COMPARISON OF TWO MATHEMATICAL MODELS FOR CORRELATING THE ORGANIC MATTER REMOVAL EFFICIENCY WITH HYDRAULIC RETENTION TIME IN A HYBRID ANAEROBIC BAFFLED REACTOR TREATING MOLASSES

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ABSTRACT

A modelling of the anaerobic digestion process of molasses was conducted in a 70-litre multistage anaerobic biofilm reactor or hybrid anaerobic baffled reactor with six compartments at an operating temperature of 26 ºC. Five hydraulic retention times (6, 16, 24, 72 and 120 h) were studied at a constant influent COD concentration of 10000 mg/L. Two different kinetic models (one was based on a dispersion model with first-order kinetics for substrate consumption and the other based on a modification of the Young equation) were evaluated and compared to predict the organic matter removal efficiency or fractional conversion. The first-order kinetic constant obtained with the dispersion model was 0.28 h^{-1}, the Peclet dispersion number being 45, with a mean relative error of 2%. The model based on the Young equation predicted the behaviour of the reactor more accurately showing deviations lower than 10% between the theoretical and experimental values of the fractional conversion, the mean relative error being 0.9% in this case.

Keywords: Anaerobic digestion - Modelling - Hydraulic retention time - Organic matter removal efficiency - Hybrid anaerobic baffled reactor.

INTRODUCTION

Anaerobic digestion of wastewaters has been considered to have a number of advantages over the conventional aerobic process. It saves the energy needed for aeration, converts organic matter into methane gas, a readily useable fuel, needs low nutrient requirement and produces low biomass. Anaerobic processes have gained popularity over the past decade, and have already been applied successfully for the treatment of many high and medium strength industrial wastewaters [1-7].

Taking into consideration the slow growth rate of many anaerobic microorganisms, particularly methanogenics, the main objectives of the efficient reactor design must be high retention time of cells with very little loss of microorganisms from the bioreactor. The technological challenge to improve the anaerobic digestion lies in enhancing the bacterial activity together with good mixing to ensure a high rate of contact between the cells and their substrate [1, 3, 7].

The anaerobic baffled reactor (ABR) consists of a cascade of baffled compartments where the wastewater flows upward through a bed of anaerobic sludge after being transported to the bottom of the compartment. The ABR does not require the sludge to granulate in order to perform effectively, although granulation can occur over time [8, 9]. Experiments with lab-scale reactors have shown that the ABR is very stable under shock loads due to its compartmentalised structure [9-11]. In addition, the ABR has many potential advantages, i.e. no requirement of biomass with unusual settling properties and low capital and operating costs coupled with mechanical simplicity [9].

In the present study, a hybrid anaerobic baffled (HABR) reactor or multistage biofilm reactor with six compartments was used. This reactor configuration can be

considered as a combination of the anaerobic baffled reactor (ABR) and upflow anaerobic fixed bed (UAFB) system. The upflow anaerobic filter basically is a contact process in which wastes pass over or through a mass of biological solids contained within the reactor by a fixed media [12]. The biomass in the reactor is attached to the media surfaces as a thin biofilm, is entrapped within the media matrix, or is held as a granulated or flocculated sludge mass beneath the media. Soluble organic compounds in the influent wastewater pass in close proximity to this biomass and diffuse into the surfaces of the attached or granulated solids where they are converted to intermediates and end products, specifically, methane and carbon dioxide [12].

Therefore, the main properties of the HABR are: lower sludge yields, and the ability to partially separate between the various phases of anaerobic catabolism [9-11,13]. The latter causes a shift in bacterial population allowing increased protection against toxic materials and higher resistance to changes in environmental parameters such as pH and temperature. The greatest advantage of this reactor configuration is probably its ability to separate acidogenesis and methanogenesis longitudinally down the reactor, allowing the reactor to behave as a two-phase system without the associated control problems and high costs.

Kinetic studies are very helpful for reproducing the operational behaviour of the anaerobic process and understanding the metabolic routes of biodegradation, while simultaneously saving time and money [14]. However, the development of an up-to-date model of organic matter anaerobic degradation is complex with considerable difficulties due to the high number of variables affecting the anaerobic system [15, 16]. For instance, it is difficult to describe the whole anaerobic process by reliable kinetics since hydrolysis of complex insoluble substrate depends on many different parameters such as particle size, production of enzymes, pH and temperature [17].

A model was developed for the anaerobic digestion of a glucose-based medium in an innovative high-rate reactor known as the periodic anaerobic baffled reactor (PABR). In this model, each compartment is considered as two variable volume interacting sections, with constant total volume, one compartment with high solids and the other one with low solid concentrations, with the gas and liquid flows influencing the material flows between the two sections. For the simulation of glucose degradation, the biomass was divided into acidogenic, acetogenic and methanogenic groups of microorganims. The model succeeded in predicting the reactor performance as the organic loading rate was gradually increased [18]. Another kinetic model for predicting the behaviour of the PABR was developed based on batch experiments using glucose as substrate [8]. The PABR may be operated as an upflow anaerobic sludge blanket (UASB) reactor, an ABR or at an intermediate mode. The key assumption of this model was that the hydraulic behaviour of a PABR was equivalent to the behaviour of CSTRs in series as regards the dissolved matter. The model adequately predicted the experimental behaviour of this glucose-fed PABR and was also used to examine the performance of this reactor as a function of the operating conditions, both for constant and varying loading rates. It was shown that the reactor would best be operated as a UASB or an ABR [8].

Another kinetic model was recently developed for explaining the performance of a four-compartment ABR, incorporating granular sludge biomass and operating at different hydraulic retention times (HRT) in the range of 3 to 24 hours using dilute aircraft de-icing fluid with total chemical oxygen demand (COD) concentrations in the range of 300-750 mg/L. However, the first-order empirical model initially developed for describing the reactor performance did not adequately predict the total COD removal efficiency in the reactor providing inconsistent results for the kinetic coefficient values

125 and no predictive correlation of these coefficients with substrate concentration,
126 hydraulic retention time and organic loading rate could be achieved [19, 20].

127 A mathematical model of the baffled reactor performance was developed and
128 applied using a concept of completely mixed reactors operating in series to describe the
129 performance of a modified laboratory-scale (150 L) ABR using molasses wastewater as
130 substrate [21]. This reactor had three chambers and a final settler. The first two
131 compartments each had a 10 cm layer of plastic media (Pall rings with a specific surface
132 area of 142 m^2/m^3) near the liquid surface. The third chamber had the upper half filled
133 with a modular corrugated block. This kinetic analysis focussed on the granular sludge
134 bed, with total mass of granular sludge as the main parameter. The model results were
135 in good agreement with the experimental data [21].

136 The Young model has been recently used to obtain the kinetic parameters of the
137 anaerobic digestion of synthetic domestic sewage in an upflow filter with corrugated
138 plastic rings as packing media at psychrophilic temperature (15-17 ºC). The flow pattern
139 observed in this reactor was intermediate between plug-flow and CSTR system,
140 although the plug-flow was predominant in this case [7].

141 However, despite the advantages offered by the hybrid anaerobic baffled reactors,
142 few mathematical analyses have been reported to date for modelling the kinetic
143 behaviour of these reactors and none of them for simulating the variation of the total
144 COD removal efficiency under several HRTs. Therefore, the main objective of this
145 work was to compare two different kinetic models in the anaerobic treatment of
146 molasses as a source of carbon: a model based on the concept of an axial diffusion or
147 dispersion model with first-order kinetics for substrate consumption and a model based
148 on a modification of the Young equation. These mathematical models have not to date
149 been reported in the literature as describing the kinetic performance of this specific type

of hybrid reactor operating under varying HRTs. The anaerobic hybrid reactor used for this purpose was composed of six sequential compartments, where each one formed a packed bed using Raschig rings as a medium for supporting the biofilm formation.

MATERIALS AND METHODS

Laboratory-scale experimental set-up

The hybrid anaerobic baffled reactor was composed of six discrete compartments with a total working volume of 70 L. The six compartments were made from "Plexiglas" with identical geometric characteristics, a total volume of 12 L and a gas accumulation space of 0.75 L for each one. The baffles inside the reactor were used to direct the flow of wastewater in an upflow mode through a series of compartments where each one formed a packed bed using Raschig Rings as a media to support the biofilm formation. The main characteristics of this Raschig Ring packing were: material, metal; nominal size, 13 mm; height, 25 mm; wall thickness, 0.8 mm; surface area, 420 m^2/m^3; and 85% porosity. The porosity of the beds was 81% and the fixed beds were placed up to a height of 40 cm from the bottom of the reactor. The beds maintained 73% porosity after cell immobilization. A schematic diagram of the experimental set-up used, including some reactor details, are shown in Figure 1.

The reactor was thermostated with a water jacket which kept the operational temperature at 26 ± 0.5ºC, and effluent wastewater from the sixth compartment was discharged. The six compartments operated only in an anaerobic regime. Sampling taps provided on the wall of each compartment allowed extraction of samples for analysis in various chambers of the biofilm reactor.

175 *Molasses used*

176 The reactor was fed with molasses as a carbon source. The characteristics of the
177 molasses used are summarized in Table 1. During the start-up period, ammonium
178 phosphate and urea were used as sources of phosphorus and nitrogen, respectively.
179 Micronutrients and trace metals, with the characteristics and composition shown in
180 Table 2, were also added during the start-up period. During the start-up period, the
181 COD:N:P ratio was 100:5:1. When a steady-state condition was achieved, the COD:N:P
182 ratio was changed to 350:5:1. In order to neutralize any volatile fatty acids (VFA)
183 accumulation and prevent acid zone forming in the reactor, sodium bicarbonate was
184 used as an alkalinity supplement. Given the appropriate pH of the influent used as feed
185 (7.4) the volume of the sodium bicarbonate solution added was very small in all cases.
186 This solution was only added during the start-up period.

187

188 *Inoculum and experimental procedure*

189

190 The microorganisms used as inoculum in the reactor came from the sludge of a lab-scale
191 ANAMMOX (Anaerobic Ammonium Oxidation)-AFBR (Anaerobic Fluidized Bed
192 Reactor) system. The reactor was initially seeded with 27 L of anaerobic sludge. The
193 basic characteristics of the inoculum used were: 1857 mg/L of total nitrogen, 967 mg/L
194 of ammonia nitrogen, a total acidity of 367 mg acetic acid/L; 96 g/L of total solid
195 content, 40 g/L of volatile solid content, 1.898 g $CaCO_3$/L of bicarbonate alkalinity and
196 a pH of 6.8.

197 At the beginning of the experiments, for effective biofilm formation on the support
198 media, the reactor was initially started by increasing the organic loading rate from 0.5 to
199 2.5 g COD/L per day in a fed-batch mode. Molasses were used as substrate during this
200 step. After a batch feeding period of two months, this same influent was used as a

second feeding step in continuous mode at a constant organic loading rate of 4 g COD/L·d for another period of two months.

The reactor was operated until a steady-state performance was reached. The bioreactor was subjected to increasing HRTs and the performance of the system was evaluated. Five HRTs (0.25, 0.67, 1, 3 and 5 days, equivalent to 6, 16, 24, 72 and 120 hours, respectively) were studied at a constant influent COD concentration of 10000 mg/L.

Analytical Methods

The COD concentration was measured by using a semi-micro method [22]. This method was very effective for COD determinations in samples with high salinity, organic matter content and nitrogenous compounds. Total VFA (TVFA) concentrations in the samples were analyzed using a titrimetric method [23]. Ammonia-nitrogen was detected by the 4500-NH_3D method, with a NH500/2 WTW ion selective electrode and WTW pH 320m. Electrodes were calibrated according to the manufacturer's procedures. BOD was measured according to standard methods [23]. Daily liquid samples were withdrawn and centrifuged at 13000 rpm for 8 min until a clear supernatant was obtained.

The steady-state values of operational parameters were taken as the average of seven successive measurements for those parameters when the deviations between the values were less than 3% in all cases.

Software used

SigmaPlot software (version 11.0) was used to elaborate all the graphs and Figures of this study and to perform the statistical analyses. Mathcad software (version 14) was used to solve the mathematical equations corresponding to the two models assessed.

RESULTS AND DISCUSSION

Operational Performance of the HABR

Figure 2 shows the variation of the pH within the different compartments of the reactor for all the HRTs studied. As can be seen, in all cases the process takes place within the most appropriate pH values for an adequate and stable anaerobic digestion [24]. An increase in pH values was observed from the first to the sixth compartments for all HRTs assessed. In addition, for a same step or compartment, the pH increased at higher HRTs. In compartments 1 and 2, and mainly at HRTs of 6 h and 16 h, acidogenesis prevailed over methanogenesis. The same behaviour was observed in compartment 3 at an HRT of 6 h. At HRTs of 72 h and 120 h and after compartment 3, pH values were always higher than 7.2.

In relation to the variations profile of TVFA concentration (Figure 3), it was observed that for all HRTs studied, the TVFA values dropped from the first to the sixth compartments. At an early stage in the process where the growth rate of acetogens is higher and the methanogens have not yet grown enough, the TVFA values are high. However, with the passing of time (once steady-state conditions were achieved) and an increase in the growth of methanogens, the TVFA values dropped with increasing HRT, and the TVFA values decreased in all chambers. At HRTs of 24 h, 72 h and 120 h and after compartment 4, the TVFA concentrations were very low, which demonstrated the almost total transformation of the organic matter into methane. At HRTs of 6 h and 16 h and for the two first compartments, the TVFA concentration was higher than 2000 mg/L, which coincides with the lower pH values observed in these two first steps (Figure 2). This shows that by reducing the contact time between the molasses and the

biomass, there was not enough time to transform the TVFAs to end products and the outflow COD was basically constituted by volatile fatty acids.

TVFA concentrations in effluents of a multistage anaerobic migrating blanket reactor (AMBR) increased from 25 to 182 mg/L as the HRT decreased from 10.3 days to 1 day when treating synthetic wastewater containing glucose as a carbon source [25]. This AMBR reactor consisted of a rectangular tank with an active volume of 13.5 L, which was divided into three compartments, which were mixed equally every 15 minutes at 60 rpm to ensure gentle mixing. Comparing these data with those obtained in the present work, it can be seen that similar TVFA values (210 mg/L) were obtained in the effluents of the HABR reactor at an HRT of 24 h.

Mathematical modelling

The fractional conversion or organic matter removal efficiency (per one) can be defined as the ratio between the amount of COD eliminated and the COD fed [26]. Figure 4 shows the evolution of the fractional conversion (X) with the HRT (h). Because the reactor has 6 compartments (equivalent to 6 stages), each stage involves a partial HRT and, thus, a partial conversion. Therefore, the total number of experimental points (partial HRTs) plotted in Figure 4 is equal to 30 (6 stages x 5 partial HRTs/stage). As can be seen, for HRTs in the interval of 0 h-20 h the conversion increased drastically with increasing HRTs. For HRTs higher than 20 h the increase in the conversion with HRT is slower with a tendency towards an asymptotic value, without reaching total conversion at a HRT as high as 120 h. This fact demonstrated the occurrence of a small fraction of the substrate that is non-biodegradable anaerobically.

In order to predict the fractional conversion or organic matter removal efficiency (per one) for HABR, two different models were evaluated and compared: an axial

diffusion or dispersion model and an empirical model based on a modification of the Young model [12].

Axial diffusion or dispersion model with first-order kinetics for substrate consumption

The formulation of a mathematical model for a complex system such as that used in the present work, in which both the overall kinetics and the flow pattern influence the process, makes necessary to assume certain significant hypotheses that allow to harmonize the model precision with its possible usefulness.

To study the kinetics of the biological reactions, a Michaelis-Menten type model has been widely proposed [27]:

$$(-r_S) = k \cdot S/(K_S + S) \qquad (1)$$

where: r_S is the substrate consumption rate, S is the substrate concentration, k is the kinetic constant and K_S is the saturation constant.

For low substrate concentrations, $K_S >> S$, and equation (1) is reduced to a first order equation: $(-r_S) = k_1 \cdot S$, where k_1 is a first-order kinetic constant.

Equation (1) has been previously proposed for anaerobic digestion of complex substrates or wastewaters with high suspended solid content, which require a hydrolysis step previous to its acidification [27-30].

In order to describe the biological reaction within each compartment of the reactor it is necessary to know the kinetic equation as well as the flow pattern. As a first approximation and given that mixing is not deliberately promoted in each step, it could be considered that the flow could behave as an ideal a “piston flow” or plug-flow. However, it should be taken into consideration that biogas generation causes an airlift effect within each compartment resulting in a certain mixing. As a consequence it is

reasonable to assume that the real flow must behave as an intermediate between completely stirred and plug-flow.

This physical situation can approximately be described by an axial or dispersion model [26, 31]. Assuming first-order kinetics for substrate consumption, the following equation was proposed for the calculation of the fractional conversion [26, 32]:

$$X = 1 - [4 \cdot a \cdot \exp(u \cdot L/(2 \cdot D))/[(1+a)^2 \cdot \exp(a \cdot u \cdot L/(2 \cdot D)) - (1-a)^2 \cdot \exp(-a \cdot u \cdot L/(2 \cdot D))]] \quad (2)$$

where:

$$a = [1 + 4 \cdot k_1 \cdot HRT\,(D/(u \cdot L))]^{0.5} \quad (3)$$

and X is the fractional conversion, u is the linear velocity, L is the bioreactor length, D is the diffusion constant, HRT is the hydraulic retention time, and k_1 is the first-order reaction rate constant. The term $(D/(u \cdot L)$ is the dimensionless dispersion number and its value is a function of the mix level in the reactor and its geometry. The inverse of the dimensionless dispersion number $((u \cdot L)/D)$ is known as the Peclet group or Peclet number [32].

A detailed scheme of the different steps made for the calculation of the first-order kinetic constant (k_1) and dispersion number ($D/(u \cdot L)$) is shown in the flow diagram included in Figure 5. Following this diagram for the calculation, the values obtained for k_1 and ($D/(u \cdot L)$) were 0.28 h^{-1} and 45, respectively.

The following equation was used as the analytical criterion for determining the ending of the calculation and consequently to determine the optimum values of the kinetic constant and dispersion number of the process:

$$\epsilon = [(\Sigma(X_{exp} - X_{model})/X_{exp})^2]^{1/2}/N \quad (4)$$

where X_{exp} and X_{model} are the experimental and theoretical conversion values respectively, the latter being calculated by using equations (2) and (3) using the

Mathcad software (version 14), ϵ is the mean relative error of the fractional conversion, its value being 0.02, and N is the number of experimental points (N=30).

As can be seen in Figure 4, only for fractional conversions higher than 0.8, slight deviations between the experimental and theoretical conversion values were observed. These deviations could be due to the increase in the experimental effluent CODs caused by the endogenous metabolism, which provokes a reduction of the experimental conversion. The proposed dispersion model (with first-order kinetics) was validated by comparing the experimental fractional conversion data with the theoretical values obtained with this model for all HRTs studied. Figure 6 shows a comparison of the experimental and simulated data obtained with this model for all experiments carried out. As can be seen, deviations equal to or lower than 20% between the experimental and theoretical fractional conversion values were obtained. This demonstrates the suitability of the proposed model to represent the performance of the HABR and, therefore, that the global or overall kinetic parameter obtained approximately represents the activity of the different microorganism populations or microbial communities involved in the anaerobic process.

On the other hand, the value of the kinetic constant, k_1, obtained with this model in the present work (0.28 h^{-1}) is higher than the specific substrate utilization rate coefficient obtained in an ABR with three chambers (0.012 h^{-1}) processing molasses wastewater (9-38 g COD/L) at OLRs of between 5-25 kg COD/m^3 d [21]. However, this constant value is lower than the maximum specific rate of substrate consumption (0.70 h^{-1}) achieved in the methanogenesis from acetate using a periodic ABR under increasing organic loading conditions (2700 to 10500 mg/L) [18].

In addition, a dispersion model was also found to be highly suitable for describing the anaerobic digestion of municipal wastewater in a novel outside cycle reactor

developed based on the characteristics of an expanded granular sludge bed (EGSB) reactor [26]. The standard deviation of the simulated data (concentration of the effluent suspended solids) was less than 6% [33]. The flow pattern and behaviour of an acidogenic UASB reactor was also successfully simulated with the dispersion model. The axial dispersion number was identified as the most important factor in the dispersion modelling of this reactor [34]. The axial dispersion model was also found to be appropriate for studying the hydrodynamic pattern of a fluidised bed reactor [35] and a rotating disc anaerobic reactor digesting acetic acid as substrate [36]. The feasibility of the dispersion model simulating the process performance in anaerobic filters was also reported in the literature [37].

According to Levenspiel [26] and taking into account the value of the dispersion number obtained (45), the flow pattern in the present bioreactor is intermediate between the plug-flow and completely stirred reactors, although it comes nearer to the completely stirred model. Similar intermediate behaviour between plug-flow and ideally mixed was also found in an ABR with eight compartments treating dilute wastewater (500 mg COD/L) at HRTs in the range of 80-10 h [38].

Empirical modified Young model

Finally, the experimental results obtained (fractional conversion values, hydraulic retention times) were fitted to the following empirical equation, which represents a modification of the Young model [12]:

$$X = a(1 - b/HRT^{c}) \qquad (5)$$

where a, b and c are empirical constants derived from a non-linear adjustment of the above-mentioned experimental value pairs (X_{exp}, $HRT_{\exp}$) by using the least-squares method.

By solving Equation (5) with the above mentioned Mathcad software, the following values for these empirical constants were obtained: a = 1.13; b = 0.73 and c = 0.29. Figure 7 shows a comparison of the experimental fractional conversion data with the theoretical curve obtained using the modified Young model. Figure 8 shows a comparison of the experimental fractional conversion values with the theoretical values obtained with this proposed model represented by Equation (5). Because the mean relative error was only 0.9%, a completely satisfactory fit was observed. This means that the relative error was much lower than that obtained with the dispersion model using first-order kinetics (2%). In addition, the Young model has three adjustment mathematical parameters while the dispersion model has only one parameter, and the higher the number of adjustment parameters, the better is the final adjustment. Therefore, of the two models proposed, the modified Young model appears to match the performance data more closely than the dispersion model hypothesis according to the mean relative errors obtained in each adjustment. Accordingly the modified Young model will be more suitable than the dispersion model to predict the behaviour of this reactor under different operating conditions.

CONCLUSIONS

The operational behaviour of a hybrid anaerobic baffled reactor treating molasses was assessed using two different kinetic models: a dispersion model with first-order kinetics for substrate consumption and a modified Young model. These models were evaluated and compared with the aim of simulating the organic matter removal or fractional conversion under different HRTs.

399 The dispersion model reproduced the experimental results with a mean relative error
400 of 2%. The modified Young model allowed a better fit of the experimental results
401 showing a mean relative error of 0.9%.
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Table 1. Characteristics of the molasses used

pH	7.4
Chemical oxygen demand (COD)	10290 mg/L
Biochemical oxygen demand (BOD_5)	3250 mg/L
Kjeldahl nitrogen	183 mg/L
Total phosphate	10 mg/L
Fe^{2+}	10 mg/L
Ca^{2+}	592 mg/L
K^{+}	32 mg/L
Alkalinity	2070 mg/L
Total solids (TS)/Volatile solids (VS) ratio	1.5

Table 2. Composition and characteristics of the micronutrients and trace elements solution

$CoCl_2 \cdot 6H_2O$	0.25 mg/L
H_3BO_3	0.05 mg/L
$FeSO_4 \cdot 7H_2O$	0.5 mg/L
$MnCl_2 \cdot 4H_2O$	0.5 mg/L
$ZnCl_2$	0.05 mg/L
$CuCl_2$	0.15 mg/L
$Na_2MoO_4 \cdot 2H_2O$	0.01 mg/L
$NiSO_4 \cdot H_2O$	0.02 mg/L
Na_2SeO_3	0.01 mg/L
$AlCl_3 \cdot 6H_2O$	0.05 mg/L
$MgSO_4 \cdot 7H_2O$	0.3 mg/L

FIGURE CAPTIONS

Figure 1. Schematic diagram of the experimental set-up used.

Figure 2. Profile of the pH variations for the different HRTs studied.

Figure 3. Profile of the TVFA variations for the different HRTs studied.

Figure 4. Variation of the experimental and theoretical values of the fractional conversion (obtained with the dispersion model using a first-order kinetics for substrate consumption) with the hydraulic retention time.

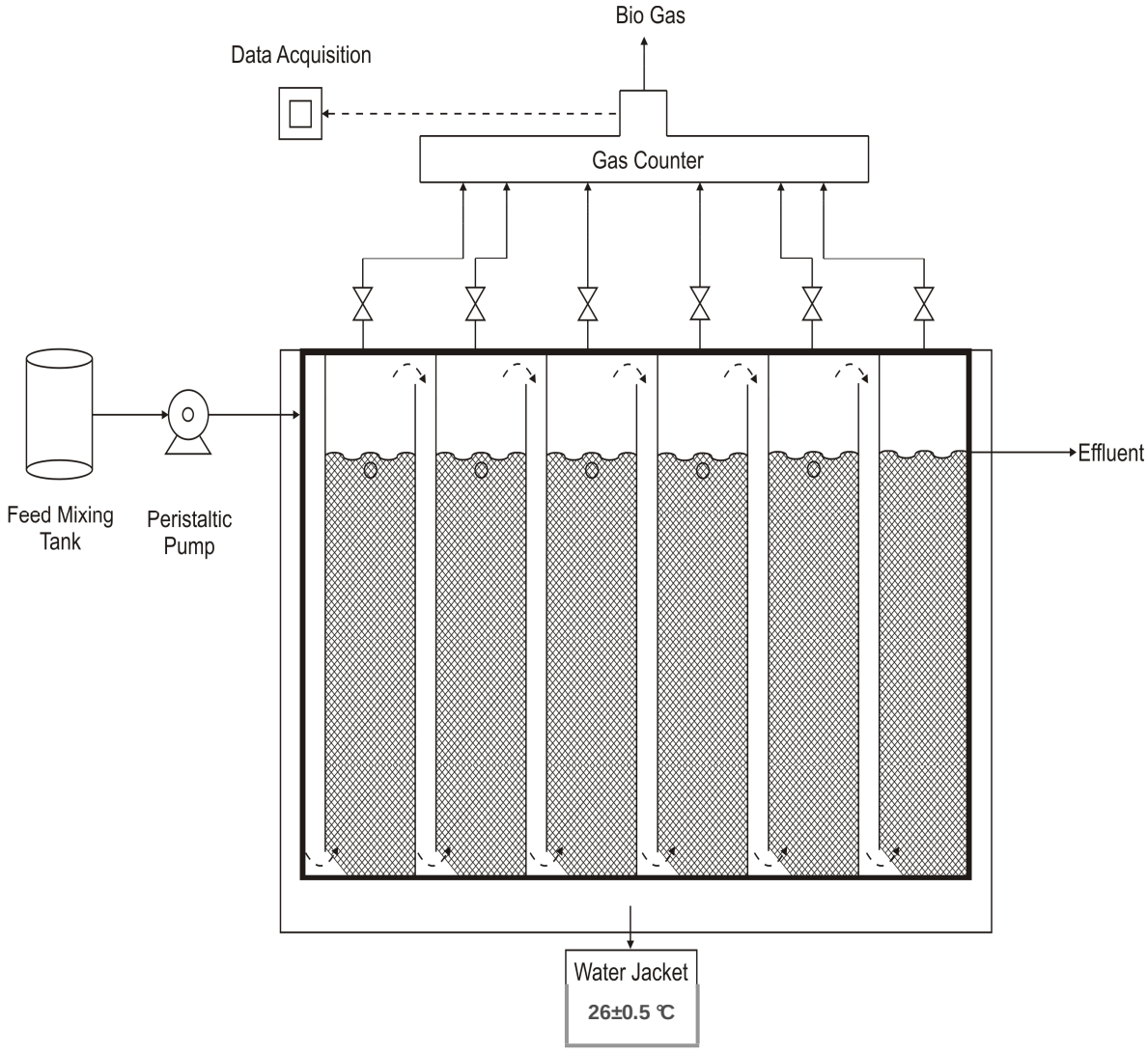
Figure 5. Flow diagram or detailed scheme of the different steps made for the calculation of the dispersion number and kinetic constant in the dispersion model.

Figure 6. Comparison of the experimental and theoretical values of the fractional conversion (obtained with the dispersion model using a first-order kinetics for substrate consumption) for all the experiments carried out.

Figure 7. Variation of the experimental and theoretical values of the fractional conversion (obtained with the empirical modified Young model) with the hydraulic retention time.

Figure 8. Comparison of the experimental and theoretical values of the fractional conversion (obtained with the empirical modified Young model) for all the experiments carried out.

Figure 1


Bio Gas
Data Acquisition
Gas Counter
Effluent
Feed Mixing Tank
Peristaltic Pump
Water Jacket
26±0.5 ℃

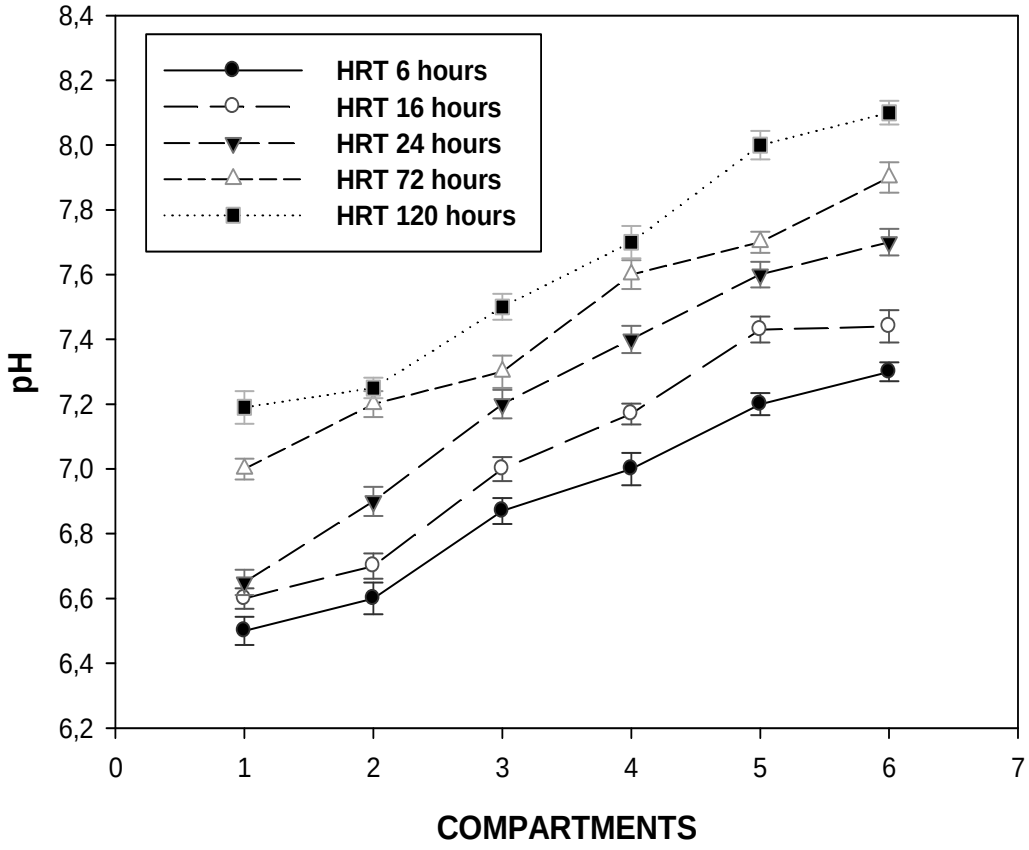

HRT 6 hours
HRT 16 hours
HRT 24 hours
HRT 72 hours
HRT 120 hours
pH
COMPARTMENTS
8,4
8,2
8,0
7,8
7,6
7,4
7,2
7,0
6,8
6,6
6,4
6,2
0
1
2
3
4
5
6
7

Figure 2

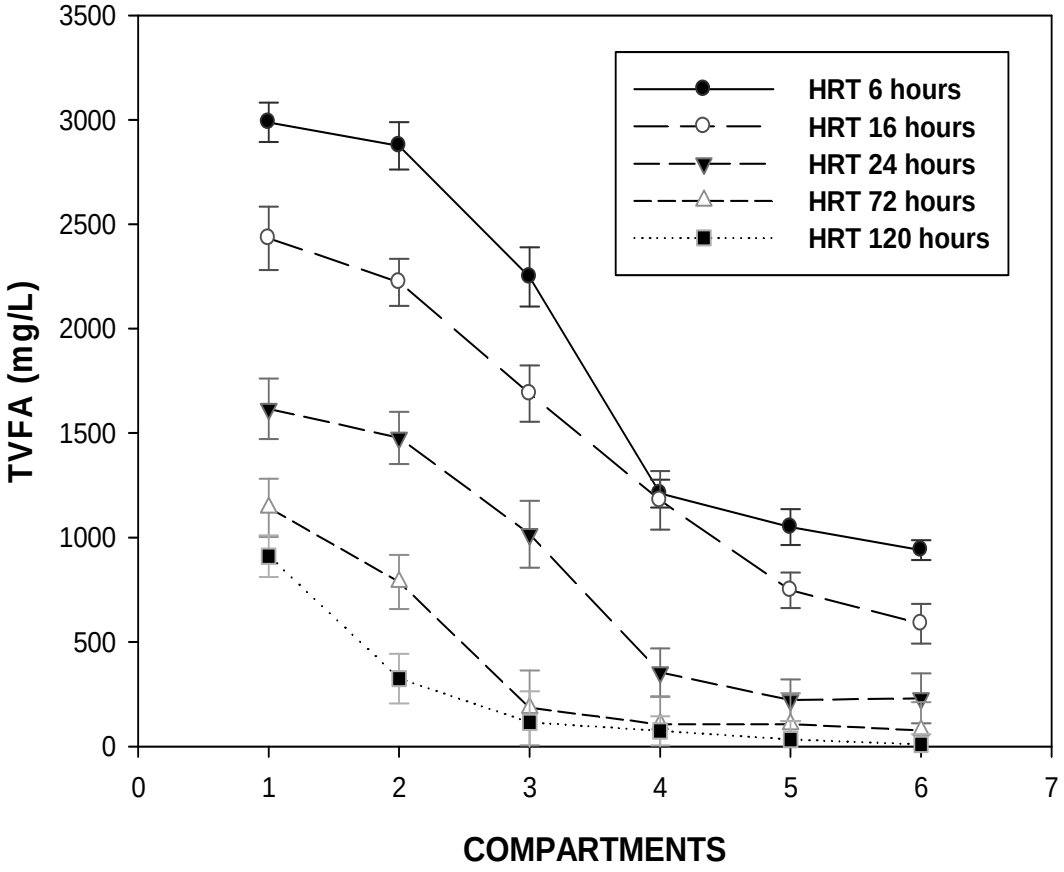

3500
3000
2500
2000
1500
1000
500
0
TVFA (mg/L)
0
1
2
3
4
5
6
7
COMPARTMENTS
HRT 6 hours
HRT 16 hours
HRT 24 hours
HRT 72 hours
HRT 120 hours

Figure 3

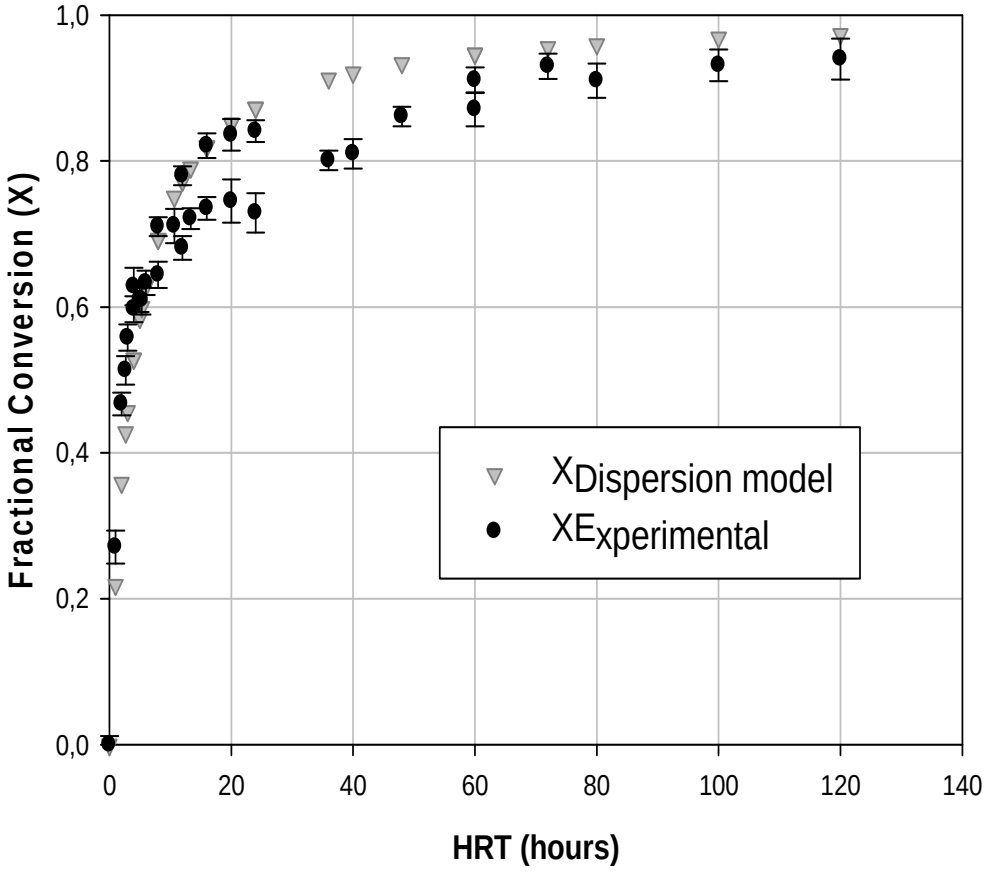
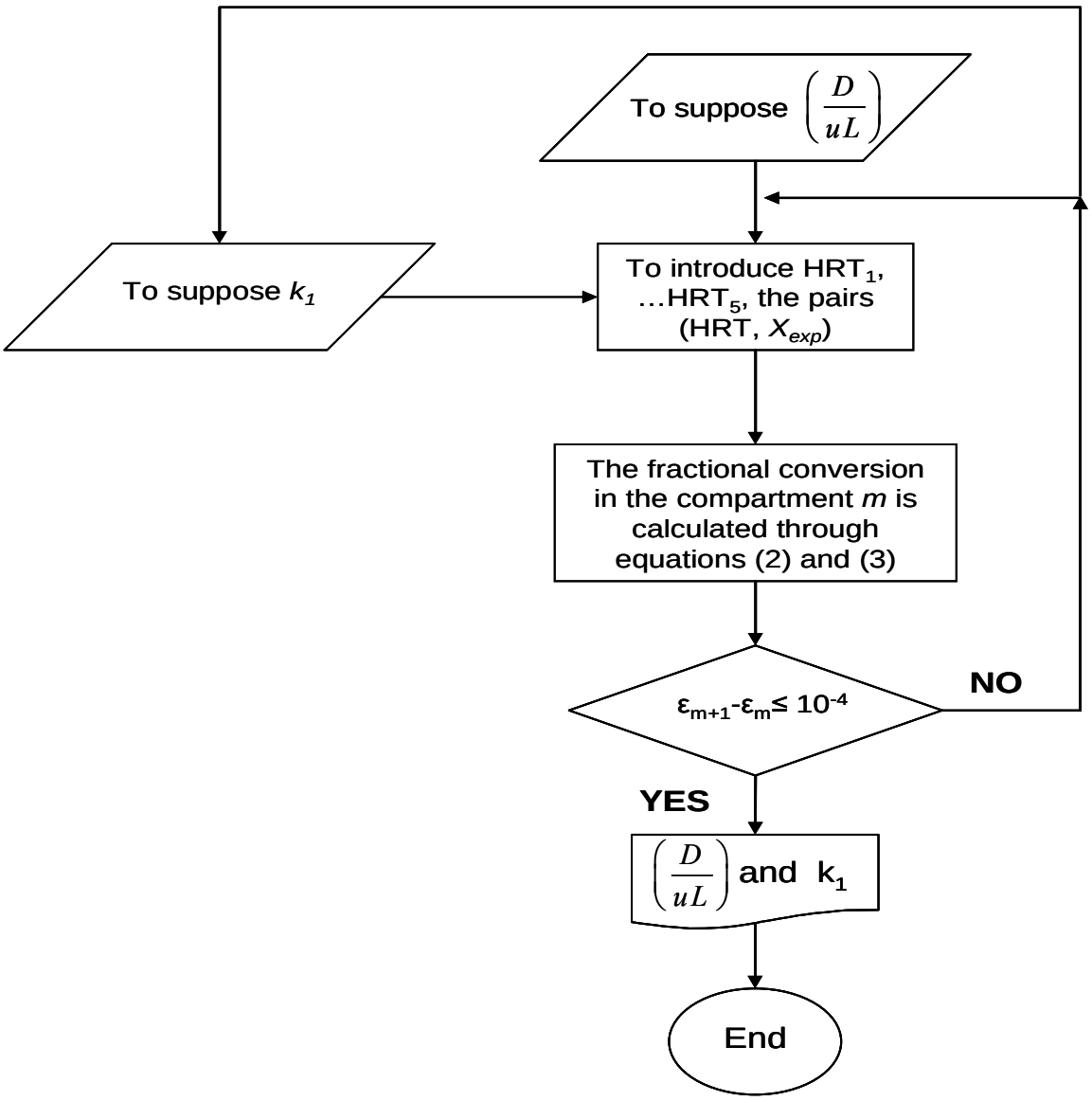

1,0
0,8
0,6
0,4
0,2
0,0
Fractional Conversion (X)
0
20
40
60
80
100
120
140
HRT (hours)
XDispersion model
XExperimental

Figure 4

699 **Figure 5**


To suppose $\left(\frac{D}{uL}\right)$
To suppose k_1
To introduce HRT_1, …HRT_5, the pairs (HRT, X_{exp})
The fractional conversion in the compartment *m* is calculated through equations (2) and (3)
$\varepsilon_{m+1}-\varepsilon_m \leq 10^{-4}$
NO
YES
$\left(\frac{D}{uL}\right)$ and k_1
End

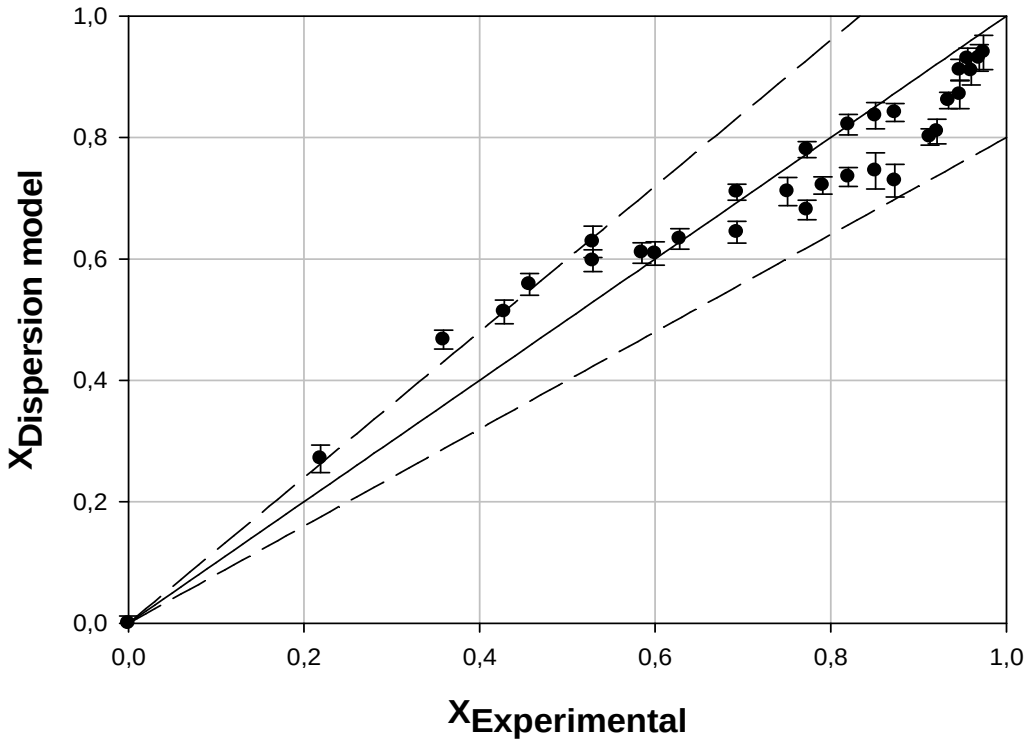

1,0
0,8
0,6
0,4
0,2
0,0
0,0
0,2
0,4
0,6
0,8
1,0
X Dispersion model
X Experimental

Figure 6

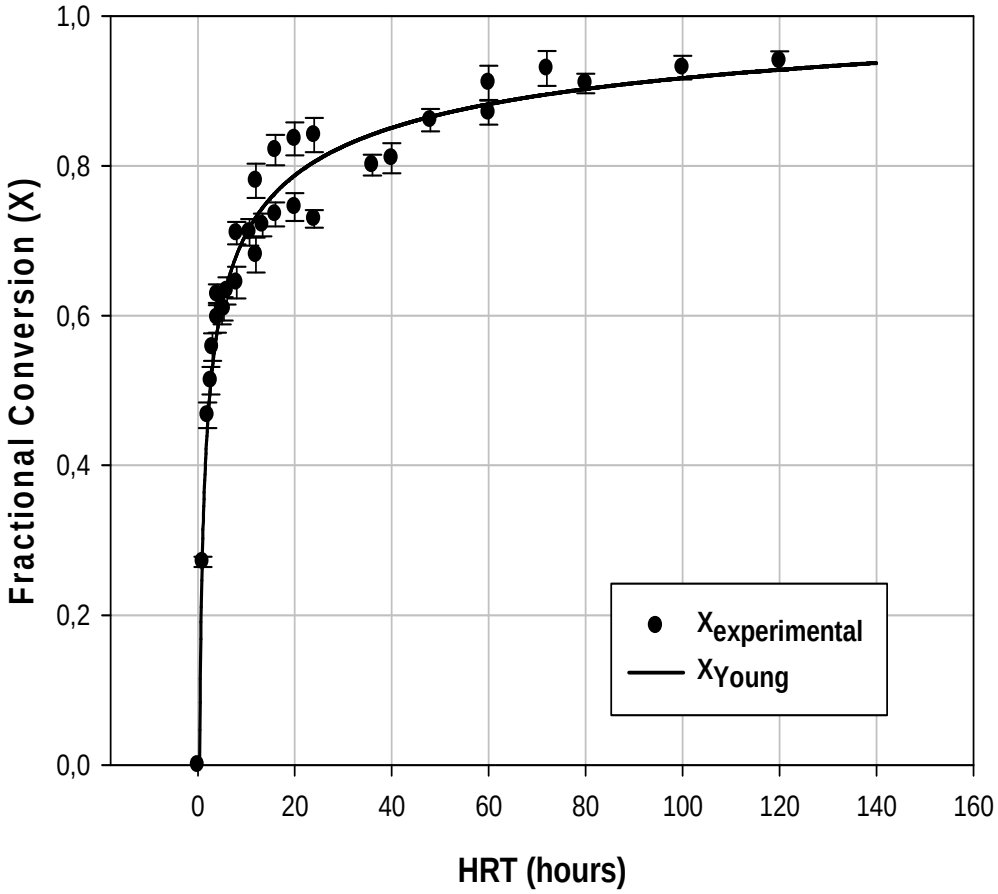

1,0
0,8
0,6
0,4
0,2
0,0
Fractional Conversion (X)
0
20
40
60
80
100
120
140
160
HRT (hours)
X experimental
X Young

Figure 7

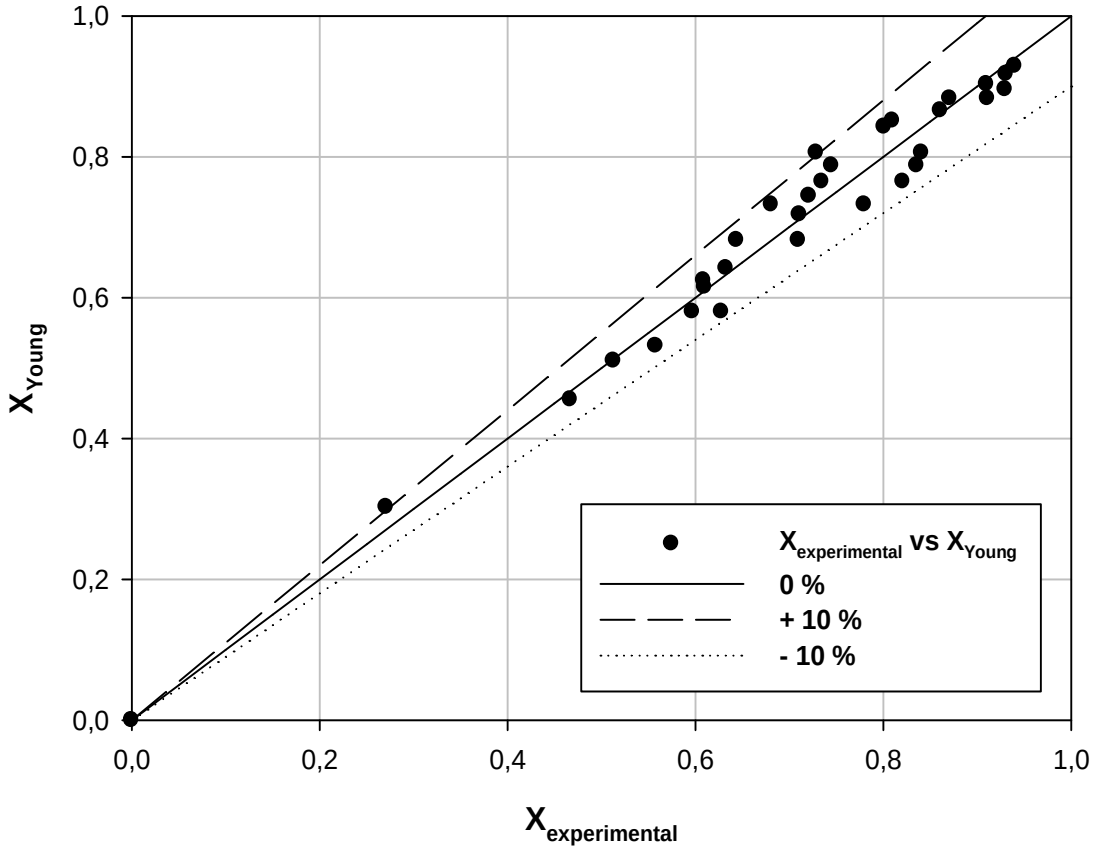

1,0
0,8
0,6
0,4
0,2
0,0
0,0
0,2
0,4
0,6
0,8
1,0
X Young
X experimental
X experimental vs X Young
0 %
+ 10 %
- 10 %

Figure 8