Assessing the energy requirements for butanol production using fermentation tanks-in-series operated under vacuum

Victor Hugo Grisales Díaz\textsuperscript{a,b,*}, Mark J Willis\textsuperscript{c}, Moritz von Stosch\textsuperscript{c}, Gerard Olivar Tost\textsuperscript{d,e}, Oscar Prado-Rubio\textsuperscript{b}

\textsuperscript{a} Universidad Libre Seccional Pereira, Facultad Ciencias de la Salud. Grupo de investigación en Microbiología y Biotecnología MICROBIOTEC, Belmonte Avenida Las Américas

\textsuperscript{b} Facultad De Ingeniería y arquitectura, Departamento de Ingeniería Química, Universidad Nacional de Colombia – Sede Manizales, Cra. 27 # 64-60, Manizales, Colombia.

\textsuperscript{c} School of Engineering, Newcastle University, NE1 7RU, Newcastle upon Tyne, United Kingdom

\textsuperscript{d} Facultad De Ciencias Exactas y Naturales, Dpto. de Matemática y Estadística, Universidad Nacional de Colombia – Sede Manizales, Cra. 27 # 64-60, Manizales, Colombia.

\textsuperscript{e} Department of Natural Science and Technology, Universidad de Aysén, Coyhaique, Chile

* Corresponding author: yhrgrisalesd@unal.edu.co; victorh.grisalesd@unilibre.edu.co
Abstract

The acetone-butanol-ethanol (ABE) fermentation from corn stover was considered. We propose, study and optimise, via process simulation (using MATLAB® and Aspen Plus®) the use of fermentation tanks-in-series and in situ product recovery by vacuum evaporation. As the operating time of continuous fermentation processes is usually limited at less than 500 h, shutdown and start-up of the reactors are considered and optimised. A multi-objective optimisation methodology that considers economics as well as the energy requirements of the process was used. The optimal configuration was found to be five fermentation tanks-in-series where the first and the last reactors are operated at atmospheric pressure, and the intermediate vessels are operated under vacuum. The economic potential using this configuration was found to be 45% higher than that of vacuum fermenters operating in parallel (continuous operating mode). Also, the total fuel requirements for ABE recovery and purification system were as low as 7 MJ kg\(^{-1}\) butanol, a reduction of between 4 and 33% when compared to a parallel configuration of batch, fed-batch or continuous fermenters. The energy efficiency of this recovery and reaction system was as high as 74% when co-generation is considered.

Keywords: multi-objective optimisation; start-up/shutdown of reactor; vacuum fermentation; butanol production; energy efficiency; tank-in-series
Nomenclature

AVA, tank-in-series using reactors operating at atmospheric pressure and under vacuum

ABE, Acetone-Butanol-Ethanol

$C_{products}$, Selling prices of the products (acetone, butanol, and ethanol)

$EE_{RR}$, the energy efficiency of the reaction and recovery system

$EP$, economic potential (USD year$^{-1}$)

$F\_P$, inlet feed-stream rate (kg h$^{-1}$)

$F\_G$, outlet gases flow of the reactor (kg h$^{-1}$)

$F\_P$, purge flow rate from the reactor (kg h$^{-1}$)

ISPR, in situ product recovery

ISPR $-$ V, ISPR by vacuum separation

$FR\_RR$, fuel requirements of the reaction and recovery system (MJ-fuel/kg-ABE)

$LHV\_P$, the lower heating value of the products (MJ-fuel/kg-ABE)

$LHV\_S$, the lower heating value of the substrate (MJ-fuel/kg-ABE)

$N_i$, amount of component ‘$i$’ in the reactor (kg)

$Q_D$, the energy requirements of the distillation system (MJ/kg-ABE).

$R$, fermentation reaction rate (kg h$^{-1}$ m$^{-3}$)

$R_w$, hydrolysis reaction rate (kg h$^{-1}$ m$^{-3}$)

$t$, fermentation time (h)

$t_a$, annualised operational time (h)

$t_r$, return on investment time (years)

$TOAC$, total operative annualised costs (USD/h)

$TIC$, total investment costs (USD)

$V\_P$, the volume of the phase liquid in the reactor (m$^3$)

$V\_T$, volume total (liquid+solid) in the reactor (m$^3$)

$\chi_A$, the mass fraction of the feed stream of the reactor $m$ (g g$^{-1}$)
Introduction

Environmental concerns and the growing number of cars on the road, e.g., 97 million units were produced in 2017, may drive the market to more efficient transportation fuels [1]. Butanol offers an attractive alternative to ethanol, as it has an energy density that is 45% higher [2]. Due to its higher energy content, butanol was approved in 2018 as a gasoline additive (up to 16% volume ratio) by the U.S. Environmental Protection Agency. Furthermore, bio-butanol may be converted to petrol, diesel, and jet fuels [3].

Butanol is conventionally produced by (acetone-butanol-ethanol) ABE fermentation. Currently, twelve butanol fermentation plants operate world-wide, one in Brazil, and eleven in China [4]. In these industrial-scale plants, fermentation tanks-in-series and batch reactors are usually used [5]. It has been suggested that the use of fermentation tanks-in-series is the better option as it gives productivities between 20 and 50% higher than that of batch operation [6,7] when between three to eight fermentation tanks are used [8,9]. Nevertheless, because of butanol inhibition, productivity is still low (<0.5 g L⁻¹ h⁻¹). Furthermore, a low substrate feed concentration (<60 g L⁻¹) must be used in order to obtain increased conversions [10–12]. As a direct result of the low substrate concentration, product recovery has significant energy requirements and consequently, high greenhouse gas emissions [2].

As high substrate concentrations are required to reduce recovery costs, in situ product recovery (ISPR) schemes are usually proposed as an alternative to conventional production methods.
In these processes, substrate concentration, butanol productivity, and product concentration increase as butanol is selectively removed from the fermentation broth [14]. Hence, in the literature, several configurations of tank-in-series and ISPR have been proposed to reduce energy requirements and increase productivity and substrate conversion. For instance, Bankar et al., proposed a two stages reaction system with ISPR by liquid-liquid extraction [15] that increases glucose conversion in 30% compared to a single chemostat, or Van Hecke and De Wever 2017 proposed a two-stage continuous reaction system using ISPR by pervaporation [16] that has a higher butanol productivity and solvent concentration than a fed-batch operation. Due to the difficulties of operating 4-6 fermentation tanks-in-series and ISPR at the laboratory and plant-pilot scale, the number reactors operated in most of the research papers are usually limited to two reactors [5]. Hence, the use of process system engineering tools offers an attractive method to optimise performance via simulation.

In this paper, vacuum recovery is selected for ISPR since it has been recently shown to be a more attractive choice than pervaporation from both an economic and operational viewpoint [17]. However, the high energy requirement of this technology is still one of the main constraints to a more extensive industrial application [18,19]. Consequently, several options, including absorption, adsorption, condensation, and compression units, have been suggested in the literature to improve vacuum fermentation. For instance, Pereira et al., (2017) studied three configurations to recover 2-butanol from the gas stream in vacuum fermentation: temperature sweep adsorption, a multi-stage compression train, and vapour absorption [18]. The energy requirements of these three schemes are between 17.6 and 25.3 MJ kg⁻¹ 2-butanol, and vapour absorption was found to be the most attractive option from an energy perspective [18].

Alternatively, heat-pump systems have also been proposed. In a heat pump system, the vapour stream is compressed, and the condensation energy is used to supply the energy required for the evaporation [20]. Mariano et al., 2011 [21] proposed a heat-pump system for ABE recovery using two compression stages, achieving energy requirements as low as 13.4 MJ kg⁻¹ butanol. However,
these energy requirements are still high, as efficient recovery methods such as pervaporation, liquid-liquid extraction or adsorption require around 8 MJ kg\(^{-1}\) butanol [22].

In previous work, substrate and several product recovery configurations using heat-pump systems have been optimised [23]. Compression was performed in three stages, and partial condensation was used. The total energy requirements (compression work and vapour required in a heat-integrated distillation system) were found to be 7.3, 7.2 and 8.7 MJ kg\(^{-1}\) butanol for batch, fed-batch and continuous reactors operating in parallel [23]. Although continuous operation, with a parallel reactor configuration, is advantageous as this requires the minimum number of reactors, this is suggested to be the least attractive option as it has the highest energy requirements and as a result, the lowest economic potential.

To overcome the high energy requirements of a continuous vacuum process with fermenters operating in parallel we investigate novel configurations, through a simulation strategy, involving the use of fermentation tanks-in-series and ISPR \(\rightarrow V\). To reduce the number of reactors operated under vacuum as both capital and operating costs of vacuum fermentation are high, novel reactor configurations were proposed. Given a series of fermentation vessels, the first reactor was demonstrated should be operated at atmospheric pressure as the butanol concentration in this reactor is low (<5 g l\(^{-1}\)), i.e. there is a threshold concentration of 4-4.8 g butanol l\(^{-1}\) below which no apparent decrease in growth arose [9]. The following reactors are then operated under vacuum to minimise butanol inhibition. Product inhibition is reduced in vacuum reactors as these in situ recovery reactors are optimise to maintain the butanol concentration in the reactor at low level (around 6 g l\(^{-1}\)).

The last reactor is a conventional reactor to increase the butanol concentration in the dilute stream from the reactor, i.e. not operated under vacuum, as high levels of butanol in the diluted stream are required to decrease the energy needs of the final purification system [13,24]. Also, to increase energy efficiency, two scenarios, without and with cogeneration of heat and power, were studied. The start-up and shutdown strategies of the reaction system are also demonstrated to be a significant factor that affects process performance when considering butanol production in a
continuous process. Start-up and shutdown are considered as the maximum operating times in actual continuous fermentative processes are limited to between 200 and 500 h because of contamination and strain stability issues [25]. This paper demonstrates that the economic potential may be increased by 16.3 % (200 h) and 6.7 % (500 h) if the reactor start-up and shutdown procedures are also optimised.

2.0 Material and methods

The focus of this work is the optimisation of the configuration of the fermenters assuming continuous operation as well as the evaluation of the energy requirements of the recovery and purification system. A general schematic of the process is shown in Fig. 1. Several fermenter configurations were optimised, both in series and parallel (Fig. 2). A start-up/shutdown strategy for those reactors was considered as it was anticipated that the relatively low maximum operating time (<500 h) will have a high impact on the yield and productivity and consequently on the economics of the process.

Fig. 1. Reaction system and downstream system studied in this paper. In situ product recovery by vacuum evaporation using tank-in-series and double-effect distillation. CS is a stripping column. C1HP and C1LP are high-pressure and low-pressure columns used to recover ABE from the vinasses, respectively. C2 is the column for ethanol and acetone purification. C3 is the column for butanol purification. D is decanter. Q is the heat of condenser-boiler. The energy values are that obtained for the best configuration studied in this paper (5AVA)

The reactor configuration has three effluent streams (1) a diluted stream (~1 wt% ABE) (2) a concentrated or condensed stream (~40 wt% butanol) (3) a gas stream (mainly CO2 and H2), see Fig. 1 and 2. In the concentrated stream, butanol composition is increased from ~8 to ~40 wt% as a result of
partial condensation (Fig. 2). The stream is then fed to a decanter where butanol concentration is further increased from ~40 wt% to ~68 wt%.

As the effect of the reactor configuration used on the energy requirements of the downstream process is also studied, once the optimisation of the reactor configuration has been performed, the total energy requirements of distillation are estimated using Aspen Plus® V10. A scheme with four columns and double-effect distillation is used to separate and purify acetone, butanol, and ethanol as this system has low energy requirements and low total annualised costs [26]. Indeed, the final purification process (4 distillation columns and 1 stripping column) has been reported as the most economical process of heat-integrated distillation for ABE purification by [26], and it has similar total energy requirements to that reported for processes involving ISPR (between 7.2 and 13.5 MJ/kg butanol).

Fig. 2. Continuous production of ABE by ISPR — V system. a) parallel operating mode b) tank-in-series. Where W is compressor work. HX is heat exchanger. CW, cold water. T is temperature. S0 and L0 are the average compositions of the substrate (cellulose plus xylose and glucose) and lignocellulose, respectively. S is the total substrate (cellulose plus xylose and glucose), and C is cellulose. F represents the mass flow. w is the mass fraction. The subscripts L and G represent liquid and gas flow, A is acetone, B is butanol, E is ethanol, W is water. The values for the energy requirements and compositions are the optimal conditions found in this work.
The minimum temperature approach in the simulation was 10°C. The pressure drop of the columns is estimated rigorously assuming using tray sieve. The Murphree plate efficiency was assumed as 0.7. Heat-loss to the environment was not considered, as we assumed that these losses are estimated assuming an efficiency to produce steam and electricity from fuel.

In the double-effect distillation scheme, the diluted stream is split into two and fed to two distillation columns one operating at low pressure (C1-LP column, 0.27 bar) and the other at 1.1 bar (C1-HP column), Fig. 1. C1-LP and C1-HP columns produce vinasses, which are used to preheat the dilute streams from the reactor. In C1-LP, the concentrated ABE at the top is condensed and then fed to a decanter. The organic phase of this decanter is then fed to C1-HP, which is used to recover acetone and ethanol. An intermediate heat exchanger in C1-HP operates as a reboiler of the vacuum columns C1-LP and the ethanol and acetone purification column (C2). Butanol is purified at 1.2 bar in C3 column (Fig. 1), and steam from the top of the column is fed to C1-HP to increase the energy integration. Finally, the gas streams are fed to a stripping column achieving a total ABE recovery of around ~98%, ethanol being the most challenging component to recover in the whole process due to its low relative volatility when compared to butanol and acetone. A decanter in C1-HP is then used to separate the butanol water azeotrope. As a result, a purity of butanol, acetone, and ethanol of 99.7, 99 and 75 wt% are obtained. As the ethanol-water azeotrope still needs to be separated, an additional step, such as extractive distillation or pervaporation, is required to purify this stream.

2.1 A kinetic model for fermentation and hydrolysis

As native and genetically engineered microorganisms that produce butanol can consume both hexoses and pentoses, low-cost lignocellulosic biomass that does not compete with food production offer economically attractive raw materials [10]. Therefore, in this work, the use of hydrolysates of detoxified corn stover was investigated as substrate as this is an inexpensive and abundant non-food substrate for ABE production [27,28].
Reactor optimisation of butanol production from hydrolysates of detoxified corn stover requires rigorous kinetic models describing fermentation and hydrolysis. In ABE fermentation, several kinetic models have been proposed using several microorganisms [29–34]. For example, a kinetic model proposed by Eom et al. (2015) was developed to study ex-situ butanol recovery by adsorption [31]. However, only glucose was used as the substrate, and only butanol and ethanol were considered as sub-products. In another example, a metabolic kinetic model describing ABE fermentation by \textit{C. saccharoperbutylacetonicum} N1-4 (ATCC 27021) was proposed by [32]. In this model, xylose consumption was not studied, and this microorganism has been reported with low productivity from glucose $< 0.1 \text{ g L}^{-1} \text{ h}^{-1}$. Although most of the metabolic models are based on the equations developed for glucose and xylose consumption by [29,30], in these models, carbon catabolite repression is not considered. In this sense, to model mixed substrate fermentation, metabolic and unstructured kinetic models were recently developed [35–37].

In order to optimise butanol production, a kinetic model is needed to describe ABE fermentation from lignocellulose that predicts the individual consumption of xylose and glucose, their co-fermentation as well as the co-production of sub-products (i.e., acetic acid, butyric acid, acetone or ethanol). The prediction of butyric and acetic acid productions are essential as both organic acids are intermediates involved in ABE production, and if they are removed, the ABE yield may be decreased [38].

In this work, a kinetic mode that was calibrated using experimental data obtained from both continuous and batch operation using \textit{Clostridium saccharoperbutylacetonicum} N1-4 (ATCC13564) was employed [35]. This microorganism can consume several sugars with solvent productivity in batch fermentation of around $0.5 \text{ g L}^{-1} \text{ h}^{-1}$ and with an overall ABE yield around $0.39 \text{ g g}^{-1}$ [10]. The model can predict butanol production using different operating strategies such as batch, continuous, biomass recirculation, and sole fermentation of xylose or glucose using high concentrations (up to 90 g L$^{-1}$) and their co-fermentation, i.e., it describes carbon catabolite repression [35]. The rates of gases...
production (H2 and CO2) were estimated using stoichiometric conversions of glucose and xylose in the synthesis of ABE [23].

In hydrolysates of detoxified corn stover, xylan is mostly converted; however, around 10% of cellulose is converted to glucose [23]. Due to this low conversion, a hydrolysis step still is required as microorganisms for ABE production are not able to consume cellulose. Here, the kinetic model for cellulose conversion was proposed by Kadam et al., 2004 [39]. This kinetic model includes glucose, cellobiose and xylose inhibitions, and it has been used in several research studies involving the production of second-generation biofuels [40,41].

2.2 Process model

In our previous work [17,23], a rigorous mathematical model was developed for reactors operating in parallel. To simulate ISPR − V using fermentation tank-in-series, the mathematical models of the reactors of our previous work were modified. Both the tank-in-series and parallel reactor configurations are shown in Fig. 2.

A constant temperature was assumed, and the reactor material balance used are as follows,

\[
\frac{dN_{m,i}}{dt} = \left( R_{m,i} \cdot V_{F(m)} + R_{w(m,i)} \cdot V_{T(m)} \right) - F_{F,m} \cdot x_{A(m,1)} - F_{P,m} \cdot x_{P(m,1)} - F_G m
\]

where \( t \) (h) is time, \( N \) (kg) and \( x \) (kg kg\(^{-1}\)) are the amount and composition of component \( i \) in reactor \( m \) and \( F_P \) is the feed flow to each reactor (kg h\(^{-1}\)), respectively. The pre-treated substrate is only fed to the first reactor, and the effluent of reactor \( m - 1 \) is used as the feed substrate to reactor \( m \), if \( m \) is greater than 1. \( R \) and \( R_w \) are the reaction rates (kg m\(^{-3}\) h\(^{-1}\)) as a result of fermentation and saccharification, respectively. \( V_F \) is the volume of the liquid phase in the reactor (m\(^3\)) and \( V_T \) is the volume of total fermentation broth (both solid and liquid phase). \( F_P \) and \( F_G \) are the outlet liquid and gas stream, respectively.

In addition to the differential equations describing the rate of change of the mass of each component,
The volume and density of the fermentation broth are assumed to change with time and are calculated assuming additive volumes.

Three phases are considered (solid, liquid and gaseous), where the reactions occur only in the liquid phase.

The liquid and gas outlet flow rates are obtained, assuming that proportional controllers regulate volume and pressure, respectively.

The saccharification model includes a dynamic term of cellulose reactivity as this term depends on cellulose conversion, which changes with time.

Gas and water reaction rates are predicted, assuming stoichiometric ratios.

In vacuum fermentation, a heat-pump system is required to reduce the energy requirements. The compressors of the heat-pump system are simulated, taking into consideration the vacuum pump efficiency proposed by [37]. Although the process studied here was continuous, as fermentation time was limited and scheduling of the reactors and its effect on the capital costs of the overall process were also studied.

In the heat-pump system, the compressors and heat-exchangers are configured in series with partial condensation of solvents (Fig. 2). Only the heat of condensation of the first stage is used to supply heat to the fermenter, while, the other stages are condensed using cooling water. The number of compression stages is three, the optimum value [23]. The temperature of reactors was assumed to be 30 °C. The heat-exchanger used to provide heat to the reactor was operated using a minimum approach temperature of 10 °C. The final pressure at the exit of the compressor series was specified as 1 atm. The volume of reactors are assumed to be equal and fixed to 1000 MM gal [42]. In the heat-pump system, a pressure drop of 0.03 bar was assumed. Most of the energy requirements of the reactor are supplied by the heat-pump system. However, energy requirements are still needed for evaporation heating. These energy requirements are of low quality as the reactor is operated a low temperature (30 °C). Sources for the energy needs required for the evaporation can be supplied by
the feed stream of the pre-treated corn stover or using the remaining vapour at a low pressure of the
distillation system. For this reason, it was assumed that the reactor does not require additional energy
requirements in the optimisation procedure. This assumption was proved once the optimisation
process is finished. As can be seen in Fig. 1 and Fig. 2, the condensation energy of the columns of low
pressure is enough to supply the additional energy requirements of the reactor for the SAVA process.

2.3. Multi-objective optimisation strategy

A parallel reactor configuration, as well as a series of two to six reactors, are economically optimised.
The number of reactors operated in either parallel or series is determined in order to achieve an ABE
production target of 40 MM kg year\(^{-1}\). The MATLAB 2019b\(^{®}\) global optimisation method that performs
multi-objective optimisation, gamultiobj (a genetic algorithm), was used (Fig. 3). The default
conditions of this function were employed. The interval for the optimal variables search studied in this
work were two times the optimal values found in our previous work multiplied by two and divided by
two [23]. The optimisations were repeated at least three times to achieve a solution close to the
optimal global value.

Two optimisation objectives \(\text{Obj}_1\) and \(\text{Obj}_2\), the maximisation of the economic potential \(\text{EP}\) and
the minimisation of the total compression work \(\text{WT}_{\text{ABE}}\), were considered,

\[
\text{Obj}_1 = \max (\text{EP}), \text{where } \text{EP} = \sum (C_{\text{ABE}} \cdot F_{\text{ABE}}) \cdot t_a - T_0 AC - T_0 C / t_r
\]

\[
\text{Obj}_2 = \min (\text{WT}_{\text{ABE}}), \text{where } \text{WT}_{\text{ABE}} = \frac{1}{F_{\text{ABE}}} \sum_{i=1}^{3} W_i
\]

where, \(F_{\text{ABE}}\) is the total flowrate of the products (kg/h), i.e. the acetone, butanol and ethanol.
Fig. 3. Optimization method used in the selection and economic and energy evaluation of the proposed configurations. 

The sales price (\(c_{ABE}\)), the total annualised operating costs (\(TOAC\)) and total investment costs (\(TIC\)) are calculated as was proposed by [23]. The compression work was estimated using the equation proposed by [43] and the isentropic efficiency was calculated as was suggested by [44]. The costs of electricity, cold water, the substrate (corn stover) and enzymes were included in the \(TOAC\) (USD per year). The investment costs of compressors, heat-exchangers and reactors are included in the \(TIC\) (USD). To estimate the \(EP\), a return time on the investment \((t_r)\) of three years was assumed. The
annualised time of operation \( t_a \) was estimated concerning the final fermentation time \( t_f \) and considering a cleaning/turn-around time of 4 h and a maintenance time of 1 day every 20 days. \( W_i \) is the compressor work of each reactor. The optimisation variables studied in this work for a given continuous processing time using the tank-in-series configuration, and \( ISPR - V \) are reported in the Table 1.

**Table 1**

<table>
<thead>
<tr>
<th>Optimisation variables and constraints for the tank-in-series reactors with ( ISPR - V )</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Optimisation variables:</strong></td>
</tr>
<tr>
<td>1) the feed flow rate of the first reactor</td>
</tr>
<tr>
<td>2) the number of parallel reactors operated in tanks-in-series</td>
</tr>
<tr>
<td>3) the set-point of vacuum pressure in the reactors</td>
</tr>
<tr>
<td>4) the exit vacuum pressures from stage one and two of compression</td>
</tr>
<tr>
<td>5) the enzymes load and composition (the ratio of enzyme load (g protein mg g(^{-1}) cellulose) being equal in the input feed stream and initial concentration in the reactor</td>
</tr>
<tr>
<td>6) the initial substrate concentrations in both the reactor and the feed-stream (the reactors had the same initial substrate concentration)</td>
</tr>
<tr>
<td><strong>Optimisation variables for the start-up:</strong></td>
</tr>
<tr>
<td>7) the initial volume of each reactor</td>
</tr>
<tr>
<td><strong>Optimisation variables for the shutdown:</strong></td>
</tr>
<tr>
<td>8) the time when the shutdown started or the time when continuous operation finished</td>
</tr>
<tr>
<td>9) the outlet flow rate of the reactors during the shutdown period</td>
</tr>
<tr>
<td>10) the reactors which volume control loop is open (control off).</td>
</tr>
<tr>
<td><strong>Constraints:</strong></td>
</tr>
<tr>
<td>1) the initial volume of the fermentation broth was restricted to be lower than 80% of the total volume</td>
</tr>
<tr>
<td>2) the ode15s function of MATLAB 2019b* was used in the simulation as the system of the ordinary differential equation is stiff. However, ode15s shows convergence issues at high cellulose conversion (&gt;~90%). These issues were mitigated when the minimum volume allowed in the reactor was limited to 20% of the total reactor volume.</td>
</tr>
<tr>
<td>3) the concentrations of non-soluble solids in the reactor was restricted to be lower than 100 g L(^{-1}).</td>
</tr>
</tbody>
</table>

The reactor start-up and shutdown strategies are proposed in order to maximise substrate conversion and minimise substrate losses. The start-up strategy consisted of using a lower reaction volume than that of the volume set-point (the volume set-point was 80% of the total reactor volume for the continuous period for all reactors). Thus, the reactor first starts in fed-batch mode, which minimises the losses of the substrate in the effluent stream of the last reactor. The continuous process only starts once the last reactor in the sequence is filled. For reactor shutdown, the continuous feed to the first reactor in the sequence is stopped, and each reactor is allowed to drain at a constant rate to provide the substrate to the other reactors. In the simulation, this constant rate of their outlet flow streams is found by optimisation as the volume control is open (control off).
2.4 Energy efficiency

The energy efficiency of the recovery, purification and reaction system ($EE_{RR}$) proposed in this work was determined in order to determine whether the energy requirements found in this paper are significant. $EE_{RR}$ was modified from the equation proposed by [26] for glucose fermentation to include a cellulosic substrate as follows,

$$EE_{RR} = \frac{Y_{ABE} \cdot (LHV_P - FR_{RR})}{LHV_{ST}} \quad (5)$$

where $Y_{ABE}$ is the yield of ABE (g g$^{-1}$) and $LHV_{ST}$ and $LHV_{PT}$ are the lower heating value of the substrate and product, respectively, and are estimated as follows,

$$LHV_{ST} = \sum x_s \cdot LHV_\Phi, \forall \Phi \in \{cellulose, glucose, xylose\} \quad (6)$$

$$LHV_{PT} = \frac{1}{Y_{ABE}} \sum Y_\Phi \cdot LHV_\Phi, \forall \Phi \in \{butanol, ethanol, acetone, hydrogen\} \quad (7)$$

where $x_s \cdot $ is the mass fraction of each substrates $\Phi$ in all the substrates, i.e., cellulose plus glucose and xylose. $Y_\Phi$ is the yield of the product $\Phi$. The products considered here where butanol, hydrogen, ethanol and acetone as these products could be used as biofuels. The $LHV_\Phi$ for butanol, acetone, ethanol and hydrogen were 34.4, 30.27 and 121.5 MJ/kg, respectively. The $LHV_\Phi$ for cellulose, glucose, and xylose is 17.4, 15.7 and 15.7 MJ/kg, respectively. As $x_s \cdot $ for glucose, xylose and cellulose were equal to 0.07, 0.40, and 0.52 in this work, respectively, $LHV_{ST}$ was equal to 16.6 MJ/kg.

In previous work, the fuel requirements of the recovery system ($FR_{RR}$) were estimated for a separate power and heat system, i.e. without cogeneration [26]. $FR_{RR}$ is estimated assuming that steam and electricity are the only energy needs that requires fuel for their production. This process without cogeneration was assumed to have efficiencies for electricity and steam production of 33% and 90% from fuel, respectively [26]. Hence, $FF_{RR}$ is the sum of the energy needs of steam and electricity divided by their respective efficiency of their production from fuel. In addition, in this work, a case scenario with energy cogeneration was considered. The total energy efficiency for this combined heat and power scenario is assumed to be 86%, where 30% of this is for electricity, and 56% is low-pressure steam production [45]. The requirements of fuel for steam production are
reduced with cogeneration as low-pressure steam is produced when electricity is made from fuel. For this reason, two scenarios must be considered, when the steam produced in the electricity manufacture is higher or equal than that needed by the recovery system and when additional steam is required to be produced from fuel. Taking these figures and scenarios into consideration, $FR_{RR}$ with and without cogeneration is proposed to be estimated as follows,

$$FF_{RR} = \begin{cases} \frac{WT_{ABE}}{0.3} \frac{Q_D}{0.9}, & \text{without cogeneration} \\ \frac{WT_{ABE}}{0.3} Q_a, & \text{for cogeneration} \end{cases}$$

$$Q_a = \begin{cases} \frac{1}{0.9} \left( Q_D - \frac{0.56 \cdot WT_{ABE}}{0.3} \right), & \text{if } \frac{1}{0.9} \left( Q_D - \frac{0.56 \cdot WT_{ABE}}{0.3} \right) > 0 \\ 0, & \text{otherwise} \end{cases}$$

where $Q_D$ is the energy requirements of the distillation system (MJ/kg-ABE), and the heat, $Q_a$, is the additional steam required for the recovery system which is not supplied by the residual vapour produced in the electricity production. The term $0.56 \cdot WT_{ABE}/0.3$ is the steam produced in the creation of electricity.

### 3.0 Results

#### 3.1. Start-up and shutdown of the fermenter tanks-in-series using ISPR $- V$

As this is the maximum operating time that has been recommended via experimental studies, the reaction time of the fermentation assumed in this work was 500 h. The simulated process consisted of five reactors in series, with three reactors operated under vacuum, while, the first and the last reactor are operated at atmospheric pressure (the notation 5 AVA was used to indicate a series of five reactors, the first being atmospheric (A), the next three operated under vacuum (V) and the last reactor being operated at atmospheric (A) conditions).

To estimate the minimum reactor operating time required to achieve the maximum economic performance for continuous operation, the evaluation of the 5 AVA configuration was performed at several reactor operating times using the optimal conditions found at 500 h. As a result of optimisation, the conversion of xylose and cellulose were around 0.95-0.99 and 0.82-0.87,
respectively (see Fig. 4 a). As would be expected, the EP increased with the operating time as a higher conversion of the substrate in the reactor is achieved (see Fig. 4 b). Interestingly, when the final operating time is higher than 500 h, a higher xylene conversion tends to be achieved with a slight reduction in cellulose conversion as the steady-state condition of the reactors is not achieved at 500 h (Fig. 5), which may be caused for the reaction rates are slow, and the control strategy, i.e., proportional controllers were used for volume and pressure control and a control strategy for concentrations was not considered.

Including the optimisation of start-up and shut down of the reactor in this work increased the EP between 16.3 % (200 h) and 6.7 % (500 h). Only when the final operating time was greater than 8000 h (approximately one year), the effect of start-up and shutdown on the EP becomes insignificant (Fig. 4 b). When the strategy of start-up and shutdown was included in the optimisation, the minimum operational time required to achieve an EP close to the maximum (a deviation of ~1.5%) was 500 h. The volume profiles of 5 AVA with an operating time of 500 h is shown in Fig. 5. It may be observed that the total time required to fill the last reactor (R5) was 47 h, i.e., the purge of R5 started at 47 h. In this sense, the occupied volume of each reactor in the start-up period was lower than in continuous operation. This strategy was employed to reduce the losses of the substrate in the last reactor (Fig. 5). During the shutdown period, the continuous feeding to the first reactor (R1) was stopped at 461 h. As
substrate feeding was stopped before the final reaction time (500 h), the shutdown strategy
maximised the cellulose conversion. Only the volume of reactors one (R1) and two (R2) in the
sequence were lower than that of set-point during shutdown period (Fig. 5), i.e. reactors R1 and R2
are used to feed reactors three (R3), four (R4) and five (R5) which are still operating as continuous
reactors.

Fig. 5 Volume profiles for the optimal strategy of operation developed in this work. 1R, 2R, 3R, 4R, and 5R are the position of the reactor in the tanks-in-series system. Only the reactors 2R, 3R, and 4R were under vacuum
The concentration of cellulose in the feed-stream was 57 g L$^{-1}$. The concentration of butanol in
reactor R1 increased during the shutdown as a vacuum is not applied, and substrate feeding was
stopped in this reactor (Fig. 6 b). While the butanol concentration in R5 is not increased during
shutdown as the substrate concentration feed to R5 was lower than 25 g L$^{-1}$ (Fig. 6 a). The average
concentration of butanol in the purge from the last reactor (R5) was ~8.5 g L$^{-1}$. The vacuum
fermenters (R2, R3, and R4) have a similar concentration of butanol in the reactor (~5 g L$^{-1}$) due to
ISPR.
Fig. 6. Concentration profiles for the optimal strategy of operation developed in this work. 1R, 2R, 3R, 4R, and 5R are the position of the reactor in the tanks-in-series system. Only the reactors 2R, 3R, and 4R were under vacuum. The grey zones are used to indicate the periods of start-up/shutdown strategies, i.e., when the volume control is open (control off). The continuous operation period is between the dashed lines.

3.2. Optimisation of proposed configurations

In this work, eight configurations were optimised, considering the start-up/shut down strategies (500 h). The configurations were one parallel production (1V) and seven tanks-in-series systems: 2V (2 reactors at vacuum), 3V (3 reactors at vacuum), 4V (4 reactors at vacuum), 5V (5 reactors at vacuum), 4AVA (2 reactors at vacuum and 2 at atmospheric pressure), 5AVA and 6AVA (4 reactors at vacuum and two at atmospheric pressure). In the AVA configurations, the first and last reactors are without in situ product recovery, i.e., conventional reactors at atmospheric pressure.

Reactors operated at atmospheric pressure are proposed in order to reduce the energy requirements of the process.

In this work, the electricity costs for beer transfer pump and compression in the distillation were not included in the optimisation. The energy requirements for beer transfer pump from 0.05 bar to 1 atm are 0.05 GJ/h, while the compression work of the distillation system was 0.04 GJ/h. These values are around 400 times lower than that of compressor 1, W1 (19.6 GJ/h for 5AVA), indicating that these terms are not necessary to be included in the optimisation procedure. Also, the cost of pumping liquids is lower than that of gasses. For instance, a beer transfer pump reported by [42] has a cost of 58165 USD (125 Hp), which is ~600 times lower than that of the heat-pump cost found in this paper.
The 3V scheme reduced the compressor work requirements by 31% when compared with parallel production (1V), and the amount of corn stover required reduced by 4%, i.e., the ABE yield was increased (Table 2). However, productivity using 3V was reduced as the total number of reactors increased from 7 to 9 to achieve the required production target (Table 2). In other words, it is more important to increase the ABE yield and reduce the compressor work than to increase productivity.

### Table 2

<table>
<thead>
<tr>
<th>Configuration</th>
<th>Number of parallel tanks-in-series</th>
<th>Total number of reactors</th>
<th>Compressor work (MJ kg⁻¹ butanol)</th>
<th>Energy requirements of distillation (MJ kg⁻¹ butanol)</th>
<th>Total Energy requirements (MJ kg⁻¹ butanol)</th>
<th>Corn stover capacity (ton dry LCB day⁻¹)</th>
<th>Yield (L ABE ton⁻¹ dry LCB)</th>
<th>Enzyme load (mg protein g⁻¹ cellulose)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 V</td>
<td>7</td>
<td>7</td>
<td>4.03</td>
<td>4.09</td>
<td>8.11</td>
<td>1983</td>
<td>255</td>
<td>8.1</td>
</tr>
<tr>
<td>2 V</td>
<td>5</td>
<td>10</td>
<td>3.23</td>
<td>4.51</td>
<td>7.74</td>
<td>1893</td>
<td>267</td>
<td>11.5</td>
</tr>
<tr>
<td>3 V</td>
<td>3</td>
<td>9</td>
<td>2.77</td>
<td>4.69</td>
<td>7.46</td>
<td>1904</td>
<td>266</td>
<td>13.1</td>
</tr>
<tr>
<td>4 V</td>
<td>2</td>
<td>8</td>
<td>2.73</td>
<td>4.62</td>
<td>7.35</td>
<td>1897</td>
<td>267</td>
<td>14.3</td>
</tr>
<tr>
<td>4 AVA</td>
<td>2</td>
<td>8</td>
<td>2.68</td>
<td>4.58</td>
<td>7.25</td>
<td>1925</td>
<td>263</td>
<td>11.8</td>
</tr>
<tr>
<td>5 V</td>
<td>2</td>
<td>10</td>
<td>2.51</td>
<td>4.72</td>
<td>7.23</td>
<td>1873</td>
<td>270</td>
<td>12.4</td>
</tr>
<tr>
<td>5 AVA</td>
<td>2</td>
<td>10</td>
<td>2.23</td>
<td>4.73</td>
<td>7.00</td>
<td>1880</td>
<td>269</td>
<td>10.8</td>
</tr>
<tr>
<td>6 AVA</td>
<td>2</td>
<td>12</td>
<td>2.27</td>
<td>4.83</td>
<td>7.11</td>
<td>1894</td>
<td>267</td>
<td>7.6</td>
</tr>
</tbody>
</table>

AVA (the first and the last reactor are operated under atmospheric pressure)
V (all reactors are operated under vacuum pressure)

Higher selectivity of butanol was achieved with a lower number of reactors in the series. For instance, the selectivity of butanol is 0.769 kg butanol kg⁻¹ ABE for 5 AVA and 0.829 kg butanol kg⁻¹ ABE for 1V. For this reason, when tanks-in-series systems were studied, the energy requirements of distillation divided by the number of kilograms of ABE produced were reduced by 5%, which is lower than that in terms of kilograms of butanol which was a reduction of 12% (Table 2). It is essential to mention that the variation in the energy requirements of double-effect distillation divided by kilograms of ABE for all the optimisations was low (<5%) and that series operating mode achieves in all scenarios evaluated lower total energy requirements (Table 2).

The process with the lowest total operating costs was 6AVA, while, the process with the lowest TIC was 4AVA (Table 3). AVA options were found to be more profitable than the exclusive use of vacuum fermenters as the operational costs (utilities and enzymes) are reduced between 9.2 and 9.5% in comparison with systems in which vacuum is applied in all the reactors (Table 3). The 6AVA
configuration required the highest number of reactors (12, Table 2), while the 6AVA and 5AVA achieved the highest economic potential. However, 5AVA would appear to be the optimal configuration as the lowest energy requirements were achieved.

<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>1V</td>
<td>30.5</td>
<td>70.8</td>
<td>33.8</td>
<td>43.6</td>
<td>15.2</td>
<td>7.9</td>
<td>32.0</td>
</tr>
<tr>
<td>2V</td>
<td>43.5</td>
<td>55.1</td>
<td>32.9</td>
<td>41.7</td>
<td>10.9</td>
<td>9.2</td>
<td>36.9</td>
</tr>
<tr>
<td>3V</td>
<td>39.2</td>
<td>47.9</td>
<td>29.0</td>
<td>41.9</td>
<td>9.2</td>
<td>10.6</td>
<td>40.0</td>
</tr>
<tr>
<td>4V</td>
<td>34.8</td>
<td>47.5</td>
<td>27.4</td>
<td>41.8</td>
<td>9.0</td>
<td>11.4</td>
<td>40.8</td>
</tr>
<tr>
<td>4AVA</td>
<td>34.8</td>
<td>46.2</td>
<td>27.0</td>
<td>42.4</td>
<td>8.9</td>
<td>9.6</td>
<td>42.6</td>
</tr>
<tr>
<td>5V</td>
<td>43.5</td>
<td>42.9</td>
<td>28.8</td>
<td>41.2</td>
<td>8.1</td>
<td>9.8</td>
<td>41.7</td>
</tr>
<tr>
<td>5AVA</td>
<td><strong>43.5</strong></td>
<td><strong>41.3</strong></td>
<td><strong>28.3</strong></td>
<td><strong>41.4</strong></td>
<td><strong>7.6</strong></td>
<td><strong>8.6</strong></td>
<td><strong>44.1</strong></td>
</tr>
<tr>
<td>6AVA</td>
<td>52.2</td>
<td>40.2</td>
<td>30.8</td>
<td>41.7</td>
<td>7.3</td>
<td>6.1</td>
<td>44.1</td>
</tr>
</tbody>
</table>

IC is the investment costs; TIAC is the total investment annualised costs divided by the return time on the investment (3 years).

The total operating annualised costs (substrate, utilities, and enzymes) and total investment annualised costs (TIAC) were 64-69% and 31-39% of the total annualised costs (TAC), respectively. The operating costs associated with the enzymes and utilities are similar for each configuration studied with each representing between 7 and 12.7% of the TAC. Substrate cost was the most critical item in the calculation of the EP and represented between 44 and 49% of the TAC. For this reason, when series production is studied, all the reactors maximise the yield (between 263 and 270 l/ton-dry-LCB, Table 2). However, as expected because of the low cost of corn stover, this impact is lower than that of conventional substrates such as corn which is around 66% of the production costs of the whole process [46].

For a given configuration, high energy requirements are required to achieve high economic potentials (Fig. 7). For instance, high yields were achieved using 1V configuration due to the high...
compressor work (4.03 MJ/kg-ABE) required to increase the yield and the productivity of ABE.

Nevertheless, the multi-objective optimisation of all the considered configurations demonstrates that the use of tanks-in-series increases the economic potential and reduces energy requirements (see Fig. 7). Furthermore, the minimisation of the energy needs and the maximisation of the EP tends also to maximise the ABE yield and productivity. Whereas, the enzyme load at optimal conditions is not necessarily increased with compressor work. 1R was the configuration that achieved the lowest enzyme load and consequently, the lowest ABE yield. In turn, SAVA was the configuration with the second-lowest enzyme load and the highest yield, the maximum economic potential, and the minimum compression work (see Fig. 7).

![Fig. 7. Pareto fronts obtained as a result of the optimisation of both compressor work and economic potential](image)

3.3 Energy efficiency

The fuel requirements for 1V reported by [23] for a continuous process with reactors operated on parallel were 13.3 MJ fuel kg\(^{-1}\) ABE (the efficiency in electricity and steam production were 0.33 and 0.9, respectively). Fuel requirements are a more convenient way to report the energy requirements as electricity and steam require different amounts of fuel for its production and acetone and ethanol are important sub-products with a high selling price (between 70% and 80% of
the selling price of butanol [13]). In this work, the energy requirements in fuel equivalents of 1V were found to be 13.8 MJ-fuel/kg-ABE. Although the energy requirements for 1V were similar than that reported by [23] (13.8 instead of 13.3 MJ fuel kg\(^{-1}\) ABE), a 21% higher EP was achieved in this work as the start-up/shutdown strategy was optimised. Additionally, the configuration proposed in this work with the highest EP (5 AVA) reduced the energy and fuel requirements reported by [23] by 14% and 33%, respectively.

The energy efficiency with a separate power and heat system was between 0.51 and 0.67 being at a minimum using the parallel system (1 V) and a maximum using 5 AVA, see Table 4. Although this energy efficiency is lower than that reported for ABE fermentation from glucose using conventional reactors and heat-integrated distillation (between 0.71 and 0.76)[26], here the raw material was a lignocellulosic substrate which it is known to have a more difficult conversion than that of glucose. To increase energy efficiency, a case scenario with cogeneration was also studied in this paper. The combined heat and power system increased the energy efficiency from 0.51-0.67 to 0.57-0.74.

Table 4. The energy efficiency of the reaction and recovery systems studied in this paper

<table>
<thead>
<tr>
<th>Configuration</th>
<th>(LHV_{PT}) (MJ fuel/kg ABE)</th>
<th>Energy of recovery (F_{RR}) (MJ fuel/kg ABE)</th>
<th>Energy efficiency (EE_{RR})</th>
<th>Energy of recovery (F_{RR}) (MJ fuel/kg ABE)</th>
<th>Energy efficiency (EE_{RR})</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 V</td>
<td>36.1</td>
<td>13.8</td>
<td>0.51</td>
<td>11.1</td>
<td>0.57</td>
</tr>
<tr>
<td>2 V</td>
<td>36.6</td>
<td>11.9</td>
<td>0.59</td>
<td>8.7</td>
<td>0.67</td>
</tr>
<tr>
<td>3 V</td>
<td>36.9</td>
<td>10.7</td>
<td>0.63</td>
<td>7.3</td>
<td>0.71</td>
</tr>
<tr>
<td>4 V</td>
<td>37.0</td>
<td>10.4</td>
<td>0.64</td>
<td>7.1</td>
<td>0.72</td>
</tr>
<tr>
<td>4 AVA</td>
<td>37.3</td>
<td>10.0</td>
<td>0.64</td>
<td>6.8</td>
<td>0.72</td>
</tr>
<tr>
<td>5 V</td>
<td>37.1</td>
<td>9.8</td>
<td>0.66</td>
<td>6.5</td>
<td>0.74</td>
</tr>
<tr>
<td>5 AVA</td>
<td>37.0</td>
<td>9.2</td>
<td>0.67</td>
<td>6.2</td>
<td>0.74</td>
</tr>
</tbody>
</table>

4.0 Discussion

Several vacuum fermenters, as well as other recovery techniques, have been proposed in the literature for butanol production. For vacuum fermentation processes, Mariano et al. 2011 and 2012 [19,21] reported energy requirements using a heat-pump system between 1.9 and 4.7 higher (13 and 32 MJ kg\(^{-1}\) butanol) than that of 5 AVA. While, Pereira et al., 2017[18] reported energy requirements
without a heat-pump (17.6 and 25.3 MJ kg⁻¹ 2-butanol) between 2.6 and 3.7 times higher than that achieved in this work by 5 AVA. The low energy requirements achieved in this work were due to 1) the use of fractional condensation, 2) consideration of AVA rectors, 3) optimisation of the heat-pump system, 4) optimisation of the start-up/shutdown of the reactors, and 5) employing an efficient distillation process, namely double effect-distillation.

In comparison, the most economical operating mode previously reported was a fed-batch with an economic potential of 37.8 MMUSD, and energy requirements of 10.5 MJ fuel kg⁻¹ ABE [23]. Hence, 5AVA increases the economic potential and reduces the fuel requirements of fed-batch for 16.6% and 12.4%, respectively. Batch reactors achieve a lower EP than fed-batch reactors; however, it achieves the lowest energy requirements of 9.6 MJ fuel kg⁻¹ ABE [23]. Nevertheless, these energy requirements were 4.4% higher than that of 5AVA. Lower energy requirements were achieved as the compression of gasses was minimised. This reduction was performed using tank-in-series operating with vacuum and atmospheric reactors, allowing a reduction of the gasses fed to the heat-pump in 70-80%.

In the literature, it has been recently suggested that the lowest energy requirements may be achieved using heat-integrated distillation and hybrid separation strategies such as extractive distillation and distillation (or stripping) and membrane technologies [26,47,48]. The energy requirements of these recovery methods are reported to be between 4 and 6 MJ fuel/kg ABE.

However, although the lowest energy requirements reported in the literature are found using these intensified techniques, ISPR, also improves reactor performance. For this reason, ISPR still is the more economical way of production [17]. Also, although the energy requirements were higher in this work, as cogeneration was proposed, the energy efficiency of 3 V, 4 V, 4 AVA, 5 V and 5 AVA were as high as that using the heat-integrated distillation schemes proposed by [26].

In comparison with other ISPR processes, the lowest energy requirements reported (8.2 MJ kg⁻¹ of butanol) by Qureshi et al. 2005 for adsorption were 1.2 times higher than that of 5 AVA. In comparison, the energy requirements using a combination of gas-stripping and pervaporation (22 MJ
butanol [49]) were 3.2-fold higher than that of 5 AVA. The energy requirements of vacuum evaporation in this work were low due to heat-integration. Although in adsorption, pervaporation and gas stripping heat-integration can be studied to reduce the energy requirements, this integration may be more complicated than that of vacuum evaporation. For instance, heat-integration may be more demanding as in pervaporation lower pressures of vacuum is needed. At the same time, it may be more challenging in gas stripping and adsorption due to the high amount of gases recycled and the high temperature of desorption required to achieve high recovery of product.

While several recovery strategies have been reported using liquid-liquid extraction, including heat-integration [13]. Direct steam extraction (DSD) was found by [13] to have lower energy requirements than that of high-temperature extraction, dual extraction, and mixture extractants with a conventional regeneration column operated at vacuum. In this system, biocompatible extractants that usually have a high boiling point (~350 °C) are used, and steam is fed directly to the extractant regeneration column, recovering ABE. In this way, the regeneration column may be operated at atmospheric pressure using low temperatures of regeneration (<160 °C), which reduces the investment costs as a lower heat exchanger area is required which increases the heat-integration capacity as an atmospheric column can be used instead of a vacuum column. The fuel requirements of DSD were found to be between 5.7 and 6.4 MJ-fuel/kg ABE. Those fuel requirements are around 1.5 times lower than those achieved in this work. However, the energy requirements are dependent of the process assumptions, i.e., the substrate concentration, substrate conversion, minimum approach temperature, butanol recovery in the reactor, efficiencies of the recovery and trays, ABE ratio, among others. For instance, the energy requirements using DSD decreases 1.2-fold if the substrate concentration rises from 200 at 300 g L⁻¹. Such a high concentration of substrate is feasible as glucose was used instead of lignocellulose. In this work, the energy requirements were found using a substrate concentration two times lower than that reported for DSD due to the constraint of water-insoluble solid concentrations (<100 g L⁻¹) in the reactor.
5.0 Conclusions

The process configuration developed and discussed in this work (5 AVA or 6 AVA) achieved the lowest energy requirements and the highest economic potential reported for vacuum fermentation. Due to the high investment cost of the compressors as well as the electricity costs, the energy requirements of the process were reduced in preference to productivity. The optimal process was found to be the use of fermentation tanks-in-series with five or six reactors (5AVA or 6AVA). The first and last reactor did not require vacuum recovery. The optimisation of the start-up and shutdown of the reactor was found to increase the economic potential of 5AVA by 3 MM/USD (assuming an operating time of 500 h). Finally, when cogeneration is considered, the energy requirements of the reaction and recovery system in fuel equivalents reduced from 9.2 to 6.6 MJ fuel/kg ABE and the energy efficiency increased from 67 to 74%. Pilot-scale experiments using corn stover are strongly recommended in future work in order to gain further understanding of the operation and economic potential of the proposed process configuration.

Funding

This research was funded thanks to COLCIENCIAS (‘Programa de estancias postdoctorales para beneficiarios de formación Colciencias en entidades del SNCTel’).

Credit authors statement


References


