Butanol production via vacuum fermentation: an economic evaluation of operating strategies

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Abstract

Butanol production from corn stover via (acetone-butanol-ethanol) ABE fermentation under vacuum was studied in this work. The reactor operating strategies considered were batch, fed-batch and continuous. The integrated reactor and vacuum separation process includes energy integration by a heat-pump system as well. A mathematical model describing the dynamics of the integrated reactors and the network of compressors and heat exchangers has been developed. The dynamic process models were used to select the optimum production strategy using economic optimisation where a methodology to determine the effect of scheduling of parallel reactor operation on the sizing of the heat-pump system was developed. The results suggest that the optimal operating mode for the integrated reaction system was fed-batch. Although the fed-batch process had the highest economic potential (37.8 MM USD), the compressor work for batch process operation (36.6 MM USD) was the lowest (1.8 MJ/kg ABE, 18% lower than that of fed-batch). Continuous process operation demonstrated the lowest economic potential (23.8 MM USD) and the highest compression work (2.87 MJ/kg ABE). The energy requirements of the purification system, a double-effect distillation process, were found to be between 3.45 and 4.14 MJ/kg ABE.

Keywords: *Butanol production; heat pump; dynamic modelling; scheduling; economic optimisation; batch, fed-batch, continuous operation*

27 **Nomenclature**

- 28 $CAPEX$, capital expenditures (USD/year)
- 29 CCR, carbon catabolite repression (-)
- 30 Cp_k , mean calorific heat capacity between the stage k and $k+1$ (MJ/kg/K)
- 31 EP , economic potential (USD/year)
- 32 F_0 , flow of feed (mmol/h)
- 33 F_p , flow of bleed or purge stream (mmol/h)
- 34 F_G , flow of gas stream (mmol/h)
- 35 K_P , proportional gain for pressure control (mmol/bar/h)
- 36 K_V , proportional gain for volume control (mmol/l/h)
- 37 H , parameter of Henry's solubility (bar)
- 38 HX , heat exchanger
- 39 I_C , investment cost of compressors (USD)
- 40 I_{HX} , installation cost of heat exchangers (USD)
- 41 I_R , installation cost of a reactor (USD)
- 42 ISPR, *in situ* recovery reactor
- 43 $LMTD$, logarithmic mean temperature difference (K)
- 44 MW , molecular weight (g/mol)
- 45 N , amount of a component in the reactor (mmol)
- 46 N_R , number of reactors
- 47 N_V , number of reactors operated under vacuum with respect to time
- 48 $OPEX$, Operating expense (USD/year)
- 49 p , parameter describing death of biomass
- 50 P , pressure (bar)
- 51 P_{set} , pressure set-point (bar)
- 52 P_V , vapour pressure (bar)

- 53 Q_{CV} , heat of condensation of the heat exchangers
- 54 r_i , rate of production / consumption of component i in the fermentation (mmol/h)
- 55 r_s , parameter relating glucose and substrate consumption
- 56 r_{wi} , saccharification reaction rate of component, i (g/h/kg)
- 57 R , global reaction rate of component i in the fermentation (mmol/h/l)
- 58 R_C , reactivity parameter for cellulose conversion (g/g)
- 59 R_{cte} , universal gas constant (J/mol/K)
- 60 R_W , global reaction rate for hydrolysis (g/kg/h)
- 61 S_M , ratio between the average and the maximum compressor work on the cycle time (-)
- 62 t , time (h)
- 63 t_d , time between the start-up of the reactors (h)
- 64 t_f , total fermentation time (h)
- 65 t_t , cycle time (h)
- 66 T , temperature (K)
- 67 T_{CR} , critical temperature (K)
- 68 V_F , total volume of liquid components in the reactor (l)
- 69 V_{set} , volume set-point (l)
- 70 V_T , total volume of broth in the reactor (l)
- 71 x , liquid composition in the reactor (mol/mol)
- 72 w , mass fraction
- 73 W , compressor work (MJ/h)
- 74 y , gas composition (mol/mol)
- 75 z , compressibility factor
- 76 Greek symbols
- 77 ω , weighting factor for consumption of glucose and xylose
- 78 ρ , density (g/l)

79	α , coefficient of activity (liquid phase)
80	γ , coefficient of activity (vapour phase)
81	η , isentropic efficiency
82	κ_r , specific heat ratio ($C_p/(C_p - R_{cte})$)
83	Subscripts
84	i , component in the biochemical network
85	j , component of the gas stream
86	k , compression stage
87	S , substrate
88	L , liquid from heat exchangers
89	Superscript
90	M , maximum
91	S , scheduling
92	A , average
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1.0 Introduction

ABE (acetone-butanol-ethanol) are alternative biofuels. Unfortunately, their production by fermentation is currently limited by butanol inhibition and low substrate yield (Raganati et al., 2016; Zheng et al., 2015). Furthermore, while lignocellulose (LCB) substrates of low cost, such as corn stover, represent attractive options when compared to conventional substrates, LCB consumption by the microorganism is known to be difficult, because of the high degree of biomass polymerisation (Ghosh et al., 2017). Therefore, in order to maximise ABE yield and reactor productivity from corn stover, pre-treatment and enzymatic hydrolysis are required. Unfortunately, the inhibition of soluble sugars during enzymatic hydrolysis (Kadam et al., 2004) reduces the conversion rate of cellulose to glucose and, in addition, carbon catabolite repression (CCR) minimises the utilization of secondary carbon sources (e.g. xylose) during fermentation (Ren et al., 2010; Servinsky et al., 2010).

Simultaneous saccharification, fermentation, and recovery offers an interesting means to intensify the process (Grisales Díaz and Olivar Tost, 2016a). As butanol inhibition is reduced by the separation units, *in situ* product recovery (ISPR) increase ABE productivity and yield when compared to conventional processing methods (Qureshi et al., 2014). The *in situ* recovery methods studied the most for bio-butanol production are gas stripping, vacuum evaporation, liquid-liquid extraction, adsorption, and pervaporation (Kujawska et al., 2015). It is important to mention that although these processes improve reactor performance (i.e., the butanol productivity may be increased more than two times (Sharif Rohani et al., 2015)), heat-integrated distillation schemes have been recently reported with lower energy requirements than that of integrated recovery and reaction methods (Grisales Diaz and Olivar Tost, 2018; Xue et al., 2013).

Pervaporation or adsorption have been shown to have a low energy requirement due to the high selectivity of some of the membranes (e.g. the butanol separation factor of poly[1-(trimethylsilyl)-1-propyne] is 104 (Claes et al., 2012)) or adsorbents (e.g. butanol is

concentrated from 0.5 to 98% by adsorption on silicalite (Milestone and Bibby, 1981)). However, the high fouling in these processes hinder their industrial use for butanol production from lignocellulose (Dubreuil et al., 2013; Xue et al., 2016). Similarly, liquid-liquid extraction have low energy requirements (8.4 MJ/kg butanol (Qureshi et al., 2005)). However, emulsion formation and control must be further studied in this multi-phase (liquid-liquid, solid and gas) process. Although it has been reported that gas stripping has a high energy requirement (23.1 MJ/kg butanol (Cai et al., 2016)), gas stripping still is frequently studied because the gases from the fermentation (CO_2 and H_2) may be use to recover the volatile components and due to the low risk of fouling (Oudshoorn et al., 2009).

Butanol selectivities in gas stripping are similar to the relative volatility of butanol under vacuum (<30, (Grisales Díaz and Olivar Tost, 2016a; Xue et al., 2014)). However, vacuum evaporation has been reported with low energy requirements 13.4 MJ/kg (Mariano et al., 2011) as the evaporation heat may be recovered using heat pump systems. Application of a heat-pump system in gas stripping is more difficult due to the elevated flow gas required. Although vacuum stripping is possibly one of the best economic and technical options, it is in an early stage of its development (Tao et al., 2014). For these reasons, bio-butanol production by vacuum evaporation is considered in this work.

The determination of the best mode of operation (batch, fed-batch or continuous) can be challenging because each mode has advantages in terms of either yield, productivity, enzyme load or energy requirements (Tao et al., 2014); and there is a significant number of operating variables that can be adjusted in order to optimise performance. In our previous work, the energy requirements and total annualised costs of continuous vacuum fermentations were determined. However, this was based upon steady-state assumptions (continuous fermenters operating for of 500 h), i.e. the implications and requirements of batch and fed-batch operation were not studied. In addition, a kinetic model that did not consider

carbon catabolite repression and gave inaccurate butanol predictions at high concentrations was used (Grisales Díaz and Willis, 2018).

The aim of this work is to select the operating strategy for ABE production from corn stover using vacuum fermentation by maximisation of the economic potential (*EP*). The *EP* was considered in order to address the conflicting criteria of productivity, yield, enzyme load, and process energy requirements. As batch and fed-batch operation are considered, scheduling of the operation of the ancillary units within the process flowsheet is also required in order to determine the operating regime that maximises the *EP*.

2.0. The process

In this work, the focus is ISPR by vacuum evaporation (ISPR-V). A set of parallel reactors are operated such that saccharification, fermentation, and recovery are achieved simultaneously. The proposed flowsheet is shown in Fig. 1. In addition to the reactor(s) a network of compressors and heat exchangers are required as the reactors operate under vacuum conditions.

A number of different configurations of this compression arrangement have been considered in the literature (Abdi et al., 2016; Mariano et al., 2011; Pereira et al., 2017) and because of the high energy requirements of the process, the energy efficiency of the vacuum fermentation needs be improved using heat pump systems. In the process flowsheet, Fig. 1, the energy obtained through condensation of the gas stream is used to provide the energy for evaporation within the reactor(s). In previous work, the use of a heat pump has been proposed where excess condensation energy is used to supply energy to the boiler of the acetone purification column. Total energy requirements of 13.4 and 22.3 MJ/kg butanol for continuous operation (Mariano et al., 2011) and between 22 and 32.4 MJ/kg butanol for batch (Mariano et al., 2012a) have been reported, demonstrating that both the final purification system and the configuration of the compression system are affected by the operational mode. Our previous work reported the use of an integrated heat pump to provide the energy of evaporation within

the reactor, the excess energy was not used in the purification system, which used a highly efficient distillation process instead (Grisales Díaz and Olivar Tost, 2016a), i.e. a double-effect distillation (DED) process. This reduced the total energy requirements for continuous operation to 5.9 MJ/kg ABE or 9.1 MJ/kg butanol (or 10.8 MJ fuel/kg ABE, calculated assuming an efficiency of production from fuel of electricity and steam of 0.33 and 0.9, respectively).

For batch and fed-batch operation, the power and the cooling water required by the heat pump system will change with respect to time because each fermentation starts at atmospheric pressure and then there is a change to vacuum (0.045-0.07 bar) operation. Altering the mode of operation, between atmospheric and vacuum has the potential to reduce the process energy requirements as the time-period and operating pressure of the system can be adjusted. Moreover, as it is assumed that there are multiple reactors operating in parallel, the optimal scheduling of these reactors will determine their optimal start-up times and periods of vacuum operation in order to minimise the investment costs of compressors as well as operating costs. Assuming batch and fed-batch operation therefore implies that the initial investment (and sizing) of the heat pump system will depend on the maximum work and condensation heat required at any given time as a result of the optimally scheduled operation. This is a complex problem as a high number of reactors will be required, because ABE fermentation has a low productivity (between 0.11 and 0.68 g/l/h (Yao et al., 2017)). For this reason, to find the maximum required compressor work, a short-cut scheduling method is proposed. To the best of our knowledge, the scheduling of dynamic ISPR by vacuum evaporation (ISPR-V) for butanol production has not been studied in the literature.

3.0. Process modelling

In order to study and optimise the performance of the reaction system a dynamic process model is required. The model needs **a)** an accurate biochemical reaction network; **b)** a kinetic description of all the reactions occurring within the system; **c)** dynamic equations (material balance expressions) describing the rate of change of the species in the reactor; and **d)** models

describing the operation of the ancillary equipment (compressors and heat exchangers) within the flowsheet (Fig. 1).

3.1 The biochemical reaction network

The proposed biochemical reaction network for simultaneous saccharification and ABE fermentation is shown in Fig. 2. The rate of production / consumption of each component in ABE fermentation and saccharification is represented by R_i (mmol/l/h) and R_{Wi} (g/kg/h), respectively. For saccharification the mass rate of production / consumption of each of the species are given by,

$$R_{W\text{ Cellubiose}} = 1.056 \cdot r_{w1} - r_{w3} \quad (1)$$

$$R_{W\text{ Cellulose}} = -r_{w1} - r_{w2} \quad (2)$$

$$R_{W\text{ Glucose}} = 1.111 \cdot r_{w2} + 1.053 \cdot r_{w3} \quad (3)$$

where the constants are calculated based in the stoichiometric reactions assumed by Kadam et al., (2004).

The global rates of reaction of CO_2 and H_2 were calculated assuming stoichiometric conversions of glucose and xylose in relation to the main products: acetone, ethanol and butanol.

Hydrogen is only produced in the stoichiometric reaction with acetone (see Fig. 2), where the stoichiometric coefficient of H_2 was four times that of the acetone, i.e. the global rate of H_2 was four times that of acetone. While, three, two and one moles of CO_2 are made for each mole of acetone, butanol and ethanol produced (Fig. 2), respectively. It was assumed that six moles of xylose has the same yield as five moles of glucose (Humbird et al., 2011) (see Fig. 2). Hence, the rate of production of H_2 and CO_2 depends on the rates of consumption of glucose and xylose,

$$R_{\text{H}_2} = 4 \cdot (r_4 + r_{11}) \cdot (r_s + 5/6 \cdot (1 - r_s)) \quad (4)$$

$$R_{\text{CO}_2} = (3 \cdot (r_4 + r_{11}) + 2 \cdot r_{14} + r_7) \cdot (r_s + 5/6 \cdot (1 - r_s)) \quad (5)$$

$$r_s = R_{\text{Glucose}} / (R_{\text{Glucose}} + R_{\text{Xylose}}) \quad (6)$$

In similar way, the mass rate of production / consumption of water is calculated in relation to the stoichiometric reactions assumed in Fig. 2,

$$R_{Water} = (-r_4 - r_{11} + r_{14}) \cdot (r_s + 5/6 \cdot (1 - r_s)) \cdot 0.018 - 0.056 \cdot r_{w1} - 0.111 \cdot r_{w2} - 0.053 \cdot r_{w3} \quad (7)$$

In addition, CCR was estimated assuming competitive carbohydrates uptake (Grisales Díaz and Willis, 2018). Therefore, the molar rate of production / consumption of each of the species are given by,

$$R_{Biomass} = r_{16} \cdot (1 - \omega) + r_8 \cdot \omega - R_{Biomass\ death} \quad (8)$$

$$R_{Biomass\ death} = (r_{17} \cdot (1 - \omega) + r_9 \cdot \omega) \cdot p \quad (9)$$

$$R_{Glucose} = -r_1 \cdot \omega \quad (10)$$

$$R_{Xylose} = -r_{15} \cdot (1 - \omega) \quad (11)$$

$$R_{Pyruvate} = -r_2 + r_1 \cdot \omega + r_{15} \cdot (1 - \omega) \quad (12)$$

$$R_{AcoA} = r_2 + r_3 + r_4 - r_5 - r_6 - r_7 \quad (13)$$

$$R_{AACoA} = r_6 - r_{10} - r_4 - r_{11} \quad (14)$$

$$R_{Acetate} = r_5 - r_4 - r_3 \quad (15)$$

$$R_{Ethanol} = r_7 \quad (16)$$

$$R_{Acetone} = r_4 + r_{11} \quad (17)$$

$$R_{Butyrate} = r_{13} - r_{12} - r_{11} \quad (18)$$

$$R_{BCoA} = r_{10} + r_{11} + r_{12} - r_{13} - r_{14} - r_{16} \cdot (1 - \omega) - r_8 \cdot \omega \quad (19)$$

$$R_{Butanol} = r_{14} \quad (20)$$

The death rate of biomass is zero for continuous operation, i.e. p is zero for continuous operation and one for fed-batch or batch. The kinetic model for ABE fermentation by *Clostridium saccharoperbutylacetonicum* N1-4 was developed in our previous work (Grisales Díaz and Willis, 2018) (Table A1). The parameters of these models are reported in the Table A2. The kinetic model gives good predictive performance in batch fermentations using both xylose and glucose as individual substrates, at high concentrations of substrate and biomass, in mixture of substrates, and in continuous operation with and without biomass recycling and

bleeding. The kinetic model describing saccharification was proposed by Kadam et al., (2004) (Table A1) and it has been widely used in a number of design and process optimisation studies (Morales-Rodriguez et al., 2011; Quiroz-Ramírez et al., 2018). The parameters of this model are reported in Table A3.

3.2. Modelling the reactor

In order to mathematically model the reactor, isothermal operation is assumed (30 °C). To describe dynamic operation of each reactor, including start-up for continuous operation, the material balances for each of the components are required. These are given by a set of ordinary differential equations (ODEs) describing the rate of change of the moles of each component ($N_{(i)}$) with respect to time (t),

$$\frac{dN_{(i)}}{dt} = R_{(i)} \cdot V_F + R_{w(i)} \cdot V_T \cdot \rho / MW_{(i)} + F_A \cdot x_{A(i)} - F_P \cdot x_{0(i)} - F_{G(0)} \cdot y_{0(i)} \quad (21)$$

Eq. (21) assumes no mass transfer resistances between the solid-liquid and liquid-gas phase. These assumptions are valid when the concentrations of non-soluble solids in the hydrolysis are lower than 20% (Hodge et al., 2008). Therefore, to minimise the effect of the mass transfer resistances in the fermentation and hydrolysis, the concentration of non-soluble solids in the reactor were specified as being lower than 100 g/l. In Eq. (21), $F_{G(0)}$ is the molar flow of gas (mmol/h). For batch operation of ISPR-V, there is no feed, i.e. the dilution rate $F_A = 0$, while, for continuous and fed-batch operation there is a feed-stream. The difference between continuous operation and fed-batch are defined in relation to the bleed stream. If the reactor has no bleed, $F_P = 0$, the reactor is fed-batch, while, continuous operation implies the presence of a bleed stream. In addition, x_0 are the mole fractions of each component in the reactor ($N_i / \sum N_i$) and x_A and y_0 are the molar compositions in the feed and the gas phase, respectively. ρ is the average density of the media (g/l) and $MW_{(i)}$ (g/mol) is the molecular weight of each component. The total volume of the fermentation broth, V_T (l) (including solids

and liquid), and liquid, V_F (l), are calculated using the densities and the moles ($N_{(i)}$ (mmol)) assuming additive volumes, i.e. an ideal mixture.

The set of differential equations is completed by,

$$\frac{dN_{(C)}^X}{dt} = x_{0(C)} \cdot F_P \quad (22)$$

$$\frac{dN_{(C)}^F}{dt} = x_{A(C)} \cdot F_A \quad (23)$$

$$R_{C(t)} = \frac{dN_{(C)}^X}{dN_{(C)}^F} \quad (24)$$

where, $x_{A(C)}$ and $x_{0(C)}$ are the molar compositions of cellulose (C) in the feed stream and reactor or bleeding, respectively, and $N_{(C)}^F$ and $N_{(C)}^X$ are the total amounts of cellulose fed and unconverted, respectively. Eq. (22) and (23) allow the calculation of, R_C , the cellulose reactivity (Eq. (24)). These differential equations are required because the substrate concentrations in the feed stream and the initial substrate concentration of the reactor may be different.

In order to determine the gas flow ($F_{G(t)}$) and the flow of the bleeding stream ($F_{P(t)}$) it is assumed that they are manipulated using a proportional control law (Grisales Díaz and Olivar Tost, 2016a),

$$F_{G(t)} = K_P \cdot (P_{(t)} - P_{set}) \quad (25)$$

$$F_{P(t)} = K_V \cdot (V_{T(t)} - V_{set}) \quad (26)$$

V_{set} and P_{set} are the volume (l) and pressure (bar) set-points and K_V (mmol/l/h) and K_P (mmol/bar/h) are the controller gains. The numerical value of the controller gains was chosen via trial and error to ensure a stable (closed loop) response.

The compositions of volatile components in the gas stream (y_i) within the reactor are calculated using the modified Raoult's law,

$$y_{(k,i)} \cdot P_{(k)} = x_{(k,i)} \cdot \alpha_{(k,i)} \cdot P_{V(k,i)} / \gamma_{(k,i)} \quad (27)$$

The coefficients of activity of the volatile components in the liquid phase (α) and the gas stream (γ) were calculated using UNIQUAC and Peng-Robinson's equation state, respectively.

Pressure (P) and the gas compositions (y) were estimated through the bubble pressure algorithm (Henley et al., 2011). Henry's law was used to estimate the concentration in the liquid phase of non-condensable gases. Hence, H_2 and CO_2 compositions were estimated as,

$$y_{(k,i)} \cdot P_{(k)} = x_{(k,i)} \cdot H_{(k,i)} \quad (28)$$

Henry's law constants for the estimation of solubility ($H_{(k,i)}$) in water of H_2 and CO_2 are reported by Sander, (2015).

3.3. Modelling of the heat exchanger and compressor units

Although the vacuum pressure and gas flow and composition changes slightly with respect to time, it was assumed that the vacuum pressure and the species flow rates into the heat-pump system were at their respective average values. This assumption allows the simulation of only one reactor in the scheduled process.

The number of compression stages used in this work was three (Fig. 1). The global material balance around the heat exchanger of stage k ($k = 1, \dots, 3$) (HX , in Fig. 1) were modelled in the same way as a conventional flash. Therefore,

$$F_{G(k)} = F_{L(k)} + F_{G(k+1)} \quad (29)$$

$$F_{G(k)} \cdot y_{G(k,i)} = F_{L(k)} \cdot x_{L(k,i)} + F_{G(k+1)} \cdot y_{G(k+1,i)} \quad (30)$$

In Eq. (29) and Eq. (30), $F_{G(k)}$ and $F_{L(k)}$ are the average flows of liquid and gas, respectively (note that all the reactors are defined as stage '0', i.e. $k = 0$). The gas ($y_{G(k,i)}$) and liquid ($x_{L(k,i)}$) compositions of components (i) in the outputs of $HX_{(k)}$ were found by simultaneously solving the equilibrium equations (Eq. (27) - Eq. (28)) and mass balance equations (Eq. (29) - Eq. (30)) through the flash algorithm proposed by Henley et al., (2011).

For ISPR-V it is assumed that there are two operating zones which are used to reduce the operational costs, one atmospheric period (no energy requirements) and a vacuum period. Vacuum fermentations for butanol production have been studied experimentally in several works (Mariano et al., 2012b; Qureshi et al., 2014) and the strategy adopted in this work is consistent with that reported experimentally where a fermentation is usually initially operated

at atmospheric conditions and then if the butanol concentration is high enough ($>5 \text{ g l}^{-1}$) vacuum operation starts. The change between atmospheric and vacuum operation is determined through optimisation of the pressure controller set-point. Eq. (27) - Eq. (30) are used for both zones of operation. The maximum pressure controller set-point is 1.1 atm. If the set-point is lower than the maximum (1.1 atm), vacuum operation is assumed and equations (Eq. (31) - Eq. (34)) are used to calculate the average compressor work and heat of condensation. In the heat-pump system, only the first stage is used to supply energy due to evaporation in the fermenter and the minimum approach temperature of its boiler-condenser was assumed to be 10°C (or T_C was 40°C). The T_C of condensers that use cold water (CW, $18\text{--}25^\circ\text{C}$) was specified as being 30°C . This allows the maximum energy requirements of cooling, $Q_{c(k)}$, of each of the heat exchangers to be calculated as,

$$Q_{c(k)} = - \sum F_{L(k)} \cdot \Delta H_{c(k,i)} \cdot x_{L(k,i)} - F_{G(k)} \cdot \int_{T_{c(k)}}^{T(k)} \sum_i C_{p_i} \cdot x_{G(k,i)} \cdot dT \quad (31)$$

$$\Delta H_{c(k,i)} = A_i \cdot (1 - T_{c(k)}/T_{CR(i)})^{n_i} \quad (32)$$

The parameters (A_i and n_i) of the volatile components are reported by Yaws and Satyro, (2009); if the component (i) are gases (H_2 or CO_2), A_i was assumed to be zero. $T_{CR(i)}$ is the critical temperature of components (i). C_{p_i} is the gaseous calorific heat capacity of the component i .

For compressor ($W_{(k)}$ (MJ/h)), the compression work was calculated using the mean properties of the gas between stage $k - 1$ and k using (Fontalvo et al., 2005):

$$W_{(k)} \cdot \eta_s = \frac{\kappa_r}{\kappa_r - 1} \cdot z \cdot R_{cte} \cdot T_{(k-1)} \cdot \left[\left(\frac{P_{(k)}}{P_{(k-1)}} \right)^{\frac{\kappa_r - 1}{\kappa_r}} - 1 \right] \quad (33)$$

where, z is the compressibility factor, R_{cte} is the constant of ideal gases ($\sim 0.083 \text{ MJ/kmol/K}$), κ_r is the specific heat ratio ($C_p/(C_p - R_{cte})$) and the isentropic efficiency (η_s) of the compressor was assumed to be 0.75 (Vane et al., 2013). However, to take into consideration the different isentropic efficiencies of vacuum pumps reported in the literature (between 0.5

and 0.75 (Sreemahadevan et al., 2018)), once the best operating mode is selected, optimisation using different η_s were performed. In addition, it has been reported that η_s for vacuum systems may be calculated as a function of the input and output pressures (Belaissaoui et al., 2016; Matsumiya et al., 2005). This is also studied in the section 4.3 using different numbers of compression stages. To the best of our knowledge the effect that these two variables have on both economic and energy performance has been not studied in the literature.

$T_{(k-1)}$ (K) and $P_{(k-1)}$ (bar) are the temperature and the mean pressure at the compressor inlet. The temperature (T) at the exit of the compressor is calculated via,

$$T_{(k)} = T_{(k-1)} + \frac{T_{(k-1)}}{\eta_s} \left[\left(\frac{P_{(k)}}{P_{(k-1)}} \right)^{\frac{\kappa_r - 1}{\kappa_r}} - 1 \right] \quad (34)$$

3.4. Scheduling of the reactors

The annual operating time (t_a) is given by,

$$t_a = 8760 \cdot (1 - 4/t_t - 1/20) \quad (35)$$

In this equation, t_t is defined as the sum of fermentation time (t_f) and the cleaning / turn-around time (which is assumed to be 4 h). An additional time for the maintenance of each reactor (independent of t_t and reactor operation mode) of 1 day every 20 days has been assumed. This allows the annual electricity consumption of the compressors and the annual cold water requirement for the heat-exchangers to be obtained by multiplying t_a by $W_{(k)}^A$ or $Q_{c(k)}^A$. $W_{(k)}^A$ and $Q_{c(k)}^A$ are the average compressor work and heat of condensation taking account of both periods of operation, i.e. atmospheric and vacuum respectively,

$$W_{(k)}^A = \frac{t_v}{t_t} \cdot W_{(k)} \quad (36)$$

$$Q_{c(k)}^A = \frac{t_v}{t_t} \cdot Q_{c(k)} \quad (37)$$

where, t_v is the vacuum time.

The compression work and heat of condensation calculated from Eq. (31) and Eq. (33) are obtained assuming that all the reactors are simultaneously operated under vacuum. Allowing the reactors to be scheduled to start at different times reduces the maximum compression work ($W_{(k)}^S$) and heat of condensation ($Q_{c(k)}^S$) required in the cycle time. This, in turn, gives a reduction in the maximum size of the units and the investment costs.

The maximum compression work ($W_{(k)}^S$) and heat of condensation ($Q_{c(k)}^S$) of the scheduled process may be calculated as,

$$W_{(k)}^S = W_{(k)} \cdot N_V^M / N_R \quad (38)$$

$$Q_{c(k)}^S = Q_{c(k)} \cdot N_V^M / N_R \quad (39)$$

where ($N_V^M \leq N_R$) is the maximum number of (the parallel) reactors that are operating under vacuum at any particular time. The value of N_V^M may be obtained using a Gantt chart assuming that all the campaigns are operated in the same way, see Fig. 3. In this example, the total number of reactors is $N_R = 12$, t_t was 80 h and vacuum was applied after 28 h, i.e. t_v was 52 h (65% of the total fermentation time).

N_V^M is a function of the time between the start-up of reactors (t_d). If we specify a normalised time between the start-up of each reactor ($t_d / t_t = 0.05$ h/h, it may be noted that for $N_V^M = 12$, there is no reduction in the amount of reactors simultaneously operating under vacuum (Fig. 3 Case A). However, when (t_d / t_t) is specified as 0.1 h/h, then $N_V^M = 9$ (Fig. 3 Case B). Therefore the adjustment (and optimisation of t_d) will give a lower N_V^M and hence minimise $W_{(k)}^S$ and $Q_{c(k)}^S$. Similarly, N_V^M is related to the vacuum time (t_v), for example, if t_v is reduced to 0.55 h/h using t_d equal to 0.1 h/h, N_V^M is increased from 9 to 11 (Case C, Fig. 3).

It is convenient to define the ratio between the average and the maximum compressor work or heat of condensation over the cycle time ($W_{(k)}^S / W_{(k)}^A$ or $Q_{c(k)}^S / Q_{c(k)}^A$, respectively) as,

$$S_{M(k)} = \frac{W_{(k)}^S}{W_{(k)}^A} = \frac{W_{(k)} \cdot \frac{N_V^M}{N_R}}{W_{(k)} \cdot \frac{t_v}{t_t}} = \frac{t_t}{t_v} \cdot \frac{N_V^M}{N_R} \quad (40)$$

This allows the Initial installation cost of compressors ($I_{C(k)}$ (USD)) and the installation cost of each of the heat exchangers ($I_{HX(k)}$ (USD)) (Kiran et al., 2012; Perkins, 2007) to be calculated in terms of $W_{(k)}^A$ and $Q_{c(k)}^A$,

$$I_{C(k)} = 4534 \cdot (W_{(k)}^A / 2)^{0.82} \cdot 2 = 4534 \cdot (W_{(k)}^A \cdot S_{M(k)} / 2)^{0.82} \cdot 2 \quad (41)$$

$$I_{HX(k)} = 21300 \cdot \left(\frac{Q_{c(k)}^S}{U \cdot LMTD} \right)^{0.65} = 21300 \cdot \left(\frac{Q_{c(k)}^A \cdot S_{M(k)}}{U \cdot LMTD} \right)^{0.65} \quad (42)$$

where, U (1 MJ/m²/K) is the overall heat transfer coefficient and $LMTD$ (10 K) is the logarithmic mean temperature difference. As U and $LMTD$ were assumed constant a more rigorous estimation of these parameters must be performed in future work. To reduce the negative effect on the EP in case of an eventual compressor failure during scheduled operation, it is assumed that two compressors (of half the size of that required) are used instead of only one compressor of full size (Eq. (41)).

3.5. Cost estimation and process optimisation

The aim is to maximise the EP (USD/year) of the integrated reactor, where EP is defined as,

$$EP = \sum F_{ABEi} \cdot ABE_i \cdot t_a - OPEX - CAPEX \quad (43)$$

where, F_{ABEi} is the flow (kg/h) of the solvents (acetone, butanol and ethanol) produced in the fermentation, ABE_i is the selling price of products (acetone, butanol and ethanol, Table 1).

The additional terms are the operating expense ($OPEX$) and the capital expenditures ($CAPEX$). Note that, during ABE fermentation from lignocellulose, butyric acid, acetic acid, H₂ and solids (residual lignocellulose and biomass) are produced as by-products. However, only ABE is considered as value added product.

The $CAPEX$ (USD/year) is calculated as,

$$CAPEX = \left(\sum I_{HX(k)} + \sum I_{C(k)} + I_R \cdot N_R \right) / t_{ri} \quad (44)$$

Here I_R is the installation cost of the reactors with N_R being the number of reactors and t_{ri} is the return time on the investment (3 years). A correlation that may be used to estimate the reactor installation cost (I_R) as a function of the volume was reported by Oudshoorn et al., (2010) (Eq. (B-1) in Table 1). Correlations used to estimate $I_{HX(k)}$ and $I_{C(k)}$ were adapted from (Kiran et al., 2012; Perkins, 2007) (Eq. (41) and Eq. (42)).

The *OPEX* (USD/year) is calculated as,

$$OPEX = C_{CS} \cdot F_0 \cdot w_{0(CS)} + C_{EZ} \cdot R_{E/C} \cdot F_0 \cdot w_{0(C)} + \sum C_{El} \cdot W_{(k)}^{oper} \cdot t_a + \sum C_{CW} \cdot Q_{c(k)}^{oper} \cdot t_a \quad (45)$$

In Eq. (45), the first term represents the operating costs associated with the use of corn stover, $w_{0(CS)}$ is the average load of corn stover (CS) on a dry basis in the feed, F_0 is the flow of corn stover on a mass basis (kg/h) and C_{CS} is the cost of corn stover (USD/kg-dry-LCB). The second term represents the cost of enzymes required for the hydrolysis of cellulose, hence, $R_{E/C}$ is the ratio of enzyme to cellulose (g/g). The third and fourth terms in Eq. (45) represent the operating costs associated with the compressors (power, W , kWh) and the heat exchangers (cold water, $Q_{c(k)}$, MJ/h), respectively. The numerical values of the various cost parameters are reported in the Table 1.

In order to maximise the *EP* of the process, the genetic algorithm function 'ga' in Matlab® was used. The variables that are adjusted during optimisation are the start-up time of the feed and the vacuum, the total time of reaction, initial concentration of cellulose in the reactor and in the continuous feed, the initial volume of fermentation, the enzyme load in the reactor and feeding stream, the vacuum pressure set-point, the pressure of the exit of compressors and the dilution rate. As the size of the heat exchanger and the compressor power depend on the pressure at the exit of compressors and operational times, these units are also optimised. Two constraints were imposed, **a)** non-soluble solids in the reactor must be lower than 100 g/l, and **b)** the concentration of corn stover in the continuous feed must be lower than 30 wt% (composition of corn stover in the pretreatment (Humbird et al., 2011)).

3.6. Implementation aspects

The annual production of ABE from corn stover was assumed to be 140,000 tonne/y. This gives a corn stover capacity around 2,000 tonne/day (Humbird et al., 2011). The initial concentration of corn stover usually used for ethanol fermentation is shown in Table 2 (Humbird et al., 2011). These concentrations are used to define the ratio of compositions in reference to cellulose concentration in the feed, on a dry-basis. While, the initial concentration of cellulose in the reactor and the feed are variables to be determined via optimisation.

ABE productivity is the conventional metric used for the scale-up of the fermentation process. ABE productivity (kg ABE/l/h) is the rate of production of ABE divided by the reactor volume (in this case, the total volume of the reactor broth was assumed to be at a maximum of 80% of the reactor volume). When ABE productivity is used to scale-up the reactor, the ratios of flow streams and reactor volume must be kept constant. As the volume of reactor does not affect the reaction rate (mass transfer resistances are not studied in this work), the productivity is constant with respect to the volume. Hence, the ABE productivity, P_{ABE} (kg/h/m³), was used as a basis for the scale-up of the reactor and this was defined as,

$$P_{ABE} = \frac{\sum 1000 \cdot N_{(ABE)}^{t=0} / MW(ABE) + \int_{t=0}^{t=t_f} \sum (F_P \cdot x_{(0,ABE)} + F_{G(0)} \cdot y_{(0,ABE)}) \cdot dt}{\max(V_T/0.8) \cdot t_f} \quad (46)$$

where, MW is the molecular weight (g/mol) of acetone, butanol and ethanol.

Given that all the reactors are assumed to be operated in the same way over the cycle time, only one reactor needs to be simulated reducing to some degree the computational effort.

Therefore, the results of the simulation of a single reactor are used to calculate the ABE productivity. The total volume of the reactors can then be found through P_{ABE} , the annual time of operation (t_a) and the annual required production of ABE (140,000 tonne/y). As the reactors all have the same size and the maximum volume of each reactor was 3785 m³ (Humbird et al., 2011), the number of reactors (N_R) that would be required to achieve the annual production target are found dividing the total volume of reactor(s) by 3785 m³. If N_R is

a decimal number, it is approximated to the next integer number and the volume of the reactor(s) is found dividing the total volume of reactor(s) by N_R . In this way, the volume of each reactor is lower than or equal to 3785 m³. Having calculated the optimal reactor volume (and the total number of reactors required), the average work of the compressors and heat of exchangers are calculated. This allows the *OPEX* and *CAPEX* to be determined as well as the *EP* given the particular set of decision variables. The optimisation algorithm then adjusts the values of the decision variables and the simulations are repeated iteratively until the maximum *EP* is achieved.

4.0 Results

4.1. Sizing of heat-pump system

In this work, a short-cut method is used to calculate $S_{M(k)}$ which is developed through the analysis of several operating strategies using a range of the number of reactors and vacuum times (an automated generation of the Gantt charts). This repeated analysis of the effect of t_d on the value of $S_{M(k)}$ is shown in Fig. 4 (a) which shows the minimum value of $S_{M(k)}$ (optimal t_d) for a range of values of parallel reactors (N_R) and vacuum times (t_v). It may be observed that very high number of reactors (>100) is necessary to totally minimise the variation in the value of $S_{M(k)}$. Furthermore, $S_{M(k)}$ may converge towards the minimum value of one (Fig. 4 (a)) for particular values of t_v and the number of possible values of t_v that make $S_{M(k)}$ converge to one increases with respect to N_R . In addition, small changes in t_v can cause large variations in $S_{M(k)}$ passing through local maxima. For this reason, instead of a minimum value of $S_{M(k)}$, here, we used the average value $S_{M(k)}$ of all t_d (t_d between zero and t_t) in the optimisation. This average $S_{M(k)}$ (Fig. 4 (b)) depends only on the number of reactors and vacuum operation, i.e. all the t_d between 0 and 1 h/h are used (and this will provide a conservative sizing of the units).

As a short cut-method to estimate S_M , it was found that S_M could be approximated with high accuracy (Fig. 4 (b), $r^2>0.99$) using the following power law correlations,

$$a_1 = 1.1174 \cdot N_R^{-0.021} \quad (47)$$

$$a_2 = 1.4701 \cdot N_R^{-0.629} \quad (48)$$

$$S_M = a_1 \cdot t_v^{-a_2} \quad (49)$$

The value of S_M calculated using Eq. (49) is used to estimate $I_{C(k)}$ and $I_{HX(k)}$ using Eq. (41) and Eq. (42), respectively. The range of t_d values that yield values of S_M lower than that obtained using Eq. (49) are calculated after the optimisation was performed (to ensure valid optimisation results are obtained).

4.2. Optimisation of operating strategies

Despite the xylose inhibition on the enzymes, for batch process operation the concentration of solids in the reactor was found to be at the maximum limit (100 g/l). This is probably because high concentrations of xylose and glucose are required for a faster growth of the biomass. The maximum concentration of active cells (2.2 g/l) was achieved around 30 h (Fig 5 (a)).

The volume of reactor(s) was found to be the maximum (3750 m³) for all the operating strategies considered. The volume decreases from 80% to 62% of the total reactor volume due to water evaporation (Fig. 5 (a)). The reduction of the reactor volume indicates that fed-batch can be a superior operating mode because a higher amount of substrate can be fed. Hence, when the fed-batch operation mode was optimised, the volume decreased from 80 to 70% (Fig. 5 (b)). As a result, the EP increased when compared to batch operation by 1.3 MM USD, Table 3. This was achieved by increasing the compressor work 1.22-fold, reducing the enzyme load 1.27-fold and reducing the number of reactors from 12 to 10. As with batch operation, it was found that the initial concentration of LCB (Fig. 5 (b)) was at the maximum bound (100 g-non-soluble-solids/l). However, the average concentration of total dry-LCB in the feed was higher, around 20.3-21.7 wt% as opposed to 18.3 wt% (batch process), see Table 3, this was because

the feed stream used a higher substrate concentration (optimisation variable) (Fig. 1). This average concentration of LCB was similar to that used in ethanol production by simultaneous co-fermentation and saccharification (20.9 wt%, Table 2).

The energy requirements of the heat-pump system for the batch process were found to be the lowest at a higher vacuum pressure (0.052 bar) and lower vacuum operational times (73.1 % of the cycle time). It was also found to be related to productivity, where a low rate of recovery of butanol decreases the energy requirement. These results are consistent with the results reported for gas stripping by Lodi et al., (2018).

In the batch process the productivity is limited by substrate availability because substrate is only fed at the start of the fermentation. Hence, it was more difficult to obtain an incremental improvement in the productivity or a reduction in enzyme load in batch process operation than a reduction in the energy requirements.

It is known that the highest productivity is achieved using continuous operation (Tao et al., 2014). These results are consistent with the results obtained in this work as the lowest number of reactors was obtained for continuous operation (8 reactors, see Table 3). However, due to the high energy requirements, the high enzyme load and low yield, continuous operation was found to be the least attractive choice both in terms of economics and energy usage, i.e. continuous operation had the highest energy requirements and the lowest *EP*.

4.3. Total energy requirements

The effect on the *EP* and the total compression work as a function of the number of stages of compression and isentropic efficiency is shown in Fig. 6. Using three stages of compression, the *EP* is reduced 18% if η_s is fixed at 0.55 instead at 0.75. The compression work reduces 28-30 % if the number of stages is two instead of one. When the number of stages of compression was increased from two to three, the total compression work reduced between 8.4 and 9.6 % (η_s between 0.55 and 0.75). Three stages was found to give the optimal *EP* ($\eta_s =$

0.75). However, the EP was only reduced 2-3% because the investment in the compressor increases as well, i.e. more compressors and heat exchangers are required.

Alternatively, the efficiency isentropic may be calculates as (Belaissaoui et al., 2016),

$$\eta_s = 0.1058 \cdot \ln\left(\frac{P_{k-1}}{P_k}\right) + 0.8746 \quad (50)$$

When Eq. (50) was used to estimate η_s the EP using three stages was similar than obtained when fixing η_s equal to 0.75 as the compression ratio P_{k-1}/P_k achieved in the optimisation (between 0.25 and 0.4) gives η_s between 0.73 and 0.78. Whilst, when the number of stages of compression was one ($P_{k-1}/P_k \cong 0.05$), η_s was reduced to 0.55. In this sense, the isentropic efficiency increases with the number of stages of compression.

The total energy requirements of the ISPR by vacuum evaporation depends on both the compression costs and the final purification system – which is performed by double-effect distillation (DED) (Grisales Diaz and Olivar Tost, 2018). Therefore, after the optimal conditions of the reactor were obtained the reactor and final separation system were simulated in Aspen Plus V9® in order to find the total energy requirements for the system and to validate the dynamic modelling results. Due to the dynamic operating conditions, the complex kinetics and the *in situ* recovery, the ISPR-V cannot be simulated in one process unit of Aspen Plus V9®. Hence, the ISPR was simulated in Aspen Plus V9® continuous mode combining a ‘stoic reactor’ followed by the ideal separator ‘Sep’ (using the vacuum recovery rates from Matlab® simulations) and the flash unit ‘Flash2’ (operating at atmospheric pressure). The differences in the compression work between those of the Matlab 2016.a® and the Aspen Plus V9® simulations were around 2.8% (results not shown).

The energy requirements of the final purification by DED were mainly a function of ABE yield, substrate concentration, and ABE recovery in the gas stream. The energy requirements of DED for the fed-batch process were the lowest, (3.45 MJ/kg ABE). However, batch operation had the lowest total energy requirement (9.6 MJ fuel/kg ABE, Table 3). Indeed, it was found that the energy requirements, when compared to continuous operation, were reduced by between 21

and 28% using fed-batch and batch operating systems. Mariano et al., (2011) reported an energy requirement for continuous operation of 13.4 MJ/Kg butanol, which was lower than that reported for batch, >22 MJ/Kg butanol. Our results differ as the substrate concentration used by Mariano et al., (2012a) in the evaluation of batch operation was 2.5 lower. In addition, the final butanol concentration in the reactor for batch and continuous operation were ~2 g/l and ~10 g/l, respectively. A higher butanol concentration decreases the energy requirements as the butanol concentration in the gas stream increases.

The total compressor work was found to be between 1.8 and 2.9 MJ/kg ABE or between 2.4 and 3.4 MJ/kg butanol. In contrast, the energy requirements of compression reported by Mariano et al., (2011) for continuous vacuum fermentation were between 4.4 and 6.5 MJ/kg butanol. In this paper, the total compressor work was around 1.3 and 1.9-fold lower than that reported by Mariano et al., (2011) as three compression stages operating under optimal conditions were used and the excess heat was not used to supply heat required for the distillation system.

In the literature, alternative processes for vacuum recovery have been proposed for butanol production. Of all the alternatives considered liquid-liquid extraction, adsorption and pervaporation have been reported with the lowest energy requirements (between 7 and 10.9 MJ fuel/kg ABE (Grisales Díaz and Olivar Tost, 2016a; Qureshi et al., 2005)). In comparison, the energy requirements of the processes studied in this paper for batch and fed-batch operation were between 9.6 and 10.5 MJ fuel/kg ABE. Hence, batch and fed-batch operation of vacuum fermentation compares favourably with alternative technologies.

5.0. Conclusions

The *EP* of three operating modes (batch, fed-batch and continuous) have been investigated. A heat-pump using a three stage compression system with partial condensation was used to minimise the energy requirements. The operating variables of the reactor (fermentation time, substrate concentrations, initial volume, enzyme load, vacuum times, dilution rate and

pressure set-point) and the heat-pump variables (pressure at the exit of the compressor one and two in the series) were optimised. Based upon the dynamic model used and the estimation of both capital and operating costs it was found that **a)** batch process operation achieved the lowest energy requirements (9.6 MJ fuel/kg ABE) because the butanol concentration in the reactor was the highest (5.8 g/l at the end of fermentation) **b)** a continuous process would have the lowest *EP* (23.6 MM USD) and the highest energy requirements (13.6 MJ fuel/kg ABE), due to the low ABE yield and that **c)** fed-batch operation was the most profitable (37.6 MM USD) because a higher productivity was achieved in comparison with batch fermentation.

The process flowsheet considered in this work assumed that the multiple reactors required to achieve the target productivity were all operated in parallel. However, it was demonstrated that the capital investment required of the heat-pump system may be minimised through the scheduling of reactor operation (sequencing the start-up time of the reactors). The optimal scheduling of the reactors combined with the proposed energy integration scheme gave process energy requirements that are similar to that achieved by units with low energy requirements.

The energy requirements are strongly dependent on the reactor operating conditions, hence, for a fair comparison is recommended that the energy analysis of the integrated reactors must be made using their optimal *EP*.

Appendix

Table A1
Biochemical reaction rates

Rate term	Eq.	Rate term	Eq.
$r_1 = \frac{V_{max1}[Glucose][Biomass]}{Km_1 + [Glucose]} \cdot I_B^{n1} \cdot F$	(1-K)	$r_{10} = \frac{V_{max10}[AACoA][Biomass]}{Km_{10} + [AACoA]} \cdot F$	(11-K)
$r_2 = \frac{V_{max2}[Pyruvate][Biomass]}{Km_2 + [Pyruvate]} \cdot F$	(2-K)	$r_{11} = \frac{V_{max11}[AACoA][Butyrate][Biomass]}{([AACoA] + Km_{b11})([Butyrate] + Km_{a11})} \cdot I_B^{n11}$	(12-K)
$r_3 = \frac{V_{max3}[Acetate][Biomass]}{Km_3 + [Acetate]} \cdot F \cdot I_B^{n3}$	(3-K)	$r_{12} = \frac{V_{max12}[Butyrate][Biomass]}{Km_{12}(1 + Km_{a12}/[Butyrate] + Km_{b12}/[Butanol]) + [Butyrate]} \cdot F$	(13-K)
$r_4 = \frac{V_{max4}[Acetate][AACoA][Biomass]}{(Km_{a4} + [Acetate]) \cdot (Km_{b4} + [AACoA])} \cdot F \cdot I_B^{n4}$	(4-K)	$r_{13} = \frac{V_{max13}[BCoA][Biomass]}{Km_{13} + [BCoA]} \cdot F$	(14-K)
$r_5 = \frac{V_{max5}[CoA][Biomass]}{Km_5 + [CoA]} \cdot F \cdot I_B^{n5}$	(5-K)	$r_{14} = \frac{V_{max14}[BCoA][Biomass]}{Km_{14}(1 + Km_{a14}/[Butyrate]) + [BCoA]} \cdot F$	(15-K)
$r_6 = \frac{V_{max6}[CoA][Biomass]}{Km_6 + [CoA]} \cdot F$	(6-K)	$r_{15} = \frac{V_{max15}[Xylose][Biomass]}{Km_{15} + [Xylose]} \cdot F \cdot I_B^{n15}$	(16-K)
$r_7 = \frac{V_{max7}[CoA][Biomass]}{Km_7 + [CoA]} \cdot F$	(7-K)	$r_{16} = \frac{V_{max16}[Xylose]}{Km_{16}(1 + Km_{a16}/[BCoA]) + [Xylose]} \cdot F \cdot I_{NH}$	(17-K)
$r_8 = \frac{V_{max8}[Glucose][Biomass]}{Km_8(1 + Km_{a8}/[BCoA]) + [Glucose]} \cdot I_{NH}$	(8-K)		

$$r_9 = V_{max9} [Biomass] \quad (9-K) \quad r_{17} = V_{max17} [Biomass] \quad (18-K)$$

$$I_B = \left(1 - \frac{[Butanol]}{[Bumax]}\right) \quad (10-K) \quad \omega = \left(\frac{[Glucose]}{[Glucose] + [Xylose]}\right)^{n_{g/x}} \quad (19-K)$$

$$I_{NH} = \left(1 - \frac{[Acetate]}{[Acemax]}\right) \left(1 - \frac{[butyrate]}{[Butmax]}\right) \left(1 - \frac{[Biomass] + [Biomass dead]}{[Bimax]}\right) \left(1 - \frac{[Glucose]}{[Glumax]}\right) \left(1 - \frac{[Xylose]}{[Xymax]}\right) \cdot I_B^{n_{19}} \quad (20-K)$$

Kinetic model of enzymatic hydrolysis of cellulose in corn stover (Kadam et al., 2004)

$$r_{w1} = \frac{k_{1r} \cdot E_{1B} \cdot R_c \cdot [Cellulose]}{1 + \frac{K_{1G2}}{[Cellulose]} + \frac{K_{1G}}{[Glucose]} + \frac{K_{21X}}{[Xylose]}} \quad (21-K) \quad r_{w3} = \frac{k_{3r} \cdot E_{2F} \cdot R_c \cdot [Cellulose]}{K_{3M} \left(1 + \frac{[Glucose]}{K_{1G}} + \frac{[Xylose]}{K_{31X}}\right) + [Cellulose]} \quad (23-K)$$

$$r_{w2} = \frac{k_{2r} \cdot (E_{1B} + E_{2B}) \cdot R_c \cdot [Cellulose]}{1 + \frac{K_{1G2}}{[Cellulose]} + \frac{K_{1G}}{[Glucose]} + \frac{K_{21X}}{[Xylose]}} \quad (22-K) \quad E_{iB} = \frac{E_{imax} \cdot K_{iad} \cdot E_{iF} \cdot S}{1 + K_{iad} \cdot E_{iF}}, i = 1, 2 \quad (24-K)$$

$$k_{ir(T2)} = k_{ir(T1)} \cdot e^{-E_{at}/R_{cte} \cdot (1/T1 - 1/T2)}, i = 1, 2, 3; 30^\circ C \leq T \leq 55^\circ C; T1 = 55^\circ C \quad (25-K)$$

[Acemax], total inhibition concentration of acetone (mmol/l); [Acmax], total inhibition concentration of acetate (mmol/l); [AACoA], concentration of acetoacetyl CoA (mmol/l); [ACoA], concentration of acetyl CoA (mmol/l); [Acetate], concentration of acetate (mmol/l); [Acetone], concentration of acetone (mmol/l); [BCoA], concentration of butyryl CoA (mmol/l); [Bimax], total inhibition concentration of butanol (mmol/l); [Biomass], concentration of biomass (mmol/l); [Bumax], total inhibition concentration of butyrate (mmol/l); [Butanol], butanol concentration (mmol/l); [Butyrate], butyrate concentration (mmol/l); [Cellulose], cellulose concentration (g/kg); [Cellobiose], cellobiose concentration (g/kg); [Glucose], glucose concentration (mmol/l); [Glucosem], glucose concentration (g/kg); [Glumax] total inhibition concentration of glucose (mmol/l); [Pyruvate], pyruvate concentration (mmol/l); [Xylose], xylose concentration (mmol/l); [Xylosem], xylose concentration (g/kg); [Xymax] total inhibition concentration of xylose (mmol/l); Ea, activation energy (J/mol); Emax, maximum amount of the enzyme that can adsorb onto of substrate (g protein/ g cellulose); F, parameter 'on-off' of the kinetic model of fermentation (zero if [Glucose]<1, else equal to 1); Kad, dissociation term for the enzyme (g protein/g cellulose); KilG, inhibition terms for glucose (g/kg); KilG2, inhibition terms for cellobiose (g/kg); kir, reaction rate terms (kg/g/h); KilX, inhibition terms for xylose (g/kg); Km, constant of kinetic model (mmol/l); K3M, cellobiose saturation term (g/kg); n_{G/x}, parameter of carbon catabolite repression; Rc, reactivity parameter of cellulose conversion (g/g); T, temperature (K); Vmax, maximum rate of reaction (h⁻¹); ω, weighting factor for preference consumption uptake; I_{N/H}, inhibition term for biomass growth (-); r_{wi}, reaction of saccharification of components i (g/h/kg); ri, reaction rate of production / consumption in the fermentation (mmol/h/l)

Table A2
Model parameters of the kinetic model of fermentation ABE

Reaction	Vmax	Km	Km _a	Km _b	n
r ₁	1.16	0.05	4x10 ¹²		1
r ₂	33.1	17.5			
r ₃	0.84	113			1
r ₄	34.1		198	8.4	2.75
r ₆	2.46	0.001			1
r ₇	50.4	2.24			
r ₈	0.92	83.7	0.0013		
r ₉	0.034				
r ₁₀	59.8	3.6			
r ₁₁	0.23		0.062	0.002	1 ^a
r ₁₂	38.3	1.36	113	679.3	
r ₁₃	6.18	2.06			
r ₁₄	146	3.65	520		
r ₁₅	1.72	0.248			0.52
r ₁₆	0.14	1.47	14.02		
r ₁₇	0.0088				

Additional parameters of the metabolic kinetic model					
Parameter	Parameter	Parameter	Parameter	Parameter	Parameter
[Bumax]	210.7	[Acmax]	142.3	[Butmax]	87.48
[Bimax]	581	[Glumax]	916	ng/x	0.096
[Xymax]	3367	n ₁₉	1.73		

a n11 is equal to 0.068 if [Butanol]<11.6 g/l

Km, kma, kmb are constants of the kinetic model (mmol/l); Vmax, maximum rate of reaction (h⁻¹); n, power term of inhibition (-); [Acemax], total inhibition concentration of acetone (mmol/l); [Acmax], total inhibition concentration of acetate (mmol/l); [Bimax], total inhibition concentration of butanol (mmol/l); [Bumax], total inhibition concentration of butyrate (mmol/l); [Glumax] total inhibition concentration of glucose (mmol/l); [Xymax] total inhibition concentration of xylose (mmol/l)

Table A3
Parameters of the sacharification model*

Parameter	Value	Parameter	Value
K _{1ad}	0.4	k _{2r}	7.18
K _{2ad}	0.1	K _{2IG2}	132
E _{1max}	0.06	K _{2IG}	0.04
E _{2max}	0.01	K _{2IX}	0.2
E _a	-23179	k _{3r}	285.5
k _{1r}	22.3	K _{3M}	24.3
K _{1IG2}	0.015	K _{3IG}	3.9
K _{1IG}	0.1	K _{3IX}	201
K _{1IX}	0.1		

K_{ad}, dissociation term for the enzyme (g protein/g cellulose); K_{IG}, inhibition terms for glucose (g/kg); K_{IG2}, inhibition terms for cellobiose (g/kg); k_{ir}, reaction rate terms (kg/g/h); K_{IX}, inhibition terms for xylose (g/kg); K_{3M}, cellobiose saturation term (g/kg); E_a, activation energy (J/mol); E_{max}, maximum amount of the enzyme that can adsorb onto of substrate (g protein/ g cellulose);

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Tables

Table 1

Cost parameters for ABE production from corn stover

Parameters	Units	Value	Eq.
Water cooling cost (Mussatto et al., 2013)	USD/MJ	2.3×10^{-4}	
Enzyme cost (Mussatto et al., 2013)	USD/kg-protein	4.24	
Corn stover cost (Tao et al., 2014)	USD/kg-dry-LCB	0.064	
Electricity cost (Mussatto et al., 2013)	USD/kWh	0.1	
Selling cost of butanol (Zaubal, 2015)	USD/kg	1.0	
Selling cost of ethanol (Zaubal, 2015)	USD/kg	0.8	
Selling cost of acetone (Zaubal, 2015)	USD/kg	0.7	
Initial installation cost of reactors (I_R) ^a (Oudshoorn et al., 2010)	USD	$27846 \cdot V_R^{0.61} + 151967$	(B-1)

V_R (m³) is the volume of reactor; ^a conversion factor of EURO to USD was 1.37

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Table 2
Characterization of corn stover^a

Component	Composition (w/w)
Xylose (SS)	0.047
Cellulose (IS)	0.061
Glucose (SS)	0.008
Other SS:IS (1:1)	0.093
Total	0.209

SS, soluble solids, IS, insoluble solids. Other: lignin, ash, ammonia, etc.

^a Initial concentration used for isobutanol or ethanol production by a simultaneous saccharification and fermentation system (Humbird et al., 2011; Tao et al., 2014))

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Table 3
The effect of operational mode on the performance of ABE production by ISPR-V

Item	Operation mode		
	Batch	Fed-batch	Continuous
ABE yield (g/g)	0.391	0.39	0.305
Butanol yield (g/g)	0.299	0.305	0.247
Cellulose conversion (%)	87.3	85.5	53.2
Xylose conversion (%)	99.9	99.3	82.6
Cycle time (h)	88.2	103.2	504
% of vacuum in the cycle time	73.1	78.9	99.2
S_M	1.17	1.16	1.08
Vacuum pressure (bar)	0.052	0.05	0.047
Butanol recovery (g/g %)	85.8	90.6	89.2
Enzyme load (g-protein/kg-cellulose)	12.7	10.2	8.5
Avg. LCB in the feeding (wt. %)	18.3	20.3	18.3
Final butanol concentration in the reactor (g/l)	5.8	4.8	3.6
Number of reactors	12	10	8
Compression work (MJ/kg-ABE)	1.8	2.2	2.87
Energy consumption of distillation (MJ/kg-ABE) ^a	3.75	3.45	4.14
Total energy requirements (MJ-fuel/kg-ABE) ^d	9.6	10.5	13.3
<i>EP</i> (MM USD/y)	36.6	37.8	23.6

^a Including ethanol dehydration requirements by a heat-integrated distillation system (Grisales Díaz and Olivar Tost, 2016b)

^b the fuel requirements were calculated assuming an efficiency of electricity and steam production from fuel of 0.9 and 0.33, respectively. These included the ABE recovery by the total recovery system (~0.98)

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Figures

Fig. 1. ISPR-V, fed-batch flowsheet. CW, cold water. HX is heat exchanger. F_L is the liquid flow after condensation. W is compressor work. w is mass fraction. T is temperature. w_s and w_c are the average compositions of substrate and cellulose in the detoxified corn stover. w_{is} and w_{ss} are the insoluble solids and soluble solids, respectively. F represents mass flow. S is the total substrate (cellulose plus xylose and glucose) and C is cellulose. The subscripts L and G represent liquid and gas flow, A is acetone, B is butanol, E is ethanol, W is water. The values of the energy requirements and compositions are the optimal conditions found in this work.

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829 **Fig. 2.** Biochemical reaction network for simultaneous saccharification and ABE fermentation. In the
830 'extracellular' reactions, the cellulose is converted to glucose and cellobiose. Glucose and xylose are
831 then fermented by *Clostridium saccharoperbutylacetonicum* N1-4. The gases were assumed to be
832 produced with respect to the stoichiometric reactions for acetone, ethanol and butanol production

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Fig. 3. Example of the graphical method used to calculate the maximum number of reactors operated in the cycle time (N_v^M). In this example, the total number of reactors, N_R is 12 and the cleaning/ turn-around time is 5% (4 h) of the cycle time (t_t), 80 h. N_v is the number of reactors operated under vacuum in a determined interval of time. Three scenarios are shown: Case A, time between the start-up of reactors (t_d) is equal to 5 % of t_t and the vacuum time of one reactor in the cycle time (t_v) was 52 h (65% of t_t). Case B, t_d is equal to 10 % (8 h) of t_t and t_v was 65% of t_t . Case C, t_v is 44 h (55% of t_t) and t_d is equal to 10%.

Fig. 4. Effect of the number of reactors (NR) on the ratio S_M . Turn-around time of the reactor (t_i) is 4 hours, t_v is the time of vacuum operation of reactor ($0 < t_v < 96$ h) and t_t the cycle time (100 h). In case (a), the minimum S_M was found using the optimal time between start-up of the reactors (t_d). Whilst in case (b), S_M was the mean value a t_d between 0 and 100% of t_{tt} . In case (b), continuous lines represent the rigorous calculation and dashed line is from a short-cut method.

Fig. 5. Batch and fed-batch profiles for butanol production from corn stover via vacuum fermentation. a) Batch. b) Fed-batch

Fig. 6. Effect of the number stages of compression on the economic potential (EP) and the total compressor work (WT). Fed-batch operating mode with optimal EP estimates for different isentropic efficiencies (η_s)