UNIVERSITY OF SÃO PAULO
SÃO CARLOS SCHOOL OF ENGINEERING
DEPARTMENT OF HYDRAULICS AND SANITATION ENGINEERING

BRUNO PESSOTTO

**DESIGN AND PERFORMANCE EVALUATION OF AN ANAEROBIC
SLUDGE BLANKET REACTOR WITH MECHANICAL MIXING SYSTEM
AND A HIGH-RATE SETTLER TREATING DOMESTIC WASTEWATER**

DOCTORAL SUPERVISOR: JOSÉ ROBERTO CAMPOS

SÃO CARLOS

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SLUDGE BLANKET REACTOR WITH MECHANICAL MIXING SYSTEM
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in partial fulfilment of the
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DOCTORAL SUPERVISOR: JOSÉ ROBERTO CAMPOS

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São Carlos

2015

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Comissão Julgadora:

Resultado:

Prof. Titular **José Roberto Campos (Orientador)**
(Escola de Engenharia de São Carlos/EESC)

Aprovado

Prof. Dr. **Adrianus Cornelius Van Haandel**
(Universidade Federal de Campina Grande/UFCG)

Aprovado

Prof. Titular **Roque Passos Piveli**
(Escola Politécnica-EP/USP)

Aprovado

Prof. Titular **Pedro Além Sobrinho**
(Escola Politécnica-EP/USP)

APROVADO

Dr. **Cristiano Luchesi Niciura**
(ENGECORPS)

Aprovado

Coordenadora do Programa de Pós-Graduação em Engenharia Hidráulica e Saneamento:

Profa. Associada **Maria Bernadete A. Varesche Silva**

Presidente da Comissão de Pós-Graduação:

Prof. Associado **Paulo César Lima Segantine**

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This achievement is dedicated to

Beatriz Schultz,

*my mother, my fortress and the most
precious treasure I own.*

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“We are governed, our minds are molded, our tastes formed, our ideas suggested, largely by men we have never heard of. [...] In almost every act of our daily lives, whether in the sphere of politics or business, in our social conduct or our ethical thinking, we are dominated by the relatively small number of persons [...] who understand the mental processes and social patterns of the masses. It is they who pull the wires which control the public mind.”

- Edward Louis Bernays

COMPOSITION BY
BRUNO PESSOTTO

“Whatever an education is, it should make you a unique individual, not a conformist; it should furnish you with an original spirit with which to tackle the big challenges; it should allow you to find values which will be your road map through life; it should make you spiritually rich, a person who loves whatever you are doing, wherever you are, whomever you are with; it should teach you what is important, how to live and how to die.”

— *John Taylor Gatto*

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ABSTRACT

PESSOTTO, B. (2015). **Design and performance evaluation of an anaerobic sludge blanket reactor with mechanical mixing system and a high-rate settler treating domestic wastewater.** Thesis (Doctoral). São Carlos School of Engineering, University of São Paulo, São Carlos, 172 p.

The UASB (Upflow Anaerobic Sludge Blanket) reactor, developed by Prof. Gatze Lettinga and his colleagues in the Netherlands more than thirty years ago, is successfully employed in many sewage treatment plants in countries with tropical and subtropical climate. However, are still reported serious operational issues related to the three-phase separators and problematic aspects involving the complexity of its influent distribution system. The focus of this research was to simplify the introduction and dispersion of the sewage in the reactor as well as enhancing the sludge retention, by replacing the characteristic piping system for a mechanical mixing apparatus and substituting the (conventional) three-phase separators by a high-rate settler, respectively. In essence, it was proposed a system partitioned into two distinct regions: one for reaction (i.e. anaerobic digestion of organic matter), in which the sludge blanket was mechanically mixed through the action of a rotating impeller, and another for the removal of suspended solids by the use of a lamella plate clarifier; a sludge recirculation system was also implanted, to pump the settled biomass back to the digestion compartment. The research was subdivided into three main phases: in the first two phases, bench-scale experiments were performed in order to obtain basic data for the design of the mechanical mixing and sedimentation devices, while the third phase consisted of the application of these technologies in a pilot-scale system with total volume of 20.8 m³ (comprised of a 17.0 m³ digestion chamber and a 3.8 m³ compartment for solids removal), as well as its performance assessment. During the development of the research, domestic sewage was applied to feed the experimental system, which was located at the School of Engineering of São Carlos (Campus I) in São Carlos (São Paulo State, Brazil). Throughout the period of effective operation, which lasted 130 days, four distinct operating modes were tested (characterized by intermittent mixing and varying hydraulic loading rates, ranging between 27.6 and 66.5 m³.d⁻¹) and the system reached maximum removal efficiencies of BOD, COD and TSS at rates of 82%, 72% and 83%, respectively. The high-rate settler presented an impressive efficiency removing suspended solids (rates up to 98%). The research showed that the new “Anaerobic Activated Sludge Blanket” system (A²SB) can provide satisfactory performance even operating without the three-phase separator and influent distribution system.

Keywords: A²SB; anaerobic activated sludge; sludge blanket; mechanical mixing; high-rate sedimentation; UASB; wastewater treatment; domestic sewage.

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RESUMO

PESSOTTO, B. (2015). **Concepção e avaliação do desempenho de um reator anaeróbio de manta de lodo dotado de agitador mecânico e decantador de alta taxa tratando esgoto sanitário**. Tese (Doutorado). Escola de Engenharia de São Carlos, Universidade de São Paulo, São Carlos, 172 p.

O reator UASB, desenvolvido pelo Prof. Gatzke Lettinga e colaboradores na Holanda há mais de trinta anos, é utilizado com sucesso em muitas estações de tratamento de esgoto sanitário em países de clima tropical e subtropical. Entretanto, ainda são relatadas sérias questões operacionais relacionadas aos separadores trifásicos e aspectos problemáticos envolvendo a complexidade de seu sistema de distribuição do esgoto afluyente. O foco da presente pesquisa foi o de simplificar a distribuição e dispersão do esgoto no reator, assim como aprimorar a retenção de lodo, mediante substituição do sistema convencional de tubos por misturador mecânico e do (tradicional) separador trifásico por decantador de alta taxa, respectivamente. Em essência, propôs-se reator dividido em duas regiões distintas: uma para reação (i.e. digestão anaeróbia da matéria orgânica), na qual a manta de lodo foi agitada mecanicamente através do movimento rotacional de um impelidor, e outra, para remoção de sólidos suspensos mediante uso de decantador com placas paralelas; um sistema de recirculação do lodo também foi implantado, para bombear a biomassa sedimentada até o compartimento de digestão. A pesquisa foi subdividida em três fases: nas duas primeiras, experimentos em escala de bancada foram realizados visando obtenção de dados básicos para dimensionamento dos sistemas de agitação e de decantação, enquanto que a terceira fase consistiu na aplicação dessas tecnologias em sistema de tratamento, em escala piloto, com volume útil total de 20,8 m³ (composto por uma câmara de digestão de 17,0 m³ e um compartimento de 3,8 m³ para remoção de sólidos), assim como na análise de seu desempenho. Durante a pesquisa, esgoto sanitário foi utilizado para alimentar o sistema experimental, instalado na Escola de Engenharia de São Carlos. Ao longo do período de operação, que durou 130 dias, quatro condições operacionais foram testadas (caracterizadas pela agitação intermitente e diferentes taxas de carregamento hidráulico, que variaram entre 27,6 e 66,5 m³.d⁻¹) e o sistema alcançou eficiências de remoção máxima para DBO, DQO e SST de 82%, 72% e 83%, respectivamente. O decantador de alta taxa apresentou surpreendente eficiência de remoção de sólidos suspensos (taxas de até 98%). A pesquisa demonstrou que o novo sistema A²SB (*Anaerobic Activated Sludge Blanket*) pode apresentar desempenhos satisfatórios, mesmo operando sem separador trifásico e sistema de distribuição do esgoto afluyente.

Palavras-chave: A²SB; lodo ativado anaeróbio; manta de lodo; mistura mecânica; sedimentação

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LIST OF ABBREVIATIONS AND ACRONYMS

ABNT	Brazilian Technical Standards Association	
ACP	Anaerobic contact process	
APHA	American Public Health Association	
BOD	Biochemical oxygen demand	mg.L ⁻¹
C	Continuous mixing	
CBL	Calculator Based Laboratory	
CFD	Computational fluid dynamics	
COD	Chemical oxygen demand	mg.L ⁻¹
CSTR	Continuous stirred tank reactor	
EESC	São Carlos School of Engineering	
FS	Fixed solids	mg.L ⁻¹
FSS	Fixed suspended solids	mg.L ⁻¹
GHG	Greenhouse gas	
GLS	Gas-liquid-solid	
HDPE	High-density polyethylene	
HLR	Hydraulic loading rate	m ³ .m ⁻³ .d ⁻¹
I	Intermittent mixing	
LATAR	Laboratory of Advanced Treatment and Reuse	
NBR	Brazilian Technical Standard	
OLR	Organic loading rate	kgCOD.m ⁻³ .d ⁻¹
PVC	Polyvinyl chloride	
RTD	Residence time distribution	
SHS	Department of Hydraulics and Sanitation Engineering	
SMA	Specific methanogenic activity	gCOD _{rem} .(gVSS) ⁻¹
TKN	Total Kjeldahl nitrogen	mg.L ⁻¹
TS	Total solids	mg.L ⁻¹
TSS	Total suspended solids	mg.L ⁻¹
TVS	Total volatile solids	mg.L ⁻¹
UAE	United Arab Emirates	
UASB	Upflow anaerobic sludge blanket	
USP	University of São Paulo	
VS	Volatile solids	mg.L ⁻¹
VSS	Volatile suspended solids	mg.L ⁻¹
WEF	Water Environment Federation	
WWTP	Wastewater treatment plant	

LIST OF NOMENCLATURE AND SYMBOLS

A_c	Cross-sectional area of each channel within the clarifier	m^2
A_e	Area effectively available for the liquid flow within the clarifier	m^2
A_{gr}	Total granule surface area	m^2
A_p	Projected, or elemental, plate area	m^2
A_R	Surface area of the plate	m^2
A_V	Cross-sectional area of the tank	m^2
b	Width of the tank	m
C	Distance between the bottom of the tank and the impeller	m
$C_{ap,bd}$	Concentration of unidentified products in the bed	$kg.m^{-3}$
$C_{ap,bl}$	Concentration of unidentified products in the blanket	$kg.m^{-3}$
C_{CH_4}	Concentration of methane (in the biogas)	$kg.m^{-3}$
$C_{g,i}$	Concentration of gas component	$kg.m^{-3}$
CH_4	Methane	
$C_{in,i}$	Concentration of influent component	$kg.m^{-3}$
$C_{out,i}$	Concentration of effluent component	$kg.m^{-3}$
$C_{sol,bd}$	Sludge concentration in the bed	$kg.m^{-3}$
$C_{sol,bl}$	Sludge concentration in the blanket	$kg.m^{-3}$
$C_{sub,0}$	Initial concentration of substrate	$kg.m^{-3}$
$C_{sub,bd}$	Concentration of substrate in the bed	$kg.m^{-3}$
$C_{sub,bl}$	Concentration of substrate in the blanket	$kg.m^{-3}$
$C_{sub,gr}$	Substrate concentration with the granules	$kg.m^{-3}$
$C_{sub,st}$	Concentration of substrate in the settler	$kg.m^{-3}$
$C_{tr,0}$	Influent tracer concentration	$kg.m^{-3}$
$C_{tr,bd}$	Tracer concentration in the sludge bed	$kg.m^{-3}$
$C_{tr,bl}$	Tracer concentration in the sludge blanket	$kg.m^{-3}$
CV	Height of the liquid above the impeller	m
D_{imp}	Impeller diameter	m
d_p	Particle size	m
D_{st}	Dispersion coefficient of liquid within the settler	$mm^2.s^{-1}$
D_T	Diameter of the tank	m
E_i	Concentration of inactive biomass	$kg.m^{-3}$

LIST OF NOMENCLATURE AND SYMBOLS

$G ; G_{med}$	Mean velocity gradient	s^{-1}
H	Effective height of the tank	m
H_2S	Hydrogen sulphide	
HLR_c	Clarifier hydraulic loading rate	$m^3.m^{-2}.d^{-1}$
K	Impeller power number	
k_{ap}	Constant of unidentified products	
K_d	Decay constant	h^{-1}
$K_{mt,gr}$	Mass transfer coefficient for the granules	$mm.s^{-1}$
L	Length of the tank	m
L_h	Hydraulic loading rate	$m^3.m^{-3}.d^{-1}$
L_o	Organic loading rate	$kgCOD.m^{-3}.d^{-1}$
L_{st}	Length of the settler	m
n	Impeller rotational velocity	rpm
n'	Velocity at the impeller extremity	$m.s^{-1}$
N_2	Nitrogen (gas)	
$NaCl$	Sodium chloride	
$NaOH$	Sodium hydroxide	
n_p	Quantity of plates	unit
P	Power applied in the liquid mass	W
P_{imp}	Power drawn by the impeller	W
$p_{g,i}$	Pressure of gas	bar
P_M	Motor power	W
P_{max}	Power drawn by the impeller (maximum value)	W
P_{med}	Power drawn by the impeller (mean value)	W
Q	Flow rate	$m^3.d^{-1}$
Q_0	Influent flow rate	$m^3.h^{-1}$
Q_{bd}	Sludge bed inflow rate	$m^3.h^{-1}$
q_{bd}	Specific substrate conversion rate in the bed	s^{-1}
Q_{bl}	Sludge blanket inflow rate	$m^3.h^{-1}$
q_{bl}	Specific substrate conversion rate in the blanket	s^{-1}
Q_{bl-bd}	Flow from sludge blanket to bed	$m^3.h^{-1}$

LIST OF NOMENCLATURE AND SYMBOLS

Q_{bp}	By-pass flow rate	$m^3.h^{-1}$
Q_c	Flow rate in each channel, formed between clarifier plates	$m^3.h^{-1}$
Q_F	Outflow rate	$m^3.h^{-1}$
q_{gr}	Substrate conversion rate within the granule	$kg.m^{-3}.s^{-1}$
Q_{rct}	Reaction zone inflow rate	$m^3.h^{-1}$
$r_{ap,bd}$	Conversion rate of unidentified products in the bed	$kg.m^{-3}.s^{-1}$
$r_{ap,bl}$	Conversion rate of unidentified products in the blanket	$kg.m^{-3}.s^{-1}$
$r_{CH4,bd}$	Methane production in the bed	$kg.m^{-3}.s^{-1}$
$r_{CH4,bl}$	Methane production in the blanket	$kg.m^{-3}.s^{-1}$
Re	Reynolds number	
R_H	Equivalent hydraulic diameter	m
$r_{sub,bd}$	Substrate conversion rate in the sludge bed	$kg.m^{-3}.s^{-1}$
$r_{sub,bl}$	Substrate conversion rate in the sludge blanket	$kg.m^{-3}.s^{-1}$
S_0	Influent COD	$kg.m^{-3}$
SF_{bl}	Short-circuiting flow to the blanket (as a fraction of the Q_0)	%
SF_{st}	Short-circuiting flow to the settler (as a fraction of the Q_0)	%
u_{st}	Flow velocity in the settler	$mm.s^{-1}$
V	Effective volume of the tank	m^3
$V ; V_R$	Reactor total volume	m^3
v_0	Sedimentation velocity	$m.h^{-1}$
v_1	Horizontal velocity of the fluid flow	$m.h^{-1}$
V_{bd}	Sludge bed volume	m^3
V_{bl}	Sludge blanket volume	m^3
v_C	Circulation velocity of the fluid	$m.s^{-1}$
V_d	Dead volume	m^3
v_H	Hazen velocity	$m.d^{-1}$
v_{ij}	Rate coefficient for component i on process j	$kg.m^{-3}$
v_p	Velocity of the liquid flow within the clarifier plates	$m.h^{-1}$
V_{rct}	Volume of the reaction zone	m^3
V_{st}	Volume of the settler	m^3
v_{up}	Upflow velocity	$m.h^{-1}$

LIST OF NOMENCLATURE AND SYMBOLS

X	Concentration of solids in the fluid	$\text{kg solids.}(\text{kg liquid})^{-1}$
$X_{\text{in},i}$	Concentration of influent particulate component	kg.m^{-3}
$X_{\text{out},i}$	Concentration of effluent particulate component	kg.m^{-3}
Y	Yield	
Z_T	Effective height of the tank	m
α	Horizontal inclination angle	degrees
θ_c	Cell residence time	d
θ_h	Hydraulic retention time	h
μ	Dynamic viscosity	N.m.s^{-1}
μ_L	Dynamic viscosity of the liquid	Pa.s
ν	Kinematic viscosity	$\text{m}^2.\text{s}^{-1}$
ξ_{sub}	Substrate conversion rate	
ρ	Specific mass	kg.m^{-3}
ρ_L	Specific mass of the liquid	kg.m^{-3}
ρ_s	Specific mass of the solid	kg.m^{-3}
Øg	Biogas flow rate	$\text{m}^3.\text{h}^{-1}$
Øg_{bd}	Biogas flow rate in the bed	$\text{m}^3.\text{h}^{-1}$
Øg_{bl}	Biogas flow rate in the blanket	$\text{m}^3.\text{h}^{-1}$
$\text{Ø}_{\text{sub}}^{\text{m}}$	Specific conversion of substrate	$\text{kg.m}^{-3}.\text{s}^{-1}$

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INTRODUCTION

1

Among all available technologies applied for domestic sewage treatment, the UASB reactor (acronym for its original name, “Upflow Anaerobic Sludge Blanket”) was, over the last two decades, the alternative that achieved more prominence amid Brazilian professionals from the academic community and other specialists working in the sanitation field. According to van Lier et al. (2010), this technology became popular in hot climate countries after the disclosure of promising results obtained in pioneering installations, built in Colombia, Brazil and India.

Recent investigations expose the predominance of inferior results presented by “contemporaneous” full-scale reactors, in comparison to the first reports (exhibited more than 20 years ago). The satisfactory operational control and the consequent substantiation of high-level performance of the UASB system — in terms of organic matter and total suspended solids (TSS) removal rates, principally — are extremely dependent on its proper design and construction. Despite the notoriety achieved, the disposition and maintenance of the elements that constitute its internal arrangement represent relatively high costs, and also imply difficulties regarding its construction, operation and maintenance (adapted from MIKI, 2010; VAN LIER et al., 2010; HEFFERNAN et al., 2011; VAN HAANDEL & VAN DER LUBBE, 2012).

Even after long period since its original disclosure (LETTINGA et al., 1980) and the execution of several subsequent studies devoted to the various aspects involved in its operation, the original configuration of the UASB reactor remains virtually unaltered^[1]. A bibliometric

¹ In his recently published book, “My anaerobic sustainability story” (LETTINGA, 2014), Gatze Lettinga — the leader of the research team that developed the UASB reactor at Wageningen University (Netherlands) — stated that the unit was initially conceived to meet the needs of the industrial pollution sector, not the public sanitation (or “municipal”) segment. The emergence and dissemination of alternative designs to better attend the domestic wastewater treatment systems (presented in Section 3.1.4.2, Chapter 3) spread in the 1990s, after the publication of works such as “Anaerobic Sewage Treatment” by van Haandel and Lettinga (1994), but did not result in significant conceptual modifications.

analysis performed by Zhang et al. (2014) identified 2,363 UASB-related outputs published in 220 journals over the last three decades (from 1983 to 2012). Scientific articles on this technology demonstrated a noteworthy quantitative increase over this period, as shown in Figure 1.1. Unfortunately, according to Mahmoud et al. (2003), while the biological processes related to the use of this technology are widely investigated, the prevalent physicochemical phenomena (and hydrodynamics) are rarely studied.

Figure 1.1 – Publications with ‘UASB’ as the searching keyword over the period between 1983 and 2012 (ZHANG et al., 2014).

Noyola et al. (2012) estimated that about 17% of Latin American municipal WWTPs use UASB units in their systems; in Brazil, specifically, the percentage would reach 30% (Figure 1.2).

Figure 1.2 – Percentage distribution of treatment technologies in Latin America and Brazil (NOYOLA et al., 2012).

O’Flaherty, Collins and Mahony (2006) stated that anaerobic reactors are usually designed using empirical criteria, which can lead to over-dimensioning of the unit volume and inferior treatment; Mostly of these systems, according to the referred authors, are operated as “black boxes”.

The UASB concept is composed of diverse elements, constituents of distinct systems that impose a high degree of complexity to the project and construction stages; the increase in the level of difficulty is directly proportional to the increment of its dimensions. The UASB reactor, unlike what is commonly disseminated in the literature, require multi-directional efforts for its efficient control, described as follows (adapted from MIKI, 2010; VAN LIER et al., 2010, VAN HAANDEL & VAN DER LUBBE, 2012):

- **Liquid phase:** – Homogeneous distribution of the influent sewage over the bottom of the unit (e.g. precise control of valves, possible need for clearing the weirs, channels, pipes, orifices etc.);
– Regulation of the effluent outlet devices (e.g. levelling of the weirs etc.).
- **Solid phase:** – Monitoring the biomass expansion (i.e. analysis of the TSS profile that establishes over the height of the reactor) and harmonious sludge withdrawal (i.e. organized valve manoeuvring etc.);
– Awareness to the scum accumulation, including the portion located under the three-phase separators (i.e. planning and implementation of procedures for removal);
- **Gaseous phase:** – Control of the undesired gaseous emissions by maintaining the biogas conveyance systems, designed to collect, transport and treat the biogas, in order to ensure its utilization and to avoid problems associated with corrosion and odours in the surrounding area of the wastewater treatment plant (WWTP).

Amidst its inherent advantages, the modern reactors (e.g. UASB, fluidized bed reactor etc.) can admit higher organic loading rates (OLRs) than those tolerated by aerobic systems, with significant abatement of organic material in reactors operated with a hydraulic retention time of just a few hours. Despite this important benefit, the overall efficiency of the technologies based on the anaerobic digestion — with regard to sewage treatment — is still insufficient to produce effluent capable of meeting the discharge standards, especially the parameters like COD and TSS (adapted from VAN HAANDEL & VAN DER LUBBE, 2012). Another important

issue related to the UASB reactor is assigned to the reaction zone (i.e. volume occupied by the active biomass in which the digestion actually occurs), generally limited to about 50% of its total volume (adapted from MAHMOUD, 2008). According to Foresti et al. (2006), research and practical expertise did not translate into improvements regarding reactor design.

These discussed findings evince the existence of some unresolved questions regarding the use of the UASB reactor for municipal sewage treatment plants, as well as the need to pursue solutions that promote cost reduction (on implementation and operation) and that convert anaerobic systems into more efficient alternatives.

In order to circumvent some of the limitations and problematic aspects observed in many existing “municipal” UASB installations, a new technology based on anaerobic digestion and high-rate sedimentation is proposed in this work. The configuration of this novel system, denominated Anaerobic Activated Sludge Blanket (A²SB), consists of an automated mechanical agitation apparatus^[1], for improved dispersion of substrate and controlled sludge blanket expansion, and a lamella settler, for enhanced removal of suspended solids. Some of the mechanisms and operational parameters that affect the physical and biological interactions within this new system were investigated. It is proposed, therefore, significant modifications in the original configuration of the UASB concept, with the following premises:

- Increase of the volumetric capacity of its reaction zone, by artificially stimulating the contact between the biomass and the influent substrate and the sludge blanket expansion through the action of a rotating impeller;
- Improve the performance in the suspended solids removal rate, through the use of high-rate sedimentation device;
- Eliminate the ‘three-phase separator’, enabling the reduction of the problems occasioned by the accumulation scum and the simplification of the biogas collection in the upper section of the reaction zone;
- Enable potential reduction of the construction costs, considering that for UASB reactors, the inflow distribution and gas collection systems assume relevant parcels in the total expenditure for implementation.

¹ The author encourages the reader not to associate the use of a mechanical equipment to a technological retrogression, considering that — depending on the imposed operating conditions — anaerobic systems may be considered, at least partially, self-sufficient technologies in terms of energy consumption.

Although, in the past, many configurations have been proposed, few are those that resemble the system developed during the present research. Nähle (1991) presented installations designed by renowned companies (i.e. Sulzer, Degrémont and Purac) that were based on the “anaerobic contact process”, originally conceived by Schroepfer et al. (1955)^[1]; despite having similar characteristics to the reactor developed in this work, such technologies made use of intense agitation for complete mixing inside the tanks, contrary to the new system, which prioritizes the stratification of biomass in accordance with the sludge blanket concept (traditionally associated with the UASB). Halalsheh et al. (2010) evaluated the application of laminar settlers positioned in the interior of UASB reactors^[2], placed under and above the three phase separators; in the study reported herein, the group of researchers decided to install the settler in a “separated” compartment, due to the high possibility of performance impairment of such device caused by the formation and movement of gas bubbles in the reaction zone.

The research was performed at the experimental WWTP located at the Campus I of the University of São Paulo, in São Carlos (São Paulo State, Brazil). A novel anaerobic reactor was designed to fit an existing installation: a concrete tank with total volume of 20.8 m³, in which a “conventional” UASB unit was previously set. After the construction of the new system, its performance was evaluated, as well as many aspects associated with its operation.

¹ Technology described in Section 3.1.3.

² A brief description of this work is presented in Table 3.7 (Section 3.1.4.5.).

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OBJECTIVES

2

The research was based on the assumption that a new wastewater treatment system, based on the anaerobic process and designed without an influent distribution system and devoid of three-phase separators, can provide similar or even superior performance to the UASB reactor.

The prime objective of this research was to propose a new treatment system, the A²SB (Anaerobic Activated Sludge Blanket) — consisted of two distinct regions: one for reaction (i.e. anaerobic reactor for the digestion of the influent organic matter) and one for removal of solids contained in the reactor effluent —, as well as its implementation (in pilot scale), operation and performance evaluation, particularly in terms of organic matter and suspended solids removal rates.

In order to fulfil the scope of the proposed work, the following specific objectives were achieved:

- Study and assessment of the effects of the mechanical mixing on the anaerobic reactor performance;
- Analyse the benefits attributed to the application of a high-rate sedimentation unit downstream of the reaction zone, especially in terms of removal efficiency of suspended solids;
- Evaluate some of the characteristics of the anaerobic sludge contained in the reactor.

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LITERATURE REVIEW

3

3.1 ANAEROBIC SYSTEMS FOR WASTEWATER TREATMENT

3.1.1 GENERALITIES

Although it is evident the importance of the fundamentals of anaerobic digestion for the development of this work, the theme will not be emphasized in this literature review; the subject is extensively discussed in the literature (VAN HAANDEL & LETTINGA, 1994; SPEECE, 2008; VAN LIER, MAHMOUD & ZEEMAN, 2008; METCALF & EDDY/AECOM, 2014).

The anaerobic treatment of wastewater and sludge is a natural process in which a variety of species — belonging to two different biological kingdoms, i.e. *Bacteria* and *Archaea* — is responsible for the conversion of organic waste into biogas and sludge. No electron acceptor (e.g. oxygen or nitrate) needs to be present or added for the process to work, as organic matter itself or the carbon dioxide resulting from its destruction serves this need. As a result, limitations associated with the oxygen mass transfer are not involved (adapted from McCARTY, 2001).

According to McCarty (1981), the formation of methane in anaerobic process has been known since the eighteenth century, and the involvement of micro-organisms became evident from the nineteenth century. After much time has elapsed, the anaerobic treatment applications have increased significantly as well as the knowledge of chemistry and microbiology associated with it.

The performance of the technologies based on anaerobic digestion has greatly improved during the last decades: while “classical” reactors (e.g. septic tanks and anaerobic ponds) present

low removal efficiencies of organic matter and suspended solids, operating under relatively high hydraulic retention times, the modern reactors obtain better performance and require considerably lower volumes. Two initiatives have made this improvement possible: (i) installation of solid-liquid separators to retain the sludge in the process (ensuring the presence of a large microbial community) and (ii) development of devices that enabled the intense contact between the substrate and the biomass. Despite advances, the implementation of these new reactors in the treatment of domestic sewage has virtually been restricted to countries with warm climate (adapted from VAN HAANDEL & VAN DER LUBBE, 2012).

With the increasing importance assigned to the anaerobic systems, it is appropriate to cite important human contributions to the understanding and advancement of this technology as well as its applications.

3.1.2 FIRST EXPERIENCES

It is believed that the first full-scale application for anaerobic treatment of wastewater was installed in the French town of Vesoul, by Jean-Louis Mouras in the 1860s. He built a tank, originally named “*Fosse Mouras: la vidangeuse automatique*” (or “Mouras’ automatic scavenger”, translated to English), in which domestic sewage from single family homes was discharged before overflowing into the local drainage system (Figure 3.1). Among its features were (adapted from DUNBAR, 1908; RIBEIRO, 1911):

- The extremities of the house connections and the outlet pipes (leading into the street drain) were always placed beneath the surface of the liquid, preventing the gases – generated during the digestion – from reaching the dwellings or the drains;
- A device made from metallic wire (similar to a cage), attached to a chain, was provided for the retention of the larger solids.

A schematic illustration showing the details of the elements that composed the apparatus created by Mouras is presented in Figure 3.2; Both illustrations (Figures 3.1 and 3.2) were inspired by the drawings presented by Dunbar (1908) and Khan (2007).

Figure 3.1 – ‘Fosse Mouras’: conceptual drawing representing its installation in France.

Figure 3.2 – ‘Fosse Mouras’: schematic illustration of the original version and its main components.

Between 1881 and 1882, the French journal *Cosmos les Mondes* published two articles on the Mouras’ tank, which was optimistically classified by an interviewee (Mr. Allain Targe) as “the most simple, the most beautiful, and perhaps, the grandest of modern inventions” and “a complete solution of the problem which for centuries had been an insolent menace hurled in the face of all humanity” (US CIRCUIT COURT, 1905). Abbé Moigno ^[1,2], author of both texts, described the unit as follows (RIDEAL, 1906):

“This mysterious contrivance, which has been used for twenty years, consists of a closed vault with a water seal, which rapidly transforms all the excrementitious matter which it receives into a homogeneous fluid, only slightly turbid, and holding all the solid matters in suspension in the form of scarcely-visible filaments. The vault is self-emptying^[3] and continuous in its working, and the escaping liquid, while it contains all the organic and inorganic elements of the faeces, is almost devoid of smell, and can be received into watering - carts for horticultural purposes, or may pass away into the sewer for use in irrigation.”

¹ MOIGNO, A.F. (1881). Hygiène domestique et industrielle, *Cosmos les Mondes*, 29 Dec 1881, p. 622.

² MOIGNO, A.F. (1882). Hygiène domestique et industrielle, *Cosmos les Mondes*, 21 Jan 1881, p. 97.

³ The statement attributed to Moigno can be considered wrong, since it was evident that the accumulated sludge in any wastewater treatment system must be periodically removed (before its useful volume is completely exhausted by the deposition of solids); this action does not occur “automatically”.

According to Dunbar (1908), the introduction of this technology in Paris was promptly supported by Moigno, but results were not satisfactory. In the city of Bordeaux it was stipulated by the “city engineer” that these units should consist of two compartments and contain a gas release device (Figure 3.3), since, contrary to the expectations of the inventor, the accumulated gas volume occasionally reached pressure able to lower the level the liquid and thus expelling solids contained inside the unit.

Figure 3.3 – ‘Fosse Mouras’: modification imposed in Bordeaux (adapted from DUNBAR 1908).

The development of the first hybrid system for wastewater treatment has been credited to William D. Scott-Moncrieff (Ashted, England), which built in 1891, according to his own words, a “cultivation filter bed”. It was a tank in which a filtering medium composed of successive layers of flint, coke and gravel was installed. Inside the unit, domestic sewage produced by 10 people was subjected to upward flow and degraded anaerobically; the effluent was then collected and conducted — by a piping system — to a device consisting of ‘oscillating trays’, in which part of the remaining organic matter was removed by aerobic process (Figure 3.4; illustration inspired by the drawings presented by its own inventor — SCOTT-MONCRIEFF, 1899). The development of organic films on the surface of the stones was reported, but unfortunately, at that time, the knowledge about the processes involved was minimal. The system was patented and, for the first time, benefits related to the retention of solids present in the sewage were identified, with the objective of promoting “hydrolytic or bacteriological” action for conversion of the organic matter into gases and soluble compounds (adapted from HOUSTON, 1894; SCOTT-MONCRIEFF, 1899; McCARTY, 2001; FANG, 2010; LETTINGA, 2010).

Figure 3.4 – The hybrid system developed by Scott-Moncrieff.

In 1895, in the city of Exeter (England), Donald Cameron built and patented the “septic tank”, with capacity to treat approx. 230 m³ of sewage per day (Figure 3.5)^[1]. According to Baker (1904), this technology was the first which design, construction and operation were based on scientific knowledge. The unit consisted of an anaerobic digestion zone preceded by a grit removal chamber. Buswell and Neave (1930) reported the basic principles inherent to the operation of the unit as follows:

¹ Illustration inspired by the drawings presented by Rideal (1906) and Khan (2007).

Figure 3.5 – Artist impression of the original Septic Tank: designed and patented by Donald Cameron.

“The septic tank is essentially a covered chamber through which sewage flows at a sufficiently reduced velocity to permit sedimentation of suspended solids. The solids then undergo biochemical degradation with the production of an inoffensive humus-like product, which is removed from the tank at suitable intervals”.

The Exeter treatment plant attracted many sanitary engineers around the world and Cameron even erected an observation chamber in the tank so visitors could climb down through an inspection chamber and see the digestion process in action from behind windows (Figure 3.6)^[1].

Figure 3.6 – Inspection chamber with windows installed inside the Septic Tank.

Due to the success of the unit, the sewage of the entire city was treated by this means. Schneider (2011) presents a testimony attributed to Cameron in which he cites fragments that formed the rationale involved in the development of the “original” septic tank:

¹ Illustration inspired by the drawing presented by Khan (2007).

“No system of treating sewage can be satisfactory which does not follow as closely as possible the lines laid down by nature. [...] I therefore set myself to devise a system in which the work of natural agents should be forwarded to the fullest extent. [...] The septic tank itself is merely a receptacle designed to favour the multiplication of micro-organisms, and bring the whole of sewage under their influence.”

William O. Travis, in 1903, developed and patented a sedimentation tank (later named “Travis tank”), consisting of two overlapping regions in which suspended solids originating from the upper channel – the “sedimentation chamber” – settled and accumulated over the steeply sloping bottom of the lower chamber (or “hydrolyzing chamber” according to the original terminology), in which occurred anaerobic degradation (Figure 3.7a). Travis believed it was desirable to allow part of the influent sewer to pass through the sludge digestion region, but this initiative caused resuspension – and consequent loss – of sludge within the effluent (METCALF & EDDY, 1922).

In 1905, Karl Imhoff, a German engineer, identified problems in a Travis tank installed in the city of Emscher (Germany) and promoted modifications to prevent disturbances caused by the flow of influent sewage inside the digestion chamber, giving rise to the “Imhoff tank” (Figures 3.7b and 3.7 c). Through the installation of baffles, the influent wastewater streamed along the sedimentation chambers and the sludge remained in the tank for a few weeks. After this period, a fraction of the biomass were discarded without notable perturbations. The produced gas bubbles emerged from the interior of the unit to an opening located in the upper region, whereby scum was also removed. The generation of gases was one of the features that most impressed the experts; considered intense, was compared to the foaming formation and boiling of water phenomena (adapted from METCALF & EDDY, 1922).

Following these developments, the anaerobic reactors were continuously improved in many European cities and in the United States; the importance of the methane generated for heating and power generation was largely recognized. Unfortunately, at that time, this technology was treated as a “black box” process, with little knowledge of the microbiology and biochemistry involved. In all aforementioned systems, no specific device was designed to

ensure or increase contact between the influent organic material and the micro-organisms, a fact which contributed to the relatively low efficiency of these technologies (30-50% removal of organic matter). From the 1930s, these systems have begun to lose ground to aerobic reactors, e.g. activated sludge system and trickling filters, better understood and more efficient options (adapted from McCARTY, 2001; VAN HAANDEL et al., 2006; FANG, 2010).

Figure 3.7 – Travis tank (a) and the Imhoff Tank (b/c); illustration adapted from McCARTY, 1981).

3.1.3 HIGH-RATE ANAEROBIC REACTORS

Anaerobic treatment systems for low strength wastewater should only be considered competitive if large volumes can be processed in a relatively short period (LETTINGA et al., 1980). According to Khanal (2008), the application of anaerobic process remained limited until 1950, year in which Stander (1950) recognized the importance of retaining solids for wastewater treatment and noted that, by separating the active micro-organisms from the effluent and maintaining them in the reactor, it has become possible to reduce the hydraulic retention time (θ_h). This was the basis for the development of high-rate anaerobic reactors, in which the cell residence time (θ_c), also known as “sludge age”, is not linked to the θ_h .

In an independent work, Schroepfer et al. (1955) developed a system denominated “anaerobic contact process” (ACP) to treat meat packing wastewater. These wastes were more diluted than the domestic sewage and, taking as reference the aerobic activated sludge process, these researchers used a sedimentation tank to collect the effluent of a continuous stirred tank reactor (CSTR) and recirculate the active biomass, conserving a high concentration of micro-organisms in the reaction compartment (Figure 3.8). A degassing unit was included in this system to prevent the formation of gas bubbles and to improve the sedimentation of the particles and, consequently, their separation from the liquid mass. This process is still used in the treatment of industrial effluents, due its high capacity for degradation of solids and fats without accumulation of scum in the reactor. Despite this advantage, this technology is being replaced by more efficient technical alternatives, such as the UASB, fluidized bed and internal circulation reactors (NADAIS et al., 2010).

Figure 3.8 – Schematic representation of the “anaerobic contact process”, as proposed by Schroepfer et al. (1955).

Years later, Stander (1966) demonstrated the validity of his own concepts in the treatment of winery wastewater using a “clarigester” ^[1] (Figure 3.9), an anaerobic reactor in which a sedimentation tank was positioned on top of a digester, for the return — and consequent accumulation — of the biomass to the bottom of the unit.

¹ The clarigester was not invented by Stander. According to a catalogue of the company that held the patent rights of this technology (DORR-OLIVER, 1959), the clarigester was originally installed in 1929 at a small municipal installation at East Rochester (State of New York, United States of America).

Figure 3.9 – The clarigester: original schematic illustration from the company that developed the unit (DORR COMPANY, 1945).

The anaerobic filter (Figure 3.10), originally developed by Coulter, Soneda and Ettinger (1957), became popular after Young and McCarty (1969) demonstrate its ability to treat synthetic wastewater containing carbohydrates and proteins, with relatively low COD (approx. $1,000 \text{ mg.L}^{-1}$). In this reactor, the influent is distributed over the bottom of the unit and flows upwards, through a bed of stones — maintained completely submerged — in which micro-organisms attached to the surfaces of the stones or suspended between its interstices degrade the substrate (adapted from VAN HAANDEL & LETTINGA, 1994). Among the advantages attributed to this technology, the lower constructive complexity stands out, as well as its superior tolerance to high organic loading rates (in comparison to other systems), the absence of agitation system, its rapid recovery after periods without substrate inflow (NADAIS et al., 2010). Its major limitations are: (i) the need for greater volume, due to the space required for disposing the filtering medium (10-40% of total volume) and; (ii) the presence of dead zones and preferential pathways due to accumulation of solids, especially in the lower region of the reactor (SHOW, TAY & HUNG, 2010).

Figure 3.10 – Anaerobic filter: illustrative scheme of its constructive details (adapted from GONÇALVES et al., 2001).

The “sludge blanket” concept of was originally applied in the “Reversed Flow Dorr-Oliver Clarigester” (Figure 3.11), which was a modified version of the anaerobic contact process (HEMENS, MEIRING & STANDER, 1962; STANDER et al., 1967). Unlike its predecessor (considered to be the ACP, not the “original” clarigester), this unit was not mechanically stirred and had an internal sludge recirculation system: the influent followed upwards through a dense liquid layer (i.e. sludge blanket), composed of micro-organisms and flocculated microbial aggregates, that accumulated in the sedimentation chamber and returned by gravity to the digestion compartment, with the aid of a scraper installed on the bottom of the settling tank (adapted from LETTINGA et al., 1980; NADAIS et al., 2010.).

In April 1970, investigations initiated at Wageningen University (Netherlands) by Professor Gatze Lettinga and his co-workers gave rise to what may be considered one of the most significant advances in wastewater treatment^[1]: the Upflow Anaerobic Sludge Blanket (UASB) reactor, initially applied to treat industrial discharges (LETTINGA, 2010, 2014). This system was designed after Lettinga glimpsed the possibility for improvement of a Reversed Flow Dorr-Oliver Clarigester installed in South Africa, regarding basically to the biomass retention in the reactor and its operational routines (adapted from HULSHOFF POL, 1989).

Figure 3.11 – The “Reversed Flow Dorr-Oliver Clarigester”: an schematic representation (adapted from CILLIE et al., 1969).

¹ Even though Lettinga was the main responsible for the rise of the UASB concept, his ideologies led him to oppose and successfully prevent the patenting of his invention (LETTINGA, 2014). This commendable decision favoured, more significantly, many citizens living in developing countries, who currently benefit from this system for the treatment of the domestic wastewater produced in their communities.

3.1.4 THE UASB REACTOR

3.1.4.1 Main characteristics

The UASB reactor is one of the technologies that presented the highest quantitative evolution in terms of implanted units throughout the recent decades, for the treatment of industrial and domestic wastewater in countries with tropical and subtropical climate. In 2009, when interviewed by Water 21 Magazine (McCANN, 2009), Lettinga stated that, at that time, there were more than 3,000 UASB reactors operating in the world.

The first experience with an UASB reactor treating in domestic sewage was reported by Lettinga et al. (1983), but its wide acceptance was stimulated by the encouraging performances observed in the first facilities built in Colombia (Cali and Bucaramanga), Brazil (Campina Grande and southern Brazilian State of Paraná) and India (Kanpur). In these locations, the maximum removal efficiencies obtained for chemical oxygen demand (COD), biochemical oxygen demand (BOD) and total suspended solids (TSS) reached 70–80%, 75–85% and 70–80%, respectively (adapted from VAN LIER et al., 2010).

In the UASB reactor, the influent sewage is distributed over the bottom of the unit and flows upwards through the digestion zone (i.e. sludge bed and/or blanket), region in which the substrate is subjected to degradation by a microbial community resulting in the anaerobic decomposition of the organic matter and, consequently, the production of biogas and sludge. After “crossing” the reaction compartment, the liquid enters the sedimentation zone through openings located between the “three-phase separators” (also referred in the literature as “gas-liquid-solid” or “GLS” separators) — the most characteristic element^[1] of the UASB reactor — which are installed on the top of the structure and basically divides the reactor into two internal “chambers”: the previously mentioned digestion zone (lower section) and settling zone (upper sector). In the settling compartment, due to the inclined walls of the separators, the area available for the liquid dispersion increases as it approaches the top of the unit, thus causing deceleration of the flow towards the effluent collection level. As a consequence of the decrement in the wastewater velocity, a parcel of the biomass particles dragged into the sedimentation zone are allowed to

¹ Due to the low growth rate presented by the anaerobic micro-organisms — caused by the inferior energy yield associated to the conversion of the organic matter into methane and carbon dioxide — the loss (or “washout”) of the sludge must be prevented. The development of a settler led to a decreasing escape of suspended solids; the sludge sent to the sedimentation compartment can flow back directly into the reaction zone by gravity, without the use of mechanical means (like the scrapers installed in the “Reversed Flow Dorr-Oliver Clarigester”) (adapted from ZEIKUS, 1977; VAN DER MEER & DE VLETTER, 1982).

settle. Hence, the existence of a settling apparatus positioned over the digestion zone enables the presence of dense sludge within the UASB reactor, while an effluent containing a relatively low concentration of suspended solids is collected. To avoid disturbances caused by the presence of gas bubbles, baffles placed underneath the GLS separators operate as gas deflectors, preventing them from entering the sedimentation compartment; the biogas, as a result, accumulates inside the hoods — also denominated as “(gas) domes”^[1] — established in the inner region of the three-phase separators and is removed from the reactor via a piping system (adapted from VAN HAANDEL & LETTINGA, 1994). A schematic representation of the UASB is shown in Figure 3.12.

Figure 3.12 – Schematic layout of an UASB reactor (LETTINGA & VINKEN, 1979).

The advantages attributed to the UASB reactor (Table 3.1) make it an attractive option for wastewater treatment, as well as the most popular choice for high rate systems based on the anaerobic process (VAN HAANDEL & VAN DER LUBBE, 2012).

¹ HASKONING & WAGENINGEN AGRICULTURAL UNIVERSITY (1994)

Table 3.1 – Main advantages, presented in the literature, associated with the UASB reactor treating sewage, in comparison to aerobic systems.

Advantages (compared to an equivalent aerobic treatment system)^[1]	
<i>Operational costs</i>	<ul style="list-style-type: none"> • Substantial savings due to the lack of electricity consuming equipment (e.g. aerators, agitators); • Lower demand for operators; • In the case of sewage treatment, optimal pH may be maintained without addition of chemicals..
<i>Implementation costs</i>	<ul style="list-style-type: none"> • Significant reduction regarding land area requirements; • Less complex infrastructure (e.g. the unit does not require power supply).
<i>Energy / Biogas</i>	<ul style="list-style-type: none"> • Rather than consuming electric energy, there is possibility of using the biogas for electricity generation; • Reduction in the emissions of greenhouse gases, when implanted initiatives for biogas management.
<i>Operational complexity</i>	<ul style="list-style-type: none"> • High-tech equipment are not required, making the system less dependent on complex and/or expensive products.
<i>Robustness</i>	<ul style="list-style-type: none"> • The system is capable of maintaining acceptable performance when subjected to hydraulic or organic overloads.
<i>Hydraulic Profile</i>	<ul style="list-style-type: none"> • Exemption of pumping operation, since the sludge returns to the digestion zone by gravity. • If properly designed, the available static head in the entrance of the influent distribution system will be sufficient to allow the uninterrupted flow of substrate along the unit.
<i>Compactness</i>	<ul style="list-style-type: none"> • It is a compact unit, usually operated with a hydraulic retention time ranging between 6 and 9 hours.
<i>Decentralized facilities</i>	<ul style="list-style-type: none"> • The application at smaller scales enables the option for decentralized treatment.
<i>Sludge management</i>	<ul style="list-style-type: none"> • The volume of sludge produced is considered inferior to that generated by aerobic processes; • The sludge is well stabilized, contains nutrients (i.e. nitrogen and phosphorus) and is easier to dewatering.
<i>Helminth removal</i>	<ul style="list-style-type: none"> • The removal of helminths eggs can be achieved in units designed and operated properly.

Sources: LETTINGA (2010); VAN LIER et al. (2010).

Lettinga et al. (1980) revealed the fundamental precepts of UASB technology as follows:

¹ Direct comparisons must be analysed with caution, due to the difference in the quality of the final effluent.

- a) *The anaerobic sludge obtains and maintains superior settling characteristics if chemical and physical conditions favorable to sludge flocculation and to the maintenance of a well flocculated sludge are provided;*
- b) *A sludge blanket (bed) may be considered as a separate – more or less – fluid phase with its own specific characteristics. A well-established sludge blanket frequently forms a rather stable phase, capable of withstanding relatively high mixing forces. The redispersion of the sludge in the liquid phase therefore may require a significant amount of mixing energy;*
- c) *The washout of discrete sludge particles (flocs) released from the sludge blanket can be minimized by creating a quiescent zone within the reactor, enabling the sludge particles to flocculate, to settle, and/or to be entrapped in a secondary sludge blanket (present in the settler compartment).*

One of the crucial advantages associated with UASB reactor is its capacity to provide conditions for the establishment and retention of large quantities of “viable” biomass, which in the form of granular or flocculent sludge with good settleability, enables the imposition of cell residence time (θ_c ; Equation 3.1) much higher than the hydraulic retention time (θ_h ; Equation 3.2) (LETTINGA et al., 1980; CHONG et al., 2012).

$$\theta_c = \frac{\text{VSS present in the reactor (kg)}}{\text{VSS removed per day (kg.d}^{-1}\text{) + effluent VSS (kg.d}^{-1}\text{)}} \quad (\text{Equation 3.1})$$

in which:

- θ_c : mean cell residence time, or “sludge age” [d].

$$\theta_h = \frac{V_R}{Q} = \frac{1}{L_h} \quad (\text{Equation 3.2})$$

in which:

- θ_h : hydraulic retention time [d];
- V_R : reactor volume [m³];
- Q : flow rate [m³.d⁻¹];
- L_h : hydraulic loading rate [m³.m⁻³.d⁻¹].

The effect of the hydraulic loading rate on the sludge retention is closely related to the agitation intensity applied in the reaction zone. In the absence of mechanical stirring or agitation caused by the gas dispersion, the wastewater may find its path flowing upwards through the channels and cracks formed in the blanket, leaving the thickened character of the main part of the sludge unaffected (LETTINGA et al., 1980).

According to van Haandel and Lettinga (1994), the hydraulic loading rate will represent the limiting factor if the performance of the treatment system becomes inadequate due to an excessive loss of active micro-organisms — original constituents of the sludge bed/blanket — carried in the effluent (i.e. “washout”). When the organisms present in the reactor are not able to convert the organic material contained in the influent (according to the desired efficiency), then the organic loading rate (L_o) may be considered excessive.

$$L_o = \frac{Q \cdot S_0}{V_R} \quad (\text{Equation 3.3})$$

in which:

- L_o : organic loading rate [kg COD.m⁻³.d⁻¹]; and
- S_0 : influent COD [kg.m⁻³].

Despite the notorious results obtained in some pioneering experiments with the UASB reactor, as previously mentioned, recent studies of investigative character suggests that the performance of novel full-scale units is inferior to those preliminarily reported. Operational performance data of some UASB reactors applied to domestic wastewater treatment — at pilot and full scales — are presented in Table 3.2. The occurrence of lower-than-expected results should not be attributed only to design flaws, but also to carelessness or neglect regarding the operation and maintenance of the systems (adapted from VAN LIER et al., 2010).

Table 3.2 - Operational performance data of UASB reactors treating domestic wastewater.

	Location (Country)	Number of reactors [units]	Volume of each reactor [m ³]	Temp. [°C]	θ_h [h]	V_{up} [m.h ⁻¹]	Removal efficiency [%]			References
							BOD	COD	TSS	
PILOT SCALE	Brazil	1	0.1	19–28	4.0	N/A	78	74	72	BARBOSA & SANT'ANNA, 1989
	Colombia	1	35.0	N/A	5.0	N/A	80	66	69	SCHELLINKHOUT & COLLAZOS, 1992
	Japan	1	1.1	9–32	6.0	0.6	53	62	88	TANDUKAR, OHASHI & HARADA, 2007
	Japan	1	20.2	16–29	9.7	1.0	75	64	68	SYUTSUBO et al., 2011
	Brazil	1	0.3	18–32	7.0	N/A	N/A	71	82	PEREIRA, 2012
	Brazil	1	14.2	N/A	10.3	0.4	67	57	62	VON SPERLING, 2015
FULL SCALE	India	1	1,200	20–30	6.0	2.3	75	74	75	DRAAIJER et al., 1992
	Colombia	3	3,360	24	5.0	NA	76	74	63	SCHELLINKHOUT & OSORIO, 1994
	Brazil	1	810	30	8.8–9.7	0.5–2.5	N/A	66	56	FLORENCIO, KATO & MORAIS, 2001
	Egypt	1	4,187	N/A	12.0	2.2	70	70	86	NADA et al., 2011
	Brazil	24	2,286	N/A	8.0	0.6	71	60	49	HEFFERNAN et al., 2011
	Brazil	16	2,029	N/A	8.0	0.5	67	67	70	HEFFERNAN et al., 2011
	Brazil	6	2,029	N/A	7.5	0.6	65	49	50	HEFFERNAN et al., 2011
	UAE	8	2,645	N/A	8.0	0.5	51	52	84	HEFFERNAN et al., 2011
	India	6	2,534	N/A	7.2	N/A	62	75	70	KHAN et al., 2014

3.1.4.2 Reactor configuration for sewage treatment

As mentioned by Chernicharo (2007a), UASB reactors were primarily designed to treat industrial wastewater and its adaptation for sewage treatment resulted in the following alterations:

- Differently from the projects developed for industrial wastewater treatment plants, which are guided by the organic loading, the design for sewage treatment is based on the hydraulic loading criteria. Because of this, the upflow velocity established inside the reactor represents a limiting factor: excessive velocities may cause the depletion of the biomass concentration in the digestion zone, thus restricting the performance of the process. As a result, to maintain the upflow velocity within reasonable range, the height of the reactor should be reduced and its cross section should be increased.
- In reactors treating industrial discharges, the influent is generally pumped into the tank via ramified piping system placed over the bottom of the reactor. The “preliminary” distribution of the influent flow in UASB systems designed for sewage treatment is located in the upper part of the structure. Hence, the effective surface area of the settler may be reduced due to the space occupied by the components of this element.
- Unlike what is often observed in industrial facilities, the installation of an equalisation basin in domestic sewage treatment plant is improbable, exposing the anaerobic reactor to hydraulic and organic overloads. The increased cross section of the structure close to the sedimentation zone may be — once again — a solution to favour lower upflow velocities during adverse operational conditions.

There are currently several designs options for this type of technology, but all require improvement concerning to the distribution of the influent sewage, retention of biomass within the reaction zone, withdrawal of the scum (especially the fraction retained under the three-phase separators), biogas collection, corrosion etc. The problems related to the implementation of the UASB system and its components are presented in Section 3.1.4.3.

The most important elements of the UASB concept and its design criteria are presented in the following sub-items. It is worth to mention that this work addresses, exclusively, the variant design of the UASB reactor designed to treat domestic sewage.

The influent distribution system:

To ensure an intimate contact between micro-organisms and the substrate, and to reduce the occurrence of dead zones along the sludge bed, the incoming wastewater must be homogeneously distributed a few centimetres above the internal base of the reactor. In order to obtain the expected result, it is necessary to design and employ a flow-splitting system able to introduce — simultaneously and equally — the sewage at various points over the bottom of the unit. In the reactor, an influent distribution apparatus is installed on top of the covering structure^[1]; it is composed by relatively small devices (denominated in literature^[2] as “distribution wells”) subdivided into compartments^[3] (“influent boxes”), each of them provided with weirs (usually triangular) and piping accessories (“lines”) extending to predetermined points located over bottom of the reactor to allow a proper dispersion of the liquid through the tank (adapted from HASKONING & WAGENINGEN AGRICULTURAL UNIVERSITY, 1994; VAN LIER et al., 2010). As stated by van Haandel & Lettinga (1994), one inlet point is sufficient to cover an area up to 4 m². To illustrate the complexity involved in the design of full-scale UASB reactors, especially with regard to the influent distribution system, Figure 3.13 shows a theoretical example of a 2,940 m³ unit (p.e. 40,000 inhabitants, approximately) to which are needed 192 sewage discharge points (according to the recommended design criteria). According to van Lier et al. (2010), the design of the distribution wells^[4] varies greatly in full-scale systems. Ideally, its shape should be symmetrical and circular, when serving a large number of inlet pipe, due to the lower constructive complexity (VAN HAANDEL & LETTINGA, 1994; MIKI, 2010; VAN HAANDEL & VAN DER LUBBE, 2012). Some of the prevailing engineering solutions in Brazil, regarding the distribution system are shown in Figures 3.14 and 3.15 (both theoretically designed with a total volume of 588 m³, to enable comparisons and to highlight the complexity of their internal arrangement).

¹ To ensure that each inlet point receives its adequate parcel of the influent flow, it is highly recommended to build the distribution system at a more elevated (hydraulic) level than the maximum liquid level established inside the reactor; feeding can then be achieved by gravity. Although the theoretical head losses over the inlet tubes are small (ranging from 7 to 15 cm), if the difference between the liquid levels set out in both structures is lower than 10 centimetres, obstructions may occur frequently due to partial clogging of the tubes and additional pressure losses at the bottom caused by concentrated sludge; when the variation is approximately 30 cm, such blockages are uncommon because the extra static head will enable some self unplugging at a sudden clogging of a tube (VAN HAANDEL & LETTINGA, 1994; HASKONING & WAGENINGEN AGRICULTURAL UNIVERSITY, 1994).

² VAN HAANDEL & LETTINGA, 1994.

³ Depending on the scale of the treatment plant, the installation of “preliminary” influent distribution chambers (placed upstream of the “distribution wells” and referred in the literature as “influent distribution channels”), might be required. In these structures — commonly found in large installations, and also located over the reactor — the incoming flow, hitherto transported by pipes with relatively large diameters, is distributed between its internal subdivisions (by the introduction of weirs) and conducted to the smaller compartments through a manifold system.

⁴ In their text, the referred authors decided designate this element as “final distribution box”.

Figure 3.13 – Influent distribution system: theoretical example of a 588 m³ UASB reactor.

Figure 3.14 – Influent distribution: main characteristics of one of the various UASB conceptions used in Brazil; (a) external layout; (b) internal arrangement of the elements that compose the feeding system.

Figure 3.15 – Influent distribution system of another UASB variant used in Brazil: (a) external layout; (b) internal arrangement of the elements that compose the feeding system.

In order to guarantee an even and consistent flow distribution, it is recommended that: (i) the verification and adjustment of flow regime in each inlet point must be assured; and (ii) eventual obstructions of these points should be easily detectable^[1] and eliminated. According to Miki (2010), the design proposals opened to criticism are those which opt for branched distribution or perforated pipe systems (e.g. manifold flow).

With respect to the pipes introduced to transport and discharge the wastewater into the reactor, the prime conditions for a proper operation are as follows (VAN HAANDEL & LETTINGA, 1994; CHERNICHARO, 2007a):

- Its diameter should be large enough to prevent obstructions by solids and to maintain the descending velocity within the range below 0.2 m.s^{-1} , so that the unwanted — but eventually present — gas bubbles can rise, flowing against the liquid current, to reach the atmosphere;
- Contrary to the recommendations regarding the necessity of maintaining reduced flow velocities in the piping system, the velocity of the liquid jet at the discharge point must be sufficiently high to promote satisfactory mixing intensities in the digestion zone, and to prevent an excessive deposition of inert solids (e.g. sand particles) over the bottom of the tank. A proposed solution is the installation of concentric reductions (e.g. nozzles) or the execution of lateral apertures in the lower extremities of the influent distribution system, limiting the cross-sectional area of the flow stream thus causing an increase in flow velocity.

As the distribution wells are placed over the main structure of the tank, the inlet piping system should either be deviated or passed through other elements that compose internal arrangement of the UASB (e.g. GLS separator, gas deflector etc.), in order to reach the bottom of the reactor (Figure 3.16). According to van Haandel and Lettinga (1994), the crossing of tubes via openings in the vertical walls is a less attractive option. The difficulty to unclog the pipes is a disadvantage of a deviated pipeline. If the tube is passed through the three-phase separator at a level bellow the gas-liquid interface, the mandatory perforation may not cause problems, and is even advantageous as it also functions as an emergency device, allowing gas escape from the

¹ According to van Haandel and Lettinga (1994), to verify the uniform division of the influent flow, individual distribution lines must be installed, i.e. it is necessary that each influent box is connected to an exclusive distribution pipe, which distributes the flow at a single feeding point.

hoods in case of obstruction in the gas withdrawal system. If the passage of the pipe must be performed at a level above the gas-liquid interface, an airtight guidance tube should be installed.

Figure 3.16 – Conflict between the influent distribution piping system and other internal elements that compose the UASB reactor: an alternative for crossing the tubes.

The gas–liquid–solid (GLS) or three phase separator

By definition, the GLS separator segregates constituents of the three phases present in the UASB reactor, i.e. the gas bubbles (G), the liquid effluent (L) and the – suspended – solids (S); it performs two important functions in this process: (i) prevent the biogas formed in the digestion zone from entering the settling compartment, collecting it in the gas hood for subsequent usage/disposal; and (ii) allow the incoming solids to settle in the quiescent environment made possible by the absence of gas bubbles^[1] (VAN LIER et al., 2010).

Therefore, it can be stated that the installation of three-phase separator implies the creation of two distinct environments in the upper region of the reactor: the settler (outer

¹ As van Haandel and Lettinga (1994) recommend, to prevent biogas — and a parcel of the solids — from invading the sedimentation compartment, a set of “gas deflectors” should be positioned underneath these gas collection elements.

region) and the gas hoods (inner zone). The pre-eminent considerations for the design of the GLS separator (contemplating its two main functions) are the following (adapted from VAN DER MEER & DE VLETTER, 1982; VAN HAANDEL & LETTINGA, 1994; CHERNICHARO, 2007a; MIKI, 2010; VAN LIER et al., 2010, VAN HAANDEL & VAN DER LUBBE, 2012):

- **Gas separation:**
 - In order to avoid the sludge flotation, the dimensions of the separator should be such that they admit the formation of a liquid-gas interface inside the gas hood, sufficiently large to facilitate the emancipation of the gas entrapped in the sludge;
 - The gas release rate^[1,2] should be adequately high to overcome the flow resistance imposed by an eventual scum layer, but low enough to permit the prompt liberation of the gas from the solid particles;
 - The gas hoods must be accessible and sufficiently large in order to offer good conditions for inspection and cleaning, considering that a fraction of the scum formed in the reactor accumulates beneath this structures;
 - With respect to the materials^[3] used in its construction, preference should be granted for corrosion resistant components;
 - In cases in which material subject to corrosion is applied, placing the device(s) underneath the liquid surface may decelerate its gradual process of destruction;

¹ Chernicharo (2007a) presents the equation by which the biogas release rate can be determined.

² According to van Haandel and Lettinga (1994) lower gas release rates are associated with the formation of scum, which may become thick and hard over time (especially under low temperatures), seriously jeopardizing the performance of the gas collection system.

³ Several materials of construction are used for GLS separators, each of them with its pros and cons. Concrete gas collection plates require extensive form work and therefore result in high labour requirements. GRP (glass-reinforced polyesters) plates are expensive, but can be pre-fabricated and are installed relatively easily. PVC (polyvinyl chloride) plates are prone to distortion or even rupture, but are relatively inexpensive to manufacture and install. The PVC-coated polyester sheets are relatively new on the market (although frequently used for inflatable gas holders) and were pioneered in Brazil. The material selection for the GLS separator will be based on the local procurement and construction costs, as well as on the skill level of the civil contractor (VAN LIER et al, 2010).

- In its conventional configuration^[1], the GLS separator is not submerged, which reduces the surface area available for settling at the overflow level. The adoption of “elevated” separators (i.e. not entirely submerged) sometimes lead to surface loads that are over 50% higher than the applied cross-sectional liquid upflow velocity. However, there are also certain advantages associated with a separator (partially) located above the reactor liquid level: (i) it acts as a support structure for the effluent collection devices and the distribution wells, and (ii) the top of its structure can be used as walkways, providing easier access to the influent distribution and effluent collection systems as these require regular maintenance;
 - In order to ensure the reliability of the biogas capture system, an overlap^[2] should be created between the deflector and the gas hood, limiting the size of the apertures that allow the flow progression into the settler.
- **Solid separation:**
- Ideally, the angle of the GLS separator wall with the horizontal axis should be higher than 50°. The choice for steeper structures imply advantages and drawbacks: a sharp inclination favours the sliding of the settled solid particles back to the digestion zone, but represent higher installation cost as it demand greater structure and consequent higher spending on material;
 - To avoid gas production in the settler, the hydraulic retention time must be relatively short and the eventual invasion of the region by the sludge blanket should be prevented (for this reason there is an optimum height for the sludge blanket);
 - The liquid velocity in the apertures is considered important because it has an effect on the amount of particles swept into the settling compartment.

¹ In their work, van Haandel and Lettinga (1994) and van Lier et al. (2010) present some conceptual engineering options for the design of the GLS separator.

² A valuable publication entitled “DECOM Manual” (HASKONING & WAGENINGEN AGRICULTURAL UNIVERSITY, 1994) expose important practical data to be applied in the design of UASB reactors, in addition to enlightening illustrations (one of them presenting the “overlap” concept).

- Consultants can opt to cover or not the settling compartment of the UASB reactor. When compared to a covered structure, the potential emission of odours is a negative aspect associated with (partially^[1]) open tanks. On the other hand, there is a hypothesis that the increased internal pressure in a covered sedimentation compartment can stimulate the dissolution of the H_2S , thus affecting the effluent quality^[2]. An unsealed construction may facilitate the access conditions for maintenance, visual inspection and collection of samples for further analysis, while a fully covered reactor might require an exhaust system to remove the gas accumulated over the settler and the installation of inspection hatches for sample collection and maintenance.

Gas deflectors

As its name implies, the installation of the gas deflectors — situated under the GLS separators — directs the gas flow to the inner region of the separators, allowing only liquid and solid particles to ingress the sedimentation zone. Different types of deflectors are constructed, differing primarily in its shape and eventual use of additional devices to orientate the gaseous flux. It is appropriate to emphasize that the sludge blanket should be maintained below the widest point of the deflector, in order to prevent biogas production beneath the settler (VAN LIER et al., 2010).

Effluent collection system

The effluent withdrawal devices are fixed structures located at the upper region of the sedimentation compartment, installed along the outer side of the gas hoods; their dimensions and positioning define the liquid surface level in the tank. As well as the reactor feeding strategy, the effluent collection should be balanced in order to maintain a good performance of the GLS separator and to avoid short-circuiting between influent and effluent flows. There two main

¹ *The gas hoods must remain airtight under any circumstances.*

² MIKI (2010).

alternatives for the collection of effluent from the top of the UASB unit: (i) effluent gutters — also referred as “launders”^[1] — endowed with triangular (generally 90° V-notches) overflow weirs and (ii) submerged perforated pipes (adapted from CHERNICHARO, 2007a; VAN LIER et al., 2010; VAN HAANDEL & VAN DER LUBBE, 2012).

The use of V-notch weirs in the gutters is preferred as they can be easily aligned^[2] and cleaned^[3,4], assuring an even liquid (up)flow in the settling compartment; of course, the dimensions of the notches should be suited to receive peak flows. To eliminate — or, at least, reduce — the problems caused by clogging of the weirs, the static head over the crest of the weir should be not lower than 25 mm. In order to facilitate their maintenance, the width the gutters should be larger than 15 cm (adapted from VAN HAANDEL & LETTINGA, 1994; HASKONING & WAGENINGEN AGRICULTURAL UNIVERSITY, 1994; VAN LIER et al., 2010).

In cases which triangular weirs are adopted, partially submerged scum baffles may be installed along the gutters to reduce operator attendance, although in Brazil, modern reactors have been designed without such elements, allowing continuous removal of the floating solids within the effluent. As the expected concentrations of (floating) solids are reduced, since the discharge occurs simultaneously with the formation of scum, the impact on the effluent quality should be minimized. Additional care regarding the gutters allude to the potential gas release, particularly hydrogen sulphide (H₂S), triggered by turbulence developed immediately downstream of the weir crest (adapted from CHERNICHARO, 2007a; 2007b).

The use of submerged perforated pipes for the collection of effluent is unconventional (VAN LIER et al., 2010), but according to Chernicharo (2007a) this alternative has shown to be very efficient, due essentially to the following features:

- As immersed elements, the preservation of uniform flow in the orifices is favoured, and the requirements related to their alignment become less important;
- The application of submerged pipes minimizes the risks associated with occurrence of

¹ CHERNICHARO, 2007a.

² When compared to another types of weirs (e.g. rectangular or trapezoidal shapes).

³ Small misalignments in the gutters and/or blockages in the weirs can represent a significant discrepancy in the flow collected at distinct points (VAN HAANDEL & LETTINGA, 1994; CHERNICHARO, 2007a; VAN LIER et al., 2010; VAN HAANDEL & VAN DER LUBBE, 2012).

⁴ Cleaning V-notches is relatively easy, but the task is threatened in the cases which the unit is fully covered by a concrete structure, even if inspection hatches are implemented. If coverage is mandatory, then light-weight modular covers are recommended (VAN HAANDEL & VAN DER LUBBE, 2012).

disturbances in the liquid flow, and consequential gas release;

- This option does not require the implementation of scum baffles, once the effluent is collected few centimetres underneath the liquid surface (region located in the upper region of the tank in which the scum layer develops)^[1], characterized by establishing an liquid-gas interface (in this case^[2], between the effluent and the atmospheric air).

The potential accumulation of solids in the orifices and inside the tubes is a drawback. As maintenance procedures are not always feasible, it is recommended that the tubes are arranged with a minimum inclination of 1%, thereby allowing self-cleaning action. It is observed therefore that both systems require provisions to allow their periodic cleaning, especially if the reactor is covered (CHERNICHARO, 2007a; VAN LIER et al., 2010).

Sludge discharge system

Essentially, the improvement of the effluent quality in the UASB reactor is directly proportional to the increment of the sludge mass within the system, when a successful retention of suspended solids is achieved. However, once the maximum capacity for sludge deposition is reached, the containment of biomass becomes inconceivable, and the loss of solids will affect the general performance of the process^[3]. For this reason, it is recommended to discharge the sludge in every occasion in which the sludge blanket equals a predetermined level. Ergo, the implementation of systems for monitoring the height of the sludge blanket and sludge withdrawal is primordial. At least two^[4,5] independent withdrawal piping systems (placed at distinct levels) should be installed^[6]: (i) one not far from the bottom of the tank and (ii) another located approximately 1.5 m above the internal base of the reactor. It is important to discharge

¹ The development of the scum layer in this region is considered disadvantage by van Lier et al. (2010).

² Under the gas hoods (i.e. the inner region of the GLS separator), the build-up of scum is observed at the boundary between the liquid and biogas (not atmospheric air).

³ By increasing values of effluent COD and TSS, mainly.

⁴ CHERNICHARO, 2007a; VAN HAANDEL & VAN DER LUBBE, 2012.

⁵ According to van Lier et al. (2010) discharge devices must be installed over at least three levels. In comparison to the recommendation made by another authors, the additional point (to perform the withdrawal) must be introduced "approximately at half way between the reactor floor and the lowest part of the GLS separator".

⁶ According to Chernicharo (2007b), each piping system placed in the lower region of the digestion zone for sludge discharge must be equipped with an individual control valve. Contrarily, the use of interconnected piping (manifold type) regulated by a single valve is allowed in the upper zone, for the removal of the less dense sludge.

sludge with a lower activity, which in the case of a municipal UASB may be located in upper layer (comprised of flocculent sludge). On the other hand, due to sedimentation of inert particulate material, the inorganic sludge fraction of the bottom sludge will gradually increase, reducing the sludge activity and increasing the density of the sludge to a certain point in which the inlet and/or sludge discharge pipes become blocked (adapted from VAN HAANDEL & LETTINGA, 1994; CHERNICHARO, 2007a; VAN LIER et al., 2010).

Collection of sludge samples

In order to evaluate the biomass expansion (i.e. sludge blanket height) inside the UASB reactor, it is indispensable to precisely determine the suspended solids (concentration) profile, established along the height of the digestion zone. The (positive/negative) difference between the measured and the — previously stipulated — maximum permitted height of the sludge blanket will determine relevance of the withdrawal procedure (required/unnecessary). In addition, the total sludge mass present in the reactor (calculated based on the collected data^[1]) and the specific methanogenic activity (SMA) of the sludge samples can be determined^[2,3] (adapted from VAN HAANDEL & LETTINGA, 1994; CHERNICHARO, 2007a; VAN LIER, et al., 2010).

The samples can be obtained by two methods: (i) through the installation of a set of tubing segments — controlled by ball valves — arranged over the height of the reactor wall^[4] or (ii) with the aid of a submersible sampler or sample extraction apparatus (allowing extraction by from any desired height), introduced in the unit via an inspection hatch implemented on top of the GLS separator^[5].

Scum removal devices

The composition of the scum layer that develops in the UASB reactor is highly dependent on the characteristics of the raw sewage and performance of the units that precede

¹ Some authors describe the procedure, e.g. van Haandel and Lettinga (1994) and Chernicharo (2007a).

² SMA tests will help the operators to define in which level the discharge must be performed, usually in the point presenting lower activity.

³ The determination method for SMA used in this work is described in Section 4.5.11.

⁴ As recommended by Chernicharo (2007a).

⁵ The best option according to van Haandel and Lettinga (1994), van Lier et al. (2010) and van Haandel and van der Lubbe (2012).

it (i.e. preliminary treatment). At the initial stages of growth, the scum presents a liquid-like consistency, but it gradually becomes very thick and viscous. According to van Lier et al. (2010), the issues related to the scum formation are the following:

- Obstruction of weirs, resulting in an irregular effluent withdrawal;
- Potential problems with odour, procreation of vectors and undesired vermin; and
- Accumulation (and eventual solidification) in the upper region of the unit, especially under the gas hoods where it may impair the biogas collection, potentially leading to an uncontrolled dispersion of bubbles, or even ruptures in the GLS separators.

As previously cited, the build up of scum in the sedimentation compartment can be mitigated even if containment baffles are not installed (i.e. through its continuous removal within the effluent^[1]), but there is an alternative method: the installation of a set of “bowls”, placed along the settler region (with intervals of 2 to 3 m) few centimetres beneath the liquid level. These devices are usually connected via piping system, which is controlled by a gate valve; when opened, the valve allows the drainage of a parcel of the liquid and the scum layer present over the settling zone^[2] (adapted from VAN LIER et al., 2010).

Under the gas hoods the scum removal is undoubtedly an even more difficult assignment. Several methods have been proposed for this purpose, each with varying degrees of success (VAN LIER et al., 2010). A popular method is the periodic removal by vacuum pumping through access points (hatches) installed on top of the GLS separator; its disadvantages are presented by van Lier et al. (2010) and van Haandel and van der Lubbe (2012). Another commonly applied method is based on the use of liquid jets^[2] directed to the liquid-gas interface created in the gas hood in an attempt to dissolve and disperse the scum. According to van Lier et al. (2010), this method has not proved to be productive; Miki (2010) cites a positive experience in one of the Brazilian municipal WWTPs visited by him.

After several technical visits, Miki (2010) expressed his satisfaction concerning the

¹ There is a risk of blockage in the cases which the flow rate (or static head) over each weir is considered extremely low (VAN HAANDEL & LETTINGA, 1994; VAN LIER et al., 2010; VAN HAANDEL & VAN DER LUBBE, 2012).

² According to van Lier et al. (2010), the scum should be discharged into the post-treatment unit. Return of this floating matter to the UASB reactor is not recommended as it will immediately result in a novel scum escalation.

¹ In Brazil it is generally noticed the use of water and treated effluent from UASB reactors or post-treatment units. Miki (2010) condemns the use of potable water for this purpose.

system implemented at the Matão municipal WWTP (Brazil) in which a routine based on the utilization of a manually operated stop-log gate is applied to the control the scum removal. When the gate is removed, the scum leaves the GLS separator and is discharged, by gravity, into an exclusive channel. This system also uses jets to the dissolution and displacement of the scum. Another Brazilian proposal was developed by Chernicharo et al. (2009a); its principle is based on the reduction or increase of the biogas pressure inside the GLS separator, in order to permit the flow of scum over a receptacle endowed with weirs, installed inside of the gas hood.

Biogas Conveyance system

The biogas conveyance system must be installed above the ground (for easier leak identification) and are subject to the same rules as those used for natural gas. Under certain conditions, extensive pipelines must be heated so that the temperature does not reach the dew point, avoiding the development of corrosive substances. All piping system must provide gentle slope to allow drainage at its lowest point (DEUBLEIN & STEINHAUSER, 2008).

The biogas composition varies according to the properties of the degraded substrate. When a dilute wastewater is treated (e.g. domestic origin), according to Noyola, Capdeville and Roques (1988), the anaerobic process will produce a mixture of gases strongly influenced by the temperature, consisting of methane (70-80%), nitrogen (10-25%) and carbon dioxide (5-10%). The relatively high nitrogen content is attributed to the dissolved N_2 in the sewage, that separates from the liquid within the reactor.

Table 3.3 – Biogas main features.

Aspect	Usual values / Characteristics
<i>Energetic content</i>	• 6,0 – 6,5 kWh.m ⁻³ .
<i>Energy equivalence, relative to other energy sources (reference: 1 m³ biogas – 60% CH₄)</i>	<ul style="list-style-type: none"> • 0,714 L petroleum; • 0,620 L kerosene; • 3,474 kg firewood; • 0,433 kg butane; • 4,698 kWh electricity.
<i>Explosive limit</i>	• 6 – 12% in air.
<i>Ignition temperature</i>	• 650 – 750°C.

(Continued on next page)

Table 3.3 (continuation) – Biogas main features.

Aspect	Usual values / Characteristics
<i>Critical pressure</i>	• 75 – 89 bar.
<i>Critical temperature</i>	• - 82,5 °C.
<i>Normal density</i>	• 1,2 kg.m ⁻³ .
<i>Odour</i>	• Rotten egg.
<i>Molar mass</i>	• 16,043 kg.mol ⁻¹ .

Sources: DEUBLEIN & STEINHAUSER, 2008; KHOIYANGBAM, GUPTA & KUMAR, 2011

Brazilian Technical Standard concerning WWTPs design

The recent review of the Brazilian technical standard ABNT NBR 12.209 (ABNT, 2011) was released in 2011, contemplating — for the first time — the guidelines for the design of UASB reactors implemented for domestic sewage treatment. This document included, among other recommendations, the following:

- Preceding the reactor, equipment for the removal of coarse solids must be installed (with aperture ranging between 6 and 12 mm, depending on the flow rate);
- The influent distribution system must be designed in way that identification of clogging points is facilitated, and the entrainment of air bubbles within the (influent) flow is avoided;
- The reactor must contain scum removal system;
- The GLS separator must be made from gas-tight and corrosion resistant material
- UASB units must be provided with a sludge sampling system, allowing extraction at different points along the height of the reactor;
- Two independent piping systems shall be installed for sludge withdrawal: (i) one close to the bottom of the unit (which must also allow complete emptying of the reactor) and (ii) other above the bottom, far between 0.8 and 1.3 m from its base. These apparatus should have at least one collection point for every 100 m² of bottom surface area.

Other values indicated by the referred technical standard (ABNT, 2011), regarding the criteria and parameters with the greatest influence on the performance of the biological stage itself, are presented in Table 3.4.

Table 3.4 – Recommendations for the design of “municipal” UASB reactors in Brazil.

Criteria / Parameter	Unity	Condition	Value
Total hydraulic retention time ^[1]	h	T>25°C	6.0
		22°C ≤ T ≤ 25°C	7.0
		22°C ≤ T ≤ 25°C	8.0
		15°C ≤ T ≤ 17°C	10.0
Upflow velocity in the digestion zone	m.h ⁻¹	Q _{mean}	≤0.7
		Q _{max}	<1.2
Hydraulic retention time within the settling compartment	h	Q _{mean}	≥1.5
		Q _{max}	>1.0
Maximum coverage area of each inlet point placed over the bottom of the reactor	m ²	–	3.0
Distance between the internal base of the unit and influent feeding points	m	–	0.1 to 0.2
Total depth of the reactor	m	–	4.0 to 6.0
Height of the digestion zone	m	–	2.5
Minimum height of the sedimentation section ^[2]	m	–	1.5 ^[2]
Inclination of the GLS separator walls in the settling zone	degrees	–	≥ 50
Minimum internal diameter of the influent distribution system tubes	mm	–	75

Source: ABNT (2011).

3.1.4.3 The main issues related to the application of UASB reactors in domestic wastewater treatment systems

There are actually various design variations for the UASB reactor, but all require improvement. In recent years, several authors dedicated themselves to disclose the difficulties and limitations associated with the use of this type of unit in sewage treatment systems, especially in developing countries such as Brazil and India. A compilation of information gathered by these researchers is shown in Table 3.5.

¹ Lower values may be admitted, provided they are justified.

² A vertical wall with a height of at least 0.3 meters should be installed.

Table 3.5 - Problematic aspects associated with the use of UASB reactors treating domestic wastewater.

Problem / Disadvantage	Description
<i>Service scale / Large units</i>	<ul style="list-style-type: none"> • The multiple inlets and outlets require continuous presence of operators for maintenance and valve actuation; ^a • Despite the economy observed in its implementation, larger reactors provide lower operational flexibility. ^b
<i>Preliminary treatment</i>	<ul style="list-style-type: none"> • To prevent accumulation of inert solids, clogging of the piping system and to reduce the risk of scum formation, the UASB reactor becomes highly dependent on the efficiency of the preliminary treatment. ^{a,b,c,d}
<i>Influent distribution system</i>	<ul style="list-style-type: none"> • The high density of feeding points makes the system susceptible to damage and obstructions; ^{a,b,d} • Blockages or failures regarding design and/or installation may cause differences in the pressure inside the pipes, compromising the homogeneous dispersion of sewage. ^{b,d}
<i>GLS separator</i>	<ul style="list-style-type: none"> • In relation to its manufacturing, there is currently no material that is presented as an ideal alternative; ^b • To ensure that this device provides an airtight environment is a costly task. ^b
<i>Effluent collection</i>	<ul style="list-style-type: none"> • There is no consensus among authors about which option is better, i.e. gutters with weirs or submerged perforated pipes. ^{a,b,d}
<i>Scum formation and removal</i>	<ul style="list-style-type: none"> • The scum accumulates in the settling compartment and under the gas hoods. The need for installation of devices for removal of this material is still discussed. ^{a,b,d}
<i>Sludge withdrawal and sample extraction</i>	<ul style="list-style-type: none"> • In units devoid of sampling points, biomass growth within the UASB can not be monitored; ^{a,b,c,d} • If the sludge withdrawal is not done regularly, the accumulation of inert solids (e.g. sand, hair, plastic, textiles etc.) will cause reduction of useful volume of the reactor; • The sludge collected by pipes installed alongside the walls of the reactor does not represent accurately the mean values assigned to other portions of biomass established on the same level from which the sample was extracted; ^d • As the concentration of decomposing micro-organisms varies along the height of the reactor, units provided with sludge discharge system installed over a single level does not offer alternatives concerning the retention of a good sludge over another, with poorer quality. ^d
<i>Cover / corrosion and odour emissions</i>	<ul style="list-style-type: none"> • The hydrogen sulphide, a biogas constituent, is the main responsible for corrosion and bad odour problems. ^{a,b,c,d}

(Continued on next page)

Table 3.5 (continuation) - Problematic aspects associated with the use of UASB reactors treating domestic wastewater.

Problem / Disadvantage	Description
<i>Cover / corrosion and odour emissions</i>	<ul style="list-style-type: none"> • The presence of a top cover is not a unanimous recommendation, as it impedes the maintenance of some internal structures. ^d • If “open” reactors offer operational advantages (e.g. visual diagnosis of effluent, scum management, unclogging of the effluent collection devices etc.), it also brings as their main inconvenience the higher gaseous emissions. ^{a,b,d}
<i>Biogas</i>	<ul style="list-style-type: none"> • Most of the systems does not make use of technologies for biogas utilization. ^{a,b,c,d}
<i>Methane loss within the effluent</i>	<ul style="list-style-type: none"> • A fraction of between 20 and 40% of the methane produced in UASB applied in the treatment of domestic sewage leaves the system dissolved in the effluent; ^c • The loss of methane has two negative effects: (i) energetic potential is lost and (ii) methane is a greenhouse gas (GHG) that, at some point after leaving the unit, will be released to the atmosphere. ^{b,c}
<i>Necessity for post-treatment</i>	<ul style="list-style-type: none"> • The effluent has lower quality compared to aerobic processes and polishing is usually required in order to achieve the discharge standards for organic matter, nutrients and pathogens. ^{a,b,d,e}
<i>Start-up procedure</i>	<ul style="list-style-type: none"> • Due to the low growth rate presented by methanogenic micro-organisms, the time required for start-up is longer than the aerobic process (when inoculum of good quality is not available). ^{b,d,e}

Sources: ^[a] MIKI (2010); ^[b] VAN LIER ET AL. (2010); ^[c] HEFFERNAN ET AL. (2010); ^[d] VAN HAANDEL & VAN DER LUBBE (2012); ^[e] CHONG ET AL. (2012).

3.1.4.4 Mathematical modelling of the UASB reactor

Anaerobic digestion modelling is a developing “tool” for assessing wastewater treatment for design, operational analysis and control. Published modelling studies for anaerobic treatment of sewage are limited in quantity and most have a relatively low structural complexity (with first order kinetics) as compared to the commonly used Monod kinetics. The types of biochemical models applied in anaerobic systems are: (i) steady-state models, (ii) with one and two step dynamic models, and (iii) fully structured dynamic models. These models, additionally, can be used in conjunction with hydraulic or biofilm variants. Another important group of models are those related to the physicochemical processes, which are to evaluate gas transfer and pH (BATSTONE, 2006).

According to Rodríguez-Gómez (2011), models include many parameters that must be determined or adopted. The development of models for this particular technology is a difficult task, due to the interdependence of processes involved and the bond between operational parameters and reactor characteristics (e.g. internal configuration, environment etc.). The basic elements for developing models for anaerobic digestion are presented in Table 3.6. Occasional studies were performed in order to better understand and evaluate the behaviour of the UASB reactor, the main features of some proposed models are presented in Table 3.7; The standardized nomenclature of the parameters and variables used in the models is presented in Table 3.8.

Table 3.6 – Basic model elements for anaerobic processes.

Elements	Description
<i>Mass balance</i>	<ul style="list-style-type: none"> It is the basis of all mathematical models, as it describes accumulation and reaction within a system in relation to flow across the system boundaries.
<i>Kinetics</i>	<ul style="list-style-type: none"> A critical part of the mass balance is the rate equation, as the other terms are fixed by hydraulics. There are a wide range of kinetic relationships used, but the most popular relationship to describe biological activity is the Monod kinetic, that can be modified to include a wide range of inhibition and regulation mechanisms as required. In sewage treatment, wherein the non-linearity prevails, a more complex kinetic should be applied for the description of the system.
<i>Physicochemical components and temperature</i>	<ul style="list-style-type: none"> The basic elements of the physicochemical system are gas flow calculation, and pH calculation. All methanogenic anaerobic models need to describe gas flow in some form. Complicated, structural models generally also include pH calculation because of its strong impact on carbon dioxide transfer and mineral solids precipitation, and biological conversions. Gas stripping is very important for predicting methane in the effluent, and product loss. Methane, and gas solubility in general increases with decreasing temperature, and this has an impact on methane release, and inhibition by soluble gases such as hydrogen sulphide and hydrogen. Diffusivity of gases decreases, and viscosity of liquids increase with increasing temperature, which means that the gas-liquid transfer can decrease significantly at lower temperatures.
<i>Biochemical structure</i>	<ul style="list-style-type: none"> The processes involved in every step of anaerobic digestion, i.e. hydrolysis, acidogenesis, acetogenesis and methanogenesis are often simplified (sometimes neglected) in the models.
<i>Hydrodynamics</i>	<ul style="list-style-type: none"> It is one of the key aspects associated with the design of anaerobic system, it has impact on the retention of solids and dispersion of the substrate, for example. Given the wide range of reactor configurations, analysis should be made case by case.

Source: BATSTONE (2006) – adapted

Table 3.7 – Some mathematical models developed to study the UASB reactor.

Reference	Description / Characteristics	Main equations
van der Meer and Heertjes, 1983	<ul style="list-style-type: none"> Using fatty acids as the main contaminant, the authors set out to describe the dynamic behaviour and distribution of both fluid and sludge in the reactor, as well as the quantification of the kinetics (using Monod equation for biological growth) of the anaerobic digestion and of the formation of sludge and methane; The sludge bed and sludge blanket were considered as separated CSTRs, while the settler was represented by a plug flow region. A bypassing flow of a parcel of the influent towards the sludge blanket was assumed. Bed and blanket were considered to be interconnected regions by the occurrence of reciprocal streams. Model of the fluid flow: <ul style="list-style-type: none"> Model for anaerobic digestion in an upflow reactor:	<ul style="list-style-type: none"> Substrate mass balance over the sludge bed: $V_{bd} \frac{dC_{sub,bd}}{dt} = Q_{bd} C_{sub,0} + Q_{bl-bd} C_{sub,bl} - Q_{bl} C_{sub,bd} - q_{bd} C_{sol,bd} V_{bd}$ Substrate mass balance in the sludge blanket: $V_{bl} \frac{dC_{sub,bl}}{dt} = Q_{bp} C_{sub,0} + Q_{bl} C_{sub,bd} - Q_{bl-bd} C_{sub,bl} - Q_0 C_{sub,bl} - q_{bl} C_{sol,bl} V_{bl}$ Total production of biogas (gas flow rate): $\phi_g = \frac{r_{CH4,bd} V_{bd} + r_{CH4,bl} V_{bl}}{C_{CH4}}$ Sludge production in the sludge bed: $V_{bd} \frac{dC_{ap,bd}}{dt} = Q_{bl-bd} C_{ap,bl} - Q_{bl} C_{ap,bd} + [k_{ap} (C_{sub,0} - C_{sub,bd})] V_{bd}$ Sludge production in the sludge blanket: $V_{bl} \frac{dC_{ap,bl}}{dt} = Q_{bl} C_{ap,bd} - Q_{bl-bd} C_{ap,bl} + [k_{ap} (C_{sub,0} - C_{sub,bl})] V_{bl}$ Specific organic load converted in the reactor: $\phi_{sub}^m = \frac{Q_0 \xi_{sub} C_{sub,0}}{V_R} \quad (\text{Continued on next page})$

Table 3.7 (continuation) – Some mathematical models developed to study the UASB reactor.

References	Description / Characteristics	Main equations
Bolle et al., 1986a	<ul style="list-style-type: none"> This model discussed the modelling of the fluid flow; Both sludge bed and blanket were described as CSTRs with short-circuiting flows (i.e. by-pass). The settler, according to the authors, act like a plug flow region; Schematic representation of the model:	<ul style="list-style-type: none"> Tracer mass balance over the sludge bed: $V_{bd} \frac{dC_{tr,bd}}{dt} = (1 - SF_{bl}) Q_0 (C_{tr,0} - C_{tr,bd})$ Tracer mass balance over the sludge bed: $V_{bl} \frac{dC_{tr,bl}}{dt} = (1 - SF_{bl}) Q_0 C_{tr,bd} + (SF_{bl} - SF_{st}) Q_0 C_{tr,0} - (1 - SF_{st}) Q_0 C_{tr,bl}$
Bolle et al., 1986b	<ul style="list-style-type: none"> A dynamic model was proposed in order to integrate the fluid flow pattern (the same presented by Bolle et al. 1986a), the kinetic behaviour and the mass transport phenomena. 	<ul style="list-style-type: none"> Substrate mass balance over the sludge bed: $\frac{dM_{s,bd}}{dt} = \left[(1 - SF_{bd}) \cdot \phi_{in} \cdot (C_{s,in} - C_{s,bd}) \right] - \frac{\mu_{bd}^A}{Y_{x,s}} C_{x,bd}$ Substrate mass balance over the blanket: $\frac{dM_{s,bl}}{dt} = (1 - SF_{bd}) \cdot \phi_{in} \cdot C_{s,bd} + (SF_{bd} - SF_{bl}) \cdot \phi_{in} \cdot C_{s,in} - (1 - SF_{bl}) \cdot \phi_{in} \cdot C_{s,bd} - r_{s,bl} \cdot V_{bl}$ <p style="text-align: right;">(Continued on next page)</p>

Table 3.7 (continuation) – Some mathematical models developed to study the UASB reactor.

References	Description / Characteristics	Main equations
Wu and Hickey, 1997	<ul style="list-style-type: none">• The UASB was divided into a reaction zone (CSTR with dead volume and bypass flow from the inlet to the outlet of the system) followed by a settler (plug flow reactor with dispersion);• Besides de hydrodynamic aspect, the model included biological reaction kinetics and mass transfer within anaerobic granules;• Representation of the hydrodynamic model:	<ul style="list-style-type: none">• Substrate mass balance in the reaction zone with granular bed:$V_{\text{rct}} \frac{dC_{\text{sub},0}}{dt} = V_{\text{rct}} E(t) - Q_0 C_{\text{sub},0}(t) - K_{\text{mt,gr}} A_{\text{gr}} \cdot [C_{\text{sub},0} - C_{\text{sub,gr}}(0,t)]$• Substrate mass balance in the settler:$\frac{\partial S_b}{\partial t} = \frac{D_p}{L^2} \frac{\partial^2 S_d}{\partial Z^2} - \frac{u}{L} \frac{\partial S_d}{\partial Z}, t > 0, 0 < Z < 1$
Batstone et al., 2002	<ul style="list-style-type: none">• Proposed by the International Water Association (IWA) Anaerobic Digestion Modelling Task Group with the objective of developing a generalised model, the “Anaerobic Digestion Model” (ADM1);• The model includes multiple steps (and variables) to describe the biochemical and physicochemical processes;• Schematic representation of the model:	

Table 3.7 (continuation) – Some mathematical models developed to study the UASB reactor.

References	Description / Characteristics	Main equations
Zhao <i>et al.</i> , 2010	<ul style="list-style-type: none"> By introducing a linear phenomenological equation into the ADM1 (Batstone <i>et al.</i>, 2002), the authors developed a model to describe the dynamic behaviour of the startup of anaerobic reactors. 	<ul style="list-style-type: none"> Mass balance for the liquid phase: $V_{\text{rct}} \frac{dC_i}{dt} = Q_0 C_{\text{in},i} - Q_F C_{\text{out},i} + \sum_{j=1-15} \rho_j V_{i,j}$ Mass balance for the solid phase: $V_{\text{rct}} \frac{dX_i}{dt} = -Q_F C_{\text{out},i} + \sum_{j=1-15} \rho_j V_{i,j}$ Mass balance for the gaseous phase: $V_{\text{gas}} \frac{dC_{\text{gas},i}}{dt} = -\phi g C_{\text{gas},j} + V_{\text{rct}} \rho_{T,i}$
Rodríguez-Gómez, Moreno & Liu, 2013	<ul style="list-style-type: none"> A model describing the main processes occurring in the UASB reactor (i.e. substrate degradation, biomass growth and the reactions that take place within the granules) was developed and, in order to “maintain simplicity and applicability of the model” (according to the authors), only the fundamental aspects were considered; The reactor was divided in several CSTRs, coupled in series; Representation of this simplified model: 	<ul style="list-style-type: none"> Mass balance of substrate for each CSTR: $\frac{dC_{\text{sub},i}}{dt} = \frac{Q_0}{V_i} (C_{\text{sub},i-1} - C_{\text{sub},i}) - K_i C_{\text{sub},i}$ Mass balance of active biomass in each CSTR: $\frac{dC_{\text{sol},i}}{dt} = K_i C_{\text{sub},i} Y - K_d C_{\text{sol},i}$ Mass balance of inactive biomass for each of the CSTRs: $\frac{dE_i}{dt} = K_d C_{\text{sol},i}$

Table 3.8 – Standardized nomenclature of parameters and variables used in the models presented in Table 3.7.

Parameter / Variable	Nomenclature	Unit
Biogas flow rate	$\dot{Q}_g$	$\text{m}^3 \cdot \text{h}^{-1}$
Biogas flow rate in the bed	$\dot{Q}_{g,bd}$	$\text{m}^3 \cdot \text{h}^{-1}$
Biogas flow rate in the blanket	$\dot{Q}_{g,bl}$	$\text{m}^3 \cdot \text{h}^{-1}$
By-pass flow rate	Q_{bp}	$\text{m}^3 \cdot \text{h}^{-1}$
Concentration of effluent component	$C_{out,i}$	$\text{kg} \cdot \text{m}^{-3}$
Concentration of effluent particulate component	$X_{out,i}$	$\text{kg} \cdot \text{m}^{-3}$
Concentration of gas component	$C_{g,i}$	$\text{kg} \cdot \text{m}^{-3}$
Concentration of inactive biomass	E_i	$\text{kg} \cdot \text{m}^{-3}$
Concentration of influent component	$C_{in,i}$	$\text{kg} \cdot \text{m}^{-3}$
Concentration of influent particulate component	$X_{in,i}$	$\text{kg} \cdot \text{m}^{-3}$
Concentration of methane (in the biogas)	C_{CH_4}	$\text{kg} \cdot \text{m}^{-3}$
Concentration of substrate in the bed	$C_{sub,bd}$	$\text{kg} \cdot \text{m}^{-3}$
Concentration of substrate in the blanket	$C_{sub,bl}$	$\text{kg} \cdot \text{m}^{-3}$
Concentration of substrate in the settler	$C_{sub,st}$	$\text{kg} \cdot \text{m}^{-3}$
Concentration of unidentified products in the bed	$C_{ap,bd}$	$\text{kg} \cdot \text{m}^{-3}$
Concentration of unidentified products in the blanket	$C_{ap,bl}$	$\text{kg} \cdot \text{m}^{-3}$
Constant of unidentified products	k_{ap}	–
Conversion rate of unidentified products in the bed	$r_{ap,bd}$	$\text{kg} \cdot \text{m}^{-3} \cdot \text{s}^{-1}$
Conversion rate of unidentified products in the blanket	$r_{ap,bl}$	$\text{kg} \cdot \text{m}^{-3} \cdot \text{s}^{-1}$
Dead volume	V_d	m^3
Decay constant	K_d	h^{-1}
Dispersion coefficient of liquid within the settler	D_{st}	$\text{mm}^2 \cdot \text{s}^{-1}$
Flow from sludge blanket to bed	Q_{bl-bd}	$\text{m}^3 \cdot \text{h}^{-1}$
Flow velocity in the settler	u_{st}	$\text{mm} \cdot \text{s}^{-1}$
Influent flow rate	Q_0	$\text{m}^3 \cdot \text{h}^{-1}$
Influent tracer concentration	$C_{tr,0}$	$\text{kg} \cdot \text{m}^{-3}$
Initial concentration of substrate	$C_{sub,0}$	$\text{kg} \cdot \text{m}^{-3}$
Mass transfer coefficient for the granules	$K_{mt,gr}$	$\text{mm} \cdot \text{s}^{-1}$
Methane production in the bed	$r_{CH_4,bd}$	$\text{kg} \cdot \text{m}^{-3} \cdot \text{s}^{-1}$

(Continued on next page)

Table 3.8 (continuation) – Standardized nomenclature of parameters and variables used in the models presented in Table 3.7.

Parameter / Variable	Nomenclature	Unit
Methane production in the blanket	$r_{CH_4,bl}$	$kg.m^{-3}.s^{-1}$
Outflow rate	Q_F	$m^3.h^{-1}$
Pressure of gas	$p_{g,i}$	bar
Rate coefficient for component i on process j	$v_{i,j}$	$kg.m^{-3}$
Reaction zone inflow rate	Q_{rct}	$m^3.h^{-1}$
Reactor total volume	V_R	m^3
Settler Length	L_{st}	mm
Settler volume	V_{st}	m^3
Short-circuiting flow to the blanket (fraction of the Q_0)	SF_{bl}	%
Short-circuiting flow to the settler (fraction of the Q_0)	SF_{st}	%
Sludge bed inflow rate	Q_{bd}	$m^3.h^{-1}$
Sludge bed volume	V_{bd}	m^3
Sludge blanket inflow rate	Q_{bl}	$m^3.h^{-1}$
Sludge blanket volume	V_{bl}	m^3
Sludge concentration in the bed	$C_{sol,bd}$	$kg.m^{-3}$
Sludge concentration in the blanket	$C_{sol,bl}$	$kg.m^{-3}$
Specific conversion of substrate	$\varnothing_{sub}^m$	$kg.m^{-3}.s^{-1}$
Specific substrate conversion rate in the bed	q_{bd}	s^{-1}
Specific substrate conversion rate in the blanket	q_{bl}	s^{-1}
Substrate concentration with the granules	$C_{sub,gr}$	$kg.m^{-3}$
Substrate conversion rate	ξ_{sub}	–
Substrate conversion rate in the sludge bed	$r_{sub,bd}$	$kg.m^{-3}.s^{-1}$
Substrate conversion rate in the sludge blanket	$r_{sub,bl}$	$kg.m^{-3}.s^{-1}$
Substrate conversion rate within the granule	q_{gr}	$kg.m^{-3}.s^{-1}$
Total granule surface area	A_{gr}	mm^2
Tracer concentration in the sludge bed	$C_{tr,bd}$	$kg.m^{-3}$
Tracer concentration in the sludge blanket	$C_{tr,bl}$	$kg.m^{-3}$
Vertical coordinate	Z	–
Volume of the reaction zone	V_{rct}	m^3
Yield	Y	–

3.1.4.5 Recently proposed configurations based on the UASB reactor

In the last decades an increasing number of studies directed at improving anaerobic systems have been observed. Recent proposals, all based on the UASB system proposed by Lettinga are presented in Table 3.9. Some variants of these systems are also schematically shown in Figure 3.17.

Table 3.9 – Proposed UASB variants applied to the anaerobic wastewater treatment.

Tecnology	Description	Reference
<i>Hybrid anaerobic reactor (1)</i>	Possibly the first proposed modification to the UASB reactor. The GLS separator was replaced by a layer composed by glass beads, in an attempt to incorporate the advantages of the UASB and the anaerobic filter into a single unit.	MAXHAM & WAKAMYIA, 1981
<i>UBF – Upflow blanket filter</i>	The upper region of the reactor was filled with floating plastic rings (contained by a screen) in an attempt to retain the biomass (Figure 3.17c).	GUIOT & VAN DER BERG, 1984
<i>Hybrid anaerobic reactor (2)</i>	Polyurethane plates were installed in the upper region of the unit; an average COD removal efficiency of 64% was achieved treating pre-settled domestic sewage.	ELMITWALLI et al., 1999
<i>UASB - septic tank</i>	A variant concept of the UASB reactor was proposed as an alternative to the septic tank (Figure 3.17e).	COELHO et al., 2003
<i>Hybrid UASB-filter reactor</i>	The GLS separator was replaced by plastic rings and operated at low temperatures (10 to 14°C); it presented worse performance compared with the UASB (with respect to COD and TSS removal).	LEW et al., 2004
<i>UAHBmod - Hybrid modified anaerobic reactor</i>	The performance of two hybrid reactors was investigated, one of them denominated “UAHBmod”, wherein the support media consisted of eucalyptus pellets.	PASSIG, 2005.
<i>UASB reactor with internal lamella settler</i>	High-rate settlers were tested in two different reactors. In the unit denominated “UASB-ESR1”, such a device was installed in the sedimentation zone, while in the “UASB-ESR2” it was set under the GLS separator (Figures 3.17a e 3.17b)	HALALSHEH et al., 2010
<i>UAFB - Upflow anaerobic fixed bed</i>	Pieces of polyethylene, randomly distributed, were fixed at the bottom of an UASB reactor operated at temperatures between 15 and 35°C.	GAO et al., 2011
<i>Double stage biogas collection UASB reactor</i>	Designed with two biogas collectors, one near the bottom of the unit and another in the upper region of the reactor (Figure 3.17d).	PEREIRA, 2012

Figure 3.17 – Different conceptions proposed as alternatives to the original UASB reactor: (a) and (b) Halalsheh et al., 2010; (c) Guiot and van der Berg, 1984; (d) Pereira, 2012 and; (e) Coelho et al., 2003.

3.1.4.6 UASB technology for sewage treatment: a critical analysis

Contrary to what is generally reported in books and academic works devoted to the subject, the author of this research — as well as his colleagues and many project engineers and operators — believe that, due to the various specifications and recommendations (technical standards, guidelines proposed by the specialized literature etc.), the internal arrangement UASB reactor becomes complex and involves relatively high costs for its construction and maintenance — if the more restrictive standards of lifespan and operational flexibility and maintenance are respected.

Owing to the limited emphasis verified in technical-scientific texts, the degree of complexity attributed to the internal arrangement of all elements that constitute the UASB reactor is still an obscure topic. Published books and articles generally make use of diagrams and other simplified schematic representations to illustrate the system (e.g. Figure 3.12), not allowing readers a detailed view of the unit. Few, too, are the works which present detailed photographic record of the inner region of the UASB. In order to demonstrate his point of view, the author of this text chose to prepare detailed schematic illustrations, in which the internal composition of the reactor is shown (Figures 3.18 and 3.19).

Figure 3.18 – Example of the complex internal arrangement of an UASB reactor (variant 1): (a) external layout and; (b) internal arrangement.

Figure 3.19 – Example of the complex internal arrangement of an UASB reactor (variant 2):(a) external layout and; (b) internal arrangement.

According to van Lier et al. (2010), in the last three decades the design of the UASB reactor was, in certain way, “standardized”, based on the 64 m³ pilot plant of installed in Cali (Colombia), following recommendations made by van Haandel and Lettinga (1994). Although many studies and surveys associated with the operation have been carried out, the advancement of knowledge regarding the microbiology processes involved in anaerobic digestion contributed little to the improvement of the reactor configuration and to the development of new operational strategies; the design of UASB reactor follows virtually unchanged^[1]. Mahmoud et al. (2003) argue that, while biological processes related to the use of UASB are widely investigated, the prevalent physical phenomena and physical-chemical mechanisms of contaminant removal are rarely studied

Regarding its application in sewage treatment systems, the slow enhancement of UASB technology can be attributed to its original proposition; the system was primarily planned to assist developing countries, appointed as a simple and low-cost technology, able to provide minimum sanitary conditions for effluent discharge. To minimize electricity consumption, maintenance requirements and considering the unavailability of skilled labour in countries where resources are scarce, the design of a treatment plant was based solely on pumping the wastewater up to its highest point (according to the WWTP hydraulic profile) and its downstream flow – by gravity – through the other units of the system. Under such conditions, the design of the UASB reactor tends to become complicated (adapted from VAN HAANDEL & VAN DER LUBBE, 2012).

Another fact, also relevant, is that the reaction zone (i.e. sludge bed/blanket) in the UASB reactor usually occupies — “only” — about 50% of its working volume (Mahmoud, 2008). The need to install the GLS separators above the sludge blanket reduces the volumetric (reactional) capacity of the reactor. As illustrated in Figure 3.20, conceptually, there is an intimate relationship between the profile of the suspended solids concentration (biomass) — established as a function of the reactor depth — and the COD removal rate^[2]. According Lettinga et al. (1980), the presence of the different types of sludge in the blanket is reflected in the solids concentration profiles determined along the height of the reactor (i.e. stratification). Despite being composed of several distinct “layers,” the sludge blanket shall be regarded as a dynamic cohesive unit, even if it extends to the settling compartment.

¹ Simple alterations in civil and hydraulic aspects are observed (as presented in Sections 3.1.4.2 and 3.1.4.5).

² The higher the suspended solids content in a particular volume, the greater is the COD removal rate observed in its interior.

Figure 3.20 – Schematic illustration of the variation of the suspended solids concentration and COD removal rate along the height of the UASB reactor (adapted from LETTINGA et al., 1980; NARNOLI & MEHROTRA, 1997).

As can be observed in Figure 3.20, almost all degradation of influent organic matter occurs in the lower section of the reactor, precisely in the volume occupied by the sludge blanket. The region occupied by the GLS separators presents weak performance in the COD removal because, unlike the reaction zone, the decrease regarding this parameter is almost entirely associated with the sedimentation of solids (i.e. the anaerobic conversion process is practically non-existent).

An important feature of the UASB reactor is the expandability of the sludge blanket, as its characteristics may be related to the ability of the system to retain sludge. Indeed, the hydrodynamic behaviour of the sludge blanket is still unclear, and its appropriate height have been determined by trial and error (LEITÃO et al., 2011).

Project guidelines for treatment systems are typically based on practical data, obtained in research and operation of full-scale units. Currently, there is no safe basis expressed in mathematical models for this purpose. Most engineers, during the design phase, apply the θ_h parameter to determine the total volume of each unit, without considering the fact that degradation of organic matter occurs only in the digestion compartment. The consequence of this fact is, conceptually, the construction of reactors with larger dimensions to those apparently sufficient; a large portion of the reactor is “sacrificed” for the installation of the GLS reactor. This questioning is a fundamental element of the design proposed in this research.

According to van der Lubbe and van Haandel (2012), even the simplest technology will fail without adequate infrastructure in terms of technical and institutional capacity. In contrast, the UASB reactor will continue to provide partial treatment levels, even when operated precariously. With the rapid social development witnessed by Brazilian, Indian and Chinese citizens, for example, improvements in technical and financial aspects are expected, as well as more restrictive limits related to the discharge standards. So, there is no more need to keep the approach around the UASB reactor in extremely low levels; significant improvements are achievable.

3.1.5 THE APPLICATION OF MECHANICAL MIXING IN ANAEROBIC REACTORS

3.1.5.1 Generalities

The UASB reactor, like most of the anaerobic technologies that preceded it, emerged without the use of equipment or procedures that could cause agitation in the reaction zone, due to the mixing provided by the intense formation of biogas, generated by the relatively high influent organic loadings (mainly from industrial sources). With the application of this technology in systems receiving low organic loads (e.g. sewage treatment), it was found that the movement of the fluid becomes an important parameter in the design and operation of these units (adapted from MICHELAN, 2006).

Given the scarcity of data and information concerning the implementation of agitation systems in reactors used for wastewater treatment, it was decided to include in the text information presented in research devoted to mixing in sludge digesters.

Conventional anaerobic sludge digesters can be stirred by gas injection, liquid jets — via pumping/recirculation, pipes with orifices, nozzles etc. — and through the use of mechanic mixers. The selection of these systems is based on installation and operational costs, maintenance requirements, process configuration and the presence of solids and scum of the influent, besides the personal experience of the consultant engineer (adapted from VESILIND, 2003).

In reactors equipped with “complementary” mixing systems, majority of the active organic matter decomposition occurs in that volume which is agitated, and, reciprocally, the quiescent zones remain stagnant and the micro-organisms contained therein are not participating in the digestion process. Ideally, the fluid flow must approach the complete mixing condition, without the occurrence of dead zones (TENNEY & BUDZIN, 1972), but practically the effect of agitation on the performance is still uncertain (TERASHIMA et al. 2009).

Karim et al. (2005) state that two important aspects of agitation are its intensity and duration. The information available in the literature regarding operational strategies, however, is contradictory. Lemmer, Naegle and Sondermann (2013) indicate that, due to the difficulty of assessing the multiphase process of synthesis of the substrates, most research related to mixing in anaerobic systems are focused on their effects over biogas production.

There are researchers who advocate the use of intermittent agitation, rather than continuous mixing (DAGUE, MCKINNEY & PFEFFER, 1970^[1]; MILLS, 1979; SMITH, HEIN & GREINIER, 1979). Other authors, like Hashimoto (1982), witnessed higher gas production in continuously stirred systems than in those who underwent intentional mixing interruptions. The impact of the mixing was also evaluated (in bench scale) by Karim et al. (2005), who demonstrated that mechanical, hydraulic and pneumatic agitation accounted for increased biogas production by 29%, 22% and 15%, respectively, in comparison to a non-agitated digester. Ben-Hasson, Ghaly and Singh (1985) observed a lower biogas production (75%) in a digester operated under continuous mixing, in comparison to a unstirred unit; Chen, Chynoweth and Biljetina (1990) also observed — using a digester with volume of 4.5 m³ — an increased gas production in the period in which the unit was not agitated. Laboratory studies applying sewage as substrate (STAFFORD, 1982; STROOT et al., 2001) showed that, in agitated continuously systems, increasing rotational speeds of the impeller did not cause an increase in gas production, however, a slight decrease was observed. Ong, Greenfield and Pullammanappallil (2002),

¹ *Dague and his co-workers (1970) attributed the superior performance presented by the system operated under intermittent agitation regime to the better solids retention caused by bioflocculation.*

analysing digesters fed with cow dung, observed no difference — in relation to the biogas production — between continuously agitated and intermittent stirred systems.

Lettinga (1981) suggested that “minimized” agitation may benefit anaerobic digestion in reactors containing high solids concentration, because there is evidence that the metabolic activity of biomass improves if they can operate in a quasi-quiescent environment. Stroot et al. (2001) found that a “minimum” mixing intensity was sufficient to adequately distribute the substrate, and suggested that continuous and “vigorous” agitation can impair the performance in digesters operating with high concentration of solids. Also, it may be attributed to the more intense agitation, the fracture of the flock structure established in the reactors, jeopardizing the syntrophic relations between micro-organisms on which the system performance depends (WHITMORE et al, 1987;. DOLFING, 1992).

Some researchers suggests that continuous agitation may not be necessary, because anaerobic micro-organisms can be more efficient in stratified environment, in which they form associations able to convert, quickly and effectively, organic matter into biogas (BRIDGEMAN, 2012). This stratification could be used as an operational advantage to retain solids larger than the average θ_h , maintaining a greater proportion of volatile solids – for more intense degradation – in the reactor (ONG, GREENFIELD & PULLAMMANAPPALLIL, 2000; KAPARAJU & ANGELIDAKI, 2007).

Despite the undeniable importance of mixing in the conversion of the substrate, the academic community has not reached the harmonic and solid understanding associated to its effects on the anaerobic digestion process. There is, therefore, the need for further research aimed at developing strategies for agitation in regard to its optimal intensity and duration (KAPARAJU et al., 2008).

3.1.5.2 The velocity gradient

In the water and wastewater treatment systems, the degree of mixing is usually measured by the velocity gradient (G). The higher the value of G, the more intense turbulence of the fluid is and, consequently, stronger are the shear forces acting on the flocks. The velocity gradient is a function of power dissipation in a given unit volume of liquid (MAHMOUD, 2002).

Camp and Stein (1943), the creators of the “velocity gradient” concept^[1], considered G as being the angular distortion of an elementary volume of water resulting from the application of tangential surface forces, expressed algebraically as (CLEASBY, 1984; DUCOSTE & CLARK, 1998; SINDALL, BRIDGEMAN & CARLIELL-MARQUET, 2013):

$$G = \sqrt{\frac{P}{\mu \cdot V}} \quad (\text{Equation 3.4})$$

in which::

- G: mean velocity gradient [s⁻¹];
- P: power applied in the liquid mass for mixing [N.m.s⁻¹];
- μ : dynamic viscosity of the fluid [N.s.m⁻²];
- V: effective volume of the reactor or tank [m³].

Theoretically, G can be calculated at any point in a tank, since the power dissipation in that place is known. In practice, the energy dissipation as well as other flow characteristics, varies across the vessel, turning the velocity gradient a function of time and location (SINDALL, BRIDGEMAN & CARLIELL-MARQUET, 2013).

In spite of being commonly used in the design of mixing tanks, it is argued that the G concept is reputed as flawed, since it is proposed to represent a complex flow field by using a single value (CLARK, 1985; KRAMER & CLARK, 1997). The distribution of velocity gradients and power inputs within a digester are not uniform, as areas of higher power input (e.g. close to impeller), are likely to experience intense turbulence and hence, velocity gradients several orders of magnitude greater than areas of lower turbulence. Although a useful parameter to approximate overall conditions, the mean velocity gradient does not sufficiently describe the fluctuations regarding the different mixing intensities established within a vessel. (SINDALL, BRIDGEMAN & CARLIELL-MARQUET, 2013). Despite this observations, it was not yet introduced in the technical literature a more accurate and practical parameter for evaluating the intensity of mixing. The optimal G values are rarely reported in the literature. WEF (1987) and Vesilind (2003) recommend velocity gradients ranging between 50 and 80 s⁻¹ for anaerobic

¹ The “G” must be considered as a mean value, since the dissipation of the power applied to the liquid mass (among other factors) does not occur homogeneously.

digesters applied exclusively on sludge stabilization. Recent studies^[1] conducted at the University of Birmingham, in England, used computational fluid dynamics (CFD) tools to suggest that lower G values are sufficient to promote most suitable production of biogas.

3.2 FUNDAMENTALS OF SOLID-LIQUID MIXING IN AGITATED VESSELS

According Oldshue (1983), the design of a mixing system includes some basic considerations. The first is the effect of the tank (i.e. reactor) in the process: its geometry may dictate the selection of the mixer and its performance. Another aspect to note is the type of impeller to be used to achieve a certain objective.

There are two main types of flow that can be developed by the action of impellers: axial and radial. The radial flow impellers move the fluid toward the tank wall in horizontal (or radial) direction. Axial flow impellers produce a flow parallel to the mixer shaft, from which the liquid moves upward along the wall of the tank and downward, when moving towards the impeller. Axial flow impellers are generally more efficient than the radial, as they produce more fluid flow per unit of applied power. Regardless of the type of flow, the installation of internal baffles along the walls of the tank — vertically — is crucial to avoid vortex formation, a detrimental phenomenon to the quality of the mixture (OLDSHUE, 1983). The flow patterns of a fluid when subjected to axial and radial regimes are shown in Figure 3.21.

Figure 3.21 – Typical flow patterns associated with the mechanical mixing of fluids: (a) radial and (b) axial (modified from OLDSHUE, 1983).

¹ SINDALL, BRIDGEMAN & CARLIELL-MARQUET, 2013.

Despite the presence of gas, formed from the anaerobic digestion and capable of influencing the suspension of solids, the goal of employing mechanical mixers in a reactor fits what the literature classifies as “solid-liquid” mixture; its main objectives are (ATIEMO-OBENG, PENNEY & ARMENANTE, 2003):

- Suspension of solids;
- Resuspension of settled solids;
- Incorporation of floating solids;
- Dispersion of solid aggregates or control of particle size from the action of fluid shear as well as any abrasion due to impacts between particles and against the impeller;
- Mass transfer across the solid–liquid interface.

3.2.1 BASIC HYDRODYNAMIC CONCEPTS OF SOLID-LIQUID SUSPENSIONS

The suspension of solids contained in a given volume of liquid requires the introduction of mechanical energy by some mode of agitation. The energy input creates a turbulent flow field in which solid particles are lifted from the bottom of the tank and subsequently dispersed and distributed throughout liquid mass. The capture of solids from the base of the reactor is accomplished by the combination of the drag and moving fluid on the particles and the bursts of turbulent vortices originating from the bulk flow in the vessel. In the displacement of mass within the reactor. For small solid particles whose density is approximately equal to that of the fluid, once in suspension, they continue to move with the liquid mass. In this case, the suspension behaves like a single liquid phase, at low solids concentrations. For denser particles, its velocities will be different from that imposed on the liquid. The drag force on the particles, caused by the fluid motion must be sufficient and directed upward to counteract the sedimentation tendency presented by these solids, caused by the action of gravity (ATIEMO-OBENG, PENNEY & ARMENANTE, 2003).

The properties of both the liquid and the solid particles influence the hydrodynamic aspect and thus the suspension. Similarly, the reactor geometry and agitation parameters are also essential. Important considerations related to the design of mixing systems are presented in Table 3.10. Some of the terminology and main dimensions considered for the design of mechanical mixing systems are shown in Figure 3.22.

Table 3.10 – Important considerations regarding mixing systems for solid suspension.

Aspects	Parameters / Considerations
<i>Physical properties of the solid</i>	<ul style="list-style-type: none"> • Specific mass, ρ_s [kg.m^{-3}] • Particle size, d_p [m]
<i>Physical properties of the liquid</i>	<ul style="list-style-type: none"> • Specific mass, ρ_L [kg.m^{-3}] • Viscosity, μ_L [Pa.s]
<i>Operating conditions of the process</i>	<ul style="list-style-type: none"> • Effective height of the tank, Z_T [m] • Concentration of solids in the fluid, X [$\text{kg solids} \cdot (\text{kg liquid})^{-1}$] • Presence or absence of gas
<i>Geometry</i>	<ul style="list-style-type: none"> • Diameter of the tank, D_T (m) • Geometry of the tank bottom • Impeller type and geometry • Impeller diameter, D_{imp} (m) • Distance between the bottom of the tank and the impeller, C (m) • Height of the liquid above the impeller, CV (m) • Type and geometry of the internal baffles
<i>Agitation conditions</i>	<ul style="list-style-type: none"> • Impeller rotational velocity, n (rpm) • Motor power, P_M (W)^[1] • Velocity at the impeller extremity, n' (m.s^{-1}) • Degree of suspension required • Liquid flow pattern • Distribution of mixing intensity within the tank

Sources: OLDSHUE, HERBST & POST, 1990; ATIEMO-OBENG, PENNEY & ARMENANTE, 2003.

According Oldshue, Herbst and Post (1990), so that the suspension of solids occurs is necessary to employ an impeller capable to promote more flow (i.e. displacement of the liquid mass) than shear. Therefore, for this application, axial flow impellers are usually applied, since they are more efficient “pumping” devices. At constant power, the larger the diameter of an impeller, the greater the liquid movement provided by it and the lower its rotational velocity within the mixing system. Despite this fact, in principle, there are practical limits regarding this parameter. Moreover, at constant power, when the impeller diameter is relatively large and its rotational velocity decreases, the torque increases, necessarily requiring the installation of a more powerful agitation equipment.

¹ $1 \text{ W} = 1 \text{ N.m.s}^{-1}$.

Figure 3.22 – Schematic representation of a mechanically agitated tank and some of its most important dimensions (adapted from OLDSHUE, 1983).

Obviously, the solid particles will settle or float depending on the differences between their properties and the characteristics of the liquid phase. The most important aspect is associated to the fact of possible agglomeration of the solids as they settle, and how easy it is to resuspend them in the moment when the agitation is increased or restored (ATIEMO-OBENG, PENNEY & ARMENANTE, 2003).

According Atiemo-Obeng, Armenante and Penney (2003), most mixing operations occur at higher velocities than the minimum required for suspension. An agitation under higher velocities increases the degree of suspension, the mass transfer rates and shearing forces, which may cause undesired friction between the particles. The degree of solids suspension in stirred tanks is generally classified into three levels, as follows (Figure 3.23):

- **Partial suspension:** characterized by the visual observation of the complete motion of all solid particles over the bottom of the tank; it excludes the formation of “fillets”, a non-cohesive particle aggregation in the corners or bottom of the vessel. Since particles are in constant contact with the base of the tank, not all of their surface area

is available for chemical reactions or mass/heat transfer;

- **Complete suspension:** characterized by the complete movement of all particles with no particle remaining over the base of the reactor for more than 1 to 2 seconds. This condition is known as “Zwietering criterion”. Under this condition, the maximum surface area of the particles is exposed to the fluid by the use of minimal agitation conditions at which all particles reach complete suspension.
- **Uniform Suspension:** corresponds to the state at which the solids concentration and the size distribution of the particles are essentially uniform throughout the tank; any subsequent increase in stirring velocity does not appreciably increase the distribution of solids in the fluid. A uniformity of 95% is often considered adequate for most applications, since the value of 100% is impractical (there is a fluid layer – a few centimetres thick – over the surface, where the particle concentration is lower due to the inferior upflow velocity established near the upper region of the vessel).

Figure 3.23 – Typical conditions for suspension of solids: (a) partial suspension, (b) complete suspension and (c) uniform suspension (adapted from LENG et al., 2008).

A summary of recommendations published in the specialized literature, which deal with the geometric relationships and other guidelines for the design of the components of a mixing system, is presented in Table 3.11.

Table 3.11 – Recommendations for the design of agitation systems.

Component	Recommendation	
<i>Tank</i>	Bottom	<ul style="list-style-type: none"> • It has an important effect on the mixing requirements, since its geometry influences the flow patterns responsible for suspension of solids. Preference is given to a rounded shape bottom (i.e. dished head); • Flat-bottom tanks require impeller velocities between 10 and 20% higher than with a rounded bottom, as they are not able to redirect the upward flow and allow the development of dead zones in the corners.
	Other geometric aspects	<ul style="list-style-type: none"> • Walls and the internal arrangement should be planned so that there is no loss in performance relative to the agitation system.
<i>Fluid Inlet</i>		<ul style="list-style-type: none"> • Must be located and directed so that the occurrence of interferences in the mixing system performance is avoided (i.e. the influent jet can not be directed toward the impeller).
<i>Impeller</i>	Quantity	<ul style="list-style-type: none"> • A single impeller is sufficient for complete suspension in tanks with rounded bottom in cases in which $Z_T/D_T < 1.3$; • Two impellers are recommended for tanks with $1.3 < Z_T/D_T < 2.5$ applied for complete suspension; • Three impellers may be required in cases in which $2.5 < Z_T/D_T$. A tank with this geometry is a poor choice for suspended solids; • Systems with multiple impellers generally are not employed for suspension of solids applications.
	Dimensions	<ul style="list-style-type: none"> • The impeller size establishes if the mixing process will be governed by the flow or turbulence. For an equal result, its diameter affects power and torque characteristics. The power and torque values are normalized when $D_{IMP} / D_T = 0.33$.
	Position	<ul style="list-style-type: none"> • For complete suspension, the distance between the impeller shaft and the bottom of the tank – expressed as “C” – should be equal to 0.33 of the C / D_{IMP} ratio.
<i>Vertical baffles</i>	Quantity	<ul style="list-style-type: none"> • For application involving solids suspension, 4 baffles must be installed.
	Dimensions	<ul style="list-style-type: none"> • The distance from the walls must be between $D_T / 72$ and $D_T / 90$.

Sources: OLDSHUE (1983); OLDSHUE, HERBST & POST (1990); ATIEMO-OBENG, PENNEY & ARMENANTE, (2003); LENG et al. (2008).

3.3 HIGH-RATE SETTLING

3.3.1 THE FUNDAMENTALS

In 1904, Hazen^[1] argued that the efficiency of sedimentation tanks mainly depended on the depth of the settler and the surface loading rate, as well as being independent of the hydraulic retention time (CULP et al., 1968). Camp (1946) described the theoretical pattern of discrete particles settling in an “ideal” rectangular sedimentation basin as narrow lines, in which all particles with similar sedimentation velocities follow parallel paths. According to this theory, the velocity of a solid particle that enters the tank is composed of two components (DEGREMONT, 2007):

- v_1 : horizontal velocity of the fluid flow, equal to Q/A (m.h^{-1});
- v_0 : sedimentation velocity (the vertical component), given by Stokes Law (m.h^{-1}).

A particle — whose path is illustrated in Figure 3.24 and that penetrates the tank in critical condition (close to the liquid surface level) — will be retained in the settler when the time required for it to reach the bottom of the vessel (or the top of the sludge deposition zone), t_1 , is lower than the time taken by the liquid to pass through the sedimentation basin (horizontal flow), t_2 .

Figure 3.24 – Scheme presenting the discrete sedimentation in a rectangular horizontal flow tank (adapted from DEGREMONT, 2007).

¹ HAZEN, A. (1904). On sedimentation. *Transactions of the American Society of Civil Engineers*, v. 53, pp. 45 – 71.

$$t_1 = \frac{H}{v_0} \quad (\text{Equation 3.5})$$

$$t_2 = \frac{L}{v_1} = \frac{L \cdot A_v}{Q} \quad (\text{Equation 3.6})$$

in which:

- H: height of the tank [m];
- L: length of the tank [m];
- A_v : cross-sectional area of the tank [m²];
- Q: flow rate [m³.h⁻¹].

Thus, when t_1 is lower than t_2 :

$$\frac{H}{v_0} < \frac{L \cdot A_v}{Q} \quad (\text{Equation 3.7})$$

$$v_0 > \frac{H \cdot Q}{L \cdot A_v} = \frac{H \cdot Q}{L \cdot b \cdot H} = \frac{Q}{A} = v_H \quad (\text{Equation 3.8})$$

in which:

- b: width of the tank [m];
- v_H : Hazen velocity or surface loading rate [m.h⁻¹].

It is noted that v_H is independent of the depth of the tank. All particles with sedimentation rates exceeding v_H will theoretically be eliminated (i.e. captured or deposited over the bottom of the tank). In practice, there are not ideal settlers, turbulence will occur within the fluid — especially at the inlet zone — as well as waves on the liquid surface, caused by the wind; convection currents influenced by ambient temperature and the density differences affect their performance. Whenever possible a “laminar” flow must be established, stable, characterized by appropriate values of the Reynolds number, established by the following equation (adapted from AZEVEDO NETTO et al., 1998 e DEGREMÓNT, 2007):

$$Re = \frac{4 \cdot R_H \cdot v_C}{\nu} \quad (\text{Equation 3.9})$$

$$R_H = \frac{\text{cross-sectional area of the flow}}{\text{wetted perimeter of the cross section}} \quad (\text{Equation 3.10})$$

in which:

- Re: Reynolds number;
- v_C : circulation velocity of the fluid [$\text{m} \cdot \text{s}^{-1}$];
- R_H : equivalent hydraulic diameter [m];
- ν : kinematic viscosity of the liquid [$\text{m}^2 \cdot \text{s}^{-1}$].

According to Degrémont (2007), conditions are considered laminar if $Re < 800$. The laminar sedimentation is based on the concepts of discrete settling and, therefore, does not consider the depth of the tank as a factor affecting the particle retention. It assumes that the surface area available for settling can be considerably expanded by overlapping cells responsible for promoting the separation of liquid and solids, arranged in the upper region of the settler. In Figure 3.25 the principle of a system of parallel plates and is displayed, and presents the idea involving the equivalent sedimentation surface area.

Figure 3.25 – Cumulative effect on the equivalent surface area promoted by the use of laminar plates (DEGREMONT, 2007).

$$A_p = A_R \cdot \cos \alpha \quad (\text{Equation 3.11})$$

in which:

- A_p : projected plate area [m^2];
- A_R : surface area of the plate [m^2];
- α : horizontal inclination angle of the plate [degrees].

According to Droste (2004), high-rate settlers are composed by evenly spaced inclined panels (e.g., flat plates, tubes etc.). The Hazen velocity is then calculated from the projected area of all the laminar elements, as follows (DEGREMONT, 2007):

$$v_H = \frac{Q}{n_p \cdot A_p \cdot \cos \alpha} \quad (\text{Equation 3.12})$$

in which:

- v_H : Hazen velocity or surface loading rate [$\text{m}^3 \cdot \text{m}^{-2} \cdot \text{d}^{-1}$];
- Q : flow rate [$\text{m}^3 \cdot \text{d}^{-1}$];
- n_p : quantity of plates [units];
- A_p : elemental area of each plate [m^2].

These pieces or modules can be made of plastic, canvas or other materials. The velocity of a solid particle, resulting from the upward liquid flow and descending (vertical) velocity of sedimentation, carries it toward the upper face of the plate; the particle then slides over the surface of the lamella and settles over the bottom of the tank. Compared to the conventional units, the relatively high surface area of the laminar settler allows the use of increased loading rates, with equal or superior performance. The Reynolds number between laminar plates is very low when compared to those of a conventional clarifier, which is in the range of turbulent flow. The turbulence effectively reduces the sedimentation rate of the particles and causes resuspension of settled particles through their dragging.

To obtain an efficient sedimentation, the settled particles must undergo morphological modifications inside the laminar modules by agglomeration between them, so that, to leave this

location (i.e. to settle), they are not carried within the effluent liquid flow or suspended from the bottom of the tank (DEGREMÓNT, 2007).

3.3.2 APPLICATION IN WASTEWATER TREATMENT

Regardless of being applied in full-scale wastewater treatment plants, there are still professionals who have not justified restrictions for the use of high-rate clarifiers in this type of installations. In view of this, in the technical literature the high-rate concept is traditionally associated with water treatment.

Despite the rare published works on high-rate clarifiers for domestic sewage treatment systems^[1], the Brazilian Standard ABNT NBR 12,209 (ABNT, 2011) presents some guidelines for the design of this type of unit (applied as secondary clarification units in systems with biofilm reactors), among them:

- Horizontal angle of inclination of the plates or ducts (α) must be between 55° and 70°;
- Length of duct or channels between 1 and 1.2 m;
- Spacing between parallel plates between 7 and 10 cm (similar for size ducts);
- The fabrication material should be inert and non-deformable, provide high mechanical resistance and present a plain surface;
- Maximum hydraulic loading rate (HLR_c) should be equal to 80 m³.m⁻².d⁻¹.

The decision to introduce a high-rate clarifier in the system developed in this research results from the full conviction of the supervisor of the author (of this work) that this type of unit is under-exploited in the wastewater treatment field. Previous Brazilian municipal WWTP projects (e.g. Piracicamirim, 1994 – Piracicaba/São Paulo State; Santa Mônica, 1997 – Campinas/São Paulo State; Jardim das Flores, 1999 – Rio Claro/São Paulo State; Águas da Serra, 2002 – Limeira/São Paulo State etc.) that were implemented have shown that this type of unit presents high performance and should be considered as an attractive alternative for inclusion in the process train of wastewater treatment plants.

¹ PLAß & SEKOULOV (1995); SALEH & HAMODA (1999); HALALSHEH et al. (2010); ABOU-ELELA et al. (2013).

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MATERIALS AND METHODS

4

4.1 THE RATIONALE

Compared to a UASB reactor of equivalent dimensions, a new conception that offers a larger region for biomass occupation may, theoretically, present a higher volumetric capacity for wastewater treatment, i.e. more mass of degraded substrate per unit volume (Figure 4.1). This is the conceptual basis on which the new system was designed. In order to promote an increased sludge blanket expansion it is necessary to better control the mixing intensity (i.e. applied energy) within the reactor, an initiative that, in this particular research, was attempted through the action of rotating axial-flow impellers, promoted by the operation of a gearmotor, and sludge recirculation (from the sedimentation compartment to the reaction zone).

Figure 4.1 – Schematic illustration showing the presumed extent of the reaction zone in different reactors: (a) traditional UASB configuration and (b) “expanded” sludge blanket in a unit without three-phase separator, as the new system proposed in this research.

Another positive aspect expected from the installation of a mechanical mixer, is its potential to replace the piping system responsible for the influent distribution inside UASB reactors, thus enabling a more efficient, economic and flexible solution for distribution and dispersion of the substrate in the lower region of the reactor (Figure 4.2). In this context, it sought to impose an appropriate mixing intensity in the reactor to favour the maintenance of a stratified sludge blanket, avoiding the complete mixing state and minimizing the washout of solids from the reaction zone.

Figure 4.2 – Conceptual comparison between the feeding system in the conventional UASB reactor (a) and the mechanical agitation system proposed in this article (b).

At this point, the author encourages the reader not to associate the use of a mechanical equipment to a technological retrogression, considering that — depending on design and operation — anaerobic systems may be considered, at least partially, self-sufficient technologies in terms of energy consumption, as the biogas can be used to generate electricity. In addition it is important to emphasize that the energy required for mixing is very low, as shown in Chapter 5.

In order to prevent disturbances in the operation of the solid-liquid separation unit caused by the expansion of the sludge blanket, it was necessary to develop a new concept for the sedimentation compartment; (In the new system) It was composed of a high rate settler in order to ensure compactness and a higher removal efficiency rate of suspended solids. A recirculation

system was also planned in order to prevent the eventual washout of solids and to maintain the maximum presence of active biomass within the system, by pumping the settled sludge back to the reaction zone.

The maturation of the design led to the non-adoption of the GLS separator concept. Without a specific device for biogas collection, this gaseous product of anaerobic digestion should accumulate in headspace of the reactor and sent to treatment and primary use as an electricity source.

4.2 RESEARCH STRUCTURE

As this work aims to demonstrate the feasibility of the implementation of mechanical agitation and high rate sedimentation systems in anaerobic sludge blanket reactors, it became necessary to study some of the physico-chemical and biological properties of the anaerobic sludge, prior to their dimensioning and installation in a pilot scale unit.

This research was developed in the academic complex which comprises the “Campus I” of the University of São Paulo in São Carlos (Department of Hydraulics and Sanitation Engineering). Results obtained in bench scale experiments performed at the “Laboratory of Micro-Sensors” located in the “Laboratory of Advanced Treatment and Reuse” (LATAR) served as subsidies for the design of a new anaerobic sludge blanket reactor, as well as its construction and operation on a pilot scale at the “Experimental Sewage Treatment Plant” (USP WWTP).

Initially, batch tests were performed in bench scale reactors to determine the mixing intensities required to maintain the stratification of the sludge blanket and also, when necessary, to promote the resuspension of settled sludge. During these assays, 2-liter flasks were partially inoculated with anaerobic sludge (collected from a UASB unit) and subjected to different stirring regimes (i.e. varying velocity gradient — or G — values) for the visual examination of its effect on the behaviour of sludge mass in terms of expansion (inside the vessels).

Later, the same bench scale units were reinoculated and, over a continuous feeding system (with domestic wastewater), subjected to different agitation regimes (caused by the action of rotating impellers), i.e. varying G values, in order to determine which operating mode

would be able to provide the best results in terms of COD removal and the maintenance of the sludge blanket integrity. The results furnished by these preliminary tests served as basic design data and criteria for the unfolding of the ensuing phase, performed in a pilot scale reactor.

From the conclusions and observations made during the initial phases of the research, a pilot scale reactor was designed and built (in a concrete tank where a UASB reactor was previously installed), with an effective volume of 17.0 m³ (followed by a 3.8 m³ compartment for the removal of suspended solids), provided with a mechanical mixing system.

After a relatively long period required for manufacture and installation of the new elements into the concrete tank, the reactor was inoculated (with anaerobic sludge from UASB reactor operated at full scale) and an initial phase of tests in pilot scale reactor was started. The period of tests lasted for 140 days, cycle during which three distinct operating modes (defined on trial and error basis) were evaluated in order to establish the preferential conditions under which the effective operation of the reactor should be performed. Concurrently with the initial period of testing in the pilot scale reactor, operational data was obtained and enabled the analysis of the unit performance in terms of COD and SST removal efficiencies.

For the start of the effective operation of the pilot scale reactor, a last inoculation was performed immediately after the closing of the initial period of tests. The start-up period, time required for the system to reach the apparent dynamic equilibrium, lasted for 43 days.

The effective reactor operation lasted for 130 days. At this phase, four distinct operating modes were tested (including the start-up stage) in order to evaluate the performance of the proposed configuration and to characterize the anaerobic sludge produced in it. As satisfactory results were verified, more challenging conditions were imposed. Auxiliary systems for removal of coarse solids, sludge recirculation and automated control of the agitation were also introduced; Improvements regarding maintenance procedures were indispensable throughout the reactor operation period.

In order to assist the comprehension regarding the structure of the work, it was decided to partition it into three main phases, as shown in Figure 4.3. All activities carried out during the research were described in detail from the Section 4.3; the results obtained at each phase enabled essential subsequent advances, that contributed to the completion of the proposed studies.

Figure 4.3 – Schematic illustration representing the general planning and development of the research.

4.3 FIRST PHASE: LABORATORY TESTS TO DETERMINE THE PROPER MIXING CONDITIONS FOR SLUDGE SUSPENSION

4.3.1 GENERALITIES

In order to maintain the stratified structure of the solid mass contained within the liquid environment established inside the reactor, it is necessary to determine the intensity of agitation, or energy, to be applied in the fluid (i.e. liquid + solids) for this phenomenon to be observed. By making use of the velocity gradient concept, tests were carried out in bench-scale reactors to investigate the more appropriate operating range for a mixing system to function economically in order to maintain a stratified sludge blanket within these small vessels.

4.3.2 DETERMINATION OF THE VELOCITY GRADIENT REQUIRED FOR SLUDGE SUSPENSION (AND RESUSPENSION)

As a subsidy for the design of the agitation system later installed in a pilot reactor, samples of anaerobic sludge, collected at the Experimental Sewage Treatment Plant (University of São Paulo, São Carlos, Brazil), were subjected to different mixing intensities during batch tests (G values ranging from 0 to 80 s^{-1}). Each of the six 2-liter units was filled with 0.4 L of sludge (inoculum) and 1.6 L of sewage (i.e. 20% and 80% of the total volume, respectively) and subsequently stirred under different regimes; the temperature was maintained at $(22 \pm 2) ^\circ\text{C}$.

Under such conditions, the analysis of the sludge behaviour aimed to determine the more appropriate G (mean velocity gradient, s^{-1}) values required to: (i) the maintenance of the solid suspension, so that a “controlled” stratification of the sludge blanket was maintained, not allowing the biomass to reach the uppermost region of the reactor through the establishment of well-defined interface, with lower concentrations of solids, and (ii) the resuspension of settled sludge, because it was found that a more vigorous agitation is required to ensure reintroduction of those solids that settle over the bottom of the unit after prolonged interruptions in the operation of the stirring system.

Due to its simplicity, i.e. small size reactors fed under batch regime (Figure 4.4), it was not possible to evaluate the influence of two important phenomena on the performance of the agitation system: the role of biogas bubbles and the operation under distinct hydraulic loading rate (characteristic in continuously fed systems).

Figure 4.4 – Schematic illustration of the assay performed to determine the minimum velocity gradients required for suspension and resuspension of the anaerobic sludge.

4.4 SECOND PHASE: OPERATION OF THE BENCH SCALE REACTORS

Based on the results obtained in the batch-fed reactor, a new phase of the research was initiated. For such tests, the 2-liter reactors (same used in the previous experiment) were modified to receive continuous feeding; the domestic wastewater was pumped from a lift station located next to the laboratory. Three reactors were inoculated with 800 mL of anaerobic sludge (each), i.e. 40% of their effective volume; no re-inoculation was performed during the assay. The velocity gradient was varied in order to evaluate the influence of hydrodynamic regime on the performance of the reactors in terms of COD and TSS removal efficiencies. In each unit, the effluent was collected through a latex hose installed in the upper part of the reactor (on the 2-liter mark). The reactors were maintained in operation for 37 days, at a temperature of $(22 \pm 2 \text{ }^{\circ}\text{C})$. A schematic illustration of the experimental apparatus is shown in Figure 4.5.

The reactors were submitted to different hydrodynamic and feeding conditions: (i) the

“reactor 1” was not subjected to agitation, while the mean hydraulic retention time (θ_h) was 8 hours; (ii) the “reactor 2” continuously stirred, with the mean velocity gradient of 20 s^{-1} and θ_h equal to 8 hours and; (iii) the “reactor 3” was submitted to the mixing conditions of the reactor 2, except in relation to the θ_h , which was set at 4 hours.

Figure 4.5 – Experimental structure assembled for the tests with the 2-L anaerobic reactors.

The existence of only four channels in the pump head limited the quantity of reactors tested. It should be noted that there was, for each experimental reactor, the precaution to fix the inlet point at the center of the reactor, close to its bottom. This objective was achieved through the production and installation of small glass tubes, as illustrated in Figure 4.6.

Figure 4.6 – Schematic illustration of the affluent distribution inside the bench scale reactors.

4.5 THIRD PHASE: DESIGN AND APPLICATION OF THE PROPOSED CONCEPT IN A PILOT SCALE INSTALLATION

4.5.1 GENERALITIES

The new reactor configuration was based on the intention to simplify the actual UASB inlet distribution system and to propose a more efficient device for the removal of suspended solids. The solution was to replace the conventional influent distribution piping system by a rotating impeller, installed in the lower region of the reactor, and to create a sedimentation compartment free from the interference of gas bubbles, endowed with a laminar settler.

The development of the preliminary tests and their results allowed a better understanding of the physical phenomena associated with the research, in a way that some concepts originally adopted were replaced with more consistent hypotheses. These “discoveries” implied, mainly, in the modification of the agitation system concept, which included a automated system able to

operate in intermittent or continuous mode and under a defined velocity range (in revolutions per minute), with flexibility. Another envisioned possibility was to implement technological components that would provide an instantaneous diagnostic and a precise control of the sludge blanket expansion, enabling adjustment of its height in order to prevent (or reduce) the undesired excessive loss of solids within the effluent flow.

4.5.2 DESCRIPTION OF THE EXPERIMENTAL SYSTEM FOR SEWAGE TREATMENT

The new system was installed in the Experimental Wastewater Treatment Plant (WWTP), located at the USP Campus in São Carlos (São Paulo State, Brazil) and occupies the same concrete tank in which a UASB reactor — designed with traditional criteria, i.e. influent distribution system, three-phase separator, gas deflectors etc. — was previously installed (Figure 4.7). The tank has square base of side 2.0 m (total surface area of 4 m²) and effective depth of 5.2 m, with a effective volume of 20.8 m³. The process train implemented at the experimental WWTP is schematically illustrated as a flowchart in Figure 4.8. Although installed, the system to transport and decontaminate the biogas, as well as the generation of electricity from it was not part of the research.

Figure 4.7 – Illustrative scheme of the previously existing UASB reactor (cross section).

Raw wastewater intake structure

A unit constructed in concrete that receives the sewage originated from the USP Campus (Area 1) in São Carlos and a fraction of the flow of domestic wastewater collected by the public sewerage system in the central region of the city. Due to the depredation by third parties — i.e. individuals not involved with the research — of a rotary drum screen initially installed to remove coarse solids, there was the need to adapt — on an emergency basis — manually cleaned screens (arranged in series) to prevent damage to the subsequent units. A schematic illustration of this component is shown in Figure 4.9.

Figure 4.9 – Raw wastewater intake: main characteristics.

Grit removal channel

A covered concrete structure composed of two chambers, with the bottom shaped as inverted truncated pyramids (Figure 4.10), in which the sand was deposited as the sewage flowed through it.

Figure 4.10 – Schematic illustration of the grit removal channel.

Figure 4.11 – Grit removal channel: main dimensions.

Pumping station

An element composed of a suction well, suction and discharge piping systems (100 mm – PVC) and centrifugal pump. The unit (Figure 4.12) was installed to transport the liquid to the buffer tank. The pump intermittent operation was controlled by a level controller (i.e. floating switch), according to the variation of the volume of sewage inside the buffer tank.

Figure 4.12 – Pumping station: main components.

Buffer tank

A structure built with reinforced concrete rings (inner diameter of 1.8 m and effective height of 4.0 m) with a total storage volume of 10.2 m³ (Figure 4.13), designed to ensure an uninterrupted supply of sewage to the reactor^[1]. From this structure, the wastewater was directed to the anaerobic reactor (through a PVC piping system with a diameter of 150 mm) with the aid of a centrifugal pump.

Figure 4.13 – Buffer tank: schematic layout.

Feed pump

The feed pump discharge flow was adjusted in accordance with the hydraulic retention time in the reactor. A PVC discharge piping system was installed, with diameter of 50 mm. A schematic representation of the pump and its discharge pipeline is shown in Figure 4.14.

Figure 4.14 – Feed pump: schematic representation of both suction and feeding lines.

¹ Due to the modest instrumentation, the operational levels of the buffer tank could not be dynamically controlled. The use of sewage for other research activities also impaired the determination of its effective hydraulic retention time.

Manually cleaned static screen

A multifunctional metal structure (Figure 4.15) designed with (i) an antechamber provided with a 30 degree V-notch weir, to receive the wastewater from the buffer tank and the recirculation sludge and to enable the flow measurement, and (ii) a stainless steel inclined bar screen (inclination of 40°), with clear spacing between bars of 3 mm — in order to meet the guidelines recommended by the Brazilian technical standard ABNT NBR 12,209 (ABNT, 2011) —, for the removal of coarse solids.

Figure 4.15 – Static screen: main features.

The new wastewater treatment system

The main characteristics of the mixing apparatus installed in anaerobic reactor, high-rate settler and sludge recirculation system are presented in Sections 4.2.3.2, 4.2.3.4 and 4.2.3.6, respectively.

Illustrations representing the external layout of the units that composed the experimental treatment system are shown in Figures 4.16 and 4.17.

Figure 4.16 – General layout showing the arrangement of units that composed the treatment system in the Phase 3 of the research.

Figure 4.17 – Photographic record of the elements that composed the Experimental Sewage Treatment Plant at USP (São Carlos, Brazil): (a) raw wastewater intake structure, grit removal channel and pumping station; (b) general view (with buffer tank); (c) and (d) static screen.

4.5.3 THE DESIGN OF THE NEW SYSTEM

4.5.3.1 *Basic design guidelines for the construction of the new system*

Basically, the new system was comprised of two distinct zones (Figure 4.18): *i*) a digestion chamber (i.e. reactor), in which the contact between biomass and substrate was stimulated through the action of a rotary impeller, and *ii*) a compartment for sedimentation, in which a lamella plate settler was installed.

The reaction zone consisted of an airtight containment vessel which was inoculated with anaerobic sludge and fed with sewage. The implementation of an agitation apparatus was planned to ensure a proper dispersion of the substrate and its contact with the micro-organisms. The mixing system consisted of a axial flow impeller fixed to a vertical axis, which was rotated by the action of a gearmotor. After going through the digestion region, the liquid flow was routed to the settling compartment via an internal (dual) piping system made of PVC (diameter of 100 mm). The exhaustion of the gases resulted from the biological decomposition occurred via proper biogas conveyance pipeline; No special device was developed for the gas collection and its accumulation occurred in the reactor headspace.

To prevent the presence of high concentrations of suspended solids in the effluent of the digestion chamber, resulting from the excessive accumulation of scum in the upper region of the reactor, orifices with diameter of 75 mm were performed in the inner (dual) piping system, responsible for conducting the liquid from the reaction zone to the settling compartment. Both pipes received circular cuts, placed 30 cm below the liquid surface level. In order to prevent the accumulated scum to form a dense solid layer and impair the gas outflow, a radial flow impeller was installed in the gas-liquid interface (fixed in the same vertical axis composing the agitation system, in the uppermost region of the reactor), so that it could promote, through rotation the rotation of the shaft, the dispersion of the these material.

Inside the sedimentation chamber, the lamella plates (made of glass, with thickness of 4 mm) were adapted into a wood support, in which they were maintained inclined — with an angle of 70 degrees relative to the horizontal axis — and with clear spacing between plates of 4 cm. The laminar settler was designed to operated under a mean surface loading rate of $82 \text{ m}^3 \cdot \text{m}^{-2} \cdot \text{d}^{-1}$ (value for “normal” operational condition, i.e. θ_h equal to 8 h). The effluent collection device — fixed over the laminar settler — consisted of a gutter endowed with rectangular weirs and

a siphon to prevent undesired gas leakage. As an initiative to minimize the odour problem and to retain the eventual biogas production in both reactor compartments (i.e. digestion and sedimentation zones), the reactor was fully covered with metallic plates.

The new system was designed to function under various operating conditions, in terms of varying hydraulic loading rates (and upflow velocities, consequently), recirculation of the settled sludge (i.e. solids accumulated underneath the settler), impeller actuation, sludge withdrawal etc.

The adopted configuration reflected the efforts made to install all the dimensioned elements inside a concrete tank that was not originally designed to receive them, thus avoiding additional expenditures with the construction of a new structure. The schematic illustration of the new system is shown in Figure 4.18, in which can be observed that the laminar settling compartment was built within the tank. The reaction and sedimentation zones were separated by metal plate fixed inside the vessel; the influent sewage was directed to the digestion region, later to be sent to the solid-liquid separation chamber.

Figure 4.18 – The new anaerobic reactor: a schematic representation.

Figure 4.19 – Schematic illustration showing the flow of materials and main units of the experimental system.

Evidently, the ideal condition would be that the sedimentation compartment was installed in the external region of the concrete structure, ensuring higher volume for the reaction zone and enabling symmetrical dissipation of the flow inside of it. The volumes occupied by each of the two compartments installed inside the concrete tank are presented in Figure 4.20.

Figure 4.20 – Volumes of the digestion zone and sedimentation compartment.

4.5.3.2 Design of the mechanical mixing system

The procedure employed in the design of the mechanical agitation system was very similar to that applied in the dimensioning of flocculation tanks (commonly used in water treatment). Table 4.1 presents the basic data used in the calculations.

Table 4.1 – Mechanical agitation: design input data.

Parameter	Symbol	Value	Unit
Diameter of the impeller ^[1]	D_{imp}	0.67	m
Effective volume of the reaction zone	V	17.0	m^3
Mean velocity gradient ^[2]	G_{med}	10 ^[3]	s^{-1}
Maximum velocity gradient ^[4]	G_{max}	80 ^[3]	s^{-1}
Coefficient - Power number ^[5]	K	0.75 ^[6]	Dimensionless
Dynamic viscosity of the fluid	μ	1.113×10^{-3} ^[7]	N.m.s^{-1}
Volumetric mass density of the fluid	ρ	1,108 ^[7]	kg.m^{-3}

¹ Defined according to the literature (Table 3.9);

² Required to maintain partial suspension;

³ Value determined in the tests performed in the previous phases of the research;

⁴ Required to the complete resuspension of the sludge settled over the bottom of the reactor;

⁵ Measure of the relative drag of the impeller;

⁶ Value according to LENG et al. (2008);

⁷ Value for sludge with a solids content of 11%.

After attentive analysis of the specialized literature, it was opted to install axial flow impellers. According to Leng et al. (2008), these units have contoured blades, specially designed to provide greater efficiency (compared to radial flow units) and to promote good dispersion of gases and low shear rate.

Power drawn by the impeller (P_{imp}), W:

$$P_{\text{imp}} = V \cdot \mu \cdot G^2 \quad (\text{Equation 4.1})$$

in which:

- V : volume of the reactor [m^3];
- μ : dynamic viscosity of the fluid [N.m.s^{-1}];
- G : velocity gradient [s^{-1}].

$$P_{\text{imp_med}}, \text{ W: } P_{\text{imp_med}} = V \cdot \mu \cdot G_{\text{med}}^2 = 17.0 \cdot 1.113 \times 10^{-3} \cdot 10^2 = 1.9 \text{ W}$$

$$P_{\text{imp_max}}, \text{ W: } P_{\text{imp_max}} = V \cdot \mu \cdot G_{\text{max}}^2 = 17.0 \cdot 1.113 \times 10^{-3} \cdot 80^2 = 121.1 \text{ W}$$

Rotational speed of the impeller (n), rpm:

$$G = \sqrt{\frac{K \cdot \rho \cdot n^3 \cdot D_{\text{imp}}^5}{V \cdot \mu}} \Rightarrow n = \sqrt[3]{\frac{V \cdot \mu \cdot G^2}{K \cdot \rho \cdot D_{\text{imp}}^5}} \therefore n = \sqrt[3]{\frac{P_{\text{imp}}}{K \cdot \rho \cdot D_{\text{imp}}^5}} \cdot 60 \quad (\text{Equation 4.2})$$

in which:

- K : power number (0.75);
- ρ : volumetric mass density of the fluid (kg.m^{-3});
- D_{imp} : diameter of the impeller (m).

$$n_{\text{med}}, \text{ rpm: } n_{\text{med}} = \sqrt[3]{\frac{1.9}{0.75 \cdot 1108 \cdot 0.67^5}} = 15.8 \text{ rpm}$$

$$n_{\max}, \text{rpm: } n_{\max} = \sqrt[3]{\frac{121.1}{0.75 \cdot 1108 \cdot 0.67^5}} \cdot 60 = 63.1 \text{ rpm}$$

4.5.3.3 Device to mitigate blockages caused by scum

In order to prevent the accumulation of scum and problems arising from the clogging caused by it (especially in relation to the biogas release), a radial flow impeller was installed in the upper region the concrete tank (Figure 4.21), i.e. at the same level occupied by the scum layer, attached to the vertical shaft that comprises the stirring system implemented in the reaction zone. The rotational movement of the impeller causes the displacement of the floating solids that accumulate near the liquid surface of the reactor, allowing the gas to leave the unit without additional difficulties by “breaking” the scum over the point at which it was inserted, creating a preferential path.

Figure 4.21 – Radial-flow impeller for mitigation of scum accumulation in the upper region of the digestion zone.

4.5.3.4 Design of the high-rate settler

The high rate clarifier — a box made of wood in which inclined glass-made lamella plates were inserted — was designed in accordance with the recommendations of the Brazilian Technical Standard ABNT NBR 12,209 (ABNT, 2011), which suggests^[1] the maximum hydraulic loading rate of 80 m³.m².d⁻¹. The option to use glass for manufacturing the lamella plates is justified by the lower roughness and reduced cost presented by this material, in comparison to PVC (plates). A schematic illustration of the clarifier is shown in Figure 4.22.

¹ According to the terminology used in this Technical Standard, the value is recommended for “secondary settlers of the lamella or tubular type” to “biological (aerobic) processes with biofilm”.

Figure 4.22 – Schematic representation of the high rate clarifier: (a) External layout and (b) internal arrangement of the plates .

Hydraulic loading rate (clarifier)

The clarifier hydraulic loading rate (HLR_c – Equation 4.4) is calculated based on the area effectively available for the upward liquid flow. In the calculation to determine the effective area, the sum of the areas occupied by the lamella plates should be deducted (Figure 4.23).

$$HLR_c = \frac{Q}{A_e} \quad (\text{Equation 4.3})$$

in which:

- HLR_c : clarifier hydraulic loading rate ($\text{m}^3 \cdot \text{m}^2 \cdot \text{d}^{-1}$);
- Q : mean inflow rate ($\text{m}^3 \cdot \text{d}^{-1}$);
- A_e : area effectively available for the liquid flow (m^2).

Figure 4.23 – Area effectively available for the liquid flow within the clarifier.

Velocity of the liquid flow within the plates

Equation 4.5 shows the calculation of the velocity of liquid flow within the channels formed between the lamella plates of the sedimentation tank; The thickness of the plates should be deducted (Figure 4.24).

$$v_p = \frac{Q_c}{A_c}$$

(Equation 4.4)

in which:

- v_p : velocity of the liquid flow within the plates (m.h^{-1});
- Q_c : flow rate in each channel, formed between plates ($\text{m}^3.\text{h}^{-1}$);
- A_c : Cross-sectional area of each channel (m^2).

Figure 4.24 – Channels formed between the lamella plates and its main dimensions.

4.5.3.5 Sample collection points

The sample collection points of anaerobic sludge located inside the digestion compartment were previously installed in concrete tank (i.e. original elements of the UASB reactor). From the manual actuation of ball valves, the samples could be collected from three points along the height of the structure: P1, P2 and P3, located 1.5 m, 3.0 m and 4.5 m from the base of the tank, respectively. Samples of the sludge accumulated over the bottom of the reactor were collected through the opening of a valve installed at its lowest level. The arrangement of collection points is shown in Figure 4.25.

Figure 4.25 – Sample collection points of anaerobic sludge within the digestion zone.

The intention to install additional points for sample collection became impracticable due to the high cost associated with the implementation of new openings in the tank wall (vertical wall of reinforced concrete with thickness of 20 cm); The greater the quantity of sample collection points, more structured are determined profiles.

4.5.3.6 Sludge management: recirculation and withdrawal systems

Predicting eventual washout of anaerobic sludge — during the stages of inoculation and operation of the reactor — and aiming to maintain a satisfactory concentration of (still) active biomass within the digestion compartment, it was decided to install a pumping system able to recirculate the sludge from the settling compartment to the chamber that precedes the static screen (Figure 4.26); Thus, the presence of a triangular weir upstream of the reactor enabled the control (i.e. measurement) of the recirculation flow, while the screen was able to retain coarse solids which somehow had been incorporated to the sludge mass.

Figure 4.26 – Sludge recirculation system: suction and discharge systems.

The settled sludge was recirculated through a system composed of helical pump, and pipelines for suction and discharge (both made of PVC, diameter ranging from 50 and 100 mm); Two sludge discharge points — controlled by globe valve actuation — were established, one upstream and other downstream of the of static screen (Figure 4.27). The pump was controlled by a frequency inverter, which permitted the variation of the rotational speed of the pump motor and, consequently, alterations in the recirculation flow rate.

Figure 4.27 – Discharge points for the recirculated sludge.

For the final disposal of the sludge, i.e. sludge conveyance from the bottom of the sedimentation chamber to its discharge point, an apparatus comprised of a 137 L tank (“disposal chamber”) and piping system (diameter varying between 50 and 100 mm, activated by ball valves) were installed (Figure 4.28). Since both disposal chamber and discharge point were positioned at a lower level than the bottom of the sedimentation compartment, the sludge flow occurred by gravity (no pumping was required).

Figure 4.28 – Schematic illustration of the chamber installed for the sludge withdrawal procedure.

4.5.4 IMPLEMENTATION OF THE PILOT SCALE REACTOR

After the definition of the internal configuration of the new anaerobic reactor, the installation of its elements in the concrete tank needed to be started. As the sludge discharge system (originally installed) was obstructed, there was the need to hire a company specializing in septic tank cleaning to remove the anaerobic sludge and other debris present in the reactor.

Following the sludge removal, the interior of the reactor was sanitized so that workers responsible for removing the internal components of the previously installed UASB reactor (i.e. three phase separator and the influent distribution system, mainly) could start their activities free of contamination. Images of the discharge operation and cleaning procedure are presented in Figure 4.29.

Figure 4.28 – Emptying and cleaning of the concrete tank: a) sludge accumulation within the UASB reactor; b) vacuum truck; cleaning performed inside the reactor to prevent contamination of workers.

The next stage of the activities consisted of the removal of the three-phase separators, gas deflectors, influent distribution system and effluent collection gutters. As the lateral gas deflectors, i.e. installed in the walls of the tank (along its square perimeter), were originally made of concrete, special equipment had to be used for its removal. Photos of the interior of the reactor before and after the removal of the internal components of the UASB are shown in Figure 4.29; the photographic record of the lower region of the tank, before the removal of the influent distribution system is shown in Figure 4.30. In the new reactor, instead of using a pipe system for the distribution of the influent sewage, a single lateral inlet point was used only to feed the unit.

Figure 4.29 – The interior of the concrete structure: a) before the removal of internal components and; b) after the withdrawal of these elements.

Figure 4.30 – Influent distribution system: a) after the anaerobic sludge withdrawal and; b) moments before the removal of the piping system.

Once the removal of all UASB internal elements was completed, the installation of the elements that integrated the new reactor was initiated. The first step was the application of a concrete waterproofing product (i.e. sealer) on the walls of the tank (Figure 4.31), and, days later, concrete fillets were executed along the corners formed between vertical walls and the base of the reactor, whose purpose was to confer a rounded shape to the bottom of the tank to avoid dead zones in the region. It is worth noting that fillets between the vertical wall were originally implemented inside the tank, eliminating the need for such interventions.

Figure 4.31 – Concrete tank after the application of waterproofing product (sealer).

The next stage was characterized by the introduction of the metallic plate which divided the reactor into two distinct regions: reaction and sedimentation compartments. The plate was fixed vertically in the concrete side walls. After this, the high-rate settler was installed, consisting of a wooden support on which glass plates (4 mm thick) were installed inclined at 70°. In Figure 4.32 images of the laminar settler are displayed, during its assembly phase and after

being arranged within the sedimentation zone.

The subsequent step involved the installation of the rotating impeller, the (hollow) vertical shaft (to which the impeller was coupled), and a lower bearing was fixed over the base of the tank for proper operation. To ensure alignment of the vertical axis during operation of the stirring system, three sliding contact bearings were positioned along the height of the reaction zone, screwed in the metallic plate. The photographic record of the installation of some components of the mixing system is shown in Figure 4.33.

Figure 4.32 – Images of the laminar settler: mounting (a, b and c) and installation phases (d), prior to the installation of the glass plates.

Figure 4.33– Installation of a mechanical agitation system: (a) vertical shaft and the hub which connects the impeller blades to the shaft; (b and d) bearing; (c) impeller base.

After installation of the mixer shaft, the reactor cover — made of a metallic sheet with slip-resistant surface — was installed, as shown in Figure 4.34. Next an opening (cut) was performed in the reactor cover, so the gearmotor (rated power of 2 cv; reduction ratio of 1:39), responsible for the rotation of the impeller, could be installed. It was carried out also the insertion of the stainless steel impeller blades on the agitator shaft, few centimetres above the base of the reactor. Images of part of the structure set up to support the gearmotor assembly and of the impeller are shown in Figure 4.35.

Figure 4.34 – Metallic cover installed in the upper region of the reactor.

Figure 4.35 – Photographic record: a) gearmotor support and; b) axial flow impeller.

To feed the anaerobic reactor, a centrifugal pump (brand/model: Thebe/AE-2) was installed, as well as its suction and discharge lines. The pump was installed to allow flexibility in the control of inflow to the unit (Figure 4.36).

Figure 4.36 – Photographic record of the centrifugal pump and other elements that composed the feeding system.

In February 2013 the installation of the equipment that composed the agitation system was performed. A control panel with a frequency inverter was installed in a maintenance room, located few meters from the reactor; the controller of suspended solids sensor was installed inside a metallic panel, mounted in one of the lateral walls of the reactor. The suspended solids sensor was installed in April 2014, due to the postponement of the reactor inoculation.

Photos of the control panel and other elements that composed the mechanical agitation system are presented in Figure 4.37.

Figure 4.37 – On site photographic record of the pilot scale system: (a) side view with the gearmotor and; (c) control panel with frequency inverter.

A tightness test was performed with water to assess the concrete tank conditions on the occurrence of leaks. A major leakage point (i.e. crack in the concrete side wall) was found, as a worrying liquid jet was established. To solve the problem, a team of experts was contracted and carried out the repair by adding a waterproofing admixture to the concrete used in the wall restoration (Figure 4.38).

Figure 4.38 – Waterproof concrete applied in the reactor wall.

Due to the permanent deactivation of a rotary screen operation placed upstream of the grit removal chamber, a static screen with 3 mm openings (Figure 4.39) was designed and built, in order to prevent the incursion of undesired solids such as hair and other materials into the reactor; to control the inflow rate, a chamber endowed with a 30-degree V-notch was installed upstream of the screen.

Figure 4.39 – Static screen implemented upstream of the anaerobic reactor.

Tests conducted by slowly filling the concrete tank with water allowed the verification of unwanted internal leakage (i.e. liquid flows between the digestion and sedimentation compartments). To solve the problem a company was hired to apply fiberglass along the corners formed in the intersections of the metallic plate and the side walls, an initiative that presented itself inefficient after new tests with water. Two more attempts were made, and after failures in both of them, such company decided to cancel the contract. Photos of the leaks and the applied fiberglass are presented in Figure 4.40.

Figure 4.40 – Application of fiberglass in an attempt to contain the internal leakage.

Due to the lack of proven effective alternatives for solving the problem, it was decided by the application of liquid rubber (usually applied as coating over the undercarriage of vehicles), an alternative that was effective during the whole operation of the reactor. After containment of leaks it was determined that the reactor was ready to be inoculated. Images of application of the liquid rubber inside the reactor are shown in Figure 4.41.

Figure 4.41 – Application of liquid rubber for the elimination of internal leaks.

4.5.5 INITIAL TESTS IN THE PILOT SCALE SYSTEM

After the completion of the pilot scale reactor construction, initial tests were performed to determine the most appropriate operating conditions. These experiments, conducted on a trial-and-error basis, were used to evaluate the behaviour of the experimental system in relation to the maintenance of the layered structure of the sludge blanket within the digestion compartment (also trying to avoid the biomass washout) and its performance in terms of COD and TSS removal.

In order to abbreviate the time required for the reactor to reach the steady state, the reaction compartment was completely filled with anaerobic sludge from a UASB reactor installed at the Monjolinho WWTP (population equivalent of 220,000 inhabitants), located in São Carlos (São Paulo State, Brazil). The photographic record of one collection campaign of anaerobic sludge held in Monjolinho WWTP is shown in Figure 4.42.

Figure 4.42 - Collection of the sludge from a UASB reactor at the Monjolinho WWTP.

During this experimental period, the TSS profile along the digestion compartment height was continuously monitored in order to evaluate the dynamic responses of biomass to the different operating conditions (e.g. varying hydraulic loading rates, mixing intensities etc.), as well as the performance of the unit. The TSS profile was a deciding factor to determine the threshold between periods of adaptation and stability of the sludge. The objective set for

these tests was to maintain the concentration of TSS in the “P1” sampling point always superior to 5,000 mg.L⁻¹. If this limit was negatively exceeded, new inoculations should be carried out and a revised operating mode proposed (in terms of HLR, mixing intensity etc.). This step was considered complete from the time that the SST profile and determinations related to the performance of the system showed satisfactory and consistent results.

4.5.6 FINAL INOCULATION AND START-UP PROCEDURE

After defining the apparently effective operating modes (i.e. feeding and mixing strategies), the unit was completely filled (for the last time) with anaerobic sludge from the Monjolinho WWTP. The photographic record of the last phase of inoculation is presented in Figure 4.43.

Figure 4.43 – Reactor inoculation by introducing anaerobic sludge from the Monjolinho WWTP.

In order to characterize the inoculum, samples were collected for laboratory testing, i.e. total suspended solids concentration, specific methanogenic activity and metal content in the sludge. During the start-up stage, as a evaluation of the system behaviour, emphasis was given to the TSS profile along the height of the reactor.

4.5.7 OPERATION OF THE SYSTEM

The period of effective operation of the reactor was characterized by the implementation of the efficient operating modes, or strategies, determined after the completion of the initial phase of tests in the unit, as well as the beginning of the sludge recirculation system operation. Due to restrictions regarding the use of the helical pump, over the 130 days of operation, the sludge recirculation was performed daily for 30 minutes; the volume recirculated (every day) was 1.4 m³.

Seeking to extract the maximum potential of the experimental system, in terms of volumetric capacity of treatment, increasingly challenging operating conditions were gradually imposed on the biomass contained within it (i.e. higher hydraulic and organic loading rates) as the unit presented satisfactory and consistent results after being subjected to certain operating mode over a few (i.e. at least two) consecutive weeks. The evaluation of the system performance was based on BOD, COD and TSS removal efficiency rates presented by it, as well as the total suspended solids profile established over the height of the digestion zone; The limit of 5,000 mg.L⁻¹ for TSS concentration at point P1 (1.5 m above the tank bottom) was maintained.

4.5.8 OPERATING PARAMETERS: SOFTWARE FOR ONLINE DATA ACQUISITION

In recent years new technologies were introduced in the wastewater treatment market, making it plausible to supervise and register, in real time, the variations associated with the behaviour of the entire process. By employing these modern equipment, diagnostics and adjustments can be performed more quickly and efficiently, since the results are emitted instantaneously.

Concurrently with the inoculation step, a specialized company was hired to develop a software able to record (in real time), with the aid of several specific devices, data on important operational aspects; it was installed in a computer placed in the maintenance room. Basically, the software was designed to allow an online diagnostic regarding important operational parameters, such as the sludge blanket expansion, influent flow rate, biogas production, temperature, gas pressure within the headspace of the digestion compartment and the mixing intensity applied inside the reactor (Figure 4.44).

Figure 4.44 – Graphical interface of the software used for data acquisition.

4.5.8.1 Suspended solids determination

Due to the slow growth of methanogens, ideal conditions should be maintained to ensure the a satisfactory concentration of anaerobic biomass within the reactor (e.g. hydraulic and organic loadings). In order to evaluate the influence exerted by some of the operating parameters (e.g. inflow rate and velocity gradient) on the dynamics of sludge blanket, a sensor able to measure the concentration of suspended solids (brand/model: Hach SOLITAX) was installed in the upper region of the reaction chamber (4.5 m from the base of the tank). The results, measured at intervals of 1 s, were recorded by the software through the aid of a multifunction data acquisition unit (NI USB-6009, Native Instruments).

4.5.8.2 Sewage inflow rate

For the online record of the inflow rate, a low-cost level measurement system — composed of a floating device (i.e. a plastic bottle filled with polyurethane foam) connected to a linear potentiometer — was installed in the chamber located upstream of the static screen, equipped with a 30° V-notch weir (Figure 4.45).

Figure 4.45 – Low-cost level measurement system installed upstream of the V-notch weir.

4.5.8.3 Gas pressure and liquid temperature inside the reaction zone

In order to measure the temperature of the liquid mass inside the reaction zone as well as the pressure of gas in the headspace of the reactor, specific sensors were installed, as shown in Figure 4.46.

Figure 4.46 – Photographic record of the sensors installed on the cover of the digestion zone to measure the liquid temperature and pressure of the gas.

4.5.8.4 Mixing intensity

The velocity gradient, calculated from the rotational velocity of the impeller and the volume of the reaction zone (among other data, as shown in Equation 4.3) was also monitored. The software was developed to allow the record of the rotational speed of the stirring system, also in real time.

4.5.9 MONITORING THE REACTOR PERFORMANCE

The performance evaluation of the reactor was accomplished by means of physical-chemical analysis, and by monitoring operating parameters to which the experimental unit was subjected, e.g. hydraulic application rate, temperature, velocity gradient etc. Temporal samplings were carried out in various points of the system in order to characterize the mass of anaerobic sludge present in the reactor and its response in terms of removal efficiency of suspended solids and organic material.

The interval between each sample collection campaign varied throughout the system operating period, usually held weekly or fortnightly. Over the initial stages of each reactor operation phase, the samples were sparser (i.e. at widely spaced intervals), due to the instability of the process. During the campaigns, samples were collected every two hours — within a period between 6:00 AM and 8:00 PM on sample extraction days — to evaluate the reactor performance in the following points:

- Sewage inlet chamber, to determine the quality of the sewage influent to the system;
- P3 (4.5 m from the base of the reactor, inside de reaction zone), to determine the characteristics of the sewage effluent from the digestion zone (or sewage influent to the sedimentation compartment);
- Effluent outlet pipe, to determine the quality of the final effluent.

For the samples collected at these points, in order to evaluate the performance of the “liquid phase” system, determinations were made regarding the following parameters:

- COD: raw and filtered samples;

- Total solids (TS), Total suspended solids (TSS) and volatile suspended solids (VSS);
- pH.

Besides the aforementioned variables, the samples were analyzed in order to also sporadically determine the biochemical oxygen demand (BOD), Total Kjeldahl Nitrogen (TKN), ammonia nitrogen, phosphate and sulphide. All analysis were performed at the Laboratory of Sanitation – SHS / USP.

In order to establish profiles and to evaluate the performance of the system, samples were collected in the following points: (i) the sewage inlet chamber, (ii) the effluent outlet pipe, (iii) sludge discharge points (i.e. bottom of the reaction zone and of the sedimentation compartment) and (iv) sludge sampling points (three points along the reactor height). The analyses and the respective methods applied are shown in Table 4.2.

Table 4.2 - Physico-chemical monitoring variables and their methodologies applied.

Variable	Analysis method	Code	Reference
Temperature	Direct measurement	–	–
pH	Potentiometric – APHA	4500–H+	APHA, 2012
COD	Colorimetric – APHA	5220	APHA, 2012
BOD	Respirometric – APHA	5210–B	APHA, 2012
TS	Gravimetric – APHA	2540–B	APHA, 2012
FS	Gravimetric – APHA	2540–E	APHA, 2012
VS	Gravimetric – APHA	2540–E	APHA, 2012
TSS	Gravimetric – APHA	2540–D	APHA, 2012
FSS	Gravimetric – APHA	2540–E	APHA, 2012
VSS	Gravimetric – APHA	2540–E	APHA, 2012
TKN	Titrimetric - APHA	4500–N _{org} B	APHA, 2012
Nitrogen (Ammonia)	Titrimetric - APHA	4500–NO ₃ ⁻	APHA, 2012
Phosphate	Colorimetric – APHA	4500–P (E)	APHA, 2012
Sulphide	Colorimetric – APHA	4500–S2	APHA, 2012
Metals	Spectrophotometric – APHA	3111–B	APHA, 2012

4.5.10 SPECIFIC METHANOGENIC ACTIVITY (SMA) TEST

The specific methanogenic activity (SMA) is a bench scale test that measures the conversion rate of soluble substrates into methane. The microbial activity test has proved important and can be used as a routine analysis to quantify the condition of the anaerobic sludge and, also, in a series of other applications, as follows (CHERNICHARO, 2007a):

- To ascertain the behaviour of the microbial community under the presence of potentially inhibiting compounds;
- To assess the degree of biodegradability of several substrates;
- To monitor behavioural changes regarding the activity of the biomass, especially those caused by the accumulation of inert material over the bottom of the reactor;
- To determine the maximum organic load tolerated by the anaerobic sludge, providing an acceleration of the start-up procedure; and
- To evaluate kinetic parameters.

Although there are various distinct methods to the realization of the SMA test, there is currently no standard procedure for its determination. In this research the following protocol was adopted (Figures 4.47 and 4.48); this routine, classified by Aquino et al. (2007) as a “direct measurement of the volume of methane”, allows the assessment of methane yield under a controlled environment, from the known values of biomass (g SSV) and substrate (g COD), it was developed from pioneering work, such as those performed by de Zeeuw (1984) and Penna (1994):

- Determine the concentration of volatile solids present in the sludge to be evaluated (g VSS.L^{-1});
- Fill the reaction flask with a pre-established volume of anaerobic sludge, calculated to obtain a final concentration of inoculum in the range between 2 and 5 g VSS.L^{-1} ;
- Put the reaction flask inside a water bath or other incubator equipment and, in order to enable the adaptation of the inoculum to the environmental conditions (periodically stirred and with temperature ranging between 30 and 35 °C) imposed during the test, maintain it under such state (and without external nutrition) for 12 to 24 hours before

the beginning of the analysis;

- Before adding the substrate, gaseous nitrogen should be applied over the headspace of the flask to remove the oxygen;
- Add the substrate (sewage) to fill the remaining volume of the reaction and close the flask with a cap;
- To ensure contact between biomass and substrate, periodically stir the flask by spinning it gently;
- Fill a Mariotte bottle with alkaline solution (e.g. NaOH 5%) and connect its cap to the reaction flask through a flexible hose;
- Place a graduated cylinder below the siphon located in the lower region of the Mariotte bottle and record the volumes of gas produced at each time interval (mL.h^{-1}) throughout the duration of the test.

Figure 4.47 – Experimental set-up required to perform the SMA test.

Figure 4.48 – Photographic record of the apparatus mounted to the realization of SMA tests on anaerobic sludge samples.

With the aid of the experimental set-up, the volume of methane was determined by periodically measuring the amount of NaOH expelled from the Mariotte bottle, caused by the depressurization that occurred inside the reaction flask due to the production of methane. Once the volume of the methane produced was known, the SMA value was calculated using the relation between the maximum rate of methane yield (expressed in $\text{g COD}_{\text{CH}_4} \cdot \text{d}^{-1}$) and the sludge mass (in terms of g SSV) used as inoculum (adapted from Aquino et al., 2007).

To calculate the specific methanogenic activity it was used the procedure described by Aquino et al. (2007).

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RESULTS AND DISCUSSION

5

5.1 INTRODUCTION

As set out in Chapter 4, the research was carried out in three main phases. In the first, bench scale studies were conducted to characterize the anaerobic sludge. These tests, due to its exploratory character, were performed without methodological rigour, intended only for observation of phenomena rarely reported in the literature (e.g. anaerobic sludge settleability and required mixing intensity within the reactor). Later, in the second phase, 2-L continuously fed bench scale reactors were maintained under different stirring regimes in order to evaluate the influence of mixing intensity on the performance of such units. Finally, the third phase consisted in the design, construction, operation and routine performance evaluation of a new anaerobic reactor applied to the treatment of domestic sewage in pilot scale. In this chapter the results of the experiments and the most relevant discussions associated with them are presented. The difficulties encountered and the findings evidenced throughout the period of operation of the experimental unit, typically related with research involving technological innovation, were also included in this text.

5.2 RESULTS AND DISCUSSION: FIRST PHASE

5.2.1 VELOCITY GRADIENTS FOR PARTIAL SUSPENSION, ENSURING A STABLE SLUDGE BLANKET, AND RESUSPENSION OF THE SLUDGE

The velocity gradients required to maintain the partial suspension, i.e. a stable sludge

blanket, and resuspension of anaerobic sludge were defined visually. From this experiment it was found that the G value required for maintaining a steady and stratified sludge blanket should be up to 20 s^{-1} ; for sludge resuspension — required after interruptions in the operation of the jar test equipment — a G value of, at least, 80 s^{-1} must be applied.

Figure 5.1 – Photographic record of the experiment conducted to investigate the sludge behaviour under different G values.

These G values were employed in the experiment of the second phase, enabling the achievement of satisfactory results; from this confirmation, they served as basic design data for the dimensioning of the agitation system installed in the pilot reactor (Phase 3). These results rely on observation and personal interpretation of the professional conducting the test (without using specific equipment) and have been shown reliable and confirmed in practice during the tests in pilot scale reactor. Unfortunately no data was found in literature concerning the mixing intensities applied in anaerobic reactors treating wastewater.

5.3 RESULTS AND DISCUSSION: SECOND PHASE

5.3.1 THE PERFORMANCE OF THE BENCH SCALE ANAEROBIC REACTORS

The second phase of the experimental procedures, performed in 2011, consisted of the operation of 2-L anaerobic reactors and its results, in terms of COD removal, are presented in Table 5.1.

Table 5.1 – Results of the experiment with the 2-L anaerobic reactors: average efficiency.

	Variable / Parameter	Unit	Total number of days elapsed from the beginning of the test				
			9	16	23	30	37
	Influent COD	(mg.L ⁻¹)	1,919	810	376	1,544	355
Reactor 1 $\theta_h = 8h$ non agitated	Effluent COD	(mg.L ⁻¹)	191	218	189	200	122
	Removal efficiency	(%)	90	73	50	87	66
Reactor 2 $\theta_h = 8h$ $G = 20 \text{ s}^{-1}$	Effluent COD	(mg.L ⁻¹)	323	277	196	551	156
	Removal efficiency	(%)	83	66	48	64	56
Reactor 3 $\theta_h = 4h$ $G = 20 \text{ s}^{-1}$	Effluent COD	(mg.L ⁻¹)	166	371	220	273	107
	Removal efficiency	(%)	91	54	41	82	70

After the conclusion of the experiment it was decided that the mean velocity gradient should be maintained below 20 s^{-1} , since the best results were achieved by the “reactor 1”, the one that was not agitated^[1].

Although apparently inconclusive, the results of the experiment were of great value. At the end of the test, it was evident that there was a loss of solids in the reactors due to the absence of a solid-liquid separation unit that would act as a barrier, capable of retaining the sludge within the system. The lower efficiency of the two “mixed” units can be attributed to the washout of solids, which caused a decrease in the concentration of active biomass. In Figure 5.2 is presented a photographic record of the sludge behaviour when: a) in idle status and; b) when subjected to agitation.

Figure 5.2 – Photographic record of the anaerobic sludge: (a) without and (b) under agitation.

¹ Although the non-stirred reactor have shown superior results, in practice it was noticed that there is need for agitation, even if minimal, in order to prevent the formation of gas pockets within the sludge bed/blanket and to ensure a satisfactory contact between biomass and substrate.

5.4 RESULTS OBTAINED IN THE THIRD PHASE

5.4.1 CHARACTERIZATION OF THE RAW WASTEWATER

The main characteristics of the raw sewage influent to the experimental wastewater treatment plant (USP WWTP) were obtained through analysis of samples collected inside the wastewater intake structure, during the effective period of system operation. The results are presented in Table 5.2.

Table 5.2 – Characterization of the raw sewage influent to the experimental WWTP.

Variable	Values (mg.L ⁻¹)		
	Median	Minimum	Maximum
BOD	306	130	331
COD (filtered samples)	255	240	257
COD (raw samples)	570	522	580
Hydrogen sulphide	0.06	0.06	0.06
Phosphate	12	12	12
TS	608	600	616
TVS	400	331	469
TFS	208	147	269
TSS	209	206	212
VSS	186	186	186
FSS	23	20	26

It is appropriate to point out that the values shown in Table 5.2 refer to the raw sewage influent to the experimental system, which has not been subjected to preliminary treatment. The purpose of presenting data relating to the raw sewage is to enable a qualitative assessment of this discharge, e.g. if it can really be classified as domestic sewage; It is possible, also, through the data, to determine the COD and solids removal efficiencies presented by the units installed upstream of the experimental system evaluated in this research. It is noted from the data presented in Table 5.2, the relationship between the mean values of BOD and COD (raw samples) is 0.54, a value which is in accordance with those associated with domestic sewage in the technical literature. Figure 5.3 shows the flowchart representing the distribution of the solids concentration in the sewage influent to the WWTP, calculated from the results shown in Table 5.2.

Figure 5.3 – Flowchart of the percentage distribution of the solids concentration in the raw sewage influent to the WWTP.

To determine the impact exerted by the presence of the preliminary treatment on the quality of the sewage influent to the pilot scale anaerobic reactor, the performance of the units that composed it was assessed in relation to their respective removal efficiencies in terms of COD (raw and filtered samples) and solids (TS and TSS). It can be concluded that the implementation of pre-treatment processes causes significant — maybe excessive^[1] — abatement in the COD, a fact that impairs the overall performance of the anaerobic reactor, in terms of COD removal efficiency, since most of the organic matter which would be removed during the anaerobic digestion process is removed before entering the pilot scale reactor.

The preliminary treatment units showed to be efficient, contributing with removal efficiencies of 42%, 22% and 26%, for COD, TS and TSS, respectively. A flowchart representing the distribution of the solids concentration in the sewage influent to the system (after undergoing preliminary treatment) is shown in Figure 5.4.

¹ Possibly the main cause of COD reduction in the sewage was an eventual occurrence of anaerobic degradation occurred within the buffer tank, since there was the potentiality of unwanted sludge accumulation in the bottom of this unit over time the influent substrate was maintained inside of it. Operational difficulties did not allow an optimum control of tank outflow.

Figure 5.4 – Flowchart of the percentage distribution of the solids concentration in the pretreated sewage influent to the pilot scale reactor.

By comparing the data shown in Figures 5.3 and 5.4, changes can be observed in the percentage compositions associated with the solids contained in the sewage samples (before and after being subjected to the preliminary treatment). In order to establish a comparison standard and thus classifying the samples of sewage influent to the experimental WWTP and to the anaerobic reactor, according to the bibliography, the typical values of some wastewater variables are presented in Table 5.3.

Table 5.3 – Typical characteristics of domestic sewage and its classification according to the literature.

Variable	Unit	Strength		
		Low	Medium	High
BOD	mg.L ⁻¹	110	190	350
COD	mg.L ⁻¹	250	430	800
TS	mg.L ⁻¹	390	720	1,230
TDS	mg.L ⁻¹	270	500	860
FDS	mg.L ⁻¹	160	300	520
VDS	mg.L ⁻¹	110	200	340
TSS	mg.L ⁻¹	120	210	400
FSS	mg.L ⁻¹	25	50	85
VSS	mg.L ⁻¹	95	160	315

Source: METCALF & EDDY, 2003.

Comparing the results shown in Table 5.2 and in Figure 5.2 with the data presented in Table 5.3, it is noted that the values for the BOD and COD indicate that the raw sewage have similar characteristics to that rated as “high strength”. However, when analysing the results for solids content, the raw sewage can be classified as “medium strength”. With regard to the pre-treated wastewater, it can generally be classified as “medium strength”, considering its composition in terms of organic material and solids content.

In order to evaluate the variation of the COD in the raw sewage over a typical day of operation (period of 24 hours), a sample collection campaign was once performed (at intervals of 4 hours). The plotted curves, representing the variation over time (i.e. presented as hourly averages) of the influent COD in the raw sewage (raw and filtered samples), are shown in Figure 5.5. The sample collection campaign began at 6:00 PM and lasted for the subsequent 24 hours.

Figure 5.5 – The typical temporal variation in the COD of the raw sewage influent to the experimental WWTP (hourly average): (a) raw samples and (b) filtered samples.

5.4.2 INITIAL TESTS IN THE PILOT SCALE REACTOR

After the first inoculation, the (preliminary) experimental period of tests lasted for 140 days, cycle during which three reinoculations were performed (approx. 15 m³ of sludge in each

opportunity) due to sludge washout. At this stage of the research, three distinct operating modes — defined by trial and error method, according to the data obtained in Phases 1 and 2 of the research — were tested in order to establish the preferential conditions under which the effective operation of the reactor should be performed (Figure 5.6 and Table 5.4). The duration of each stage was defined according to the need for alterations in the operating mode of the system, executed to preserve a satisfactory volume of active biomass within the reaction zone.

Figure 5.6 - Stages set during the exploratory testing and their main operating characteristics.

Table 5.4 – Operating modes imposed during the initial tests on pilot scale reactor.

Stage	Duration (days)	Q ^[1] ($m^3 \cdot d^{-1}$)	θ_h (h)	v_{up} ^[2] ($m \cdot h^{-1}$)	G ^[3] (s^{-1})	Agitation mode ^[4]	Number of inoculations
A	74	55.5	9.0	0.58–0.81	10	C	1
B	40	37.0	13.5	0.39–0.54	2.5	I (on/off: 15/15)	2
C	26	37.0	13.5	0.39–0.54	2.5	I (on/off: 2/18)	1

Concurrently with the initial period of testing in the pilot scale reactor, operational data — including operating temperature (Figure 5.7) — was obtained and enabled the analysis of the unit performance in terms of COD and SST removal efficiencies (Figures 5.8 and 5.9). This stage was important because it allowed the definition of the best operating procedures to allow the maintenance of a stratified sludge blanket and lower occurrence of solids washout within with

¹ The inflow rate — as well as the intensity of agitation — was reduced in an attempt to reduce the solids washout.

² Since the surface area along the height of the reaction zone is not constant, the range of values for the upflow velocity (v_{up}) inside this compartment is presented in Table 5.4.

³ The G value refers only to the period in which the agitation system was on.

⁴ C: continuous mixing / I: intermittent mixing.

the effluent. The tests carried out during these stages enabled the verification of the influence of the mixing intensity on the overall efficiency and operational stability of the system; No sludge recirculation was performed during these initial period of tests.

Figure 5.7 – Fluctuations of the liquid and atmosphere temperatures during the initial period of tests in pilot scale reactor.

Figure 5.8 – Reactor performance during the initial period of tests: COD removal (filtered samples).

Figure 5.9 – Reactor performance during the initial period of tests: TSS removal.

In Figure 5.8 it is observed that, although it varies greatly during the initial test period (31 to 69%), the best result related to the COD removal efficiency was obtained after adopting intermittent agitation with longer intervals between activation of the agitation system (Stage C). It was found that longer interruptions in the rotational movement of the impeller are beneficial to system performance, because it allowed the maintenance of a higher biomass concentration in the digestion compartment (as seen in Figure 5.10).

Regarding the TSS removal (Figure 5.9), it was found that continuous agitation impaired the satisfactory control of this operating parameter. It was also observed that the intermittent mode operation of the mixing system with 20-minute cycles (2-minute operation at each interval of 18 minutes) favoured the “effective” performance of the solid-liquid separation unit, characterized by the ratio between the influent and effluent TSS concentrations of the samples collected in the settling compartment.

At the end of stage C, it can be concluded that the recirculation of the settled sludge (i.e. solids deposited in the bottom of the sedimentation compartment) to the digestion compartment was essential because, besides providing protection against the solids washout, it allows operational flexibility and the maintenance of a higher concentration of active biomass

in the reaction zone. Another conclusion was that the maintenance of the sludge blanket of the stratified structure is fundamental for the achievement of the desired reactor performance standards, since the complete mixing pattern proved to be unstable in the operational control aspect.

To investigate the influence of the different modes of operation on biomass behaviour (or sludge blanket stability), samples were collected along the height of the digestion compartment and other key points of the experimental system and the respective TSS and VSS concentrations were determined; The results are shown in Figures 5.10 and 5.11. The statistical analysis on the variation in the VSS:TSS ratio of each sample collection point, over the period of tests, is shown in Figure 5.12.

Figure 5.10 – Concentrations of TSS and VSS during the experimental period of tests: influent sewage, final effluent and sedimentation compartment.

Figure 5.11 – Concentrations of TSS and VSS during the preliminary period of tests: bottom of the digestion compartment, P1, P2 and P3.

Figure 5.12 – The statistical analysis on the VSS:TSS ratio of each sample collection point, over the experimental period of tests in the pilot scale reactor.

The analysis of the charts shown in Figure 5.10 highlights the instability in controlling the quality of the final effluent (in terms of TSS concentration) in the period in which the system was subject to continuous stirring and the improvement achieved after adoption of intermittent regimen (of agitation). A note becomes opportune: the disparate point found in the chart referring to the suspended solids content in the settled sludge (114th day) is attributed to a failure in the collection of the sample, i.e. it was performed after the daily sludge withdrawal operation.

In Figure 5.11 is observed that during the “Stage A”, characterized by continuous agitation and higher hydraulic loading rate (compared to the subsequent stages), the sludge loss becomes

apparent. The rapid decline in observed in the TSS content of the sludge collected over the reactor bottom indicates that such operating mode could hardly be maintained, due to the loss of sludge. Some sporadic high TSS concentrations observed in P1 (60th day) and P3 (48th day) points were, apparently, consequences of the detachment of the biomass contained over the base of the unit. After further reinoculations and modifications in the mixing regime, the values related to TSS concentrations were maintained relatively stable, although the values in P1 did not reach the desired value of 5,000 mg.L⁻¹.

The VSS:TSS ratio in the influent and effluent were relatively high, reaching 98% (Figure 5.12). In Figure 5.13 is shown the variation in the SST concentration profile within the reaction zone during the initial period of tests.

Figure 5.13 – The variation of the TSS concentration profile inside the digestion compartment throughout the experimental period of tests.

Although the reactor contained few points for collection of samples along the height of the digestion compartment, it can be observed in the profiles shown in Figure 5.12 that the vast

majority of the existing biomass within the unit was established at its lower region (up to 1.5 m level). The increase in TSS concentrations in sludge collected on the bottom of the reaction zone in the final stage of the experimental period (106th, 113th and 140th days), can be attributed to the absence of sludge withdrawal operations in this part of the system.

At the end of initial testing period it was concluded that the recirculation of sludge deposited in the sedimentation compartment back to the digestion compartment and the intermittent mixing were essential for the achievement of better results, since these initiatives enabled the maintenance of a higher concentration of active biomass in the digesting compartment and operational flexibility to system (e.g. allowing some control over the quality of the final effluent in terms of its TSS concentration). The installation and start of the sludge recirculation system operation significantly alter the unit configuration and thus defined a new stage of the research, regarded as the effective period of operation of the pilot scale reactor.

5.4.3 FINAL INOCULATION AND START-UP PROCEDURES

For the start of the effective operation of the pilot scale system, a final inoculation was performed immediately after the closing of the initial period of tests; The capability of system to retain the biomass would be — finally — assessed from this moment.

In the anaerobic sludge used as inoculum, the mean concentrations of total suspended solids and volatile suspended solids were 22,389 mg.L⁻¹ and 13,940 mg.L⁻¹ respectively, resulting in a VSS:TSS ratio of 0.62. This flocculent sludge presented a specific methanogenic activity of 0.034 g CH₄ COD.g⁻¹ VSS.d⁻¹. The initial stage of effective operation of the reactor lasted for 43 days, a period characterized by the presence of high concentration of solids in the effluent, due to the previous introduction of significant amounts of anaerobic sludge (in the last inoculation). During this stage (Stage 1), the system was maintained with hydraulic retention time of 18 hours and intermittent agitation, with cycles of 30 minutes (3 minutes of operation with 27-minute intervals off) and G value of 2.5 s⁻¹.

From the moment in which the final effluent presented a stable condition in terms of solid concentration (visually defined), samples were collected and evaluated (40th day). The confirmation of satisfactory results (Section 5.4.4) stimulated the beginning of a new stage (Stage 2), in which the reactor was subjected to a higher hydraulic loading rate (equal to 13.4 h).

5.4.4 EFFECTIVE OPERATION OF THE SYSTEM AND ITS ROUTINE MONITORING FOR PERFORMANCE EVALUATION (THIRD PHASE)

5.4.4.1 Description of system operation phase

The effective reactor operation lasted for 130 days, without the need of reinoculations. At this phase, four distinct operating modes were tested (including the start-up stage) in order to evaluate the performance of the proposed configuration (Figure 5.14 and Table 5.5). To extract the best possible performance of the unit during the regular period available to the conclusion of the Thesis and to allow comparisons with other systems based on the anaerobic digestion, the duration of each stage was defined according to the results obtained in laboratory tests; As satisfactory results were verified, more challenging conditions were imposed.

Figure 5.14 - Stages set during the system operation and their main operating characteristics.

Table 5.5 – Main features of each stage defined during the operation pilot scale system.

Stage	Duration (days)	Q ($m^3 \cdot d^{-1}$)	θ_h (h)	$v_{up}^{[1]}$ ($m \cdot h^{-1}$)	G ^[2] (s^{-1})	Agitation mode ^[3] (min)
1	43	27.6	18.0	0.29–0.40	2.5	I (on/off: 3/17)
2	29	37.1	13.4	0.39–0.54	2.5	I (on/off: 3/17)
3	14	50.1	10.0	0.52–0.73	2.5	I (on/off: 3/17)
4	44	66.5	7.5	0.69–0.97	2.5	I (on/off: 3/17)

¹ Since the surface area within the reaction zone is not constant, the range of values for the upflow velocity (v_{up}) inside this compartment is presented in Table 5.6.

² The G value refers only to the period in which the agitation system was on.

³ I: intermittent mixing.

As noted in Table 5.5, the influent flow was progressively enhanced in order to evaluate the performance of the treatment system to be subjected to increasingly critical conditions; The agitation conditions were maintained unchanged. In Table 5.6 the data related to sludge recirculation is presented, procedure performed daily over a period of 30 minutes.

Table 5.6 – Data regarding the recirculation procedure: flow rate and recirculation ratio.

Stage	Sewage inflow rate (m ³ .d ⁻¹)	Sludge recirculation flow rate (m ³ .d ⁻¹)	Recirculation ratio ^[1] (%)
1	27.6	1.4	5.1
2	37.1	1.4	3.8
3	50.1	1.4	2.8
4	66.5	1.4	2.1

5.4.4.2 Overall performance of the new system

The operational data — including operating temperature (Figure 5.15) — was obtained and enabled the analysis of the system performance in terms of COD, BOD and SST removal efficiencies (Figures 5.16, 5.17 and 5.18, respectively).

Figure 5.15 – Fluctuations of liquid and atmosphere temperatures during the operation of the pilot scale system.

¹ Recirculation ratio = recirculation flow rate : inflow rate.

Figure 5.16 – System performance during the effective system operation phase: COD removal (raw and filtered samples).

Figure 5.17 – System performance during the effective system operation phase: BOD removal.

From the results shown in Figures 5.16 and 5.17 it can be observed that the system presented a satisfactory performance in terms of organic matter removal efficiency, even with the gradual increase in the hydraulic loading rate (remarkable results were achieved while the reactor was operated with θ_h of 7.5 hours, i.e. Stage 4); The values remained within the range commonly reached by UASB reactors: 51–72% for COD (raw samples) and 72–82% for BOD. The BOD:COD ratio varied in the range of 0.53–0.64 for the influent and between 0.25 and 0.53 for the final effluent.

A limit to the COD value of the final effluent was found, as the results of the tests carried out with different organic and hydraulic loading rates indicated that the effluent COD (raw samples) did not negatively exceed the threshold of 100 mg.L⁻¹ (it varied in the range of 105–193 mg.L⁻¹ over the referred period). From this evidence it can be assumed that the performance of the unit — in terms of COD removal efficiency — only was not higher because of the relatively low content of organic matter in the (pre-treated) sewage influent to the reactor (the values varied between 321 mg.L⁻¹ and 426 mg.L⁻¹).

Figure 5.18 – System performance during the effective reactor operation phase: overall and effective TSS removal efficiency.

With regard to the TSS removal (Figure 5.18), the use of intermittent stirring and sludge recirculation proved to be a more suitable operational strategy, since goods results were verified

from the stage 2 onwards, in both overall and effective efficiencies (66 to 83% and 89 to 98%, respectively). In certain way, the possibility of removal of the settled sludge (via continuous or intermittent recirculation) prevents an excessive build up of solids and its dragging by the upward flow established in the settling zone. Just as in the initial period of testing, samples were collected in order to evaluate the stability of the sludge blanket; The TSS and VSS concentrations were determined. The results are shown in Figures 5.19 and 5.20. The statistical analysis on the variation in the VSS:TSS ratio of each sample collection point, over the reactor operation phase, is shown in Figure 5.21.

Figure 5.19 – Concentrations of TSS and VSS during the system operation phase: influent sewage, final effluent and in the sludge collected within the sedimentation compartment.

Figure 5.20 – Concentrations of TSS and VSS during the system operation phase along the height of the digestion compartment.

Figure 5.21 – The statistical analysis on the VSS:TSS ratio of each sample collection point, over the system effective operation phase (Phase 3).

The results presented in Figure 5.19 indicate a significant improvement in the quality of the final effluent, in terms of concentration of suspended solids. It was verified, over the last 100 days of operation of the system, that the TSS and VSS concentrations remained constantly below 40 mg.L⁻¹, a result that can be considered excellent, hardly achieved by UASB reactors operating under similar conditions.

In Figure 5.20 it can be observed that the initiative to recirculate the sludge from the sedimentation compartment to the area of digestion, assured the maintenance of biomass in the system, without the need for reinoculation. Attention should be given to the SS concentrations

of collected at point P1, which remained well above the pre-set limit of $5,000 \text{ mg.L}^{-1}$, defined as control condition during the whole period of tests in the experimental system. At the end of the operation phase it could be observed an increase in the SS concentrations at P2 and P3, which might be influenced by the increased temperatures during this period, resulting in high microbial activity and thus enhanced biogas production. The variation of the SS concentration at P3 justifies the need for a clarifier, since the high solids content would impair the effluent quality.

The analysis presented in Figure 5.21 indicates (again) the occurrence of very high values for the VSS:TSS ratio for both influent and effluent, which still presented values of 98% The results also indicate the increase of the VSS:TSS ratio along the height of the reactor (from bottom to P3).

Figure 5.22 – Photographic record of sample filtration assay for determination of TSS: (a) influent; (b) the effluent from the reaction zone and; (c) final effluent.

5.4.4.3 The Performance of the Reaction Zone and Further Investigations

The performance of the reaction zone and the behaviour of the sludge blanket contained therein was evaluated in search of information that could indicate the best operating strategy to be adopted, especially with regard to the agitation regime. In addition to the data previously presented (Sections 5.4.4.1 and 5.4.4.2) it was decided to investigate some aspects considered to be relevant for understanding the processes and phenomena occurred inside the digestion compartment.

TSS concentrations

Since the TSS concentration in the influent is relatively low and no specific device for sludge retention was installed inside the digestion compartment of the reactor, an abatement on the suspended solids content of its effluent is not expected and the dragging of solids is a natural occurrence (Figure 5.23). Clearly the function of retaining the solid particles dragged within the effluent will be performed by the unit attached to the reactor, which in the case, was a high-rate clarifier.

Figure 5.23 – TSS concentrations within the reaction zone.

In order to demonstrate the distribution of biomass within the digestion compartment and its variation along the unit operation period, the TSS profiles determined during this phase are presented in Figure 5.24.

Figure 5.24 – The TSS profiles during the system operation phase.

The chart presented in Figure 5.24 indicates a significant reduction in the TSS concentration in the sludge collected over the bottom of the reactor from the 55th day. However, samples collected on the 124th and 130th days indicated a trend of increasing TSS concentration in P3 (placed 4.5 m over the base of the unit), an increase probably caused by the displacement of a portion of the biomass originally established in the lower region of the reactor. Apparently, these changes are associated with the higher mixing intensities caused by the increased upflow velocities and biogas production^[1] — induced by the gradual increment in the hydraulic and organic loading rates, respectively — imposed during the development of the research.

Cell residence time (θ_c)

From the SSV profiles for each stage which constituted the effective period of operation of the reactor, it was possible to estimate the amount of active biomass present in the reactor and, consequently, the cell retention time (Table 5.7).

Table 5.7 – Data regarding the recirculation procedure: flow rate and recirculation ratio.

Stage	Sewage inflow rate (m ³ .d ⁻¹)	VSS mass within the reactor (kg)	VSS mass carried out within the effluent (daily) (kg)	VSS mass discharged (daily) (kg)	θ_c (d)
1	27.6	261.25	0.72	–	363
2	37.1	261.62	1.19	–	220
3	50.1	263.11	1.50	–	175
4	66.5	219.63	2.13	–	103

Despite being operated for a short time, the system demonstrated its ability to retain biomass efficiently since that, even operating under θ_h of 7.5 h, it showed higher θ_c value than those presented by UASB reactors operated under similar conditions.

Organic matter removal

The COD determination of the samples collected along the reactor height allowed the examination of the behaviour of the unit in relation to its performance regarding the removal of organic matter as the substrate flows through the digestion compartment. Due to the high

¹ A higher intensity of agitation may cause disturbances in the sludge mass located in the bottom of the reactor, causing the expansion of that portion of the biomass in a way that these solids are able to reach higher sectors within the reaction zone.

TSS concentration within the reaction zone, COD could only be determined by use of filtered samples. The results are shown in Figure 5.25, in a chart that presents the COD and VSS concentration at each point.

Figure 5.25 – COD (filtered samples) and VSS concentration along the system.

The analysis of the chart shown in Figure 5.25 clearly allows the establishment of a similar trends between the organic matter removal rate and the VSS concentration; the greater the presence of active biomass in the reactor at a point, the greater the reduction in COD therein (e.g. the highest organic matter removal rates occurred at the lower region of the reactor, between the bottom and P1). Based on the same chart it can also be hypothesized that the increase in COD verified between the influent box and the bottom of the reactor is apparently attributed to the solubilization of components contained in the sewage.

The pH profile

The monitoring of the pH of samples collected along the height of the reaction zone allowed the construction of a chart (Figure 5.26).

Figure 5.26 – The pH and the VSS concentrations along the system.

The sudden drop in pH verified between the influent box and the reactor base indicates a probable intense acidogenesis, characterizing one of the early stages of degradation via anaerobic process.

The H₂S profile within the system

The H₂S profile of in the system, created through the analysis of samples collected along the system and the height of the digestion zone. is presented in Figure 5.27.

Figure 5.27 – The H₂S profile and the VSS concentrations along the system.

The concentrations of hydrogen sulphide over the system are influenced by the presence of biomass. The sudden increase in the H₂S concentration between the sewage inlet box and the bottom of the reactor confirms a probable intense acidogenesis at that point.

The phosphate profile

In Figure 5.28 a chart containing data related to the phosphate concentrations measured at each point of the system is shown; including those located along the height of the digestion compartment.

Figure 5.28 – The PO₄³⁻ profile and the VSS concentrations along the system.

As well as other variables, phosphate also varies in correspondence to the activity of VSS concentration. An apparent intense acidogenesis should also influence the sudden increase observed in the PO₄³⁻ content between the sewage inlet box and the bottom of the reaction zone.

As known by the professionals involved in wastewater treatment of sewage, the anaerobic process presents low efficiency associated with phosphorus removal. In the region of low pH and low H₂S concentration there is more solubilization of phosphorus, or phosphate decomposition. Apparently high solubilization occurs, and then, when the pH rises slightly, adsorption in the solid mass (micro-organisms + solids) which forms the flocs present in the reactor.

5.4.5 THE PERFORMANCE OF THE HIGH-RATE SETTLER

During the reactor operation phase the high-rate clarifier showed satisfactory results; His “effective” performance was determined by performing comparison between the TSS concentrations in the influent (reaction zone effluent) and the final effluent. The TSS removal efficiency during this period ranged from 89 to 98%.

In order to present the conditions imposed on the high-rate clarifier, the Hazen velocity (v_H) and the hydraulic loading rates (HLC_c) applied are presented in Table 5.8, calculated by applying the Equations 3.12 and 4.3, respectively; The upflow velocity within the channels formed by the plates was calculated based on Equation 4.4.

Table 5.8 – Hydraulic conditions imposed on the high-rate clarifier.

θ_h (h)	Flow rate ($\text{m}^3 \cdot \text{d}^{-1}$)	v_H (Equation 3.12) ($\text{m}^3 \cdot \text{m}^{-2} \cdot \text{d}^{-1}$)	HLC_c (Equation 4.3) ($\text{m}^3 \cdot \text{m}^{-2} \cdot \text{d}^{-1}$)	Velocity of the liquid within the plates ($\text{m} \cdot \text{h}^{-1}$)
18.0	27.65	4.53	37.57	1.61
13.4	37.15	6.09	50.48	2.17
10.0	50.11	8.22	68.09	2.92
7.5	66.53	10.91	90.39	3.88

Despite being designed to operate under hydraulic loading rates of up to $80 \text{ m}^3 \cdot \text{m}^{-2} \cdot \text{d}^{-1}$, the high-rate settler presented impressive performance when subjected to rates of $90 \text{ m}^3 \cdot \text{m}^{-2} \cdot \text{d}^{-1}$ (TSS removal efficiencies shown in Figure 5.18).

5.4.6 THE RESULTS OF THE COMPLEMENTARY ASSAYS FOR SLUDGE CHARACTERIZATION

5.4.6.1 Specific Methanogenic Activity

The results of the assays for determination of the specific methanogenic activity of the sludge collected inside the digestion compartment are presented in Table 5.9.

Table 5.9 – Specific methanogenic activity: results of the sludge collected within the reactor.

Sample	Days of operation	θ_h (h)	VSS ($\text{mg} \cdot \text{L}^{-1}$)	SMA ($\text{g COD}_{\text{rem}} \cdot (\text{g VSS})^{-1}$)
P0 (bottom)	84	10.0	44,784	0.037
P1 (1.5 m)	84	10.0	20,962	0.022
P0 (bottom)	130	7.5	44,876	0.041
P1 (1.5 m)	130	7.5	14,252	0.043

Although no tendency was identified, it was observed that both samples showed equivalent results at the 130th day, operating under θ_h of 7.5 hours. Although it contained a higher concentration of biomass, the sludge collected over the bottom of the unit may have its syntrophic relationship impaired by the presence of inert solids, for example.

5.4.7 CRITICAL ANALYSIS AND DISCUSSION

By analysing the results presented by UASB reactors treating domestic sewage all over the world, it can be observed that the performance of the new system — in terms of removal of suspended solids and organic material — is comparable to these units operated under similar conditions (Table 3.2). Due to a number of factors the system could not be operated for longer period, but the initial results showed its high potential for application in wastewater treatment systems. The results obtained with the implementation of the new system are promising; The results could have been better if the influent organic load to the system and influent TSS concentrations were higher; The low organic content within the substrate did not allow higher removal rates.

In order to better understand the sludge blanket behaviour, more sample collection points should have been installed along the height of the digestion compartment (at least at intervals of 0.5 m), a fact made impossible due to the high cost. The existence of only four points for sample extraction did not allow the determination of more accurate profiles, which would be of great value. A factor which stands out as a positive aspect of this system, in relation to UASB units, is the operational flexibility made possible through the recirculation of the settled sludge (in the sedimentation chamber) to the digestion zone. Operating in conjunction with the agitation system, the initiative to reintroduce the biomass in the reaction medium can provide higher removal rates of organic matter by extracting the maximum potential associated with anaerobic micro-organisms, without impairing the performance of the clarifier.

In order to illustrate the application of A²SB technology on a larger scale, it is shown in Figure 5.29 the theoretical design of the mechanical agitation system of a unit designed to serve a population of approx. 40,000 inhabitants (four impellers would be installed to replace the traditional influent distribution system).

Figure 5.29 – Scale-up exercise: theoretical example of the replacement of the UASB influent distribution system (a) by four rotating impellers (b).

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CONCLUSIONS AND RECOMMENDATIONS

6

Even in its early stages of development, the implementation of the system disclosed in this research proved its applicability in the treatment of sewage, with satisfactory performance even under the most critical condition under which it was operated (θ_h 7.5 h). The installation of devices such as mechanical mixer, high-rate clarifier and recirculation system, allowed this sludge blanket reactor to operate without the three-phase separator and influent distribution system.

The intermittent operation of the impeller (mixing system) under relatively low velocity gradients was presented as the best operating alternative, allowing sludge retention in the system and hence higher removal rates of organic matter and suspended solids.

The high-rate clarifier presented a high efficiency at removing suspended solids, i.e. up to 98% considering the influent from the digestion zone, when operated under θ_h of 7.5 h and temperature of $(29.1 \pm 0.6)^\circ\text{C}$, showing that the reactor can operate properly even without the phase separation.

The introduction of the sludge recirculation system (from the sedimentation compartment to the reaction zone) allowed the maintenance of satisfactory amount of biomass within the system (and no washouts — Figure 6.1), and preventing its accumulation to impair the performance of the settler. The new system, Anaerobic Activated sludge Blanket (A^2SB), promoted removal of organic matter at rates up to 78% (BOD) and 72% (COD), when operated under θ_h of 7.5 h and temperature of $(29.1 \pm 0.6)^\circ\text{C}$.

Figure 6.1 – Schematic illustration of the research stages: main characteristics of the distinct operation modes and inoculations.

The proposed system allows flexibility and great potential for automation of its operation by the introduction of technological components (available in the market, such as sensors) that could enable real-time diagnostics, for the optimal control of mechanical agitation and recirculation system.

From the results and conclusions presented, the author of the study makes the following recommendations for the development of future research based on the technology proposed in this work::

- The use of Computational Fluid Dynamics (CFD) for the search of the best mixing alternatives;
- The use of the A2SB system in large scale units or even at full-scale WWTPs;
- The study of the physical properties of the sludge, e.g. viscosity, specific mass, and its relations with the solids content presented by it;
- The evaluation of the use of plastic carriers for biofilm fixation within the anaerobic reactor;
- The development or improvement of the automatic controlling system.

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