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Abstract: The simulated moving bed reactor (SMBR) is a device in which reaction and separation processes take place simultaneously. The separation of products allows higher conversion and high-purity product can be also obtained. In this work, a mathematical model has been presented to predict the behavior of the SMBR in the sucrose inversion process. For this process, the triangular region which defines operating conditions to recover high-purity products in SMBR has been obtained using two modeling strategies. The set of partial differencing equations is solved by

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finite volume method. The influence of some operation conditions on the reactor performance is analysed for the sucrose inversion process.

Keywords: Simulated moving bed reactor, sucrose inversion, SMBR modeling strategies, computer simulation

INTRODUCTION

The use of separation processes by adsorption in industrial scale has grown in the last years in several areas with the coming of advanced technologies, allowing highly efficient processes in comparison to other classical separation methods. The simulated moving bed (SMB) technology, nowadays one of the most advanced separation techniques, has been successfully applied in various areas: petrochemical (separation of xylene isomers), pharmaceutical (separation of enantiomers) and food (separation of sugars) industries, fine chemical, and others (1–5). There are some benefits obtained by the use of SMB units in several processes—high purity and recovery of the species, better use of the solid inventory and lower eluent consumption with a smaller dilution of the products, smaller equipments—with inherent properties such as its continuous nature and its counter-current operation without the true flow of the solid phase. This last quality is obtained by an original system of valves that, combined with one or more packed bed columns, simulates the counter-current flow of the current phases. Due to the SMB technology potential, it has been recently introduced in the area of reaction engineering, appearing in the so-called simulated moving bed reactor (SMBR) (6). The reaction and separation processes take place simultaneously inside the SMBR, being therefore a strong competitor to processes in which reactors are followed by separation units. The main application of this type of reactor is in processes where the reactions are strongly limited by the chemical equilibrium. SMBR can also be applied in processes where the removal of inhibitors, catalytic poisons, and acceptors products improve the reaction (7).

Figure 1 presents a scheme of a SMBR in the configuration of four sections in closed circuit, which is a type of configuration successfully applied in reaction process of estherification and enzymatic reactions. Other configurations are possible depending on the character of the reactive process and of the separation that is being considered. The unit is composed by several subsections (fixed bed columns) divided in sections by the inlet streams (feed and eluent) and outlet streams (raffinate and extract). The subsections may be packed with adsorbent and catalyst particles (8), or catalysts with adsorbents properties (9) or even just adsorbents in the case of a homogeneous reaction in the fluid phase (10). To simulate the counter-current operation, these external streams are simultaneously shifted from its current position to a subsequent subsection in direction of the fluid flow at

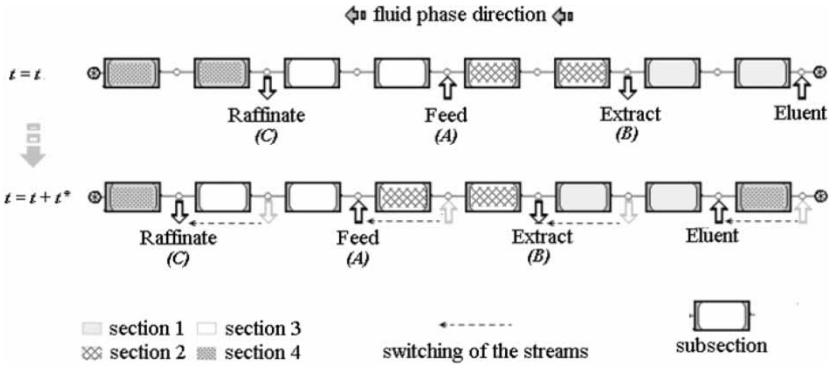

Figure 1. Representative scheme of simulated moving bed reactor (SMBR)—closed circuit/four sections.

fixed time intervals (switching time). Therefore, a rearrangement of the sections takes place at each new switching interval. In each section, a set of functions should be executed so that the reaction process is totally finished, leading to high reactant conversion, and so that the separation process is proper, with the products being transported to the raffinate stream (species with lower affinity to the adsorbent) and to the extract stream (species with higher affinity with the adsorbent) with a high level of purity. As an example, the decomposition reaction $A \leftrightarrows B + C$ catalyzed by particles with catalytic and adsorptive properties is considered. In this case the reaction product B is more strongly adsorbed species, while product C has lower affinity with the solid matrix. The reagent A should be completely consumed in sections 2 and 3, so that it does not contaminate the outlet streams and so that the conversion is complete. As for the liquid mass flow of products B and C, these species are collected in the extract and raffinate streams, respectively, which is possible due to the adsorption of B and desorption of C.

In section 1, the species B should be desorbed and recovered in the extract stream and the regeneration of the solid will be achieved; in section 4, the species C should be adsorbed and the eluent regenerated.

The success of the operations accomplished in SMBR, like in SMB units to separation, depends on the proper choice of the stationary phase used and on the correct definition of the operation conditions, particularly of the inlet and outlet flow rates of the reactor streams and switching time. Modeling and dynamic simulation of SMBR have been gaining attention, especially because these tools facilitate the choice of optimized parameters, for the maximum performance of the reactor.

Many mathematical formulations presented in the literature, considering the reaction and adsorption phenomena, can be interpreted as different versions of the general model, which takes into account all the important

effects of nonideality of the columns: axial dispersion, diffusion in the (micro- or macro-) pores, and resistances to mass transfer between the fluid and solid phases.

Different ways to describe these effects distinguish the various mathematical approaches. In a more general scale, most of the models of the reacting SMB units reported in the literature use models of transport, dispersion, and equilibrium. These models are based on the hypothesis of instantaneous equilibrium of adsorption between the phases and a linear driving force approximation (*LDF*) for the mass transfer from fluid phase to the solid phase (8–13). In some work the bi-*LDF* approximation was also applied (10).

For the resolution of the partial differential equations obtained from the mathematical formulation of these units, it is necessary to use numerical techniques. Among the numerical methods, the method of finite difference and the method of the orthogonal collocation have been often applied.

In this work is presented a mathematical model to predict the reaction and separation process to the sucrose inversion in SMBR. As numerical technique is used volume finite method. Two modeling approaches are discussed to achieve the triangular region in sucrose inversion process: the SMBR which represents the actual configuration and another that is based on the equivalent true moving bed reactor (TMBR). The intention is to show that the choice of one or the other modeling strategy on SMBR may be as significant to achieve performance requirements as it was to SMB separator. The influence of operation parameters on behavior of the SMBR in this enzymatic process is analyzed.

DESCRIPTION AND MATHEMATICAL FORMULATION OF THE PROCESS

The sucrose inversion process is applied to the production of fructose and glucose syrup in high concentrations for the food industry. The sucrose is converted in both sugars by action of mobilized enzyme *invertase* through an irreversible kinetics given for

$$\text{sucrose} + \text{water} \xrightarrow{\text{enzyme}} \text{fructose} + \text{glucose} \qquad (1)$$

It has been shown that the SMBR technology can offer important advantages in the sucrose inversion (7, 14). In the literature, experimental results have been presented showing that it is possible to recover both sugars in high purity; the design methodology has been proposed to the SMB reactor (10). Recently, the sucrose inversion process has been also applied in two systems that are modifications of traditional SMBR: one based on nonsynchronous shift of the inlet and outlet ports (*VARICOL* process) and another with

distributed feed systems (15). The system studied in this work is shown in Fig. 1, where the sucrose is fed by the feed stream and transformed into glucose and fructose. The used eluent is the enzyme diluted in water that eventually can be adsorbed on external surface of the adsorbent. The stationary phase is composed of particles that adsorb only the products of the inversion. The glucose is the less adsorbable product, and the fructose is the most adsorbable species.

The modeling of this system, as well as of other systems that use the SMB technology, can be obtained using two different approaches: one describes the process from the analogy with a true moving bed, where there is flow of both phases; and the other considers the true conception of reactor operation, i.e., the permutation of the inlet and outlet streams of the unit. The use of the first approach requires that equivalence relationships of geometric and dynamic nature are obeyed. It has been shown that this methodology is adequate to predict the operation in cyclic steady-state of processes where the unit sections contain three or more subsections. The steady state of the processes in SMB units is a cyclic state. In each section of the reactor, an identical transient state occurs during each interval of permutation time. In general, this dynamic behavior cannot be approached by an equivalence model, making it necessary to use a time-dependent model for an appropriate description of its behavior.

The SMBR model equations for the processes have been obtained from mass balances of chemical species conservation in the different phases and can be represented by the generic form of Eq. (2). The corresponding terms for the phases are given in Table 1. The differential mass balance equations for chemical species i are written for each subsection of a SMBR (denoted by subscript k).

$$\frac{\partial \phi}{\partial t} + \frac{\partial}{\partial z}(U_k \phi) = \frac{\partial}{\partial z}\left(C_1 \frac{\partial \phi}{\partial z}\right) + S^{\phi} \tag{2}$$

Table 1. Identification of terms in generic equation according to the considered phase

Phase	ϕ	U_k	C_1	S^{ϕ}
Interparticular fluid[a]	$C_{i,k}$	v_k	D_k	$(1 + \vartheta K_{En})R_{i,k} - \vartheta k_{fp}(C_{i,k} - \bar{C}_{pi,k})$
Intraparticular fluid[b]	$\bar{C}_{pi,k}$	0	0	$\frac{k_{fp}}{\varepsilon_p}(C_{i,k} - \bar{C}_{pi,k}) - \frac{k_c}{\varepsilon_p}(q^*_{i,k} - \bar{q}_{i,k})$
Solid[b] (within the particles)	$\bar{q}_{i,k}$	0	0	$k_c(q^*_{i,k} - \bar{q}_{i,k})$

[a]For sucrose, k_{fp} is equal zero; for enzyme, S^{ϕ} is equal zero.
[b]Only fructose and glucose are present.

The reaction rate is described by a *Michaelis-Menten* equation in Eq. (3)

$$R_{Su,k} = -k_r \frac{C_{Su,k} C_{En,k}}{K_M + C_{Su,k}} \tag{3}$$

For intraparticle fluid phase, the mass transfer resistance has been taken into account by a global rate coefficient k_{fp}, which includes both the effects of external film and intraparticle diffusion resistances. This model allows the use of any adsorption equilibrium relation.

The initial conditions are

$$t = 0: \qquad C_{i,k} = \overline{C}_{pi,k} = \overline{q}_{i,k} = 0 \tag{4}$$

The boundary conditions are specified as

$$v_k C_{i,k}|_{z=0^-} = v_k C_{i,k}|_{z=0^+} - D_k \left.\frac{\partial (C_{i,k})}{\partial z}\right|_{z=0^+} \tag{5}$$

$$\left.\frac{\partial C_{i,k}}{\partial z}\right|_{z=L_C} = 0 \tag{6}$$

$$\overline{C}_{pi,k} = \overline{C}_{pi,k+1}|_{z=0} \tag{7}$$

$$\overline{q}_{i,k} = \overline{q}_{i,k+1}|_{z=0} \tag{8}$$

Different boundary conditions for the subsections are established for the entrance and exit points of species. Therefore, the mathematical model is complemented.

- In the entrance point of the feed stream,

$$C_{i,k+1}|_{z=0} = \frac{v_{Fe}}{v_3} C_{i,Fe} + \frac{v_2}{v_3} C_{i,k} \tag{9}$$

- In the exit point of the products and in the points between the columns inside the section,

$$C_{i,k+1}|_{z=0} = C_{i,k} \tag{10}$$

- In the entrance point of the eluent stream,

$$C_{i,k+1}|_{z=0} = \frac{v_4}{v_1} C_{i,k} + \frac{v_{El}}{v_1} C_{i,El} \tag{11}$$

For each subsection, the boundary conditions are changed after the end of each switching time. In view of the time dependence of the boundary conditions, the reactor reaches a cyclic steady state. This set of model equations can be also used to represent the equivalent TMBR model since the generic velocity for the intraparticular fluid phase is equal to the velocity of the solid phase. In this case, the subscript k represents the four

sections of the TMBR instead of SMBR subsections. The equivalent solid velocity is given by

$$u_S = \frac{L_C}{t^*} \tag{12}$$

where L_C is the length of one column and t^* is the switching time. The equivalent counter-current liquid velocity in section k to the TMBR is related to the corresponding liquid velocity in the SMBR by

$$v_k^{TMBR} = v_k^{SMBR} - u_S \tag{13}$$

RESULTS AND DISCUSSION

In order to solve numerically the model equations presented to the SMBR, computational algorithms have been developed in FORTRAN language. Partial differential equations are discretized using Finite Volume Method in which discrete equations are obtained by conservation balances on elementary volumes. Modified Strongly Implicit Method (*MSI*) is used to solve ODE systems. In this development the co-localized variable arrangement is also used in computational grids and interpolation functions from *WUDS* scheme (Weight Upstream Differencing Scheme). Details can be found elsewhere (16).

The dynamics of the reactor for sucrose inversion are presented by concentration profiles based on values of the operating conditions listed in Table 2. The equilibrium model concept was used to provide adequate values for the operation conditions. In Table 2 values of model parameters used in numerical simulations are also presented. Adsorption equilibrium data, axial dispersion, mass transfer, and reaction parameters were experimentally obtained from literature (10).

Reactors that use SMB technology reach a cyclic steady state, its behavior time dependent at every moment. The prediction of reactor behavior description can be accomplished the three ways: the exact evolution of the concentration profiles, the average concentration profile, and the instantaneous concentration (evaluated in a required time of the switch time interval). In Fig. 2a, it is possible to verify the initial variation of glucose and fructose concentrations, in raffinate stream, during the switching time. Figure 2b presents sugars concentration in raffinate stream when the reactor has already reached cyclic steady state. The condition of cyclic steady state was obtained approximately after 12h or 180 switching cycles. The average and instantaneous (in the half of the switch time interval) concentration profiles can also be verified in Fig. 2. As the number of subsections in the SMBR increases, the average and instantaneous (in the half of the switch time interval) concentrations profiles become closer.

Table 2. Operation conditions and model parameters in the sucrose inversion process in a SMB reactor (10)

	Model parameters	
Operation conditions	Adsorption parameters	Reaction parameters
$C_{Su,Fe}$: 80.0 kg/m^3		
$C_{En,El}$: 0.04 kg/m^3	K_{Gl}: 0.17	K_M: 23.0 kg/m^3
Q_{Fe}: $6.03 \cdot 10^{-8}$ m^3/s	K_{Fr}: 0.43	k_r: 0.84 s^{-1}
Q_{El}: $18.96 \cdot 10^{-8}$ m^3/s	K_{En}: 5.00	
Q_{Ra}: $9.21 \cdot 10^{-8}$ m^3/s	ε; ε_p: 0.4; 0.1	Reactor design
Q_{Ex}: $15.78 \cdot 10^{-8}$ m^3/s	Peclet (Pe): 1500	N. of columns: 12 (3-2-5-2)
$Q_{Recycled}$: $4.00 \cdot 10^{-7}$ m^3/s	k_{fp}: $4.12 \cdot 10^{-2}$ s^{-1}	Length (L_C): 0.29 m
t^*: 204.0 s	k_c: $2.50 \cdot 10^{-2}$ s^{-1}	Diameter: 0.026 m

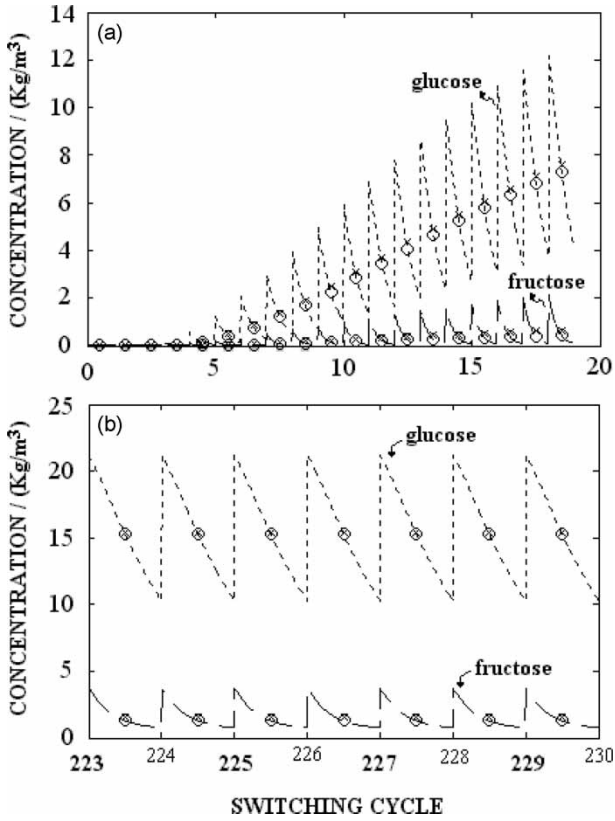

Figure 2. Concentration profiles of the products of the sucrose reaction in the raffinate stream in SMBR four sections. (a) Beginning of the operation in unit; (b) cyclic steady state condition. (– – –) Exact transient concentration profile in the unit, (x) average concentration profile in the switch interval, (⊖) instantaneous concentration profile in the half of the switch interval.

Figure 3a shows average concentration profiles in the SMBR for the case described in Table 2. Experimental and numerical results are plotted. Both numerical profiles are obtained from equivalent TMBR model. One can verify a good agreement between numerical profiles, even with different numerical methodologies: finite volume and finite element collocation method; and also good agreement between the theory and experimental. According to literature, the experimental points show concentrations of samples collected in half of the switching time. The good agreement obtained among the numerical and experimental profiles is related to the high number of subsections in the reactor. In Fig. 3b, numerical profiles obtained by two strategies of solution of the process in SMBR are compared.

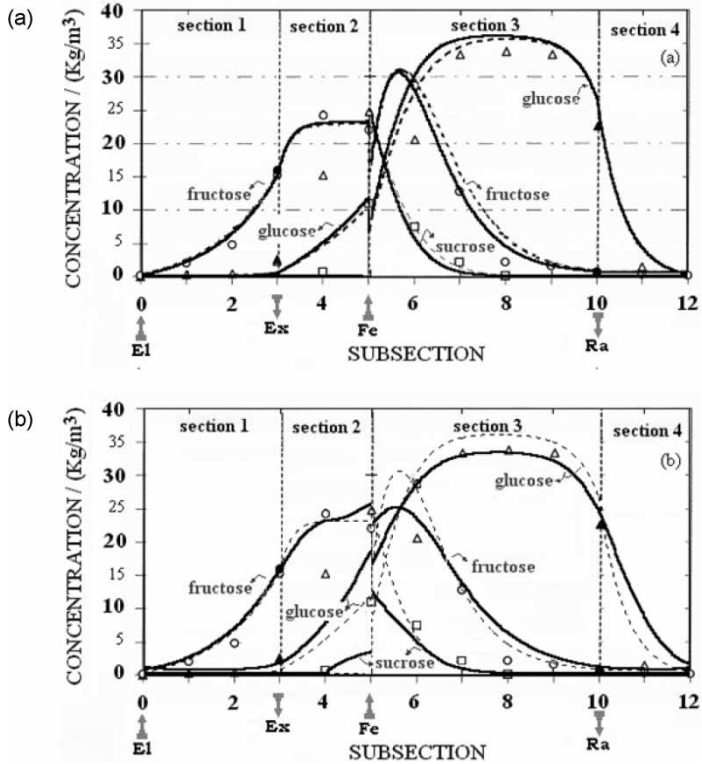

Figure 3. Average concentration profiles of the inversion process in the SMBR in the cyclic steady state; (○), (□), and (△) experimental points for the fructose, sucrose, and glucose, respectively (10). (a) (– – –) Numerical profiles obtained from literature (10), and (——) numerical concentration profiles obtained in this work (both using TMBR equivalent model). (b) (– – –) Numerical profiles using TMBR equivalent model; and (——) numerical concentration profiles using the model of actual SMBR conception.

By analysis of Fig. 3b, a good agreement is observed among the profiles. However, it is verified that the equivalent TMBR model is not able to predict some reagent in section 2. On the other hand, the intermittent model shows this occurrence, which is due to the jump of the inlet feed stream from one subsection to the next one in the direction of fluid flow when one switching cycle is completed. As sucrose is not adsorbed (homogeneous reaction) and is totally consumed inside sections 2 and 3, there is no problem in the current process. However, there are some situations where such information is important. For example, in SMBR with low flow rates in section 2, one can have contamination of the extract by the reactant.

To analyze the verified difference between SMBR behaviors using one or the other model strategy, the region of separation in m_2 vs. m_3 plane, where one can obtain high purity of products, is investigated. Figure 4 shows the

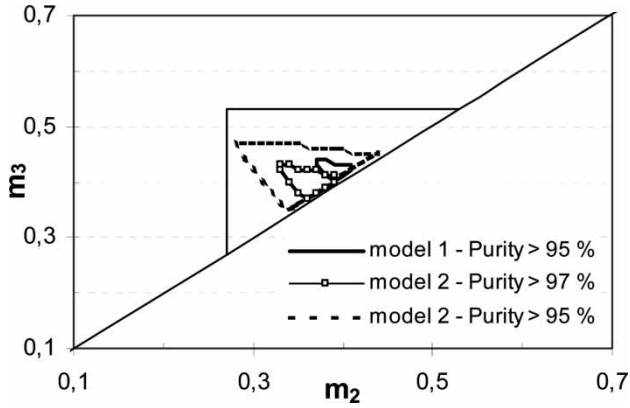

Figure 4. Region of reaction and separation (high product purities) in an $m_2 \times m_3$ plane for the SMBR in inversion sucrose process. Model 1 = actual SMBR conception model; Model 2 = model of equivalence between SMBR and TMBR. Process parameters are listed in Table 2, except $C_{En,El} = 0.20\,\mathrm{kg/m^3}$.

triangular region, based on the equilibrium theory, of the complete separation in the plane of m_2 and m_3 parameters—Eq. (14)—where simple separations of species would be possible—without mass transfer resistances or chemical reactions. It is known that both the mass transfer resistances and the presence of a reaction inside of units may have influence in the separation area where a complete separation is expected to occur. Using the model parameters and reactor design listed in Table 2 (except, $0.20\,\mathrm{kg/m^3}$ enzyme concentration), the regions of separation for the sucrose inversion were found by performing successive simulations using the cyclic steady-state form of the actual SMBR conception model (Model 1) and the steady-state form of the TMBR model (Model 2). The m_1 and m_4 parameters are equal to 0.635 and 0.216, respectively.

In Fig. 4, it is shown that different regions where both extract and raffinate purity is above 95% (Models 1 and 2) and 97% (Model 2), as indicated. The region obtained from Model 1 is narrower than that obtained from Model 2 for case of purities above 95% in the SMBR. It means that if one takes any operation condition defined by region of separation designed from Model 2 to operate a SMBR (to this process, at least), one cannot reach the required performance. Therefore, the use of a TMBR equivalent model can lead to significant differences between estimated and experimental reactor performance; however, as a starting point, this strategy may be used.

$$m_j = \frac{Q_j' t^* - \varepsilon V_C}{(1-\varepsilon) V_C} \tag{14}$$

For the analysis of the SMBR with four sections for the sucrose inversion process, some numerical studies were carried out by varying some operating

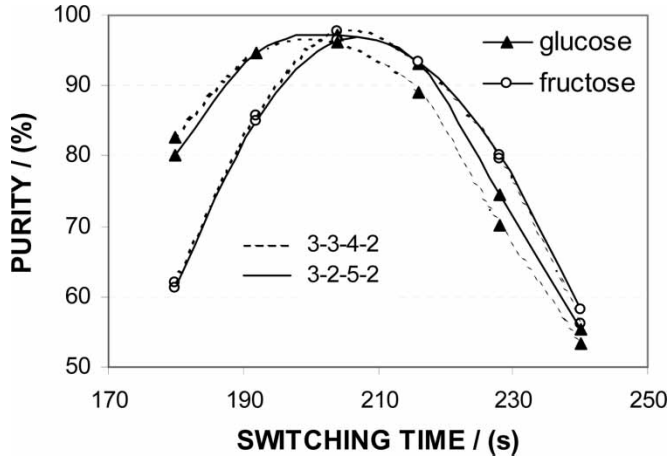

Figure 5. Effect of the switching time on product purity for two different SMBR configurations in inversion sucrose process. (○) Fructose, (▲) Glucose. Configuration (——) 3-2-5-2, (– – –) 3-3-4-2 (based on the values shown in Table 2).

conditions presented in Table 2. Purities of products in outlet streams are used to verify SMBR performance.

Figure 5 describes effects of changing the switching time of external streams. An increase or decrease of the switching time interval leads to a decrease of the purity of both products—glucose and fructose. Only in a narrow range of switching time can one obtain the maximum purity for both products. The decrease of switching time produces an influence in the net flow of species in opposite direction to the fluid phase. This will affect the proper accomplishment of functions of each section in reactor. The effect on fructose purity in extract stream is larger than that on glucose purity in raffinate stream. In section 2, glucose that it should be desorbing is being taken toward "solid flow" and contaminating the extract stream. In raffinate stream, for this time, a decrease also occurs on glucose purity, due to the improper displacement of an amount of fructose from section 1 to section 4 until raffinate stream. In the last one, the effect is smaller because fructose must cross by section 1, in which there is a high flow rate, and section 4 to finally contaminate raffinate stream. It has been shown that the reactor configuration can be important to obtain maximum purity of the recovered species (17). For configuration 3-3-4-2 the purity of the recovered glucose is inferior to that for configuration 3-2-5-2 for higher switching times. This occurs in consequence of the decrease of the section 3 length—reaction zone—that does not adsorb the total amount of fructose produced before the end of this section.

In Fig. 6 the influence of extract and feed flow rate on the purity of the products is shown. In regard to the effect of changing the extract flow rate,

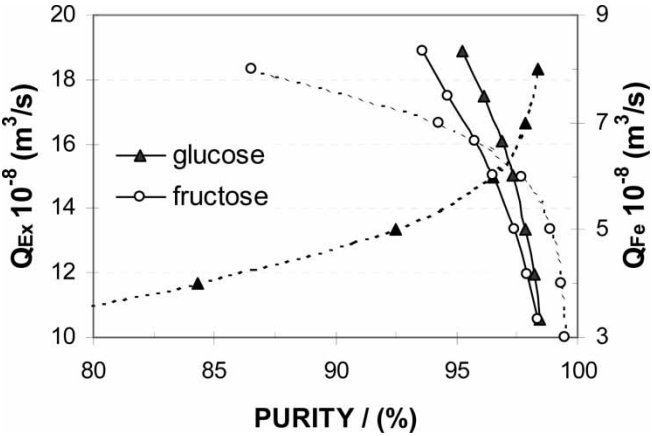

Figure 6. Effect of the (– – –) extract flow rate and (——) feed flow rate on product purity for SMBR in inversion sucrose process. (○) Fructose, (▲) Glucose (based on the values shown in Table 2).

one or the other product purity can be increased depending on extract flow rate magnitude. Interstitial velocities of fluid phase in sections 2 and 3 are changed when the extract flow rate changes. Lower extract flow rates mean higher interstitial velocities in these sections. So, in this situation, the species net flow in sections 2 and 3 is in direction of the fluid phase flow leading to lower glucose purity due to the undesirable displacement of fructose in direction to the raffinate stream. The concentration profile of sucrose is also moved forward in direction to fluid phase (homogeneous reaction) and the peaks of products will necessarily be advanced. Then, the length of the section 3 cannot be able to adsorb the whole fructose produced before the raffinate stream.

In Fig. 6, it is also possible to verify that higher feed flow rate provides lower purity of both reaction products. Higher feed flow rates lead to lower interstitial velocities in sections 1 and 2, and this will influence the species net flow in direction opposite to the fluid phase. So, in this time, the fructose purity in extract stream will be decreased due to contamination of glucose in this stream. Considering that the section used to regenerate the solid phase did not properly accomplish its function, the contamination of the raffinate stream by fructose is also observed.

The reaction of the sucrose should preferentially occur in section 3, and this will depend on the enzyme concentration. Or for constant enzyme concentration, the sucrose concentration is high enough to be completely consumed in section 3. Moreover, section 3 must be able to adsorb fructose so that it does not contaminate the stream that collects the glucose. Figure 7 presents the effect on performance of the reactor when the enzyme

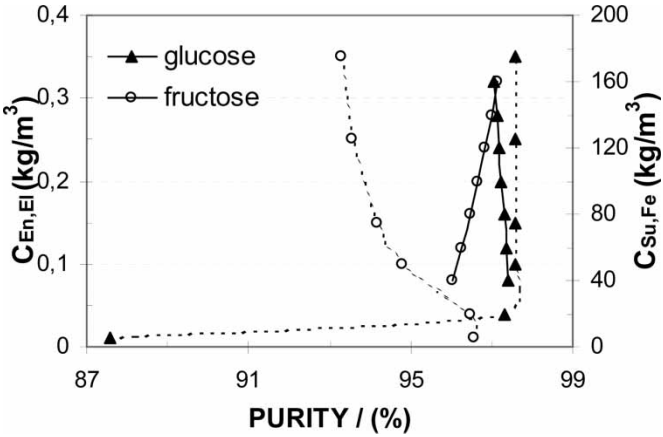

Figure 7. Effect of the (– – –) enzyme concentration—in eluent stream—and (——) sucrose feed concentration—in feed stream—on product purity for SMBR in inversion sucrose process. (○) Fructose, (▲) Glucose (based on the values shown in Table 2).

concentration in the eluent stream and the sucrose concentration in the feed stream are changed.

Figure 7 exhibits the influence of the amount of enzyme inside of the SMBR in the recovery of high-purity products. The smaller enzyme concentration in the reaction section, the slower the sucrose consumption. With this, the glucose and fructose profiles are taken in direction of the fluid flow. The section 3 will not be capable of adsorbing the fructose produced in the final length of the section, and it will contaminate the raffinate stream. However, the fructose is collected with larger purity since the glucose goes away from the collection point. There is tendency for a constant purity for each one of the species with the gradual increase of the enzyme concentration. This is due to the fact that the total amount of sucrose is consumed near the entrance of the feed stream to produce the sugars that will be separated as in a SMB unit. It should be noticed that the reaction rate is directly proportional to the enzyme concentration. For the sucrose concentration in the feed stream, the reply of the reactor is similar to the case of the analyzed enzyme. For a given amount of enzyme, larger sucrose concentrations in the feed correspond to larger demand in the consumption time of this reactant. Then, sucrose can be just totally consumed at the end of the reaction zone (section 3), not being possible to adsorb all produced fructose in time. So, this last can contaminate the raffinate stream. In some situations, sucrose will contaminate glucose due only to the partial consumption of the sucrose in the reaction zone. For the optimization of the sucrose inversion process in SMBR, one of the more important aspects is the interdependence between feed concentration of sucrose and enzyme, regarding productivity and costs of this unit.

CONCLUSION

The SMBR technology has been used in the enzymatic reaction and separation of sugars in the sucrose inversion process for obtaining high-sugar syrups. The process was mathematically modeled through conservation balance of the chemical species in SMBR in the phases: inter- and intraparticle fluid phase and solid phase. To solve the set of equations, it was considered both the actual operation system of the reactor and the model of TMBR equivalent.

The numerical concentration profiles obtained in this work were compared with experimental and numerical results from literature and they were shown to be in agreement. It was observed that the intermittent moving bed model reveals some important aspects in the evolution of the reaction process in SMBR, which cannot be possible to predict with a continuous moving bed model. It has been shown that this last one can be used to select operation conditions of a SMBR, but in start point of operation. To obtain an answer more faithful of the reactor, regarding to its performance, the intermittent moving bed model is the reliable option.

By numerical experimentation, it was possible to evaluate the influence of some operation parameters in the sucrose inversion process using a SMBR four sections; namely, flow rate of inlet and outlet streams of the reactor, switching time, unit configuration and reactant, and enzyme concentration in inlet streams. The analysis of the behavior showed that as well as in a SMB separation unit, the operation is quite sensitive to the changes in the operation conditions, being the reaction a factor of additional complexity. Once the definition of a strict purity specification, and then the adequate configuration, has been determined, the choice of the best conditions can be performed, based on process simulation, supported by SMBR package developed.

NOMENCLATURE

C	fluid phase concentration, kg/m^3
$\overline{C}_p$	average fluid phase concentration within pores, kg/m^3
D	axial dispersion coefficient, m^2/s
k_{fp}	global rate coefficient, s^{-1}
k_c	solid rate coefficient, s^{-1}
k_r	reaction rate constant, s^{-1}
K	equilibrium constant, dimensionless
K_M	*Michaelis-Menten* constant, kg/m^3
L_C	subsection length, m
m	operational parameter in moving bed units, dimensionless
Pe	number of *Peclet*, dimensionless
$\bar{q}$	mean solid phase concentration averaged over the particle, kg/m^3
q^*	mean solid phase concentration in equilibrium with average fluid phase concentration within pores, kg/m^3

Q	flow rate, m^3/s
Q'	flow rate in SMBR, m^3/s
v	interstitial fluid velocity, m/s
V_C	subsection volume, m^3
R	reaction rate, kg/m^3 s
U_k	generic velocity in Eq. (2), m/s
u_S	interstitial solid velocity, m/s
t^*	switching time, s
t	time, s
z	axial coordinate, m

Greek Symbols

ε	bed porosity, dimensionless
ε_p	particle porosity, dimensionless
ϑ	adsorbent/fluid volume ratio, dimensionless

Subscripts and Superscripts

El	eluent
En	enzyme
Ex	extract
Fe	feed
Fr	fructose
Gl	glucose
i	chemical species
j,k	SMBR or TMBR sections and subsections
Ra	raffinate
Su	sucrose

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