds the increase of *pH* can be considerable.

The influence of temperature on the decay constant can be described by an Arrehnius expression, in which normally with the k_d value at 20 ^oC as reference:

$$k_{dT} = k_d \theta^{(T-20)} \tag{4.13}$$

where:

k_{dT} value of the decay constant for any temperature (day^{-1});

k_d value of the decay constant at a reference temperature, normally 20 ^oC (day^{-1});

θ experimentally determined temperature dependent coefficient.

Table 4.5. Values of the decay constant kd and the temperature dependent coefficient θ reported by different authors.

Reference	Reference temperature (^0C)	k_d value (day^{-1})	θ value
Marais, 1974	20	2.6	1.20
Sherry and Parker, 1979	20	1.5	1.06
Klock, 1971	20	1.1	1.07
Yanez, 1993	20	0.84	1.07

Table 4.5 summarizes values of k_d and θ calculated from the results of some research workers. The value $\theta = 1.07$ has been found most frequently.

With respect to the influence of depth on the k_d value few data have been reported in the literature. van Haandel and Lettinga, 1994 indicated a hyperbolic relationship between the value of k_d and the pond depth (H). von Sperling, 1999, on the basis of data reported by several authors, proposed a similar relationship in which the value of k_d varies with $H^{-0.96}$.

4.2.3 Nutrient Removal

The removal efficiency of nutrients in conventional ponds is low and therefore cannot guarantee an effluent quality that is compatible with the standards required by authorities. However, in polishing ponds the *pH* can rise and nutrient removal will proceed through physical and chemical processes. The *pH* increases without the need to apply any alkali, provided the photosynthetic CO_2 consumption rate exceeds the CO_2 production rate by biological oxidation of organic material. At *pH* values exceeding 9, ammonium predominates in the gaseous (NH_3) form and thus nitrogen will be removed by ammonium desorption. At high *pH* also considerable removal of phosphate can be achieved due to the formation of poorly soluble phosphates that can precipitate, e.g. apatite ($Ca_{10}(OH)_2(PO_4)_6$) and struvite ($Mg(NH_4)PO_4$). For the precipitation, phosphorus must be present in the form of PO_4^{3-}, which is only the case at a high pH. Moreover calcium and/or magnesium must be present.

4.3 Experimental Investigation

In order to evaluate the behaviour of polishing ponds with respect to removal of organic material (BOD_5 and COD), total suspended solids (TSS), helminth egg (HE) and faecal coliform (FC), as well as nutrients (nitrogen and phosphorus), four polishing pond systems were operated, at pilot scale: PP1, PP2, PP3 and PP4 with depths of 0.65, 1.65, 2 and 3 meters, respectively. These systems were fed with a continuous flow of anaerobically pre-treated sewage. Table 4.6 shows the main dimensions of the ponds systems. All ponds received effluent from a UASB reactor (also at pilot scale) treating municipal sewage. A photograph of all experimental units, including UASB reactors and ponds is shown in Figure 4.4.

Table 4.6. Dimensions of the polishing pond systems used in the experimental investigation.

Code	Volume (m^3)	Depth (m)	Area (m^2)	Material
PP1	32.5	0.65	50 (= 5 lanes of 10 m x 1 m)	Brickwork
PP2	5	1.65	3 (= 2.0 m diameter)	Glass fibre
PP3	6	2.0	3 (= 1.75 m x 1.70 m)	Brickwork
PP4	9	3.0	3 (= 1.75 m x 1.73 m)	Brickwork

The UASB reactor was constructed in polyester reinforced with glass fibre with a volume of 5000 litres. The UASB reactor treated raw sewage taken from the main outfall of the sewer system of Campina Grande (latitude 7°13' South; longitude 35°52’ West and 550 m above sea level). The reactor was operated at R_h in the range of 3 to 6 hours. During the entire period of the investigation there was no intentional discharge of excess sludge from the UASB reactor. Consequently the ponds received the raw effluent containing sludge particles, washed out of the UASB reactor. These particles settled in the pond systems.

Figure 4.4. Experimental system: UASB reactors and polishing ponds.

Since efficient *FC* removal was required, PP1 was designed to minimise mixing. The pond was subdivided into five lanes each with a length of 10 m and a width of 1 m, which corresponded to a total length/width ratio of 50/1, as illustrated in Figure 4.5 and 4.6. Figure 4.5 also shows the connection between one lane and the next one. The connections were always operated in the sense top layer to bottom layer, although other options were possible. There was a head loss of 1 cm between the lanes to avoid blockages and reflux in the series. The other pond systems were conventional single units with different depths.

The investigation with the system UASB+PP1 was carried out in three phases characterised by different R_h values in the pond. In order to define R_h during the first phase the following assumptions were made: the *FC* concentration in the UASB effluent is equal to 10^7 CFU/100ml; desired removal efficiency in the pond system is 99.99%; dispersion number $D = 0.02$ (calculated from Equation (4.11)) and decay constant $k_d = 2.0$ day^{-1} (determined experimentally for the depth of 0.65 m in batch tests). With the aid of Equation (4.8) the required liquid retention time was calculated as $R_h = 5$ days, and this value was adopted during the 1st phase of the investigation.

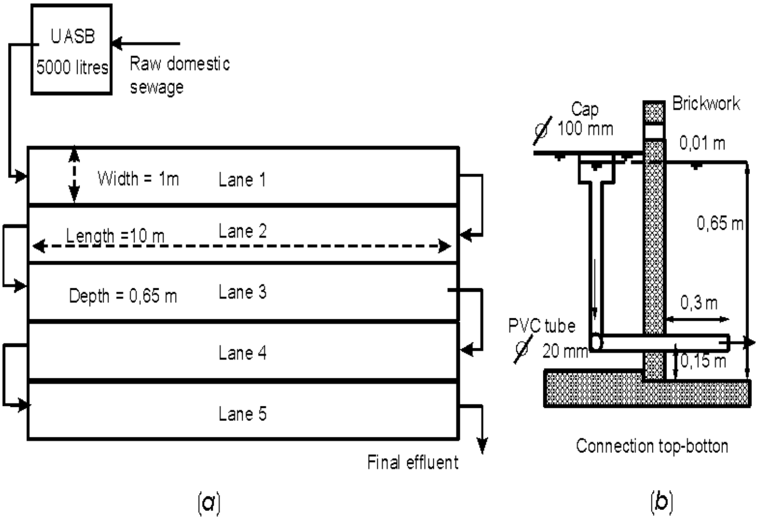

Figure 4.5. Experimental set-up of the UASB + PP1 system (***a***) and detail of the connection between the different lanes of the pond system (***b***).

Figure 4.6. The 5 lanes polishing pond PP1

During the operation it was observed that the actual removal efficiency of the *FC* never approached the intended value of 99.99%. For this reason the retention time was increased to 7.5 days (2nd phase) and, once there was evidence that even this was too short to produce the required effluent quality, it was increased to a liquid retention time of 15 days (the 3rd and final phase).

Values of the decay constant were obtained from results of batch experiments in which equal volumes of the contents of the five lanes were mixed in a vessel with a depth of 0.65 m, with a volume fraction of 10% of raw sewage in order to increase the *FC* concentration. The vessel was exposed to the sun during a period of 4 days (which was sufficient for a high *FC* removal efficiency) and 4 daily samples were taken for *FC* determinations. From the data obtained at the end of the experimental tests it was possible to determine the decay constant as the slope of the semi log plot of the number of *FC* as a function of the time of exposure (Equation 3.7).

Additionally, other experiments were carried out to determine the influence of the pond depth on the value of the decay constant. Ponds PP2, PP3 and PP4 were operated at a constant hydraulic loading rate of 0.22 $m^3.m^{-2}.day^{-1}$ until a steady state established as reflected by a constant value of the chlorophyll a concentration in the effluent. Then the feeding was interrupted in order to move to a batch operation of the pond and the *FC* concentration was measured as a function of time, always sampling at several depths. From the results of each pond operation, the decay constants were determined, using the same procedure as described above for the 0.65 m batch pond. The results of the experiments show that the *FC* decay in the batch ponds followed an exponential expression and the value of the decay constant appeared to be inversely proportional to the depth over the investigated range of reactor height (0.65 to 3 m).

At the end of each of the 3 phases of investigation of the system PP1, tracer studies were carried out to determine the dispersion number and the dead volume fraction. The tracer used in the studies was potassium chloride. A slug of 3.5 kg of commercial KCl, (sufficient for an average concentration of about 50 mg K/*l* in the volume of 32.5 m^3 of PP1, was added to the pond influent. The relatively high tracer concentration was chosen because potassium was present in the sewage at a concentration of 10 to 15 mg/*l*.

The experiments in the four PP systems were also conducted with the objective of assessing the influence of depth on the removal of OM and *TSS*. In the case of the shallow ponds (PP1), the final effluent quality, in this respect, was hampered by the high production rate of algae, which was possible because the sunlight could penetrate virtually until the bottom of the pond. The operation of pond systems PP2, PP3, PP4 was realised in parallel to that of PP1 and with liquid retention times of 7, 14, 21 and 28 days.

4.4 Results

The results are presented in Tables 4.7 to 4.12. Table 4.7 shows the average BOD_5 and *COD* values of raw and filtered samples as well as the *TSS* values, obtained during all operational periods of systems of the deep pond systems PP2, PP3 and PP4. Tables 4.8, 4.9 and 4.10 show the results for PP1 at R_h 5, 7.5 and 15 days. The R_h values of each separate lane are also indicated. Along with the data of BOD_5 and *COD* (unfiltered and filtered samples, except UASB effluent samples which refer to settled sample in an Imhoff cone) and *TSS*, also the average values of *TKN*, ammonium ($N\text{-}NH_3$), total phosphorus (P_{total}), *pH*, total alkalinity (*TAlk*), chlorophyll a (*Chl a*), helminth eggs (*HE*) and faecal coliforms (*FC*) are presented. All parameters were determined according to procedures described in Standard Methods for the Examination of Water and Wastewater(APHA, 1995) except *Chl a* and *HE*, which were determined according to Jones, 1979 and Ayres and Mara, 1996, respectively.

Table 4.11 shows the experimental values for the dispersion numbers and the dead volume fractions, calculated from the tracer profiles of the residence time distribution for the three applied retention times. In the same table are data from the observed residual *FC* fraction. Also the experimental values of the decay constant, determined in batch ponds with a depth of 0.65 m and a temperature of 25 ± 2 ^oC is shown. Having determined the actual retention time, the actual pond volume was calculated using Equation (4.9). Finally, with the aid of the actual and nominal volumes of the pond, the dead volume fraction was calculated.

Table 4.11 also contains the values of the decay constant k_d, determined from data obtained from batch 0.65 m deep pond. These values were determined during each of the three phases of the investigation with PP1. Under the conditions of the investigation the value of the decay constant (k_d) found was 2.3 day^{-1} for a depth of 0.65 m and a temperature of 25 ^oC. Table 4.12 shows the experimental data of the decay constant determined in ponds with different depths. For all these determinations the ponds were operated in batch mode.

Table 4.7. Average BOD_5 and COD values for unfiltered and filtered effluent samples and TSS values for samples from unfiltered effluent from the pond systems PP2, PP3 and PP4, for different values of R_h.

	BOD_5						COD						TSS		
Parameter	Unfiltered			Filtered			Unfiltered			Filtered					
PP	2	3	4	2	3	4	2	3	4	2	3	4	2	3	4
Depth (m)	1.65	2.0	3.0	1.65	2.0	3.0	1.65	2.0	3.0	1.65	2.0	3.0	1.65	2.0	3.0
R_h = 7 days	89	97	86	58	73	56	185	194	215	101	86	115	110	168	164
R_h = 14 days	54	65	72	38	37	42	163	168	169	85	72	79	108	130	129
R_h = 21 days	38	48	66	26	26	44	140	145	155	69	65	75	99	102	106
R_h = 28 days	32	35	58	28	18	27	112	106	132	52	61	70	91	61	87

Table 4.8. Performance of polishing pond system PP1 for a constant flow of 270 *l*/h (R_h = 5 days), treating effluent from a UASB reactor operated at R_h =3 hours. Operational period: February to June 1999. Temperature of the pond 25 ± 2 °C.

Parameter	Unit	Raw sewage	Digested sewage	Pond lane				
				1	2	3	4	5
DO_{min}	mg/*l*	-	-	<1	<1	<1	<1	<1
$DO_{\text{máx.}}$	mg/*l*	-	-	1.5	2.5	>20	>20	>20
TSS	mg/*l*	422	190	58	99	130	103	61
BOD_5	mg/*l*	396	169	87	71	55	54	59
$BOD_{5,f}$	mg/*l*	-	105*	64	36	40	32	35
COD	mg/*l*	712	254	187	223	256	212	188
COD_f	mg/*l*	-	-	112	87	78	88	102
pH	-	7.4	6.9	7.1	7.3	7.7	8.1	8.3
TKN	mg/*l*	56	54	51	51	47	45	43
NH_3	mg/*l*	39	46	48	45	43	39	36
P_{total}	mg/*l*	7.1	5.4	5.9	5.2	6.0	5.6	5.7
$TAlk$	mg$CaCO_3$/*l*	384	432	436	413	397	391	397
Chl a	µg/*l*	-	-	123	464	916	1466	1702
HE	No./*l*	214	82	8	5	0	-	-
FC	CFU/100 ml	7.4E7	2.3E7	4.7E6	2.9E6	1.1E6	3.3E5	7.5E4

*$BOD_{5,set}$, settled sample

Table 4.9. Performance of polishing pond system PP1 for a constant flow of 180 l/h (R_h = 7.5 days), treating effluent from a UASB reactor operated at R_h =3 hours. Operational period: June to December 1999. Temperature of the pond 25 ± 2 °C

Parameter	Unit	Raw sewage	Digested sewage.	Pond lane				
				1	2	3	4	5
DO_{min}	mg/l	-	-	<1	<1	<1	<1	<1
$DO_{\text{max.}}$	mg/l	-	-	<1	2.8	>6.8	>20	>16
TSS	mg/l	536	177	189	107	127	110	62
BOD_5	mg/l	408	142	73	57	66	62	44
$BOD_{5,\text{f}}$	mg/l	-	96*	58	36	47	34	32
COD	mg/l	683	329	306	346	285	287	221
COD_f	mg/l	-		124	156	177	142	93
pH	-	7.2	6.7	7.0	7.4	7.7	7.9	8.0
TKN	mg/l	66	57	59	54	57	55	42
NH_3	mg/l	45	44	41	39	40	40	29
P_{total}	mg/l	7.2	6.7	7.0	7.4	7.7	7.9	8.0
$TAlk$	mgCaCO$_3$/l	285	310	352	348	346	341	333
$Chl\ \underline{a}$	µg/l	-	-	1149	2157	2154	2680	1282
HE	No./l	209	76	4	0	-	-	-
FC	CFU/100 ml	7.3E7	1.4E7	4.5E6	8.7E5	2.2E5	3.4E4	6.9E3

*$BOD_{5,\text{set}}$, settled sample

Table 4.10. Performance of polishing pond system PP1 for a constant flow of 90 l/h (R_h = 15 days), treating effluent from a UASB reactor operated at R_h =6 hours. Operational period: January to May 2000. Temperature of the pond 25 ± 2 °C.

Parameter	Unit	Raw sewage	Digested sewage.	Pond lane				
				1	2	3	4	5
$DO_{\text{mín}}$	mg/l	-	-	2.6	3.1	1.6	1.5	1.4
$DO_{\text{máx.}}$	mg/l	-	-	8.1	9.5	3.4	4.1	4.6
TSS	mg/l	438	99	52	44	40	44	35
BOD_5	mg/l	312	150	70	64	27	32	24
$BOD_{5,\text{f}}$	mg/l	177	63*	48	36	15	22	18
COD	mg/l	522	192	227	187	127	137	108
COD_{f}	mg/l	194	144*	155	119	69	66	53
pH	-	7.4	6.9	8.1	8.6	8.3	8.3	8.5
TKN	mg/l	56	58	54	46	57	55	42
NH_3	mg/l	41	47	43	33	30	26	20
P_{total}	mg/l	7.4	7.6	6.9	6.0	5.2	5.7	5.2
$TAlk$	mgCaCO$_3$/l	371	395	357	327	319	294	304
$Chl\ \underline{a}$	µg/l	-	-	984	939	291	238	249
HE	No./l		52	<1	-	-	-	-
FC	CFU/100 ml	3.1E7	2.4E6	4.3E5	8.0E4	7.7E3	5.0E2	5.8E2

*$BOD_{5,\text{set}}$, settled sample

Table 4.11.Values for the decay constant (k_d), the dispersion number (D), the fraction of dead volume (f_d) and experimental and theoretical values for the removal efficiency of FC, in polishing pond PP1, as a function of the R_h .

Parameter	5 days $^{(*)}$	7.5 days $^{(*)}$	12 days $^{(*)}$
Dispersion number	0.14	0.18	0.16
Dead volume fraction	0.12	0.18	0.15
Actual liquid retention time(day)	4.4	6.5	10.2
Experimental decay constant (day^{-1})	2.2	2.1	2.0
Experimental residual FC fraction	3.3E-3	4.9E-4	2.1E-4
Theoretical residual FC fraction (Wehner-Wilhelm)	3.4E-3	8.1E-4	1.2E-4
Theoretical residual FC fraction (plug flow)	6.2E-5	3.2E-7	1.4E-9
Theoretical residual FC fraction (complete mixing)	9.3E-2	6.3E-2	4.6E-2

(*)R_h total .

The data in Table 4.11 allow the calculation of the theoretical value of the remaining FC fraction assuming (1) complete mixing, (2) plug flow and (3) partial mixing using Equations (4.5), (4.7) and (4.8), respectively. The results of the calculations are also presented in Table 4.11. When the calculated values are compared, it can be seen that the partial mixing model of Wehner and Wilhelm gives an adequate estimate of the three cases. By contrast, the complete mixing model underestimates and the plug flow model overestimates grossly the FC removal efficiency.

Table 4.12. Experimental values for the decay constant k_d as a function of the depth H of a pond operating in batch mode.

H (m)	k_d (day^{-1})	$k_d.H$
0.40	4.0	1.60
0.65	2.3	1.50
0.80	2.0	1.60
1.5	0.7	1.05
2.0	0.6	1.20
3.0	0.5	1.50

Table 4.12 shows experimental values for the decay constant, determined in batch ponds with different depths, ranging from 0.40 to 3.0 m. The data show that the product $k_d.H$ does not seem to depend on the depth of the pond, from whence it can be concluded that the decay constant value at any particular temperature is inversely proportional to the depth of the pond. This is an important conclusion, because it means that a pond with a certain area will have a constant FC removal efficiency independent of the depth, because a decrease in the decay constant when the depth is increased, is compensated by the longer retention time, which is a result of the larger volume, due to the increased depth.

During the months of March, April and May 2000, when PP1 was operating at R_h = 15 days, the algae were classified and their populations estimated twice per week. Samples of the entire column of each lane were taken and the algae were allowed to settle after immobilisation with formaldehyde (2% solution). Part of the sediment was transferred to a Sedgwick Rafter cell and examined with the aid of a microscope. The data in Table 4.13 show which were the most important genera identified in the samples of the different lanes (L1 to L5). The most frequently observed genera were *Euglena*, *Chlorella* and *Phacus*. The relative and absolute numbers of these genera varied significantly not only spatially in the five lanes but also with time over the three months of observation. In the first lane the genus *Euglena* predominated with 40 to 61.4% of the total algae counted. In L5 *Chlorella* was most frequent accounting for 42.3 to 66% of the counted algae.

Table 4.13. Average monthly values for the relative proportions of different genera of alga in each lane of PP1 in the months of March April and May 2000, when PP1 operated with R_h = 15 days (T = 25 ± 1 °C).

		values in %						
Lane	Month	Euglena	Chlorella	Oscillatoria	Chlorococcum	Phacus	Chlamydomonas	Pyrobotrys
	March	46.8	1.6	0	0	38.4	3.2	16.4
L1	April	40	16	1.9	0.1	18.9	13.1	10
	May	61.4	10.4	0.33	0.17	5.8	1.9	20
	March	35.6	14.1	0	0	37.1	0	13.2
L2	April	44.7	17.6	0	0.6	19.4	11.3	6.4
	May	56.5	19.8	1.3	2.5	8.3	1.6	10
	March	16	43.2	0	0	34.3	0	6.5
L3	April	21	32.6	0	20.5	18.7	2.2	5
	May	45.8	11.5	0.2	30.4	1.3	0.8	10
	March	19	54.8	0	0	26.2	0	0
L4	April	12.3	36.8	0	32	12.7	5.2	1
	May	39.6	24	0	33.2	1.4	0.5	1.3
	March	13.5	66	0	12.2	8.3	0	0
L5	April	17	42.3	0	25	13.7	1	1
	May	10.9	48.5	0	36.3	0.1	0.3	3.9

Oscillatoria is a cyanobacterie.

4.5 Discussion of the Results

Organic Material and Suspended Solids Removal

Before discussing the performance of the polishing ponds with respect to the removal of organic material and suspended solids, it is opportune to highlight the excellent performance of the UASB reactors. In Tables 4.8, 4.9 and 4.10 it can be noted that the BOD_5 removal efficiency of the UASB reactor was on average 73 % and 76 % in phases 1 and 2 (R_h =3 hours) and 80% in phase 3 (R_h = 6 hours). These values refer to the settled effluent, which in this case is relevant, since settling would take place in the pond.

As a result of the high UASB reactor efficiency the polishing pond operated with loads compatible with those applied to maturation ponds of conventional waste stabilization pond systems. Having established that the organic material removal is not the determining factor of the PP design, this unit can therefore be designed to remove efficiently the pathogens and/or nutrients. If the primary objective is pathogen removal, the conventional configuration AnP + FP + MP is not optimal and can be conveniently substituted by UASB + PP, where PP is a single unit with minimum mixing intensity. In other words the UASB reactor substitutes effectively an AnP + FP of the conventional WSP system.

Figures 4.7 (***a***, ***b*** and ***c***) show the drop in the BOD_5, COD and TSS concentrations of the unfiltered effluent of the shallow (PP1) and deep pond (PP2, PP3 and PP4) as functions of the liquid retention time at a temperature of 25 ± 2 ^oC. Figure 4.8 (***a*** and ***b***) shows the drop in the filtered BOD_5 and COD in the same shallow and deep ponds. It can be noted that the BOD_5 of anaerobically pre-treated sewage was readily reduced in the PP from its value of 154 mg/*l* (average of the values presented in Tables 4.8, 4.9 and 4.10) to values in the range of 70 to 87 mg/*l* in the first lane of the shallow pond PP1. This can be attributed partially to settling of the flocs of anaerobic sludge leaving the UASB reactor. After this physical process the BOD_5 reduction was smaller and more gradual until final values in the range of 24 to 59 mg/*l* were obtained. Figure 4.7 and 4.8 (***a***) clearly illustrate the different aspects of BOD_5 removal in the polishing ponds:

- the first order model often used to describe the (total) BOD_5 removal in WSP is inadequate for polishing ponds; the decrease of BOD_5 with R_h certainly does not proceed exponentially. The experimental data rather indicate a zero order kinetics, where the decrease of BOD_5 proceeds linear with time;
- the rate of the BOD_5 drop both in the raw and the filtered samples was higher in the shallow pond than in the deep ponds. In the samples of unfiltered effluent the observed decrease of BOD_5 in PP1 exceeds by 50 to 55% the decrease in the deep PPs. This observation is in contradiction with the expectation with that in deep ponds the rate of photosynthesis would decrease, resulting in low BOD_5 (algae) production so that the BOD_5 removal would be higher than in shallow ponds. From a

investigation conducted in facultative ponds with depths between 1.0 and 2.0 meters (Table 4.3) and with anaerobically digested sewage as influent (Pearson et al., 1996) concluded that neither depth nor the configuration of the ponds change their performance with respect to the removal of BOD_5;

- the filtered BOD_5 in the effluent of polishing ponds comprises 50 to 75 % of the unfiltered BOD_5, indicating that a large proportion of the BOD_5 (25 to 50 %) consists of particulate material, principally algae, as it is illustrated in Figure 4.8.

The use of R_h as an independent variable (Figures 4.7 and 4.8) is not convenient because the performance of the ponds also depends on their depth. The influence of the depth can be eliminated when the superficial loading rate is used as an independent variable. In Figure 4.10 (***a***, ***b***, ***c***) the BOD_5 and *COD* values (raw and filtered samples) and the *TSS* values of the effluent of the polishing ponds PP1, PP2, PP3 and PP4 were plotted as functions of the superficial loading rate, ($l_s = H/R_h$, where H is the depth of the pond). It can be noted that for the same l_s, the values of BOD_5, *COD* and *TSS* tend to be the same in deep and in shallow ponds.

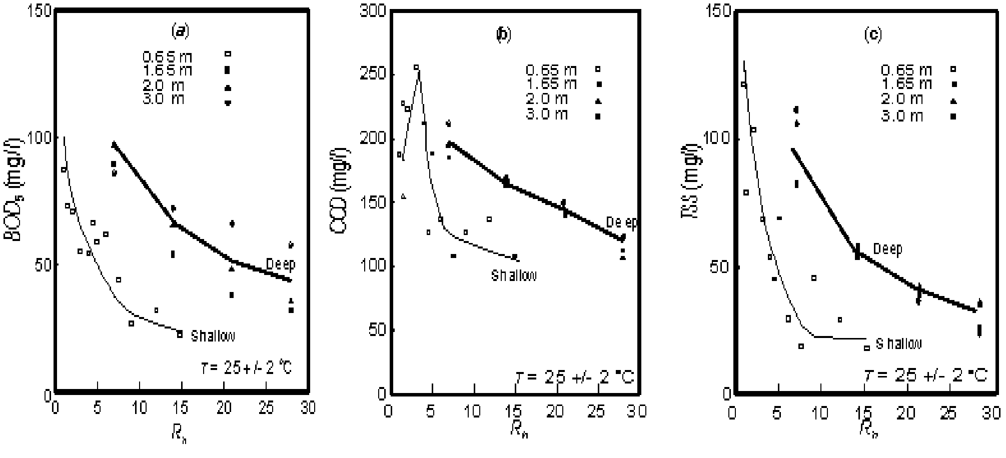

Figure 4.7. Performance of the shallow PP1 (depth of 0.65) and deep PP2, PP3 and PP4 (depths of 1.65, 2.00 and 3.00 m), operated with continuous flow, with regards to OM removal: BOD_5(***a***) and *COD* (***b***) of unfiltered effluent samples and *TSS* removal (***c***), as a function of the R_h.

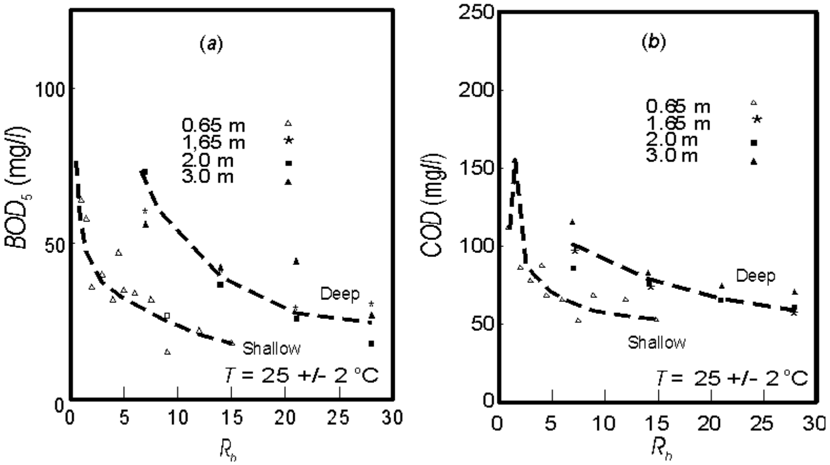

Figure 4.8. Performance of the shallow PP1 (depth of 0.65) and deep PP2, PP3 and PP4 (depths of 1.65, 2.00 and 3.00 m), operated with continuous flow, with regard to OM removal: BOD_5 (***a***) and *COD* (***b***) of filtered effluent samples, as a function of the R_h.

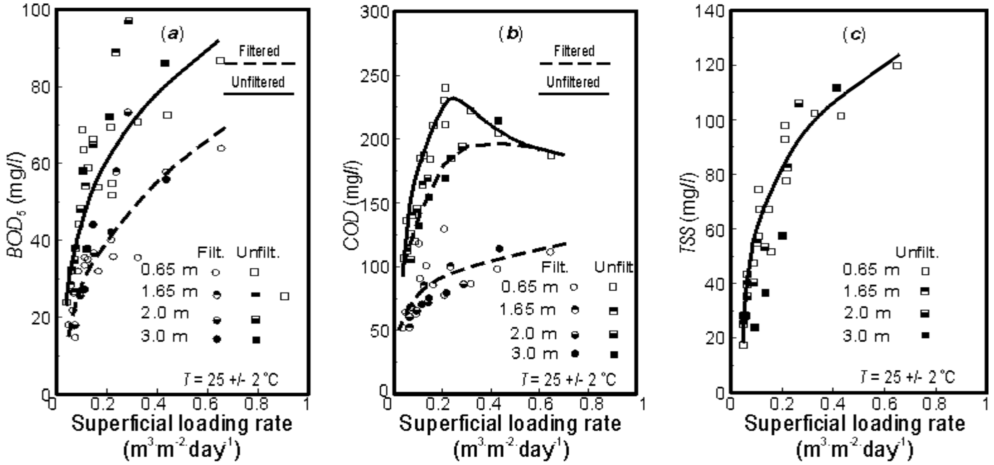

Figure 4.9. Performance of the shallow PP1 (depth of 0.65) and deep PP2, PP3 and PP4 (depths of 1.65, 2.00 and 3.00 m), operated with continuous flow in terms of BOD_5 (***a***) and *COD* (***b***) removal in unfiltered (drawn curves) and filtered (interrupted curves) effluent samples and of *TSS* (***c***) removal as a function of the superficial loading rate (l_s).

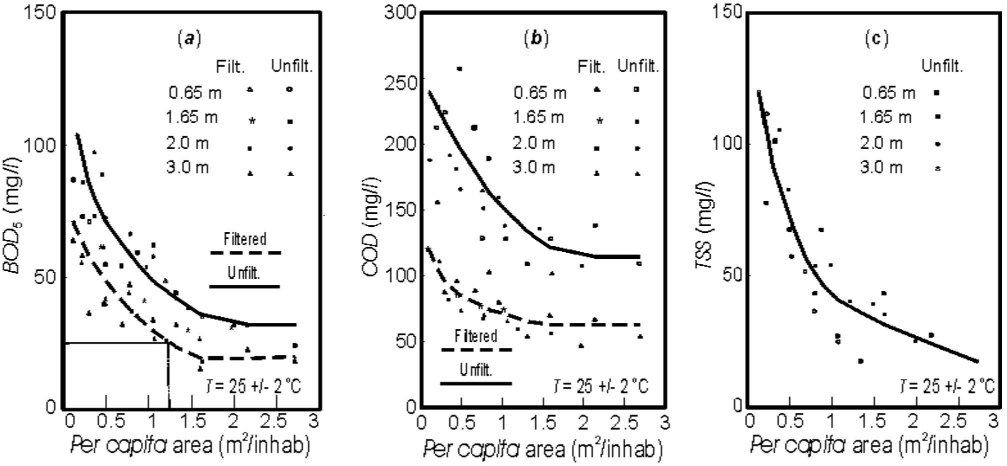

Figure 4.10. Performance of the shallow PP1 (depth of 0.65) and deep PP2, PP3 and PP4 (depths of 1.65, 2.00 and 3.00 m), operated with continuous flow in terms of BOD_5 (***a***) and *COD* (***b***) removal in unfiltered (drawn curves) and filtered (interrupted curves) effluent samples and of *TSS* (***c***) removal as a function of the *per capita* area assuming a contribution of 100*l*.inhab^{-1}.day^{-1}.

In Figures 4.10 (***a***, ***b*** and ***c***), another way of presenting the data is shown, using the *per capita* area of the pond as the independent variable. In these figures it is clear that the values of BOD_5, *COD* and *TSS* in the PP effluent depend on the *per capita* area but not on the depth of the ponds. It can be seen in Figure 4.10(***a***) that for a desired filtered BOD_5 concentration of 25 mg/*l* in the effluent, in accordance with the CEC standards (CEC, 1991), the required area would be 1.3 m^2 per inhabitant, for a *per capita* contribution of 100 *l*/day

With regard to *COD*, it can be noted that this parameter behaves qualitatively in a similar way to the BOD_5. There is a *COD* decrease with increasing R_h, except in PP1, where the *COD* of the anaerobically pre-treated sewage increases initially and only then decreases gradually with increasing R_h. The increase of *COD* in PP1 can be attributed to the photosynthetic production of algae. Pearson and König (1986) showed that 1 mg *Chl a* corresponds to a *COD* of 300 mg/*l*. For a concentration in the range of 1000 to 3000 µg *Chl a* in FP and MP, the *COD*, in the pond could increase up to 900 mg/*l*. Another relationship between *COD* and *Chl a* (Equation (4.14)) was proposed by König (1984) after a study with a combined culture of algae.

$$DQO(mg/l) = 74.36 + 240.10X \tag{4.14}$$

where X is *Chl* *a* in mg/*l*

In actual fact these very high values of *COD* do not occur in WSP systems as can be noted in the example of Table 4.3. In that WSP system the effluent *COD* of the tertiary maturation ponds was a factor 0.80 to 1.34 times the effluent *COD* in the AnP of the series. The fact that the effluent *COD* in Figures 4.7 to 4.10 is not compatible with the chlorophyll a concentration in the pond can be explained on the basis of applied sampling procedure. To determine the *Chl* *a* concentration samples were withdrawn over the entire depth of the liquid column and at different points of the pond, because it is known that the spatial and temporal distribution of the algae is not uniform. Thus the apparent algal concentration in the pond may not be equal to the concentration in the effluent. Thus it is possible that ponds with a high algal concentration can have a low *COD* concentration in the effluent.

The decrease of *COD* in PP can be observed better in Figure 4.10, where the *COD* in unfiltered samples is approximately 220 mg/*l* in a pond with a *per capita* area less than 0.5 m^2/inhab and this value is reduced to 120 mg/*l*, for the range of 1 to 2 m^2/inhab. To explain this decrease of *COD* in PP several physical mechanisms like flocculation, settling and migration from the surface to the bottom of the pond must be considered: the high transparency of the pre-treated sewage (the values of UASB effluent turbidity were in the range of 30 NTU throughout the investigation) allowed a high incidence of sunlight in the PP and caused an intensification of the migration of algae from the top to the bottom of the pond.

Due to the abundant presence of algae in the effluent of the PP, the value of the effluent *COD* remained high and, only rarely, an unfiltered effluent *COD* of less than 100 mg/*l* was observed (Figure 4.10(***b***)). By contrast, for the filtered *COD* values well below 60 mg/*l* can be expected, when the *per capita* area is in the range of 1.0 a 1.5 m^2.

In terms of suspended solids removal the polishing pond is not particularly efficient once again because of the presence of algae. However the quality is compatible with the recommendations of the Council of the European Communities for discharge into surface waters, where the limit of *TSS* is ≤ 150 mg/*l* (Table 4.1). From the results in Figure 4.7(***c***) it can be seen that the concentration of suspended solids decreases with time, but also depends on the depth. From Figure 4.10(***c***) it follows that the required *per capita* area to obtain a low *TSS* effluent concentration of 30 mg/*l* in the PP is about 1.5 m^2, independent of the pond depth.

It is interesting to note that a small but stable algae population (concentration of *Chl* *a* of

123 μg/*l*) established in the first lane of polishing pond PP1, although the liquid retention time in this pond was only 1 day. The algae concentration tended to increase during the first 5 days and then decreased as can be seen in Tables 4.9 and 4.10. This decrease influenced other parameters *TSS*, *COD* and BOD_5 but also the dissolved oxygen (*DO*) concentration (measured 5 cm below the surface and at 8 a.m.), as can be seen in Table 4.10, when R_h in each lane was 3 days. During the 3rd phase of PP1 operation (R_h = 15 days), small crustacea called *Daphnia*, predators of algae, established themselves on several occasions and during long periods in the ponds, especially in lanes 3, 4 and 5. The population of *Daphnia* varied severely without an identified reason.

Helminth Eggs Removal

PP1 was very efficient in the removal of helminth eggs (HE). During the first phase of operation (total R_h = 5 days) no *HE* were found in the effluent of the 3rd lane which corresponded to R_h = 3 days. Also in the 2nd (R_h = 7.5 days) and 3rd (R_h = 15 days) phases *HE* were not found in the effluents of the 2nd and 1st lanes respectively also corresponding to R_h =3 days. These observations lead to the conclusion that the removal of *HE* was complete when the surface loading rate was less than 0.22 m/day.

Considering that the flow regime in the pond was laminar (Reynolds number < 0.5) the settling velocity of the *HE* can be calculated by Stokes' law (Equation (4.15)):

$$v_s = \frac{g.(S_p - 1)d_p^2}{18.\nu} \tag{4.15}$$

where:

g gravity constant (9.81 m/s^2);

S_p relative density of the *HE*;
HE density /water density (ρ_{HE}/ρ_{H2O});

d_p diameter of the *HE*;

ν kinematical viscosity of water .

With the aid of Equation (4.15) and considering the maximum and minimum diameter of *Ascaris* (40 and 70 μm) and a relative density of 1.11 the calculated settling velocity is between 6.4 and 14.3 mm/min or 9 and 21 m/day for a temperature of 25 °C. These values exceed significantly the determined vertical settling velocity of 0.22 m/day. This velocity matches closely the determinations by Ayres and Mara, 1996: average settling velocities of 20, 16 and 6 mm/min for *Ascaris lumbricoides*, *Trichuris trichiura* and hookworm eggs, respectively in standing of water. It can be concluded that the settling of *HE* in ponds is severely hindered, probably by the vertical mixing induced by the mechanical action of the wind, biogas bubbles, thermal mixing and organisms present in the pond. Even so, the removal of *HE* is complete after a relatively short retention time and normally will not be the limiting factor with respect to the required retention time of a PP when a particular final effluent quality is to be obtained.

On the basis of the experimental data for PP1, it can be estimated that, for a *per capita* wastewater contribution of 100 *l*/day, a *per capita* area of 0.1/0.22 = 0.45 m^2 is required for the removal of *HE* in a PP. In practice the required area for the correction of other parameters is larger and hence it is to be expected that the effluent of a PP will be substantially free from helminth eggs. However, upon settling the *HE* are incorporated in the bottom sludge, where they can remain viable for very long periods (several years in the case of *A. lumbricoides*). Although the development of these eggs is interrupted in the anaerobic environment of the bottom sludge, the eggs of *A. lumbricoides* can resume their development as soon as they enter in contact with an aerobic environment (Cram, 1943). This is an aspect that must be taken into consideration when considering the possibility of utilization of the bottom sludge as an organic fertilizer in agriculture.

Faecal Coliform Removal

The experiments conducted to evaluate the decay of faecal coliforms in PP can be divided into two parts (1) for the experimental determination of the decay constant in batch ponds and (2) for determination of the *FC* removal efficiency in flow-through ponds (PP1). In the first part the depths of the batch ponds were different, because there were indications that depth would affect the value of the constant. The experiments conducted in ponds with depths of 0.65, 1.65, 2 and 3 m showed that, in all cases, the decay of *FC* was approximately exponential, at least until 10^3 CFU/100 ml had been reached. Also it was established unequivocally that the value of the decay constant depended on the pond depth (Table 4.12). In Figure 4.11 the experimental values of k_d have been plotted in relation to the pond depth. In good approximation, the data conform to a hyperbolic equation and therefore the value of the constant can be written as:

$$k_{dH} = k_{d1}/H \tag{4.16}$$

where:

k_{dH} decay constant for any pond depth : $0.65<H<3$ m (day^{-1});

k_{d1} decay constant in a pond with 1 m depth;
1.36 /day for a temperature of 25 ^{o}C (prevailing water temperature during the investigations);
H depth of the pond (m).

For other temperatures the k_d values can be corrected by applying Equation (4.13).

The reduction of *FC* undoubtedly is the most important parameter of pathogen removal in polishing ponds. Since the *FC* are more resistant than other micro organisms, their reduction is a clear indication that other harmful organisms will not be present at concentrations that can lead to infection. Since it has been established that the decay (death) rate of *FC* is a first order process the most efficient flow regime to reduce their concentration is plug flow, (Equation (4.7)). However in the experimental investigation the exponential decay was observed only in the tests with batch ponds not in flow-through ponds. The system PP1 was designed to achieve a 99.99 % removal efficiency at a retention time of 4 to 5 days but, in reality, 9 to 12 days were required as shown by the data in Tables 4.9 and 4.10, which is more than double the value calculated for exponential decay.

Basically three factors influence the decay of *FC* in polishing ponds: (1) the influent *FC* concentration which, in turn, depends on the concentration in the raw sewage and the removal efficiency in the anaerobic pre-treatment reactor, (2) the value of the decay constant (k_d) which depends on several factors among which are depth and temperature and (3) the flow regime in the pond, notably the mixing intensity.

In order to explain the deviation of the *FC* removal efficiency from the expected value, experiments were carried out to determine the value of the decay constant (described in Section 4.3) and the value of the dispersion number D in the pond.

The values obtained for the dispersion number (D) indicate a high dispersion, in each of the 3 phases, even though the pond was constructed specifically to induce a flow regime that would approach plug flow. The determined dispersion numbers were: 0.14; 0.16 and 0.18 for R_h values of 5, 7.5 and 15 days, respectively (see Table 4.11). These values are 7 to 9 times larger than would be expected from the Agunwamba's model (Equation (4.11)). The large dispersion number must be attributed to some kind of mechanical mixing, even though the pond had laminar flow, e.g. enforced by wind (Campina Grande is known for its mild breeze) and rising gas bubbles likely consisting of oxygen (supersaturation during daytime especially in the top layer) or biogas produced in the anaerobic zone in the bottom sludge. Also a variety of macro organisms that emerged in the ponds may induce mixing due to their motion. The measured dead volume fraction was also surprisingly high (≥ 12%, Table 4.11), even though attempts were made to reduce it as much as possible by forcing the flow from the top to the bottom at each connection between the lanes (Figure 4.5), so as to avoid short circuiting as much as possible.

In Figure 4.12 the *FC* removal efficiencies in the different lanes and during the different phases have been plotted as a function of the liquid retention time (semi log scale). In the same diagram the theoretical profiles of *FC* decay with R_h, according to the Wehner and Wilhelm equation (Equation (4.8)) have also been plotted for different values of the dispersion number. A value of 2.1 day^{-1} has been adopted from the average experimental value of the decay constant (Table 4.11). It can be noted that the theoretical curve for a dispersion number $D = 0.18$, most closely describes the experimentally observed behaviour. From Figure 4.12 it can be estimated that the R_h for a *FC* removal efficiency of 99.99 % would be 9.8 days whereas R_h for exponential decay (observed in the batch ponds) was only 4.3 days. This leads to the conclusion that due to imperfections in the flow regime of the ponds the liquid retention time (and hence the volume) have to be a factor 9.8/4.3 = 2.3 larger than the theoretical value in a pond without mixing ($D = 0$).

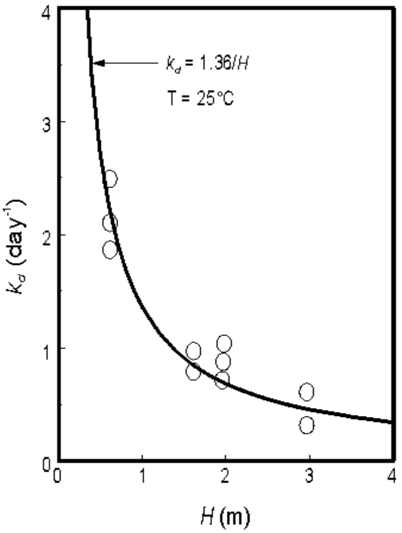

Figure 4.11. Graphical representation of the experimental values of the decay constant in ponds with different depths ($T = 25\ ^oC$).

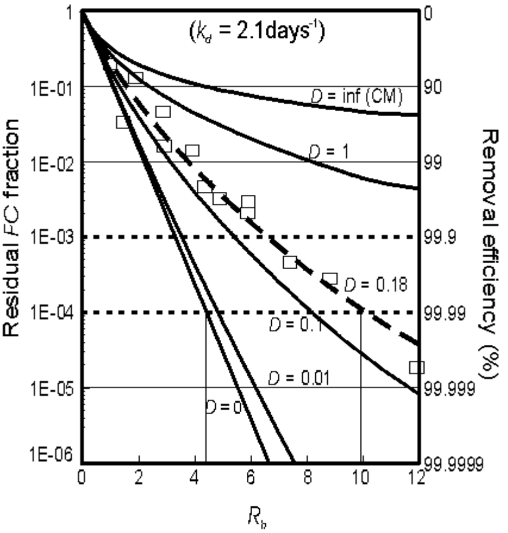

Figure 4.12. Experimental relationship between R_h and the *FC* removal efficiency in flow-through ponds with a depth of 0.65 m. ($T = 25\ ^oC$).

The experimental data show very clearly that the *FC* removal in polishing ponds can be described by the Wehner and Wilhelm equation (Equation (4.8)) and therefore it is a first order process. However it was also shown that it is not possible to operate a pond with a low mixing intensity, even though the pond was specifically designed to induce a plug flow regime. The considerable mixing led to a substantial reduction of the *FC* removal efficiency. If the mixing intensity is known, it can be calculated how much the liquid retention time must be increased to attain the desired removal efficiency, in spite of the mixing. However the mixing intensity, quantified by the dispersion number, can only be determined after construction of the pond. The methods to predict the dispersion number in ponds are still very inaccurate as was shown in this investigation and are of little use in practice.

The validity of the first order model of the decay constant could be verified for all investigated polishing ponds, including batch and flow-through ponds with partial mixing. In all cases the removal of *FC* was properly described by equations derived from the basic differential equation of the first order model for the different pond types.

Nutrient Removal

During the experimental investigation no significant removal of nitrogen and phosphorus in the flow-through polishing ponds was observed. This may be attributed to the fact that the pH in the ponds did not rise considerably; a high pH is a prerequisite for substantial nutrient removal. The highest pH was observed in the last lane of the shallowest pond, PP_1, where a maximum of 8.6 was measured. This value is insufficient for complete nitrogen removal and too low for significant phosphorus removal. Three reasons may have contributed to the low nutrient removal in the pond PP1: (1) If the pond is not very shallow, the required time to increase pH is long, even under the most favourable conditions (2) in the ponds the rate of photosynthesis was reduced due to the continued presence of *Daphnia* and (3) The alkalinity of the sewage was high (350-400 mg $CaCO_3/l$, so that a high buffer capacity existed. Biological consumption of carbon dioxide is the main process for the establishment of a high pH value in polishing ponds. In a parallel research study the processes that affect pH in polishing ponds were identified and the rates at which these processes developed were determined experimentally. It was shown that efficient nutrient removal is feasible in shallow ponds operated in sequential batch mode (to avoid the establishment of a *Daphnia* population).

4.6. Conclusions

The most important observation of the present experimental investigation is that efficient anaerobic pre-treatment of sewage has a profound and positive impact on post treatment in ponds: since the removal of organic material from anaerobically pre-treated sewage becomes relatively unimportant, for irrigation purposes, the pond can be designed based only on *FC* removal. To achieve an efficient *FC* removal there is no need, nor any advantage in having the conventional configuration of a series of waste stabilization ponds. Using a continuous flow regime, the post treatment can be carried out in a single maturation pond or, to use a more appropriate name, in a polishing pond.

The required size of a flow-through polishing pond is determined by the need for efficient *FC* removal, which is mainly determined by the hydraulic regime, the temperature and the concentration in the anaerobically pre-treated effluent. The required reduction in BOD_5, *TSS* and *HE* concentrations required a much shorter liquid retention time in a single pond system.

The validity of the first order model of the decay constant was verified for all investigated polishing ponds, including batch ponds and flow-through ponds with partial mixing. In all cases the removal of *FC* was properly described by equations derived from the basic differential equation for the first order model for the different pond types.

Under the conditions of this experimental investigation ($T = 25\ ^{o}C$) the required liquid retention time for a 4 log (99.99%) removal efficiency of *FC*, in a specially designed flow-through polishing pond (length/width ratio = 50, depth = 0.65 m), was about 10 days. The correction of BOD_5, *TSS* and *HE* was largely completed after a liquid retention time of 3 days.

Although PP1 had been designed to operate under plug flow conditions there were imperfections in the hydrodynamic regime as became apparent from the large values of the dispersion number (0.12 to 0.18) and the dead volume fraction (12 to 18 %). As a result the required time for 99.9 % *FC* removal in PP1 was about 10 days, more than double the theoretical value ($R_h < 5$ days) for a true plug flow regime, with exponential decay under the same conditions of depth and temperature.

The experimental value of the dispersion number *D*, in the PP1 was much larger than the theoretical value calculated from empirical models (Equations (4.11) and (4.12)). This may be due to the fact that these equations were derived from the operation of conventional

WSP systems, where a large dispersion number is a pre-condition for good performance of the pond. These models do not appear to be useful to describe the hydrodynamics of ponds, in which mixing is minimized. The experimental results indicate that a dispersion number of less than 0.1 is not really feasible in a flow-through polishing pond.

The batch pond studies allowed the determination of the decay constant value for *FC* and the influence of depth on the value of the constant. The experimental results suggest a hyperbolic relationship between the value of the decay constant and the pond depth: $k_{dH} = k_{d1} / H$

Under the conditions of the investigation, an effluent quality compatible with the standards for unrestricted irrigation was obtained in a flow-through polishing pond with retention time of 10 days, compared to a required retention time of 20 to 30 days in a conventional waste stabilization pond system.

Along with the very important reduction of liquid retention time (and hence area) of the PP compared to the conventional WSP, another advantage is that the PP does not generate the bad odours that at least occasionally emanate from anaerobic ponds in a conventional WSP system.

Due to the fact that the decay constant is inversely proportional to the liquid retention time, this parameter is not suitable as an unequivocal design parameter: the required liquid retention time for a particular *FC* removal efficiency is inversely proportional to the depth. By contrast surface loading rate and the *per capita* area of a polishing pond is a more adequate parameter, because these values do not depend on the pond depth, at least not for the investigated range from 0.65 to 3 m.

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Chapter 5

Post Treatment of UASB Reactor Effluent in Sequential Batch Polishing Ponds

5.1 Introduction

As it was not possible to achieve exponential decay in polishing ponds being fed continuously, these were then operated in batch mode as sequential batch reactors (SBR). In this chapter it will be shown that, under batch feeding, exponential decay of *FC* did occur and a high removal efficiency of *FC* (99.99%) was obtained in ponds with a short liquid retention time (4 to 5 days).

5.2 Considerations about *FC* Removal in Sequencing Batch Polishing Ponds

Having verified experimentally that in flow-through ponds mixing is inevitable and, for that reason, the *FC* removal efficiency is reduced so that the required liquid retention time is much larger than the theoretical minimum, it is concluded that the only way to have exponential decay of the *FC* in ponds is to operate these in batch mode. In batch ponds the surviving fraction of *FC* can be described by as Equations (4.7) and (4.16) at a temperature of 25 ^{o}C:

$$\frac{N_e}{N_i} = \exp\left(-k_{dH}.R_H\right) = \exp\frac{-k_{d1}}{H}.\frac{AH}{Q}$$

or

$$\frac{N_e}{N_i} = \exp\left(\frac{-1.36}{\lambda_s}\right) \tag{5.1}$$

The efficiency *E* can be expressed as a function of the applied surface loading rate (λ_s):

$$E = 1 - \frac{N_e}{N_i} = 1 - \exp\left(-\frac{1.36}{\lambda_s}\right)$$

or

$$\lambda_s = \frac{-1}{0.7\ln(1-E)} \tag{5.2}$$

In Figure 5.1 the *FC* removal efficiency is plotted as a function of the hydraulic surface loading rate for a temperature of 25 ^{o}C (k_{d1} = 1.36/day). In the same figure the maximum surface loading rate for completely *HE* removal (0.22m/day) is also indicated.

Figure 5.2 shows the *FC* removal efficiency as a function of the *per capita* area, assuming a *per capita* contribution of 100 *l*/day. The required area per inhabitant to remove *HE* is also indicated. The curve of Figure 5.2 is Equation (5.3), which expresses the relationship between the *per capita* area and the *FC* removal efficiency:

$$A_{inhab} = \frac{Q_{inhab}}{\lambda_s} = -0.7Q_{inhab}\ln(1-E) \tag{5.3}$$

where Q_{inhab} is the *per capita* flow.

The relationships that allow determination of the *FC* removal efficiency as a function of the surface loading rate or the *per capita* area show that this efficiency does not depend on the depth.

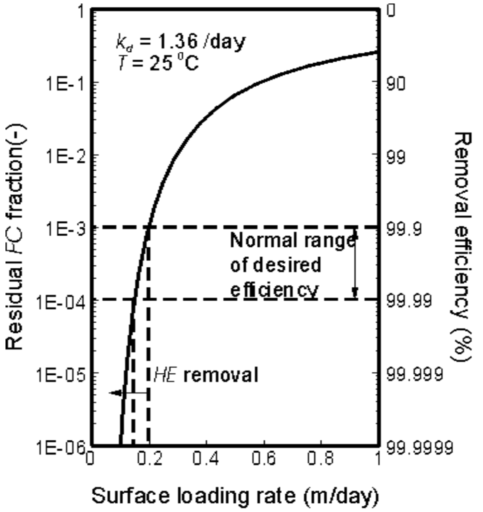

Figure 5.1. *FC* removal efficiency as a function of the hydraulic surface loading rate in batch ponds (the maximum rate for complete removal of *HE* is also indicated).

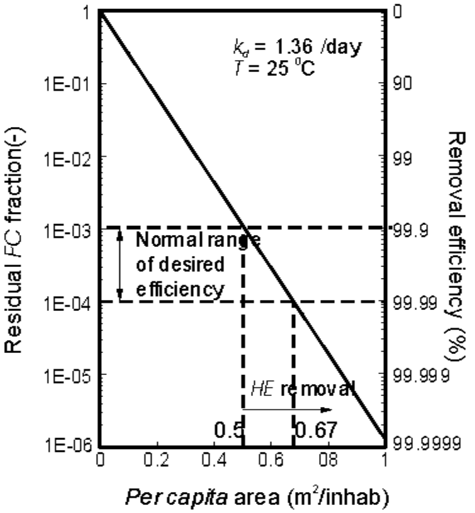

Figure 5.2. *FC* removal efficiency as a function of the *per capita* area (contribution of 100 *l*/day) in batch ponds (the minimum area for complete *HE* removal is also indicated).

When Figures 5.1 and 5.2 are analyzed, it is seen that if a *FC* removal of more than 99.99% is to be achieved, the maximum surface loading rate for *FC* removal is smaller than the maximum hydraulic surface loading rate for *HE* removal and the minimum *per capita* area is higher for *FC* removal than for *HE* removal. The required area per inhabitant for 99.99 % *FC* removal is 0.67 m^2 (Figure 5.2), which represents only 1/4 of the area normally used for conventional WSP's. It is interesting to note that it would be practically impossible to have such a *per capita* area for conventional WSP's: the organic load would be excessive and the entire pond system would be anaerobic, resulting in an unsatisfactory removal of organic material, which would be the main objective of a conventional pond.

When Figures 5.1 and 5.2 are compared with Figures 4.10(***a***), it can be noted that the required area for 99.99 % *FC* removal (0.67 m^2/inhab) is practically half the value necessary to produce an effluent with a filtered BOD_5 concentration of 25 mg/*l* (1.3 m^2/inhab).

5.3 Operation of Sequencing Batch Polishing Ponds

In this sub section the configuration and operation of sequencing batch polishing ponds

(SBPP) for the post treatment of effluent of UASB reactors are described. The objective of these ponds is to assure the exponential decay of *FC*, thus minimizing the liquid retention time.

Figure 5.3 shows schematically two possible SBPP configurations. The difference between the configurations is the presence of a transfer pond (TP). In the 1st configuration, Figure 5.3(**a**), the SBPP system is composed of a series of *M* ponds, all operating in parallel and each with a volume equal to the daily production of sewage. Every day the digested sewage is discharged into a particular pond, filling it. The digested sewage flow for the next day is then diverted to another pond in the sequence, which has been emptied previously. Through this procedure each pond receives, during 1 day, the daily flow of digested sewage with a frequency of 1 per *M* days. The value of *M* must be sufficient to attain the desired effluent quality. For this configuration it is considered that the treatment initiates effectively at the end of the feeding day, just after having completed its volume.

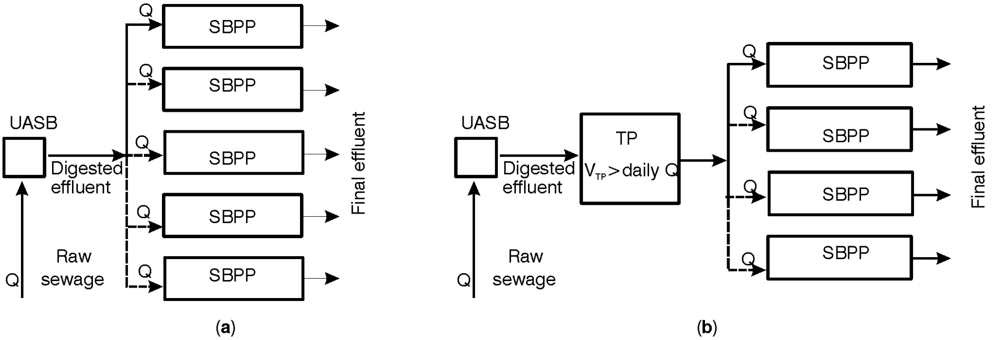

Figure 5.3. Configurations for sequencing batch polishing ponds, without (a) and with a transfer pond (b).

In the 2nd configuration, represented in Figure 5.3(**b**), there is a transfer pond in which the anaerobically pre-treated sewage production over a day is accumulated. Once per day, always in the morning, so that there is full benefit of the sun for photosynthesis, a batch of accumulated digested sewage is discharged into one of the SBPP, which has again a volume equal to the daily sewage production. This batch will remain in the pond for a sufficient period to acquire the desired effluent quality, whereupon it will be emptied and be ready to receive a new batch. Due to the fact that with the presence of the transfer pond the SBPP can be filled rapidly, the number of required ponds *N* will be one less than in the case without the transfer pond: $N = M - 1$.

The criterion for dimensioning the SBPP in the 1st configuration is straightforward and is established by the minimum liquid retention time to effect the required *FC* removal, since the correction of other undesirable constituents like residual BOD_5 and *TSS* and *HE* will require a shorter liquid retention time. The R_h for *FC* removal can be calculated immediately from the basic kinetic equation (Equation 4.3). The solution for a batch pond is the same as for a plug flow pond (Equation 4.7), which means that the decay of *FC* in a batch is exponential with time. Once the values of the *FC* in the digested sewage and the desired *FC* concentration in the effluent are known the R_h can be calculated for any value of the decay constant. In the case of the 2nd configuration the calculation is less straightforward because first the *FC* concentration in the effluent of the transfer pond must be calculated. This concentration will depend on the R_h of the digested sewage in the TP.

The volume of accumulated digested sewage in the TP varies between a minimum, V_{min}, just after a discharge to a SBPP, and a maximum, V_{max}, just before such a discharge. It will prove to be convenient to introduce the ratio *R* as the proportion between the V_{min} and the daily flow: R = V_{min}/Q. Similarly the R_h can be equated to the ratio between the maximum volume and the flow: $R_h = V_{max}/Q = R+1$.

The basic differential equation (Equation 4.3) can also be used to calculate the removal of *FC* in the transfer pond. This removal will be low because there is a continuous inflow of digested sewage into the TP with a high concentration (N_i). If complete mixing is assumed in the TP, there will be a cyclic profile in the *FC* concentration: the *FC* concentration at the end of the day will be equal to the concentration at the beginning (instantaneous filling of the SBPP is assumed) and, at some moment during the day, the number of *FC* in the TP will be maximum. Just after the discharge, when the volume of sewage in the TP is minimum, the concentration of *FC* increases, because the rate of introduction with the influent is greater than the decay rate. The increasing *FC* concentration leads to a higher decay rate and the increase of the volume of liquid in the TP leads to a decrease of the rate of increase of the *FC* concentration. Thus, at some stage, the rate of decay will become greater than the rate of increase due to the feed and the *FC* concentration in the TP will start to decrease with time. The solution of the differential equation, with the boundary condition that the *FC* concentration in the TP is equal at the beginning and the end of a cycle of 1 day, is given by:

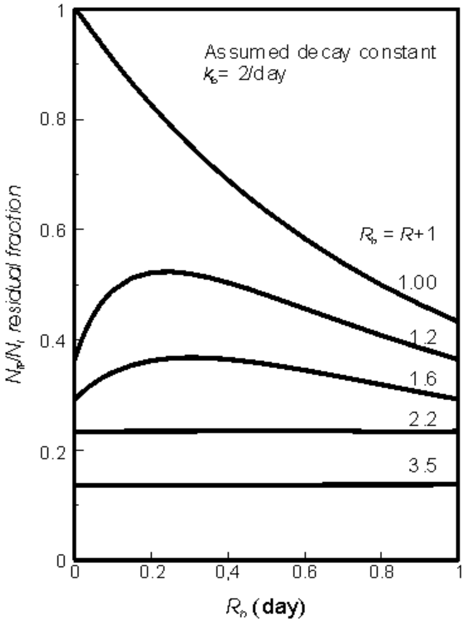

Figure 5.4. Variation of the residual *FC* fraction in a TP for different values of R_h=R+1.

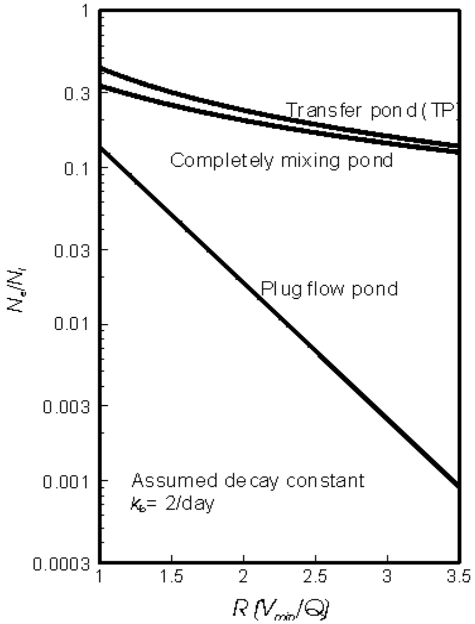

Figure 5.5. Residual *FC* fraction in completely mixed TP as well as in complete mix ponds with a constant volume and plug flow or batch ponds. (SBPP), as functions of R_h.

$$\frac{N_{TP}}{N_i} = \frac{1 - \exp\left(-k_d t\right)}{\left[k_d\left(R+t\right).\left(1 - R\left(\exp\left(-k_d\right) - 1\right)\right)\right]} \tag{5.4}$$

where:

N_{TP} *FC* concentration in the TP;

N_i *FC* concentration in the influent of the TP;

R ratio between the minimum volume of liquid in the TP (t = 0) and the daily flow of sewage (R>0).

t time elapsed since the last discharge ($1 \geq t \geq 0$).

Equation (5.4) shows that the value of the ratio $R = V_{min}/Q$, (with dimension time) determines the residual *FC* fraction N_{TP}/N_i in the TP. Figure 5.4 is a graphical representation of the residual fraction of *FC* in the TP over a period of 1 day between two successive discharges for R ratios between 0 and 2.5 (and R_h between 1 and 3.5). When R = 0 the volume of the TP is equal to the daily flow and the unit becomes effectively a SBPP with a R_h of 1 day because at the end of the filling period the volume is completely emptied. For R > 1 the residual *FC* fraction depends on the value of this ratio. Naturally the larger the fraction in the TP effluent, the bigger the volume of the SSBP receiving this effluent will have to be. Thus the question presents itself, which is the optimal value of the ratio R.

Figure **5.5** is a graphical representation of the residual *FC* fraction in the discharges of a TP (Equation (5.4)) for different values of the liquid retention time $R_h = R + 1$. For comparison the values of the residual *FC* concentrations in flow through ponds with complete mixing (Equation (4.4)) and with a plug flow regime (Equation (4.7)) are also represented. It can be noted that the TP is even less efficient than the completely mixed flow-through pond. This is due to the fact that the volume of the former is not fully used. In Figure 5.5 it can be seen that the *FC* removal efficiency in the range of R_h from 1 to 3.5 days is of the order of 60 to 85 %. The difference of the performance of TPs with the minimum (R_h = 1day) and maximum retention time (R_h = 3.5 days) in terms of *FC* removal is so small that for all practical purposes it may be ignored. Thus purely from the point of view of *FC* removal efficiency, the optimal value of the TP volume is equal to the daily flow of digested sewage. In that case the removal efficiency of the *FC* is given by the Equation (5.4) for $R = 0$ and t = 1 day: N_{TP}/N_i = (1-exp$(-k_d))/kd$ and the subsequent SBPP can now be designed by calculating the required retention time:

$$R_h = \frac{1}{k_d}\left(\ln\frac{N_i}{N_e} + \ln\frac{1-\exp^{-k_d}}{k_d}\right) \tag{5.5}$$

However other considerations may lead to a design with a larger volume for the TP. A liquid retention time of only 1 day in the TP will be insufficient for the algae to develop, so that the pond would only be an accumulation and settling tank. At larger retention times algae would develop and these could not only assist in the BOD_5 removal of in the TP, but also the BOD_5 and *TSS* removal in the subsequent SBPP would be more rapid as the influent would already contain algae. In addition the accumulation of bottom sludge (and particularly the excess sludge from the UASB reactor) would occur mainly in the TP, which would facilitate its removal and use.

5.4 Experimental Investigation and Results

An experimental investigation was carried out to evaluate the behaviour of sequential batch polishing ponds and to determine the optimal value of the liquid retention time of the transfer pond. The experimental system at pilot scale was composed of a UASB reactor, several TPs and SBPPs. The same UASB reactor described earlier was used, now with R_h =5 hours. The UASB effluent (digested sewage) was used to feed the TPs and from these the sewage was transferred to the SBPPs. The TPs were brickwork tanks with a working depth of 0.65 meters and a volume of 1400 litres.

Table 5.1 shows the changes in the digested sewage characteristics in a SBPP after having been fed with UASB effluent with a volume of 1400 litres (1st configuration, without a TP). These changes were observed during the operational period of 5 days, in samples taken after gentle mixing of the pond contents: the *DO* concentration measured at 8 a.m. in the top section (5 cm from the surface) of the liquid layer varied from <1 mg/*l*, on the first day after feeding the SBPP, to > 20 mg/*l*, on the fifth and last day of operation. Although there was no detectable *DO* concentration (on the first day and the algae population was very small), no bad odours were perceived. The temperature at mid depth remained approximately constant (25 °C), but at the surface there were large variations from a minimum of 20 to a maximum of 35 °C.

The decay of the *FC* was, in good approximation, exponential and the value of the decay constant that was calculated from the data (k_d = 2.3 days^{-1}), was compatible with earlier results (Section 4.4) as well as those of other research workers under similar condition (von Sperling, 1999 and Yanez, 1993). It should be noted that there was little variation in the BOD_5 and *TSS* concentrations in the pond after the third day of operation. In separate tests it was established that there were no *HE* in the effluent after 5 days of operation but these were present in the solids on the bottom of the TP after discharge of the effluent.

Table 5.1. Main characteristics of UASB reactor effluent and its evolution in a SBPP during a treatment period of 5 days (depth = 0.65 m, volume = 1400 litres).

Parameter	Unit	UASB effluent	R_h (days)				
			1	2	3	4	5
FC	CFU/100 ml	7.8E7	1.3E7	4.0E5	2.5E4	5.6E3	6.3E2
ln(*FC*)	-	7.89	7.11	5.60	4.40	3.75	2.79
BOD_5	mg/*l*	66	52	72	65	55	52
TSS	mg/*l*	89	67	56	65	72	59
COD	mg/*l*	145	128	181	198	156	125
pH	-	6.9	7.1	7.2	7.2	7.3	7.3
TAlk	$mgCaCO_3$/l	355	358	342	348	348	322
TKN	mgN/l	52	49	52	55	50	45
P_{total}	mgP/l	8.4	8.0	8.0	8.2	8.2	7.8
Chl. a	μg/*l*	-	-	120	670	1221	1867

The data in Table 5.1 show that the *FC* removal efficiency for R_h = 5 days was of the order of 99.999%, sufficient to meet the most stringent standards for irrigation. However it was insufficient to alter considerably the *pH*: its value increased slowly from 6.9 to 7.3. For that reason the chemical equilibrium in the pond was almost unaffected and no removal of nutrients was observed. It was concluded that the SBPPs were very suitable for the production of water for irrigation because they are extremely efficient in the removal of pathogens and, at the same time, they preserve the nutrients and maintain the *pH* in the neutral range, even in shallow ponds. The short R_h, apart from being a factor in cost reduction, is important because it limits the evaporation losses of water. Thus indirectly the SBPPs contribute to the preservation of a third important factor for irrigation: water with low salinity.

In order to establish the optimal R_h for transfer ponds, five units were operated in parallel each with a volume of 1400 *l* and a depth of 0.65 m. The ponds were operated with different R_h and R values: the minimum liquid volume varied from 0 to 2.5 times the applied daily flow. Due to the variation in the R value, the applied flow varied from 1400*l*/day to 400 *l*/day, so that R_h varied from 1 to 3.5 days.

The operation of the transfer ponds can be summarized as follows: at 8 a.m. a volume corresponding to the daily flow was discharged semi instantaneously from the TPs and then they were fed at constant flow until they reached a maximum volume again 24 hours later. This procedure was followed every day. Samples from the discharged volume were separated and used for analyses.

Table 5.2 shows the values of the flows applied to the different TPs and the main characteristics of the effluent from the TPs. The data show the *FC* removal efficiency was low and erratic as was to be expected from theory (Equation (5.4)). It is interesting to note that a liquid retention time of 1.6 day was sufficient to maintain an algal population of approximately half the maximum value at 2.2 days (TP_2), however for a retention period of 1.2 days the algae population was very small. For this reason tentatively it is suggested that a R_h of 1.6 days be applied for transfer ponds. The volume of such a pond would only be a fraction 1/1.6 = 62 % greater than the absolute minimum value. The residual volume after discharge (38 %), would be sufficient to inoculate the digested sewage (to be introduced) with an algal population and start the process of photosynthesis. The photosynthetic activity was sufficient to suppress the bad odours occasionally emanating from the UASB effluent. At the same time the extra volume would be used for the accumulation of solids from the UASB reactor, which could be returned to the UASB reactor or disposed to land.

The 2nd configuration (TP+SBPP) was tested using a TP with R_h = 1.6 days and a SBPP with a depth of 0.4 m and R_h = 5 days, with the objective of evaluating *FC* removal. Table

5.3 shows the data from these systems.

According to the data in Table 5.3, the time in the SBPP required to obtain a concentration of less than 1000 *CFU*/100 ml was only 4 days, which is compatible with the theoretical prediction assuming exponential decay. The increase in *pH* was quite significant: values of *pH* > 9.0 were obtained with a retention time of 5 days. This can be attributed to the development of intense photosynthetic activity, which in turn is due to the inoculation of the digested sewage with algae in the transfer pond. Due to the pH increase there was also significant removal of the macronutrients nitrogen and phosphorus.

Table 5.2. Applied daily flows retention times and the main effluent characteristics of five transfer ponds with volumes of 1400 *l* and depths of 0.65 m.

	Unit		TP_1	TP_2	TP_3	TP_4	TP_5
Daily applied flow	*l*/d	Influent	400	650	900	1150	1400
R_h	day		3.5	2.2	1.6	1.2	1.0
FC	CFU/100 ml	5.4E7	3.8E5	6.0E7	3.8E6	1.4E7	8.8E6
HE	no./*l*	56	0	4	6	13	5
BOD_5	mg/*l*	93	58	66	81	73	71
TSS	mg/*l*	78	67	52	57	52	58
COD	mg	135	121	142	157	124	110
pH	-	6.9	7.4	7.3	7.2	7.0	7.0
TAlk	$mgCaCO_3/l$	357	340	351	369	349	355
TKN	mg*N*/*l*	46	42	46	44	41	46
P_{total}	mgP/l	7.1	7.3	7.0	7.0	7.3	7.2
Chl. a	µg/*l*	-	1182	1553	658	117	-

Table 5.3. Variation in quality parameters in a SBPP, fed from a TP with R_h = 1.6 days, during 5 days of operation (depth = 0.4 m, samples were taken at 8 a.m. after gently mixing).

Parameter	Unit	Operational time (days)					
		0	1	2	3	4	5
FC	CFU/100 ml	7.1E6	9.7E5	1.2E5	4.7E3	5.8E2	1.1E2
log(*FC*)	-	6.85	5.98	5.08	3.67	2.76	2.04
DO	mg/*l*	0	<1	3.4	17.2	>20	>20
BOD_5	mg/*l*	77	56	49	49	52	42
TSS	mg/*l*	86	57	46	52	43	56
COD	mg/*l*	155	121	147	167	144	129
pH	-	7.1	7.3	7.8	8.6	8.8	9.5
TAlk	mgCaCO$_3$/l	345	355	351	326	289	270
TKN	mg*N*/*l*	51	52	49	48	34	12
NH_3	mg*N*/*l*	42	38	34	24	11	1.8
P_{total}	mgP/l	7.8	8.0	7.7	7.9	6.8	5.7
Ortho-P	mgP/l	5.3	5.4	4.9	4.5	3.9	2.2
Chl. a	µg/*l*	740	1028	1273	1851	1537	894

5.5 Discussion of the SBPP Results

When polishing ponds are used to improve the quality of anaerobic effluents with the intention of using the final effluent for irrigation, the process that determines the required size of the PP is the removal efficiency of *FC* because this process is slower than the removal of BOD_5, *TSS* and *HE*. The most important reasons for using sequential batch ponds is that in these the *FC* removal rate is maximum and hence higher than in a flow through pond of the same dimensions. For the experimental conditions (temperature = 25 ± 2°C, depth = 0.60 m) the required minimum liquid retention time required to obtain 99.99% removal efficiency in a SBPP was only 4 to 5 days, whereas in a flow-through PP with an optimized design to minimize mixing a liquid retention time of 10 days was required. It is concluded that the SBPP can be operated at half the retention time of optimally designed flow-through PPs and yet produce the same hygienic effluent quality.

The results of the experimental investigation presented in this section show that the decay of *FC* in SBPP was exponential as predicted by theory and that the decay rate was the maximum that can be obtained. The good correlation obtained between the experimental results and the prediction on the basis of kinetic theory of a first order process (Equation (4.7)), indicates that the theory can be used for the design of SBPP. To define the dimensions of the pond unequivocally it is also necessary to know the value of the decay constant, which is influenced by temperature and depth. The results of the two tested configurations using SBPP with (Figure 5.3(b)) or without (Figure 5.3(a)) a transfer pond show that:

- in case of the 1st configuration, for kd = 2.3/day and a desired *FC* removal efficiency of 99.99 %, the required retention time is Rh = 4 days. Taking into account that the filling time of a pond is 1 day and knowing that the *FC* removal during filling is negligible (Figure 5.4) it is concluded that 5 SBPP ponds are required each with a volume equal to the daily flow. Hence the total retention time would be 5 days. In practice the number would probably be higher to accommodate unforeseen contributions like rainwater or an increase in the contributing population and also to have a reserve treatment capacity for maintenance of the ponds;
- with respect to the 2nd configuration, the experimental results show that the *FC* removal in the TP is very small as foreseen by Equation (5.4) so that, for practical purposes, the *FC* removal in the TP can be neglected. Thus in the subsequent SBPP the required removal efficiency would still be 99.99% and hence the retention time Rh = 4 days. Considering that the filling of the pond from the TP would be almost instantaneous, there would be a need for 4 SBPP units again, each with a volume equal to the daily flow. Under the conditions of the investigation, the optimal liquid retention time for the TP would appear to be 1.6 days, which allows the presence of an active and stable algae population. In this case the total retention time would have to be 4 + 1.6 = 5.6 days, which is only marginally more than the total retention time of 5 days in the first configuration, so that in terms of required area and construction costs the two alternatives are approximately equivalent.

The inclusion of a transfer pond can have some practical advantages:

- the algae present in the effluent of the TP will act as an inoculum for the SBPP, so that in these photosynthesis can take place as soon as they are filled;
- the TP will have a sufficient liquid retention time for the removal of the settleable solids in the digested sewage, including a considerable part of the HE thus concentrating an important fraction of these solids in a single pond rather than being spread over all the ponds and thus simplifying the cleaning operation. This is particularly important when there is no excess sludge withdrawal from the preceding UASB reactor;
- if the liquid retention time is sufficient for the presence of an expressive algae population, the photosynthetic oxygen production can be used to oxidise sulphide, eventually present in the digested sewage. In this case bad odours are avoided which

is an important consideration if the pond system is to be constructed near or within an urban area;

- the TP will also act as an equalization tank, dampening fluctuations in the quality of the digested effluent and absorbing hydraulic overloads, which can include raw sewage when the flow to the UASB reactor is excessive. The rapid transfer of the overload to the SBPP (and eventual reserve units) is an effective means of minimizing the impact of hydraulic shock loads.

A disadvantage of the sequencing batch pond compared to flow through ponds is that the operation of the former is more complicated, because every day the digested sewage must be transferred to a different pond and every day the contents of one pond must be discharged. This can either be done manually by operators or by a simple automatic control device. Even if the operation is automatic there will be a need to control the operation to make sure that the equipment is properly working. Depending on the topography of the pond area the transfer of the liquid between ponds may be done by gravity of by pumping, but even in the latter case the operational costs are very low.

At any rate, in all design cases it will be necessary to weigh the advantage of a large reduction in the area required for the pond system against a loss of operational simplicity. In the decision to use either flow-through or sequencing batch ponds for post-treatment of digested sewage the important variables are the availability and price of land at the location where the system is to be constructed.

5.5. Conclusions

The validity of the first order model was verified for all investigated polishing ponds, which included batch ponds and transfer ponds. In all cases the removal of *FC* was properly described by equations derived from the basic differential equation of the first order model for the different pond types.

Two different configurations of batch ponds were operated: (1) five parallel batch ponds sequentially fed with digested effluent from a UASB reactor and (2) a transfer pond receiving digested effluent transferring to four sequential batch ponds. It was shown that the first configuration requires a marginally smaller area, but the second configuration has operational advantages: (1) all discharged UASB sludge in the effluent is accumulated in the transfer pond and (2) the risk of odours in the pond system is reduced by having the transfer pond, where photosynthesis (and oxygen production) takes place.

The minimum liquid retention time for the transfer pond was 1.6 days; at smaller liquid retention times it was not possible to maintain the algae population and an anaerobic environment developed.

At the environmental temperature of 25 °C it was possible to obtain a 99.99 % *FC* removal efficiency in five parallel ponds each with five days retention time and in a system with a transfer pond with a retention time of 1.6 days and 4 parallel batch ponds each with a retention time of 4 days.

The required area of the sequential batch pond systems for a 4 log removal efficiency of faecal coliforms was approximately 0.67 m^2/hab. This area was also sufficient to obtain complete removal of *HE* and substantial reductions in the BOD_5 and *TSS* concentrations.

Even at the short retention time of 5 days partial nutrient removal in the batch ponds is feasible.

5.6 References

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Chapter 6

Sludge Accumulation in Polishing Ponds Treating Anaerobically Pre-Treated Sewage

6.1 Introduction

Settling of particulate material present in raw sewage is one of the main processes for removing organic material, from the liquid phase, in conventional waste stabilization ponds (WSP's) systems. In the typical WSP configuration, the settleable solids will be removed to become part of the sludge formed at the bottom of the anaerobic pond. The removal efficiency of the organic material and suspended solids in the anaerobic pond is in the range of 50 to 70% (Vincent *et al*., 1963) and (Yanez, 1993). A considerable part of the settled organic solids is transformed into biogas via anaerobic digestion processes proceeding in the system. The non-biodegradable organic and inorganic residue accumulates at the bottom of the pond, reducing gradually its volume, so that the pond performance tends to deteriorate with time. As a result, the bottom sludge of anaerobic ponds needs to be removed every 2 to 5 years, depending on the raw sewage characteristics, the liquid retention time and the average temperature. For this reason, often two anaerobic ponds are constructed, so that one can be taken out of operation for desludging. This increases the space requirements of the total treatment system.

Although settling and subsequent anaerobic digestion are the most important processes in anaerobic ponds of conventional WSP´s, the accumulation of the bottom sludge has also negative aspects. The removal of the bottom sludge, represents a major contribution to the overall operational costs. Moreover, in the process of anaerobic decomposition of the solids, sulphide is produced and part of this compound is released as hydrogen sulphide gas and thus is a major factor contributing to bad odours that usually emanate from conventional waste stabilization pond systems.

Data for the accumulation of bottom solids vary widely, as they depend on the composition of the influent and the operational conditions of the pond. In temperate climates, the rate of bottom sludge accumulation varies with the season with a steep accumulation during the period with low temperatures and a decrease of the accumulated sludge mass during the summer as a result of anaerobic digestion (Marais, 1966). In tropical regions, anaerobic digestion takes place throughout the year and a continuous slow bottom sludge accumulation occurs.

The sludge accumulation is often expressed in terms of volume per inhabitant per annum with a reported range from 0.03 to 0.09 $m^3.inh.^{-1}.annum^{-1}$ (Arceivala, 1985). Moreover, it has been observed that the concentration of the bottom sludge varies considerably, i.e. reported values are between 52 g/*l* (Paing *et al.*, 2001) and 120 g/*l* (Saqqar and Pescod, 1995). From the published data, it can be estimated that the amount of accumulated bottom sludge, in the case of raw sewage, ranges from 157 to 266 $mg.l^{-1}$ of influent (Van Eck and Simpson, 1966; Paing *et al.*, 2001) and (Yanez, 1993). Unfortunately the rate of bottom sludge accumulation is often given in terms of increase in the bottom layer thickness, which is only a useful parameter if the area of deposition is also provided, but this is generally not the case.

When ponds are used for post treatment of anaerobically pre-treated sewage, the nature and the concentration of the settleable matter is very different from that in raw sewage. In tropical regions, when using an efficient anaerobic reactor like, for instance, a UASB reactor, the removal of suspended solids achieved in the UASB reactor operated at retention times in the range of R_h = 4 to 6 hours will be in the range of 70% to 80% (van Haandel and Lettinga, 1994). Most of the settleable solids in the UASB effluent are anaerobiccally stabilised sludge particles. The production of these solids in the UASB unit is in the range of 100 to 150 mg *TSS*/*l* (van Haandel and Lettinga, 1994). Under steady state conditions, if no excess sludge is discharged, the average (over a sufficiently long period of time) settleable sludge concentration in the effluent will approach the net (excess) sludge production in the UASB reactor. The settleable solids concentration in the anaerobic effluent is much lower when periodically the excess sludge is discharged from the UASB reactor, usually well below 100 mg/*l* (Cavalcanti *et al.*, 1999). This value is very much lower than the suspended solids concentration in raw sewage, which typically is in the range of 400 to 600 mg/*l*. The accumulated sludge is always smaller than the total amount of suspended solids in the influent, not only because not all suspended solids will be removed by settling but also because a significant fraction of the influent solids is biodegradable and therefore will be converted into volatile fatty acids (VFA) and biogas by anaerobic digestion in the pond. On the other hand, the configuration of well designed polishing ponds for anaerobically pre-treated sewage is very different from that of a conventional waste stabilization pond system. Organic material removal in polishing ponds is a secondary consideration; normally the main objective is pathogen removal. As was demonstrated in Chapter 4, this can best be achieved in a single polishing pond, especially if this pond is subdivided into sections or lanes by appropriately placed baffles, so that mixing is minimized. The settleable solids of anaerobically digested sludge will tend to deposit in the first lane of the polishing pond. If any significant accumulation of bottom sludge occurs in the subsequent lanes, these solids predominantly will originate from processes in the polishing pond, such as algae which grow quite well as a result of good conditions for photosynthesis.

6.2 Experimental Investigations and Results

The bottom sludge accumulation was evaluated in the pilot scale polishing pond PP1, fed with pre-treated sewage from a UASB reactor. As mentioned in Chapter 4, the UASB reactor treated raw municipal sewage with an average temperature of 25°C and at a very short liquid retention time (R_h) of 3 hours. As a result of the presence of an efficient phase separator design (Chapter 2) the operational stability and treatment efficiency were high, leading to removal efficiencies for settled *COD* and BOD_5 in the range of 70% to 80% and 80% to 85 %, respectively. Therefore, the settled effluent BOD_5 and *COD* concentrations were in the range of 60 – 80 mg/*l* and 150 - 200 mg/*l* respectively

Since the UASB reactor was operated without intentional sludge discharges, the sludge mass in the effluent equals the net sludge production (corresponding to the excess sludge). The average total suspended solids concentration was 122 mg TSS/*l* with an organic material fraction of 65% and the volume of the settled solids amounted to 1.2 m*l*/*l*. Stability tests on the UASB effluent solids showed that anaerobically biodegradable fraction was 15% to 25%. The hygienic quality of the pre-treated sewage was still poor with a faecal coliform concentration in the range of 10^7 to 10^8 CFU/100m*l* and an average concentration of helminth eggs of 208 per litre.

The polishing pond PP1, which was described in Chapter 4, had a working volume of 32.5 m^3 and an area of 50 m^2 and had a concrete bottom and plastered brick walls. The unit was composed of five parallel lanes, operated in series, each lane was 10 m long, 1 m wide and 0.65 m deep. Figure 4.4 shows a schematic layout of the polishing pond, as well as the interconnections between the lanes. The pond was operated over a period of 1 year with liquid retention times of 5 to 7.5 days, so that over a year approximately 2000 m^3 of pre-treated sewage was polished in the pond. Thus the mass of suspended solids over the operational period of 1 year was estimated at 2000*x*0.122 = 244 kg *TSS*.

In order to determine the accumulated mass and volume of the bottom sludge after one year of operation the following procedure was applied: the influent flow was stopped and the supernatant was slowly siphoned off from each lane, taking care that no bottom sludge was lost in the process. The remaining thick sludge water mixture showed clear density stratification. In order to obtain representative samples, the bottom sludge was thoroughly mixed and samples were taken for the characterization of physical, chemical and biological properties of the solid fraction of the bottom sludge in each lane. Physically, the appearance of bottom sludge was similar to biological sludge: it did not contain any macroscopic particles with firm mechanical strength like those found in bottom sludges of anaerobic ponds. Apparently these particles had been removed and retained in the preceding UASB reactor.

The data in Table 6.1 summarize the values of the bottom solids mass, composition and properties in each of the five lanes, as well as in the total sludge mass. The real moisture content of the bottom sludge was lower than values mentioned in Table 6.1 (last line) because some water had to be left when siphoning off the supernatant of the pond in order to avoid sludge loss. The actual concentration of the bottom layer of the accumulating sludge was estimated by settling sludge samples in an Imhoff cone for 24 hours and determining the solids content. An extremely high *TSS* content of about 10% in all lanes (i.e. 90 % of moisture content) was found. Consequently the *TSS*-concentration of the bottom sludge in the pond can be estimated at about 100 g/*l*.

In the experiments Standard Methods procedures (APHA *et al.*, 1995) where used when these were available. Chlorophyll a was determined by the extraction method described by (Jones, 1979). The number of helminth eggs was estimated by using the Yanko recovery method described in USEPA (Yanko, 1987). Sludge stability was determined after incubating sludge samples at 30 °C for one month after dilution with pond effluent and it is represented as the VSS ratio after and before incubation. The biodegradable sludge fraction was calculated both from the measured methane production (van Haandel and Lettinga, 1994) and the reduction of volatile suspended solids. It was observed that hydrolysis and digestion of organic material in the bottom sludge of all lanes was insignificant: less than 5% of the volatile solids were digested during the sludge stability test.

Table 6.1. Mass and characteristics of sludge accumulated during 1 year at the bottom of a 32.5 m^3 polishing pond (PP1), with a yearly average R_h = 6 days and a volume of 2000 m^3.

Parameter	Unit	Lane 1	Lane 2	Lane 3	Lane 4	Lane 5	Total
Dry sludge mass	kg *TS*	77.3	13.9	21.5	16.4	12.0	141.0
Volatile sludge mass	kg *VS*	41.2	7.7	13.8	11.5	8.2	84.5
Volatile fraction	kg*VS*/kg*TS*	0.53	0.55	0.64	0.70	0.68	0.58$^{(1)}$
Nitrogen	g*TKN*/g*VS*	0.054	0.085	0.065	0.081	0.090	0.066$^{(1)}$
	g*TKN*/*gTS*	0.029	0.047	0.047	0.057	0.061	0.039$^{(1)}$
Phosphorus	g*P*/g*VS*	0.013	0.032	0.020	0.022	0.023	0.018$^{(1)}$
	g*P*/g*TS*	0.007	0.028	0.013	0.015	0.016	0.011$^{(1)}$
Chlorophyll a	g/g*SV*	0.0	38.9	44.1	38.4	45.6	-
Stability	mg*VS*/g*VS*	32	50	45	35	32	-
Helminth eggs	n^o/g*TS*	2071	2151	362	0	0	$1.9x10^{8(2)}$
Faecal coliforms	n^o/g*TS*	2960	3580	-	6060	3980	-
Moisture content$^{(3)}$	(%)	95.8	94.6	95.1	95.5	94.2	-

$^{(1)}$ average weight, $^{(2)}$ total number and $^{(3)}$ from settling test.

6.3 Discussion

Experimental data show that the accumulation of bottom sludge in a polishing pond, used for post treatment of anaerobically pre-treated sewage, is significantly smaller than generally observed in a pond system treating raw sewage, even despite the fact that no excess sludge was discharged from the UASB reactor pre-treating the raw sewage. This is to be expected because major fraction of the biodegradable organic material present in the raw sewage is converted in biogas in the anaerobic reactor. Moreover, it also is quite obvious that the relatively high sludge accumulation in the 1st lane of the pond (where more than half of the sludge accumulation takes place) mainly can be attributed to the satisfactory flocculation and settling characteristics of the sludge particles present in the UASB effluent. The sludge accumulation in the pond can be reduced substantially by applying regular sludge discharges from the UASB (Chapter 3).

Considering the results, it appears that the amount of inorganic sludge in the influent of the

1st. lane (0.35*x*244 = 85 kg) exceeds the amount of accumulated inorganic bottom sludge in this lane (77.3*x*0.47 = 36 kg), for the average TSS concentration of 122 mg/*l* in the 2000 m^3 that were polished in the pond and 60% VSS/TSS ratio. Apparently not all the suspended solids settled out in the first lane and/or some inorganic material (ammonia, phosphate, sulphide) was released from settled organic macro molecules, when they are decomposed. Interesting is the fact that the much lower organic fraction of the accumulated sludge in the 1st. lane as compared to that in the other ones. This can be attributed to the higher mineral suspended solids concentration in the influent of the first lane.

Even though in the 1st. lane more solids accumulated than the subsequent ones and the amount of sludge solids in this lane was more than half the total, the amount of bottom sludge in lanes 2 to 5 was much higher than expected on the basis of the amount of settleable matter in the liquid phase. The effluents of the different lanes were virtually free from settleable solids (in all cases, settleable solids tests with Imhoff cone yielded less than 0.1 ml/*l*). Apparently the sludge accumulated in lanes 2 to 5 occurs as a result of spontaneous flocculation of finely dispersed (poorly settling) suspended solids. These suspended solids consisted predominantly of algae, because these were abundant in the liquid phase. The occurrence of spontaneous flocculation of algae is supported by the following two observations: (1) when exposing a sample of the liquid phase from lanes 2 to 5 in a beaker to sunlight a green precipitate formed within a period of 1 to 2 hours at the bottom of the beaker, indicating that indeed a flocculation induced sedimentation of algae occurred and (2) Chlorophyll a was detected in the bottom sludges of lanes 2 to 5, and this was not the case for the sludge in lane 1, although algae also were clearly present in the liquid phase of this lane.

Even under the unfavourable operational conditions of the total system, considering (1) the short UASB liquid retention time applied in the UASB-reactor and (2) the fact that no excess sludge was discharged from the UASB reactor, the rate of bottom sludge accumulation was very low. On the basis of the experimental data, it was estimated an accumulation rate of 70 mg TS/*l* of anaerobically pre-treated sewage and a volumetric bottom sludge accumulation rate of only 700 ml per m^3 of influent, assuming a sludge concentration of 100 g TS/*l* (the approximate value after 1 year of operation). Adopting a yearly *per capita* sewage contribution of 40 m^3 (equivalent to 110 $l.inh^{-1}.day^{-1}$) the bottom sludge production would amount to 28 $l.inh^{-1}.annum^{-1}$. This value is a factor 2 - 4 smaller than the values reported for raw sewage treatment in anaerobic ponds (Arceivala, 1985 and Gonçalves *et al.*, 2000).

The volumetric accumulation rate depends on the applied liquid retention time in the pond. The volume of pre-treated sewage in PP1, during 1 year of operation at an average liquid retention time of 6 days, was about 2000 m^3 or, in terms of volume per m^3 of pond, equal to 2000/32.5 = 61 m^3 per annum. Consequently a volume of 0.7 *l*/m^{3}x61 m^3= 43 litres bottom sludge will accumulate per m^3 of pond, i.e. the bottom sludge accumulated at a rate of only 4.3% per year of the total pond volume. It is likely there will be no need to desludge the polishing pond during its useful life span of, say, 20 years. And this is for the situation in which discharge of excess sludge from the UASB reactor was omitted. It can be concluded that periodical discharges of excess sludge from the UASB reactor become optional when polishing ponds are used for post- treatment. However in cases where the excess sludge represents an attractive by-product (for example, for sale as organic fertilizer) it is better to apply regular discharges of excess sludge from the UASB-reactor.

The pond performance and its operational stability will not be affected by the introduction of excess sludge from the UASB-reactor. Hence it is possible to operate the UASB reactor without sludge discharge and accumulate the excess sludge in the PP. This could constitute an attractive and simple procedure to reduce operational costs substantially.

Table 6.2. Observed bottom solids accumulation per unit volume of treated sewage and *per capita*.

Parameter	Unit	Lane1	Lane2	Lane3	Lane4	Lane5	Total
Accumulation per m^3 of digested sewage	g/m^3	38.6	7.0	10.8	8.2	6.0	70.5
	m*l*/m^3 (@ 100 g/*l*)	386	70	108	82	60	705
Accumulation *per capita* and per annum	kg.inh^{-1}.annum^{-1}	1.54	0.28	0.43	0.33	0.24	2.82
(@ 40 m^3.inh^{-1}.annum^{-1})	*l*/inh^{-1}annum^{-1} (@100g/*l*)	15.4	2.8	4.3	3.3	2.4	28

The results of the sludge stability test (Table 6.1) show that less than 5% of the organic material was biodegradable. This excellent stability in fact was expected given the fact that they originate from the UASB reactor and the long period of time they are exposed to anaerobic digestion in the pond as bottom sludge.

It can be observed, from the average, for all lanes, of organic solids fraction (amounting to 0.58 of the total solids) and the nitrogen and phosphorus content of the volatile matter that there is a relatively high proportion of organic material and nutrients in the sludge (6.6% and 1.8% for N and P, respectively). These values, which were very similar to those normally found in biological sewage sludge (less than 10% of N and 2.5% of P), together with its excellent stability, make the sludge an attractive organic fertilizer and/or soil conditioner. However, the hygienic quality of the sludge is still poor, because the number of helminth eggs in the bottom sludge was very high with an average value of 651 eggs per gram of dry sludge, which corresponds to 45 eggs per litre of treated influent, about 50% of the number of helminth eggs in the pre-treated sewage. This egg density is higher than some data for sewage treatment plants in the south of Brazil which showed egg density of 100 to 1000 per kg *TS* (Tsutiya, 2001). Even though according to investigations by Gonçalves *et al.* (2000), only a fraction of 0.02 to 0.10 of the eggs in bottom solids are viable (there are no data, from this present study, about the egg viability), the number is still significantly higher than most standards. EPA (1992) uses a standard of 1 viable helminth egg per 4 g of sludge for use in mechanized agriculture, without direct contact. Based on this, and assuming that these standards are realistic, it can be concluded that the use of the bottom sludge in agriculture might only be acceptable after thorough treatment to improve its hygienic quality

6.4 Conclusions

From the experimental investigation, a pilot sewage treatment plant, comprising a UASB reactor as pre-treatment unit and a polishing pond as the post-treatment step, it can be concluded that under tropical conditions:

- unless the excess sludge can be used for agricultural purposes, there is no need for regular discharge of excess sludge from the UASB reactor: the polishing ponds can accommodate excess sludge without affecting their performance or operational stability;
- the rate of solids accumulation in the bottom sludge in polishing ponds is much lower than generally found in anaerobic ponds treating raw sewage, despite the fact that excess sludge was not discharged from the pre-treatment unit;
- the observed bottom sludge accumulation over a period of 1 year, amounted to 70 mg/*l* of influent, which was a factor 2 to 4 times smaller than the value found for anaerobic ponds treating raw sewage;
- under the investigated conditions the bottom sludge was well stabilizing: less than 5% of the VSS was digested during stability test;
- the hygienic quality of the bottom sludge was still rather poor: more than half of the helminth eggs of the UASB effluent fed to the pond persisted in the bottom sludge;
- the organic fraction of the bottom solids as well as the nutrient concentrations were

high (58% volatile matter, 0.066 g*TKN*/g*VS* and 0.018 g*P*/g*VS*), so that the sludge after removal of the pathogens would appear to be useful as an organic fertilizer.

6.5 References

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Chapter 7

Summary and General Conclusions

7.1 Chapters Summary

CHAPTER 2 - Influence of the phase separator design on the performance of the UASB reactor and on excess sludge production. (Proceedings of the 7th Latin American Congress on Anaerobic Digestion, October 2002, Merida, Mexico).

The most typical and essential aspect of the UASB reactor is the phase separator device that avoids the sludge particles from escaping together with the effluent. An efficient phase separator will allow a large sludge mass to be retained in the reactor and this is advantageous for the performance in two ways: (1) the more efficient the sludge retention, the less suspended solids will be present in the effluent and this, in itself, will also increase the BOD_5 and COD removal efficiencies and (2) the average sludge retention time or sludge age increases, so that the hydrolysis process of particulate influent material will be more complete. As a consequence, the reactor produces less sludge but, at the same time, the bacterial sludge mass is increased.

In Chapter 2 it was shown that the efficiency of the phase separator can be increased by having additional parallel plates above the conventional device, thus forming a high rate settler. From settling theory it can be deduced that the efficiency of the plates is increased when: (1) the height of the zone with parallel plates increases, (2) the distance between plates decreases and (3) the angle of the plates decreases. In practice the height is limited by the construction cost of the reactor; the distance between the plates must be sufficient for maintenance and the angle to allow the retained sludge to slide readily back into the digestion zone.

The experimental data was obtained from 2 pilot scale (1500 litres) UASB reactors, fed with raw sewage. One of two reactors was adapted by the inclusion on an additional solid – liquid high rate separator: parallel plates were added to the conventional phase separator. The parallel plates had a depth of 0.35 m, and were placed at a distance of 7 cm apart and at an angle of 45^o.

The experimental data showed that the fundamental operational variable to describe the behaviour of the UASB reactor is, in fact, the sludge age: at any particular temperature, the effluent quality and the quality and quantity of excess sludge will be equal in two UASB reactors if the sludge age in these is the same. Hence the removal efficiency, in principle, is independent of the liquid retention time. The liquid retention time only has importance because it must be sufficient for the reactor to accommodate the sludge mass corresponding to the sludge age.

CHAPTER 3.- Excess sludge discharge frequency for UASB reactors. (Published in 1999 by Elsevier Science Ltd/IAWQ - Water Science Technology, Vol.40, No. 8, pp 211-219).

Like all biological wastewater treatment systems, the UASB reactor produces sludge that accumulates in the reactor. Since the storage capacity is limited, sludge will be washed out together with the effluent once the reactor is full. To avoid this it is necessary that excess sludge be discharged periodically from the reactor before its storage capacity is exhausted. For minimum operational costs, it is preferable to have large discharges with a low frequency. On the other hand the discharges cannot be excessive in order to avoid deterioration of the reactor performance after the discharges.

An experimental investigation was carried out at pilot scale to establish the influence of the magnitude of excess sludge discharge on the performance of UASB reactors, treating municipal sewage. UASB reactors were operated at liquid retention times (R_h) of 4 and 8 hours and excess sludge discharges of varying magnitude were applied. The performance and operational stability of these reactors after the discharges were observed and the sludge accumulation and unintentional discharge (wash-out) in the effluent were determined. During periods of steady state without excess sludge discharge, the net sludge production was determined from the settleable solids in the effluent.

The data show that for discharges of up to 50% of the sludge mass in reactors with R_h = 4 hours and 60% with R_h = 8 hours, the reactor efficiency and operational stability were affected very little and only during the first few days after the discharge. Discharges of up to 80 % of the sludge did not cause instability, but a reduction in the *COD* removal efficiency was observed during 1 to 2 weeks after the discharge. Independent of the magnitude of sludge discharge, the effluent *pH* remained in the narrow range of 6.8 to 7.0 and the ratio between bicarbonate alkalinity, *BAlk* (average 275 mg $CaCO_3/l$) and volatile fat acids, *VFA* concentrations (average of 28 mg *HAc*/*l*) was always very high so that there was never a danger of souring the reactor contents. Based on the obtained experimental data an operational procedure for optimization of excess sludge discharges was developed.

It was believed that along with the favourable natural conditions during the experimental investigation (temperature of 25 ± 2 ^{o}C) the observed reactor stability can be attributed to a large extent to the design of the phase separator used, which is much more efficient than the conventional one. It is possible that results with a conventional UASB reactor would have been different.

CHAPTER 4 - Post treatment of UASB reactor effluent in flow-through polishing ponds. (Published in 2001 by Elsevier Science Ltd/IAWQ - Water Science Technology, Vol.44, No. 4., pp 237-245).

Polishing ponds are used to improve the quality of effluents from efficient anaerobic sewage treatment plants, like UASB reactors, so that the final effluent quality becomes compatible with legally binding standards. The residual organic material and suspended solids concentrations in the digested sewage are reduced, but often the main objective of polishing ponds is to improve the hygienic quality, measured by the concentration of two indicator organisms: helminths eggs (*HE*) and faecal coliforms (*FC*). The *FC* removal normally is the slowest process and for that reason becomes the main design criterion for a polishing pond. By contrast in conventional waste stabilization ponds (WSP's) systems the organic material removal is the governing design parameter.

The feasibility of operating a polishing pond, 0.65 m deep and 50/l length/width ratio, for the post-treatment of UASB effluent was shown in Chapter 4 and the final effluent quality as a function of the liquid retention time and the pond depth is discussed. The results also show that the coliform removal kinetics is a first order process, properly described by the

Wehner and Wilhelm equation for flow through-ponds and by exponential decay for batch ponds. Tracer studies in the flow-through pond to evaluate the hydrodynamic regime showed that there was moderate mixing and a relatively high dead volume fraction, even though the pond had been designed for plug flow.

To maximize the *FC* removal efficiency, one of the four operating polishing pond was constructed with the objective of approaching a plug flow regime. However, the observed efficiency was well below the expected value for all liquid retention times, which was attributable to imperfections of the flow regime. From tracer studies it was established that the dispersion number was in the range of 0.14 to 0.18, which means that moderate mixing occurred, even though the pond was designed to avoid mixing as much as possible. Also the dead volume fraction ranging from 0.12 to 0.18 was quite considerable. Due to these imperfections the required liquid retention time for an effluent to be used for unrestricted irrigation required a retention time of about 10 days. Experiments with batch ponds showed that the decay constant in the pond with a depth of 0.65 m was 2.0 to 2.3 day^{-1}, so that in the case of exponential decay (in an ideal plug flow reactor or a batch reactor) the required time for efficient *FC* removal (99.99%) would be only 4 to 5 days. Thus the actual retention time required in the flow-through pond was twice the value of the minimum retention time. Although the plug flow regime could not be approached in practice, the required retention time of 10 days is still very much shorter than the value used in conventional WSP´s (20 to 30 days).

CHAPTER 5 – Post-treatment of UASB reactor effluent in sequential batch polishing ponds. (Proceedings of the VI Latin-American Workshop and Seminar on Anaerobic Digestion – V.1 pp 360-367 – November, 2000 - Recife-Brazil).

In Chapter 4 it was established that a polishing pond could be used for post treatment of digested sewage to correct the deficiencies in the anaerobic pre-treatment: residual BOD_5 and *TSS* concentration and helminth eggs number were reduced to CEC recommendations (CEC, 1991)[1] and WHO (WHO, 1989)[2] guidelines for unrestricted irrigation values in less then 5 days, but faecal coliform removal took longer (10 days).

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[1] CEC (1991). Council Directive 91/271/CEE of 21 May 1991, concerning urban wastewater treatment. Journal Official of the European Communities, no. L 135. 40-52.

[2] WHO (1989). Health guidelines for the use of waste water in agriculture and aquaculture. Geneva, Switzerland., World Health Organization.: 72.

In Chapter 5 the behaviour of ponds fed with sequential batches of digested sewage was evaluated. In these ponds the expected exponential decay was indeed observed. Two different configurations of sequential batch ponds were tested. The first one was comprised of five parallel batch ponds each with a volume equal to the daily flow. They were fed sequentially and the required effluent quality was obtained after 5 days. The second system had a transfer pond which received the UASB effluent continuously and from which a series of four batch ponds were fed. The two systems had equivalent performances but the second one has practical advantages.

CHAPTER 6 – Sludge accumulation in polishing ponds treating anaerobically digested sewage. (IAW publishing 2001 - Water Science and Technology, Vol 45, no.1 pp 75-81).

When ponds are used for sewage treatment, settleable solids will form a steadily growing bottom sludge layer, which reduces the effective volume. Eventually this sludge must be removed to ensure that the pond maintains the required retention time to keep performing properly. The settleable solids may either be present in the influent or they are formed during the treatment as a result of algae sedimentation.

If raw sewage is treated in a conventional waste stabilization pond system, the rate of bottom sludge accumulation is in the range of 150 to 250 mg/*l*, which settles mainly in the anaerobic pond, thus reducing its effective volume. In order to avoid a reduction of its performance the anaerobic pond needs to be desludged every 2 to 5 years. If pre-treatment in an UASB reactor is applied to treat raw sewage, the polishing pond influent concentration and nature of the settleable solids will differ radically from those in ponds treating raw sewage. The removal efficiency of suspended solids in a well performing UASB system is of the order of 70 to 80 % and this figure is even higher for the settleable solids (80 to 95 %). The escaping settleable solids are mostly sludge particles that could not be retained in the treatment unit. Due to the reduction in the settleable solids concentration, the rate of solids accumulation in the bottom sludge may be expected to be smaller in polishing ponds treating digested sewage than in anaerobic ponds treating raw sewage. Also, the sludge particles are much less subject to anaerobic degradation than the mainly biodegradable suspended solids present in raw sewage, so that odour is likely to be a smaller problem.

An investigation was carried out to evaluate bottom sludge accumulation in a polishing pond used for the post-treatment of UASB effluent. The mass and composition of the bottom sludge formed in a polishing pond was evaluated after the pilot scale pond had been in operation for 1 year and about 60 m^3 of digested sewage had been treated per m^3 of pond. The bottom sludge mass represented a solids accumulation of 70 g/m^3 of digested sewage. About half of these solids were the result of the settling of influent solids in the first part of the pond, while the other half was attributable to settling of algae, formed in the pond. It is concluded that the bottom sludge growth in a polishing pond is so low, that desludging during the useful life span of the pond will most likely not be necessary. This leads to the important conclusion that excess sludge discharge from UASB reactors (a major factor in operational costs) may be omitted, if a polishing pond is used for post treatment.

The bottom sludge had a high volatile solids concentration (58%) and the macronutrient fractions were also high (3.9% *N* and 1.1% *P* of the *TS* mass. The hygienic quality of the bottom sludge was very poor: about half the influent helminth eggs during one year of operation were found in the bottom sludge and the faecal coliform concentration was very high.

7.2 General Conclusions

In tropical regions sewage can be treated in a system composed of a UASB reactor and a polishing pond in series (UASB+PP). In such a system it is possible to produce an effluent that can be used for unrestricted irrigation in accordance with guidelines of the World Health Organization. The experimental investigation has led to the following conclusions:

- the most important parameter that influences the efficiency of a UASB reactor is the sludge age, which in turn is strongly affected by the phase separator design;
- in a UASB reactor equipped with an efficient separator, more than 80% of the *COD* and *TSS* of the sewage influent, can be removed, even if the reactor is operating at a short liquid retention time ($R_h \geq 4$ hours);
- to improve the UASB effluent quality, excess sludge can be discharged periodically . The size of the maximum sludge mass discharge can be determined experimentally, but in general more than 50% of the maxim sludge mass can be discharged without affecting significantly the performance of the UASB reactors;
- even if the excess sludge is not discharged from the UASB reactor, due to its high physical effluent quality (low suspended solids concentration and turbidity), a post-treatment in a polishing ponds can be applied, whose sizes are determined by the need not for organic material stabilization but for an efficient *FC* removal;

- the hydrodynamic regime in flow-through polishing ponds in practice does not approach plug flow: uncontrollable factors like wind, microbial movement and rising gas bubbles induce mixing, which is also apparent from the value of the dispersion number, calculated from retention time distribution profiles. This means that the liquid retention time for *FC* removal in a flow-through polishing pond must be at least twice the theoretical minimum value, observed in ponds operated in batch mode;
- the *FC* removal efficiency was closely predicted by the Wehner and Wilhelm equation for a first order process in partially mixed ponds;
- due to the fact that the required liquid retention time for a particular *FC* removal efficiency is inversely proportional to the depth, the liquid retention time is not suitable as a unequivocal design parameter. By contrast, surface loading rate and the *per capita* area of a polishing pond are more adequate parameters, because the values do not depend on the pond depth, at least for the investigated range from 0.65 to 3 m.

For each of the important parameters to be corrected in PP (*FC*, BOD_5, *TSS* and *HE*) there is a relationship between the removal efficiency and the *per capita* area of the pond. With respect to this dimension parameter, Figure 7.1 shows the *per capita* area necessary to attain a certain standard. For a temperature of 25 ± 2 ^{o}C the following relationships were experimentally established, assuming a *per capita* contribution of 100 *l*/day:

- A_1 = 0.5 m^2/inhab the minimum for the removal of *HE*. The mechanism of *HE* removal is discrete settling. The settling efficiency is independent of depth. The maximum superficial loading rate for *HE* removal is 0.22 m/day;
- A2 = 0.67 m2/inhab the minimum area for a FC removal efficiency of 99.99 % in sequential batch polishing ponds (SBPP). In flow-through ponds with a carefully designed to minimize mixing the required area (A4 = 1.3 m2/inhab) can be expected to be at least double the value for the SBPP;
- A_3 = 0.95 m^2/inhab the minimum area for a $BOD_5 \leq 25$ mg/*l*, determined in filtered samples, which satisfies the standards of the CEC;
- A_5 = 1.4 m^2/inhab the minimum area for *TSS* = 30 mg/*l* for ponds with depth from 1.65 to 3 m deep, a concentration considered to be very low for raw samples due to the presence of algae. In shallow ponds (0.65 m deep) the required area is larger (A_6 = 1.9 m^2/inhab) due to algae concentration.

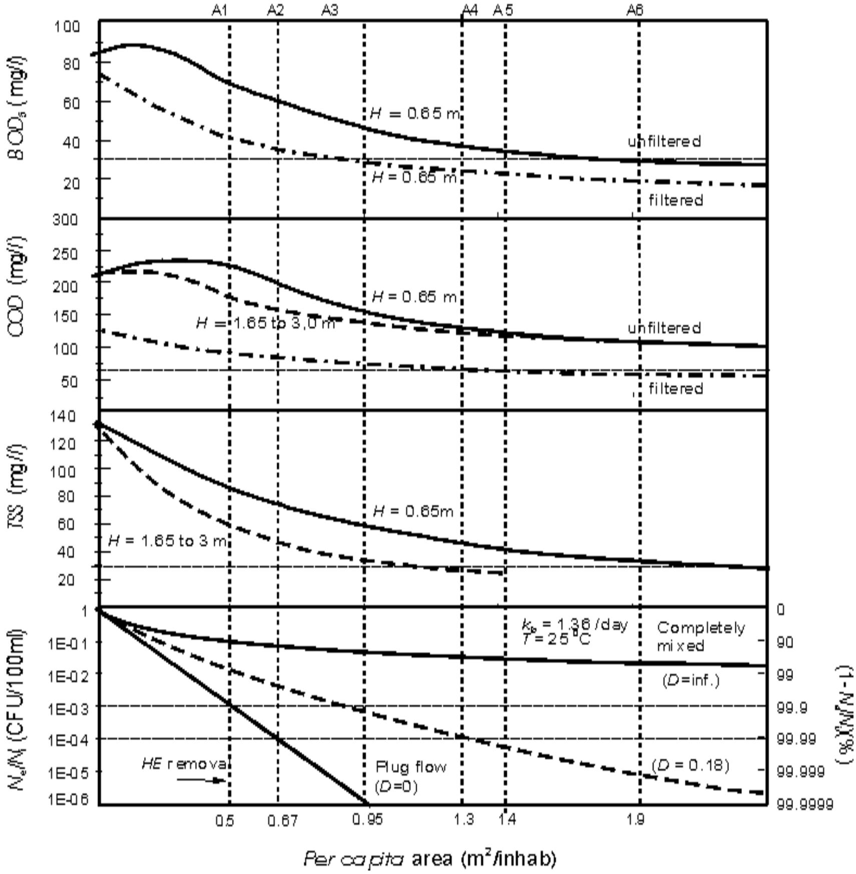


Figure 7.1. *Per capita* area and BOD_5, *TSS, FC* and *HE* relationships for a desired removal, considering a *per capita* contribution of 100*l*.inhab^{-1}.day^{-1} of 25 °C.

Hoofdstuk 7

Samenvatting en Algemene Conclusies

7.1 Samenvatting van de Hoofdstukken

HOOFDSTUK 2 – Invloed van het ontwerp van de fasenscheider op de efficiëntie van de USB reactor en op de slibproductie (Gepresenteerd op het 7^de^ Latijns-Amerikaans congres over anaërobe vergisting, Merida-Mex. 2002).

Het meest typerende en essentiële onderdeel van de UASB reactor is de fasenscheider, die onder meer verhoedt dat slibdeeltjes ontsnappen in het effluent. Met behulp van een efficiënte fasenscheider kan een grote slibmassa in de reactor worden gehouden en dit heeft een dubbel voordeel: (1) hoe beter de slibretentie, des te minder slibdeeltjes zullen in het effluent aanwezig zijn, hetgeen op zichzelf leidt tot een hogere efficiëntie van de BZV_5 en CZV verwijdering en (2) de gemiddelde slibverblijftijd of slibleeftijd neemt toe, zodat de hydrolyse van particulair influent materiaal meer compleet zal zijn. Daardoor wordt er minder slib geproduceerd en ter zelfder tijd wordt de bacteriële slibmassa vergroot.

In Hoofdstuk 2 wordt aangetoond dat de efficiëntie van de conventionele UASB fasenscheider kan worden verhoogd door erboven parallelle platen te plaatsen, zodat in feite een bezinker met hoge snelheid wordt gevormd. Met behulp van de bezinkingstheorie kan worden afgeleid dat de efficiëntie van de platen hoger is naarmate (1) de hoogte van de zone met parallelle platen toeneemt, (2) de afstand tussen de platen kleiner is en (3) de hoek met de horizon vermindert. In de praktijk wordt de hoogte van de zone met platen gelimiteerd door economische overwegingen, terwijl de afstand voldoende moeten zijn voor gemakkelijk onderhoud en de hoek van de platen het mogelijk moet maken dat de bezonken slibdeeltjes vlot terugglijden naar de vergistingzone van de UASB reactor.

Twee UASB reactors, elk met een volume van 1500 litre, werden bedreven met ruw rioolwater om de experimentele gegevens te verkrijgen. Eén van de twee reactoren was uitgerust met een conventionele fasenscheider, terwijl de andere bovendien nog paralelle platen boven de fasenscheider had. De zone met platen was 35 cm hoog en deze waren geplaatst op een afstand van 7 cm en met een hoek van 45^o.

De experimentele data toonden aan dat de slibleeftijd de fundamentele operationele variabele is die de werking van de UASB reactor bepaalt. Bij een bepaalde temperatuur zullen de effluentkwaliteit zowel als de samenstelling en hoeveelheid van het geproduceerde slib in twee UASB reactors gelijk zijn, als de slibleeftijd in deze dezelfde is. Daarom is de efficiëntie van de verwijdering van organisch materiaal in principe onafhankelijk van de vloeistof verblijftijd. Deze heeft alleen belang omdat ze voldoende moet zijn om de slibmassa die overeenkomt met de slibleeftijd in de reactor te kunnen houden.

HOOFDSTUK 3. - Frequentie van het aflaten van spuislib voor UASB reactors. (Gepubliceerd 1999 door Elsevier Science Ltd/IWA - Water Science Technology, Vol.40, No. 8, pp 211-219).

Zoals alle biologische systemen voor afvalwaterbehandeling, produceert ook de UASB reactor slib dat zich ophoopt in de reactor. Aangezien de capaciteit om slib in de reactor te houden beperkt is, zal het nodig zijn om periodiek spuilib af te laten. Om de kosten van deze operatie te minimaliseren verdient het de voorkeur om een grote hoeveelheid slib te spuien met een lage frequentie. Van de andere kant moet de hoeveelheid spuislib natuurlijk niet zo groot zijn dat de werking van de reactor verslechtert, nadat het is afgelaten.

Een experimenteel onderzoek werd uitgevoerd om vast te stellen wat de invloed de hoeveelheid spuislib op de werking van demonstratie UASB reactors, waarin huishoudelijk afvalwater werd behandeld. De reactors werden bedreven met verblijftijden van 4 en 8 uur, wat respectievelijk de minimale en maximale waarden zijn die in de prakrijk gebruikelijk zijn. De efficiëntie en de stabiliteit van de reactoren werden bepaald. Ook werden de accumulatie van slib in de reactoren en de ongewilde lozing van slib in het effluent gemeten. Gedurende periodes met “volle” reactoren werd de slibproductie bepaald met behulp van de concentratie van bezinkbare stoffen in het effluent.

De resultaten tonen aan dat het lozen van 50% van de slibmassa in de reactoren met R_h = 4 uur en van 60% in die met R_h = 8 uur geen merkbaar effect had op de efficiëntie van de reactoren en dat ook de operationele stabiliteit nauwelijks werd beïnvloed en dan alleen nog gedurende de eerste dagen na het spuien van het slib. Zelfs het aflaten van 80 % van de slibmassa veroorzaakte geen instabiliteit, maar de efficiëntie van de CZV verwijdering verminderde enige weken. De pH van het effluent en in de reactoren veranderde nauwelijks en bleef tussen 6,8 en 7,0 onafhankelijk van de grootte van de gespuide slibmassa. De verhouding tussen de bicarbonaat alkaliniteit (gemiddeld 275 mg $CaCO_3/l$) en de concentratie van vluchtige vetzuren (gemiddeld 28 mg HAc/l) was altijd groot, zodat op geen enkel moment het gevaar bestond dat de inhoud van de reactor zou verzuren.

Het is waarschijnlijk dat naast de hoge temperatuur (gemiddeld 25 ± 2 ^oC) de reactor stabiliteit ook werd bevorderd door het ontwerp van de fasenscheider die veel efficiënter was dan de conventionele. Mogelijkerwijs zouden de resultaten met een conventionele UASB reactor anders zijn geweest.

HOOFDSTUK 4 – Nabehandeling van UASB reactor effluent in nabehandelingsvijvers met een continu debiet. (Gepubliceerd in 2001 door Elsevier Science Ltd/IAWQ - Water Science Technology, Vol.44, No. 4., pp 237-245).

Nabehandelingvijvers worden gebruikt om de kwaliteit te verbeteren van het effluent van efficiënte anaërobe behandelingsystemen zoals de UASB reactor, zodat de samenstelling van het uiteindelijke effluent strookt met wettelijke of gewenste normen. De resterende concentratie van organisch materiaal en zwevende stoffen in het vergiste rioolwater is normaliter klein. In het algemeen is het voornaamste doel van de nabehandelingvijvers dan ook het verbeteren van de hygiënische kwaliteit, die bepaald wordt door de concentratie van twee indicator organismen: wormeieren en fecale coliformen. Gewoonlijk is de verwijdering van de coliformen het langzaamste proces, waardoor het de voornaamste grondslag wordt voor het ontwerp van nabehandelingvijvers. Dit staat in tegenstelling met het ontwerp van conventionele stabilisatievijvers, waar het voornaamste criterium (nomen est omen) de verwijdering van het organisch materiaal is.

In dit hoofdstuk wordt aangetoond dat het mogelijk is de nabehandeling van vergist rioolwater uit te voeren in één enkele vijver. De effluentkwaliteit van de vijver als functie van de vloeistof verblijftijd en de diepte wordt besproken. De resultaten in het hoofdstuk tonen ook aan dat de kinethiek van de fecale coliformen verwijdering goed beschreven wordt als een eerst orde proces. Een van de vier vijvers die gebruikt werden in het onderzoek was zodanig gebouwd, dat verwacht werd dat het hydrodynamische regime dat van een propstromer zou benaderen. Hiervoor werd een ondiepe vijver (0,60 m) gebruikt met een grote lengte/breedte verhouding (50/l). De waargenomen FC verwijdering bleek echter veel kleiner te zijn dan de verwachte efficiëntie, wat was toe te schrijven aan imperfecties in de vloeistofstroming: Studies met tracer toonden aan dat het dispersie getal in de vijver een waarde van 0,14 tot 0,18 had, wat betekent dat gematigde menging optrad. Ook de fractie dood volume (stagnante zones in de vijver) was aanzienlijk: 0,12 tot 0,18. Door deze imperfecties was de benodigde verblijftijd voor de productie van een effluent dat voor irrigatie zonder restricties kan worden gebruikt ongeveer 10 dagen. Batch experimenten toonden evenwel aan dat de vervalconstante van de FC onder de omstandigheden in de vijver ongeveer 2,0 tot 2,3 per dag was, zodat bij exponentiële vermindering van de FC concentratie (zoals die in een ideale propstromer of een batch vijver zou plaatsvinden) slechts 4 tot 5 dagen nodig zouden zijn om de gewenste hygiënische kwaliteit te bereiken. De werkelijke benodigde verblijftijd in de vijver was dus het dubbele van het theoretische minimum. Hoewel het niet mogelijk bleek het regime van een ideale propstromer te benaderen, was de benodigde verblijftijd van 10 dagen toch veel korter dan de waarde die normaliter in conventionele stabilisatie vijvers wordt gebruikt (20 tot 30 dagen). De resultaten toonden ook aan dat de verwijdering van wormeieren al compleet was na R_h = 3 dagen.

HOOFDSTUK 5 – Nabehandeling van UASB reactor effluent in sequentiële batch vijvers (Gepresenteerd op het 6^de^ Latijns-Amerikaans congres over anaërobe vergisting – Vol 1 pp 360-367 – November, 2000 - Recife-Brazil).

In Hoofdstuk 4 werd vastgesteld dat een nabehandelingvijver kan worden gebruikt om de hygiënische kwaliteit van UASB effluent te verbeteren en tezelfdertijd ook de overblijvende BZV_5 en TSS concentraties te verlagen, maar de verblijftijd voor FC verwijdering (10 d) was twee maal langer dan het theoretische minimum. Daarom werd het gedrag van vijvers bestudeerd waarin deze werden gevoed met opeenvolgende ladingen (batches) UASB effluent. Immers in dergelijke vijvers zal de FC concentraties exponentieel afnemen en de tijd die nodig is om irrigatie water te bereiden zal minimaal zijn. De batch vijvers vertoonden inderdaad de verwachte exponentiële afname van de FC concentratie. Twee verschillende configuraties werden getest: de eerst had vijf vijvers (elk met een volume gelijk aan het dagelijks debiet) die op opeenvolgende dagen werden gevoed en waarin na vijf dagen de gewenste effluent kwaliteit werd bereikt. De tweede had een overslag vijver waarin het UASB effluent stroomde en van waaruit vier parallelle batch vijvers werden gevoed. De twee systemen hadden een vergelijkbaar rendement, maar de tweede configuratie heeft praktische voordelen.

HOOFDSTUK 6 – Slib accumulatie in nabehandelingvijvers. (IAW publishing 2001 - Water Science and Technology, Vol 45, no.1 pp 75-81).

Wanneer vijvers worden gebruikt voor de behandeling van huishoudelijk afvalwater, bouwt zich een sliblaag op de bodem, die bestaat uit de bezinkbare deeltjes. Daardoor wordt de vloeistof verblijftijd in deze vijvers geleidelijk korter. Na verloop van tijd moet dit bodemslib worden verwijderd om te garanderen dat de verblijftijd voldoende is voor de gewenste efficiëntie. De bezinkbare deeltjes kunnen aanwezig zijn in het influent (in het bijzonder als er geen slib wordt gespuid van de voorafgaande UASB reactor) of ze kunnen worden gevormd door de flokkulatie van algen.

Wanneer ruw huishoudelijk afvalwater wordt behandeld in een conventioneel systeem van stabilisatievijvers, is de te verwachten accumulatie van slib op de bodem van de vijver van de orde van 150 tot 250 mg/l, die voor het grootste gedeelte in de anaërobe vijver zal worden afgezet. Daardoor is het nodig dat het bodemslib in zo’n vijver elke 2 tot 5 jaar moet worden verwijderd, anders wordt de verblijftijd te kort. Wanneer efficiënte voorbehandeling in een UASB reactor wordt toegepast, zal het influent van de nabehandelingvijver een geheel andere concentratie en samenstelling hebben en daardoor zal ook het bodemslib kwantitatief en kwalitatief anders zijn.

Een onderzoek werd uitgevoerd om de hoeveelheid en de samenstelling van het bodemslib in een nabehandelingvijver te bepalen. De vijver werd gedurende een jaar bedreven met een verblijftijd tussen 5 en 7,5 dagen zodat ongeveer 60 m^3 anaëroob vergist afvalwater werd nabehandeld per m^3 vijver. De massa van het bodemslib vertegenwoordigde een vaste stof accumulation van 70 g per m^3 behandeld afvalwater. Ongeveer de helft van deze vaste stof was het gevolg van de bezinking van UASB slibdeeltjes, terwijl de andere helft het resultaat was van de bezinking van geflokkuleerde algen die in de vijver gevormd werden. Daaruit kon worden geconcludeerd dat de opeenhoping van bodemslib in een nabehandelingvijver zo langzaam geschiedt dat het waarschijnlijk niet nodig zal zijn om het slib te verwijderen gedurende de tijdsduur dat de vijver in werking is (b.v. 20 jaar). Dit leidt tot de belangrijke conclusie dat het spuien van slib van een UASB reactor niet nodig is als een vijver wordt gebruikt voor nabehandeling en er geen zinvol gebruik van het spuislib voorhanden is.

Het bodemslib had een hoog gehalte aan organische stoffen (58%) en de fracties macronutriënten waren ook aanzienlijk (3,9% *N* en 1,1% *P* van de *TS* massa). Het bodemslib was stabiel en kon direct worden gedroogd zonder problemen. De hygiënische kwaliteit van het slib was zeer laag: ongeveer de helft van de wormeieren in het influent gedurende het jaar dat de vijver in werking was, werden in het bodemslib aangetroffen en ook de FC concentratie was zeer hoog. Dit is een aspect dat in acht moet worden genomen wanneer men overweegt om het bodemslib als meststof in de landbouw te gebruiken.

7.2 Algemene Conclusies.

In tropische gebieden kan de behandeling van huishoudelijk afvalwater worden uitgevoerd in een systeem dat is samengesteld uit een UASB reactor en een nahehandelingvijver. Met zo’n systeem kan men een effluent kwaliteit produceren die overeenkomt met de normen van de wereldgezondheidsorganisatie (WHO) en dat gebruikt kan worden voor irrigatie zonder restricties. Het experimenteel onderzoek heeft geleid tot de volgende conclusies:

- de meest belangrijke grootheid die de efficiëntie van een UASB reactor beïnvloed is de slibleeftijd, die op zijn beurt voornamelijk wordt bepaald door het ontwerp van de fasenscheider;
- in een UASB reactor met een efficiënte fasenscheider kan meer dan 80 % van het influent *CZV* en *TSS* worden verwijderd, zelfs wanneer de vloeistofverblijftijd kort is ($R_h \geq 4$ uur);
- om de effluentkwaliteit te bevorderen kan periodiek spuislib worden geloosd. De grootte van de massa slib die kan worden geloosd kan experimenteel worden vastgesteld, maar in het algemeen kan meer dan 50% worden gespuid zonder merkbare gevolgen voor de het rendement of de stabiliteit van de UASB reactor;
- zelfs wanneer geen spuislib wordt geloosd is de kwaliteit van het UASB effluent zodanig dat de nabehandeling kan worden uitgevoerd in een enkele vijver. Door de lage restconcentraties van organisch materiaal en zwevende stoffen is de verlaging van deze gemakkelijk, zodat de limiterende grondslag voor ontwerp van de nabehandelingvijvers de verwijdering van pathogene organismen, en meer specifiek die van Fecale Coliformen wordt;
- het blijkt niet mogelijk om een propstroom regime te benaderen in een nabehandelingvijver met een continu influent debiet. Oncontroleerbare factoren zoals wind, thermische menging, opstijgende gasbelletjes en beweging van de aanwezige

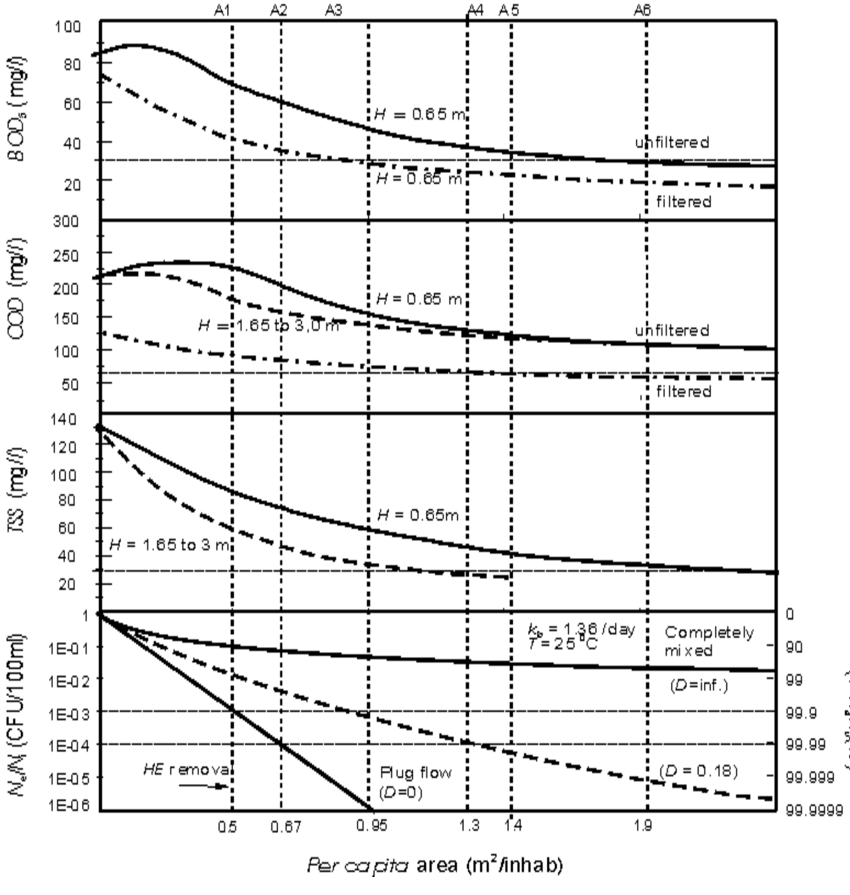
organismen leiden tot aanzienlijke menging, zoals ook blijkt uit de waarden van het dispersiegetal dat kan worden berekend uit de perfielen van de verblijftijd distributie in zulke vijvers. Daardoor was onder de omstandigheden van het experimenteel onderzoek de benodigde verblijftijd om de FC efficiënt te verwijderen het dubbele van de minimale tijdsduur die kan worden waargenomen in een vijver die batchgewijs wordt bedreven;

- de experimentele efficiëntie van de *FC* verwijdering was in goede overeenkomst met de theoretische waarde die met de Wehner en Wilhelm vergelijking kan worden berekend voor een eerste orde proces in een reactor met partiële menging;
- er werd vastgesteld dat de FC vervalconstante omgekeerd evenredig is met de diepte van de vijver. Daardoor is de gemiddelde verblijftijd in een vijver geen ondubbelzinnige grootheid, wanneer men een vijver wil ontwerpen met een bepaalde FC verwijderingsrendement. Daarentegen zijn de oppervlaktebelasting en en het *per capita* oppervlak van een nabehandelingvijver onafhankelijk van de diepte (tenminste voor de onderzochte waarden van 0,65 tot 3 m)en daardoor meer zinvol als ontwerp grootheden.

Voor elk van de belangrijke effluent parameters die gecorrigeerd dienen te worden in een nabehandelingvijver (*FC*, *BZV*$_5$, *TSS* and *wormeieren*) bestaat er een betrekking tussen het verwijderingrendement en het *per capita* vijveroppervlak. In Figure 7.1 is de effluentkwaliteit uitgezet als functie van het *per capita* vijveroppervlak. Voor de omstandigheden gedurende het experimenteel onderzoek (temperatuur van 25 ^{o}C, *per capita* afvalwaterproductie van 100 *l*/d) kunnen de volgende belangrijke waarden worden aangeduid:

- A_1 = 0,5 m^2/inw is het minimum voor de verwijdering van wormeieren. Het mechanisme van de verwijdering is discrete bezinking. Het bezinkingsrendement is onafhankelijk van de diepte. De maximale oppervlaktbelasting is 0,22 m/dag;
- A_2 = 0,65 m^2/inw is het minimale oppervlak om 99,99% van de *FC* te verwijderen batchgewijze bedreven vijvers. In vijvers met een continu debiet treedt menging op en daardoor is het minimale oppervlak op z’n minst twee maar zo groot als in batchgewijs bedreven vijvers. Een minimaal oppervlak van A_4 = 1,3 m^2/inh moet worden verwacht voor dit type vijver;
- A_3 = 0,95 m^2/inw is het minimaal oppervlak om een gefiltreerde *BZV*$_5$ $\leq$ 25 mg/*l*,te verkrijgen wat voldoet aan de normen van de Raad van de Europese Commissie;

- A_5 = 1,4 m^2/inw is het minimaal oppervlak voor *TSS* = 30 mg/*l* in vijvers met een diepte van 1,65 tot 3 m. Deze zeer lage concentratie kan in ondiepe vijvers (0,65 m) slechts worden bereikt als een groter oppervlak wordt gebruikt (A_6 = 1,9), wat te wijten is aan de aanwezigheid van algen.



Figuur 7.1. *BOD*$_5$, *TSS*, *FC* en *HE* concentratie als functie van het per capita oppervlak van nabehandelingvijvers (een afvalwater productie van 100l.inw^{-1}.dag^{-1} en een m^2/inw temperatuur van 25 ^{o}C zijn verondersteld.)

Capítulo 7

Sumário e Conclusões Gerais

7.1 Sumário dos Capítulos

CAPÍTULO 2 – Influence of the phase separator design on the performance of the UASB reactor and on excess sludge production. (Anais do 7º. Simpósio Latino-Americano sobre Digestão Anaeróbia, Mérida, Yucatán, México, outubro de 2002).

Reatores UASB se caracterizam principalmente pelo separador de fases, nele contido, que evita que partículas de lodo escapem junto com o efluente. Um separador de fases eficiente possibilita a retenção de uma grande massa de lodo no reator, melhorando o desempenho deste de duas maneiras: (1) quanto mais eficiente for a retenção de lodo menos sólidos suspensos estarão presentes no efluente e, conseqüentemente, maior será a eficiência de DBO_5 e DQO e (2) o tempo médio de retenção de lodo ou idade do lodo aumenta, de modo que o processo de hidrólise da matéria particulada afluente será mais completo e o lodo acumulado mais estável. Como consequência o reator produz menos lodo, aumentando, no entanto, a massa de lodo ativo (bactérias).

Mostra-se, no Capítulo 2, que a eficiência do separador de fases pode ser melhorada pela aralelas acima do separador convencional, configurando-se, assim, um separador de alta taxa. Da teoria de sedimentação pode ser deduzida que a eficiência das placas paralelas aumenta quando: (1) a altura da zona com placas paralelas aumenta, (2) a

Os dados experimentais foram obtidos da operação de 2 reatores UASB construídos em escala piloto (1500 litros) e alimentados com esgoto bruto municipal. Um dos dois reatores foi adaptado para ter um separador sólido-líquido de alta taxa, através da colocação de 11 placas paralelas em adição ao separador de fases convencional. As placas paralelas com altura de 0,35 m e distanciadas de 7cm uma das outras, foram colocadas formando um ângulo de 45º com a base do reator.

Dos dados experimentais, pode-se concluir que a variável operacional que descreve o comportamento de reatores UASB é de fato, a idade de lodo: para uma determinada temperatura, a qualidade do efluente e a qualidade e quantidade de lodo de excesso seriam iguais nos dois reatores se a idade de lodo em ambos fosse a mesma. Desta forma, a eficiência de remoção, em princípio, é independente do tempo de retenção do líquido. O líquido tem importância apenas porque define o volume do reator que, por sua vez, tem que ser suficiente para acomodar a massa de lodo correspondente a idade de lodo.

CHAPTER 2. - Excess sludge discharge frequency for UASB reactors. (Publicado em 1999 por Elsevier Science Ltd/IAWQ - Water Science Technology, Vol.40, No. 8, pp 211-219).

Como todos os sistemas de tratamento de águas residuárias o reator UASB produz lodo que nele se acumula. Uma vez atingida a capacidade de armazenamento o lodo produzido excedente é descarregado junto com o efluente. Para evitar que o lodo seja carreado junto com o efluente, torna-se necessário descarregar periodicamente o lodo de excesso, antes que a capacidade de armazenamento do reator seja atingida. Para se ter custos operacionais mínimos é preferível que sejam feitas grandes descargas e com menor frequência. Por outro lado, a magnitude das descargas não pode ser excessiva, evitando, dessa maneira, que o reator apresente um baixo desempenho após a descarga.

Foi realizada uma investigação experimental em escala piloto para estabelecer a influência da magnitude de descargas de lodo de excesso no desempenho de reatores UASB, tratando esgoto municipal. Foram operados dois reatores UASB com tempo de retenção do líquido (R_h) de 4 e 8 horas e aplicadas descargas de lodo de excesso de várias magnitudes. Durante toda a investigação foram observados o desempenho e a estabilidade operacional dos reatores e determinadas a massa de lodo acumulada e carreada (wash-out) com o efluente. Durante os períodos de estado estacionário, estabelecidos entre duas descargas consecutivas, a produção de lodo era determinada a partir dos sólidos sedimentáveis carreados com o efluente.

Os dados mostram que para descargas de até 50% da massa de lodo acumulada no reator operado com R_h = 4 horas e 60% da do reator operado com R_h = 8 horas, a eficiência e estabilidade operacional foram pouco afetadas e apenas durante os primeiros dias após as descargas. Descargas de até 80% da massa de lodo causaram uma diminui

DQO durante as primeiras duas semanas após as descargas sem, contudo, causar instabilidade operacional: independente da magnitude da descarga de lodo, o pH efluente se manteve na faixa de 6,8 a 7,0 e a razão entre a alcalinidade de bicarbonato (valor médio de 275 mg $CaCO_3/l$) e a concentração de ácidos graxos voláteis (média de 28 mg*HAc*/*l*) foi sempre muito alta e, desta forma, nunca houve perigo de azedamento do conteúdo do reator. Com base nos dados obtidos, foi desenvolvido um procedimento operacional para otimização das descargas de lodo de excesso.

Acredita-se que, juntamente com as condições naturais favoráveis durante a investigação experimental (temperatura de 25 ± 2 ^{o}C), a estabilidade operacional do reator pode ser atribuída à configuração, na forma de braço, do separador das fases sólido-líquido, que se mostrou muito mais eficiente que o convencional. É possível que, tratando-se de um UASB com separador convencional, os resultados obtidos fossem diferentes.

CHAPTER 4 - Post treatment of UASB reactor effluent in flow-through polishing ponds. (Publicado em 2001 por Elsevier Science Ltd/IAWQ - Water Science Technology, Vol.44, No. 4., pp 237-245).

Lagoas de polimento podem ser usadas para melhorar a qualidade de efluentes de reatores anaeróbios eficientes, tal qual o reator UASB, adequando-os aos padrões legais desejados. Embora a concentração de matéria orgânica e de sólidos suspensos no esgoto digerido seja reduzida, as lagoas de polimento geralmente têm, como principal finalidade, que promover um efluente de boa qualidade higiênica, medida por dois organismos indicadores: ovos de helmintos (*OH*) e coliformes fecais (*CF*). A remoção de CF normalmente é um processo lento e, por essa razão, torna-se o principal critério de dimensionamento dessas lagoas. Em enção hidráulica de 10 dias foi ainda muito menor que aquele necessário em sistemas de lagoas de estabilização (20 a 30 dias).

CHAPTER 5 – Post-treatment of UASB reactor effluent in sequential batch polishing ponds. (Anais do VI Seminário Latino-americano sobre Digestão Anaeróbia – Vol. 1 pp 360-367 – Novembro, 2000 - Recife- Brasil).

No Capítulo 4 ficou estabelecido que lagoas de polimento podem ser usadas para o pós-tratamento de esgoto digerido para corrigir as deficiências do processo anaeróbio: a concentração residual de DBO_5 e SST e o número de ovos de helmintos foram reduzidos para os valores recomendados, para irrigação irrestrita, pela Comunidade Européia (CEC, 1991)[1] e pela Organização Mundial de Saúde (WHO, 1989)[2], em menos que 5 dias, sendo necessário, contudo, para os CF um tempo de detenção hidráulica bem mais longo (10 dias).

No Capítulo 5 se faz uma avaliação do comportamento de lagoas de polimento alimentadas em bateladas seqüenciais (LPBS) com esgoto digerido em um reator UASB. Da operação destas lagoas foi observado um decaimento exponencial dos coliformes fecais. Duas diferentes configurações foram investigadas: a 1ª configuração consistia de 5 LPBS operadas em paralelo, cada uma com volume igual ao fluxo diário de esgoto digerido. As lagoas eram alimentadas em bateladas, permanecendo o esgoto digerido em cada uma delas por 5 dias, quando, então, se obtinha o efluente final desejado. Na 2ª configuração 4 LPBS eram alimentadas da mesma forma que as da 1ª configuração contudo diretamente de uma lagoa de transbordo ou de transferência, a qual recebia continuamente o efluente de um reator UASB. As duas configurações apresentaram desempenhos equivalentes, mas a 2ª configuração apresentou vantagens de ordem prática.

CHAPTER 6 – Sludge accumulation in polishing ponds treating anaerobically digested sewage. (Publicado pela IAW em 2001 - Water Science and Technology, Vol 45, no.1 pp 75-81).

Quando lagoas são usadas para o tratamento de esgoto, a sedimentação dos sólidos suspensos irá formar, gradativamente, uma camada de lodo, reduzindo o volume efetivo da lagoa. Eventualmente esse lodo pode ser removido para assegurar o tempo de detenção hidráulica requerido e, assim, manter o desempenho desejado. Os sólidos sedimentáveis podem tanto estar presentes no afluente como ser formados durante o tratamento na lagoa, como resultado da floculação de algas.

Se esgoto bruto é tratado em sistemas de lagoas de estabilização, a taxa de acumulação de lodo é da ordem de 150 a 250 mg/*l*, sedimentando, o lodo, principalmente na lagoa

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[1] CEC (1991). Council Directive 91/271/CEE of 21 May 1991, concerning urban wastewater treatment. Journal Official of the European Communities, no. L 135. 40-52.

[2] WHO (1989). Health guidelines for the use of waste water in agriculture and aquaculture. Geneva, Switzerland. World Health Organisation.**:** 72.

anaeróbia, reduzindo, assim, o seu volume efetivo. Para evitar que o desempenho da lagoa anaeróbia seja reduzido é necessário que sejam aplicadas descargas de lodo a cada 2 a 5 anos. Se o esgoto bruto é pré-tratado em um reator UASB, a lagoa de polimento receberá um afluente com concentração e natureza de sólidos sedimentáveis completamente diferentes daquele em lagoas tratando esgoto bruto. A eficiência de remoção de sólidos suspensos em um reator UASB de bom desempenho é na ordem de 70 a 80% , sendo maior para sólidos decantáveis (80 a 95%). Os sólidos suspensos carreados pelo efluente do reator são, na sua maioria, partículas grandes que podem ser retidas na unidade de pós-tratamento. Devido à redução da concentração de sólidos suspensos no reator UASB, pode ser esperado que a taxa de acumulação de lodo em lagoas de polimento tratando esgoto digerido seja bem menor que aquela em lagoas anaeróbia tratando esgoto bruto. Também as partículas de lodo digerido que chegam à lagoa de polimento já estão bem estabilizadas (baixa fração de sólidos suspensos biodegradáveis), de modo que não haverá problemas de maus odores.

Uma investigação foi realizada para avaliar a acumulação de lodo em uma lagoa de polimento usada para o pós-tratamento do efluente de um reator UASB. A massa e a composição do lodo acumulado na lagoa foram avaliadas após a lagoa ter sido operada por 1 ano e ter tratado cerca de 60 m^3 de efluente digerido por m^3 da lagoa. A massa de lodo acumulada representava uma concentração de $70 g/m^3$ de esgoto digerido. Cerca da metade desse lodo era resultado da sedimentação dos sólidos presentes no afluente e se encontrava na primeira parte da lagoa, enquanto que a outra metade foi atribuída a sedimentação de algas formadas na lagoa. Constatou-se que a massa de lodo no fundo da lagoa de polimento aumentava de forma tão lenta que provavelmente não seria necessária descarga de lodo ao longo de sua vida útil. Dessa constatação se chega a uma conclusão importante, por ser um alto fator de custo operacional: descargas de lodo de reatores UASB podem ser omitidas quando a unidade de pós-tratamento é uma lagoa de polimento.

O lodo acumulado no fundo da lagoa tinha uma alta concentração de sólidos voláteis (58%) e as frações de macro nutrientes também eram altas (3,9 % N e 1,1 % P por massa de ST). O lodo acumulado era estável e podia ser drenado diretamente para leitos de secagem sem problemas. O lodo apresentou uma pobre qualidade higiênica: cerca da metade dos ovos de helmintos presentes no afluente de 1 ano de operação foram encontrados no lodo acumulado, sendo a concentração de coliformes fecais também muito alta.

7.2 Conclusões Gerais

Em regiões tropicais esgotos municipais podem ser tratados num sistema formado por um reator UASB seguido por lagoa de polimento (UASB+LP). Em tais sistemas é possível produzir um efluente que pode ser usado na irrigação irrestrita por estar de acordo com as recomendações da Organização Mundial de Saúde (OMS). A investigação experimental desenvolvida com o sistema UASB+LP levou às seguintes conclusões:

- o parâmetro mais importante que influencia o desempenho dos reatores UASB é a idade de lodo, a qual, por sua vez é fortemente determinada pelo configuração do separador de fases;
- um reator UASB provido de um eficiente separador de fases pode remover eficientemente mais de 80% da *DQO* e *SST* do esgoto bruto afluente, operando com um baixo tempo de detenção hidráulica ($R_h \geq 4$ hours);
- para melhorar a qualidade do efluente de reatores UASB, o lodo de excesso deve ser descarregado, periodicamente, do reator. A massa máxima de lodo a ser descarregada pode ser determinada experimentalmente, mas, em geral, mais que 50% da massa máxima de lodo acumulada podem ser descarregados sem que o desempenho do reator seja afetado significativamente;
- mesmo não sendo descarregado lodo de excesso do reator UASB, por apresentar boa qualidade (baixa concentração de sólidos suspensos e baixa turbidez) o efluente pode ser aplicado diretamente numa lagoa de polimento, cujas dimensões não são mais determinadas pela necessidade de se estabilizar a matéria orgânica e sim de se obter *CF*;
- o regime de escoamento em lagoas de polimento de fluxo contínuo, na prática, não pode ser aproximado ao de fluxo pistão: fatores não controláveis como vento, movimento de organismos e desprendimento de bolhas de gás induzem considerável mistura, determinada pelo número de dispersão, o qual é calculado a partir de perfis da distribuição do tempo de detenção hidráulica. Como conseqüência, o tempo de detenção do líquido para remover *CF* numa lagoa de polimento de fluxo contínuo deve ser pelo menos duas vezes o valor mínimo teórico observado em lagoas de fluxo pistão ou alimentadas em bateladas;
- a eficiência de remoção de *CF* observada foi bem próxima a predita pela equação de Wehner e Wilhelm para processos de 1^a ordem em lagoas com mistura parcial;
- devido ao fato de ser o tempo de detenção hidráulica, requerido para uma determinada *CF*, inversamente proporcional a profundidade da lagoa, o tempo de detenção do líquido é um parâmetro de dimensionamento não adequado

como também ambíguo. Em contraste, a taxa de carga superficial e a área *per capita* da lagoa de polimento são parâmetros mais adequados, porque não dependem da profundidade da lagoa, pelo menos para a faixa investigada de 0,65 a 3 m.

Para cada um dos principais parâmetros que devem ter seus valores corrigidos na lagoa de polimento (*CF*, *DBO*$_5$, *SST* e ovos de helmintos (*OH*)), existe uma relação entre a eficiência de remoção e a área *per capita* da lagoa, conforme mostra a Figura 7.1. Para uma temperatura de 25 ^{o}C as seguintes relações foram experimentalmente estabelecidas, considerando-se uma contribuição *per capita* de 100 *l*/dia:

- A_1 = 0,5 m^2/hab, a área *per capita* mínima para remover completamente ovos de helmintos. O mecanismo de remoção é a sedimentação discreta que independe da profundidade da lagoa. A carga superficial máxima para remoção de ovos de

- A_2 = 0,67 m^2/hab, que corresponde à área mínima para uma eficiência de remoção 99,99% em lagoas de polimento alimentadas em bateladas seqüenciais (LPBS). Em lagoas de fluxo contínuo, criteriosamente dimensionadas para minimizar mistura, a
  $_4$ = 1,3 m^2/hab) deve ser no mínimo o dobro da área necessária para LPBS;
- A_3 = 0,95 m^2/hab, a área mínima para se obter um efluente com *DBO*$_5$ $\leq$ 25 mg/*l*, determinada em amostras filtradas, atendendo assim aos padrões do *Council European Communities* (CEC, 1991);
- A_5 = 1,4 m^2/hab é a área mínima para se obter um valor de *SST* = 30 mg/*l* para lagoas com profundidade de 1,65 até 3m, uma concentração que pode ser considerada muito baixa para amostras brutas devido à presença de algas. Em lagoas rasas (0,65m de profundidade) a área requerida é maior (A_6 = 1,9 m^2/hab) devido à alta concentração de algas decorrente da alta taxa de atividade fotossintetizante.

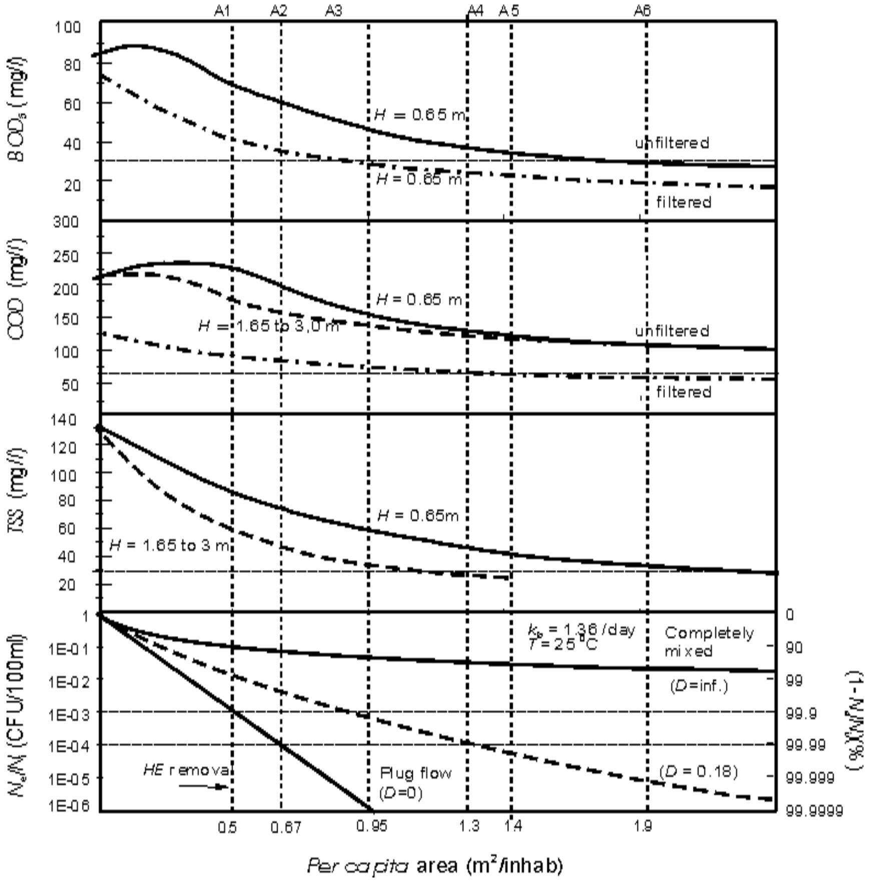

Figura 7.1.: Relação entre a área *per capita* e a remoção desejada de DBO_5, *SST, CF* e *OH*, considerando uma contribuição *per capita* de 100*l*.hab^{-1}.dia^{-1} e a 25°C.

Acknowlegements

According to the National Service for Information, an entity of the Planning Ministery, in 1996 in Brazil only 35% of the population was served by a sewerage network and only 10% of the collected sewage was treated. From the five administrative regions in Brazil (South, South-East, North, North-East and Centre-West) the Northern and North Eastern regions had the lowest coverage of sewage collection with only 2% e 15% respectively of the population. With the aim of finding technically and financially feasible solutions for these problems in Brazil, a national research program of basic sanitation (PROSAB) was created in 1996. Within the program four networks of cooperative institutions were formed to develop research into four topics: water treatment and supply, sewage treatment, solid waste treatment and sludge treatment. The cooperative networks of each theme have been receiving financial support from FINEP (entity for financial support of studies and projects, linked to the Ministery of Science and Technology), CNPq (the National Council for Research) and CEF (the National Housing Bank).

The Federal University of Campina Grande (UFCG) has been a member of PROSAB, and since 1997 is developing investigations into sewage treatment, and more specifically research projects concerning the operation and design of UASB reactors and post-treatment of anaerobic effluents using polishing ponds among other systems. The development of this thesis was possible due to the financial support of PROSAB and the contributions of the members of the cooperative network during the periodic discussions and for which the author wishes to express her gratitude.

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List of Symbols

q	temperature dependency coefficient (-)t
ν	cinematic viscosity of water (m^2/s).
Balk	bicarbonate alkalinity (mg as $CaCO_3/l$).
BOD_5	biodegradable oxygen demand, 5 days (mg/*l*).
CFU	colony forming units (-).
COD	chemical oxygen demand (mg/*l*).
D	dispersion number (-).
f_{ana}	*COD* influent fraction converted into bacterial mass (-).
f_{be}	*ratio COD* biodegradable/ total *COD* in the effluent (-).
f_{cat}	*COD* influent fraction catabolised (-).
f_{cv}	conversion factor between *COD* and *VSS* = 1,5 *gCOD/gVSS*.
f_d	dead volume fraction (-).
f_{us}	non biodegradable and soluble *COD* fraction in the effluent (-).
k_d	value of the decay constant at a reference temperature, normally 20 ^oC (day^{-1}).
k_{dH}	decay constant for any pond depth: $0,65<H<3$ m (day^{-1}).
k_{dT}	value of the decay constant for any temperature (day^{-1});
ME_t,	daily total excess sludge discharge or daily total sludge production (kg).

mS_{be}	fraction of the biodegradable *COD* influent in the effluent (-).
mS_{d}	fraction of influent *COD* digested in the reactor (-).
mS_{e}	fraction of influent *COD* present as organic material in the settled effluent (-).
mS_{x}	fraction of influent *COD* transformed into volatile sludge (-).
$mS_{x,min}$	minimum sludge production (kg)
mS_{xa}	fraction of influent *COD* converted into bacterial mass (-).
mS_{xi}	influent COD fraction discharged as inert material in the excess sludge (-).
MX_{t}	total sludge mass in the reactor (kg).
MX_{v}	volatile sludge mass in the reactor (kg).
N	*FC* concentration (no. *CFU*/100 ml).
N_{e}	effluent *FC* concentration (CFU/100ml).
N_{i}	influent *FC* concentration (CFU/100ml).
Q_{inhab}	per capita flow (m^{3}.inhab^{-1}.day^{-1}).
r_{d}	decay (death) rate of the *FC* (no./m^{3}.day).
R_{h}	liquid retention time (hour, day).
R_{s}	sludge age (day).
s_{c}	critical settling velocity for floc retention by conventional phase separator (m/s).
$s'c$	critical settling rate (m/s).
S_{re}	raw effluent *COD* concentration (mg/*l*).
SS	suspended solids (mg/*l*).
S_{se}	settled effluent *COD* concentration (mg/*l*).
TSS	total suspended solids (mg/*l*).
S_{ti}	influent *COD* concentration (mg/*l*).
$TAlk$	total alkalinity (as mg CaCO3/*l*).
VFA	volatile fatty acids concentration (mg *HAc*/*l*).
VSS	volatile suspended solids (mg/*l*).

X_t	total sludge concentration (kg *TSS*/m^3).
X_v	total volatile sludge concentration (kg VSS/m^3).
X_{ve}	concentration of volatile sludge in the effluent (kg VSS/m^3).
Y	yield coefficient or bacterial mass production per unit metabolized organic material (g *SS*/g*COD*).
Y_{ap}	apparent yield coefficient (g *TSS*/g*COD*).
Y_t	specific total sludge production (g *TSS*/g*COD*).
Y_v	specific volatile sludge production (g *VSS*/g*COD*).

Author’s Biography

The author of this thesis graduated in Civil Engineering at the Federal University of Paraiba where she also obtained her M.Sc. certificate. Since 1977 she has been a member of the Civil Department of the Federal University of Paraiba where she has been teaching Water and Wastewater Treatment, supervising student’s research work and conducting research into wastewater treatment.