



Purification of organic acids by chromatography with strong anionic resins: Investigation of uptake mechanisms



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ABSTRACT

Bio-based organic acids are promising renewable carbon sources for the chemical industry. However energy-consuming purification processes are used, like distillation or crystallization, to reach high purities required in some applications. That is why preparative chromatography was studied as an alternative separation technique. In a previous work dealing with the purification of lactic, succinic and citric acids, the Langmuir model was insufficient to explain the elution profiles obtained with a strong anionic resin.

Consequently the Langmuir model was coupled with a usual ion-exchange model to take into account the retention of their conjugate bases (<2%), which are commonly neglected at low pH (<1.5). Elution simulations with both uptake mechanisms fitted very well with experimental pulse tests. Only two parameters were optimized (equilibrium constant of acid uptake and ion-exchange selectivity coefficient of conjugate base) and their value were coherent with experimental and resin suppliers' data.

These results confirmed that the singular tailing and apparent delay observed with succinic and citric acids can be explained by the high affinity of succinate and citrate for resin cationic sites. The model was implemented in a preparative chromatography simulation program in order to optimize operating parameters of our pilot-scale ISMB unit (Improved Simulated Moving Bed). The comparison with experimental ISMB profiles was conclusive.

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1. Introduction

In the frame of sustainable development, biomass valorization as a renewable carbon source for the chemical industry is growing. Indeed, fossil resources are running out in the medium to long term. Moreover carbon balances have to be improved so as to reduce the CO₂ level in the atmosphere, whose rapid rise is linked to global warming. In the famous report of the Pacific Northwest National Laboratory (PNNL) and the National Renewable Energy Laboratory (NREL) [1], bio-based organic acids are identified as very promising sources of renewable carbon. Many compounds can be produced from these molecules in the chemical industry, particularly in the polymer industry. Up to now, the production of bio-based organic acids keeps small because they are still uncompetitive compared to the corresponding petroleum-derived products [2]. That is why

the reduction of their production cost is a major concern of the chemical industry.

Bio-based organic acids are usually produced by microbial fermentation for the food industry but further purification techniques are required for others applications which need a much higher purity, particularly polymer synthesis. Currently, distillation [3,4], esterification [5–7] and crystallization [8] are commonly used to reach very high purities (>99.5%) [9]. However, these techniques are energy-consuming and give quite low recovery rates. Preparative chromatography is a promising solution for downstream processes because its energy, chemical and water consumptions are generally quite lower and consequently effluent volumes too. Initially developed for extraction and purification of complex vegetal mixtures [10], this separation method is spreading in different fields: biotechnology [11,12], pharmaceutical [13,14] and the sugar industry [15], for instance. Indeed, the same separation quality may be reached with better yields, lower costs and impacts on the environment. However, for several years now, the state of the world economy has not been favorable to industrial investments for the improvement of downstream purification processes. Moreover,

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chromatography needs to prove itself for high purity application (>99.5%) at the industrial scale.

In the frame of organic acids purification in fermentation broths, chromatography is often performed with ion-exchange resins to separate mineral salts, sugars and organic acids. The eluent is an aqueous solution whose pH can be adjusted with the inorganic acid or base corresponding to the counter-ions fixed on the resin so as to keep its composition constant. The separation of compounds is based on affinity differences for the resins. Higher is the affinity, slower migrates the compound along the resin bed. In the sugar industry, cationic resins are used to separate sucrose from salts as well as sugars from each other. The glucose–fructose separation is among the most famous applications of preparative chromatography. Otherwise, anionic resins are performed for the citric acid purification. Uptake and slowdown mechanisms are not well known and complex, coupling size exclusion, ligand exchange, hydrophilic or hydrophobic interactions and ion-exchange [16,17].

In a previous work dealing with the purification of lactic, succinic and citric acids by chromatography with a strong anionic resin (quaternary ammonium), the Langmuir model fitted well with the experimental equilibrium isotherms obtained by frontal analysis [18]. However, the corresponding affinity coefficients were significantly lower than those obtained by pulse tests for succinic and citric acids. Their peaks had a singular tailing and were delayed compared to simulated ones with the Langmuir model and experimental coefficients of isotherms. Indeed, the Langmuir model was insufficient to describe the organic acids retention on the strong anionic resin and frontal analysis gave underestimated affinity coefficients.

These results lead us to suspect the impact of conjugate bases (dissociated organic acids) retention by ion-exchange, despite their very low concentrations. Indeed, the citrate anion affinity is high for quaternary ammonium anionic resin according to selectivity coefficients given by resins producers. Usually, this uptake mechanism is not taken into account because the strong anionic resins are performed at such acid pH (1–1.5) that only 0.2–0.4% of lactic acid, 0.1–0.2% of succinic acid and 1.2–2.3% of citric acid are dissociated. Thus, the organic acids dissociation is generally not considered although such small amount of conjugate bases may have a significant impact if they are strongly retained.

The present work aimed at modelling uptake mechanisms on a strong anionic resin. The Langmuir model, commonly used for the retention of organic acids, was coupled with a usual ion-exchange model for the retention of conjugate bases. Simulation results were compared with experimental pulse tests to check the model. The final goal was to improve the accuracy of a preparative chromatography simulation program in order to optimize the operating parameters of our pilot-scale ISMB unit (Improved Simulated Moving Bed).

2. Chromatography modelling

2.1. Mass transfer model

The Martin and Synge plate model [19] was chosen to describe the elution of organic acids through a strong anionic resin bed. In each plate of volume V , the mobile phase takes up a volume εV and the stationary phase a volume $(1 - \varepsilon)V$, with ε the bed voidage. Usually, for preparative chromatography, volumes are expressed in BV (bed volume) [20–22], corresponding to the ratio of volume to the resin bed volume. Both phases are always supposed in equilibrium in all plates (nonlinear equilibrium in our case). The axial dispersion is characterized by the plate number N .

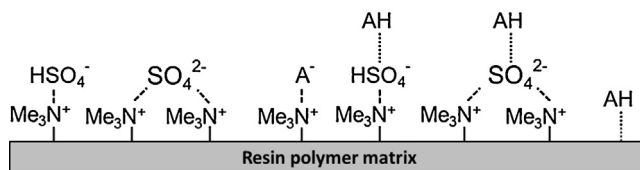


Fig. 1. Illustration of the six compounds considered in the stationary phase.

2.2. Uptake mechanisms—reactions in stationary phase

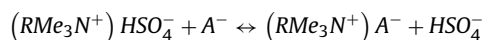
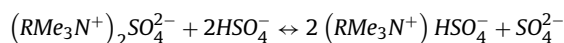
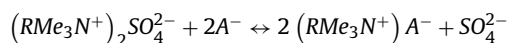
The strong anionic resin is a styrene-divinylbenzene resin functionalized with quaternary ammonium groups RMe_3N^+ . The amount of these cationic sites corresponds to the ion-exchange capacity $q_{\text{RMe}_3\text{N}^+}$ of the resin, expressed in $\text{mol L}^{-1}_{\text{resin}}$. The resin is initially in the sulfate form, namely with the counteranions HSO_4^- and SO_4^{2-} in equilibrium. The proportion of both forms depends on their concentration in the mobile phase, so on its pH. The concentrations of hydroxyl anion OH^- in both phases were considered negligible at $\text{pH} < 1.5$.

The proposed model was based on two classical uptake mechanisms, with some assumptions:

- Uptake of the organic acid in the dissociated form A^- by ion-exchange with resin counteranions HSO_4^- and SO_4^{2-} .
- Retention of the organic acid in the molecular form AH according to the Langmuir model, by weak bonds (hydrogen or van der Waals), on free sites either on counteranions HSO_4^- and SO_4^{2-} (weak nucleophiles, cf. UOP patents [23,24]) or on the polymer matrix whose capacity q_{mat} is also expressed in $\text{mol L}^{-1}_{\text{resin}}$. We supposed that the AH retention does not modify the selectivity coefficients of counteranions.

Consequently, 6 compounds were considered in the stationary phase: HSO_4^- , SO_4^{2-} , A^- , $\text{AH-site}(\text{HSO}_4^-)$, $\text{AH-site}(\text{SO}_4^{2-})$ and $\text{AH-site}(\text{matrix})$ (Fig. 1).

The ion-exchange reactions correspond to the following equilibria:



Since the affinity of counteranions HSO_4^- and SO_4^{2-} for resin cationic sites was supposed not influenced by AH uptake, their equilibrium constants can be written as follows:

$$K_{\text{A}^-/\text{SO}_4^{2-}} = \frac{[\text{SO}_4^{2-}] q_{\text{A}^-}^2}{[\text{A}^-]^2 q_{\text{SO}_4^{2-}}^2} \quad (1)$$

$$K_{\text{HSO}_4^-/\text{SO}_4^{2-}} = \frac{[\text{SO}_4^{2-}] q_{\text{HSO}_4^-}^2}{[\text{HSO}_4^-]^2 q_{\text{SO}_4^{2-}}^2} \quad (2)$$

$$K_{\text{A}^-/\text{HSO}_4^-} = \frac{[\text{HSO}_4^-] q_{\text{A}^-}}{[\text{A}^-] q_{\text{HSO}_4^-}} = \sqrt{\frac{K_{\text{A}^-/\text{SO}_4^{2-}}}{K_{\text{HSO}_4^-/\text{SO}_4^{2-}}}} \quad (3)$$

where $q_{\text{HSO}_4^-}$ and $q_{\text{SO}_4^{2-}}$ represent the HSO_4^- and SO_4^{2-} contents in the stationary phase (free or bound to AH), q_{A^-} the A^- content and $[\text{X}]$ the concentration of X in the mobile phase.

The ion-exchange equilibrium constants K can be estimated from data of resin suppliers by Equations (5)–(7). For example, the Dow Chemical Company gives the selectivity coefficients of anions

X[−] for a strong anionic resin in the hydroxide form, $k_{X/OH}$, defined as follows:

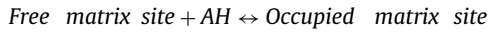
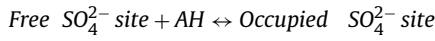
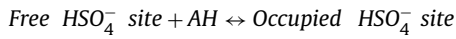
$$k_{X/OH} = \frac{q_{X-} [OH^-]}{[X^-] q_{OH-}} \quad (4)$$

$$Ke_{HSO_4^-/SO_4^{2-}} = \frac{k_{HSO_4/OH}^2}{k_{SO_4/OH}} \quad (5)$$

$$Ke_{A^-/SO_4^{2-}} = \frac{k_{A/OH}^2}{k_{SO_4/OH}} \quad (6)$$

$$Ke_{A^-/HSO_4^-} = \frac{k_{A/OH}}{k_{HSO_4/OH}} \quad (7)$$

According to the Langmuir model, the AH uptake reactions on free sites on counteranions HSO_4^- and SO_4^{2-} and on the polymer matrix correspond to the following equilibria:



The corresponding equilibrium constants K_s are defined as follows:

$$K_{s_{AH/HSO_4^-}} = \frac{q_{AH-HSO_4}}{[AH] q_{HSO_4^-}^{\text{free}}} \quad (8)$$

$$K_{s_{AH/SO_4^{2-}}} = \frac{q_{AH-SO_4}}{[AH] q_{SO_4^{2-}}^{\text{free}}} \quad (9)$$

$$K_{s_{AH/mat}} = \frac{q_{AH-mat}}{[AH] q_{mat}^{\text{free}}} \quad (10)$$

where q_{AH-HSO_4} , q_{AH-SO_4} and q_{AH-mat} represent the AH concentrations bound to uptake sites respectively on counteranions HSO_4^- and SO_4^{2-} and on the polymer matrix, meanwhile $q_{HSO_4^-}^{\text{free}}$, $q_{SO_4^{2-}}^{\text{free}}$ and q_{mat}^{free} represent the concentrations of corresponding free uptake sites.

Mass balances on the three kinds of uptake sites lead to the following equations:

$$q_{HSO_4^-}^{\text{free}} = x q_{HSO_4^-} - q_{AH-HSO_4} \quad (11)$$

$$q_{SO_4^{2-}}^{\text{free}} = y q_{SO_4^{2-}} - q_{AH-SO_4} \quad (12)$$

$$q_{mat}^{\text{free}} = q_{mat} - q_{AH-mat} \quad (13)$$

where x and y correspond to the uptake capacity of counteranions HSO_4^- and SO_4^{2-} , namely the number of molecule that can be retained per counteranions. Indeed, we assumed that several molecules can eventually be retained on each counteranion. These numbers are not necessarily integers and can be smaller than 1 if one molecule AH occupies several counteranions HSO_4^- or SO_4^{2-} for steric reasons.

For each type of uptake sites, the combination of one mass balance with its corresponding equilibrium constant gives the following usual equations of the Langmuir model. They relate concentration of AH retained respectively on the three kind of sites to AH concentration in mobile phase and total concentration of corresponding sites.

$$q_{AH-HSO_4} = \frac{K_{s_{AH/HSO_4^-}} [AH] (x q_{HSO_4^-})}{1 + K_{s_{AH/HSO_4^-}} [AH]} \quad (14)$$

$$q_{AH-SO_4} = \frac{K_{s_{AH/SO_4^{2-}}} [AH] (y q_{SO_4^{2-}})}{1 + K_{s_{AH/SO_4^{2-}}} [AH]} \quad (15)$$

$$q_{AH-mat} = \frac{K_{s_{AH/mat}} [AH] q_{mat}}{1 + K_{s_{AH/mat}} [AH]} \quad (16)$$

2.3. Reactions in mobile phase

At low pH (<1.5), organic acids are mainly in the molecular form AH, however the retention of the dissociated form A[−] was studied (for polyacids like succinic or citric acids, only the monodissociated form was considered). So 5 species were considered in the mobile phase: H⁺, HSO_4^- , SO_4^{2-} , AH and A[−]. The main reactions are both acid-base equilibria, whose equilibrium constants K_a are defined as follows:



$$K_{a_{AH}} = \frac{[A^-] [H^+]}{[AH]} \quad (17)$$

$$K_{a_{HSO_4^-}} = \frac{[SO_4^{2-}] [H^+]}{[HSO_4^-]} \quad (18)$$

The pKa of lactic, succinic and citric acids and hydrogen sulfate at 60 °C (operating temperature) are respectively 3.86, 4.13, 3.05 and 2.39.

2.4. Plate mass balances

The plate content is represented by 11 unknown variables (5 in the mobile phase and 6 in the stationary phase) that obey 7 independent equilibrium relations (Eqs. (1) and (2) and (14)–(18)). Consequently, 4 mass balances are necessary to solve the system and estimate the content of each plate at time “t” from the composition of all plates at time “t − 1”.

Mass balance of ion-exchange site RMe_3N^+ whatever the time t:

$$2q_{SO_4^{2-}t} + q_{HSO_4^-t} + q_{A^-t} = q_{RMe_3N^+} \quad (19)$$

Mass balance of sulfate group SO_4 between t − 1 and t:

$$\begin{aligned} \varepsilon V \left(\frac{d[HSO_4^-]_t^n}{dt} + \frac{d[SO_4^{2-}]_t^n}{dt} \right) + (1 - \varepsilon) V \left(\frac{dq_{SO_4^{2-}t}^n}{dt} + \frac{dq_{HSO_4^-t}^n}{dt} \right) \\ = Q \left([HSO_4^-]_t^{n-1} - [HSO_4^-]_t^n \right) + Q \left([SO_4^{2-}]_t^{n-1} - [SO_4^{2-}]_t^n \right) \end{aligned} \quad (20)$$

Mass balance of acids (H⁺ element) between t − 1 and t (water dissociation negligible):

$$\begin{aligned} \varepsilon V \left(\frac{d[H^+]_t^n}{dt} + \frac{d[HSO_4^-]_t^n}{dt} + \frac{d[AH]_t^n}{dt} \right) \\ + (1 - \varepsilon) V \left(\frac{dq_{HSO_4^-t}^n}{dt} + \frac{dq_{AH-HSO_4t}^n}{dt} + \frac{dq_{AH-SO_4t}^n}{dt} + \frac{dq_{AH-mat}^n}{dt} \right) \\ = Q \left([H^+]_t^{n-1} - [H^+]_t^n \right) + Q \left([HSO_4^-]_t^{n-1} - [HSO_4^-]_t^n \right) \\ + Q \left([AH]_t^{n-1} - [AH]_t^n \right) \end{aligned} \quad (21)$$

Mass balance of organic compound A (molecular and dissociated forms) between $t - 1$ and t :

$$\begin{aligned} \varepsilon V \left(\frac{d[A^-]_t^n}{dt} + \frac{d[AH]_t^n}{dt} \right) \\ + (1 - \varepsilon)V \left(\frac{dq_{A^-}^n}{dt} + \frac{dq_{AH-HSO_4^-}^n}{dt} + \frac{dq_{AH-SO_4^{2-}}^n}{dt} + \frac{dq_{AH-mat}^n}{dt} \right) \\ = Q \left([A^-]_t^{n-1} - [A^-]_t^n \right) + Q \left([AH]_t^{n-1} - [AH]_t^n \right) \end{aligned} \quad (22)$$

In the last three mass balances, the terms on the left-hand side of the equal sign correspond respectively to the accumulation terms in the mobile phase and in the stationary phase, meanwhile the expressions on the right-hand side correspond to the difference between inputs (from the $(n - 1)^{th}$ plate) and outputs (from the n^{th} plate). Q represents the flowrate.

The proposed model depends on 10 parameters whose value is more or less known accurately according to data from resin suppliers, literature or our own experimental results:

- Plate number: N ,
- Resin ion-exchange capacity: $q_{RMe_3N^+}$ expressed in $\text{mol L}^{-1}_{\text{resin}}$,
- Polymer matrix uptake capacity: q_{mat} expressed in $\text{mol L}^{-1}_{\text{resin}}$,
- HSO_4^- uptake capacity: x ,
- SO_4^{2-} uptake capacity: y ,
- Equilibrium constant of ion-exchange between A^- and HSO_4^- : $K_{e_{A^-/HSO_4^-}}$,
- Equilibrium constant of ion-exchange between A^- and SO_4^{2-} : $K_{e_{HSO_4^-/SO_4^{2-}}}$,
- Equilibrium constant of AH uptake on HSO_4^- counteranions: $K_{s_{AH/HSO_4^-}}$,
- Equilibrium constant of AH uptake on SO_4^{2-} counteranions: $K_{s_{AH/SO_4^{2-}}}$,
- Equilibrium constant of AH uptake on resin polymer matrix: $K_{s_{AH/mat}}$.

2.5. Numerical methods

The finite-difference method with implicit scheme was used to discretize the 3 differential equations (Eqs. (20)–(22)). The Newton-Raphson iterative method was chosen to solve the nonlinear system of 4 remaining equations after simplification. The model and its numerical solving were computed using the software MATLAB®. The initial conditions (concentrations in both phases) need to be given as well as the boundary conditions: composition of the feed (organic acid solution then eluent) as a function of time.

3. Results and discussion

3.1. Simulation with only the Langmuir model and coefficients of equilibrium isotherms

First, the simulations of organic acids elution through a strong anionic resin bed were performed considering that AH was only retained on the polymer matrix, according to the Langmuir model. We used the coefficients of equilibrium isotherms obtained previously by frontal analysis (Table 1). So, the elution profiles were only influenced by 3 parameters: the plate number N , the polymer matrix uptake capacity q_{mat} equated to q_{max} and the equilibrium constant of AH uptake on resin polymer matrix $K_{s_{AH/mat}}$ equated to $K_{s_{AH}}$.

For lactic and succinic acids, the resin uptake capacity q_{max} and the equilibrium constant $K_{s_{AH}}$ have significant error margin

Table 1

Resin uptake capacity q_{max} and equilibrium constant $K_{s_{AH}}$ of 3 organic acids estimated by frontal analysis at 60 °C for a strong anionic resin [13].

| | $q_{max} \text{ mol L}^{-1}_{\text{resin}}$ | $K_{s_{AH}} 10^{-1} \text{ L}_{\text{solution}} \text{ mol}^{-1}$ |
|---------------|---|---|
| Lactic acid | 6.8 ± 2.6 | 1.4 ± 0.5 |
| Succinic acid | 2.7 ± 0.9 | 7.2 ± 2.5 |
| Citric acid | 2.5 ± 0.2 | 9.0 ± 1.7 |

because it was not possible to get enough experimental points in the nonlinear region of equilibrium isotherms [18].

According to our previous study [18], the plate number N was set to 250 for the 3 organic acids. It corresponds to the experimental value obtained by pulse tests with lactic acid whose symmetric profile is the most appropriate to characterize axial dispersion. The plate number N was considered equal for all species. Indeed, the study of axial dispersion (Van Deemter curves) in this work showed that the molecular diffusion is negligible compared to the mechanical dispersion and the influence of mass transfer limitation, which were quite similar between compounds in our experimental operating conditions.

These conditions were applied in the model so as to compare with experimental results: 0.05 BV of 60 g L⁻¹ organic acid solution, whose pH was adjusted to 1.4 with sulfuric acid, were eluted with a 2.5 g L⁻¹ sulfuric acid solution. The temperature was 60 °C and the flow rate was 0.5 BV h⁻¹.

For lactic acid, the simulation results are quite close to the experimental ones (Fig. 2). So, the lactic acid uptake mechanisms by a strong anionic resin can be well fitted with only the Langmuir model. However, this model is insufficient to simulate elution of succinic and citric acids. Indeed, the simulation gives much earlier and narrower peaks than the experiments, particularly for citric acid. Another uptake mechanism must then be responsible for the delay and peak tailing observed experimentally and should be taken into account.

3.2. Simulation with taking into account the retention of the dissociated form A^-

Simulations of organic acids elution were performed with the proposed model which takes into account the retention of the molecular form AH according to the Langmuir model and the retention of the dissociated form A^- by ion-exchange with resin counteranions.

To simplify, as a start, we considered that the molecular form AH was retained only on free uptake sites of counteranions HSO_4^- and SO_4^{2-} according to UOP patents [23,24]. Moreover, we supposed that the equilibrium constants of AH uptake on both counteranions were equal. They were estimated first from the experimental equilibrium constant $K_{s_{AH}}$ (Table 1), then optimized by fitting simulation results to experimental profiles (least squares method). They were expected lower than $K_{s_{AH}}$ since it includes also the contribution of ion-exchange in retention.

According to the supplier, the ion-exchange capacity $q_{RMe_3N^+}$ is about 1.35 mol L⁻¹_{bed}, equivalent to 2.4 mol L⁻¹_{resin}. The uptake capacities of counteranions HSO_4^- and SO_4^{2-} , x and y , were supposed equal and defined from the experimental uptake capacity q_{max} (Table 1). For instance, q_{max} of lactic acid is twofold to threefold higher than $q_{RMe_3N^+}$, so x and y were supposed equal to 2 (Table 3). It could be related to both nucleophile carbonyl groups of counteranions HSO_4^- and SO_4^{2-} while lactic acid gets only one electrophile hydroxyl group. It means that 2 molecules of lactic acid can be retained on each counteranion. For succinic and citric acids, x and y were supposed equal to 1 (Table 3).

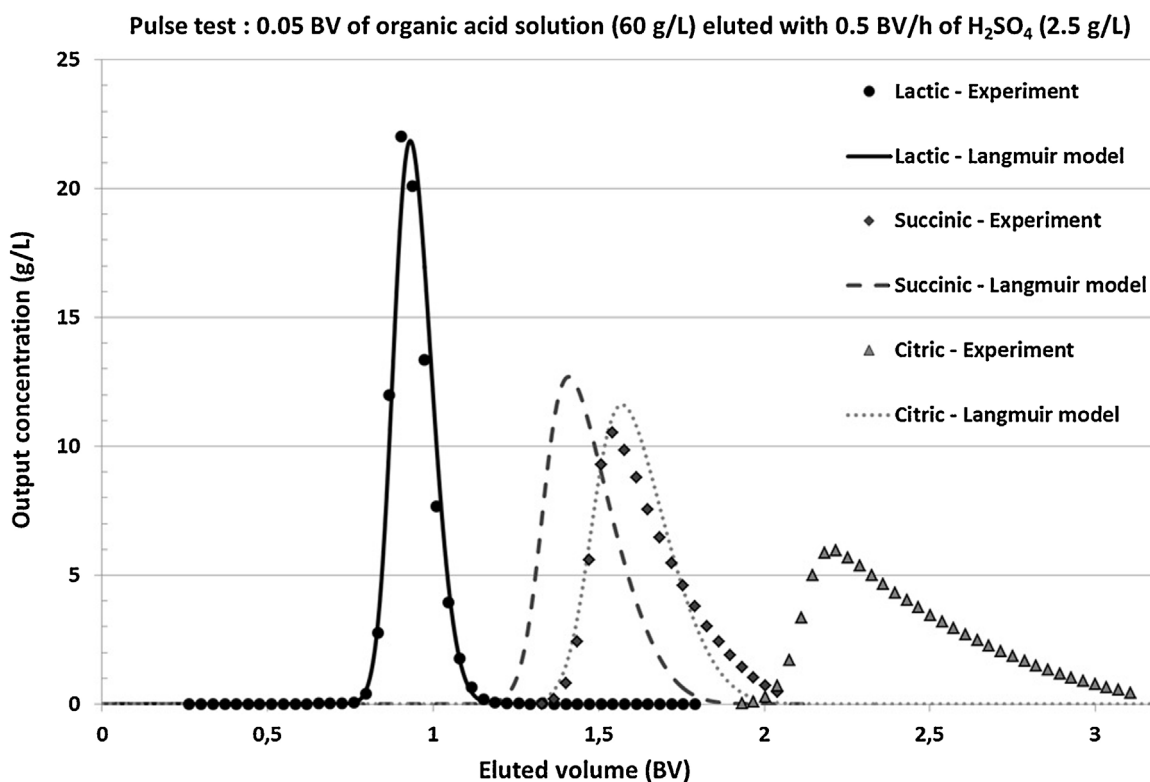


Fig. 2. Comparison of experimental pulse test profiles with simulated ones considering the Langmuir model with experimental coefficients of equilibrium isotherms.

Table 2

Selectivity coefficients of some anions (compared with OH[−]) on a strong anionic resin (Data from Dow Chemical Company).

| | HSO ₄ [−] | SO ₄ ^{2−} | Citrate ^a | Formate | Acetate | Propionate |
|------------|-------------------------------|-------------------------------|----------------------|---------|---------|------------|
| $k_{X/OH}$ | 85 | 150 | 220 | 4.6 | 3.2 | 2.6 |

^a Valence not mentioned.

Table 3

Optimized parameters for simulation of lactic, succinic and citric acids elution with the proposed model which also takes into account A[−] ion-exchange.

| | | Lactic acid | Succinic acid | Citric acid |
|---------------------|--------------------------------------|-------------|---------------|-------------|
| N | – | 250 | 250 | 250 |
| $q_{RMe_3N^+}$ | mol L ^{−1} _{resin} | 2.4 | 2.4 | 2.4 |
| $k_{A/OH}$ | – | 2.5 | 330 | 110 |
| $k_{HSO_4/OH}$ | – | 85 | 85 | 85 |
| $k_{SO_4/OH}$ | $L_{resin} L_{solution}^{-1}$ | 150 | 150 | 150 |
| K_{SAH/HSO_4^-} | $L_{solution} mol^{-1}$ | 0.225 | 0.82 | 0.70 |
| $K_{SAH/SO_4^{2-}}$ | $L_{solution} mol^{-1}$ | 0.225 | 0.82 | 0.70 |
| x | – | 2 | 1 | 1 |
| y | – | 2 | 1 | 1 |

The selectivity coefficients $k_{X/OH}$ were first estimated from data of the Dow Chemical Company (Table 2), then optimized by fitting simulation results to experimental profiles (least squares method).

The plate number N was still set to 250 for the 3 organic acids. Values of simulation parameters for the 3 organic acids were gathered in the Table 3.

Fig. 3 shows that the proposed model fitted very well the experimental profiles with coherent value for each parameter compared to experimental and resin supplier's data. The delay and peak tailing observed mainly with succinic and citric acids can be reproduced with this model.

It confirms that the retention of the dissociated organic acids by ion-exchange with resin counteranions must be considered despite

of the very low dissociation of the acid at acidic pH (0.1–2%). Indeed, the conjugate base of some organic acids, like citrate, can have a very high affinity for resin cationic sites, so the organic acid can be much more retained than expected. Moreover, it shows that assumptions added in this simulation to simplify the model were acceptable.

3.3. Implementation in a ISMB chromatography simulation program

This uptake model was developed in order to improve simulation of preparative chromatography like ISMB (Improved Simulated Moving Bed) that is currently used at pilot scale in our laboratory. We aimed at predicting the ISMB performance more accurately with strong anionic resins and optimizing operating parameters according to the wanted recovery rates and purities.

An ISMB unit is composed of 4 columns in series filled with the stationary phase. The ISMB process consists in changing sequentially position of inputs (feed and eluent) and outputs (extract enriched with the more retained component, succinic acid, and raffinate enriched with the less retained, salts and lactic acid). An operating step is composed of four different sub-steps, where each column corresponds to one zone:

- 1) Feed input in the zone 3 and raffinate output from the zone 3.
- 2) Eluent input in the zone 1 and extract output from the zone 1.
- 3) Eluent input in the zone 1 and raffinate output from the zone 3.
- 4) Recirculation loop with all zones connected in series.

After one step, each zone is shifted by one column: the zone 1 becomes the zone 2, the zone 2 becomes the zone 3, etc. So, a cycle is performed after 4 steps. ISMB performance depends on different volumes eluted during these four sub-steps.

ISMB simulations were performed with the Langmuir model whose coefficients were optimized previously from an experimen-

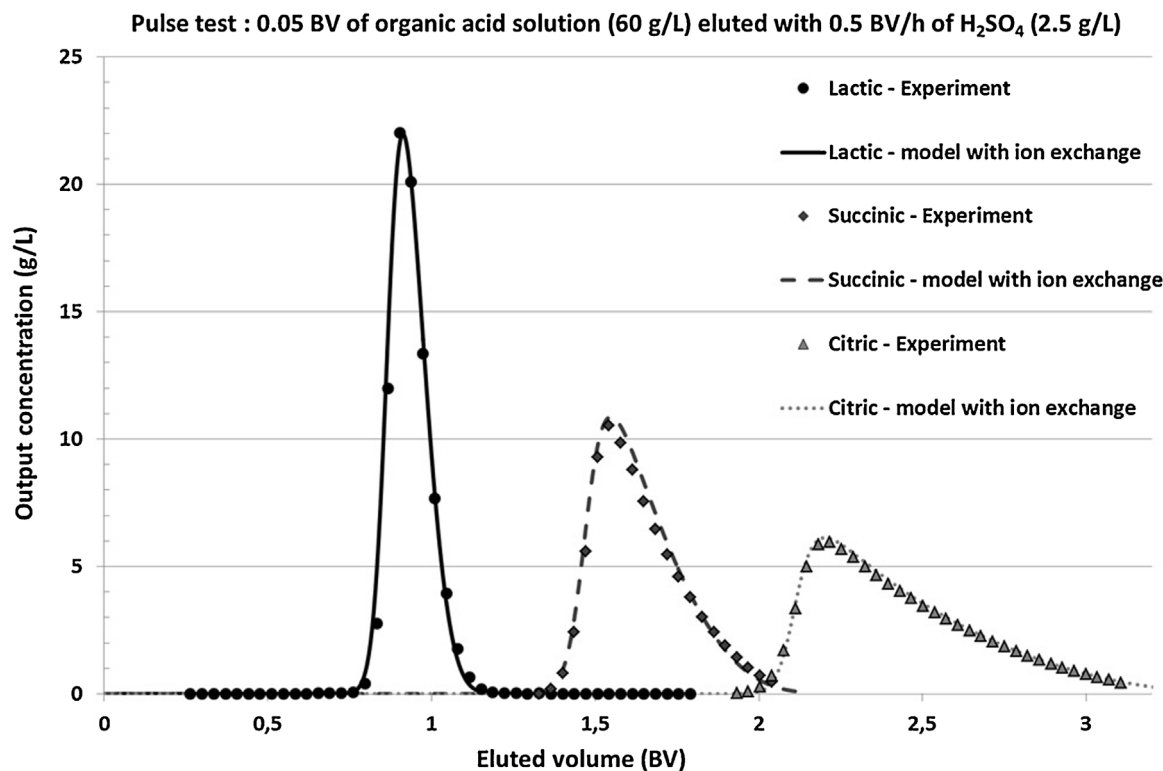


Fig. 3. Comparison of experimental pulse test profiles and simulations ones considering the Langmuir model combined with ion-exchange of conjugate bases.

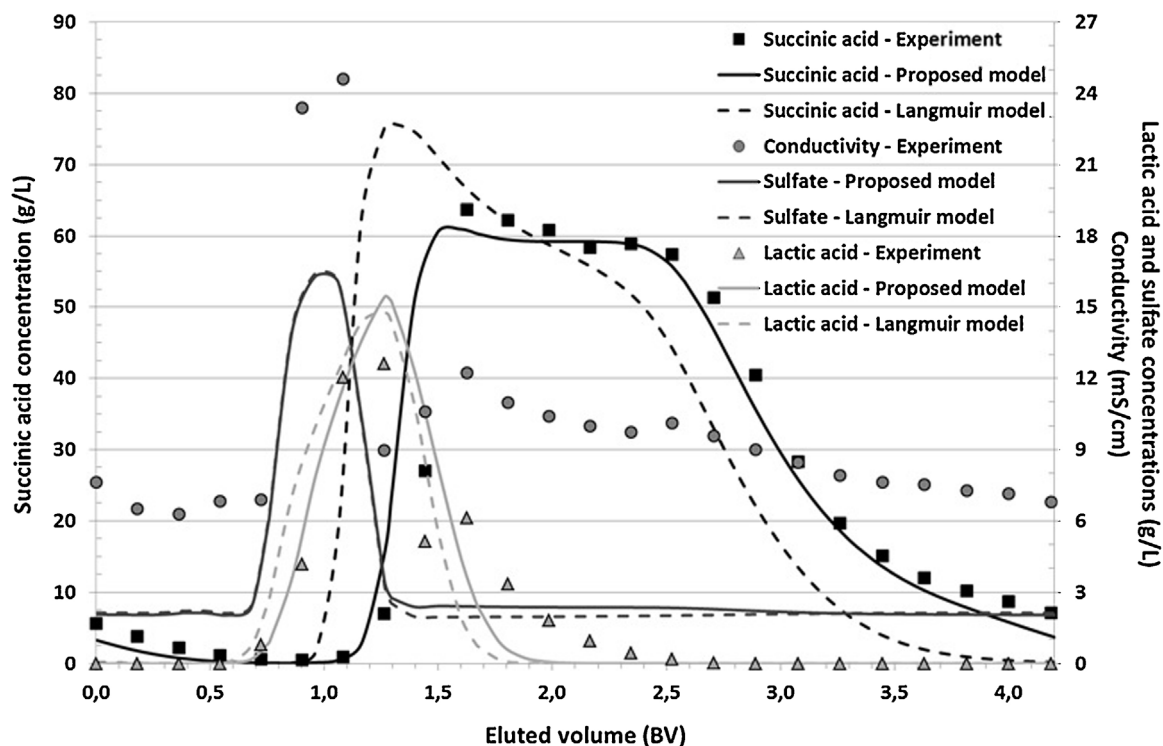


Fig. 4. Comparison of experimental ISMB profiles with simulated ones, first with the model which is developed, then with the Langmuir model only.

tal pulse test at pilot scale, then with the model developed whose coefficients were also optimized in the same way. The feed was mainly composed of sulfate salts (16 g/L), succinic acid (65 g/L) and lactic acid (14 g/L) while the eluent was a 2.5 g/L sulfuric acid solution.

Fig. 4 shows that the proposed model fitted better ISMB experimental profiles than the Langmuir model. The accuracy of ISMB simulation with strong anionic resin was improved by taking into account the retention of dissociated organic acids by ion-exchange, despite their very low concentrations.

4. Conclusion

Preparative chromatography is an efficient separation technique which is spreading in different industrial fields. For several years now, anionic resins have been performed for the citric acid purification, however uptake mechanisms are still not well known. In a previous work [18], peaks corresponding to succinic and citric acids exhibited singular tailing and delay which could not be explained only by the Langmuir model.

The uptake mechanisms supposed in the proposed model are relevant to explain the characteristic elution profiles of some organic acids obtained in pulse tests. To sum up, we assume that the acid molecular form can be retained on resin counteranions or polymer matrix while the dissociated form can be fixed to cationic sites by ion-exchange. Usually, the organic acids dissociation is not considered at such low pH (<1.5), but we proved that even a very small proportion of dissociated form can have a significant impact when it is strongly retained. This is the reason why the Langmuir model is sufficient for lactic acid because of the low affinity of lactate but insufficient for citric acid since citrate affinity is very high for quaternary ammonium anionic resins.

The developed model was implemented in a preparative chromatography (ISMB) simulation program in order to optimize the operating parameters of our pilot-scale unit and improve estimation of its performance. Experimental ISMB profiles confirmed that the proposed model was better than the Langmuir model to describe the organic acids retention on strong anionic resins.

Next, a plug flow reactor model will be used to take into account different axial dispersions of compounds. Moreover, it would be interesting to further investigate so as to support the assumptions which were made to simplify the model and eventually to refine it.

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