Study of Simulated Moving Bed Reactor (SMBR) for \textit{para}-xylene production

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Abstract

In the context of the new Process Intensification paradigm, units that integrate reaction and separation play a decisive role \textsuperscript{(1)}. The viability of the multifunctional Simulated Moving Bed Reactor (SMBR) technology to the \textit{p}-xylene production process was studied through simulation and an original experimental setup, the One-Column Reactor pilot. Simulation of different SMBR configurations demonstrated that the maximal productivity attained by the SMBR can be superior to the SMB’s, according to the feed composition. When accepting a 5% loss in productivity relative to the SMB, it is possible to achieve a 48% decrease in Recycle Ratio for a 6-reactor configuration and 46% for a 3-reactor. Compared with the SMB the Desorbent consumption can be reduced in 2% (6-reactor) or 4% (3-reactor) but theoretically a new desorbent has to be employed that increases distillation costs. Thus far the pilot unit was operated in sole separative mode for a preliminary validation. The current performances were inferior to the expected (63.5% purity and 67.0% yield). However, the storage tank operation and the management of dead volumes still need to be optimized to avoid significant dispersion phenomena.

Keywords: \textit{p}-xylene; Simulated Moving Bed Reactor; One-Column Reactor; reactive separation; flow rate optimization; Process Intensification

1. Introduction

Process Intensification is a growing effort in the Chemical Industry. This recent paradigm searches the optimization of capital, energy, environment and safety by radically reducing the physical size of the plant. In this context, multifunctional reactors, which combine reaction and separation steps in a single unit, constitute a central tool \textsuperscript{(1)}

Currently, the production of \textit{p}-xylene comprises two separate units. The first involves its separation by adsorption from a xylene isomeric mixture using the Simulated Moving Bed (SMB) technology. The second is an isomerization reactor, where \textit{p}-xylene is formed from the \textit{para}-depleted effluent of the separation unit, being however the conversion limited by the chemical equilibrium. The equilibrium mixture is then recycled to the separation unit and constitutes the major part (3/4) of the SMB feed \textsuperscript{(2)}

The Simulated Moving Bed Reactor (SMBR) technology combines chemical reaction with adsorptive separation. The \textit{in-situ} separation of \textit{p}-xylene allows the displacement of the isomerization equilibrium and consequently the reduction of the recycle current, resulting in less equipment and energy costs.

\textbf{Fig. 1.} Effects of the reduction of the recycle current of the \textit{p}–\textit{xylene} process aimed by the integrated SMBR technology.
1.1 The Simulated Moving Bed (SMB) principle

The SMB is a continuous separation adsorption process where the solid adsorbent is fixed in a series of columns. To maximize the driving force, a solid-fluid countercurrent is simulated by the discrete switch of the system’s inlets and outlets along the adsorber in the direction of the fluid flow.\(^{(3)}\)

The SMB implementation follows the concept of relative motion. In a true countercurrent system, the solid in the signaled column in fig. 2 moves downwards in the direction of the Desorbent port in order to flow in the opposite direction of the fluid. In an SMB it is the Desorbent port that is shifted upwards to approach the solid in the said column.

![Fig. 2. First 3 steps of a 12-step Simulated Moving Bed system with preferential adsorption of A.](image)

Most SMB systems have four zones delimited by the inlets and outlets of the process. Each zone has a specific function.

Let A be the preferentially adsorbed species and B the less adsorbed. The function of Zone II is the removal of B from the adsorbent by the desorbent to prevent the contamination of the extract. In Zone III component A must be adsorbed and dragged by the solid so that it does not contaminate the raffinate. The role of Zone IV is the regeneration of the desorbent to be recycled to Zone I. This is achieved by the adsorption of B. Similarly, the role of Zone I is to regenerate the adsorbent before entering Zone IV. This is achieved by the desorption of A.

1.2 The One-Column setup: a simpler SMB alternative

The One-Column system\(^{(4)}\) reproduces the SMB operation with a single adsorptive column and a number of tanks equal to the number of steps in the SMB cycle.

The principle of the simplified technology consists in having one column reproducing the role of one particular SMB column through a complete cycle. To mimic the interaction with the SMB columns upstream and downstream, the effluent exiting the One-Column in step \(n\) is stored in tank \(n\) and reinjected in the column the next cycle, in step \(n-1\). After a certain number of cycles, the One-Column analogue will attain a similar periodic steady-state.

![Fig. 3. Complete cycle for a 4-column SMB and equivalent One-Column.](image)
The One-column analogue has only been studied by simulation but it is a promising substitute for SMB since the equipment costs are significantly lower and the repacking is simpler.

1.3 The Simulated Moving Bed Reactor (SMBR)

The Simulated Moving Bed Reactor technology allows the integration of reaction in the merely separative SMB process. Numerous studies presented in open literature apply the SMBR technology to $A \leftrightarrow B + C$ or $A + B \leftrightarrow C + D$ equilibrated reactions where the reaction products are separated by the adsorbent (5) (6) (7). In these cases the reaction may be integrated in all the system since the in-situ separation of the products avoids the undesired reverse reaction.

For $A \leftrightarrow B$ equilibrate reactions, where the desired compound A is also the most adsorbed, the reaction has to be segregated from the vicinity of the Extract outlet to avoid the conversion of A into B and the consequent contamination of the product. In 1983, Hashimoto et al. (8) developed an original scheme that confined the reaction to zone III, where B is the main component. Purely reactive columns were intercalated with the separation columns in this zone and accompanied the switching of the external streams.

In 2004, Minceva et al. (9) proposed the adaptation of Hashimoto’s scheme to the p-xylene process (Fig. 4). The investigation was limited to simulation, the isomer ethylbenzene was neglected in the feed and p-diethylbenzene was considered as desorbent. Nonetheless, these first results revealed the potential of the SMBR technology for the p-xylene process.

The main objectives of this article are (i) to expose the latest simulation results regarding the viability of the SMBR technology to the production of p-xylene; (ii) to present the current advances on the development of a novel experimental setup, the One-Column Reactor, created for the study of the p-xylene SMBR.

![Fig. 4. Adaptation of Hashimoto’s scheme to the p-xylene process. The arrow represents the direction of the fluid flow and port switch.](image)

2. Modeling the SMBR for p-xylene production

In this work the SMBR model developed by Bergeot (10) was employed. This model considers a 3-scale mass transfer in the adsorbent beds - extragranular phase, adsorbent macropores and adsorbent micropores – and employs the following assumptions:

(i) Axial dispersion plug flow for the bulk phase;
(ii) Linear Driving Force approximation for the macro and microporous transfer rates;
(iii) Adsorption on the external surface of the adsorbent crystallites negligible compared to the internal surface;
(iv) Competitive Langmuir adsorption equilibrium;
(v) Molecules inside the micropores all in the adsorbed phase;
(vi) Particle geometry and porosity constant along the column and identical internal fluxes in all the grain;
(vii) Isothermal operation;
(viii) Negligible pressure drop;
(ix) Constant velocity.
The reactors description assumes ideal plug flow of the fluid phase and neglects diffusion and adsorption phenomena in the kinetics of the reaction. The liquid phase isomerization is modeled according to the triangular scheme proposed by Cappellazzo and co-workers.\(^{(11)}\)

![Fig. 5. Simplified reaction scheme representing the xylene isomerization reactors.](image)

In this model the SMBR periodic steady state is computed by an equivalent True Moving Bed Model (TMBR) approach.

The SMBR model equations were implemented in the free scientific software Scilab\(^{(10)}\) and were solved numerically by the DASSL subroutine\(^{(12)}\), a widely used code for the resolution of ordinary differential-algebraic systems of equations (ODAEs).

To use DASSL the model equations were transformed by introducing dimensionless variables and the systems of partial differential and algebraic equations (PDAEs) were reduced into ODAEs in time through the discretisation of the axial domain.

Table 1 summarizes the main model parameters used in the simulations.

**Table 1.** Main model parameters used in SMBR simulations (R = nr. reactors).

<table>
<thead>
<tr>
<th>Number of adsorbent beds</th>
<th>24</th>
</tr>
</thead>
<tbody>
<tr>
<td>Configuration</td>
<td>5-9-(7+R)-3</td>
</tr>
<tr>
<td>Switching time</td>
<td>71 s</td>
</tr>
<tr>
<td>Temperature</td>
<td></td>
</tr>
<tr>
<td>Adsorbers</td>
<td>200 °C</td>
</tr>
<tr>
<td>Reactors</td>
<td>300 °C</td>
</tr>
<tr>
<td>Feed composition</td>
<td></td>
</tr>
<tr>
<td>PX</td>
<td>0.24</td>
</tr>
<tr>
<td>MX</td>
<td>0.31</td>
</tr>
<tr>
<td>OX</td>
<td>0.34</td>
</tr>
<tr>
<td>EB</td>
<td>0.11</td>
</tr>
</tbody>
</table>

### 3. Simulation Results

#### 3.1 Operating Regions for a 6-reactor SMBR

The operating region that guarantees minimum levels of extract purity and yield was determined by simulation of the SMBR system for multiple pairs of extract and feed flow rates. At this step, the desorbent and pumparound flow rates were fixed (D=100cm\(^3\)/min; PA=176cm\(^3\)/min).

Purity and yield were defined according to:

\[
\text{pur}_X = \frac{x_{PX}}{x_{PX} + x_{MX} + x_{OX} + x_{EB}} \quad (1)
\]

\[
Y = \frac{x_{PX} Q_X}{(x_{PX}^2 + x_{MX}^2 + x_{OX}^2) Q_F} \quad (2)
\]

![Fig. 6. Extract-Feed regions for different demands on Yield for the 6 reactor SMBR configuration (thin lines) versus the SMB triangle (thick lines).](image)

Firstly, Fig. 6 reveals the influence of the reaction on the purity attained in the extract. In the SMB configuration the adsorbent is occupied by MX, OX and EB. To force the desorption of these impurities, the extract flow rate must be reduced to increase zone 2 velocity. For the SMBR configuration, the PX produced in the reactors occupies the liberated capacity and the required decrease in the extract rate is smaller.

With respect to the SMBR yield behavior it is specially interesting to observe that, to achieve an augmentation, the feed flow rate has to be reduced. This means that the productivity will fall with the increase of yield.

#### 3.2 Influence of the Feed Composition on the operating regions

Fig. 7 compares the triangle tips of the SMB and 6-reactor SMBR for (a) a feed with PX over the equilibrium composition (Table 1) and (b) a feed with PX under the equilibrium composition\(^{(13)}\). It is possible to conclude that with the over-equilibrium PX feed the maximum productivity achieved by the SMBR is inferior to the SMB, but with the under-equilibrium PX feed the maximum productivity achieved by the SMBR becomes superior.
This phenomenon is caused by the first reactor. As it is close to the feed inlet, in the first case it receives a PX over-equilibrium mixture and consumes PX instead of producing. The reduction of PX around the feed inlet favors the adsorption of impurities. Thus in this case, the SMBR extract rate has to be reduced to increase zone II velocity and attain the desired purity. The consequence is the said decrease in productivity.

3.3 Effects of the flow rates on the Productivity and Recycle Ratio

Fig. 8 compares the evolution of the SMB and 6-reactor SMBR productivities with the decrease of feed rate for a 99.8% purity. It can be observed that, except for extreme feed rates, the SMBR productivity surpasses the SMB due to the almost flat top of the SMBR productivity curve. The reason for the minor productivity decrease is the production of PX in the 6 reactors, which compensates for the reduction of feed. The practical consequence of this behavior is that with a slight reduction in PX production, the xylenes fed could be strongly diminished being therefore expected a significant decrease in the raffinate flow rate.

3.4 Influence of reactor number and position

To study the possibility of reducing costs certain reactors were eliminated. Table 2 presents the configurations tested.

Table 2. SMBR configurations studied.

<table>
<thead>
<tr>
<th>Zone 3 configuration</th>
<th>1 R</th>
<th>2 R</th>
<th>3 R end</th>
<th>3 R intercalated</th>
<th>4 R end</th>
<th>4 R intercalated</th>
<th>5 R</th>
</tr>
</thead>
<tbody>
<tr>
<td>4+R+3</td>
<td>2+R+2+R+3</td>
<td>4+R+1+R+1+R+1</td>
<td>2+R+2+2+R+1</td>
<td>3+R+1+R+1+R+1+R+1</td>
<td>2+R+2+R+1+R+1+R+1</td>
<td>2+R+1+R+1+R+1+R+1+R+1</td>
<td></td>
</tr>
</tbody>
</table>

The operating regions defined by a 26.2% yield are depicted in Fig. 10 for these configurations.
It is interesting to note the shape of the purity lines. In fact all these lines start by being almost vertical but under a certain value of feed bend to the interior of the triangle. Except for the configurations where the reactors are placed at the end, with the increase in the number of reactors the length of the straight part increases and the curvature becomes softer. This happens since more PX is created. As explained, the PX created occupies the adsorbent capacity liberated by the reduction of feed and prevents the adsorption of impurities, maintaining similar purity levels. Under the said values of feed, the PX formation is limited and the adsorption of impurities increases, demanding lower extract rates.

The 3R_end and 4R_end regions reveal the importance of the reactor positioning. In these cases the purity line curvature is more accentuated than the corresponding intercalated configurations. This happens because the PX production is distant from the feed inlet and thus the created PX cannot avoid the adsorption of the impurities entering the process.

With respect to the Recycle Ratio reduction, as less PX is formed with the elimination of reactors, the gains relative to the SMB diminish as well, as may be confirmed in Table 3 for a productivity 5% less the SMB value.

These results show that the passage from SMB to SMBR allows a significant decrease in fraction of feed treated in the recycle loop. Also, it can be stated that the difference in recycle ratio between the configurations 3R_intercalated, 4R_intercalated, 5R and 6R is not superior to 6%.

Table 3. Recycle Ratios for each reactor configuration with a 5% loss in Productivity compared to the SMB process.

<table>
<thead>
<tr>
<th></th>
<th>1R</th>
<th>2R</th>
<th>3R_end</th>
<th>3R_intercalated</th>
<th>4R_end</th>
<th>4R_intercalated</th>
<th>5R</th>
<th>6R</th>
<th>SMB</th>
</tr>
</thead>
<tbody>
<tr>
<td>RR</td>
<td>2.25</td>
<td>1.98</td>
<td>1.92</td>
<td>1.78</td>
<td>1.77</td>
<td>1.72</td>
<td>1.67</td>
<td>1.70</td>
<td>3.28</td>
</tr>
</tbody>
</table>

3.5 Minimization of the Desorbent requirement

The minimization of the Desorbent rate was applied to the 3R_intercalated and 6R configurations. The depart conditions corresponded to a 99.8% purity and a 5% loss in productivity relative to the SMB. For comparison, the same optimization procedure was applied to the SMB at maximal productivity.

Fig. 11 presents, for these points, the operating regions for the Desorbent and Pumperaround flow rates defined by the minimum levels of purity and yield accepted. These plots demonstrate that the Desorbent minimization achieved with the passage from SMB to SMBR is modest for the 6R configuration (less than 2%) but slightly more interesting in the 3R_intercalated case (4%).
Comparing to the 6R configuration, the reduction attained with the 3R<sub>intercalated</sub> is more important due to the superior separation efficiency in zones II and III. At high productivity rates, the adsorbent capacity is close to its limit. Reducing the number of reactors reduces the PX produced in zone III and thus facilitates the separation. Since the 3R<sub>intercalated</sub> performances in zones II and III are superior, the difference until the minimum levels of purity and yield is larger. Therefore, it is possible to reduce more the desorbent rate.

In order to substitute the SMB p-xylene process for a SMBR technology, the Desorbent should be changed from p<sub>DEB</sub> to toluene. As with the xylenes, p<sub>DEB</sub> isomerizes and this would difficult the SMBR operation.

Relative to the SMB processes operated with toluene the SMBR seems advantageous but the same may not be true when the SMB desorbent is p<sub>DEB</sub>. Contrarily to p<sub>DEB</sub>, toluene is lighter than the xylene isomers. This change in volatility results in larger energy consumption in the Extract-Desorbent and Raffinate-Desorbent distillations since the desorbent, which is in greater proportion, is now recovered in the distillate.

4. Description of the One-Column Reactor pilot unit

The One-Column Reactor pilot was developed from the One-Column concept<sup>(4)</sup> by placing a catalytic reactor downstream to the adsorption column. When operating in the reactive mode, the effluents of the adsorption column resulting from Zone III passages enter the isomerization reactor. During the solely separative steps, a bypass allows the reactor not to be used.

The adsorber is operated in liquid phase at 175°C and approximately 10 bar; 200°C and 30 bar when the reaction is integrated. The adsorbent used is BaLSX, a low silica X zeolite exchanged with barium.

The reactor is packed both with HZSM-5 zeolite as catalyst and an inert solid. It is operated in liquid phase at 300°C and 30 bar.

24 storage tanks serve to recover the effluents from the adsorption column, or the reactor. A last tank acts as a reserve of desorbent to compensate variations in the other 24 tanks.

An automatic sampler allows the determination of the concentration profiles exiting the column. The entire operation of the pilot is controlled by an automate accessible through an interface.

The feed of the process is a mixed xylene charge. p-Diethylbenzene and Toluene were consecutively used as desorbents.

5. Experimental Results

The current advances in the implementation of the One-Column Reactor pilot still concern its operation in sole separative mode. The aim is the pilot’s preliminary validation by comparison to the results of an existing SMB pilot.

5.1 Results from the operation with 12 equivalent adsorbent beds, p<sub>DEB</sub> and simple tank level

The operation of the One-Column Reactor with an equivalent 12-bed configuration allows splitting the effluent of each column passage by 2 storage tanks. If the tanks have an ideal degree of mixture, important performance gains are achieved when the effluents were divided<sup>(4)</sup>. The level contained in each tank corresponded to the quantity necessary for one step, calculated
according to the respective flow rate and step period or switching time ($t_s$). This simple level corresponds, in a 12 bed configuration, to storing a half profile in each tank. An extra 50 cm$^3$ volume was added in each tank to ensure that the pump remained primed.

Fig. 12 presents the system concentration profile obtained in the beginning (continuous lines) and end (dots) of this validation test. Table 4 presents the corresponding performances.

### Table 4 Initial and final performances of the OCR unit in the first validation test

<table>
<thead>
<tr>
<th></th>
<th>Initial</th>
<th>Final</th>
</tr>
</thead>
<tbody>
<tr>
<td>Purity (%)</td>
<td>28.0</td>
<td>64.4</td>
</tr>
<tr>
<td>Yield (%)</td>
<td>11.6</td>
<td>55.9</td>
</tr>
</tbody>
</table>

The final performances were still far from the industrial 99.9% purity and 97% yield. In the pilot unit, much of the PX fed in bed nr.7 was still driven to zone 3, being lost in the raffinate. Also, there is a non negligible amount of impurities passing from zone 3 to 2, impeding the increase of the extract.

The approach to the permanent state was evaluated by mass balances on all the components. In all the analyses performed in this test, none of them closed even with one week stabilization times.

5.2 Results from the operation with 12 equivalent adsorbent beds, TOL and double tank level

A double tank level was employed in this test to avoid the utilization of an extra volume of security. Residence Time Distribution tests revealed that the tanks’ hydrodynamic was closer to that of a piston and, thus being, the extra length stores in fact the end of the previous profile resulting in the injection of a profile in the wrong order.

For the future integration of the reaction in the process, the desorbent was changed to toluene. Literature results $^{(14)}$ showed that the desorbent requirement increases when using toluene.

The final performances (Table 5) were inferior to the previous test, which means that the new flow rates were not the adequate for the new desorbent.

### Table 5 Final OCR performance in the second validation test

<table>
<thead>
<tr>
<th></th>
<th>Final</th>
</tr>
</thead>
<tbody>
<tr>
<td>Purity (%)</td>
<td>64.7</td>
</tr>
<tr>
<td>Yield (%)</td>
<td>27.3</td>
</tr>
</tbody>
</table>
Also noteworthy is the fact that doubling the tank level, i.e. increasing the rigor of the operation, did not reflect positively in the performances. On the contrary, this modification resulted in a slower system stabilization because the column effluents were only reinjected after two cycles.

5.3 Results from the operation with 24 equivalent adsorbent beds, TOL and double tank level

For the future integration of the reaction a 24-bed configuration is more adequate since the 12-bed only permits 2 passages in the reactor per cycle. As the flow rates were kept the same as in the previous test, doubling the adsorbers also implied doubling of the treatment capacity, from which one expected an improvement in the separation performance.

Fig. 13 presents the system concentration profile obtained in the beginning (continuous lines) and end (dots) of this validation test. Table 6 presents the corresponding performances.

![Table 6 Initial and final performances of the OCR unit in the third validation test](image)

<table>
<thead>
<tr>
<th>Purity (%)</th>
<th>Initial</th>
<th>Final</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>61,0</td>
<td>63,5</td>
</tr>
<tr>
<td>Yield (%)</td>
<td>40,1</td>
<td>67,0</td>
</tr>
</tbody>
</table>

Despite doubling the separation capacity relative to the previous test, the performances in the beginning of the test did not behave as desired and the concentration profile (continuous lines) remained shifted to the raffinate.

The displacement of PX to the extract and the higher plateau in zone 2 achieved by the end of the test were due to augmentations of the extract rate up to the limit when purity started to be affected. In the last concentration profile we can observe in zone 3 we can observe the establishment of the EB, MX and OX plateaus. In zone 2, it is also clear the establishment of the PX plateau and the fact that the increase of the impurity trail was proportionally smaller than the PX augmentation.

The small influence of zone 2 deceleration in the impurities’ trail suggests their presence in zone 2 might be caused not only by the solid drag but also by the tanks and the important number of lines, valves and fittings existing in this pilot, which aggravate the dispersion.

Again in this test, any mass balance closed. For a difference similar to the first test more than two weeks were needed. On the other hand, the performance gain relative to the first test was small. These facts suggest that the measures taken to improve the separation efficiency (24 beds; double tank level) were not useful due to less favorable response time/benefit ratio.

![Table 7 Final flow rates (g/min) and switching time (s) of each validation test](image)

<table>
<thead>
<tr>
<th>12 bed pDEB</th>
<th>12 bed TOL</th>
<th>24 bed TOL</th>
</tr>
</thead>
<tbody>
<tr>
<td>Simple level</td>
<td>Double level</td>
<td>Double level</td>
</tr>
<tr>
<td>D</td>
<td>26,10</td>
<td>55,20</td>
</tr>
<tr>
<td>X</td>
<td>12,96</td>
<td>26,30</td>
</tr>
<tr>
<td>F</td>
<td>8,56</td>
<td>17,10</td>
</tr>
<tr>
<td>PA</td>
<td>93,20</td>
<td>171,71</td>
</tr>
<tr>
<td>t_s</td>
<td>262,8</td>
<td>142</td>
</tr>
</tbody>
</table>

![Fig. 13. Comparison between the concentration profiles from the beginning of the test – continuous lines - and the end – dots.](image)
6. Conclusions

The application of the Simulated Moving Bed Reactor technology to the production of \(p\)-xylene was proposed. The study here presented included (a) the simulation of different SMBR configurations aiming the optimization of the operating conditions and (b) the operation of a One-Column pilot unit for its validation as a SMB studying tool.

Firstly, the simulation results revealed that a 48% reduction in the SMB Recycle ratio relative to the SMBR could be achieved accepting only a 5% loss in productivity. In addition, the simulation results showed the influence of the feed source on the viability of the SMBR over the SMB. Nonetheless, only one feed source is disadvantageous for the SMBR. The minimization of desorbent consumption revealed that the gains in toluene achieved by the passage from SMB to SMBR were between 2% and 4%. However relative to the SMBS using \(p\)-diethylbenzene the SMBR with toluene is in disadvantage due to its higher distillation costs. Up to date, the impossibility of using \(p\)-diethylbenzene in the SMBR has not been proved.

Finally it was seen that the performances of an SMBR system with 3 reactors regularly spaced and a 6-reactor system differed in the order of 2%.

The One-Column Reactor pilot unit was for the moment operated in a sole separative mode for its preliminary validation. The current performances were inferior to the expected (63.5% purity and 67.0% yield). However, the storage tank operation and the management of dead volumes still need to be optimized to avoid significant dispersion phenomena. The attempts to enhance the separation efficiency with a 24-bed configuration and double tank level required significant stabilization times and the performance improvements were minor.

7. Nomenclature

- D: Desorbent
- EB: Ethylbenzene
- F: Feed
- \(k\): Kinetic constant (s\(^{-1}\))
- MX: \(m\)-Xylene
- OCR: One-Column Reactor
- OX: \(o\)-Xylene
- PA: Pumparound
- pDEB: \(p\)-Diethylbenzene
- PX: \(p\)-Xylene
- Q: Flow rate (kg/s)
- R: Reactor
- RR: Recycle Ratio
- SMB: Simulated Moving Bed
- SMBR: Simulated Moving Bed Reactor
- TMBR: True Moving Bed Reactor
- TOL: Toluene
- \(t_s\): Switching time (s)
- \(x\): Adimensionalized fluid concentration
- Y: Yield (%)

Subscripts and superscripts

- D, X, F, R: Desorbent, Extract, Feed, Raffinate

8. References