EPAS n.v.
Eco Process Assistance
De Prijkels • Venecoweg 19 • B-9810 Nazareth
Tel. +32 9 381.51.30
Fax +32 9 221.82.18
www.epas.be • epas@epas.be

ENGINEERING OF THE WASTE COMPARTMENT

ESA contract 15689/01/NL/ND

TECHNICAL NOTE 71.1

Requirements, Performances and Sizing of the MELiSSA Waste Compartment

Version : 2
Issue : 1

	Name	Signature
Prepared by:	Noëlle Michel	
	Farida Doulami	
	Veronik Hermans	
Approved by:	Dries Demey	

18 July 2003

DOCUMENT CHANGE LOG

Version	Issue	Date	Observation
1	0	30 July 2002	Draft
2	1	18 July 2003	Final

DISTRIBUTION LIST

Quantity	Company/Department	Name
4	ESA	C. Lasseur
1	EPAS	D. Demey
		H. Vanhooren
		F. Doulami
		V. Hermans
		N. Michel
1	ADERSA	J. Richalet
1	UAB	J.Albiol
		F. Godia
1	UBP	C.G. Dussap
		L. Poughon
1	VITO	H. De wever
1	LabMET	W. Verstraete
1	U. Guelph	M.A. Dixon
1	SCK	M. Mergeay

CONTENT

LIST OF FIGURES

LIST OF TABLES

1. Introduction

The liquefying compartment in the MELiSSA loop is responsible for the biodegradation of human faecal material and other wastes generated by the crew. The volatile fatty acids and ammonia produced during the anaerobic fermentation process are fed to the second phototrophic anoxygenic compartment with *Rhodospirillum rubrum*. The CO2 that is produced is supplied to the photosynthetic *Arthrospira platensis* compartment and to the higher plant compartment.

The liquefying compartment is the first step in the MELiSSA loop and determines the fraction of organic wastes that can be recycled in the loop.

At the pilot plant of the University 'Autonoma' of Barcelona, three compartments of the MELiSSA loop (compartment II, III, IVa) were connected at lab scale and these three compartments will be validated at pilot scale in 2003. In order to validate the whole MELiSSA loop, it is necessary to construct a first compartment at pilot scale.

In this technical note the general requirements, the performances and the sizing of the anaerobic waste compartment are presented. The concept of the waste compartment proposed in the EWC proposal is represented in Figure 1.

2. Requirements

The reactor needs to be operated at 55°C (thermophilic conditions) and at a pH of 6 to avoid methane production and pathogens proliferation. The reactor will be operated in anaerobic conditions. The biodegradation efficiency of the reactor should be at least 55 %. The reactor needs to be provided with several interfaces among which the solid loop with the waste material, the gas loop to transport the produced CO_2 to compartments C IVa and C IVb and a liquid loop including the the filtration unit. The latter will be equipped with a membrane system. The task of the membrane is to separate the non-biodegradable organic matter from the soluble fractions expressed as volatile fatty acids (VFA), ammonium and minerals. It is very important, for safety reasons and to avoid contamination of the photoheterotrophic compartment, to retain the bacteria and viruses present in the bioreactor by the membrane filtration. For this purpose an ultrafiltration membrane technology will be the most appropriate. Table 1 summarises the requirements of the reactor.

Table 1. Reactor requirements

Parameter	Requirement	Reason
1) Reactor objectives		
Feed = FM 1pers/d + plants + toilet paper = 210 gDW/d	to degrade it	optimisation of CI
2) Reactor content		
[DM]	<50g/L	optimisation of FU
[N]	<3g/L	avoid acidogenic bacteria inhibition
SRT	>20d	optimisation of acidogenesis
VFA	maximise production	optimisation of acidogenesis
CO_2	maximise production	optimisation of acidogenesis
3) Process parameters		
pH	<6,0	inhibition of methanogenesis
T	55°C	optimisation of acidogenesis inhibition of pathogens
O_2	absent	optimisation anaerobic process

3. Substrate composition and preparation requirements

3.1 Substrate composition

The MELiSSA Pilot Plant will need a standardised composition of the waste. At the MELiSSA meeting on the 29-30th of November and at the EWC Progress meeting on the 16th of January the waste composition was determined. The waste will consist of faecal material of one man and non-edible parts of higher plants, when the Higher Plant Compartment is calculated to provide 20% of a one-man diet.
The urine produced by one man will not be treated in the first compartment due to the risk of high ammonia concentrations in the reactor, which are toxic for the bacteria.
One man a day produces 30 g DW of faecal material and about 180-190 g DW of non-edible parts of higher plants.
The selected plant material, are wheat straw, lettuce and beet. The total amount and the ratios are represented in Table 2. Per day about 54 g DW of non edible parts of lettuce, 54 g DW of non edible parts of wheat straw, 54 g DW of non edible parts of beet and 18 g of toilet paper need to be processed in the reactor.

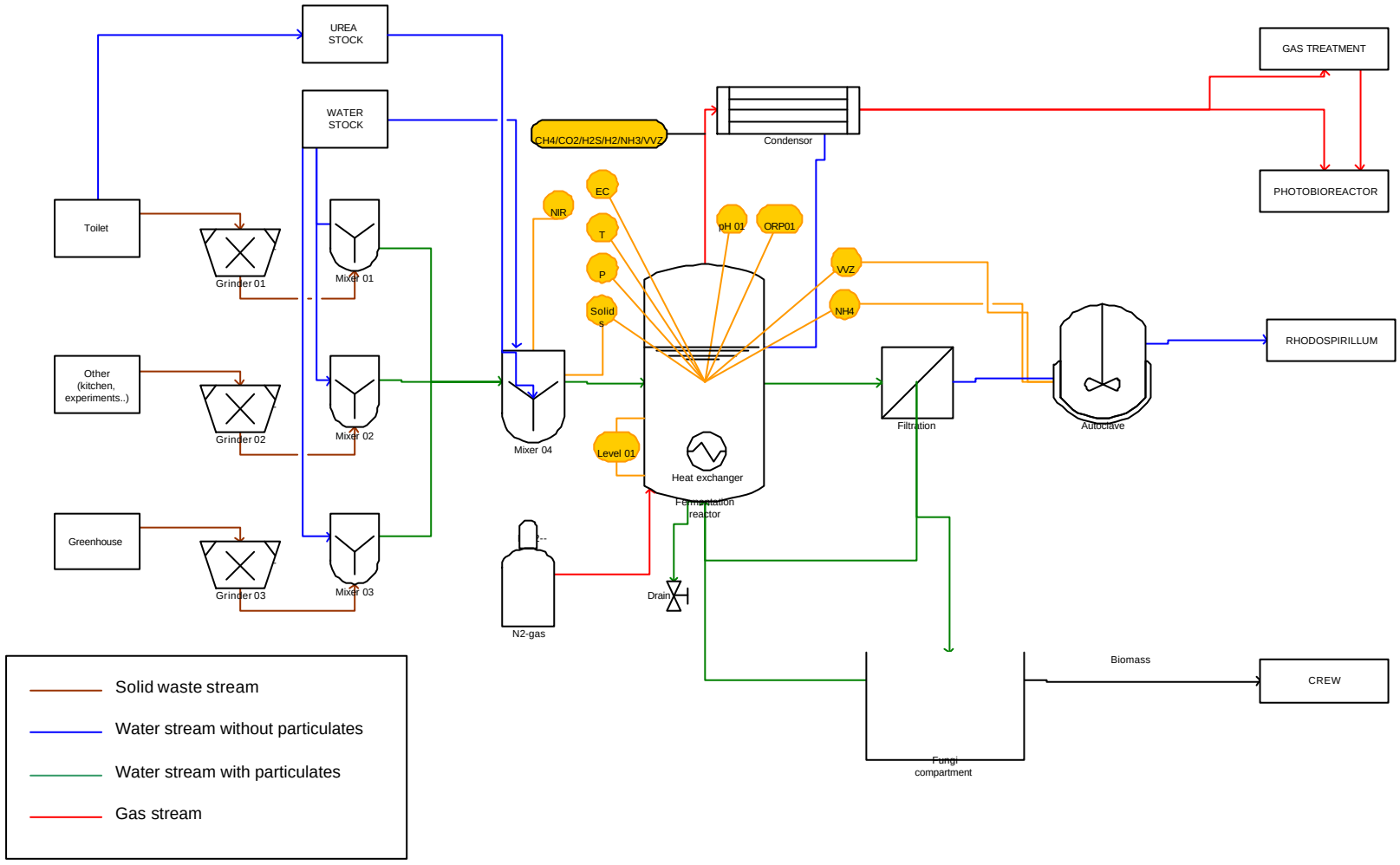

Figure 1. Concept of the MELiSSA anaerobic waste compartment (CI)

Table 2. Composition of feed of compartment I

Material	Amount (g/d)	DW	Percentage (%)
Lettuce	54		30
Beet	54		30
Wheat Straw	54		30
Toilet paper	18		10
Total plants and paper	***180***		***86 %***
Faecal material	***30***		***14 %***
Total amount of material	**210 DW g/d**		**100 %**

Since 94.5% of the lettuce is water, 910 g lettuce (wet weight) need to be processed a day in order to reach the fixed amount of 54 g DW. This corresponds with the processing of two crops a day or 14 crops a week.

3.2 Substrate preparation requirements

The non-edible parts of the selected higher plants need to be pre-treated, before introducing the material into the first compartment. It is important to provide a material with a diameter of around 0.5 mm to avoid clogging of the filtration unit and to facilitate the biodegradation of the material by the anaerobic bacteria.
During the MELiSSA meeting on the 29-30th of November it was decided to pre-treat the plant material in its natural state, meaning wet, for the non-edible parts of the beet and the lettuce and dry for the wheat straw.
The plant waste will be grounded first and then frozen to preserve the material. Part of the waste will be dried for analysis purpose only.

4. Process engineering

4.1 MELiSSA loop scheme

Figure 2, recapitulates the conditions required for the running of the pilot plant

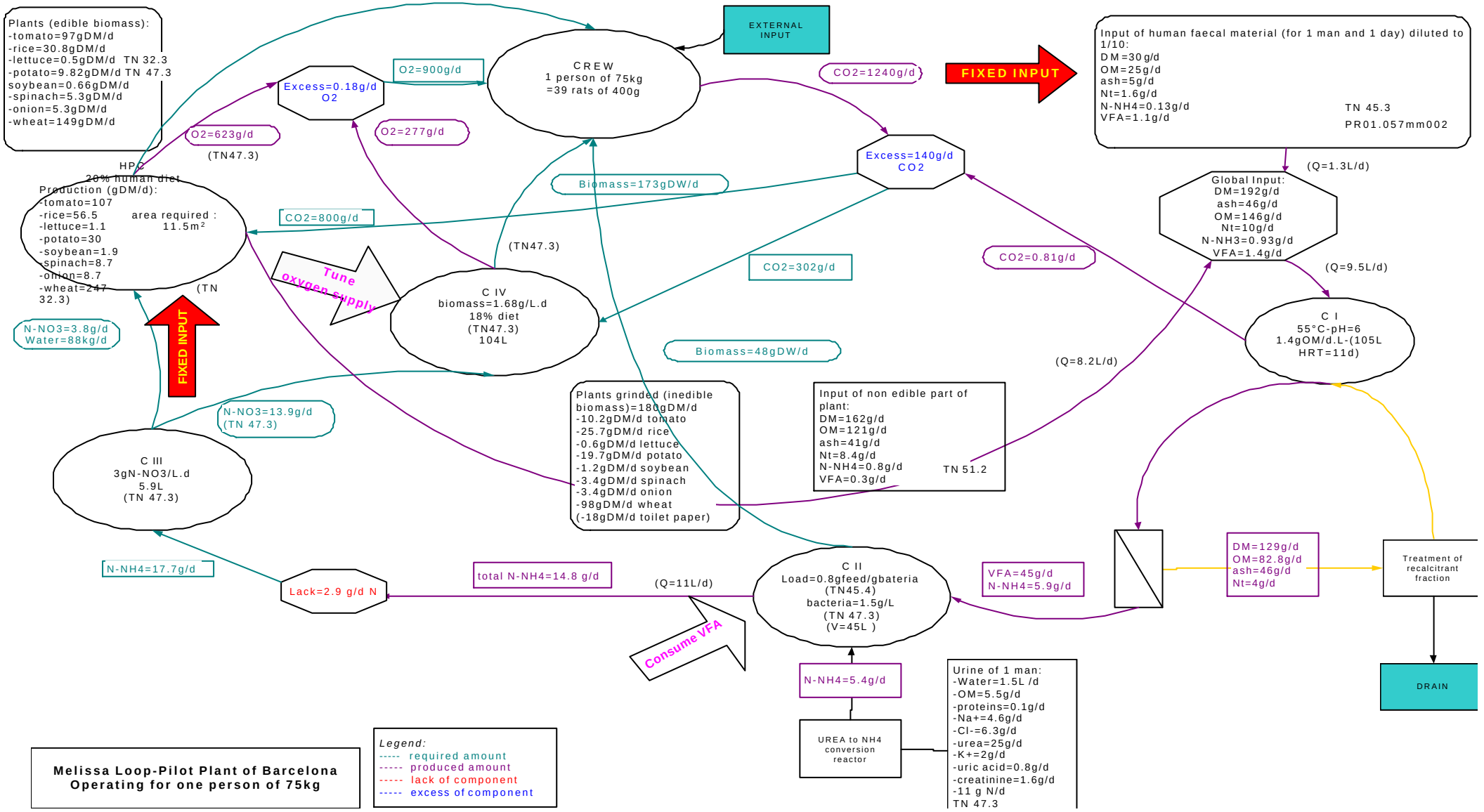

Figure 2. MELiSSA loop scheme

The aim of the loop is to provide sufficient oxygen for one man of 75kg, and to produce 20% of his diet. The loop is here equilibrated to provide enough oxygen needed by one man and enough carbon dioxide needed by the algae and plants from the determined influent. It implies a lack of nitrogen for the Higher plants compartment. The corresponding algae production from compartment IV represents then 18 % of the human diet.

4.1.1 Compartment I (Liquifying reactor)

4.1.1.1 Input

The global input was determined during the EWC progress meeting on the 16th of January and consists of:
-Human faecal material of one man per day: 30gDW/d
-Non-edible parts of plants (edible parts must cover 20% of the diet of one man per day): 162gDW/d.

4.1.1.1.1 Faecal material (FM)

Table 3 shows the composition of faecal material.

Table 3. Composition of human faecal material diluted to 1/10: (TN 45.3)

Human faecal material components	Concentrations (g/L)
dry matter	23
ash	3.7
OM	19.3
N total	1.24
NH4-N	0.1
VFA	0.85

The mass flow rates x of each component of human faecal material can be determined using the formula:

$$x(g/d) = \frac{C_x(g/L) \cdot DM(g/d)}{C_{DM}(g/L)}$$

where: C_x = concentration of X
DM = mass flow rate of dry matter
C_{DM} = concentration of dry matter

Table 4 gives the mass flow rates of faecal material compounds.

Table 4. Mass flow rates of human faecal material

Human faecal material components	Mass flow rates (g/d)
dry matter	30
ash	4.8
OM	25.2
N total	1.62
NH4-N	0.13
VFA	1.11

4.1.1.1.2 Non edible parts of plant material (NEPM)

A previous experiment (TN 51.2) is used for the calculations. The input consisted in a mix of 0.05gDM/d faecal material and 0.8gDM/d NEPM, meaning a ratio FM/NEPM=0.84, with a flow rate of 0.043L/d.
The composition and flow rates of this mix (mass flow=concentration x volumetric flow) are presented in Table 5.

Table 5. Composition and flow rates of the mix (TN 51.2)

Components	Concentrations (g/L)	Mass flow rates (g/d)
dry matter	20	0.9
ash	4.9	0.2
OM	15.1	0.65
N total	1.04	0.05
NH4-N	0.1	0.004
VFA	0.08	0.003

In this mix, the NEPM represents 16/17. The composition and flow rates of NEPM in this mix are presented in Table 6.

Table 6. Composition and flow rates of NEPM (TN 51.2)

Components	Concentrations (g/L)	Mass flow rates (g/d)
dry matter	18.8	0.8
ash	4.6	0.2
OM	14.2	0.6
N total	1	0.04
NH4-N	0.1	0.004
VFA	0.07	0.003

As the flow rate of dry matter of NEPM in the pilot plant is fixed to 162g/d, the flow rates of each compound can be calculated (Table 7).

Table 7. Flow rates of NEPM in the pilot plant

Components	Mass flow rates (g/d)
dry matter	162
ash	40
OM	122
N total	8.4
NH4-N	0.83
VFA	0.62

4.1.1.1.3 Global input

The global input corresponds to the sum of faecal material and non-edible plant material.

Table 8. Global mass flow rates in the pilot plant

Components	Global mass flow rates (g/d)
dry matter	192
ash	45
OM	148
N total	10
NH4-N	1
VFA	1.7

In this case, the volumetric flow rate Q (L/d) is:

$$Q=Q_{NEPM}+Q_{FM}= q_{FM}/C_{FM}+q_{NEPM}/C_{NEPM}$$

Where: q_{FM} = massic flow rate of FM (g/d)
C_{FM} = concentration of FM in the influent (g/L)
q_{NEPM} = mass flow rate of NEPM (g/d)
C_{NEPM} = concentration of NEPM in the influent (g/L)
Q = 9.5L/d

The load of the reactor is estimated to be 1.4gOM/d.L (where OM = Organic Matter). The volume (V) can be calculated by dividing the global organic matter flow rate by the load:

$$V = 148/1.4 = 105L.$$

This corresponds to the following hydraulic residence time (HRT):

$$HRT = V/Q = 11.2 d.$$

4.1.1.2 Output

4.1.1.2.1 Output of faecal material

It can be estimated from the experiment described in TN 45.3.

Table 9. Calculation of the volumetric flow rates for the output of FM

Components	Input (g/d) (TN45.3)	Output (g/d) (TN 45.3)	Input pilot plant (g/d)	Output pilot plant (g/d)
dry matter	23	18	30	23.5
ash	3.7	3	4.8	3.9
OM	21.8	15	25.2	19.6
N total	1.24	1.25	1.62	1.63
NH4-N	0.1	0.7	0.13	0.9
VFA	0.85	2.4	1.11	3.13
CO_2 (g/L reactor)	0	0.6	0	0.77

4.1.1.2.2 Output of NEPM

Based on the previous experiment TN 51.2, the flow rates of NEPM output can be determined. The concentrations taken into account are averages of results between 26th and 68th days, which correspond to a representative period.

The mass flow rates (q) are determined with the formula:

$$q\ (g/d) = C \times Q$$

where: C = concentration of the compound (g/L)
Q = volumetric flow rate (L/d)

Table 10. mass flow rates of NEPM (TN 51.2)

Components	Output (g/L)	Output (g/d)
dry matter	14.6	0.63
ash	5	0.22
OM	14.1	0.61
N total	1	0.043
NH4-N	0.6	0.026
VFA	0.6	0.026

The output of NEPM in the mix corresponds to the global output minus the output of FM.

Table 11. Output of NEPM (TN 51.2)

Components	Output (g/d)
dry matter	0.6
ash	0.21
OM	0.58
N total	0.04
NH4-N	0.024
VFA	0.021

This mix corresponds to 0.8gDM/d NEPM; in the pilot plant, the flow of NEPM is 162 gDM/d. Table 12 gives the deduced output.

Table 12. Output of NEPM in the pilot plant

Components	Output (g/d)
dry matter	119
ash	42.2
OM	77
N total	8.2
NH4-N	4.9
VFA	41.7

According to the results of TN 51.2, the biogas production was 2800mg in 150 days, with 46% CO2. The flow rate of CO2 is thus in the pilot plant:

$$CO2 = 0.009 g/d \times (16/17) \times (162/192) = 0.007 g/d$$

4.1.1.2.3 Global Output

The global output mass flow rate corresponds to the sum of faecal material and non-edible plant material (Table 13).

Table 13 Global mass flow rates of the output in the pilot plant

Components	Output (g/d)
dry matter	142.7
ash	46.1
OM	96.6
N total	9.8
NH4-N	5.83
VFA	45
CO2	0.81

4.1.2 Compartment II

4.1.2.1 Input

As a membrane filtration unit follows the compartment I, only the soluble organic matter joins the second compartment, meaning the volatile fatty acids (45g/d) and the ammonia (5.8g/d). Moreover,

the urine of one person per day could be treated in compartment II (40.4 gDM/d), towards to reduce the lack of nitrogen in the loop. This represents a global flow rate of 91.3 gDM/d.

As the load is estimated to 0.8 gfeed/gbacteria (TN 45.4) and the concentration in bacteria is estimated to 1.5 g/L (TN 47.3), the volume required for compartment II should be 91.3/0.8/1.5 = 76L.

4.1.2.2 Output

According to TN 47.3, the decomposition of urine provides 12.7 gN-NH4/d. Furthermore, 0.03gN-NH4/h are produced from 0.32gVFA/h, meaning a production of 4.5gN-NH4/d from the input of compartment I. All the volatile fatty acids are supposed to be consumed.

4.1.3 Crew

In the pilot plant, for safety reasons the crew is represented by rats. The number of rats which oxygen consumption corresponds to the consumption of one man per day must be determined.

According to the Chambure (1992) the O_2-consumption of 1 rat is about 1 L /(kg.h) (normal laboratory conditions; restful rat). According to Preud'homme (1997) the O_2-consumption of a rat varies between 0.8 L /(kg.h) and 1.7 L /(kg.h) (depending on the activity of the rat). For the calculations the highest value (1.7 L/(kg.h)) was taken.

According to the Chambure (1992), the O_2-consumption of 1 man can vary between 0.2 L / (kg.h) in restful position and 4 L /(kg.h) in hard working conditions. The value of 0.35 L/ (kg.h) was arbitrarily chosen. Assuming that the average body weight for men is 75 kg, the O_2 consumption is thus 630 L / (person.d). This corresponds to the consumption of 39 rats of 400g: 900 g O2/d. With a respiratory ratio of 1, the crew produces 1240 g CO2/d.

4.1.4 Higher Plants Compartment (HPC)

According to the repartition of the eight plants chosen for the HPC in TN 32.3, the following results are obtained:

Table 14 Calculation of amounts of plants required in the pilot plant

Plants	% of dry matter of total edible plants	Weight of DM edible plant (g/d.man)	DM Waste/ DM edible	Weight of DM inedible plant (g/d.man)	Total weight of plant (g/d.man)	Weight of inedible plants required in the pilot plant (gDM/d)	Weight of edible biomass (g/d)	Weight of total biomass (g/d)
Tomato	0.8	3.4	9.6	33	36	10.2	97	107
Rice	16	70	1.2	83	152	26	30	56
Lettuce	0.5	2.2	0.9	1.8	4	0.6	0.5	1.1
Potato	30	127	0.5	63	190	20	9.8	29
Soybean	1.6	6.9	0.6	3.8	10.7	1.2	0.7	1.9
Spinach	1.6	6.9	1.6	10.8	18	3.4	5.3	8.7
Onion	1.6	6.9	1.6	11	18	3.4	5.3	8.7
Wheat	48	207	1.5	316	523	100	150	247
Total	100	430		523	952	162	300	460

According to TN 47.3, 18.5 gDM/d of total plants require 32 gCO2/d and produce 25 gO2/d, meaning in the MELiSSA loop a consumption of 800 gCO2/d and a production of 623 gO2/d.

For 18.5 gDM/d of total plants, 0.15gN-NO3/d are required, meaning a consumption of 3.8 gN-NO3/d.

4.1.5 Compartment IV

As no specific requirement has been determined for the production of the fourth compartment until now, the biomass production is fixed arbitrarily to cover 18% of the diet of the crew, which allow to equilibrate the O_2/CO_2 balance. This production corresponds to 175 g/d (as 40% of the diet corresponds to 10 gbiomass/rat of 400 g.d). The resulting production of O_2 is 277 g/d and the consumption of CO_2 is 302 g/d. Furthermore, an amount of 13.9 gN-NO3/d is required (based on TN 47.3).

4.1.6 Compartment III

The compartment III is supposed to produce 3 gN-NO3/L.d (TN 47.3). It should provide an amount of 17.7 g/d of N-NO3 which is required by the HPC and compartment IV, this implies a volume of 5.9L. The resulting balance for nitrogen is slightly negative: there is a lack of 0.5 gN/d.

This scheme allows to show how the loop can be equilibrated: all the parameters must be evaluated to make sure that there is no lack of any components in the loop. To work on it more easily, tables in Excel were established to see directly the consequence of the change of one parameter on the rest of the loop.

4.2 Dynamic modelling of the waste compartment

A simple model can be written to represent the digestion occurring in the first compartment, according to the IAWQ models.

4.2.1 Nomenclature and parameters involved in the model

The following nomenclature is used with the IAWQ models:

Table 15. IAWQ nomenclature

Nomenclature	Meaning	Input	Output
X	solid components		
X_A	autotrophic biomass		
X_B	biomass		
X_E	enzymes		
X_H	heterotrophic biomass		+
X_M	methanogens		
X_{ON}	organic nitrogen	+	
X_{PAO}	phosphorus accumulating organisms		
X_{PHA}	polyhydroxyl-alkanoates		
X_{PP}	polyphosphate		
X_S	insoluble organic carbon	+	
X_{Ac}	acidogenic biomass		
S	components dissolved in wastewater		
S_A	Volatile fatty acids	+	+
S_F	fermentable chemical oxygen demand	+	
S_N	nutrient		

S_{NH}	ammonia	+	+
S_{NO}	organic matter containing nitrogen	+	+
S_O	dissolved oxygen		
S_{PO4}	dissolved phosphorus		
S_S	soluble carbon	+	+

All parameters used for the model are summarised in Table 16.

Table 16. Parameters used in the model

Parameters	Values	Units	References
Reaction rates			
• r_{XAc}		g X_{AC}/d	
• r_{XE}		g X_E/d	
• r_{Xs}		g Xs/d	
• r_{Ss}		g Ss/d	
• r_{XON}		g X_{ON}/d	
• r_{SF}		g S_F/d	
• r_{SNO}		g S_{NO}/d	
• r_{SA}		g S_A/d	
• r_{SNH}		g S_{NH}/d	
• r_{CO2}		g CO_2/d	
Maximum specific growth rate			
• μ_m	3	d^{-1}	Design of anaerobic processes, 1992
Volumetric flow rates			
• q		L/d	
• q_{FDC}			
• q_{drain}			
Volume of the reactor			
• V		L	
Dilution rate			
• D		d^{-1}	
Efficiency of the FDC			
• h_{FDC}			
Decay of bacteria			
• d_{Ac}	0.0625	-	
Half-saturation constant / Affinity constant			

• K_{SNO}	0.2	g S_{NO}/L	Design of anaerobic processes, 1992
• K_{SF}	0.2	g S_F/L	Design of anaerobic processes, 1992
Kinetic constants			
• $k_{Ac,Xs}$		g S_F/ g Xs . d	
• $k_{Ac,Ss}$		g S_F/ g Ss . d	
• $k_{Ac,Xon}$		g S_{NO} / g X_{ON} . d	
• $k_{Ac,Xs0}$	2.25	d^{-1}	
• $k_{Ac,Ss0}$	1.4	d^{-1}	
• $k_{Ac,Xon0}$	1.1	d^{-1}	
Inhibition constant			
• K_i	0.33	g S_A/L	
Yields			
• Y_{SF}	2.7	g X_{Ac}/ g S_F	Angelidaki *et al.*, 1993
• Y_{SNO}	18	g X_{Ac}/ g S_{NO}	Angelidaki *et al.*, 1993
• Y_{SA}	0.915	g X_{Ac}/ g S_A	Angelidaki *et al.*, 1993
• Y_{SNH}	0.09	g X_{Ac}/ g S_{NH}	Angelidaki *et al.*, 1993
Fraction of inert matter			
• f_{Xs}	0.9	-	
• f_{Ss}	0.3	-	
• f_{Xon}	0.74	-	

4.2.2 Description of the digestion process

The waste input in the system consists of a mix of human faecal material and inedible parts of plants produced in the High Plants Compartment. It contains insoluble organic matter (X_S) like polymers (lignin, cellulose, pectin, proteins, lipids, starch…), insoluble organic nitrogen (X_{ON}) like proteins, fermentable soluble matter (S_F) (sugars…) and soluble carbon (S_S).
X_S, S_S and X_{ON} are hydrolysed into smaller molecules (respectively S_F and S_{NO}) by the bacterial enzymes. Then, fermentable soluble molecules S_F are consumed with nitrate S_{NO} by the acidogenic bacteria and converted into volatile fatty acids (S_A), NH_4^+(S_{NH}) and CO_2. The process is stopped at this stage of the fermentation by implementing low pH 5.8 – 6 to avoid methanogenesis.
The CO_2 produced is then supplied to the Higher plants chamber and to the algae compartment. The liquid effluent of compartment I goes through a membrane filtration unit which separates particulates. The solid components such as biomass and insoluble organic matter which were not biodegraded are recycled to the first compartment. A fraction can first join a Fibre Degradation Compartment (FDC) in order to improve the efficiency of the system. Moreover, when it is necessary a part of the solids can be completely removed. The soluble phase containing VFA, NH_4^+ and nitrogenic organic matter joins the compartment II. Acetogenesis and methanogenesis do not occur in this model, as they are wanted to be inhibited in the compartment I of MELiSSA. The model takes into account a continuous mode (Figure 3).
The balances of each component are established and allow to write the matrix.

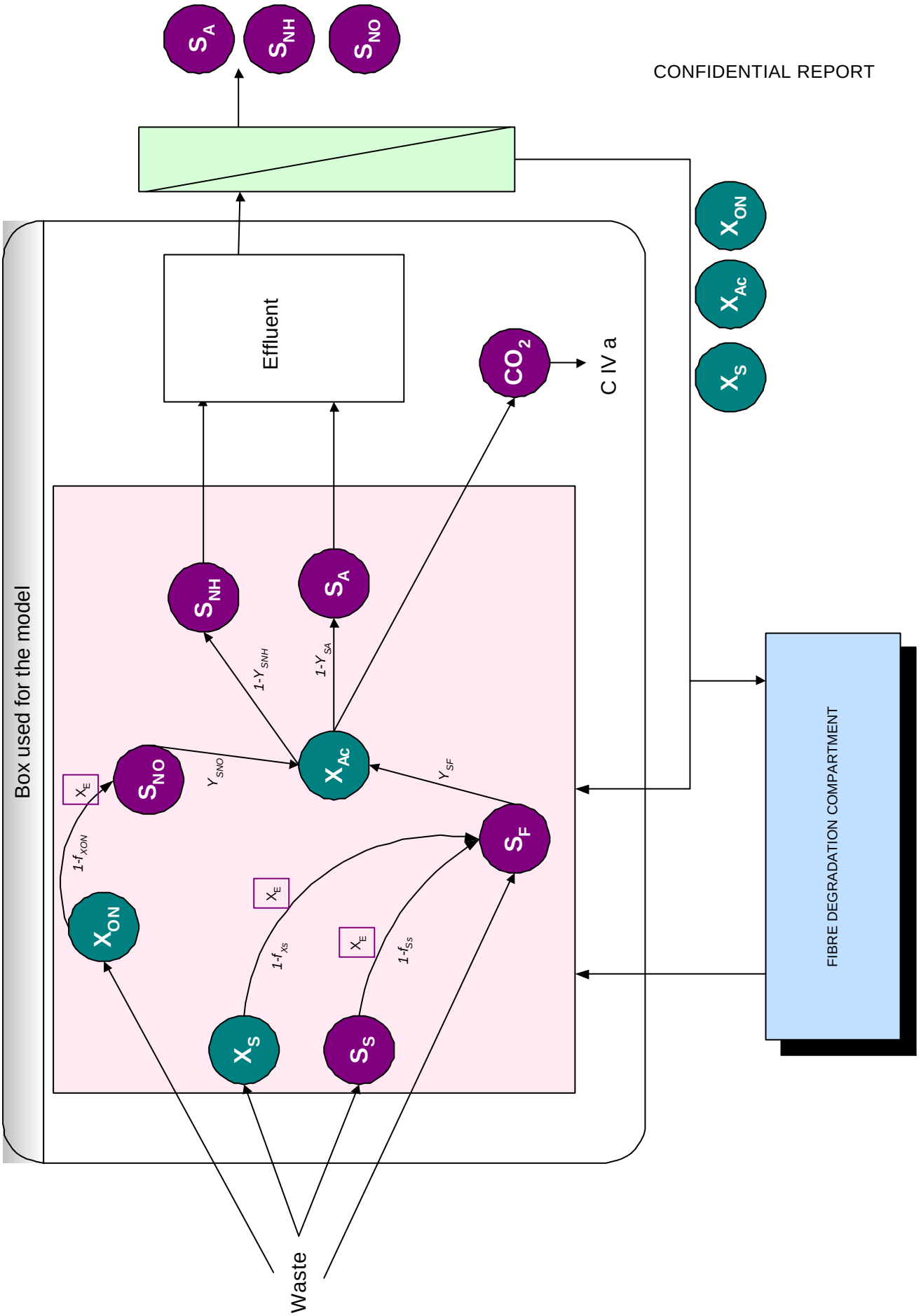

Figure 3. The digestion process in the first compartment using the IAWQ nomenclature

4.2.3 Balances

4.2.3.1 Enzymatic hydrolysis

The enzymatic hydrolysis can be summarized by the following equations:

$$X_E + aX_S + bS_S \rightarrow cS_F \quad \text{eq(1)}$$

$$X_E + dX_{ON} \rightarrow eS_{NO} \quad \text{eq(2)}$$

where : X_E = Enzymes concentration
X_S = Insoluble organic carbon concentration
S_S = Soluble carbon concentration

4.2.3.1.1 Substrates

Insoluble organic carbon

X_S is hydrolysed by X_E into S_F (fermentable chemical oxygen demand):

$$\frac{d(VX_S)}{dt} = Vr_{X_S} + X_{S0}q - q_{FDC}h_{FDC}X_{S0} - q_{drain}X_{S0} \quad \text{eq(3)}$$

$$\frac{dX_S}{dt} = -(1-f_{Xs})k_{Ac,XS}X_S + X_{S0}D - \frac{q_{FDC}}{V}h_{FDC}X_{S0} - \frac{q_{drain}}{V}X_{S0} \quad \text{eq(4)}$$

where: V = Volume of the reactor
r_{Xs} = X_S reaction rate
X_{S0} = Initial X_S concentration
q = Volumetric flow rate entering in the reactor I
f_{Xs} = Fraction of inert matter in X_S
$k_{Ac,Xs}$ = Kinetic constant

D = Dilution rate = $\frac{q}{V}$

q_{FDC} = Volumetric flow rate going to the FDC
q_{drain} = Volumetric flow rate of drain
h_{FDC} = Efficiency of the FDC

The insoluble organic carbon is recycled in compartment I after being separated by the membrane unit. The only output of X_S occurring corresponds to its degradation in the FDC and its simple removal in the drained fraction.
The reaction rate corresponds to the hydrolysis of Xs by hydrolytic enzymes for synthesising S_F. A proportion of X_S is hydrolysed (1-f_{Xs}), the rest (f_{Xs}) stays in the reactor.
The volatile fatty acids inhibit the enzymatic hydrolysis:

$$k_{Ac,XS} = k_{Ac,XS0}\frac{K_i}{S_A + K_i} \quad \text{eq(5)}$$

where: $k_{Ac,Xs}$ = kinetic constant
$k_{Ac,XS0}$ = non inhibited conditions kinetic constant
K_i = inhibition constant

(Angelidaki *et al.*, 1993)
The validity of the previous equations can be checked by writting the dimension equations ; the units of each member of the equation are developed to see if they are corroborating :

$$\left[k_{Ac,XS}\right] = \left[\frac{gS_F}{gX_S d}\right] = \left[\frac{gS_F}{gX_S d}\frac{gS_A}{L}\frac{L}{gS_A}\right] = \left[\frac{gS_F}{gX_S d}\right] \quad \text{eq(6)}$$

$$\left[\frac{dX_S}{dt}\right]=\left[\frac{gX_S}{Ld}\right]=\left[\frac{gS_F}{gX_S d}\frac{gX_S}{L}+\frac{gX_{S0}}{Ld}\right]=\left[\frac{gS_F}{Ld}+\frac{gX_{S0}}{Ld}\right] \quad \text{eq(7)}$$

Soluble organic carbon

S_S is hydrolysed by X_E into S_F:

$$\frac{d(S_S)}{dt}=Vr_{Ss}+(S_{S0}-S_S)q-q_{drain}S_S \quad \text{eq(8)}$$

where: r_{Ss} = S_S reaction rate
S_{S0} = Initial Ss concentration

$$\frac{dS_S}{dt}=-(1-f_{Ss})k_{Ac,SS}S_S+D(S_{S0}-S_S)-\frac{q_{drain}}{V}S_S \quad \text{eq(9)}$$

where: f_{Ss} = Fraction of inert matter in S_S
$k_{Ac,Ss}$ = kinetic constant

$$k_{Ac,SS}=k_{Ac,SS0}\frac{K_i}{S_A+K_i} \quad \text{eq(10)}$$

where: $k_{Ac,Ss0}$ = non inhibited conditions kinetic constant

Dimension equation:

$$\left[\frac{dS_S}{dt}\right]=\left[\frac{gS_S}{Ld}\right]=\left[\frac{gS_F}{gS_S d}\frac{gS_S}{L}+\frac{gX_{Ac}}{Ld}+\frac{gS_S}{Ld}\right]=\left[\frac{gS_F}{Ld}+\frac{gX_{Ac}}{Ld}+\frac{gS_S}{Ld}\right] \quad \text{eq(11)}$$

S_S can leave the system in the flow joining the second compartment and in the drain.

Insoluble organic nitrogen

X_{ON} is hydrolysed by X_E into S_{NO}:

$$\frac{d(VX_{ON})}{dt}=Vr_{X_{ON}}+X_{ON0}q-q_{FDC}h_{FDC}X_{ON}-q_{drain}X_{ON} \quad \text{eq(12)}$$

where: r_{XON} = X_{ON} reaction rate
X_{ON0} = Initial X_{ON} concentration

$$\frac{dX_{ON}}{dt}=-(1-f_{XON})k_{Ac,XON}X_{ON}+DX_{ON0}-\frac{q_{FDC}}{V}h_{FDC}X_{ON}-\frac{q_{drain}}{V}X_{ON} \quad \text{eq(13)}$$

where: $k_{Ac,\ XON}$ = Kinetic constant

$$k_{Ac,XON}=k_{Ac,XON0}\frac{K_i}{S_A+K_i} \quad \text{eq(14)}$$

where: $k_{Ac,\ XON}$ = Non inhibited conditions kinetic constant
As solids, X_{ON} have a behaviour similar to X_S.
Dimension equation:

$$\left[\frac{dX_{ON}}{dt}\right]\left[\frac{gX_{ON}}{Ld}\right]=\left[\frac{gS_{NO}}{gX_{ON}d}\frac{gX_{ON}}{L}+\frac{gX_{ON}}{Ld}\right]=\left[\frac{gS_{NO}}{Ld}+\frac{gX_{ON}}{Ld}\right] \quad \text{eq(15)}$$

Products

Fermentable COD

$$\frac{d(VS_F)}{dt}=Vr_{S_F}+(S_{F0}-S_F)q+\frac{3}{5}q_{FDC}h_{FDC}X_S-q_{drain}S_F \quad \text{eq(16)}$$

where: r_{SF} = S_F reaction rate
S_{F0} = Initial S_F concentration

$$\frac{dS_F}{dt} = \underbrace{(1-f_{Xs})k_{Ac,XS}X_S}_{1^{st}\text{ term}} + \underbrace{(1-f_{Ss})k_{Ac,SS}S_S}_{2^{nd}\text{ term}} \underbrace{-\frac{\mu_m}{Y_{SF}}\frac{S_F}{K_{SF}+S_F}X_{Ac}}_{3^{rd}\text{ term}} + D(S_{F0}-S_F) + \frac{3}{5}\frac{q_{FDC}}{V}h_{FDC}X_S - \frac{q_{drain}}{V}S_F \quad \text{eq(17)}$$

where: μ_m = Maximum specific growth rate

Y_{SF} = yield = $\frac{dX_{Ac}}{dS_F}$

K_{SF} = Affinity constant

S_{F0} = Initial concentration of S_F

The first and the second terms correspond to the enzymatic hydrolysis of X_S and S_S into S_F, ; the third one concerns the consumption of S_F by the acidogenic bacteria for their growth.

Some fermentable COD is produced in the FDC from the degradation of X_S. It can be considered that 60% of X_S degraded is converted into organic molecules, the remaining 40% disappearing in CO_2 and biomass.

Dimension equation:

$$\left[\frac{dS_F}{dt}\right] = \left[\frac{gS_F}{Ld}\right] = \left[(1-\frac{gX_E}{gX_S})\frac{gS_F}{gX_S d}\frac{gX_S}{L} + (1-\frac{gX_E}{gS_S})\frac{gS_F}{gS_S d}\frac{gS_S}{L} - \frac{1}{d}\frac{gS_F}{gX_{Ac}}\frac{gS_F}{L}\frac{L}{gS_F}\frac{gX_{Ac}}{L} + \frac{gS_F}{Ld}\right]$$

$$= \left[(1-\frac{gX_E}{gX_S})\frac{gS_F}{Ld} + (1-\frac{gX_E}{gS_S})\frac{gS_F}{Ld} + \frac{gS_F}{Ld}\right] \quad \text{eq(18)}$$

Soluble nitrogen

$$\frac{d(VS_{NO})}{dt} = Vr_{S_{ON}} + (S_{NO0}-S_{NO})q + \frac{3}{5}q_{FDC}h_{FDC}X_{ON} - q_{drain}S_{NO} \quad \text{eq(19)}$$

where: r_{SNO} = S_{NO} reaction rate

S_{NO0} = Initial S_{NO} concentration

$$\frac{dS_{NO}}{dt} = \underbrace{(1-f_{XON})k_{Ac,XON}X_{ON}}_{1^{st}\text{ term}} \underbrace{-\frac{\mu_m}{Y_{SNO}}\frac{S_{NO}}{K_{SNO}+S_{NO}}X_{Ac}}_{2^{nd}\text{ term}} + D(S_{NO0}-S_{NO}) \quad \text{eq(20)}$$

where : Y_{SNO} = yield = $\frac{dX_{Ac}}{dS_{NO}}$

K_{SNO} = Affinity constant

The first term corresponds to the hydrolysis of X_{ON} in S_{NO}, the second corresponds to the growth of X_{AC} using S_{NO} as a substrate. Some soluble nitrogen is produced in the FDC from the degradation of X_{ON}. It can be considered that 60% of X_{ON} degraded is converted into soluble nitrogen, the remaining 40% disappearing in CO_2 and biomass.

Dimension equation:

$$\left[\frac{dS_{NO}}{dt}\right]\left[\frac{gS_{NO}}{Ld}\right] = \left[\frac{gS_{NO}}{gX_{ON}d}\frac{gX_{ON}}{L} - \frac{1}{d}\frac{gS_{NO}}{gX_{Ac}}\frac{gS_{NO}}{L}\frac{L}{gS_{NO}}\frac{gX_{Ac}}{L} + \frac{gS_{NO}}{Ld}\right] = \left[\frac{gS_{NO}}{Ld}\right] \quad \text{eq(21)}$$

4.2.3.2 Acidogenesis

The acidogenesis can be summarized by equation 22 :

$$X_{Ac} + fS_F + gS_{NO} \rightarrow hS_A + iS_{NH} + jCO_2 \quad \text{eq(22)}$$

4.2.3.2.1 Acidogenic biomass

Balances are established using Monod model for the growth of bacteria.

$$\frac{d(VX_{Ac})}{dt} = Vr_{X_{Ac}} \quad \text{eq(23)}$$

$$\frac{dX_{Ac}}{dt} \approx 0 \quad \text{eq(24)}$$

The biomass is considered as a constant to simplify the model.

4.2.3.2.2 Substrates

Fermentable COD

See 1.1.4.1.2.

Soluble nitrogen

See 1.1.4.1.2.

4.2.3.2.3 Products

Volatile fatty acids

S_A is formed from S_F and S_{NO}:

$$\frac{d(VS_A)}{dt} = Vr_{S_A} + (S_{A0} - S_A)q - q_{drain}S_A \quad \text{eq(25)}$$

where: r_{SA} = S_A reaction rate
S_{A0} = Initial S_A concentration

$$\frac{dS_A}{dt} = (1-Y_{SA})\frac{m_m}{Y_{SF}}\frac{S_F}{K_{SF}+S_F}X_{Ac} + (1-Y_{SA})\frac{\mu_m}{Y_{SNO}}\frac{S_{NO}}{K_{SNO}+S_{NO}}X_{Ac} + D(S_{A0}-S_A) - \frac{q_{drain}}{V}S_A$$

eq(26)

where: Y_{SA} = yield = $\frac{dX_{Ac}}{dS_A}$

Dimension equation:

$$\left[\frac{dS_A}{dt}\right] = \left[\frac{gS_A}{Ld}\right] = \left[(1-\frac{gX_{Ac}}{gS_A})\frac{1}{d}\frac{gS_F}{gX_{Ac}}\frac{gS_F}{L}\frac{L}{gS_F}\frac{gX_{Ac}}{L} + (1-\frac{gX_{Ac}}{gS_A})\frac{1}{d}\frac{gS_{NO}}{gX_{Ac}}\frac{gS_{NO}}{L}\frac{L}{gS_{NO}}\frac{gX_{Ac}}{L} + \frac{gS_A}{Ld}\right] \quad \text{eq(27)}$$

$$= \left[(1-\frac{gX_{Ac}}{gS_A})(\frac{gS_F}{Ld} + \frac{gS_{NO}}{Ld}) + \frac{gS_A}{Ld}\right]$$

Ammonia (NH_3)

$$\frac{d(VS_{NH})}{dt} = Vr_{S_{NH}} + (S_{NH0} - S_{NH})q - q_{drain}S_{NH} \quad \text{eq(28)}$$

where: r_{SNH} = S_{NH} reaction rate
S_{NH0} = Initial S_{NH} concentration

$$\frac{dS_{NH}}{dt} = (1/6.25)\frac{\mu_m}{Y_{SNO}}\frac{S_{NO}}{K_{SNO}+S_{NO}}X_{Ac} + D(S_{NH0} - S_{=nh}) - \frac{q_{drain}}{V}S_{NH} \quad \text{eq(30)}$$

$$Y_{SNH} = \frac{dX_{Ac}}{dS_{NH}} \quad \text{eq(31)}$$

To simplify the model, S_F is considered as carbohydrates containing no nitrogen: the ammonium is only synthesised from S_{NO}.
S_{NH} is usually expressed in gN/L. To have correct units, the factor depending on S_{NO} can be divided by 6.25 (which is the ratio of N in proteins).

Dimension equation:

$$\left[\frac{dS_{NH}}{dt}\right] = \left[\frac{gS_{NH}}{Ld}\right] = \left[(1-\frac{gX_{Ac}}{gS_{NH}})\frac{1}{d}\frac{gS_{NO}}{gX_{Ac}}\frac{gS_{NO}}{L}\frac{L}{gS_{NO}}\frac{gX_{Ac}}{L} + \frac{gS_{NH}}{gX_{Ac}}\frac{gX_{Ac}}{Ld} + \frac{gS_{NH}}{Ld}\right]$$
$$= \left[(1-\frac{gX_{Ac}}{gS_{NH}})\frac{gS_{NO}}{Ld} + \frac{gS_{NH}}{gX_{Ac}}\frac{gX_{Ac}}{Ld} + \frac{gS_{NH}}{Ld}\right] \quad \text{eq(32)}$$

CO_2

For this first approach, the carbon dioxide gas is considered as inexistant. No gas-liquid equilibrium is taken into account.

4.2.4 Matrice

For each step of the process, the kinetics (reaction rates) can be represented depending on the different components. Table 17 gives the reaction rate of each component and the kinetics for each process considered in the model.

Table 17. Matrice

Process	Components					
	X_{Ac}	X_S	S_S	X_{ON}	S_F	S_{NO}
Hydrolysis of insoluble organic carbon		$-(1-f_{Xs})$			$(1-f_{Xs})$	
Hydrolysis of insoluble organic nitrogen				$-(1-f_{XON})$		1
Hydrolysis of soluble organic carbon			$-(1-f_{Ss})$		$(1-f_{Ss})$	
Growth of acidogenic bacteria on carbon substrate					$-\frac{1}{Y_{SF}}$	
Growth of acidogenic bacteria on nitrogen substrate						$-\frac{1}{Y_{SNO}}$

Process	Components			Kinetics
	S_A	S_{NH}	CO_2	
Hydrolysis of insoluble organic carbon				$k_{Ac,XS0}\frac{K_i}{S_A+K_i}X_S$
Hydrolysis of insoluble organic nitrogen				$k_{Ac,XO0}\frac{K_i}{S_A+K_i}X_{ON}$
Hydrolysis of soluble organic carbon				$k_{Ac,SS0}\frac{K_i}{S_A+K_i}S_S$
Growth of acidogenic bacteria on carbon substrate	$\frac{(1-Y_{SA})}{Y_{SF}}$		$\frac{(1-Y_{SF}-Y_{SA})}{Y_{SF}}$	$m_m\frac{S_F}{K_{SF}+S_F}X_{Ac}$
Growth of acidogenic bacteria on nitrogen substrate	$\frac{(1-Y_{SA})}{Y_{SNO}}$	$\frac{(1-Y_{SNH})}{Y_{SNO}}$	$\frac{1}{Y_{SNO}}-\frac{1}{Y_{SA}}-\frac{1}{Y_{SNH}}$	$m_m\frac{S_{NO}}{K_{SNO}+S_{NO}}X_{Ac}$

5. Sizing of the compartment

5.1 Introduction

Based on the previous balances and model, a program was written using Matlab. With this model, simulations were performed, using different scenarios, to have a first estimation of the volume of the reactor.

5.2 The program

To start with simple cases, the biomass is considered as a constant (X_{AC} = 1.8 g/L), and the gas-liquid equilibrium and CO_2 production are not represented. The simulation is done on a period of 90 days (3 months). A preliminary study was performed with a program simulating a lab scale reactor operating in the Melissa conditions at EPAS. The comparison with the experimental results allowed to calibrate the model. The modelled configuration of the first compartment in the pilot plant (Figure 4) takes into account the presence of a membrane filtration unit crossed by the effluent. A part of the effluent can also be sent to a Fibre Degradation Compartment in order to improve the degradation of fibres. Another part of the effluent can also be removed from the system (drain).

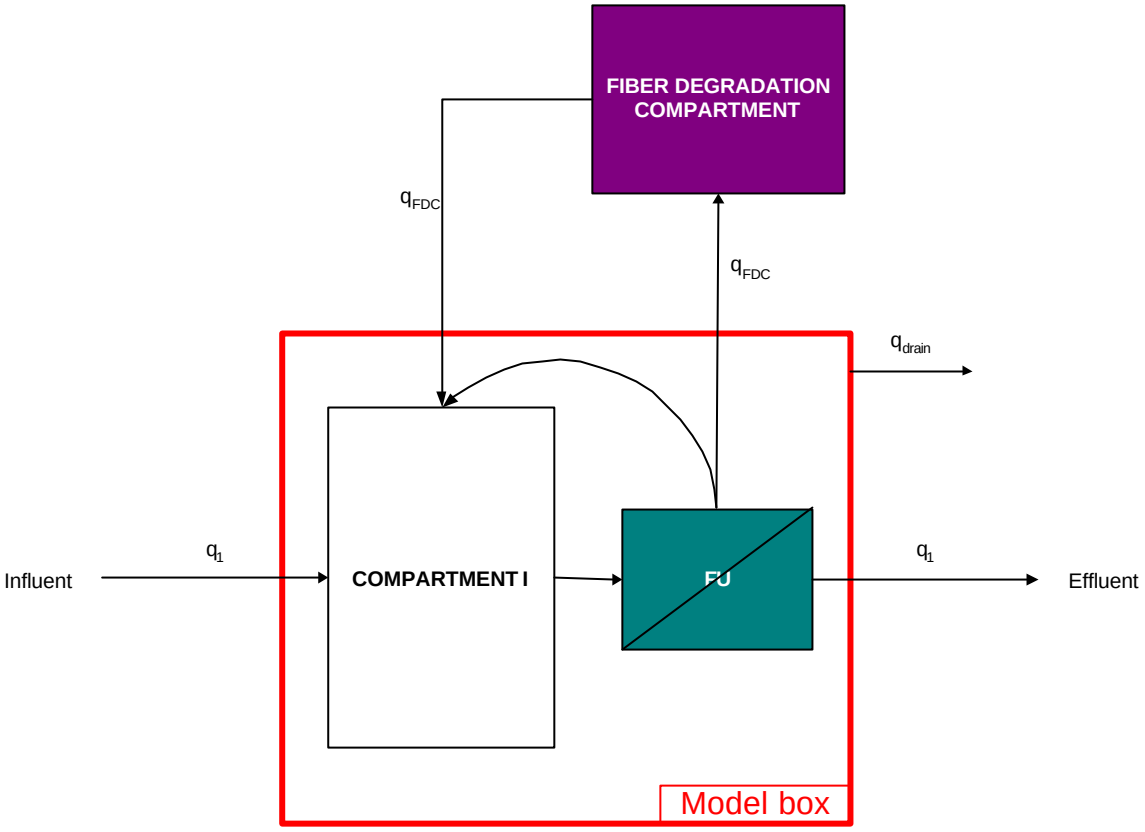

Figure 4. General configuration of the first compartment in the model

5.2.1 Running of the program

The program considers the compartment I running in continuous mode. It runs as a dynamic model: the user can enter initial values such as the massic flow of compounds, the dilution rate and the volumetric flow rate in a text file which is read by the program. Then it gives back curves showing the evolution of the concentrations of each component with the initial conditions previously fixed. The corresponding data are also available in an excel file (see Figure 5).

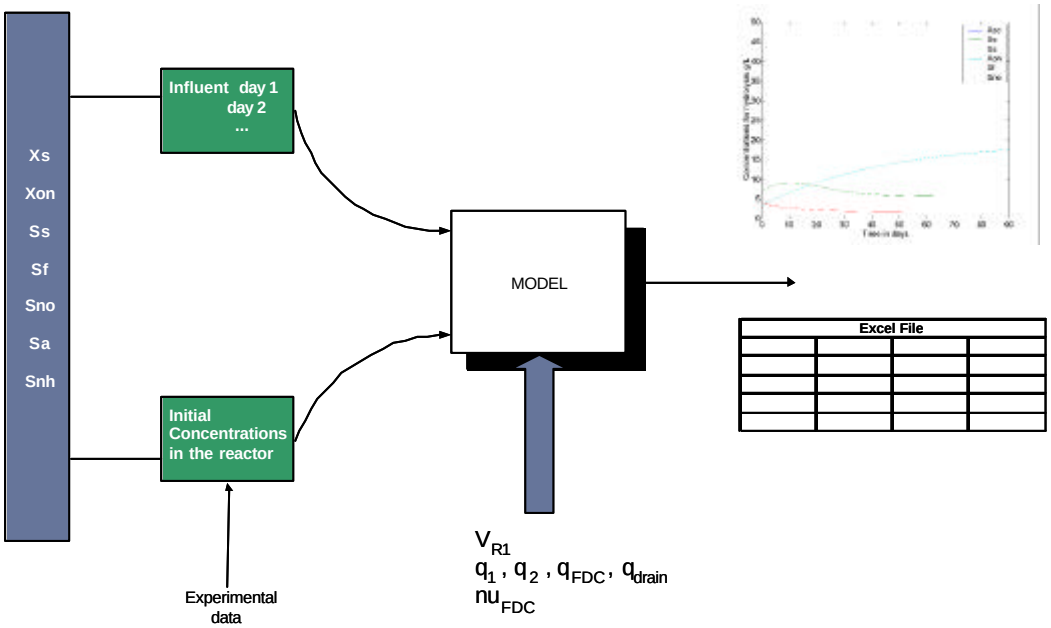

Figure 5. Model use

5.2.2 Function of definition of the system of differential equations:

```
%Equa-diff
% Modelling of compartment I
%
%-----------------------------------------------------------
%
function dxdt = equadiffFDC( t, x )
global p
%
% x = [ 1=V 2=Xac 3=Xs 4=Ss 5=Xon 6=Sf 7=Sno 8=Sa
%     9=Snh 10=CO2 ];
% p = [ 1=mu_m 2=K_Sno 3=K_Sf 4=f_Xs 5=f_Ss 6=k_AcXs_0 7=k_AcSs_0
%       8=k_AcXon_0 9=Y_Sf 10=Y_Sno 11=f_Sa 12=Y_Sa 13=Y_Snh 14=d_Ac 15=Ki
%       16=D 17=Xs_0 18=Ss_0 19=Xon_0 20=Sf_0 21=Sno_0 22=Sa_0 23=Snh_0
24=f_Xon 25=q
%       26=nu_FDC 27=q1 28=q_FDC 29=q_drain];
%
%-------------------------------------------------------------------
%
%Kinetic of enzymatic hydrolysis of Xs
HydXs  = (p(6) * p(15) / ( x(8) + p(15) ) )* x(3) ;
%Kinetic of enzymatic hydrolysis of Ss
HydSs  = (p(7) * p(15) / ( x(8) + p(15) ) )* x(4) ;
%Kinetic of enzymatic hydrolysis of Xon
HydXon = (p(8) * p(15) / ( x(8) + p(15) ) )* x(5) ;
%Kinetic of consumption of Sf by the bacteria
ConsoSf  = p(1) * x(6) * x(2) / ( p(3) + x(6) );
%Kinetic of consumption of Sno by the bacteria
ConsoSno = p(1) * x(7) * x(2) / ( p(2) + x(7) );
%
dxdt = zeros( 10, 1 );
%Volume (constant)
dxdt(1) = 0;
%Xac = biomass
dxdt(2) = 0;

%Xs
dxdt(3) = - ( 1 - p(4) ) * HydXs ...
   + p(27) * p(17) / x(1) - p(28) / x(1) * p(26) * x(3) - p(29) / x(1) *
x(3) ;

%Ss
dxdt(4) = - ( 1 - p(5) ) * HydSs + p(27) * ( p(18)- x(4) ) / x(1) - p(29) /
x(1) * x(4) ;

%Xon
dxdt(5) = - ( 1 - p(24) ) * HydXon + p(27) * p(19) / x(1) - p(28) / x(1) *
p(26) * x(5) - p(29) / x(1) * x(5) ;

%Sf
dxdt(6) = ( 1 - p(4) ) * HydXs + ( 1 - p(5) ) * HydSs -  ConsoSf / p(9) ...
      +  p(27) * ( p(20)- x(6) ) / x(1) + p(28) * 3 / 5 * p(26) * x(3) /
x(1) - p(29) / x(1) * x(6);

%Sno
dxdt(7) = ( 1 - p(24) ) * HydXon - ConsoSno / p(10) ...
         + p(27) / x(1) * ( p(21) - x(7) ) + 3 / 5 * p(28) / x(1) * p(26) *
x(5) - p(29) / x(1) * x(7) ;

%Sa
```

```
dxdt(8) = ( 1 - p(12) ) * ConsoSf / p(9) +  ( 1 - p(12) ) ...
        * ConsoSno / p(10) + ( p(22) - x(8) ) * p(16)  - p(28) / x(1) *
x(8) - p(29) / x(1) * x(8) ;
%Snh
dxdt(9) = (1/6.25)*( 1 - p(12) ) * ConsoSno / p(10)  ...
         + ( p(23) - x(9) ) * p(16) -p(28) / x(1) * x(9) - p(29) / x(1) *
x(9)  ;
%CO2
dxdt(10) = 0;
%
%---------------------------------------------------------------
%end
```

5.2.3 Principal program solving the system

```
%
%Compartment I in continuous mode
%
% - constant volume
% - well mixed
% - Monod kinetic
% - continuous configuration
% - inhibition by VFA
% - Units : days,g,L
%-------------------------------------------------------------
% parameters
mu_m = 3;
K_Sno = 0.2;
K_Sf = 0.2;
f_Xs = 0.9;
f_Ss = 0.3;
k_AcXs_0 = 2.25;
k_AcSs_0 = 1.4;
k_AcXon_0 =1.1;
Y_Sf = 2.7;
Y_Sno = 18;
f_Sa = 0.7;
Y_Sa = 0.915;
Y_Snh = 0.09;
d_Ac = 0.0625;
Ki = 0.33;
s = load('influent.txt');
D = s(8);
q = s(9);
%Concentrations in the influent
Xs_0 = s(1)/q;
Ss_0 = s(2)/q;
Xon_0 = s(3)/q;
Sf_0 = s(4)/q;
Sno_0 = s(5)/q;
Sa_0 = s(6)/q;
Snh_0 = s(7)/q;
f_Xon = 0.74;
nu_FDC = s(12);
q1=q;
q_FDC = s(10);
drain = s(11);
%
%
global p
p = [ mu_m K_Sno K_Sf f_Xs f_Ss k_AcXs_0 k_AcSs_0 k_AcXon_0 Y_Sf Y_Sno f_Sa
```

```
Y_Sa Y_Snh d_Ac Ki D Xs_0 Ss_0 Xon_0 Sf_0 Sno_0 Sa_0 Snh_0 f_Xon q nu_FDC
q1 q_FDC drain];
%-----------------------------------------------------------
%
% Initial states in the reactor
V = q/D;
Xac = 1.8;
Xs = 12;
Ss = 3.8;
Xon = 3.3;
Sf = 0.36;
Sno = 0.678;
Sa = 4.5;
Snh = 0.87;
CO2 = 0;
 %
xo = [ V Xac Xs Ss Xon Sf Sno Sa Snh CO2 ];
%-----------------------------------------------------------
%integrate the state equation
to = clock;
days = 90;

[ t, x ] = ode45( 'equadiffFDC', [ 0 days ], xo );

secs = etime( clock, to )
%
%-----------------------------------------------------------
%
%plot the results
%Fig 1 : Hydrolysis
font = 15;
figure;
plot( t, x(:,2) , t, x(:,3), t, x(:,4) , t, x(:,5), t, x(:,6), t, x(:,7));
h = legend('Xac','Xs', 'Ss', 'Xon', 'Sf', 'Sno');
axis( [ 0 days 0 50 ] );
set ( gca, 'fontsize', font );
xlabel( 'Time in days' );
ylabel( 'Concentrations for hydrolysis g/L' );
%
%Fig 2 : Acidogenesis
font = 15;
figure;
plot( t, x(:,2) , t, x(:,6), t, x(:,7) , t, x(:,8), t, x(:,9), t, x(:,10));
h = legend('Xac','Sf', 'Sno', 'Sa', 'Snh', 'CO2');
axis( [ 0 days 0 20] );
set ( gca, 'fontsize', font );
xlabel ( 'Time in days' );
ylabel ( 'Concentrations for acidogenesis(g/L)');
%

% Save the results in an excel file

% Check the information to save
if exist('x')==0 , return, end;

%Open a window to save the results in an excel file
[Resultats ]=uiputfile('*.xls','save as');

   if ischar(Resultats); %Check the existence of the file
      if exist('x_profile')==1; % Check the existence of the matrice to
save
```

```
          sauve=[t x_profile]; % Prepare the results: time (column),
results (1 per column)
          save([Resultats],'sauve','-ASCII','-DOUBLE','-TABS'); % Save in a
text format, data separated by tabulation
      else;
         sauve=[t x];% Save in a text format, data separated by tabulation
         save([ Resultats ],'sauve','-ASCII','-DOUBLE','-TABS')
      end;

  clear sauve;
  end;

clear Resultats;

%-------------------------------------------------------------

% the end!
```

5.3 Calculations of the figures needed by the matlab program:

5.3.1 Hydrolytic constants

They are determined from using the model in the lab configuration and according to bibliographic data.

5.3.2 Composition of the feed

The amounts of each component that should be fed to the reactor every day can be estimated for the substrate previously defined (containing the faecal material (FM) of one man per day and 162 gDW/d of inedible parts of plants (NEPM)). They are calculated from experimental measurements: total and soluble nitrogen (Nt_{total} and $Nt_{soluble}$), total and soluble ammonium (NH_{4total} and $NH_{4soluble}$), VFA, dry matter (DM), ashes (Ash), short time biochemical oxygen demand (BOD_{short}), total and soluble chemical oxygen demand (COD_{total} and $COD_{soluble}$) (see Figure 6 and Table 18).

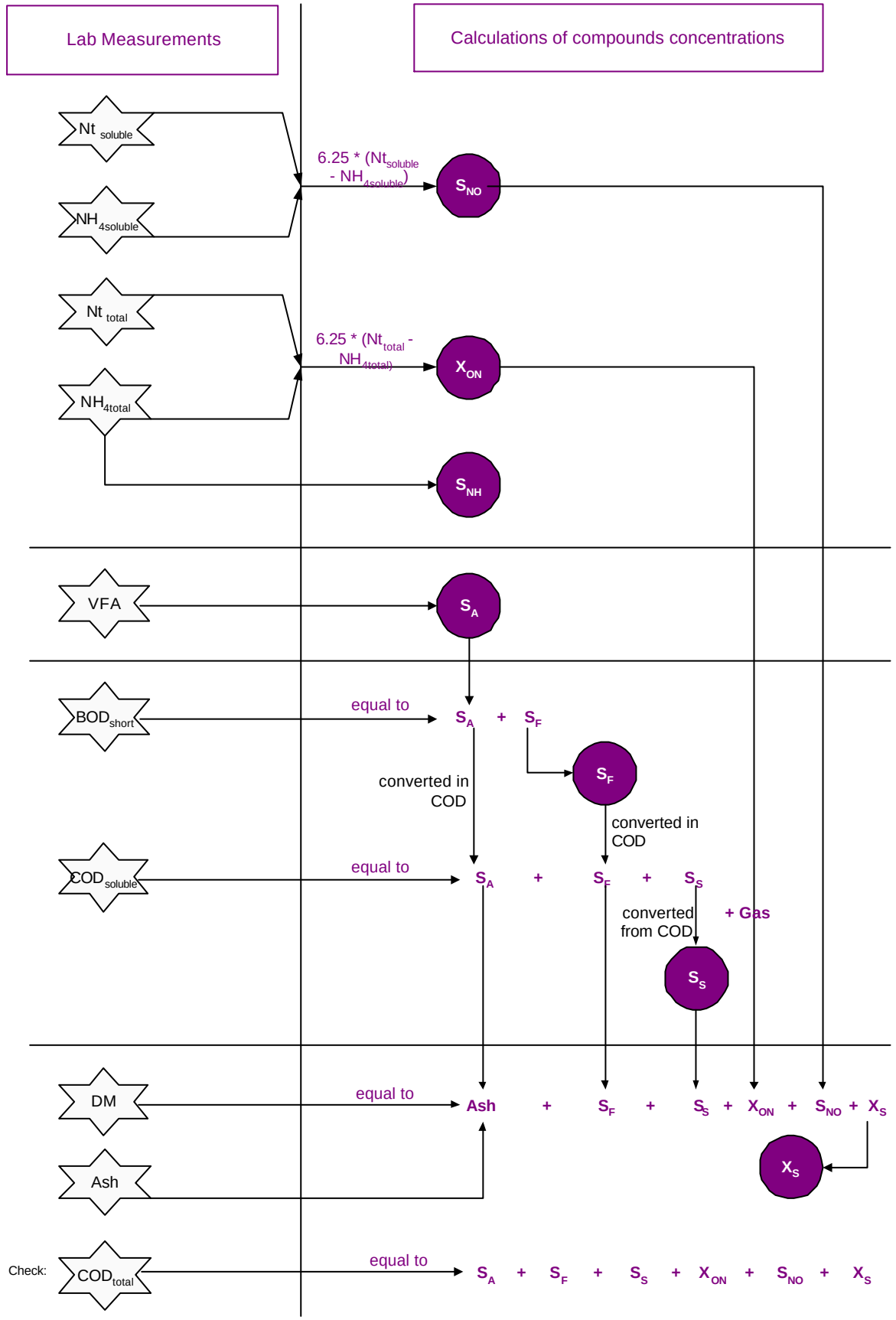

Figure 6. Calculation of concentrations from experimental measurements

Table 18. Feed composition

	Lab Reactor	**Pilot Reactor**
Total dry weight load (gDW / d)	2.56	192
	Feed Concentration (g / L)	**Load in the pilot reactor (g / d)**
X_S	13.2	63.4
S_S	11	53
X_{ON}	7.2	34.5
S_F	0	0
S_{NO}	2.1	9.9
S_A	0.53	2.6
S_{NH}	0.12	0.6

5.3.3 Degradation ratio

The proportions of substrates that can be degraded must be determined to allow correct simulations. They are also deduced from experiments, by comparing the concentrations in the influent and in the effluent (see Table 19).

Table 19. Fractions of inert matter

f_{XS}	90%
f_{SS}	30%
f_{XON}	74%

5.4 Models validation

An official model aiming to predict behaviour of compartment I is developed at UBP. It is an advanced model, wanted to be improved in the future with experimental results obtained from the prototype and the pilot reactors, to be used to control the process in the Melissa pilot plant. In parallel, the simple model developed at EPAS is intended as a simple tool to dimension the pilot reactor, in order to obtain immediately some feelings about the evolution of critical parameters (Figure 7).

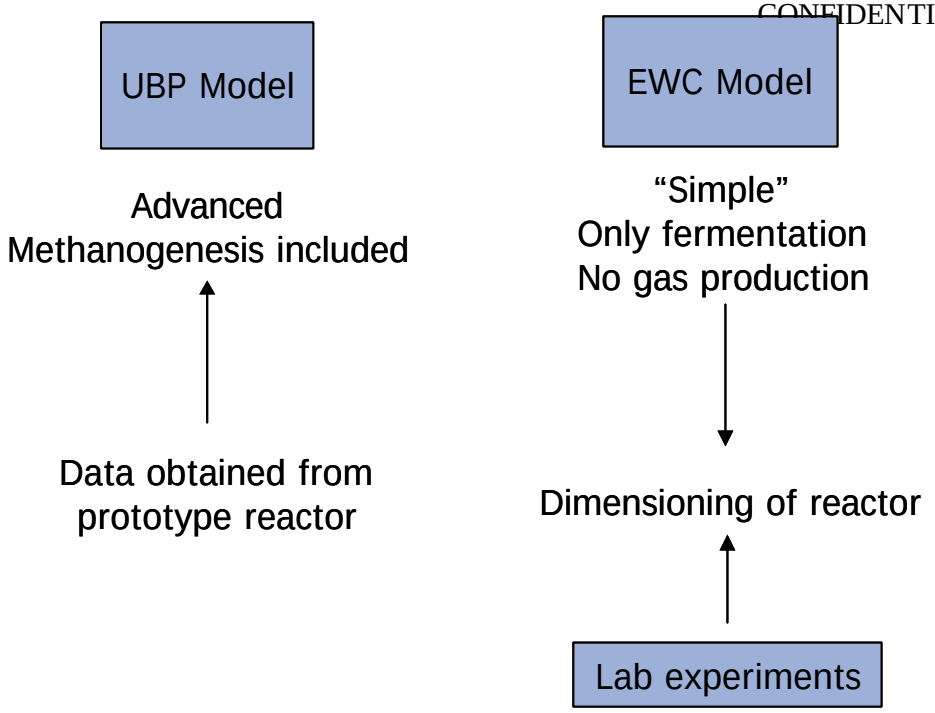

Figure 7. UBP and EWC Models

As the models will be used to work on the concept of the pilot reactor, it is essential to validate them by checking that they can give a good approach of experimental results. To join this aim, a simulation is realised with the EWC and the UBP model, with a concrete set-up from which experimental results are available. The process can be described as following:

Reactor volume = 1.5L
HRT = 23d
Influent composed of faecal material and plants:

2.3gOM/d
Bx = 0.05 gOM/gDM.d (where Bx = Mass load)
Bv = 1.53 gDM/L.d (where Bv = Volumetric load)

Initial state in the reactor: reactor already fed with faecal material and plants

The experimental concentrations obtained in steady state were compared with the ones obtained using the two models. Table 20 shows the comparison of results and the deviation for each component.

Table 20. Comparison of experimental and model-predicted results

	EWC_Model	Deviation with experimental results (%)	UBP_Model	Deviation with experimental results (%)	Experimental results
Xs (g/l)	11,78	0,1	11,8	0,3	11,77
Ss	2,92	2,7	3,16	5,3	3
Xon	5,5	3,8	5,2	1,9	5,3
Sf	0,1	23,1	0,18	38,5	0,13
Sno	0,86	4,4	0,57	36,7	0,9
Sa	2,37	2,9	2,94	20,5	2,44
Snh	0,21	48,8	0,54	31,7	0,41
sum	23,74	0,9	24,39	1,8	23,95
DM					31.9

As shown in Table 20, the deviations for both models are globally acceptable. Higher deviations are obtained mainly for S_F and S_{NH}. It has to be noticed that the dry matter concentration measured on the effluent is higher than the sum of each components describing the matter in the model (+ 8 g/L), which means that this list is not representative of the totally of the matter. The difference corresponds to the biomass, salts and other compounds which are not taken into account in the model.

A second check can be performed on the evolution of the total dry matter in the reactor. It is described by the dry matter content measured on the effluent for the experimental results, and on the sum of each component concentration for the model results. A difference around 8 g/L can be expected, related to the difference mentioned above. The compared evolutions (only with EWC model) are shown in Figure 8).

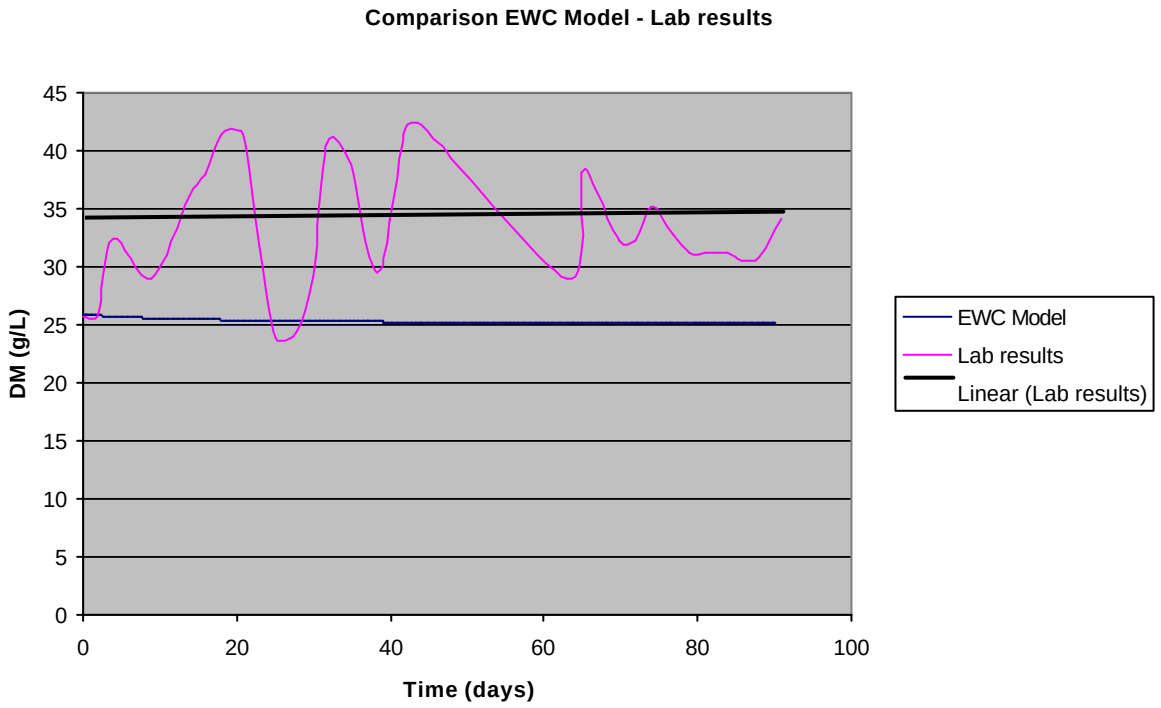

Figure 8. Comparison of DM evolution with EWC Model and lab results

A difference around 8 g/L is indeed obtained between the two evolutions. The profile is similar. From these results the EWC Model can be validated. The UBP Model will be further calibrated in the future with experimental results obtained with the prototype reactor by UBP.

5.5 Scenarios: Results and discussion

Some simulations were made with a substrate containing only faecal material; these preliminary results allowed to check the coherence of the model, by comparing the results with the one obtained by practical experiments on the first compartment.

Two factors of variation were used:

- the volume of water used to dilute the substrate to feed the reactor : 2.5 to 20L/d
- the dilution rate D, and as a consequence the volume of the reactor : 1 to 1000L.

Two main criteria have importance:

- the running of the first compartment must be optimised, meaning to maximise the biodegradation efficiency: the highest production of VFA is wanted.
- the dry matter concentration must be lower than 5% to insure a good working of the membrane filtration unit.

It appeared that the volume of the reactor has not a big influence on the efficiency of the VFA production; as a consequence, the simulations were focused on volumes of 25, 50 and 100 L only, which seem to be practically usable to build a prototype reactor.
For each dilution volume (2.5, 10, 20L), a simulation was done for each volume of the reactor (25, 50, 100L).

5.5.1 S_A

Table 21. Evolution of VFA (g/L) depending on the volumetric flow rate and the reactor volume

S_A	Reactor volume (L)		
Volumetric flow rate (L/d)	25	50	100
2.5	3	4.1	4.7
10	0.72	1.2	1.3
20	0.35	0.5	0.6

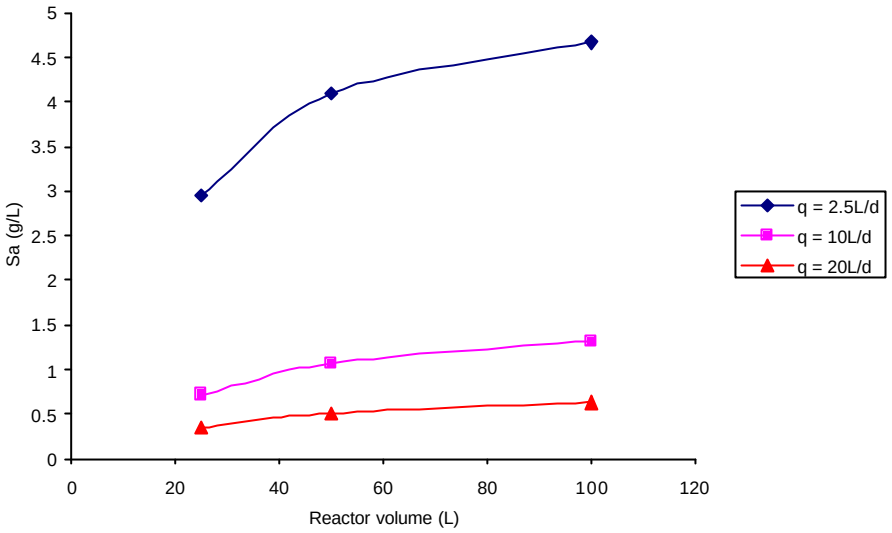

Figure 9. Evolution of S_A (g/L) depending on the volumetric flow rate and the reactor volume

S_A is the most important component to take into account. Indeed, the best production of this component is wanted in the first compartment. First, it can be noticed that the volume of the reactor seems to have no significant influence on the production of volatile fatty acids. This result is a little amazing : it was expected that a lower volume of the reactor would not allow a sufficient residence time to obtain the optimal amount of S_A. As a consequence, the volume of the prototype reactor will be chosen from other criteria, e.g. the amount of organic matter (for the using of the membrane filtration unit).
Furthermore, the simulation shows that the concentration of S_A decreases when the volumetric flow rate is increased : the more the precursors of the volatile fatty acids are diluted, the less these volatile fatty acids produced are concentrated.

5.5.2 Dry matter

Considering the evolution of the sum of the components allows to characterize the load of dry matter in the reactor, which is an important parameter to take into account for using the membrane filtration unit (Table 22).

Table 22. Evolution of Dry Matter (g/L) depending on the volumetric flow rate and the reactor volume

DM	Reactor volume (L)		
Volumetric flow rate (L/d)	25	50	100
2.5	185	111	64.7
10	64.4	43.5	27.4
20	40.7	26.3	16.4

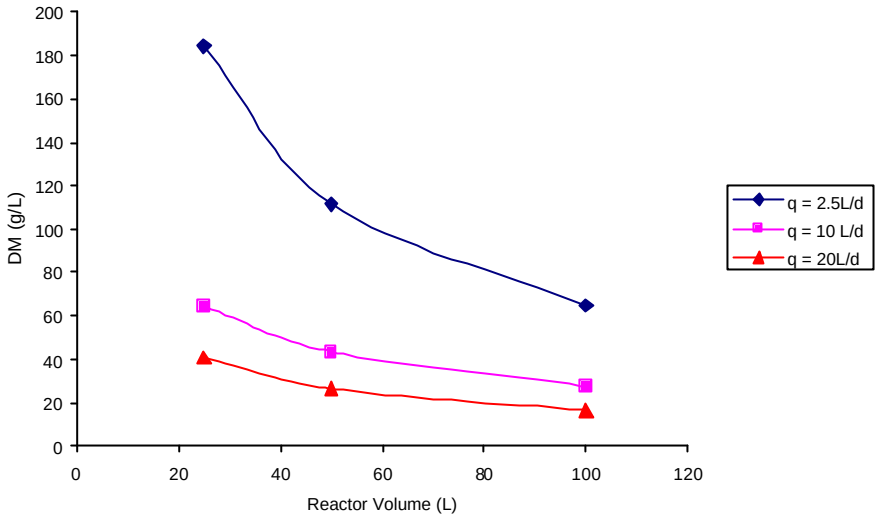

Figure 10. Evolution of Dry Matter (g/L) depending on the volumetric flow rate and the reactor volume

The dry matter content increases when the flow and the volume decrease. Indeed, in that case the matter is more concentrated.

5.5.3 S_F, S_{NO}

Table 23. Evolution of S_F (g/L) depending on the volumetric flow rate and the reactor volume

S_F	Reactor volume (L)		
Volumetric flow rate (L/d)	25	50	100
2.5	12.4	0.76	0.16
10	3.88	0.92	0.19
20	1.90	0.65	0.18

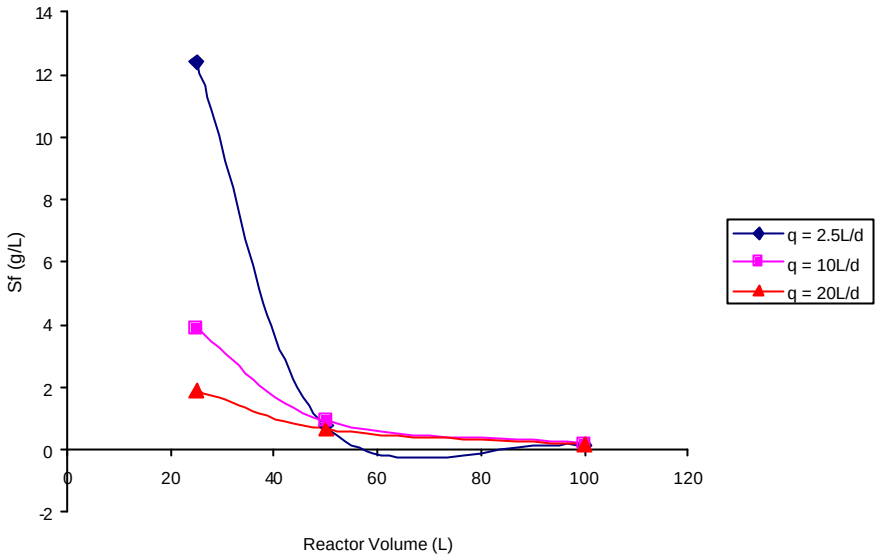

Figure 11. Evolution of S_F (g/L) depending on the volumetric flow rate and the reactor volume

Table 24. Evolution of S_{NO} (g/L) depending on the volumetric flow rate and the reactor volume

S_{NO}	**Reactor volume (L)**		
Volumetric flow rate (L/d)	25	50	100
2.5	13.40	8.79	2.93
10	3.73	3.03	1.78
20	1.88	1.56	0.98

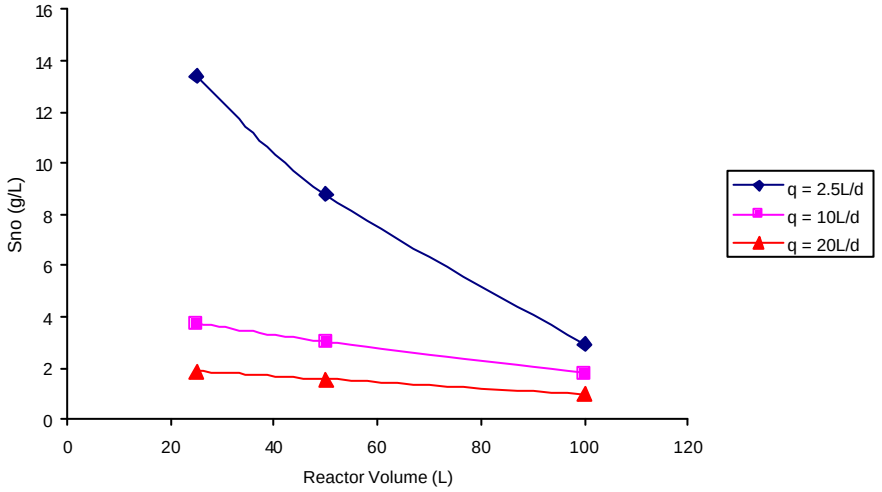

Figure 12. Evolution of S_{NO} (g/L) depending on the volumetric flow rate and the reactor volume

S_F and S_{NO} have the same evolution: their concentrations decrease when the volumetric flow rate increases, but stay at low value, due to the fact that they are intermediate products almost totally consumed immediately after their formation. The decrease is more marked with lower reactor volume: the impact of a little change is more obvious in little volumes.

5.5.4 Xs, Xon, S_S

Table 25. Evolution of X_S (g/L) depending on the volumetric flow rate and the reactor volume

X_S	Reactor volume (L)		
Volumetric flow rate (L/d)	25	50	100
2.5	98.60	59.80	33.75
10	35.85	23.90	14.31
20	23.20	14.30	8.28

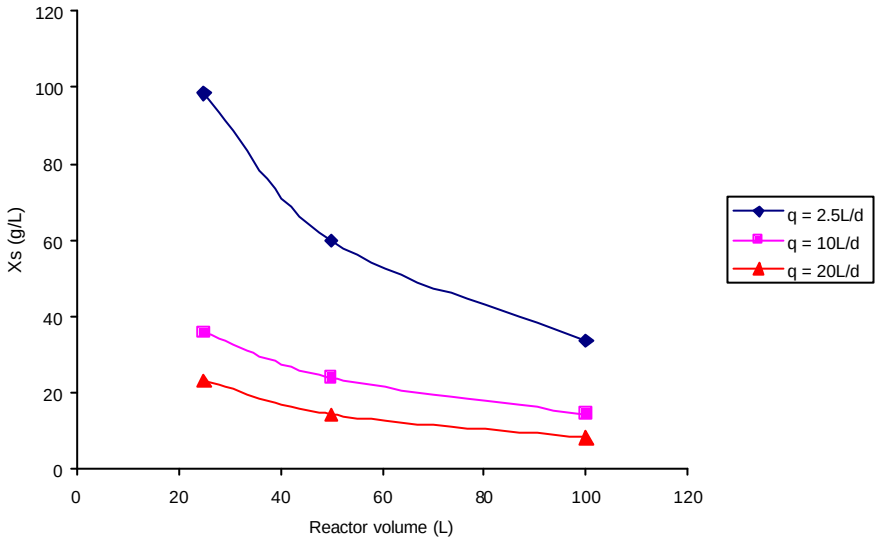

Figure 13. Evolution of X_S (g/L) depending on the volumetric flow rate and the reactor volume

Table 26. Evolution of X_{ON} (g/L) depending on the volumetric flow rate and the reactor volume

X_{ON}	Reactor volume (L)		
Volumetric flow rate (L/d)	25	50	100
2.5	44.40	27.28	15.14
10	15.37	10.20	6.08
20	9.93	6.14	3.53

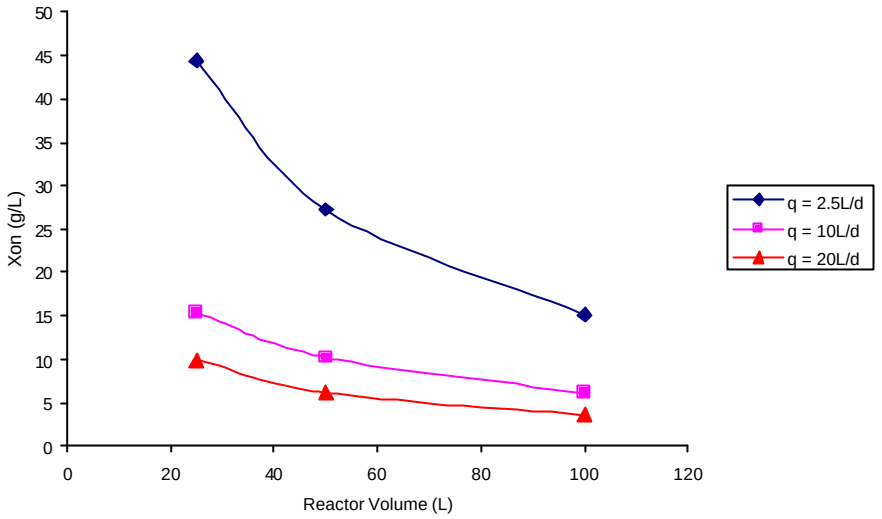

Figure 14. Evolution of X_{ON} (g/L) depending on the volumetric flow rate and the reactor volume

Table 27 Evolution of S_S (g/L) depending on the volumetric flow rate and the reactor volume

S_S	**Reactor volume (L)**		
Volumetric flow rate (L/d)	25	50	100
2.5	10.70	8.54	5.84
10	2.99	2.46	1.79
20	1.66	1.35	0.99

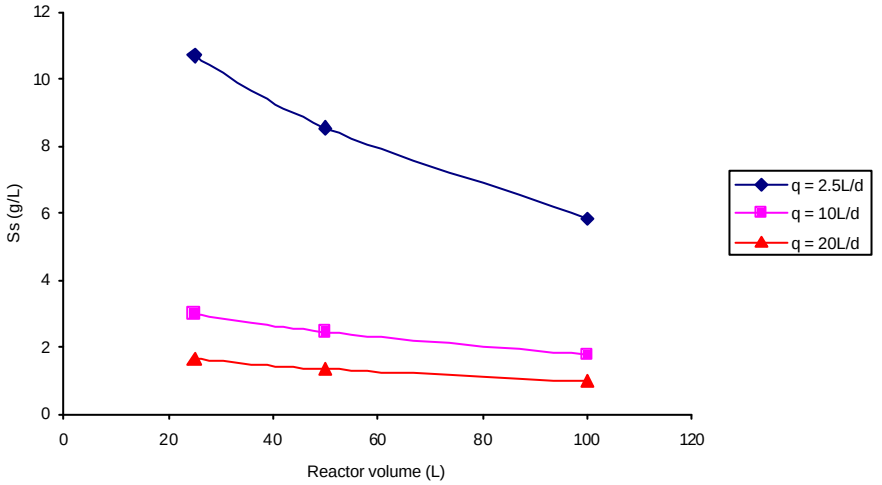

Figure 15. Evolution of S_S (g/L) depending on the volumetric flow rate and the reactor volume

These two insoluble components and the soluble organic matter vary in the same way. Their concentrations decrease when the reactor volume and the volumetric flow rate increase. They come from the influent, an increase of the reactor volume or the flow dilutes these substrates.

5.5.5 S_{NH}

Table 28. Evolution of S_{NH} (g/L) depending on the volumetric flow rate and the reactor volume

S_{NH}	Reactor volume (L)		
Volumetric flow rate (L/d)	25	50	100
2.5	0.27	0.32	0.43
10	0.07	0.08	0.09
20	0.03	0.04	0.05

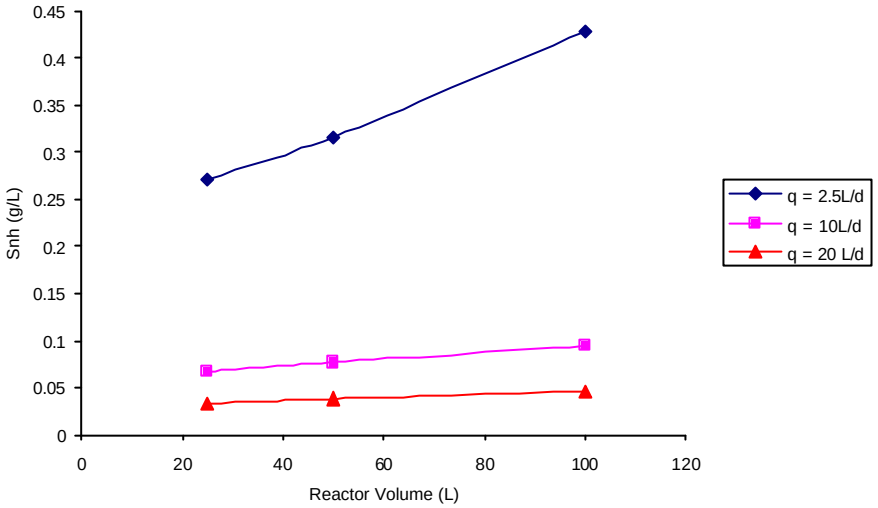

Figure 16. Evolution of S_{NH} (g/L) depending on the volumetric flow rate and the reactor volume

Its concentration decreases when the flow rate increases until a stable value, with an evolution similar to S_A. For big volumetric flow rates (more than 20L/d), the concentrations seem to reach a stable value.

5.5.6 First proposition and scenarios

According to the previous results, the choice of the reactor volume will not be determined by the production of volatile fatty acids because of their relative independence. Other criteria should be taken into account, e.g. the load of organic matter contained in the reactor, which must be lower than 5% to allow an efficient use of the membrane filtration unit. The volume has thus to be sufficient high. Nevertheless, a bigger volume should be safer for the MELISSA cycle because it allows softening external variations, but for studying the behaviour of the compartment, a smaller volume could be more interesting to evaluate the impact of variations which are more obvious in this case. According to the previous results, a reactor volume of 100L fed with a flow of 10L/d is proposed. Then some scenarios are carried out in order to evaluate the impact of the variation of:

- dry matter content at equilibrium in the reactor
- relation with the Fibre Degradation Compartment
- load of feed.

Indeed, the proposed concept will have to be able to assume variations and extreme conditions. Scenarios are carried out with three FDC efficiencies (0.2, 0.5 and 0.8), and with the nominal load of feed and its double. The flow that has to be sent to the FDC is determined in order to stabilise the matter concentration around 25g/L and 50 g/L in the reactor (see Table 29). Depending on the capacity of the FDC, a drain could be used as an additional output of solids to stabilise the reactor. By these means the proposed concept can soften the variations.

Table 29. By-pass flow sent to the Fibre Degradation Compartment

	FDC Efficiency		
Melissa massic load	0.2	0.5	0.8
Nominal load	q_{FDC} = 1 L/d	q_{FDC} = 0.6 L/d	q_{FDC} = 0.45 L/d
Nominal load · 2	q_{FDC} = 5 L/d q_{drain} = 4.5 L/d	q_{FDC} = 4.25 L/d q_{drain} = 4 L/d	q_{FDC} = 4 L/d q_{drain} = 3.5 L/d

a) DM stabilised at 25 g/L

	FDC Efficiency		
Melissa massic load	0.2	0.5	0.8
Nominal load	$q_{FDC} = q_{drain}$ = 0 L/d		
Nominal load · 2	q_{FDC} = 5 L/d	q_{FDC} = 3 L/d	q_{FDC} = 2 L/d

b) DM stabilised at 50 g/L

Where Melissa nominal load x 1 = 192 gDW/d

Melissa nominal load x 2 = 384 gDW/d

6. Compatibility with the MELiSSA pilot plant

The anaerobic waste compartment needs to be equipped with interfaces in order to connect this compartment to the entire MELiSSA loop to obtain a closed loop. The VFA and ammonium produced in the first compartment need to be transported to the second compartment, using the liquid loop. The CO_2 produced need to be transported to the 4th compartment, by means of the gas loop. Therefore, it is necessary to know the connection ports of the other two compartments (comp.II and comp. IV).
Ammonia and VFA will be measured in both the first and the second compartment. The ammonia will moreover be measured in the third compartment. The VFA analyser and the Ammonium analyser can therefore be shared.

7. General Conclusions

The liquefying compartment is responsible for the biodegradation of human faecal material, non-edible parts of higher plants and other wastes generated by the crew. The waste is composed of faecal material of one man, toilet paper and non-edible parts of higher plants (wheat, beet and lettuce), when the Higher Plants Compartment is calculated to provide 20% of a one-man diet. In Table 30 the ratio of the waste substrate is shown.

Table 30. Composition of the substrate

Substrate	Amount (gDW/d)
Lettuce	54
Beet	54
Wheat	54
Toilet paper	18
Faecal material	30
Total amount of substrate	210

The waste compartment and the substrate, that needs to be biodegraded, need to come up to several requirements. In Table 31 the general requirements of the waste compartment and the waste substrate is represented. The waste compartment consists of several units, like filtration unit, liquid loop, gas loop and solid loop.

Table 31. Requirements of bioreactor, filtration unit and substrate preparation

Requirements
Bioreactor
• Temperature: 55°C
• Condition: anaerobic
• Biodegradation efficiency of at least 55%
• PH: 6
Filtration unit
• Retentate: bacteria and non biodegraded organic matter
• Permeate: VFA and ammonium
Substrate preparation
• Lettuce, beet: wet
• Wheat straw: dry
• Pre-treatment: final diameter of material 0.5 mm

Mass balance on the waste compartment.

The values given to each parameter in the general mass balance scheme, as shown in Figure 17, are based on the model studied above. For each compartment or loop which is connected to the waste compartment, a general concept is given mainly the input flow rate and the volume of the reactor.

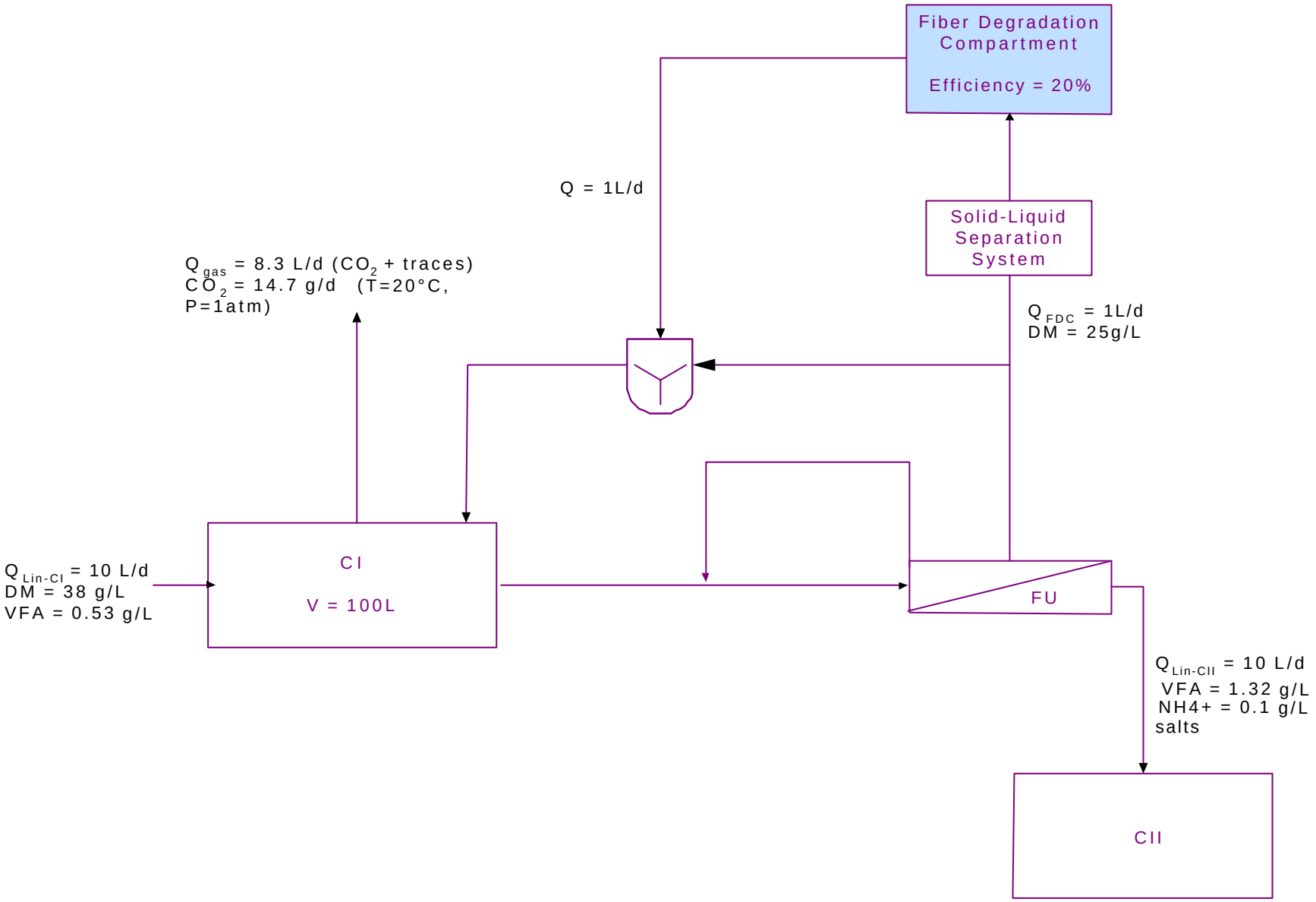

Figure 17. Mass balance of the Waste Compartment

To optimise the filtration unit, which can a maximum of 5% (W/V) solids filtering, and to ensure its long term running, 1L per day of the solids fed to the system should be withdraw from the reactor CI and processed in the FDC (according to the model), the aim being to stabilise the concentration of matter in CI at 25 g/L (2.5% W/V). The treated waste is recycled to CI using a dilution tank. This configuration assumes a FDC efficiency of 20%, which seems realistic.

The gas flow produced by CI is estimated from previous laboratory experiments at EPAS.n.v. location.

Table 32. Nomenclature

CI	Compartment I
Q_{Lin-CI}	Volumetric liquid flow rate influent to CI
$Q_{Lout-CI}$	Volumetric liquid flow rate effluent from CI
Q_{gas}	Volumetric gas flow rate from CI
FU	Filtration Unit
Q_{Lin-FU}	Volumetric liquid flow rate influent to FU
$Q_{Lout-FU}$	Volumetric liquid flow rate effluent from FU
CII	Compartment II
$Q_{Lin-CII}$	Volumetric liquid flow rate influent to CII
FDC	Fibre Degradation Compartment
$Q_{Sin-FDC}$	Volumetric solid flow rate influent to FDC
Q_R	Volumetric recycling flow rate
TSS	Total Suspended Solids
VFA	Volatile Fatty Acids

8. References

MELiSSA Technical Notes:

TN 25.31: PEREZ J. and MONTESINOS J.L., 1996 "Nitrifying compartment studies, Starting of the nitrifying reactor", contract 11549/95/NL/FG.

TN 32.3: POUGHON L., June 1997 " Including of a higher plants chamber in the MELISSA loop, description of a HPC for MELISSA loop steady state simulations", PO 161081.

TN 40.2 : CLOUTIER G., DIXON M.A., June 1998 "Composition of Edible and Non-Edible Fractions of Crops in the MELiSSA Menu", PO 171684.

TN 43.2: HERMANS V., DEMEY D., May 1999 "MELISSA demonstration reactors", contract 12922/98/NL/MV.

TN 45.2: POUGHON L., April 2000 "Behaviour of the nitrifying compartment fed with organic substrates", PO 161031

TN 45.3: POUGHON L., 2000 "Modelling the liquefying compartment", Memorandum of understanding ECT/FG/MMM/97.012.

TN 45.4: FAVIER-TEODORESCU L., PONS A., POUGHON L., June 1999 "stoichiometric analysis of *R.rubrum* growth on different carbon substrates", Contract 13323/98/NL/MV

TN 47.3: GODIA F., ALBIOL J., CREUS N., PEREZ J., December 2000 "Preliminary review of the pilot plant final loop".

TN 51.2: HERMANS V., DEMEY D., March 2000 "Biodegradation of the non edible parts of higher plants", contract 12922/98/NL/MV.

Other references:

PR01.057mm002: Minutes of meeting, 16/01/2002.

ANGELIDAKI I., ELLEGAARD L., AHRING B.K., 1993, "A mathematical model for dynamic simulation of anaerobic digestion of complex substrates: Focusing on ammonia inhibition"

DEMEY D, 1998 "MELiSSA Demonstration Breadboard, BIORAT" contract BRT-TN-001-EP.

HAMMES F., DEMEY D., HERMANS V., VERSTRAETE W., 1998 "Status and future needs : biological techniques (Introductory Lecture)"

KOBAYASHI M., KURATA S., 1978 "Mass culture and cell utilization of photosynthetic bacteria", Process Biochem., 13, pp.27-30.

LASSEUR C., BINOT R.A., 1991 "Control systems for artificial ecosystems. Application to MELiSSA", 21st International Conference on Environmental Systems", San Francisco (USA).

MALINA J.F., POHLAND JR.F.G., 1992 "Design of anaerobic processes for the treatment of industrial and municipal wastes", Water Quality Management Library, Technomic publishing CO., p23-24.

PREUD'HOMME F., DEMEY D., CORNET J.F., WOOD P., WOLFF H., AERTS J.M., NIJS K., 1998 "MELiSSA demonstration breadboard: Sizing of the consumer compartment", BIORAT Technical note 1.

OLSSON G., NEWELL B., 1999 "Wastewater treatment systems: Modelling, diagnosis and control", IWA Publishing, 1999.

VRATI S, 1984 "Single cell protein production by photosynthetic bacteria grown on the clarified effluent of biogas plant", Appl. Microbiol. Biotechnol., 19, p199.

Web-sites:

-http://perso.libertysurf.fr/pst/svtiufm/respire.htm -Les briques de l'univers, 1997, LE CAM A., PALIN SAINTE AGATHE R.,

http://www.ac-versailles.fr/etabliss/herblay/briques/fr/fr_o.htm. -Canadian Journal of Soil Science, february 2001, "Root-to-straw ratios – influence of moisture and rate of N fertilizer", C.A. Campbell and R. de Jong :

http://www.nrc.ca/aic-journals/2001ab/cjss01/feb01/cjss00-027.html