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Analysis of the high-fructose syrup production using reactive SMB technology

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Abstract

This paper deals with the production of high-fructose syrup (HFS) from glucose isomerization using reactive simulated moving bed technology. The reversible reaction is catalyzed by the immobilized enzyme glucose isomerase. In a simulated moving bed reactor (SMBR), reaction and separation processes can be coupled to achieve complete reactant conversion. The isomerization kinetics is experimentally determined at 328 K by the Lineweaver–Burk technique. Basic adsorption data for the sugar isomers (glucose and fructose) were obtained with cationic exchange resin as adsorbent. A mathematical model based on the analogy with true moving bed reactor and its numerical solution based on finite volume method were used for the prediction of the behaviour and performance of a SMBR. A new SMBR configuration for glucose isomerization is proposed and the effect of process parameters on its performance is addressed.

Keywords: Reactive simulated moving bed; High-fructose syrup; Isomerization kinetics; Sugars adsorption; Numerical simulation

1. Introduction

A mixture comprising glucose, fructose and minor amounts of oligosaccharides is known commercially as high-fructose syrup (HFS); typical syrups contain 42% and 55% fructose. HFS is an alternative to sucrose, since its sweetness is comparable to that of common sugar. HFS is used widely as nutritional sweetener (soft drinks, deserts, jams, dairy products) contributing for many useful physical and functional attributes of food and beverage applications, such as sweetness, flavour enhancement, colour and flavour development, freezing-point depression, and osmotic stability [1].

The fast growth in demand for fructose is attributed to several factors: it has a more refreshing flavour than that offered by sucrose, can be produced from starch (substrate available in food material) at lower cost and has the important advantage of lower risk for diabetic people or with some other metabolic disorders [2].

A typical process for production of fructose syrups uses alpha-amylase to liquefy starch and then glucoamilase to saccharify the hydrolyzed starch for the content of 94% dextrose. This resulting product will be directed for isomerization [3]. The isomerization process leads to a mixture of glucose and fructose. Nowadays the conversion of glucose to fructose has been subject of many studies in which the main objective is to maximize the yield and conversion of the isomerization reaction and to minimize costs of all types: additives, energy consumption, purification stages, among others.

Although different ways are presented in the literature for conversion of these isomers – application of acid solutions [4], strongly alkaline ion exchange resins [5,6], zeolites or hydrotalcites [7,8] – the use of the enzyme technology has been the major industrial application in the HFS production by isomerization of glucose to fructose. In this context, the enzyme glucose isomerase, mainly in immobilized form, has been used industri-

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Nomenclature

c	fluid phase concentration (mol/L)
d	subsection diameter (m)
D	axial dispersion coefficient (m^2/s)
E_T	total enzyme concentration (mol/g_{En})
K_e	reaction equilibrium constant
K_{mf}, K_{mr}	Michaelis Menten forward and reverse reaction constant (L/mol)
k_r	reaction rate constant (s^{-1})
k_T	global rate coefficient (s^{-1})
k_1, k_2	forward reaction rate constant (s^{-1})
k_{-1}, k_{-2}	reverse reaction rate constant (s^{-1})
L	subsection length (m)
m	operational parameter of moving bed units
N_C	number of subsections
q	average adsorbed phase concentration (mol/L)
q^*	adsorbed concentration in equilibrium with average fluid phase concentration within pores (mol/L)
Q	flow rate (mL/min)
r	reaction rate (mol/s g_{En})
u_S	interstitial solid velocity (m/s)
v	interstitial fluid velocity (m/s)
V_C	Subsection volume (m^3)
V_{mf}, V_{mr}	forward and reverse maximum reaction rate, respectively (mol/s g_{En})
t	time (s)
t^*	switching time (s)
W	mass of catalyst (g)
X	reaction conversion
z	axial coordinate (m)

Greek symbols

α	separation factor
ε	bed porosity
ε_r	reactor porosity
ϕ	parameter in Eq. (17)
ρ_{fb}	fixed bed density (g_{En}/m^3)

Subscripts

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

ally for operations with plug flow reactors (PFR) [9]. Some of the important parameters of this enzymatic isomerization process that have been discussed are temperature [10], pH [11], concentration of additives, substrate protection and by-product production.

The amount of fructose produced by the enzymatic isomerization reaction is related to the equilibrium constant of the reaction of glucose to fructose, which results approximately in an equimolar mixture of the sugars at 60 °C. For many commercial purposes, 42% fructose syrup is totally sufficient; though, for obtaining higher fructose syrup, adsorption columns packed with cationic exchange resins, or zeolites, in the Ca^{2+} form is frequently used. The simulated moving bed (SMB) technology has been successfully used for the separation of fructose and glucose since the coming of the process SAREX, one of the processes developed by Universal Oil Products using the SORBEX technology [12].

One common industrial practice is an isomerization reactor being fed with glucose and the reactor product directed to a SMB. This last operation unit will produce a rich-fructose (higher than 90% in purity) extract stream [13–16]. In raffinate stream, rich-glucose is recovered and can be recycled eventually (Fig. 1).

Obviously, the need for fructose enrichment plant from a mixture of sugars due to the reversible reaction occurring in reactor can be seen as an additional cost. At the present time, the enzyme technology faces the idea of 55% or higher fructose syrups production directly from enzymatic reactors as a challenge to be overcome.

On the other hand, the promising reactive simulated moving bed technology allows the combination of a chemical or biochemical reaction process and a chromatographic separation process in a single unit operation. The advantages of this integration were recognized in the early 1960s, when the first patents of the integration of chemical reaction and separation processes appeared [17].

In the field of the reactive chromatography, the chemical or the biochemical reactions are accomplished in the presence of a stationary phase, which separates the reaction products. In such a SMB reactor, reactions can occur either in the mobile phase or stationary phase. The consequences of the coupling reaction and separation are mainly that a high purity of the products can be obtained and, in situation limited by the chemical equilibrium, the reactions can be taken towards the complete conversion of the reagents. Ray et al. [18] discuss some other situations in which this integration can also be applied with advantages to exothermic and endothermic reactions. Other possible applications of the chromatographic reactors include the inhibitors removal, acceptors products, and catalytic poisons to yields improvement of the revenue of the main reaction [19].

Some classes of reactions for which the reactive chromatography with SMBR technology have been investigated are: alkylation [20], esterification and transesterification [21–23], hydrogenation [24,25], and also some enzymatic reactions [26–29].

The goal of this work is to analyze the combination of the enzymatic reaction of glucose isomerization with the isomers separation process in a single unit using the potentialities of the SMBR technology. A first research seeking the integration of reaction and adsorption for HFS production from glucose isomerization with SMB technology was made by Hashimoto et al. [26,27]. In that work an alternative SMBR configuration has been presented combining adsorbers and bioreactors in which HFS (45–65% in purity) can be produced starting from

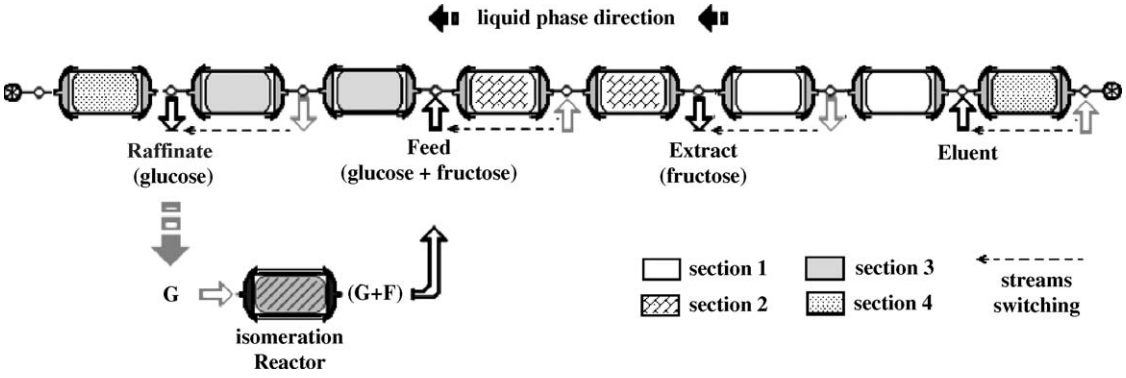

Fig. 1. Isomerization reactor before a simulated moving bed unit.

an equimolar mixture of sugars – glucose and fructose – with eluent consumption lower than in the traditional process (enzymatic reactor plus SMB). Zhang et al. [30] have presented an optimization study for the Hashimoto's SMBR system to obtain higher productivity of HFS (55% fructose) using minimum eluent. In that case, systematic multi-objective optimization was performed to improve system performance operated at different temperature and feed composition. Toumi and Engell [31] have investigated a SMBR process for glucose isomerisation in which the adsorbent and the catalyst are both in all columns.

In this work, a new SMBR configuration, whose conception is derived from Hashimoto's system, is proposed for operating this enzymatic isomerization and obtain fructose syrups. In this configuration, a glucose solution feeds the system in a way which is different from those reported in literature. Some experimental studies are described to measure the kinetics of the isomerization reaction in a stirred batch reactor and the adsorption kinetic and equilibrium of the sugars isomers on cationic exchange resins in a fixed bed column. The mathematical model of the SMBR system is presented and the numerical simulation is carried out using the measured and published parameters to discuss some aspects related to the SMBR configuration in HFS production. Simulation is also employed to determine a region spanned by flow ratios where high-fructose syrup is achieved. The effect of some important operating variables – streams flow rates and switching time – on the performance of the proposal SMBR configuration is addressed.

2. Mathematical model

Fig. 2 is an illustration of the SMBR system designed by Hashimoto et al. [26]. The reactors, packed with catalytic particles, accompany the switching of the external streams – feed, extract and eluent – in the switching time intervals. The unit has three sections and a maximum conversion of the glucose in Section 3 must be reached for not contaminating the product stream (extract).

To analyze the SMBR system of Fig. 2 and some of its variations, two models can be used to calculate concentration profiles and to obtain unit performances: the model of intermittent moving bed (considering the external streams and reactors switching) and the equivalence model to a true moving bed unit. In this analysis, the equivalence model has been chosen, but a comparison of results obtained for this system using these two strategies can be found in Borges da Silva et al. [32]. The SMBR technology is described as if a true flow of the stationary phase takes place in the unit: it requires smaller computational time. Moreover, it has been shown that this methodology is adequate to predict cyclic

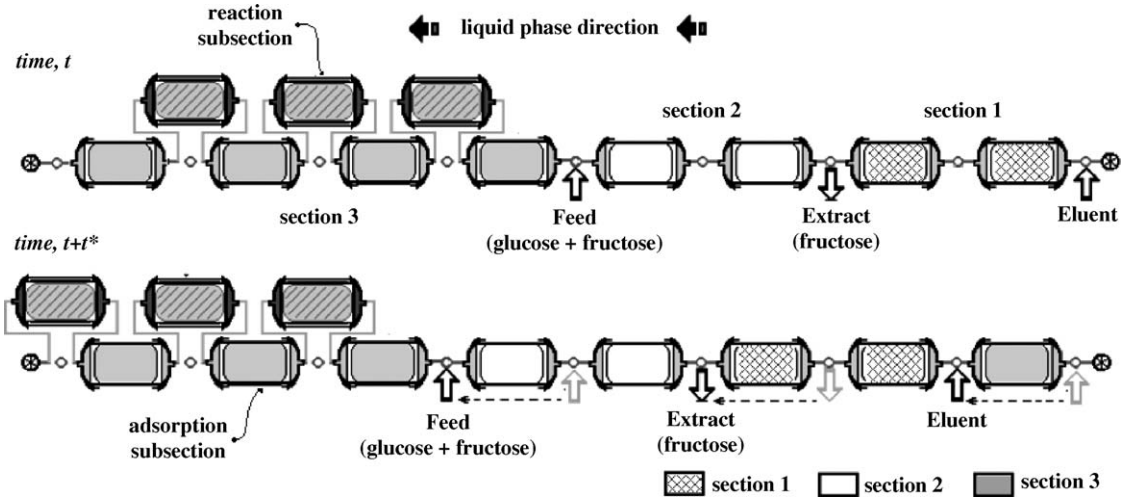

Fig. 2. System: SMBR unit combining adsorbers and bioreactors for glucose isomerization [26].

steady-state in processes where the unit sections contain three or more subsections [33,34]. The use of this approach requires that equivalence relationships of geometric and dynamic nature are obeyed. Particularly in the scheme of Fig. 2, the equivalence relationships should be determined with caution. The flow rate in reactors and adsorbers of Section 3 of the simulated moving bed reactor in relation to the flow rate in these subsections in an equivalent true moving bed reactor (TMBR) should be well observed for not altering residence times of the species in the reactive and adsorptive subsections. Lode et al. [35] have shown that the TMBR model cannot be applied to SMBR units, because the TMBR and SMBR units exhibit different residence time distributions and, hence, lead to different degrees of conversion.

In the current SMBR unit, the 'countercurrent movement' of the phases occurs only in adsorbers, while in reactors, which switch according to the external streams, the solid phase is not in movement, i.e., it is stationary in this operation process. In order to use a TMBR equivalent model to predict the behaviour and the performance of this configuration attention must be paid to the followed equivalences for subsections in Section 3: (i) for adsorbers: $v_k^{\mathrm{SMBR}} = v_k^{\mathrm{TMBR}} + u_{\mathrm{S}}$; (ii) for reactors: $v_k^{\mathrm{SMBR}} = v_k^{\mathrm{TMBR}}$.

The governing equations for the TMBR equivalent mathematical model consider the continuous fluid flow with axial dispersion for both bulk fluid phase in reactors and adsorbers, and the linear driving force mechanism to describe the intraparticle mass transfer rate in the adsorbers. This mathematical model is based on the following assumptions: the process is isothermal, the absence of radial gradients and velocities along a SMBR section are constant.

The mass balance equations for the species i in SMBR subsections are:

- Fluid-phase mass balance in a volume element of the subsection k

  Reactors

$$\frac{\partial c_{ik}}{\partial t} + v_k \frac{\partial c_{ik}}{\partial z} - D_k \frac{\partial^2 c_{ik}}{\partial z^2} = \left(\frac{1}{\varepsilon_{\mathrm{r}}} \delta_i \rho_{\mathrm{fb}}\right) r_{i,k} \tag{1}$$

  where $r_{i,k}$ represents a isomerization reaction rate and the value of δ_i must be -1 to glucose or $+1$ to fructose. It is supposed an insignificant resistance to the mass transfer of species between liquid and solid phases present in the reactors.

  Adsorbers

$$\frac{\partial c_{ik}}{\partial t} + \frac{1-\varepsilon}{\varepsilon}\left(\frac{\partial q_{ik}}{\partial t} - u_{\mathrm{S}} \frac{\partial q_{ik}}{\partial z}\right) + v_k \frac{\partial c_{ik}}{\partial z} - D_k \frac{\partial^2 c_{ik}}{\partial z^2} = 0 \tag{2}$$

- Particle mass balance

$$\frac{\partial q_{ik}}{\partial t} - u_{\mathrm{S}} \frac{\partial q_{ik}}{\partial z} = k_{\mathrm{T}}^{i}(q_{ik}^{*} - q_{ik}) \tag{3}$$

For adsorber solid phase, mass transfer resistances have been taken into account by a global rate coefficient k_{T}^{i} for the LDF approximation. The model allows using any adsorption equilibrium relation (q_{ik}^{*}).

Table 1
Variables of performance for a binary mixture in extract stream of a SMBR system

Purity (%)	$\frac{c_{\mathrm{F}}}{c_{\mathrm{F}} + c_{\mathrm{G}}} \times 100$
Eluent consumption (L/g)	$\frac{Q_{\mathrm{El}} + Q_{\mathrm{Fe}}}{c_{\mathrm{F}} Q_{\mathrm{Ex}}}$
Productivity (g/L h)	$\frac{c_{\mathrm{F}} Q_{\mathrm{Ex}}}{V_{\mathrm{S}}}$

Initial conditions:

$$t = 0: \quad c_{ik} = q_{ik} = 0 \tag{4}$$

Boundary conditions:

$$z = 0,\ t > 0: \quad c_{ik}|_{z=0^-} = c_{ik}|_{z=0^+} - \frac{D_k}{v_k} \left.\frac{\partial c_{ik}}{\partial z}\right|_{z=0^+} \tag{5}$$

$$\left.\frac{\partial (q_{ik})}{\partial z}\right|_{z=0} = 0 \quad \text{(except for reaction subsections)} \tag{6}$$

$$z = L,\ t > 0: \quad \left.\frac{\partial c_{ik}}{\partial z}\right|_{z=L} = 0 \tag{7}$$

$$q_{ik}|_{z=L} = q_{i(k+1)}|_{z=0} \quad \text{(for Sections 1 and 2)} \tag{8}$$

$$q_{ik}|_{z=L} = q_{i(k+2)}|_{z=0} \quad \text{(for adsorption subsections in Section 3; } k+2 = 1 \text{ when } (k+2) \text{ is higher than maximum number of subsections)} \tag{9}$$

For completing the model, mass balances at the nodes of the TMBR equivalent should be used in the boundary condition given by Eq. (5), in agreement with the location of the considered subsection (k).

- For columns inside of a section and for the extract node:

$$c_{i(k+1)}|_{z=0} = c_{ik} \tag{10}$$

- For the eluent node (with pure eluente):

$$c_{i(k+1)}|_{z=0} = \left(\frac{v_4}{v_1}\right) c_{ik} \tag{11}$$

- For the feed node:

$$c_{i(k+1)}|_{z=0} = \left(\frac{v_{\mathrm{Fe}}}{v_3}\right) c_{i\mathrm{Fe}} + \left(\frac{v_2}{v_3}\right) c_{ik} \tag{12}$$

The performance of the SMBR system for binary mixture is evaluated by purity, eluent consumption and productivity. These variables are calculated as shown in Table 1.

3. Experimental

3.1. Glucose isomerization reaction

The experiments of reaction kinetics of the glucose isomerization were carried out in batch jacket reactors. The catalyst

was a commercial Sweetzyme IT with particle size in the range 0.4–1.0 mm, supplied by Novozymes A/S (Bagsvaerd, Denmark). The Sweetzyme IT particle has a porous structure in which glucose isomerase enzyme is immobilized.

One of the possible reaction mechanisms to describe the kinetics of the glucose isomerization catalyzed by glucose isomerase enzyme is the reversible Briggs–Haldane mechanism [36], given for

$$G + E \underset{k_{-1}}{\overset{k_1}{\rightleftharpoons}} GE \underset{k_{-2}}{\overset{k_2}{\rightleftharpoons}} F + E$$

where parameters k are rate constants of the elementary reactions.

The global reaction rate of the glucose isomerization based on the reversible Briggs–Haldane mechanism is expressed as

$$r = E_T \frac{k_1 k_2 c_G - k_{-1} k_{-2} c_F}{(k_{-1} + k_2) + k_1 c_G + k_{-2} c_F} = \frac{K_{mr} V_{mf} c_G - K_{mf} V_{mr} c_F}{K_{mr} K_{mf} + K_{mr} c_G + K_{mf} c_F} \quad (13)$$

where Michaelis constants K_m and maximum velocities V_m of the forward and reverse reaction are

$$K_{mf} = \frac{k_{-1} + k_2}{k_1}, \qquad V_{mf} = k_2 E_T \quad (14)$$

$$K_{mr} = \frac{k_{-1} + k_2}{k_{-2}}, \qquad V_{mr} = k_{-1} E_T \quad (15)$$

The equilibrium constant that indicates a favorable toward to the enzymatic reaction – forward or reverse – is

$$K_e = \frac{k_1 k_2}{k_{-1} k_{-2}} = \frac{K_{mr} V_{mf}}{V_{mr} K_{mf}} \quad (16)$$

The kinetic parameters of the global reaction rate can be obtained with Lineweaver–Burk method and mass balance of the species in the reactor.

Glucose and fructose solutions were prepared in concentrations of 0.5, 1.0, 2.0 and 3.0 mol/L. In these reagent solutions was added 20 g/L heptahydrated magnesium sulphate ($MgSO_4 \cdot 7H_2O$) for enzyme activity and stability. At first, all the solutions were degassed for, at least, 2 h to guarantee that the dissolved air did not interfere in the reaction for obtaining of the initial reaction rate. It was used a 0.05 mol/L tris buffer solution, so that the final pH of the reagent solution kept between 7.7 and 7.8 (at 298 K).

A volume of 60 mL of reagent solution and 1 g of Sweetzyme IT was used in each experiment. The reactors were stirred at 150 rpm and temperature was maintained at 328 K. Samples of 50 μL were withdrawn from the reactor at specific time intervals and diluted to a measurable concentration range in a RI (Refraction Index) detector (Gilson, model 131). The fructose and glucose concentrations were determined by HPLC (high performance liquid chromatography) using the RI detector.

The obtained kinetic parameters are shown in Table 2. The value of K_e, equilibrium constant, indicates a slightly favorable direction towards the fructose production under employed experimental conditions. In Table 2, it can be seen some kinetic parameters of glucose isomerization already published. Although the experimental conditions were different and the enzyme used in this work has higher activity, the parameters are comparable, confirming the validity of the procedure adopted for estimate of the isomerization kinetics.

For a batch reactor, the mass balance could be written in terms of conversion as

$$\frac{W}{V} t = \left(\frac{\phi_3}{\phi_2} + \frac{\phi_1 \phi_4}{{\phi_2}^2} \right) \ln \left(\frac{\phi_1}{\phi_1 - \phi_2 X} \right) - \frac{\phi_4}{\phi_2} X \quad (17)$$

where W is the catalyst mass in the reactor and X is the reactant conversion. The ϕ parameters are

$$\phi_1 = \frac{V_{mf} K_{mr} c_{Go} - V_{mr} K_{mf} c_{Fo}}{c_{Go}}, \qquad \phi_2 = V_{mf} K_{mr} + V_{mr} K_{mf} \quad (18)$$

$$\phi_3 = \frac{K_{mf} K_{mr} + K_{mr} c_{Go} + K_{mf} c_{Fo}}{c_{Go}}, \qquad \phi_4 = K_{mf} - K_{mr} \quad (19)$$

Since in the application of the Linewaever–Burk plots technique only initial experimental points of the concentration evolution as a function of time are considered in the evaluation of the kinetic parameters, Eq. (17) is used to obtain these parameters using all the points of the experimental curves. In order to obtain an initial estimation of the kinetics parameters Linewaever–Burk technique is used. A nonlinear regression method is performed, specifying the function of Eq. (17), such that the goal of the regression was to determine the values of the parameters that minimized the sum of the residual values for the set of experimental data. In this procedure, the equilibrium constant is kept equal to 1.04 (value from experimental results).

Fig. 3 presents some experimental results and predicted profiles by simulation using the global reaction rate – Eq. (13) – with the parameters listed in Table 3. As illustration, both glucose and fructose production profiles are shown for the isomerization reaction using initial feed reagent solution of 0.5 and 1.0 mol/L

Table 2
Kinetic parameters of the glucose isomerization

Kinetic parameters	V_{mf} (μmol/min g_{En})	V_{mr} (μmol/min g_{En})	K_{mf} (mol/L)	K_{mr} (mol/L)	K_e
Chen and Wu[a] [11]	2.92×10^2	1.87×10^2	0.70	0.45	0.99
Convert and Borghi[b] [36]	3.95×10^2	2.59×10^2	0.70	0.45	0.98
This work[c]	3.15×10^2	2.76×10^2	0.26	0.24	1.04

[a] Chen and Wu [11]—Swetase enzyme (250 IGUI/g); $T = 333$ K; pH 8.25.
[b] Convert and Borghi [36]—Sweetzyme T enzyme (350 IGUI/g); $T = 333$ K; pH 8.25.
[c] Present work—Sweetzyme IT enzyme (400 IGUI/g); $T = 328$ K; pH 7.7–7.8.

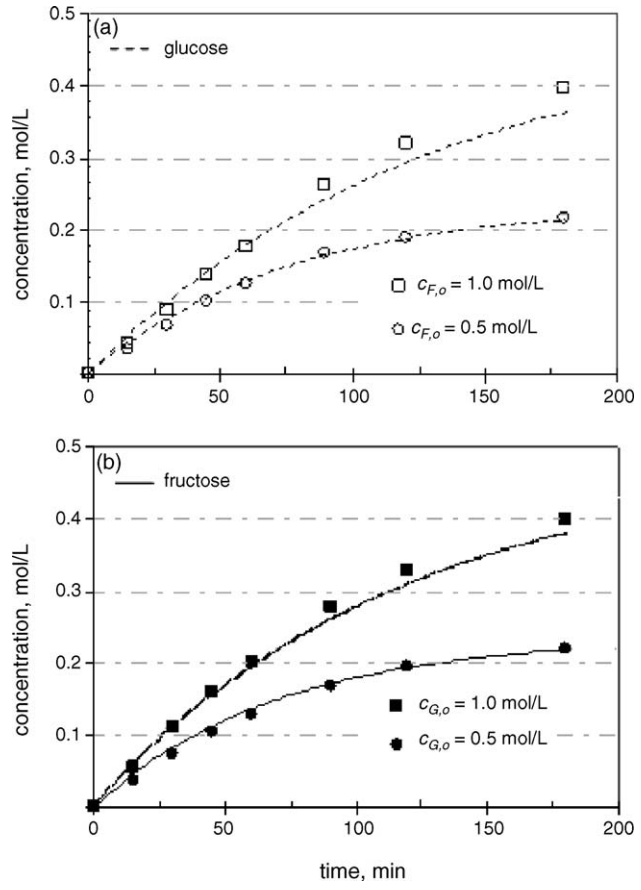

Fig. 3. Product concentration profiles of isomerization reactions catalyzed by immobilized glucose isomerase (Sweetzyme IT) with initial sugar solution of 1.0 and 0.5 mol/L for each case. Production of: (a) glucose—reverse reaction; and (b) fructose—forward reaction. Reaction conditions: 328 K and pH 7.7.

concentration. The calculated curves represent the experimental behaviour fairly well, and then the obtained global reaction rate can be useful in predicting the glucose isomerization kinetics.

The influence of the resistance to the external and internal mass transport of the sugars in the conversion of this enzymatic reaction is shown in Borges da Silva et al. [37]. It has been experimentally verified that, under the operating conditions applied in this work, the diffusion of species in both external film and intra-particular region in the porous particles has not significant effect in the reactive process.

3.2. Isomers separation

The measurement of single-component breakthrough and elution curves of the species allow the knowledge and determination of the adsorption equilibrium isotherms and adsorption kinetics. The adsorption parameters were determined by the chromatographic method of the frontal analysis (FA) carried out in a jacket fixed bed column (column Merck Superformance 26–30 cm length × 2.6 cm i.d.).

The concentration range of 0.2–1.0 mol/L was investigated at 328 K, the same temperature used to obtain the reaction. The adsorbent was a cationic exchange resin (supplied by FINEX Co., Finland) in the Ca^{2+} and Mg^{2+} forms (358 μm size). In the experiments in which the cationic resin was used in commercial form – Ca^{2+} form – the mobile phase was a 0.05 mol/L tris buffer solution (pH 7.7–7.8 at 298 K); though, in those experiments in which the resin was in Mg^{2+} form, 20 g/L heptahydrated magnesium sulphate was added to the mentioned mobile phase. The adsorption experiments have been performed at feed flow rate 5 mL/min and the height of the bed was 11.2 cm for resin in Ca^{2+} form and 11.8 cm for resin in Mg^{2+} form. The fructose and glucose concentrations were measured by analysis in HPLC using a RI detector.

Fig. 4 shows the adsorption equilibrium isotherms for both ionic forms of the exchange resin. In the experimental conditions used in this work, both sugars are linearly adsorbed in the analyzed concentration range. Some published works report non-linear isotherms for these sugars isomers in systems operating at high liquid concentration [38–40]. Concerning to the glucose and fructose separation, the adsorption data are described in Table 3. The adsorption kinetics are represented by a global mass transfer coefficient for each species. The suggested kinetic adsorption is fitted to the experimentally measured breakthrough curves. The breakthrough profiles are calculated using an axial dispersed plug flow model to the liquid phase and it is assumed mixed diffusion control of micro-pores and macro-pores—linear driving force approximation (LDF). The estimation of k parameters in LDF model is found from a 'best fit' procedure in which the sum of residual between the experimental and calculated

Table 3
Summary of adsorption data for glucose–fructose mixtures on various kinds of adsorbents

Reference	T (K)	Adsorbent	Eluent	K_G/k_T^G (min^{-1})	K_F/k_T^F (min^{-1})
Hashimoto et al. [26]	323	Zeolite Y: Ca^{2+}, Mg^{2+}	Buffer solution (20 mol/m^3 *tris* HCl + 10 mol/m^3 $MgSO_4$)	0.586/0.41[a], 0.423/–	0.686/0.41[a], 0.488/–
Ruthven and Ching [39] (c_i = up to 1.8 mol/L)	328	Zeolite Y Ca, Ca^{2+} resin	Water	0.365/–, 0.067/–	$0.675–5 \times 10^{-4}\, c_G - 4 \times 10^{-3}\, c_F$, $0.12 + 2 \times 10^{-3}\, c_G + 1 \times 10^{-3}\, c_F$
Azevedo [41] (c_i = up to 0.17 mol/L)	303, 323	Dower Monosphere 99/Ca	Water	0.28/1.06, 0.27/1.89	0.6/0.75, 0.53/1.33
Altenhoner et al. [42] (c_i = up to 2 mol/L)	312	Imac HP 1320 Resin (Ca^{2+})	Water	0.34/–	0.56/–
This work (c_i = up to 1 mol/L)	328	FINEX CG11CS Resin: Ca^{2+}, Mg^{2+}	Solution tris 0.05 M, solution tris 0.05 M + $MgSO_4 \cdot 7H_2O$ 20 g/L	0.319/2.92, 0.295/9.45	0.482/2.25, 0.354/8.85

FINEX resin: Ca^{2+} Form, $\alpha_{F/G} = 1.51$; Mg^{2+} Form, $\alpha_{F/G} = 1.20$.
[a] Rate expression based on fluid concentratrion.

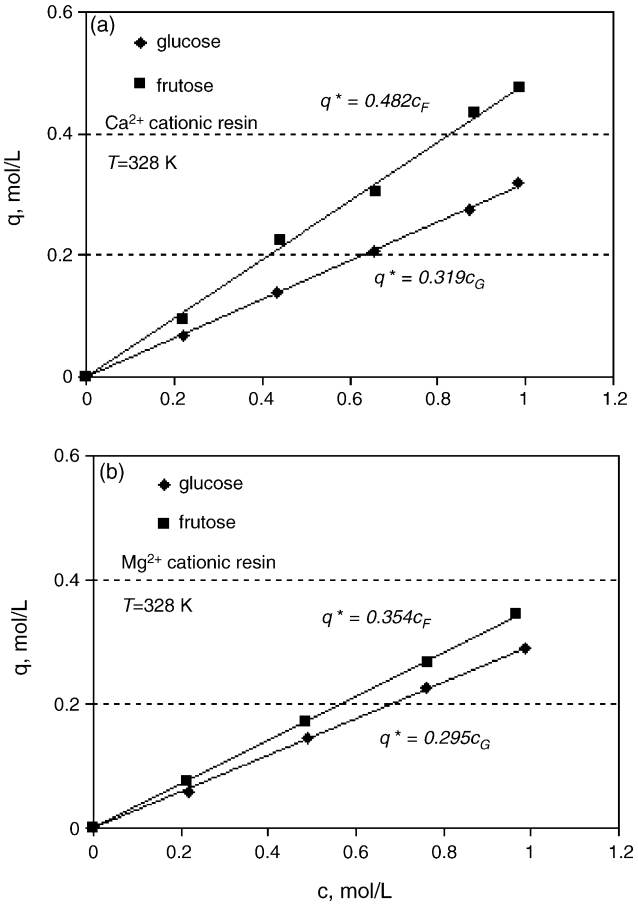

Fig. 4. Adsorption isotherms for fructose and glucose at 328 K: (a) resin in Ca^{2+} form; (b) resin in Mg^{2+} form.

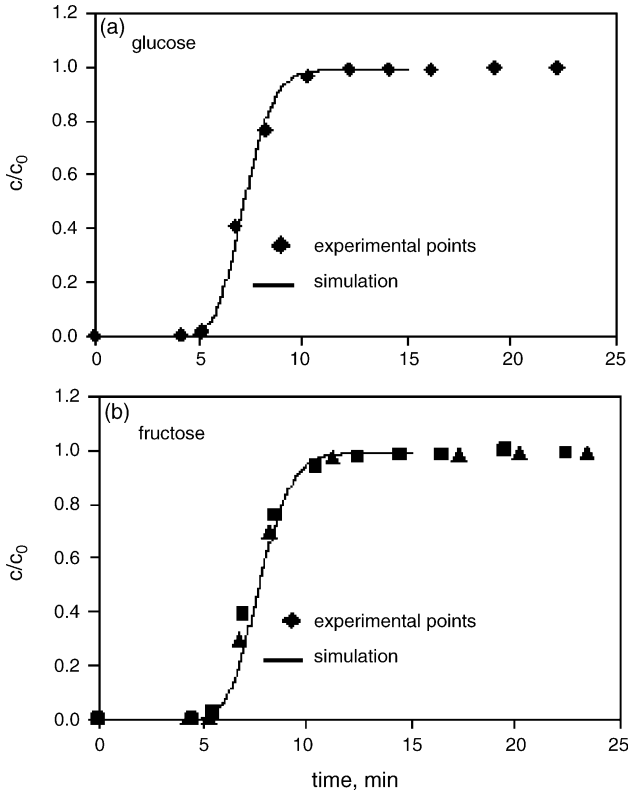

Fig. 5. Experimental data (symbols) and numerical simulation (solid line) of the breakthrough curves (dimensionless concentration vs. time) at 328 K: (a) glucose (c_{Go} = 0.219 mol/L –♦–) and (b) fructose (c_{Fo} = 0.216 mol/L –■–; 0.989 mol/L –▲–). Feed flow rate: 5 mL/min. Adsorbent: cationic resin in Mg^{2+} form.

concentration have been minimized. Fig. 5 shows the adsorption data (symbols) for glucose and fructose from the FA measurements to the cationic exchange resin in Mg^{2+} form (at 328 K). The experimental data fit well to the model of adsorption kinetics.

4. Results and discussion

4.1. Validation of the numerical methodology

The focus of this work is the fructose production in high-purity (HFS) by operating a reactor using the technology of simulated moving bed for glucose isomerization. So, only one device would be operating the reaction and separation stages simultaneously. The reaction of glucose isomerization is catalyzed by the action of the enzyme glucose isomerase (Sweetzyme IT). According to the supplier [43], this immobilized enzyme has an optimum performance in a range of temperature from 328 to 333 K (pH 7.2–7.5). The enzyme performance means a trade-off between high activity and high stability. So, the adsorption and desorption processes of both sugars should also occur in this temperature range in the unit. In general, the adsorbents used to separate these sugar isomers are zeolites *Y* (or *X*) or ionic exchange resins, both in the Ca^{2+} form, in fact due to the property of fructose to complex with calcium ions. In the isomerization reactor, the magnesium ion activates and stabilizes enzyme glucose isomerase and a suitable amount of this ion should therefore be present in the feed solution. On the other hand, the calcium ions act as an enzyme inhibitor by displacing the magnesium ion activator from isomerase molecule, interfering in the yield of the reaction (Novozymes A/S). In order to carry out the isomerization process in one device only, it is necessary to find an adsorbent – different of the zeolite or resin in Ca^{2+} form – for the separation of the sugars, since the Mg^{2+} ion can exchange with the Ca^{2+} ion, and have a detrimental effect on the reaction stage. In a previous study [26], it was shown to be possible to use the mentioned adsorbent not in Ca^{2+} form, but in the Mg^{2+} form. The separation factors for both ionic forms are different and relatively smaller for the magnesium form—as it can be verified by the analysis of Table 3. However, the inhibition problem of the glucose isomerase in the reactions for the presence of calcium ion is outlined.

In application of the SMBR technology in the glucose isomerization, let us take into consideration the configuration of the system suggested by Hashimoto et al. [26,27]—Fig. 2. It is known that the efficiency and successful operation of units using the SMBR technology are extremely dependent on the correct choice of operational conditions, particularly of the flow rates in each one of the unit sections and of the switching time ('velocity of stationary phase').

Table 4
Operating conditions and parameters of the model to the glucose isomerization in a SMBR [26,46]

Operational conditions		Parameters	
c_{Fo}	1.0 M	k_T	0.41 min^{-1}
c_{Go}	1.0 M	K_F	0.686
Q_{Fe}	0.14 mL/min	K_G	0.586
Q_{El}	0.43 mL/min	k_r	0.0617 min^{-1}
Q_3	4.00 mL/min	N_C	21 (8*A*-7*R*-3-3)
t^*	3.0 min	L_C	10.2 cm
ε; ε_r	0.4	d	1.38 cm

Kinetic reaction [46]: $r_i = k_r(c_G - c_F)$.

Regarding the separation regions for SMB units it is known that [44,45]:

$$m_j = \frac{Q_j^{SMBR} t^* - \varepsilon V_c}{(1-\varepsilon)V_c} = \frac{Q_j^{TMBR}}{Q_s} \quad (20)$$

where m_j is the flow rate ratio parameter representing the ratio between the net fluid flow rate and the solid flow rate in the jth section. Flow constraints can be introduced as

$$m_1 = \beta_1 K_F \quad (\beta_1 > 0) \quad (21)$$

$$K_G\beta_2 = m_2 < m_3 = \beta_3 K_F \quad (\beta_2 > 0;\ \beta_3 < 0) \quad (22)$$

where β are safety margin parameters.

These flow constraints were used by Hashimoto et al. [26] when they presented simulation studies for the isomerization process in the system of Fig. 2. Later Ching and Lu [46] shown that these constraints do not need to be strictly satisfied because it is not just the separation that is taking place in unit but also the reaction. Then, with a new performance parameter to be increased – the conversion, some restrictions in Eqs. (18) and (19) can be violated. The parameters used in the experimental and numerical simulation studies of Hashimoto et al. [26] and Ching and Lu [46] are described in Table 4.

Based on the operating conditions given in Table 4, one can compare in Table 5 the values of the net flow parameters presented by Hashimoto et al. [26] and those by Ching and Lu [46], when the flow restriction in Section 2 is violated.

Table 5
Values of the flow parameters for operation of the system illustrated in Fig. 2

	No. of section		
	1	2	3
Hashimoto et al. [26]			
m_j	0.785	0.598	0.645
Ching and Lu [46]			
m_j	1.096	0.481	0.609

The simulated profiles of average concentration of the sugars along the SMBR sections, in the process of glucose isomerization, using the TMBR equivalent model, are presented in Fig. 6. The model equations were discretized by using the finite volume method [47] and computational algorithms have been developed in FORTRAN language. Modified strongly implicit method (MSI) is used to solve ODE systems [48]. In this development it is also used the co-localized variable arrangement in computational grids and interpolation functions from WUDS scheme (weight upstream differencing scheme) [49]. The convergence of the solution is evaluated by increasing the number of control volumes and decreasing the integration time until no change is observed between different runs. A typical computational grid used in the simulations was 100 control volumes in axial direction with time integration step of 1 s. Run times were typically 5–10 min in a Pentium IV 2 GHz processor.

According to Fig. 6, a good agreement is verified between simulated profiles, calculated by Ching and Lu [46] and in this work, validating the numerical methodology development. The unit performance is evaluated through the purity of the fructose collected in extract stream and it was 53.3%. In Fig. 6, an undesirable reaction of fructose isomerization into glucose occurs at the beginning of Section 3 [46]. However, with the conditions suggested by Ching and Lu [46], this situation in Section 3 does not happen, according to the results shown in Fig. 7, where the product recovered in extract was fructose with 55.2% purity.

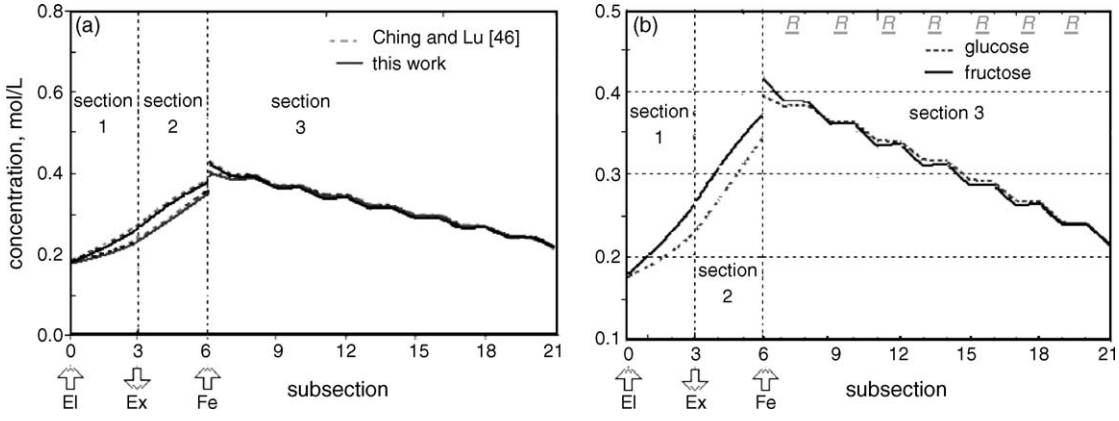

Fig. 6. (a) Comparison between average concentration profiles in SMBR at cyclic steady state: (- - -) Ching and Lu [46]; (—) this work. (b) Average concentration profiles of glucose and fructose in this reactive and separative process. Purity (Ex): 53.3%.

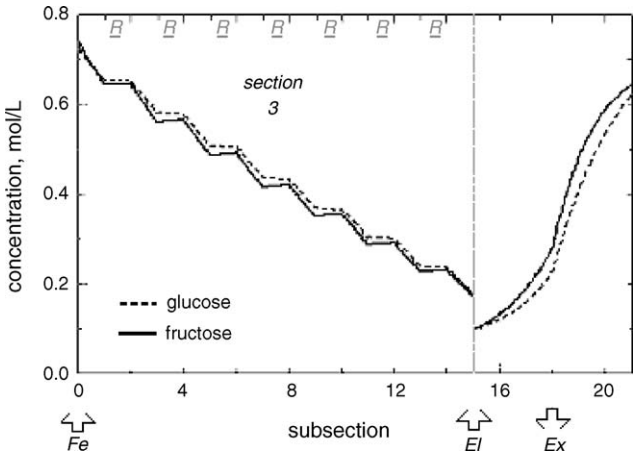

Fig. 7. Concentration profile in SMBR. Operation conditions: Table 4, except $Q_{Fe} = 0.5$ mL/min; $Q_{El} = 1.48$ mL/min; $Q_3 = 3.9$ mL/min. Flow parameters proposed by Ching and Lu [46]. Purity (Ex): 55.2%.

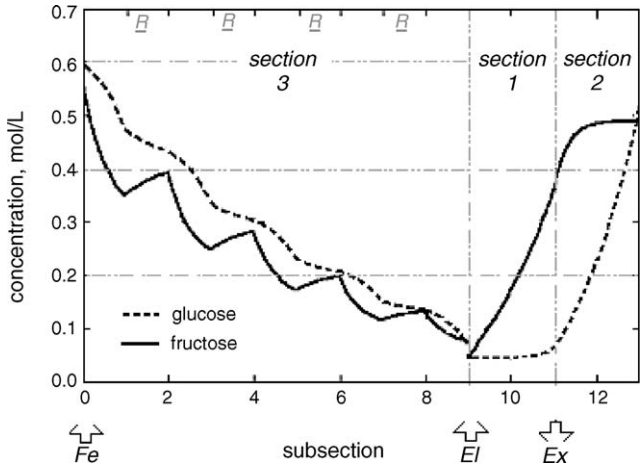

Fig. 8. Concentration profile in SMBR. Operating conditions listed in Table 6. Purity (Ex): 85.0%. Resin in Ca^{2+} form.

Table 6
Operating conditions and model parameters to the glucose isomerization in a SMBR

Operational conditions[a]	
c_{Fo}	1.0 M
c_{Go}	1.0 M
Q_{Fe}	1.50 mL/min
Q_{El}	5.60 mL/min
Q_3	25.5 mL/min
t^*	3.84 min
ε; ε_r	0.4
N_C	13 (5*A*-4*R*-2-2)
L_C	30 cm
d	2.6 cm

[a] Form for resin in Ca^{2+} form.

4.2. A new SMBR configuration

The operating conditions suggested for the SMBR illustrated in Fig. 2 are listed in Table 6. In order to define adequate operating conditions for the SMBR the constraints of the equilibrium theory are used, formulated for linear equilibrium non-reactive case [50]. However, in view of the operation of the SMB equipment existing in our laboratory, some of the equipment limitation have to be taken into account, such as total number of columns (12) and recycling pump working within the range of 20–120 mL/min.

In this SMBR system, the lowest flow rate occurs in Section 2 and then, to overcome any difficulty related to the mass transfer resistance, it is assumed 24 mL/min. By imposing a safety margin β_1 of 1.1, the switching time can be calculated from Eq. (20). All the other section flow rates are calculated using Eq. (20). The inlet – feed and eluent – and collect streams are calculated from the differences between neighbouring section flow rates. The reaction rate is given by Eq. (13) with the kinetic parameters presented in Table 2. The adsorption parameters are given in Table 3.

The analysis of Fig. 8 shows that glucose is transformed into fructose in fixed bed reactor, and then the species enter an adsorption column where a differential of concentration difference will be created for the mixture of the isomers entering the next reactor. The equilibrium situation is practically reached in the end of the last fixed bed reactor. At the end of Section 3, the glucose concentration is sufficiently low so that the liquid stream can be recycled to Section 1, without problems of high contamination of the fructose recovered in extract stream. The purity of the product obtained with these conditions described in Table 6 was 85.0% in fructose.

At this time, the question to be discussed should not concentrate on the largest purity found, because the reaction kinetics and adsorption parameters are different from those applied in published works. Our attention is focused on the potential that this technology can offer. It could be interesting to operate the

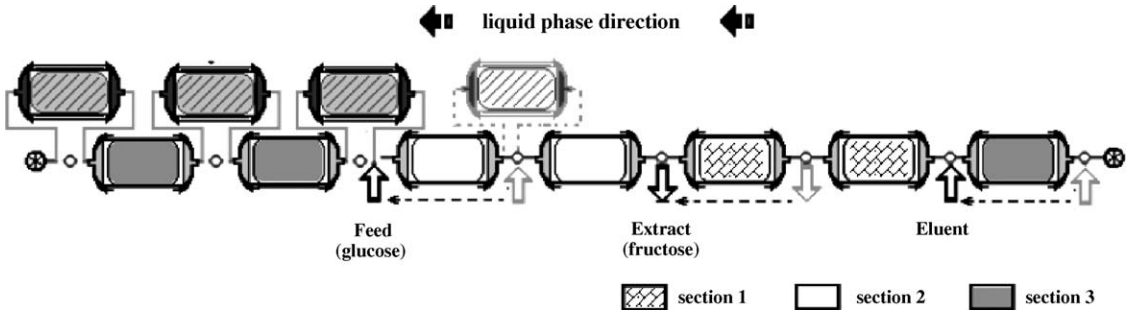

Fig. 9. New proposal of SMBR configuration for glucose isomerization.

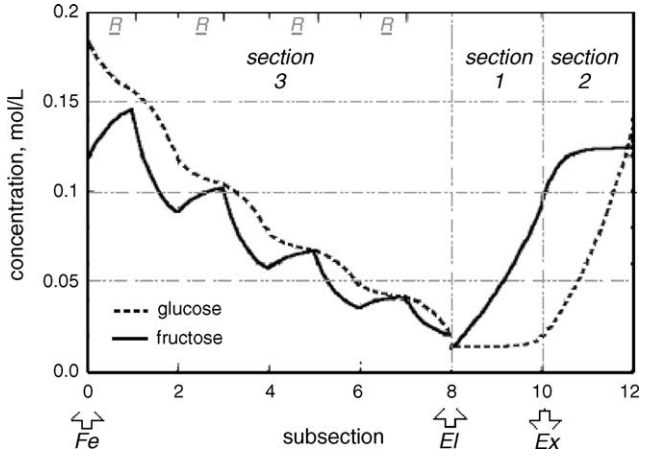

Fig. 10. Concentration profile in the new arrangement of SMBR. Feed stream: solution of glucose 1 mol/L. Pu_{Ex}: 83%. Operating conditions: $Q_{Fe} = 1.5$ mL/min; $Q_{EI} = 5.6$ mL/min; $Q_{Ex} = 7.10$ mL/min; $t^* = 3.84$ min. Operation parameters: $(m_2; m_3) = (0.298; 0.358)$.

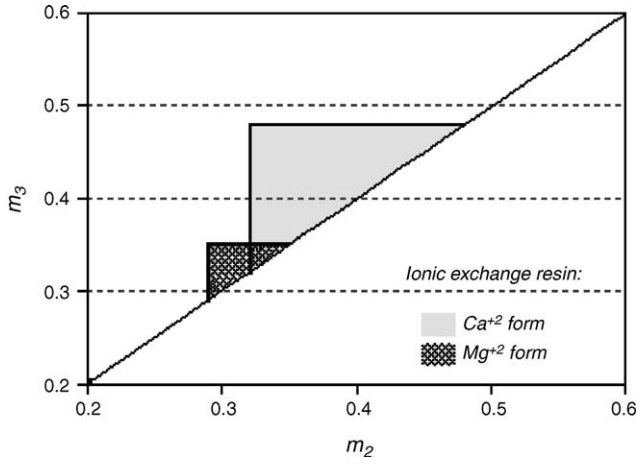

Fig. 11. Triangular region of complete separation of glucose and fructose – equilibrium theory – for a SMB packed with ionic exchange resin in Mg^{2+} and Ca^{2+} form.

system of Fig. 2 not only in view of eluent economy, as it is discussed in the literature, but instead operating with a feed containing only glucose and not a mixture of glucose and fructose coming from previous reactor. This way, the need of the enzymatic reactor preceding the SMBR unit would be eliminated. Hence, the SMBR unit should have, after the feed stream, not an adsorber subsection, but a reactor, as suggested in Fig. 9.

Fig. 10 presents the behaviour of the new SMBR configuration proposed here for a feed of glucose 1 mol/L. The operating conditions are listed in Table 6, except that now the number of columns in Section 3 is 12. It can be verified in Fig. 10 that the equilibrium condition between the sugar isomers in the last reaction subsection is reached. The consumption of glucose is enough to not damage the product purity in extract. The fructose purity obtained for this case was 84%, and the product appears in low concentration.

4.2.1. Triangular region (m_2; m_3) to glucose isomerization in SMBR system

For the convenient operation of the reactive SMB unit illustrated in Fig. 9, the separation of glucose and fructose on the ion exchange resin in Mg^{2+} form was used. This is because the presence of Ca^{2+} ions inhibits the action of the glucose isomerase enzyme in the conversion of glucose. Then, with the adsorption data obtained for the cationic resin in Mg^{2+} form, the operation conditions were designed and a new set of model parameters were introduced in this numerical investigation. The flow constraints for using this last adsorbent are given by

$$m_1 > 0.354 \tag{23}$$

$$0.295 < m_2 < m_3 < 0.354 \tag{24}$$

It is important to have in mind that operation restrictions imposed in the SMB separation unit do not always need to be respected for the operation of a reactive SMB. Fig. 11 shows the triangular region, according to the equilibrium theory [44,45,50], where a simple separation of the species would be possible using the cationic resin in Mg^{2+} form.

In order to achieve the operation conditions that can be applied in the proposal reactive SMB unit for reaching HFS, some simulations are accomplished and the results are shown in a $(m_2; m_3)$ plan. This $(m_2; m_3)$ plot is an important tool in the choice of best operating condition. Since Section 1 is an adsorbent regeneration section, in which no reaction happens, the value of m_1, and consequently of the flow rate in Section 1 (maintained the switching time), should be high enough so that it guarantees the desorption of the retained species—fructose. The operating conditions that are possible to obtain HFS could be, therefore, represented in a $(m_2; m_3)$ plan for a given m_1 defined, or in a volumetric region in which m_1 assume different values—three-dimensional graph in $(m_1; m_2; m_3)$ space [41]. The plan or volumetric region is determined for one switch time interval. The switching time has crucial function in adequate operation of the unit, interfering in the net liquid flow of the species in each one of the SMBR sections.

The flow rate in Section 1 is chosen to be 29.3 mL/min (having set m_1 as 0.52 based on the restriction of Eq. (23)) and, keeping the value of β_1 as before, the switching time was equal to 3.87 min. Fig. 12 presents areas in m_2 versus m_3 plot in which there are different process performances in the SMBR unit. The m_2 and m_3 points are defined in way they are on parallel lines to diagonal of the plan that defines the condition of minimum feed flow rate ($m_3 \rightarrow m_2$) [51]. The concentration of glucose in feed stream is 1 mol/L.

For the construction of Fig. 12 six parallel diagonals were used starting from the line defined by $m_2 = m_3$. The lines far from this diagonal line mean that higher feed flow rate can be processed and higher productivity. The $(m_2; m_3)$ coordinate in the region indicate different unit performances in obtaining high fructose in extract. The points represented by (■) and (×) provide operation conditions to collect a product of purity higher than 55% in fructose, while the points (◯) a product of purity lower than 55% in fructose.

From the results shown in Fig. 12, one can observe that outside the area of complete separation, in the case of a separative SMB, it is possible to obtain products with certain degree of

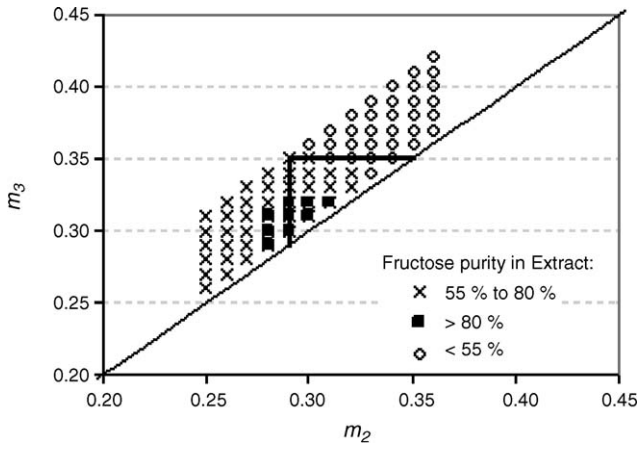

Fig. 12. Plan of (m_2; m_3) for the process of glucose isomerization using the resin in Mg^{2+} form. Fructose in extract stream with purity: (×) higher than 55% and lower than 80%; (■) higher than 80%; (◯) lower than 55%. Conditions: $m_1 = 0.52$; $t^* = 3.87$ min and $c_{Go} = 1$ mol/L.

purity (larger than 85%), taking into account the specifications of purity requested for commercial HFS (for example, 55% and 90%). It should be remembered that the equilibrium theory does not include the resistances to the mass transfer of the species and that the area of complete separation for non-reactive SMB, considering the mass transfer resistances is more reduced [52].

Fig. 13 presents areas of the (m_2; m_3) plan corresponding to operational conditions for which the purities of 72% and 90% in fructose can be reached in extract stream. The values of the switching time and m_1 parameter are 3.87 and 0.52 min, respectively. The other parameters are listed in Table 6. The feed stream of the unit is only composed of glucose solution 1 mol/L.

4.3. Discussion of process characteristics

In Fig. 13 it has been disclosed that it is possible to use the system shown in Fig. 9 for glucose isomerization to obtain high purity fructose. The analysis is made using experimental data from reactive and separation processes taken separately and then, by mathematical modelling and simulation, these

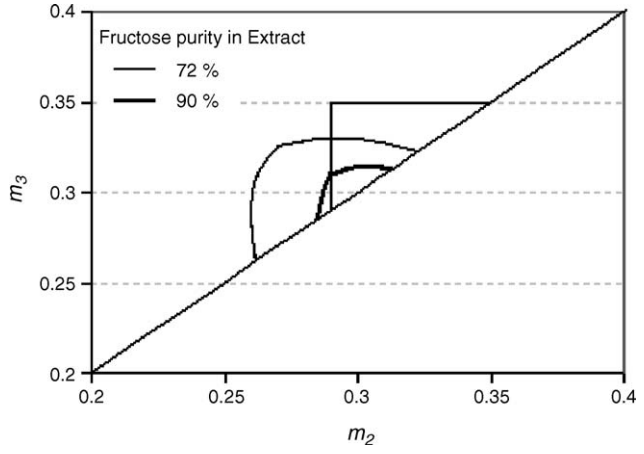

Fig. 13. Region of the (m_2; m_3) plan to obtain fructose purities of (——) 72% and (▬▬) 90% for glucose isomerization in SMBR (system of Fig. 9).

processes are grouped in a single operation unit. Considering industrial information, some important aspects related to HFS production employing the glucose isomerase enzyme and further purification stage for adsorption columns must be addressed to allow practical application. When entering the system (first reactor), the feed solution (glucose) should be refined to remove any particulate matter which can plug the columns or soluble impurities – calcium ions, peptides, oxygen, oxidant products – which may inhibit the enzyme [2,53]. Therefore, the solution must be filtered, carbon treated, and ion exchange before entering the first reactor of the system. However, the other reactors into Section 3 of the system will receive effluent from others subsections – adsorbers – that may not eventually meet the same specifications standardized in the beginning. Additionally, even adsorption subsections (ionic exchange columns) will be fed with untreated effluent proceeding directly from reactors. Actually, in the current industrial units, which combine processes are in series, the effluent from reactors is directed for separation unit only after some cleaning treatments with anionic and cationic exchange resins and carbon are carried out to desalts and to decolorize the reaction product [3].

The problem of by-products (salts, organic acids, proteins) and colour formation is linked to the temperature, pH and duration of the isomerization process. The quality of the final product – high-fructose syrup – suffers a negative effect when the solution remains during long time in conditions as alkaline pH-values and elevated temperatures [54]. In batch isomerization in which normally longer reaction time is required compared to continuous isomerization, colour and by-products are formed more frequently [55]. The isomerized glucose syrup is almost colorless and has few by-products when plug flow reactors are used and operated in adequate reaction conditions [2,55]. One needs to evaluate experimentally if the nature of the produced by-products in reactors could affect the efficiency of the separation operation taking place in the adsorbers. Other ways of coupling the glucose isomerization and isomers separation have been studied experimentally [26,31] and the above discussed problem has not been addressed, in despite of its importance.

Another aspect that must not be forgotten is the deactivation of the enzyme. After starting the isomerization process in reactors packed with immobilized enzyme the initial activity of the enzyme decays following some days of operation [43]. The decay rate depends on the operating conditions and it seems to be quite slow in operations occurring with no pH drops or at low temperatures (at around 55 °C) [56]. To describe activity decay of different glucose isomerase enzymes with time, either first-order kinetics models [10,57] or more sophisticated models [58–60] have been used. The loss of activity of the enzymes immobilized on a solid matrix is due to the leaching, denaturation, poisoning or damage to particles [57,59]. In order to avoid a decrease on the glucose conversion during the isomerization and therefore to maintain the desired degree of conversion, some ideas have been suggested, such as: (i) to supply enzyme periodically according to a predetermined schedule by an amount which can be predicted from decay models of the enzyme reactor system, (ii) to gradually decrease the amount and/or concentration of glucose in the feed in accordance with the decrease in

activity of the enzyme [10], or (iii) to reduce the flow rate during the run to kept conversion constant [55,56]. In these two last alternatives, when the activity is very small compared to the initial activity (or when the flow rate is much lower than the initial value), the best solution is to discard the enzyme and replace it. Thus, for design purposes, it is also a crucial point to know the decay of the enzyme activity.

Because of the deactivation of the biocatalyst in the reactors of the proposed system, one can suggest the substitution of reactors in which the required conversion is not achieved any more by other ones packed with fresh enzyme. Alternatively, one can change liquid and solid flow rate, trying to keep the design values of m parameters, in such way that it is possible to maintain the desired level of conversion for a longer period of time. It is obvious that none of the mentioned alternatives is straightforward task and perhaps the cost of the process will increase. Hence, as a continuation of this work, the economic evaluation of the proposed system should be performed. A comparison of the productivity, eluent consumption, costs with stages of evaporation, etc., achieved with the configuration showed in Fig. 9 and those performance parameters reached by current industrial processes (reactor plus separation unit) could better illustrate the potential of this new configuration. However, the possibility of glucose isomerization in an integrated system of reaction and of separation producing high-fructose syrup from pure glucose solution has been demonstrated. The suggested system can also be useful in any other processes of reversible reactions of type $A \rightleftarrows B$, to exceed the equilibrium conversion.

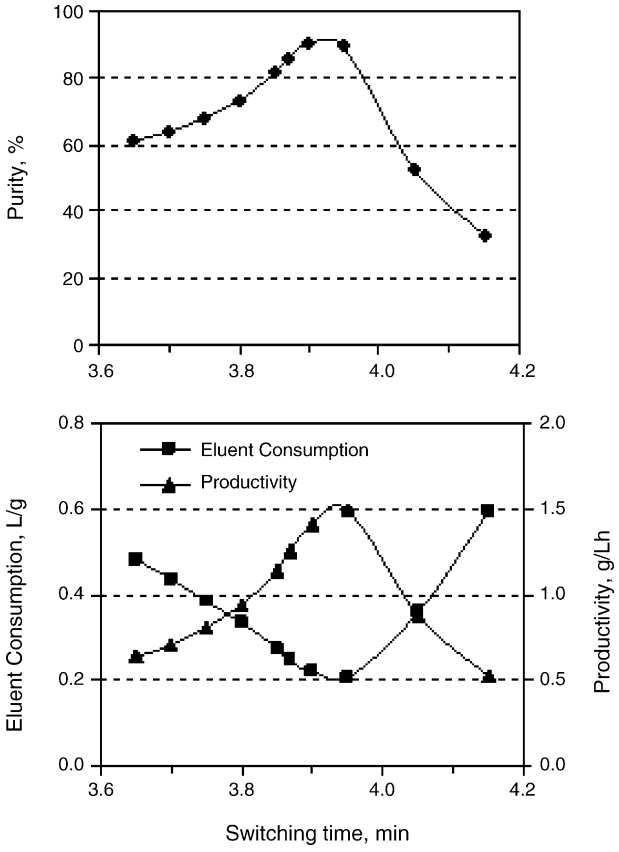

Fig. 14. Influence of the switching time (t^*), on the performance of the SMBR for the glucose isomerization.

4.4. Influence of operating conditions

The effect of some process variables (switching time, extract flow rate and feed flow rate) on the performance of the separation/reaction process is examined by numerical study. The performance of simulated moving bed is evaluated by three performance criteria: purity, eluent consumption and productivity—Table 1. In all situations analyzed, the adsorption data (equilibrium and kinetics) are those of the resin FINEX in Mg^{2+} form and the columns are 30 cm length × 2.6 cm i.d., arranged as 4rea-4ads-2-2 in the system. The recycle flow rate is kept constant equal to 23.765 mL/min.

Fig. 14 presents the effect of the switching time on the process performance. The operation conditions are $Q_{Fe} = 0.371$ mL/min, $Q_{El} = 5.560$ mL/min and $Q_{Ex} = 5.931$ mL/min. The m-parameters are changing as function of the switching time.

From the analysis of Fig. 14, and considering the purity of the product stream, it is noticed that there is an optimum switching time corresponding to the maximum of product purity. The decrease of this operating parameter changes the net flow of the species in opposite direction to the flowing phase. This will affect the performance of all the sections in relation to their functions. On the other hand, an increase of the switching time influences the net flow of the species in the same direction of the flowing phase. As a result, the glucose species that should be transformed and adsorbed in Section 3 will go to Section 1 and will contaminate the fructose in the extract current.

In this analysis, the profile of the solvent consumption and adsorbent productivity in Fig. 14 depends on the behaviour of the sugars concentration in the exit stream, since the feed and the eluent streams, and the adsorbent volume in the subsections, are kept constant. It is observed that there is a region where the eluent consumption is minimum and the productivity is maximum in the system, around to 3.9 min.

The influence of the variation of the extract and feed flow rate are presented in Figs. 15 and 16, respectively. The results presented in Fig. 15 shows the effects of the changes of the extract flow rate; beyond already mentioned operational variables, the feed flow rate and the switching time are also maintained constant and equal to 0.371 mL/min and 3.87 min, respectively.

Fig. 15 shows that the extract purity decreases at lower extract flow rates while there is no significant changes at higher flow rates. If fact when the extract flow rate is changed, at fixed feed flow rate, recycle flow rate and switching time, only the interstitial velocity of the flowing phase in Section 1 is changed, as a consequence of the mass balance in the unit nodes. Therefore, in this situation, the m_2 and m_3 parameters are constant and equal to 0.281 and 0.296. Section 1 is basically responsible for the adsorbent regeneration by the fructose desorption.

One can notice that there is no reaction in Sections 1 and 2 of the proposed system; it works as a separative SMB in which for complete separation the restrictions of m_1 and m_2 in Eqs. (23) and (24) must be valid. However, a 100% purity fructose

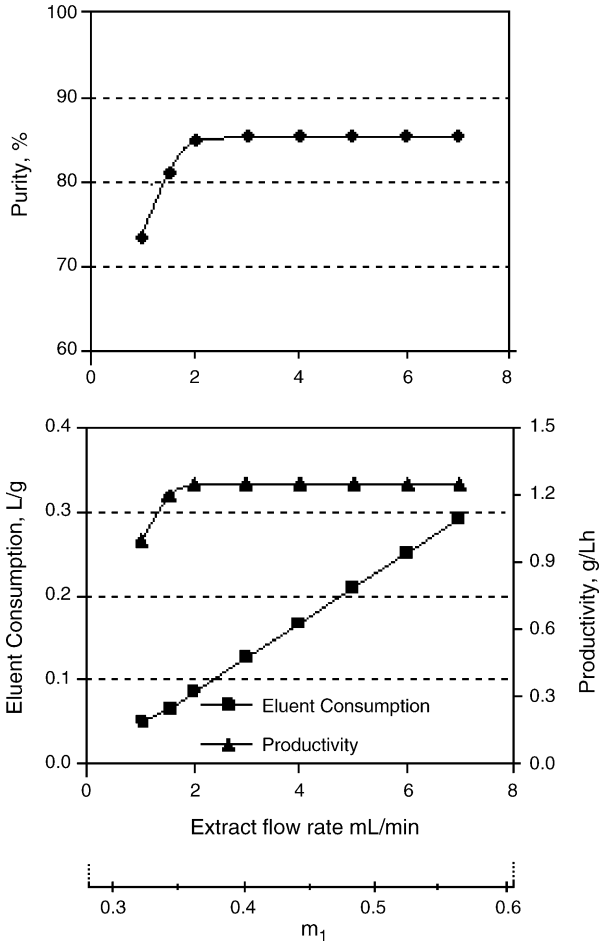

Fig. 15. Influence of the extract flow rate (for constant feed flow rate) on the performance of SMBR for glucose isomerization.

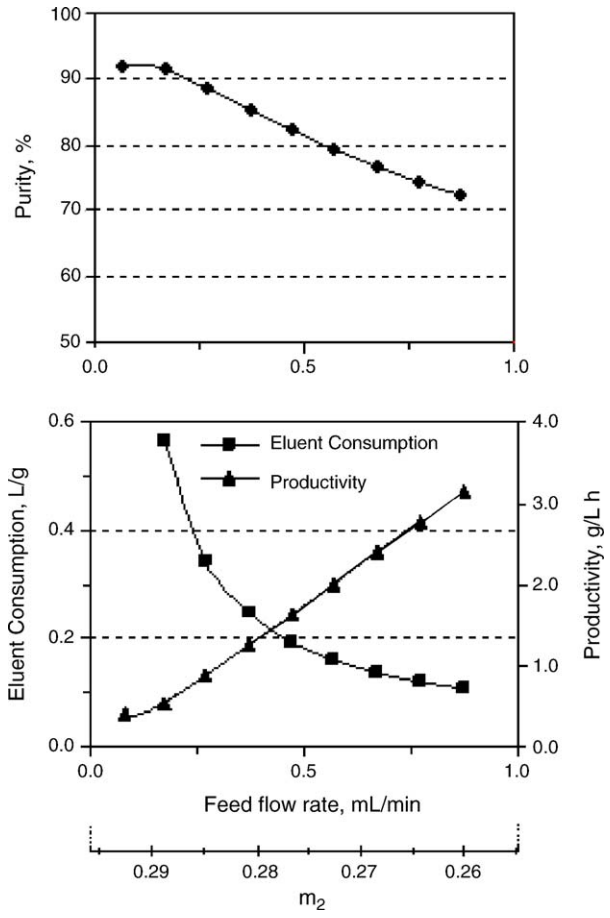

Fig. 16. Influence of the feed flow rate (for constant eluent flow rate) on the performance of the SMBR for glucose isomerization.

would only be possible if glucose was converted completely in Section 3 and this could occur if system had an infinite number of coupled reactor/adsorber in that section.

In the condition statement above, lower extract flow rate implies lower flow rate in Section 1 (and so lower m_1). Too low values of m_1 will allow contamination of extract by glucose. For higher values of m_1 the contamination of final product is due to the glucose not converted from Section 3. Thus, when liquid flow rate in Section 1 is too increased, there is not significant variation in the fructose purity. It means that the region in which the operating conditions allow to obtain high-fructose purities in the space (m_1; m_2; m_3) would not change with higher values of m_1 parameter.

Fig. 16 shows the effect of the feed flow rate on the performance of the process; when the eluent flow rate and the switching time are maintained constant (equal to 5.56 mL/min and 3.87 min, respectively). So, the parameters m_1 and m_3 are constants in this analysis.

One can see that fructose purity decreases in extract stream when the feed flow rate increases. In such case, and keeping the eluent flow rate constant, the extract flow rate must also increase and the interstitial velocity in Section 2 becomes lower. So, the net flow of both species is influenced and they move in opposite direction to the flowing liquid phase. The glucose species, which should be desorbed in Section 2, reaches to the extract outlet and leads to a reduction in fructose purity. On the other hand, when the feed flow rate is decreased, the liquid flow rate in Section 2 is larger influencing the net flow of both species, which move in direction of the flowing liquid phase in this section, increasing the fructose purity.

With regard to the solvent consumption and the productivity, one can notice that both are favourable when the feed flow rate increases; what means, they are minimum and maximum, respectively. The product concentration is larger for higher feed flow rate, but at the cost of a reduction of fructose purity in the extract stream.

5. Conclusion

The process of glucose isomerization in the SMBR was analyzed to obtain the production of HFS (high/higher fructose syrup). The kinetics of the glucose isomerization was obtained with the technique of Linewaever–Burk using the enzyme immobilized glucose isomerase (Sweetzyme IT) as catalyst.

The adsorption experiments allowed getting kinetic and equilibrium parameters for glucose and fructose in the ion exchange resin. The capacity of the sugars separation was investigated for the cationic resin in the commercial ionic form – Ca^{2+} – and

in the form Mg^{2+}. To integrate the enzymatic reaction and the adsorptive separation in the same unit, it is necessary to use the resin in magnesium form, because the presence of ions calcium inhibits the enzymatic action in the glucose conversion.

The adsorption kinetics was represented by the linear driving force model where the global mass transfer coefficient, containing the resistances to the mass transfer in the adsorbed process, has been obtained from a 'best fit' procedure in which the sum of residual values is minimized. The values of global mass transfer coefficient of the sugars, when applied the adsorbent under magnesium form, were much larger than the corresponding values when the resin was under calcium form. In the range of investigated concentration (0 to 1 mol/L), the adsorption equilibrium isotherms for the two sugars in both ionic forms of the resin showed a linear behaviour. The adsorption equilibrium constant for fructose, using the adsorbent in the form Ca^{2+}, was much larger than that obtained when the resin was in the form Mg^{2+}. The separation factor calculated for the resin in the form Ca^{2+} was of 1.52 and in the form Mg^{2+} was of 1.20.

A numerical methodology was developed to predict the behaviour and the performance of reactors and reactive units of simulated moving bed through the presented model. In this work, alternative configuration of SMBR were evaluated for glucose isomerization in order to obtain fructose in requested purities. A new system was suggested and it was shown to be able to obtain the reaction product with purity above 90%. This system uses the idea of Hashimoto et al. [26]: a group of fixed bed reactors is incorporated to Section 3 of a SMB with three sections; however, the configuration of the unit was modified in a way that was possible to feed the system with glucose only. The system can use all eluent of the feed solution of the glucose, because there is not raffinate stream, and then the amount of additional pure eluent for adsorbent regeneration can be reduced.

The effect of some operating conditions (switching time, extract and feed flow rates) were investigated in view of process optimization, and the process performance analyzed through variables as product purity, eluent consumption and productivity. It was verified that there are combinations of operating variables where high purity in the extract stream, and good productivity and low solvent consumption can be obtained. With this study, it is possible to establish the adequate operation conditions for a certain application of the unit. Considering the results obtained in this work, it is possible to apply this methodology for other reversible reactions of the type $A \rightleftarrows B$, and to obtain some performance in the process involving separation and reaction in SMBR technology.

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