Open Loop Optimal Operation and Sensitivity Analysis of a Continuous Biobutanol Fermentation Process with Ex-Situ Adsorption Recovery

Bocun Kim*, Hong Jang*, Jay H. Lee*

* Department of Chemical and Biomolecular Engineering, Korea Advanced Institute of Science and Technology, Daejeon, 305-701, Republic of Korea (Tel: 042-350-3926; e-mail: jayhlee@kaist.ac.kr).

Abstract: Biobutanol is considered to be an attractive biofuel due to its chemical similarity to gasoline, but low fermentation performance caused by butanol toxicity stands as a major obstacle in taking it to full commercialization. As a solution, an ex-situ recovery system with periodically switched adsorption column can be used to maintain the butanol concentration in a fermerter below the critical level. Due to the complex nature of the operation manifested by the periodic operation and the resulting cyclic steady state (CSS), it becomes a challenge to determine the optimal operation strategy of the system. In this study, an optimization problem is formulated to determine the operating condition to maximize a profit function at the CSS for a given feed concentration. This is followed by an open-loop sensitivity analysis to investigate the effect of uncertainties in the model parameters and adsorbent states on performance of the determined recipe.

Keywords: Continuous biobutanol fermentation; Ex-situ adsorption recovery; Cyclic steady state; Nonlinear optimization; Open-loop sensitivity

1. INTRODUCTION

Biobutanol produced by Acetone-Butanol-Ethanol (ABE) fermentation is a promising renewable energy source. However, production through the traditional batch fermentation suffers from low product concentration and low volumetric productivity due to the toxicity of butanol (Green, 2011, Jones & Woods, 1986). To overcome this limitation, several previous papers have suggested the fermentation process integrated with a butanol recovery system, using various separation techniques such as adsorption, pervaporation, liquid-liquid extraction, and gas stripping (Ezeji & Li, 2010, Vane, 2008). The recovery system is intended to maintain the butanol concentration below the threshold of toxicity by removing the produced butanol from the fermentation broth as the fermentation progresses. Among the tested recovery technologies, adsorption is particularly attractive given its simplicity and high energy efficiency (Oudshoorn et al., 2009, Qureshi et al., 2005).

Recently, (Eom et al., 2015) has proposed a fermentation process integrated with an ex-situ butanol recovery (ESBR) system for continuous biobutanol production, and developed a dynamic model of the system. The proposed system includes a fermenter and stirred-tank-type adsorption columns filled with adsorbents highly selective for butanol. The produced butanol is removed while the fermentation broth is circulated between the fermenter and the adsorption column. Switching the saturated adsorption column to a new one enables continuous biobutanol production at a high butanol concentration level by circumventing the butanol inhibition effect, resulting in an increased volumetric productivity. Due to the periodic switching of the adsorption columns, the operation is cyclic and the system converges to a Cyclic Steady State (CSS), in which a same dynamic pattern is repeated. As CSS is an inherent characteristic of the ESBR system, the determination of the operation condition should take the CSS behavior of the system into account.

To achieve an optimal operation of the adsorption based ESBR system ultimately, two distinct periods of development should be considered: the adaptation period where operating recipes are continually improved off-line based on collected operation data, and the stable operation period where operating recipes finally get fixed and just fine adjustments are done on-line to compensate for various disturbances. To establish a strategy for recipe adjustment during the adaption period, effect of uncertainties in the model parameters on the CSS condition should be analyzed first. During this period, parameters of the dynamic model may need to be adjusted and recipe optimization is repeated. After entering the stable period (cell concentration reaches about 8 g/L), a model-based controller should be designed to on-line revise the cyclic operation strategy in consideration of disturbances on operating variables and other uncertainties (e.g. adsorbent degradation).

In this study, as a basis for designing the overall adaption strategy, we define the optimization problem in terms of productivity and loss, and solve the optimization at the CSS. In addition, an open-loop sensitivity analysis at the CSS with respect to model parameter variations is carried out to determine the important parameters to adapt.

Section 2 presents a description of the ESBR system and its dynamic model. In Section 3, the optimization problem is defined, and the optimization methods are introduced. In addition, an open-loop sensitivity analysis is discussed. The results and discussion of the simulation study are reported in
Section 4. Section 5 contains a summary and conclusion of this work.

2. PROCESS DESCRIPTION AND DYNAMIC MODEL

2.1 Process description

During the initial phase of the operation, the ABE fermentation occurs only in the fermenter without feeding and circulation until the butanol concentration reaches a specific level decided based on the consideration for the butanol toxicity and the adsorption efficiency. Next, the cyclic operation begins which involves the following three steps (corresponding to one cycle) (Fig. 1).

1) Filling of the adsorption column with the fermentation broth ($F_{rc} \neq 0$ and $F_{cr} = 0$)

2) Circulation ($F_{rc} = F_{cr} \neq 0$)

3) Restoring the fermenter volume

if $V_{r}<V_{r,0}$, $F_{rc} = 0$ and $F_{cr} \neq 0$

if $V_{r}>V_{r,0}$, $F_{rc} \neq 0$ and $F_{cr} = 0$

In the continuous cyclic operation, the feeding into the fermenter continues to keep the glucose concentration for the growth and production, and the circulation of the fermentation broth occurs between the fermenter and the adsorption column to maintain the butanol concentration below the threshold level. When the adsorbent is saturated and the butanol concentration reaches a specific value, the fermenter volume is restored to the initial level ($V_{r,0}$) in the first step of a cycle, the column is switched to a refreshed one, and the cycle is repeated. In the meantime, the remaining broth in the saturated column is transferred to the harvest tank, and the saturated adsorbents are regenerated by steam for reuse in next cycle.

Fig. 1 Fermentation process integrated with ex-situ adsorption recovery.

To maintain the optimal fermentation condition, the concentrations of glucose, butanol, and ethanol in the fermenter have to be controlled within some proper ranges. Appropriately controlled glucose concentration in the fermentation broth has a favorable effect on the cell growth and productivity but too high glucose concentration can cause a major loss. The concentrations of butanol and ethanol are required to be below the threshold levels of toxicity.

In the ESBR system, the feed rate and the circulation rate are the major operational degrees of freedom. The feed rate directly affects the glucose concentration and cell concentration in the fermenter; a proper feed rate gives a positive influence on fermentation performance but an excessive feed rate gives a large glucose loss and a severe dilution effect. Increase in the circulation rate widens the range of butanol concentration level during the circulation as it speeds up the adsorption. Though the feed concentration is also a decision variable, an operation strategy is first designed assuming a given concentration level in this study because of the highly negative correlation between the feed concentration and the feed rate.

2.2 Dynamic model

A dynamic model of the ESBR by adsorption system requires kinetic models for the product adsorption in the adsorption column and the ABE fermentation, which occurs in the adsorption column as well as in the fermenter. A kinetic model for multicomponent adsorption in the ABE fermentation was developed by using the extended Langmuir theory (Eom et al., 2013).

More recently, an ABE fermentation kinetic model was developed using the Monod/Luedeking-Piret model (Eom et al., 2015). The rate of the cell growth is described in consideration of the three inhibitory effects of substrate, product, and cell mass. Kinetic models for the glucose consumption and product formation are established with respect to a growth-associated part and a non-growth-associated part. Only butanol and ethanol are considered as products because the other products are produced only in small quantities. Detailed models and model parameters can be found in (Eom et al., 2015).

Based on these kinetic models, mass balances for cell, glucose, and products can be constructed for both the fermenter and adsorption column. Eq. (1) ~ Eq. (3) shows the mass balance equations in the fermenter including the effect of dilution and circulation. In the adsorption column, changes in the components’ concentrations by the adsorption as well as the circulation are introduced (Eq. (4) ~ Eq. (6)).

$$
\frac{dX_r}{dt} = \left( \mu_{net,r} - D \right) X_r + \left( F_{r,c} X_{ad} - F_{r,c} X_r \right) \frac{1}{V_r}
$$

(1)

$$
\frac{dS_r}{dt} = \left( S_j - S_i \right) D + \left( F_{r,c} S_{ad} - F_{r,c} S_r \right) \frac{1}{V_r} - r_{s,X} X_r
$$

(2)

$$
\frac{dC_{r,j}}{dt} = r_{r,j} X_r + \left( F_{r,c} C_{ad,j} - F_{r,c} C_{r,j} \right) \frac{1}{V_r} - C_{r,j} D
$$

(3)

$$
\frac{dX_{ad}}{dt} = \mu_{net,ad} X_{ad} + \left( F_{r,c} X_r - F_{r,c} X_{ad} \right) \frac{1}{V_{ad}}
$$

(4)

$$
\frac{dS_{ad}}{dt} = \left( F_{r,c} S_{ad} - F_{r,c} S_{ad} \right) \frac{1}{V_{ad}} - r_{s,X} S_{ad}
$$

(5)

$$
\frac{dC_{ad,j}}{dt} = r_{r,j} X_{ad} + \left( F_{r,c} C_{ad,j} - F_{r,c} C_{ad,j} \right) \frac{1}{V_{ad}} + \frac{dA_j}{dt}
$$

(6)

where $X_r$ and $X_{ad}$ (g/L) are the concentrations of the cell mass in the fermenter and in the adsorption column, respectively. $V_r$ is the volume of the fermenter, and $V_{ad}$ is the volume of
the adsorption column (Fig. 1). $S$, $C_{r,B}$ and $C_{r,E}$ are the concentrations of the substrate, butanol and ethanol in the fermenter, respectively. $F$ is the feed rate into the fermenter, and $S$ is the glucose concentration in feed so $D = F/V$. $r_{j,S}$ and $r_{j,B}$ indicate the specific rate of glucose consumption and product formation $(j= B$ (butanol) and $E$ (ethanol)) in the fermenter, respectively. Concentration (or variable) with subscript $ad$ means the same variable but in the adsorption column. $dA/dt$ is the rate of concentration change of component $j$ by adsorption in the adsorption column. $F_{ad}$ and $F_{ev}$ are the flow rates of circulation from the fermenter to the adsorption column and vice versa; they have the same value during the circulation.

3. OPERATION OPTIMIZATION AND OPEN-LOOP SENSITIVITY ANALYSIS

3.1 Cyclic steady state (CSS)

Due to the periodic switching of the adsorption column at the end of a cycle, the ESBR system shows cyclic patterns in the concentration trajectories. After a sufficient number of cycles, the concentration profiles approach a CSS, in which the end states of a cycle are identical to the beginning states of the next cycle. The successive substitution method is the conventional way to determine the CSS condition (Croft & LeVan, 1994). In dynamic simulation, the terminal state of a current cycle, calculated from the initial state of the current cycle by using the dynamic model introduced in Section 2, is substituted for the beginning state of the next cycle. This substitution is repeated until the difference between the entire state trajectories $(\forall t \in [0, t_f])$ of two successive cycles is smaller than a tolerance ($\epsilon_{CSS}$).

$$x^i(t_f) = H(x^i(0))$$

$$x^{i+1}(0) = x^i(t_f)$$

$$\frac{x^{i+1}(t) - x^{i+1}(t_f)}{x^{i+1}(t_f)} \leq \epsilon_{CSS} \forall t \in [0, t_f], i = X, S, \text{and } B$$

where $x$ represents the vector of system states. $H$ is the mapping (defined by the dynamic model) between the initial state $x^0(0)$ and the end state $x(t_f)$ of $k$th cycle. $k$ is the cycle index, and $t_f$ is the time period of a cycle. $x^i(t)$ indicates the simulated concentration of the $i$th component at time $t$ of the $k$th cycle. ($j=X$ (cell), $S$ (glucose), and $B$ (butanol))

3.2 Optimization problem

For optimizing the operation variables at the CSS of the ESBR system, the objective function should be determined with consideration for butanol productivity and glucose loss. At the CSS condition, profit can be expressed as the subtraction of the glucose loss per hour from the butanol productivity per hour. Therefore, the objective function ($J$) to be maximized is chosen as below:

$$J = w_1 \left( g_a(t_f) \times \frac{m}{t_f} + \frac{C_{ad,B}(t_f) \times V_{ad}(t_f)}{t_f} \right) - w_2 \left( S_{ad,B}(t_f) \times V_{ad}(t_f) \right)$$

The butanol productivity over one cycle consists of the adsorbed butanol on the adsorbents and the remaining butanol (not adsorbed to the adsorbents) in the broth discharged from the adsorption column at switching time ($t_f$). The amount of adsorbed butanol is calculated by multiplying the amount of adsorbed butanol per unit mass of adsorbent ($g_a$) per cycle by the mass of adsorbent ($m$). The quantity of remaining butanol in the discharged broth equals to the butanol concentration in the adsorption column ($C_{ad,B}$) multiplied by the liquid volume of the column ($V_{ad}$) at the switching time. The glucose loss refers to the remaining glucose in the broth discharged from the saturated adsorption column at the end of a cycle. This can be obtained in a similar way to the calculation of the quantity of remaining butanol in the discharged broth. Therefore, to maximize the profit at the CSS, we have to determine the operation variables that maximize an appropriate balance between the butanol productivity and minimize the glucose loss.

At the switching time, the discharged broth from the saturated column is moved to a harvest tank to convert the remaining glucose into products through additional fermentation. However, the delay in production by the remaining glucose has to be considered as a loss of production time. The weighting coefficients, $w_1$ and $w_2$, are chosen as 1 and the glucose conversion yield coefficient of the ABE fermentation, 0.35, respectively. To make the sum of the weighting coefficients equal to one, $w_1$ and $w_2$ are normalized as follow:

$$w_1 : w_2 = 1 : 0.35 = 0.74 : 0.26$$

As mentioned before, there are constraints on the concentrations of glucose and products in the fermenter, which can be described as

$$1 \leq S (g/L) \leq 10$$

$$0 \leq C_{r,B} (g/L) \leq 5$$

$$C_{r,B} - C_{ad,B} \leq 0.1$$

Eq. (11) and Eq. (12) represent the maximal allowable concentrations of glucose $S$ (g/L), and ethanol $C_{r,B}$ (g/L) in the fermenter, respectively, which are determined based on insights and experiences. Since the adsorption column is switched when the butanol concentration reaches a specific value, a separate constraint on the butanol concentration is not required. The last requirement, Eq (13), means that the difference between the butanol concentrations of the fermenter $C_{r,B}$ (g/L) and of the adsorption column $C_{ad,B}$ (g/L) has to be smaller than ten percent of $C_{r,B}$. Note that $C_{r,B}$ is always higher than $C_{ad,B}$ due to the adsorption occurring in the column. If the difference becomes larger than this, the cost of product recovery can be excessive due to the low $C_{ad,B}(t_f)$.
The optimization problem with the nonlinear system dynamic model at the CSS can be stated as follows:

\[
\begin{align*}
\text{Max} & \quad J(x,u,q) \\
\text{s.t.} & \quad G'(x,u,q) = 0 \\
& \quad \left| \frac{x_i^k(t) - x_i^{k+1}(t)}{x_i^k(t)} \right| \leq \epsilon_{CSS} \quad \forall t \in [0, t_f], \; i = X, S, \text{and } B \\
& \quad u_l \leq u \leq u_u
\end{align*}
\] (14)

where \( J \) is the objective function (Eq. 10). Optimization variable vector \( u \) includes the feed rate \( F \) and the circulation rate \( F_{cr}/F_r \) with a lower bound \( u_l \) and an upper bound \( u_u \). \( G=0 \) is the system model, and \( D \leq 0 \) refers to the inequality constraints specified earlier. Other design variables represented by \( q \) are assumed to be fixed \textit{a priori} based on insights and experiences obtained from the laboratory and pilot experiments. For instance, the ratio of the volume of fermenter to the adsorption column without the adsorbent are chosen to be 10 so they are set as 200 L and 20 L, respectively.

3.3 Optimization method

In the operation optimization, the system is assumed to have converged to a Cyclic Steady State (CSS) for given operation variables and the objective function is computed at the CSS. The optimization problem in the ESBR system is highly nonlinear due to the nonlinear objective function and terms describing the nonlinear kinetics of the extended Langmuir isotherm and the ABE fermentation. Therefore, convergence of the optimization problem can be quite sensitive to the initial guess. For this reason, a good initial guess in the feasible region should be chosen first.

To identify the feasible region and a good initial guess of the optimization variables, i.e., the feed rate and the circulation rate, for a given feed concentration, a uniform grid search is performed with the search size of 0.1 L/h for the feed rate and 10 L/h for the circulation rate. The search range of the feed rate and the circulation rate are determined as 5 L/h – 15 L/h and 150 L/h – 250 L/h, respectively. Infeasible regions of the search domain (e.g., those that do not converge at all or lead to a CSS condition violating the constraints) are discarded, resulting in a feasible operation region for a given feed concentration. Therefore, the initial guess is selected as the point giving the maximum value of the objective function within the feasible region. Since the number of grid points grows exponentially in the search dimension, the grid search is practical only for searches in one or two dimensions (Neumaier, 2004, Strongin & Sergeyev, 2000).

For solving the constrained nonlinear optimization problem with the good initial guess, \texttt{fmincon} in MATLAB R2014b Optimization Toolbox based on the SQP (Sequential Quadratic Programming) algorithm is employed.

3.4 Open-loop sensitivity analysis

Since the number of parameters is large compared to the measurement data that can be collected during the operation, modification of all parameters seems impractical. Through an open-loop sensitivity analysis at the CSS condition, we can identify the significant parameters as well as evaluate the effect of the parameters’ variations on the CSS condition quantitatively.

The reference CSS trajectory (or the reference objective function value) is derived from the dynamic simulation using the optimal choice of operation variables obtained from the optimization. The terminal state of the reference CSS is used for the initial state in the analysis. The new CSS condition with perturbed parameters is also determined by the successive substitution introduced in Section 3.1, and its end state and objective function value are compared to those of the reference. Endpoint deviations (ED) of the three main state variables (cell, glucose and butanol) and objective function value resulting from each parameter perturbation are calculated as below:

\[
\text{ED}_i^p (\%) = \frac{y_i(p+\Delta p,t_f) - y_i^{ref}(p,t_f^{ref})}{y_i^{ref}(p,t_f^{ref})} \times 100 \tag{15}
\]

where \( y_i^{ref}(p,t_f^{ref}) \) and \( y_i(p+\Delta p,t_f) \) represent the simulation result, e.g., a component’s concentration in the fermenter at the terminal time of a cycle or the objective function value, of the reference CSS and the new CSS associated with perturbed parameter \( p+\Delta p \), respectively.

4. RESULT AND DISCUSSION

4.1 Dynamic simulation and optimal operating condition

Dynamic simulation for the analysis of the CSS behavior is executed according to the operation procedure described in Section 2.1 with the dynamic model presented in Section 2.2. For the dynamic simulation and optimization, we set the initial concentrations of cell, glucose, butanol, and ethanol in the fermenter based on the experiment and design requirements as \( X_{c,0} = 8.5 \text{ g/L}, \; S_{x,0} = 10\text{ g/L}, \; C_{c,0} = 7\text{ g/L}, \) and \( C_{G,E,0} = 2 \text{ g/L}, \) respectively. An ordinary differential equation (ODE) solver, \texttt{ode45} in MATLAB, is used for the integration of the differential mass balances Eq (1) – (6).

Fig. 2. Concentration profiles in the fermenter at the CSS

Fig. 2 shows the concentration profiles at the CSS resulting from the dynamic simulation with the feed concentration of 200 g/L, the feed rate of 11 L/h and the circulation rate of
250 L/h. With tolerance for the CSS convergence set as $10^{-3}$, the system approaches the CSS after 33 cycles. Under the CSS for the specified rates of feed and circulation, the concentrations of cell and glucose are almost constant, whereas the concentrations of butanol and ethanol show pronounced time-varying cyclic behavior (Fig.2).

At the beginning phase of the circulation step, the concentrations of the products in both the fermenter and the adsorption column decrease rapidly due to the rapid adsorption by fresh adsorbents in the column. After about an hour, the concentrations of the products increase gradually since the capacity of the adsorbents is exceeded. Thus, they show a cyclic pattern at the CSS as shown in Fig. 2.

The operation optimization is conducted for several feed concentrations: 150, 200, and 250 g/L. The results of the optimization for three different feed concentrations are summarized in Table 1, and Fig. 3 shows the optimal objective function values for the given feed concentration values. As the feed concentration decreases, the optimal feed rate increases, as expected, whereas there is no apparent correlation between the feed concentration and the optimal circulation rate. Since high circulation rate speeds up the adsorption and reduces the time for the filling of the adsorption column, the values of butanol productivity and objective function increase as the circulation rate increases. Note that, a fermenter volume at the switching time is higher than its initial level with the feed concentration of 150 g/L due to the large feed rate needed.

![Fig. 3. The objective function values for different feed concentrations (with w1=0.74 and w2=0.26)](image-url)

<table>
<thead>
<tr>
<th>Feed concentration (g/L)</th>
<th>Feed rate (L/h)</th>
<th>Circulation rate (L/h)</th>
</tr>
</thead>
<tbody>
<tr>
<td>150</td>
<td>13.77</td>
<td>250</td>
</tr>
<tr>
<td>200</td>
<td>11.44</td>
<td>250</td>
</tr>
<tr>
<td>250</td>
<td>7.81</td>
<td>250</td>
</tr>
</tbody>
</table>

Table 1. The optimized operating variables

Fig. 3. The objective function values for different feed concentrations (with w1=0.74 and w2=0.26)

To explain the optimization results, we plot the trends of butanol productivity and glucose loss separately under each optimal CSS condition for the given feed concentrations in Fig 4. As shown in Fig. 4, the value of glucose loss decreases as the feed concentration increases. A lower feed rate at a higher feed concentration leads to a larger return of the broth from the adsorption column to the fermenter at the switching time, resulting in a smaller $V_{ad}(t)$. According to Eq (10), a reduction in $V_{ad}(t)$ leads to decreases in butanol productivity and glucose loss. When we compare the values of butanol productivity and glucose loss for the feed concentration of 200 and 250 g/L (Fig. 4), the decrease in butanol productivity dominates over the decrease in glucose loss as the feed concentration increases. Thus, the objective function value of at the feed concentration of 200 g/L is higher than that at 250 g/L. On the other hand, the larger feed rate with the feed concentration of 150 g/L causes a huge dilution effect on the cell concentration. For this reason, the value of butanol productivity decreases when the feed concentration decreases further from 200 g/L to 150 g/L. As a result, the feed concentration of 200 g/L gives the highest optimal objective function value (Fig. 3).

![Fig. 4. Butanol productivity (square) and glucose loss (triangle) for different feed concentrations)](image-url)

4.2 Open-loop sensitivity analysis at CSS

There are 16 parameters in the ABE fermentation kinetic model: the parameters for the Monod equation ($\mu_0$ and $K_b$), the coefficients for the inhibition effects of substrate ($K_i$), butanol concentration ($P_B$, $i_B$, $P_{\beta_B}$, and $i_{\beta_B}$) and cell concentration ($P_X$ and $i_X$), the cell death rate constant ($K_d$), the substrate consumption parameters ($\alpha_{cO_B}$ and $\beta_{cO_B}$), and the yield coefficients and other parameters related to the product formation ($\alpha_B$, $\beta_B$, $\alpha_E$ and $\beta_E$). In addition, the adsorption kinetic model has 6 parameters: the coefficients for the maximum per unit mass of adsorbent ($q_{E,m}$ and $q_{L,m}$), the adsorption-equilibrium constant ($B_E$ and $B_L$) and the adsorption kinetics ($k_0$ and $k_0$). In addition, two initial state variables for the adsorbent ($q_{E,0}$ and $q_{L,0}$) are important as they indicate the amount of the residual products after the regeneration. We perform the open-loop sensitivity analysis with ±10 % variations in each of the 24 parameters. The reference CSS condition is the optimal CSS condition of the feed concentration of 200 g/L resulting from Section 4.1.

The feasibility and ED of the objective function value calculated by using Eq (15) for each parameter perturbation are reported in Table 2. In the result, the CSS conditions of four perturbation cases turn out to be infeasible, as they violate the required upper limit of the glucose concentration.
in the fermenter; these include the lower values of the maximum growth rate $\mu_m$, the maximum cell concentration $P_X$, $\alpha_{\text{Glu}}$, and the higher value of the initial state of adsorbent for butanol $q_{B,0}$. Besides the infeasible cases and the intrinsic parameter $\mu_m$, six parameters including $k_p$, $P_X$, $i_X$, $\alpha_{\text{Glu}}$, $\alpha_B$, $q_{B,m}$ have significant effects on the objective function value at the CSS (Table 2). Since the changes in $k_p$, $P_X$, $i_X$, $\alpha_{\text{Glu}}$, and $\alpha_B$ indicate a metabolism change (e.g. the phase transition from acidogenesis to solventogenesis) during the initial operation, they would have to be regarded as critical parameters in the adaptation period. Parameter $q_{B,m}$ corresponding to the adsorbent capacity for butanol adsorption can be considered as a disturbance to represent the adsorbent degradation in the stable period.

5. CONCLUSIONS

The cyclic steady state (CSS) behavior of the butanol fermentation with an ESBR system is analyzed. Based on the insight gathered, dynamic optimization is performed to determine the optimal feeding rate and circulating rate leading to the most profitable CSS condition for a given feed concentration. Through the open-loop sensitivity at the CSS, significant model parameters are identified and the effects of their variations on the objective function value and constraint satisfaction are evaluated. This work provides insight and a basis for developing the strategies of the adaptation and control for the ESBR system.

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REFERENCES


### Table 2. Results of open-loop sensitivity analysis for 24 parameters’ variation at the CSS

<table>
<thead>
<tr>
<th>Parameters</th>
<th>Feasibility</th>
<th>ED of objective function value (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>$\mu_m$</td>
<td>Infeasible in the case of -10%</td>
<td>3.6/-114</td>
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<tr>
<td>$K_S$</td>
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<td>-0.5/0.4</td>
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<tr>
<td>$K_f$</td>
<td></td>
<td>0.2/-0.1</td>
</tr>
<tr>
<td>$k_d$</td>
<td></td>
<td>0.8/-3.5</td>
</tr>
<tr>
<td>$P_B^{'}$</td>
<td></td>
<td>0.1/-0.1</td>
</tr>
<tr>
<td>$i_B$</td>
<td>Infeasible in the case of -10%</td>
<td>-0.1/0</td>
</tr>
<tr>
<td>$P_X$</td>
<td>Infeasible in the case of -10%</td>
<td>4/-21.8</td>
</tr>
<tr>
<td>$i_X$</td>
<td>feasible</td>
<td>-4.3/2.4</td>
</tr>
<tr>
<td>$P_B$</td>
<td></td>
<td>0/0</td>
</tr>
<tr>
<td>$i_B^{'}$</td>
<td></td>
<td>0/0</td>
</tr>
<tr>
<td>$\alpha_{\text{Glu}}$</td>
<td>Infeasible in the case of -10%</td>
<td>-5.4/-17.4</td>
</tr>
<tr>
<td>$\beta_{\text{Glu}}$</td>
<td>feasible</td>
<td>0/0</td>
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<tr>
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</tr>
<tr>
<td>$\beta_B$</td>
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<td>0/0</td>
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<tr>
<td>$\alpha_E$</td>
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<td>$\beta_E$</td>
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<tr>
<td>$q_{B,m}$</td>
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<td>1.2/-3.6</td>
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<td>Infeasible in the case of +10%</td>
<td>8.5/0</td>
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<td>$q_{E,m}$</td>
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<tr>
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<tr>
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