New Control Strategies for Quality of the Separated Products of a Butylene-Butane Distillation Column

Internal Model Control Algorithm

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This paper presents the design of an advanced control system for products quality of a butylene-butane distillation column (BBDC). After thorough analysis of the current control structures under the UNISIM® simulation program and using steady-state performance criteria, the LB control structure was chosen for BBDC. For the development of the automatic control systems for the products quality the multicomponent distillation process was adapted to a pseudo-binary distillation process. Starting from this assumption, there were obtained dynamic models of the process for the input-output channels reflux flowrate – concentration of the pseudo-light component and bottom product flowrate – concentration of the pseudo-heavy component. For concentration controllers associated with the two separated products was used an original Internal Model Control algorithm. The dynamic simulations of the proposed advanced control structure under the SIMULINK® program proved that this technique is more suitable for the studied column than the conventional decentralized PI control for step changes in setpoints and disturbances.

Keywords: distillation column, multicomponent distillation, control structure selection, internal model control, dynamic simulation

Distillation remains the most chosen separation method for liquid mixtures - with over 40,000 columns in operation around the world [1]. The major drawback is the large energy requirements, as distillation can generate more than 50% of plant operating cost [2].

In previous researches [3-6] most of the work has been focused on dual composition control for binary distillation columns with inferential control or online measurements. Control strategies which are suitable for binary distillation columns can be applied to multicomponent distillation columns if their behaviour is not highly nonlinear [7]. The importance of the product quality control loop for a distillation column is always written in the production program in a refinery. In recent years, decentralized PI and MPC control strategies are presented for both binary and multicomponent columns [8 - 12].

The butylene-butane distillation column hosts a multicomponent distillation process. The control of a multicomponent column is more difficult than the control of a conventional binary distillation column because there is more interaction among control loops. Multicomponent distillation columns have been well known to have a difficult controller design problem, resulting from its characteristics, such as complex dynamics, high nonlinearity and interaction between the control loops [13].

The advanced control strategy applied to distillation columns that made an impact on the industrial scale is the Model Predictive Control with its variant Internal Model Control IMC [14-18]. The purpose of the paper is to properly implement the IMC technique which is economically important due to the controller effect on product quality, production rate, and energy usage.

Starting from the industrial process, the authors had studied the possibility of implementing products quality control systems for the butylene-butane distillation column, using the internal model control algorithm techniques.

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Experimental part

Butylene-butane mixture distillation process

The industrial process

Butylene-butane distillation process takes part of the gas concentration (GASCON) unit, an important element of the catalytic cracking plants from a refinery. In table 1 are presented some characteristics of the butylene-butane distillation column from a catalytic cracking plant.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Column height</td>
<td>48.8 m</td>
</tr>
<tr>
<td>Column diameter</td>
<td>3.5 m</td>
</tr>
<tr>
<td>Number of trays</td>
<td>100</td>
</tr>
</tbody>
</table>

The feed of the column is the C4 fraction from the bottom of the depropanizer column, and can be introduced into the column through several inlets, on trays 35, 43 or 51, depending on the raw material composition. The overhead product is calor gas composed of i-butane, i-butylene, and the bottom product is a mixture of n-butane, cis-butene and trans-butene. In figure 1 is presented the scheme of the C4 separation column. The feed, both in flowrate (fig. 2) and concentration (fig. 3), has important variations, which contributes to obtaining products with different qualities from the required specifications.

Quality specifications for separated products from multicomponent distillation columns

The quality of the separated products from a distillation column is expressed by the chemical concentration of the distillate, \( x_{ci} \), \( i = 1,...,nc \), and of the bottom product, \( x_{bi} \), \( i = 1,...,nc \), where \( nc \) is the number of chemical components in the mixture. This way of specifying the quality is...
customized according to the distillation column type. In the following will be analyzed the specifications for binary fractionating columns (ethylene-ethane, propylene-propane) and also for multicomponent distillation columns (propane-butane, butylene-butane).

Usually, for binary mixtures, the chemical composition is expressed by the concentration of the most volatile component, which leads to distillate quality expressed as $x_D = x_{D1}$ and the bottom product quality as $x_B = x_{B1}$. In these conditions, the quality control of one or both products means the concentration control of the most volatile component from the mixture, the quality specifications having the form $x_i^D = x_{iD1}$ and, respectively $x_i^B = x_{iB1}$. The index $i$ refers to the setpoint of the automatic system for product quality control.

For multicomponent mixtures, the problem of the quality of the separated products in the distillation column has a different approach. In this case, quality specifications like $x_i^D = x_{iD1}$ or $x_i^B = x_{iB1}$ cannot be applied. Instead, is used the concentration of an impurity in distillate and (or) in bottom product [19]. The impurity (concentration of the unwanted component in the product) can be measured more accurately than the purity (concentration of the desired component in the product), which would allow superior performances of the automatic control systems for product quality. However, this approach cannot be generalized. In Table 2 are presented the quality specifications for the separated products in an industrial butylene-butane distillation column (the distillate and the bottom product). Characteristic of these specifications is that there are not fixed values imposed for the chemical components concentrations and the sum of the concentrations is not 100 mole %. In this case, the concentration control of a chemical component in distillate or in bottom product is not an automation solution.

### Analysis of the current control structure

The distillation column from figure 1 has five control systems associated with the first hierarchical level of automatic control. From these five systems two are level control systems (in the reflux tank and in the column bottom) and the third system is for column pressure control. The last two automatic systems are used for separated product quality control. In reality, the reflux flowrate and condensate flowrate control systems are not included in some automatic systems for quality control for the separated product, the quality specifications being assured exclusively by the operators and is guided only by laboratory analyses, with a frequency of 3 analyses/day. Consequently, for the studied distillation column, the quality of the separated products is not controlled.

The distillation process is a multivariable system and normally the quality of the products can be controlled using a multivariable control system. However, in practice, this

<table>
<thead>
<tr>
<th>Component</th>
<th>Specification</th>
</tr>
</thead>
<tbody>
<tr>
<td>Name</td>
<td>Type</td>
</tr>
<tr>
<td>i-butane</td>
<td>desired</td>
</tr>
<tr>
<td>i-butylene</td>
<td>desired</td>
</tr>
<tr>
<td>n-butane</td>
<td>impurity</td>
</tr>
<tr>
<td>cis-butene</td>
<td>impurity</td>
</tr>
<tr>
<td>trans-butene</td>
<td>impurity</td>
</tr>
<tr>
<td>i-butane</td>
<td>impurity</td>
</tr>
<tr>
<td>i-butylene</td>
<td>impurity</td>
</tr>
<tr>
<td>n-butane</td>
<td>desired</td>
</tr>
<tr>
<td>cis-butene</td>
<td>desired</td>
</tr>
</tbody>
</table>

### Table 2

**Quality Specifications for a Butylene-Butane Distillation Column**
automation solution is not widespread due to the process nonlinearity. Often is preferred the consideration of the distillation process as a system composed by two mono-variable subsystems, which allows the design of a control system for distillate quality and another for bottom product quality. The interaction between these two subsystems depends on the pair of variables selected as control signals within the two subsystems. In table 3 are presented pairs of variables that can be used as control signals.

The selection of the pair of variables which assure the lowest degree of interaction is done using the relative gain array (RGA). An element of RGA represents the ratio of the steady-state gain between the controlled variable $y_i$ and control agent $u_j$ when the other control agents are constant and the steady-state gain between the same variables when the other controlled variables are constant [20]:

$$\lambda_{y} = \left(\frac{\partial y_i / \partial u_j}{\partial y_i / \partial u_j}\right)_{y_i}$$  \hspace{1cm} (1)

Calculation of the RGA elements for the pairs of variables from table 3 allows the identification of that pair of variables which can lead to the lowest degree of interaction and which will assure an efficient control of the separated products quality.

**Mathematical modeling and numerical simulation of the distillation column**

The development of the mathematical control model for a distillation column requires a study to highlight the steady-state and dynamic response of the distillation column to changes of the control signals and disturbances.

**Steady-state modeling**

For steady-state modeling and simulation of the distillation process, there was used the UNISIM® program. In figure 4 is represented an image of the simulator for the steady-state of the industrial butylene-butane distillation column. The simulator was used to determine the RGA elements (1). The numerical values of the derivatives can be evaluated as ratios between the steady-state variations of the concentrations and the variations of the control signals used in the product quality control systems. The obtained results are presented in table 4. Taking into consideration the closest value to 1 of the $\lambda_{y}$ element from the RGA, but also the industrial implementation criterion for the distillation column presented in figure 1, resulted the fact that the best structure for product quality control for the butylene-butane distillation column is LB [21].

Consequently, as a result of steady-state modeling and simulation of the distillation process, a new structure was selected in order to implement products quality control systems (fig. 5).

**Dynamic modeling**

For the dynamic simulation of the butylene-butane distillation column was used the mathematical model from Skogestad [13]. This dynamic model is dedicated for binary distillation columns and is based on the following assumptions: constant pressure; constant relative volatility; total condensation; constant molar flowrates;

### Table 3

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Control signal for distillate quality control</th>
<th>Control signal for bottom product quality control</th>
</tr>
</thead>
<tbody>
<tr>
<td>LB</td>
<td>Reflux flowrate (L)</td>
<td>Bottom product flowrate (B)</td>
</tr>
<tr>
<td>LV</td>
<td>Reflux flowrate (L)</td>
<td>Flowrate of the heating agent from the reboiler (V)</td>
</tr>
<tr>
<td>SV/B</td>
<td>Reflux ratio (S)</td>
<td>The ratio between the heating agent from the reboiler and the bottom product flowrate (V/B)</td>
</tr>
<tr>
<td>DV</td>
<td>Distillate flowrate (D)</td>
<td>Flowrate of the heating agent from the reboiler (V)</td>
</tr>
<tr>
<td>DV/B</td>
<td>Distillate flowrate (D)</td>
<td>The ratio between the heating agent from the reboiler and the bottom product flowrate (V/B)</td>
</tr>
</tbody>
</table>

Fig. 4. UNISIM program used for distillation column simulation
negligible vapour holdup; linearized dynamic of liquid flowrate.

Given the problem of quality specifications and the particularities of the mathematical model, the multicomponent mixture was assimilated to a binary mixture, table 5.

<table>
<thead>
<tr>
<th>Chemical component</th>
<th>Pseudo-component</th>
<th>Concentration in distillate</th>
<th>Concentration in bottom product</th>
</tr>
</thead>
<tbody>
<tr>
<td>i-butane</td>
<td>light</td>
<td>$x_D$</td>
<td>$x_B$</td>
</tr>
<tr>
<td>i-butylene</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>n-butane</td>
<td>heavy</td>
<td>$1 - x_D$</td>
<td>$1 - x_B$</td>
</tr>
<tr>
<td>cis-butene + trans-butene</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Skogestad’s model was adapted for the butylene-butane distillation column, taking into consideration the pseudo-binary mixture defined in table 5.

Within the dynamic modeling and simulation one purpose was the determination of some simplified dynamic models of the distillation process, models which will be used within the algorithms for product quality control. The process outputs (the concentration of the light pseudo-component in distillate $x_D$ and the concentration of the light pseudo-component in bottom product $x_B$) are dependent on the four inputs (reflux flowrate $L$, bottom product flowrate $B$, feed flowrate $F$, and concentration of the light pseudo-component in feed $x_F$), according to figure 6.

In figures 7-8 are presented dynamic responses of the outputs, $x_D$ and $x_B$, generated by step changes of the inputs $L$ and $B$. The initial steady-state of the process is defined by: feed flowrate - 4.34 kmole/min; feed concentration -

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**Table 4**

VALUES OF THE RGA ELEMENTS FOR DIFFERENT CONTROL STRUCTURES

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Value of the $A_{ij}$</th>
<th>Observations</th>
</tr>
</thead>
<tbody>
<tr>
<td>LB</td>
<td>0.5841</td>
<td>Structure suitable for industrial implementation</td>
</tr>
<tr>
<td>LV</td>
<td>41.8353</td>
<td>Very large interaction</td>
</tr>
<tr>
<td>SV/B</td>
<td>2.7674</td>
<td>Large interaction</td>
</tr>
<tr>
<td>DV</td>
<td>0.4905</td>
<td>Poor structure in terms of level control in the reflux tank</td>
</tr>
<tr>
<td>DV/B</td>
<td>0.5328</td>
<td>Poor structure in terms of level control in the reflux tank</td>
</tr>
</tbody>
</table>

**Table 5**

CONSIDERATION OF THE MULTICOMPONENT MIXTURE AS A PSEUDO-BINARY MIXTURE

![Figures 5-8](image-url)
Based on the time responses of the outputs to step changes of the inputs have been determined dynamic models on each input-output channel:

\[ T_1 T_2 \frac{d^2 \Delta y}{dt^2} + (T_1 + T_2) \frac{d \Delta y}{dt} + \Delta y = K_p \left( T_1 \frac{d \Delta u}{dt} + \Delta u \right), \]  

(2)

where: \( \Delta y \) is the output variation (\( \Delta x_D \) or \( \Delta x_B \)), \( \Delta u \) - the input variation (\( \Delta L \) or \( \Delta B \)), \( K_p \) - the process gain, \( T_1 \), \( T_2 \) and \( T_3 \) - time constants.

The transfer functions associated to the differential equations (2) have the form

\[ G_p(s) = \frac{K_p(T_2 s + 1)}{(T_1 s + 1)(T_2 s + 1)}. \]  

(3)

The determination of the time constants and of the gain was done using numerical and graphical identification techniques, the values of these parameters for channels \( L \rightarrow x_D \) and \( B \rightarrow x_B \) being presented in table 6.

**Table 6 MODELS PARAMETERS**

<table>
<thead>
<tr>
<th>Input-output channel</th>
<th>Parameters</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>( K_p )</td>
</tr>
<tr>
<td>( L \rightarrow x_D )</td>
<td>0.2635</td>
</tr>
<tr>
<td>( B \rightarrow x_B )</td>
<td>1.3992</td>
</tr>
</tbody>
</table>

The validation of the dynamic models was realized by comparing the simulated process dynamic with the dynamic obtained using model (3), for step changes of the inputs. The standard deviation between the simulated process response and the model response has values in \([10^{-7} ... 10^{-5}] \) mole fr. The very small values of the standard deviation validate the dynamic models based on the transfer function (3), and these models can be used in any other analysis of this process.

In conclusion, the dynamic modeling and simulation of the distillation process had as main result the determination of the dynamic models on the control signals - outputs channels, models which will be used for the design of the controllers for products quality and also for the validation through simulation of the proposed control system.

**The proposed control system**

Automatic system design

In order to develop the mathematical control model and the control structure the authors considered the following requirements of the automatic system and process particularities:

- the consideration of the mixture to be separated as a binary mixture;
- the use of quality specifications specific to binary mixtures;
- the use of LB structure as a guarantee of low degree of interaction between control signals;
- quality control of one product (distillate, bottom product);
- the use of an internal model controller.

For the butylene-butane distillation column were proposed the control structures from figure 9. The structures have a hierarchical organization with two levels: at level 1 is the conventional automation according to figure 5 and at the second level is implemented an internal model controller for quality control of the distillate (fig. 9a) or the bottom product (fig. 9b).

**Development of the control algorithm**

According to the formulated requirements, it was chosen an internal model controller for the implementation of the second level of hierarchical control. This control algorithm represents an alternative to the proportional-integral-derivative (PID) algorithm and contains in its structure a model of the controlled process, materialized by transfer function \( G_m(s) \), figure 10.

It is considered that the process model implemented in the controller, \( G_m(s) \), defines a stable and overdamped process and has the transfer function [22]

\[ G_m(s) = \frac{K_m}{(T_m s + 1)^2}, \]  

(4)

where: \( K_m \) is the gain and \( T_m \) - model time constant.

The model parameters are determined as follows [22]:

a) the gain \( K_m \) is equal to the process gain \( K_p \)

\[ K_m = K_p; \]  

(5)

b) the model time constant \( T_m \) is considered to be the sixth part of the process step response settling time \( T_s \)

\[ T_m = \frac{T_s}{6}. \]  

(6)

The assumption that the process is stable and overdamped is not found in the process response at reflux flowrate change, the response being in this case underdamped and with overshoot, figure 7. Because the standard method of internal model control is applicable...
only to a overdamped process, it was necessary the introduction in series with the process of a filter
\[ G_f(s) = \frac{1}{T_f s + 1}. \]  

The tuning of the filter time constant \( T_f \) was done by dynamic simulation of the filter-process ensemble, only for changes in reflux flowrate \( L \), so that the process response to be overdamped (without overshoot). The best response was obtained for \( T_f = 55 \) min, the filter transfer function becoming
\[ G_f(s) = \frac{1}{55 s + 1}. \]

After identifying the compensated process the following model parameters were obtained: \( K_m = 0.2635 \) and \( T_m = 25 \) min. These values will be used as starting parameters in the tuning of the controller associated with the distillate quality control.

On the channel \( B - x_B \), the process response fit the pattern for the use of standard procedure of internal model control. The model parameters resulted from identification are: \( K_{m2} = 1.5098 \) and \( T_{m2} = 77 \) min.

Simulation of the proposed control system

The simulation of the control system had as goal the numerical validation of the proposed automation solution. For this simulation was used SIMULINK®, the simulation program structure being presented in figure 11. The program can simulate both the dynamic of the distillate quality control system and the dynamic of the bottom product quality control system.

For both proposed control systems the goal was the tuning of the internal model controller, so that the system response to be without overshoot, and the transient time to be as small as possible [23].

The tuning of the controller used for distillate quality control

The initial values of the internal model tuning parameters are: \( K_m = 0.2635 \) and \( T_m = 25 \) min. Starting from these values, were done dynamic simulations of the distillate quality control system, in which the tuning parameters were modified individually. In figures 12-13 are presented dynamic variations of the \( x_D \) concentration and the controller output, obtained for different values of the parameter \( K_m \): 0.3; 0.2635; 0.2; 0.1. Similar simulations were done for tuning of the parameter \( T_m \), the tested values being: 35; 25; 22 and 20 min. As a result of the simulations were obtained quasi optimal values for the tuning parameters of the controller used for distillate quality, namely: \( K_R = 1, K_m = 0.2, T_m = 22 \) min.

The tuning of the controller used for bottom product quality control

For this controller the initial values of the internal model tuning parameters are: \( K_{m2} = 1.5098, T_{m2} = 77 \) min. The simulation aimed to study the dynamic behaviour of the control system to changes of the parameters \( K_{m2} \) and \( T_{m2} \). For parameter \( K_{m2} \) were tested the values: 3; 1.5098; 1.4; 1, and the parameter \( T_{m2} \) was tested in the range \([60...85]\) min, figures 14-15. As a result were obtained the best tuning parameters for the controller used for bottom product quality, namely: \( K_R = 1, K_{m2} = 1.4, T_{m2} = 70 \) min.

Testing of the control system to disturbances changes.

The two control systems developed by the authors were tested to step changes of the two disturbances, the feed flowrate \( F \) and feed concentration \( x_F \).
A. The distillate quality control system. The increase/decrease of the column feed flowrate led to a small change in the concentration, but after a while it returns to the initial value, figure 16. A similar situation occurs in the case of feed concentration changes, figure 17.

B. The bottom product quality control system. Changes of the two disturbances lead in this case to much higher transient times than in the previous case, due to the fact that the time for liquid to flow to the column bottom is much higher and the number of trays covered by the liquid flow is also higher. Because each tray represents an accumulation point, with specific delay, the process time constant increases and consequently the transient time increase also. Figures 18 and 19 present the way the effects of the disturbances $F$ and $x_F$ are rejected by the control system.

Analyzing figures 16-19 it can be observed that when the disturbances change, the internal model controller associated with any of the two concentrations act in such way that the deviations produced by these changes of the disturbances are eliminated. Thus, it can be considered that the two systems proposed for separated products quality control had been validated.

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Fig. 14. Time evolution of $x_B$ for $T_M$ tuning

Fig. 15. Control signal variation for $T_M$ tuning

Fig. 16. Time evolution of $x_D$ for a 3% change in $F$

Fig. 17. Time evolution of $x_D$ for a 3% change in $x_F$

Fig. 18. Time evolution of $x_B$ for a -3% change in $F$

Fig. 19. Time evolution of $x_B$ for a -3% change in $x_F$

Fig. 20. Hierarchical control system structure
Industrial implementation
The proposed hierarchical control system has the structure presented in figure 20.

The hierarchical control system from figure 20 has two hierarchical levels: level 1 which is represented by the LB structure together with the conventional automation (the pressure, level and flowrate control systems) and level 2 which contains two internal model controllers for the quality control of the separated products. The controllers outputs represent setpoints for the reflux flowrate control system and respectively for the bottom product flowrate control system.

Conclusions
The steady-state simulation of the butylene-butane distillation process allowed the choosing of the reflux flowrate and bottom product flowrate as control signals based on the relative gain array values. For the dynamic simulation the multicomponent distillation process was adapted to a pseudo-binary process. The results of the dynamic simulations were used for the identification of the simplified dynamic models of the distillation process on the input-output channels reflux flowrate – concentration of the pseudo-light component and bottom product flowrate – concentration of the pseudo-heavy component. For the distillate and bottom product quality control was chosen an internal model control algorithm. The dynamic simulation of the developed control systems allowed the tuning of the two controllers associated with the two separated products. The control systems were tested at setpoints and disturbances step changes and the simulation results validated the proposed control structure and algorithm.

References
23. POPESCU, M., Contribuții la dezvoltarea unui sistem ierarhizat de conducere distribuită a proceselor de fracționare, PhD Thesis, Univ. Petrol-Gaze din Ploiești, 2012

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