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Preliminary design for simultaneous saccharification and fermentation stages for ethanol production from sugar cane bagasse

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ABSTRACT

The interest of ethanol production from agro-industry residues as raw material has deserved great attention since these feedstocks may constitute an alternative to fossil fuels as well as represent beneficial effects both from economic and environmental point of view. The sugar cane bagasse (SCB) represents an available and cheap opportunity of raw material. The use of lignocellulosic biomass as a feedstock needs a pretreatment stage to breakdown lignocellulosic complex into lignin, cellulose and hemicelluloses, then the hydrolysis stage converts cellulose into fermentable glucose, and finally, in order to increase the glucose yield, simultaneous saccharification and fermentation (SSF) process is carried out. In this work a preliminary design for these stages is proposed. A mixed integer linear programming model is formulated in order to obtain the optimal number of parallel units in each stage and their sizes (plant configuration and capacity). The production planning along the time horizon of one year is determined, given by the number of batches to be processed and its size. Different production scenarios are analyzed, for which the investment cost is minimized, and economic profitability indicators are calculated.

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1. Introduction

Bioethanol can be produced from any lignocellulosic biomass, and sugar cane bagasse (SCB) is presented as an available and cheap opportunity of raw material. According to Hofsetz and Silva (2012), and Dantas et al. (2013), per ton of sugar cane, approximately 0.27–0.3 tons of bagasse are obtained. Although the SCB is usually burnt at the sugar mill for steam and electricity production, there still remains certain amount of residues available for second generation bioethanol production (Albernas-Carvajal et al., 2014; Narra et al., 2016).

Fuel ethanol from SCB can be obtained through pretreatment, enzymatic hydrolysis and fermentation process stages. These stages frequently involve thermo-mechanical and chemical pretreatments followed by either chemical or enzyme-catalyzed hydrolysis to release

fermentable sugars (Patel et al., 2017). Enzymatic depolymerization of polysaccharides is selective towards sugar production, but less amenable to industrial scale implementation due to the enzyme loading requirements and associated costs. Once the pretreatment is performed, there are several ways for carrying out saccharification and fermentation. There are four strategies documented for fermentation of cellulosic hydrolysate, namely, separate hydrolysis and fermentation (SHF), simultaneous saccharification and fermentation (SSF), simultaneous saccharification and co-fermentation (SSCF) and consolidated bio-processing (CBP) processes (Santosh et al., 2017). In the SHF, hydrolysis of lignocellulosic materials and fermentation of sugars are separately carried out. An advantage of the SHF process is that both hydrolysis and fermentation steps can be performed under their optimal conditions, mainly pH and temperature (Huang et al., 2015) and substrate concentration (Balat, 2011) while allowing cell recycling after fermentation, but the capital cost is high because two unit operations are needed (Neves et al., 2016).

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One of the limitations when using cellulase in SHF is the rates reduction due to end product (cellobiose and glucose) inhibition. The operation of integrated membrane reactors for the simultaneous cellulose hydrolysis and the removal of inhibition product can be performed using different configurations, as describe [Andrić et al. \(2010\)](#):

1. Simultaneous lignocellulose hydrolysis and removal of produced glucose in a stirred reactor (STR) for achieving additional conversion due to the available volume and/or unconvertible fraction discharge.
2. Integrated reaction vessel STR with ultrafiltration (UF) membrane.
3. Separate reaction vessel with:
 - a. UF membrane ([Henley et al., 1980](#); model system 3 from [Ghose and Kostick 1970](#)); combined with packed bed reactor (PBR) in [Yang et al. \(2006\)](#).
 - b. Ordinary (model system 1 from [Ghose and Kostick, 1970](#)) and special membrane reactor (with adsorption of substrate and enzymes from [Bélafi-Bakó et al., 2006](#)).
 - c. Microfiltration (MF) and UF for recovery of bounded and soluble enzymes, respectively ([Knutsen and Davis, 2004](#)).
 - d. UF and MF membrane for recovery of enzymes and removal of lignin reach fraction, respectively ([Ishihara et al., 1991](#)).

4. Separate reaction vessel with UF membrane with shell immobilized β -glucosidases (BG), in 1-stage, and 2-stages ([Klei et al., 1981](#)).

The specific mechanical design features in connection to the membranes are usually employed in order to reduce the negative effect that the hydrolysis reaction mixture has on membrane flux. However, as [Andrić et al. \(2010\)](#) affirm, a number of important challenges must be recognized to accomplish successful operation:

- The outlet (permeate) glucose concentrations is generally too low. To fully alleviate product inhibition by glucose product removal, the glucose levels in the reactor — and hence in the permeate, must be very low, typically less than 10 g/L to have an effect.
- Membrane fouling is pronounced a high substrate (dry matter) levels, and currently only low to medium dry matter levels are possible to process.
- Discharge of the non-convertible fraction which accumulates during extended reaction, especially with fed-batch operation is difficult. Removal of the lignin fraction prior to hydrolysis might be helpful.
- Scale-up of the membranes as well as the membrane reactors.

Simultaneous saccharification and fermentation (SSF) overcomes the problem of rates reduction due to end product (cellobiose and glucose) inhibition by hydrolyzing cellulose and fermenting the hydrolysis product at the same time ([Balat, 2011](#)). This process is also advantageous because both stages are carried out in a single reactor, but one of them can occur at conditions far from ideal. For this reason, there has been a great effort to isolate or develop enzymes that work best at pH and temperature conditions used for fermentation or organisms that are thermotolerant or even thermophilic, so that SSF could be performed at the optimal enzyme temperature of around 50 °C ([Neves et al., 2016](#)).

The SSF technology was studied by several authors employing different feedstock. [Scordia et al. \(2013\)](#) study the use of giant reed biomass pretreated with dilute oxalic acid. [You et al. \(2016\)](#) make a comparative study of SSF byproducts from sugarcane bagasse employing different pretreatment alternatives, while [Narra et al. \(2016\)](#) perform a comparison between SHF and SSF using dilute acid pretreated lignocellulosic biomass (rice straw, wheat straw, and sugarcane bagasse). [Franceschin et al. \(2008\)](#) present a complete model for ethanol production from corn and analyze different technical and economical scenarios.

SSF as its name indicates, has the advantage of executing simultaneously the enzymatic hydrolysis and fermentation stages on the same processing unit (bioreactor). This technology has been improved to include the co-fermentation of multiple sugar substrates, i.e., simultaneous saccharification of both cellulose (to glucose) and hemicellulose (to xylose), and co-fermentation of both glucose and xylose by genetically engineered microbes in the same broth ([Balat, 2011](#)). The glucose

produced from hydrolysis is simultaneously metabolized by microorganism to produce bioethanol as a product ([Triwahyuni et al., 2015](#)). Then, it represents a promising technology because end-product inhibition of cellulose in the hydrolysis step is minimized (as glucose is converted as it is released), and therefore higher solid levels are allowed ([Mesa et al., 2011](#)).

In addition, the SSF process is more economical, when compared with the SHF, because it can increase the ethanol concentration in the fermentation broth ([Wyman et al. 1992](#)), decreases the risk of contamination ([Balat, 2011](#)), needs minor enzyme dosage, and reduces the total process time ([Tomás, 2010](#)), and therefore, the energy costs during distillation are reduced ([Huang et al. 2015](#)). Moreover, the number of process equipment is reduced and consequently the investment cost ([Wahono et al., 2014](#)). Also, less energy is required and therefore production costs are also decreased and more sustainable process is achieved. Nevertheless SSF also presents some inconvenience, the optimal conditions of pH and temperature in hydrolysis and fermentation stage are different and commitment conditions between both should be established ([Tomás, 2010](#)).

The design problem deals with the plant configuration, i.e. number of units for each process stage, and units' sizing. This problem is usually modelled with constant size factors and fixed operating times ([Ravemark and Rippin, 1998](#); [Moreno et al., 2009](#); [Fumero et al., 2012](#)) or with dependent variables in more detailed models ([Bathia and Biegler, 1996](#); [Corsano et al., 2006](#); [Albernas-Carvajal et al., 2014](#)).

Production planning determines how to produce the required production according to the available resources in the desired horizon time. It involves the number and size of batches to be produced as well as the process cycletime decisions. In single product plants, as addressed in this work, no tradeoff among products are presented, and therefore the production planning problem is simpler. In [Albernas et al. \(2014\)](#) a similar problem was faced, but for pre-fermentation and fermentation stages operating in separated way. In that work, non linear programming (NLP) model were used and therefore, the global optimality cannot be assured.

Mathematical modeling and computing simulation can be exploited to maximize ethanol production in SSF, as was demonstrated by [Mutturi and Lidén \(2014\)](#) and [Paulova et al. \(2015\)](#).

In general, there are a few papers related to SSF process design. [Balat \(2011\)](#) affirms that the lignocellulosic biomass is the most promising feedstock considering its great availability and low cost, but the large-scale commercial production of fuel bioethanol from lignocellulosic materials has still not been implemented. [Elumalai and Thangavelu \(2010\)](#) determine the optimization variables for SSF using sugarcane bagasse, and evaluate some kinetics models, but the process design is not taken into account. [Akhtar et al. \(2016\)](#) make a characterization of microwave-alkali-acid pre-treated rice straw and scale up the shake-flask SSF under optimal conditions, but the industrial levels are not considered.

On the other hand, [Llamoso et al. \(2015\)](#) present a detailed design of SSF-distillation through a prototype based on laboratory results, where the equipment costs are included, but the pretreatment stages are not detailed.

In this work a preliminary design for SSF stages to ethanol production is proposed. The design problem is addressed through fixed size factors and processing times. With the aim of obtaining the fixed size factors, the results of a previous work ([Albernas-Carvajal et al., 2016](#)), which includes experimental data and kinetic models, are used. Therefore, the present work represents a real industrial case of an emerging technology. Although, a the fixed size factor approach is a simple formulation, due to the limited related work on SSF design, the proposed preliminary design represents an original contribution. A mixed integer linear programming (MILP) model is proposed and therefore, the optimal solution can be obtained. Acid and alkaline pretreatments and SSF stages are designed in order to assure the continuous feed to the distillation stage. In this way, units duplication is considered for reducing the process cycletime and consequently the idle times. The details of distillation stage were previously reported by [Albernas et al. \(2012\)](#).

The optimal number of parallel units in each stage and their sizes (plant configuration and capacity) is determined through the model.

The production planning along the time horizon of one year is also decided, given by the number of batches to be processed and its size. A formulation based on fixed sizes factors and times for discontinuous stages is proposed, which were obtained by experimental results of Albernas (2014) and Albernas-Carvajal et al. (2015). The model is implemented and solved in GAMS (Rosenthal, 2013). Different production scenarios are analyzed, for which the investment cost is minimized, and economic profitability indicators are calculated.

2. Process description

Fig. 1 shows the stages involved in ethanol production process from bagasse sugar cane considered in this work, where SSF technology is taken into account.

Bagasse sugar cane is usually used for producing electricity for the sugar cane plant, but there is an excess of bagasse that can be used for ethanol production taking advantage of their fermentable sugaring substrate. The bagasse is stored with nearly 60% of humidity.

The process involves two bagasse pretreatment stages or hydrolysis, with the objective of incrementing the global balances of the downstream conversion processes. The fiber structure is broken during the pretreatment for removing the lignin and the hemicelluloses in order to facilitate and increase the enzymatic activities. Thus, biomass is fractionated into their main components (cellulose, hemicelluloses, and lignin), cellulose crystallinity is reduced and the accessible surface area is increased (Dantas et al., 2013). A furfural-rich condensate is obtained through the dilute acid hydrolysis. For this stage, the biomass is diluted with sulfuric acid with 1% (w/v) in dry fiber, 175 °C and 9 atm of steam, 1:1 (w/v) of solid-liquid ratio and residence time equal to 40 min. After washing stage, a solid rich-glucose solution is obtained for continuing with the pretreatment second stage. The solid is treated with an ethanol solution: 30% v/v, 185 °C of temperature, 1:7 kg/L solid-liquid relation, and 60 min. of residence time. In this stage, the ethanol loss is equal to 1% with respect to the dry fiber (Albernas-Carvajal et al., 2016). Then filtration is carried out for reducing 20% the lignin content with respect to the initial raw material.

Finally, SSF process is performed using a mixture of enzymes composed of a cellulolytic complex CellicRCTec2 and β -glucosidase (NS50010), provided by Novozymes A/S (Denmark), which was thoroughly studied by Bansal et al. (2009). Fermentation is carried out by *Saccharomyces cerevisiae* yeast No. 1701 (DER CIEMAT Culture Collection, Spain). SSF is performed in 100 mL Erlenmeyer flasks containing 25 mL of medium in 0.05 M tampon sodium citrate with 4.8 of pH, 30–35 °C of temperature and incubated in a rotatory shaker (Certomat-R, B-Braun, Germany) at 35 °C, 150 rpm, for 50 h, with 10 or 15% (w/v) of solid load (Albernas-Carvajal et al. 2016).

In this work, as was previously mentioned, the distillation and rectification stages are not considered.

3. Problem statement

The design model for pretreatment (acid and basic hydrolysis) and SSF stages is presented in this section. The objective is to develop a batch design and planning approach in order to implement this technology in existing ethanol plants, assuring the optimal loading to the distillation units, which is a continuous stage. The fixed size factors and time formulation is used for this purpose. A key issue in this approach is

the size factor calculation. For this task, experimental data from Albernas-Carvajal et al. (2016) were used for obtaining the streams involved in the process and describing the mass balances. Through these mass balances, the size factors were calculated.

Let t_j be the processing time and S_j the size factor for stage j . It is worth to mention that during the processing time, no materials enter or leave the process unit, and the size factor represents the size needed at stage j to produce 1 mass unit of final product. The problem consists in determining the number of batches of hydrolyzed bagasse, N_b , its size, B , and the optimal design for the involved stages (number of duplicated unit per stage, M_j , and its size, V_j), in order to reach the required production Q in the horizon time H at a minimum cost.

4. Model formulation

For the optimal design, a mathematical programming approach is proposed. According to Grossmann and Sargent (1979), the unit size for stage j , is calculated by:

$$V_j \geq S_j B \quad (1)$$

where B represents the final product batch, in this case, the hydrolyzed bagasse.

A batch of final product is produced each CT hours, that is, the process cycle time. A total of N_b batches are produced on the time horizon H .

Then, the total produced amount Q is given by the equation:

$$Q = B N_b \quad (2)$$

This amount must be reached on the time horizon:

$$CT N_b \leq H \quad (3)$$

where the cycletime is defined as:

$$CT \geq \frac{t_j}{M_j} \quad (4)$$

M_j represents the number of out-of-phase parallel units at stage j , which is an optimization variable. In this way, the zero-wait transfer policy between stages is adopted in order to ensuring the quality of these operations, given the nature and characteristics of microbiological sugary substrates.

The objective function is the minimization of investment cost, calculated according to the number of units duplicated out of phase and its sizes:

$$C_{inv} = C_{ann} CCF \sum_j \alpha_j M_j V_j^{\beta_j} \quad (5)$$

where C_{ann} is a constant that annualizes the inversion, CCF is a capital charge factor for the investment cost, and α_j y β_j are cost coefficients.

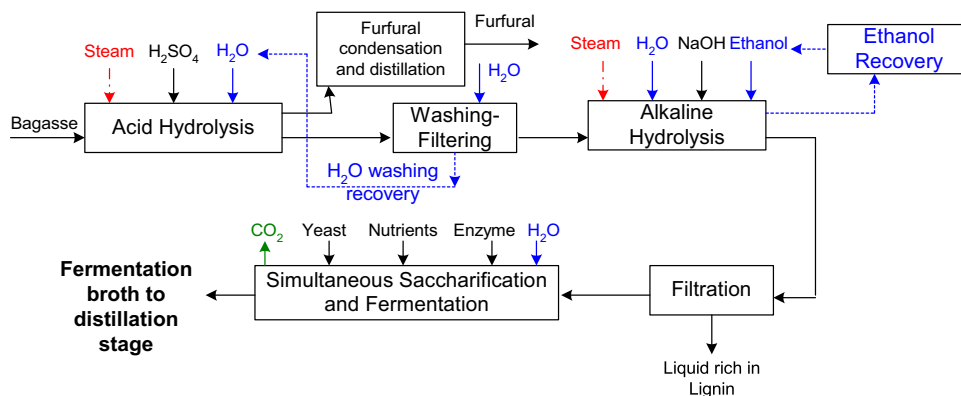


Fig. 1 – Stages considered in the design model.

Therefore, the overall model is a mixed integer non linear (MINLP) formulation expressed as:

$$\begin{aligned} \text{Min } & C_{ann} CCF \sum_j \alpha_j M_j V_j^{\beta_j} \\ \text{s.t. : } & V_j \geq S_j B \\ & CT N b \leq H \\ & CT \geq \frac{t_j}{M_j} \\ & Q = B N b \\ & V_j, B, CT \geq 0; N b, M_j \in \mathbb{Z}^+ \end{aligned} \quad (\text{MINLP})$$

In order to overcome the difficulties associated to MINLP resolutions approaches and assure the global optimal solution, a model reformulation is proposed.

Let V_{Fix_j} be the set of discrete unit sizes for stage j , $V_{Fix_j} = \{VF_{j,1}, VF_{j,2}, \dots, VF_{j,p_j}\}$. Then, a binary variable z_{jp} , which takes value 1 if unit size p is selected for stage j , is defined:

$$V_j = \sum_{p=1}^{p_j} VF_{jp} z_{jp} \quad (6)$$

$$\sum_{p=1}^{p_j} z_{jp} = 1 \quad (7)$$

Eq. (6) defines the unit size of stage j , while Eq. (7) states that only one size from the set V_{Fix_j} can be selected. On the other hand, as Nb is an integer variable, it can be rewritten for linearizing Eq. (2):

$$N b = \sum_{n=1}^{N^{UP}} n x_n \quad (8)$$

$$\sum_{n=1}^{N^{UP}} x_n = 1 \quad (9)$$

Therefore,

$$Q \leq \sum_{n=1}^{N^{UP}} n B x_n \quad (10)$$

where x_n is a binary variable that takes 1 if n batches are produced, and 0 otherwise. Bx_n is the new continuous variable representing the product $B \cdot x_n$, such that:

$$Bx_n \leq B^{UP} x_n \quad (11)$$

$$Bx_n \leq B \quad (12)$$

$$Bx_n \geq B - B^{UP}(1-x_n) \quad (13)$$

Also, Eq. (3) can be linearized using Eq. (14), through a new continuous variable CTx_n , which satisfied Eqs. (15)–(17):

$$\sum_{n=1}^{N^{UP}} n CTx_n \leq H \quad (14)$$

$$CTx_n \leq CT^{UP} x_n \quad (15)$$

$$CTx_n \leq CT \quad (16)$$

$$CTx_n \geq CT - CT^{UP}(1-x_n) \quad (17)$$

Finally, the integer variable M_j can be expressed as:

$$M_j = \sum_{k=1}^{M_j^{UP}} k y_{jk} \quad (18)$$

$$\sum_{k=1}^{M_j^{UP}} y_{jk} = 1 \quad (19)$$

in order rewrite the objective function as:

$$C_{ann} CCF \sum_j \sum_p \sum_k \alpha_j k y_{jk} z_{jp} VF_{jp}^{\beta_j} \quad (20)$$

As can be noted, Eq. (20) has the bilinearity $y_{jk} z_{jp}$. Then, a new continuous variable, yz_{jkp} , is defined with the objective of reach a linear formulation:

$$yz_{jkp} \leq y_{jk} \quad (21)$$

$$yz_{jkp} \leq z_{jp} \quad (22)$$

$$yz_{jkp} \geq y_{jk} + z_{jp} - 1 \quad (23)$$

Table 1 – Model parameters for process design.

Stage	Inlet flux (m ³ /d)	SF _j (m ³ /kg)	α _j	β _j
Acid hydrolysis	2290.86	2.48 · 10 ⁻³	1560	0.68
Alkaline hydrolysis	1319.00	1.43 · 10 ⁻³	1560	0.68
SSF	2843.52	3.08 · 10 ⁻³	14	0.72

$$0 \leq y_{z_{jpk}} \leq 1 \quad (24)$$

Therefore, the overall linear problem is formulated by (MILP):

$$\begin{aligned} \text{Min } & C_{ann} CCF \sum_j \sum_p \sum_k \alpha_j k y_{z_{jpk}} V_{F_{jp}}^{\beta_j} \\ \text{S.t.: } & V_j \geq S_j B \\ & V_j = \sum_{p=1}^{P_j} V_{F_{jp}} z_{jp} \\ & Q = \sum_n n B x_n \\ & \sum_n n C T x_n \leq H \\ & C T \geq \sum_k \frac{t_j y_{jk}}{k} \\ & \sum_p z_{jp} = 1, \sum_k y_{jk} = 1, \sum_n x_n = 1 \\ & Y Z_{jpk} \leq y_{jk} \\ & Y Z_{jpk} \leq z_{jp} \\ & Y Z_{jpk} \geq y_{jk} + z_{jp} - 1 \\ & B x_n \leq B^{UP} x_n \quad C T x_n \leq C T^{UP} x_n \\ & B x_n \leq B \quad C T x_n \leq C T \\ & B x_n \geq B - B^{UP}(1 - x_n) \quad C T x_n \geq C T - C T^{UP}(1 - x_n) \\ & \sum_n B x_n = B \quad \sum_n C T x_n = C T \\ & B, B x_n, C T, C T x_n \geq 0 \quad x_n, y_{jk}, z_{jp}, Y Z_{jpk} \in \{0, 1\} \end{aligned} \quad (\text{MILP})$$

5. Study cases

The model was implemented and solved in GAMS. It involves 34,341 equations, 5141 binary variables, and 11,446 continuous variables. The resolution time is about 30 s for the studied cases.

In order to determine the size factors for the involved stages, experimental analysis, according to that described in Section 2, were carried out (Albernas-Carvajal et al., 2016), and the proper mass balances were calculated. The model parameters are summarized in Table 1. The processing times used in that study are adopted in this work, and are displayed in Table 2. The capital charge factor for the investment cost, CCF, is equal to 1.413. The considered C_{ann} coefficient is equal to 0.1, i.e. ten years.

A set of fifteen unit sizes is considered for each process stage. For acid and alkaline hydrolysis the sizes are {30, 45, 50, 60, 70, 80, 90, 100, 110, 120, 130, 140, 150, 160, 180}, while for SSF are {150, 175, 185, 200, 205, 210, 220, 225, 230, 240, 245, 250, 255, 260, 270} all measured in m³. Moreover, up to 32 parallel units for each stage is considered, but it would be expected that only SSF stage might use a considerable number of parallel units due to its operating time (time limiting stage). The horizon

Table 2 – Processing times for each stage.

Operation	Acid pre-treatment (min)	Alkaline pretreatment (min)	SSF (h)
Loading time	45	45	3
Heating time	35	40	
Reaction time	12.46	0.33	50
Cooling time	20	35	2
Unloading time	45	45	

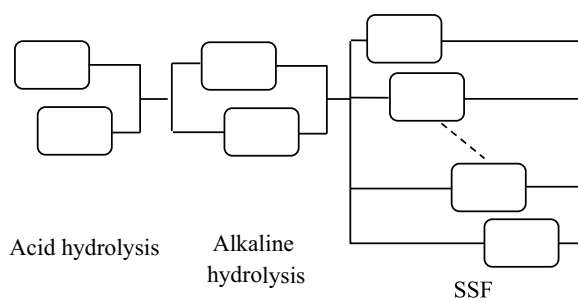


Fig. 2 – Optimal plant design considering SSF reaction time equal to 50 h.

Table 3 – Economical analysis for SSF reaction time equal to 50 h.

	SSF operation time equal to 50 h
Net present value (\$)	5,954,677.37
Internal rate of return (%)	61
Pay back period (years)	3.1
Profitability (\$/year)	1,633,501.97
Total production cost (\$/year)	5,866,498.03
Unitary cost (\$/hl)	39.11

time is equal to 7200 h and the annual required production is 38426.175 kg/h of fermentation broth.

In Section 5.1, the results for the presented model data are analyzed, and according to that, a new study case is performed and presented in Section 5.2.

5.1. Results for SSF reaction time equal to 50 h

The optimal plant design for this case involves two parallel units out of phase for the pretreatment stages (acid and alkaline hydrolysis) and 32 units in parallel for the SSF. Fig. 2 shows the plant configuration while Fig. 3 displays the production planning. The unit duplication allows reducing the production cycle time, which is equal to 1.719 h and is reached at the SSF stage. In this way, the required production is fulfilled in the horizon time through 4156 batches. It is worth noting that the maximum number of available units for SSF is used. If fewer units in parallel are utilized, the cycle time is increased and the production cannot be satisfied in the fixed horizon time. Also, for fewer duplicated units, bigger unit sizes are needed, which is not possible due to limited capacity are adopted according to the usual industrial policies. In this case, the acid pretreatment adopt the maximum available unit size.

The annualized investment cost for this solution is \$428508, and for the obtained design it is possible to calculate the profitability dynamic indexes, like NPV and IRR among others. Hydrous ethanol with 93.5°GL is produced and the selling price is 50 \$ per hl. In Table 3, the economical analysis of this alternative is presented. It can be noted that to produce

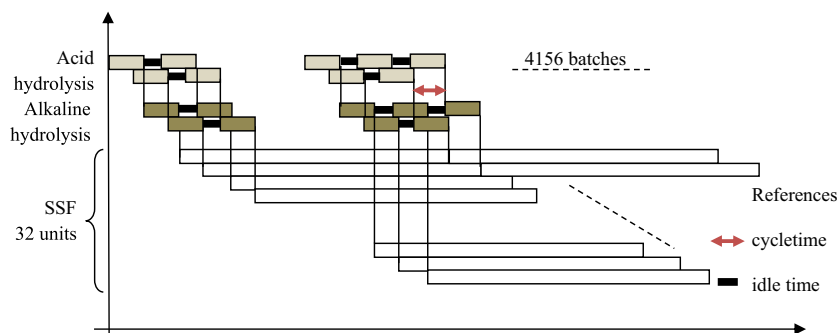


Fig. 3 – Production planning considering SSF reaction time equal to 50 h.

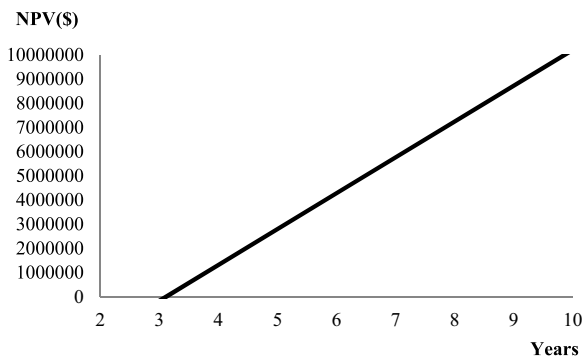


Fig. 4 – NPV vs. time considering SSF reaction time equal to 50 h.

ethanol through this alternative is profitable. Fig. 4 depicts the NPV vs. time (in years) profile, where the process feasibility can be noted, with a payback period of 3.1 years. Anyway, the long reaction time of SSF affects the total processing time and therefore, the process design.

On the other hand, from the experimental results (Albernas-Carvajal et al., 2016), it was determined that the ethanol concentration growth at the SSF stage between 29.5 and 45.5 h, increases slowly from 19.88 g/L to 22 g/L. Fig. 5 shows the ethanol growth through the SSF processing time.

Therefore, it is reasonable to reduce the reaction time of SSF stage, and analyze the profitability of the process. In the next section, this case is presented.

5.2. Results for SSF reaction time equal to 30 h

The model is performed for all the parameters defined in the previous section except for the SSF reaction time which is fixed to 30 h.

The optimal design for this case consists of 2 parallel units out of phase for acid and alkaline pretreatment stages, and 22 units out of phase for the SSF stage. The cycle time is equal to 1.591 h and 4461 batches are needed to fulfill the required production. The plant flowsheet and the Gantt chart are similar to that presented in Figs. 2 and 3 respectively, except for the number of units for the SSF stage.

In Table 4, the selected unit sizes and the occupied size in each one for both studied cases is shown. The occupied size represents the batch volume in each stage.

In Table 5, the economical results for this case are summarized. As can be noted, the process is also profitable, with a payback period of 2.5 years as depicts Fig. 6.

When the SSF reaction time is reduced from 50 to 30 h, the ethanol concentration is only 2 g/L decreased, but from the technological point of view, some advantages are reached: the

Table 4 – Optimal design characteristics for both studied cases.

50 h of SSF			30 h of SSF		
Number of units	Size (m ³)	Occupied size (m ³)	Number of units	Size (m ³)	Occupied size (m ³)
Acid pretreatment			Acid pretreatment		
2	180	165	2	160	154
Alkaline pretreatment			Alkaline pretreatment		
2	100	95	2	90	89
SSF			SSF		
32	210	205	22	200	191

Table 5 – Economical analysis for SSF reaction time equal to 30 h.

	SSF operation time equal to 30 h
Net present value (\$)	6,697,136.59
Internal rate of return (%)	75
Pay back period (years)	2.5
Profitability (\$/year)	1,718,820.18
Total production cost (\$/year)	5,781,179.82
Unitary cost (\$/hl)	38.54

stage working time is 20 h reduced, less units are used for the SSF, the cycle time is also reduced, and therefore more batches can be produced in the horizon time. When decreasing the batch sizes, also the unit sizes are decreased and therefore, a lower investment cost is obtained. The investment cost is reduced to \$358951, which represents a reduction of 16.23%. Thus, the entire dynamic economic factors (NPV, IRR and payback period) are improved.

It is worth highlighting that the economic indexes for this technology are more favorable than SHF technology addressed by Albernas-Carvajal et al. (2014), since the ethanol unit price is reduced from \$41.97 to \$38.54 per hl, for the same conditions, for processes attached to a distillery. This result is similar to that reported by Lux Research (2016) and Sapp (2016), and smaller than that obtained via simulation by da Silva et al. (2016).

5.3. Parametric analysis

In this section, some remarks about the two studied cases are stated. As it can be observed, the SSF processing time is a critical model parameter, since when it is reduced, the number of parallel unit at SSF stage is also reduced as well as the investment cost. Table 6 shows this effect when the model is executed for different SSF processing times.

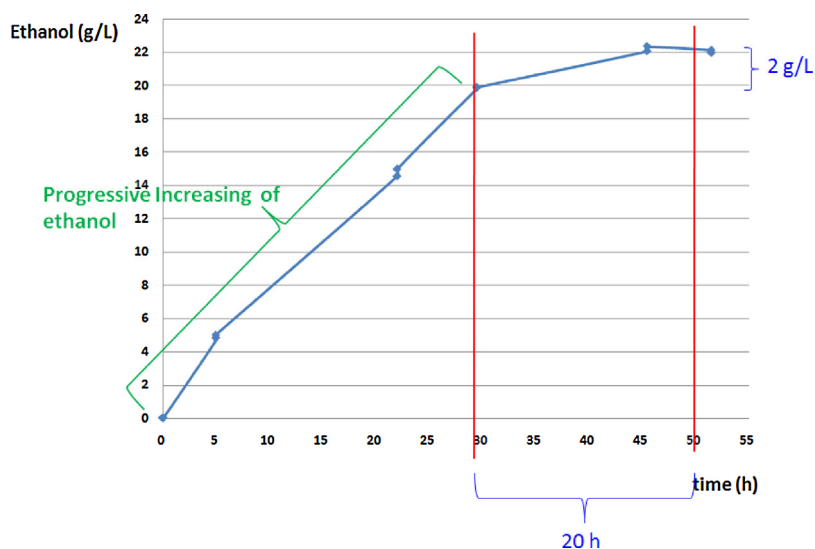


Fig. 5 – Ethanol concentration vs. SSF reaction time.

Table 6 – SSF optimal configuration for different processing times.

t_{SSF} (h)	M_{SSF}	V_{SSF} (m ³)	CT (h)	Nb	C_{inv} (\$)
35	26	185	1.5	4680	363,466
40	28	200	1.6	4480	385,110
45	31	200	1.6	4468	398,190



Fig. 6 – NPV vs. time considering SSF reaction time equal to 30 h.

It is worth to mention that a disadvantage of this approach is that the use of fixed times for the process stages, do not allow to find the optimal processing time for producing a given amount of product. In a more detailed formulation, processing time can be depended on the ethanol concentration, for example, through the corresponding kinetic equation. But this type of models correspond to non linear programming formulations (NLP or MINLP), where the global optimum solution cannot be assured. Given the complexity to represent the kinetics through algebraic equation for formulation the optimization model, the preliminary design proposed in this work as a first approximation approach is a valid representation.

6. Conclusions

In this work, a mathematical framework is presented for determining the optimal design of pretreatment and SSF stages of an ethanol plant from sugarcane bagasse. The model allows obtaining the number of units for each stage and their sizes, as

well as planning variables as the number and size of batches to be produced in the horizon time.

The formulation serves as a tool for analyzing different production scenarios like processing time variations and the use of out-of-phase parallel units, and permits assessing the different tradeoff among design and planning variables (number of batches vs. units sizes, units duplication vs. processing cycletime, etc.). All these elements can be used for determining the optimal investment cost and therefore, the economic indexes can be evaluated. From results analysis, it was concluded that considering the reaction time of SSF stage equal to 30 h is more profitable and technologically feasible than when this stage works 50 h.

Although the limitation of the fixed time formulation, considering the lack of works on plant design for this emerging technology to produce ethanol from lignocellulosic biomass, according to the results obtained from the different studied cases, this approach represents a good approximation for the preliminary design.

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