Application for Testing Control Configurations of Binary Distillation Columns

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Abstract: The paper addresses the problem of testing various control configurations for binary distillation columns. Analyzing from plantwide control point of view the place of distillation column within the plant, the result will be the best control configuration. The proposed application *Test BCC* is a very useful tool to evaluate the resulted control configuration for distillation columns. First, it is presented the selection of best control configuration from plantwide control point of view. The resulted control configuration is applied to a real column or to a simulation module via DAQ modules. The two-point composition control problem of binary distillation columns is addressed. The control configuration uses a multivariable controller formed by a parametric decoupler and two monovariabile PI controllers. The example of propane - propylene splitter is used to demonstrate the application efficiency, identifying the best control configurations from potential structures. The limits of this approach for control configuration selection are also analyzed. A closed-loop dynamics study formally shows the potential of the proposed application, matching or improving the behavior obtained with previous schemes.

Key-words: plantwide control, composition control, dynamic simulations, distillation columns.

1 Introduction[1]

The state of the art dynamic simulators for distillation columns are time consuming and requires both process and control knowledge to properly configure ([3], [10], [23], [26]). A reduced order model proposed by Skogestad captures the main dynamic features of distillation columns ([18], [19], [20], [21]).

Even that with such a simulator one can study from simple control schemes to double ratio configurations, the majority of industrial distillation columns are operated in one-point control mode, with one loop on automatic and one manipulated variable adjusted manually, according to inventory-based supervisory control schemes ([4], [11], [15], [17], [24]). In the presence of feed changes in quality and quantity, the operation tends to yield overpurified products with excessive high energy consumption ([5], [11]). The possibility of energy savings, tighter quality control, and reduced utility consumption motivates the consideration of multiple-input, multiple-output (MIMO) control schemes ([2], [6], [13], [14], [25]). Since distillation columns exhibit complex nonlinear dynamics with strong and asymmetric input-output coupling, especially in the high-purity case, the design of a two point composition controller is regarded as a difficult task ([1], [16], [21]). The problem has been addressed with a diversity of techniques: linear decentralized ([7], [20]), multivariable (MIMO) ([12], [13], [14]), and model predictive control (MPC) ([15], [19]) and geometric ([1], [2], [8], [9]) schemes. However, the applicability of MIMO techniques is still questioned on the basis of complexity, reliability, maintenance, and cost ([12], [17]). In the design of a MIMO controller the choice of structure and algorithm is usually taken on unit based approach ([11], [17]). In this work, it is presented an application for binary distillation columns that tests the behavior of different control schemes. The proposed application is extended to meaningfully compare and choose among various structures (in manual mode, one point control or two point control). These comments motivate the need to develop simpler and more systematic control structure and algorithm two point composition control schemes for distillation columns and a unified framework to connect the diversity of related linear and nonlinear control techniques. In this work, the dual composition control problem of binary distillation columns is addressed. The starting point was the study of the dynamics of binary distillation columns by the means of state of the art rigorous dynamic simulators. Then, a reduced model is used to test in offline mode the distillation column behavior, yielding the limiting behavior attainable with any controller. A rigorous closed-loop state dynamics assessment formally shows the key recovery feature and provides tuning rules for the parametric decoupler (in two point control case), and for the two monovariabile controllers that resemble the ones employed in industrial practice to implement linear single-input, single-output (SISO) controllers and filters.

[1] This work was supported in part by the Romanian National Authority for Scientific Research under Grant 3145/13.10.2005.

2 Proposed application

In the last years it was emphasized a new research direction in chemical process control field the so called integrated design and control of chemical processes ([3], [6], [12]). The already designed plants can be improved choosing the best control configuration using plantwide control procedures. The goal of plantwide control studies is to design a control system that can achieve the operating requirements of the complete plant optimal conditions. Plantwide control does not meant the tuning and studying the behavior of each of these loops, it correspond to the control philosophy of the overall plant with emphasis on the structural decisions [19].

The distillation columns are one of the most important processes in the petroleum refineries. The objective of these units is to obtain products from splitting raw materials of low commercial value. The objectives of this paper are the presentation of proposed application *Test BCC* (**Test**ing the **B**est **C**ontrol **C**onfiguration) for **b**inary **d**istillation **c**olumns (BDC) and demonstrate its efficiency on propane-propylene splitter from the separation part of the FCCU (also called GPU – Gas Processing Unit).

The starting point is implementing at a minimum functional level the working framework to develop the application of generating the best structure. This means the design of these basic components: simulation module of binary distillation columns *Sim BDC*, the simulation module of decoupling control composition loops, and the composition control modules. The last two components are the main parts of *Test BCC* and will be described in the following.

At this stage of the project after a thoroughgoing study on mathematical formalism used for distillation columns the result is a distillation model suited for control purposes. This model is resulted after studying the dynamic behavior of distillation columns with different control configuration using Dynsim® and Aspen Hysys® environments.

This paper also deals with the design issues (detailing the requirements for the proposed application and the evaluation methods of the results, detailing the initial project specifying the new implementation requirements) and determining a simple way to describe the requirements of such application so that it can represent a useful guide for the process engineer designer.

Distillation plants are from economical point of view, one of high energy consumption units from a refinery. The developing of chemical industry is tightly linked with the emerging of advanced chemical process control systems, while the energy consumptions are considerably reduced. The complexity of the distillation process leads to an increase effort in finding the best control configuration.

The proposed application tests the already chosen best control configuration for binary distillation columns.

In plantwide control context, the problem of control structure design the choice is made in 5 steps [19]:

1. selecting controlled variables;
2. selecting manipulated variables;
3. selecting measured variables;
4. selecting control configuration;
5. selecting and tuning the controller type.

On the other side, it appears as viable alternative, the principle of *feedback optimizing control structures*, proposed by Stephanopoulos and Ng [22] as formal framework for identifying the controlled variables. Another major problem to be solved in this context is the translation of implicit operational objectives into setpoints for feedback control loops. The hierarchical approach is an efficient mechanism of treating the complexity of the control problem by:

(i) specifying the control objectives with different time scales;
(ii) modeling from abstract to detailed level;
(iii) selecting the control variables and manipulated variables;
(iv) facilitating the configuring of control structures.

Plantwide control field is in an early stage of its development. Although the distillation process is well known and deeply theoretically investigated, its complexity and the magnitude of material and energetic fluxes associated justify the need of ongoing efforts to build mathematical models as efficient as possible from control perspective. That is why, for the distillation process is needed the efficient solving of the modeling issues of this process from plantwide control point of view. A plantwide control solution, namely decentralized control structure represents an alternative to the use of centralize multivariable controller with very large dimension. Binary distillation columns are high energy consumers, any improvement in their operation being very important for a refinery.

The proposed method takes into account the place of the studied distillation column. Distillation columns usually take part into a large complex formed by other process unit. From this point of view we can characterize a distillation column as the first process unit, intermediate or final distillation column. An intermediate column influences downstream units if there are no recycle streams; otherwise it has effects on both downstream and upstream unit.

Final columns do not disturb any upstream or downstream unit. From plantwide control strategy, there is no restriction imposed to this final column, the best control structure for it being generated by steady state criteria (e.g. RGA) usually combined with dynamic simulations with rigorous simulation tools.

The first distillation column is usually forced to refine the main flux formed by a mixture from which the final products of the unit will be recovered. The throughput manipulator place also influences the control decisions to be taken. The control structure chosen must have good features regarding the effect of the main disturbances that affect a distillation column, namely feed flowrate F and its composition x_F (Fig.1).

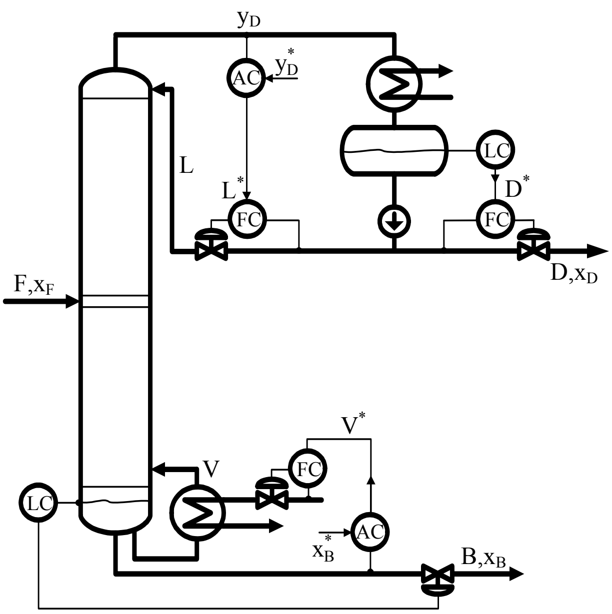

Fig.1. Manual LV configuration for a BDC.

Also, the proposed structure for the first distillation column must avoid a structure that uses the ends as manipulating variables, because the final products of the first columns represent feed streams for downstream units. Usually distillate is the most important product of the columns, so the main flux of the plant will be directed to the tops of the distillation columns.

The structure of an intermediate distillation column will depend on the existence of the recycle streams. After studying various process units from Romanian refineries, in the case of no recycle streams we recommend a control structure with minimum influence for downstream units.

In Fig.1 it is considered a BDC with one feed. From systemic perspective it has five manipulated variables $u = (L, V, D, B, Q_c)^T$ representing the reflux flowrate, the boilup flowrate, distillate flowrate, bottom product flowrate and cooling agent flowrate. The system outputs are also five $y = (x_D, x_B, M_D, M_B, p)^T$ represented by light component composition in distillate and bottom product, the holdups on reflux drum and in the bottom of the column, and the column pressure. In the case of a distillation column the first thing to be done is stabilizing the column by its inventory control loops (three decentralized monovariabile control loops: one for pressure and two for level) with the outputs:

$$y_2 = (M_D \;\; M_B \;\; p)^T . \qquad (1)$$

The non-controlled variables are the compositions:

$$y_1 = (x_D \;\; x_B)^T . \qquad (2)$$

The three monovariabile control loops associated to y_2 weakly interact between them and can be considered independents. There are more possibilities to choose the u_2 manipulated variables (therefore for u_1, too). These control configurations are named from u_1 inputs remained for composition control loops. The LV configuration from Fig.1. refers to control system with

$$\begin{cases} u_1 = (L \quad V)^T \\ u_2 = (D \quad B \quad Q_c)^T \end{cases} \qquad (3)$$

The most known control configurations for distillation columns are LV, DV, LB or double ratio configurations etc. The proposed application can test these configurations in no composition control, one point composition control or two point composition control.

For two point composition control, *Test BCC* provides the possibility of designing a robust decoupler, parametric type, its implementation needing two static parameters, two deadtimes and two dynamic time constants, that are drawn from plant data and/or simulation packages. These parameters have direct physical meaning and results from process features.

The decoupling of composition control loops will be applied to a partial control system, inventory control loops closed. The process to be controlled becomes two input - two output multivariable process.

The decouplers should never be used when the model has large RGA elements compared to one (LV configuration – Fig.1). The main advantage of this structure is that the tuning problem of the multivariable controller is reduced to tuning independent monovariable controllers associated to each control loop (Fig.2).

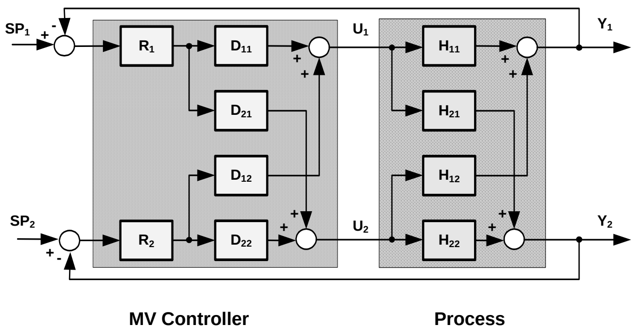

Fig.2. Multivariable control system with decoupling controller.

In order to design the decoupler, each input-output channel of the process must be characterized by three parameters, which are determined experimentally: the proportional gain $(K_p)_{ij}$, deadtime $(\tau_p)_{ij}$ and transient time $(T_t)_{ij}$:

$$\text{Process}: \begin{bmatrix} (K_{p11}, \tau_{p11}, T_{pt11}) & (K_{p12} \; \tau_{p12}, T_{pt12}) \\ (K_{p21}, \tau_{p21}, T_{pt21}) & (K_{p22}, \tau_{p22}, T_{pt22}) \end{bmatrix}.$$

The decoupler has the following structure:

$$D = \begin{bmatrix} \dfrac{e^{-\tau_{11}s}}{T_{11}s+1} & \dfrac{k_{12}e^{-\tau_{12}s}}{T_{12}s+1} \\ \dfrac{k_{21}e^{-\tau_{21}s}}{T_{21}s+1} & \dfrac{e^{-\tau_{22}s}}{T_{22}s+1} \end{bmatrix}, \qquad (4)$$

where

$$k_{12} = -K_{p12} / K_{p11},\ k_{21} = -K_{p21} / K_{p22}. \quad (5)$$

These features simplify the decoupler structure and lead to a decoupled process with faster dynamics. The decoupling problem is solved based on the idea of compensating the effects of two parallel opposite input-output channels, with almost the same proportional gains, deadtimes and transient times.
From this idea, the time constants of the decoupler will be computed, according to process dynamics, as follows:

$$T_{pt22} \le T_{pt21} \Rightarrow T_{11} = 0,\ T_{21} = (T_{pt21} - T_{pt22})/4;$$

$$T_{pt22} \ge T_{pt21} \Rightarrow T_{21} = 0,\ T_{11} = (T_{pt22} - T_{pt21})/4;$$

$$T_{pt11} \le T_{pt12} \Rightarrow T_{22} = 0,\ T_{12} = (T_{pt12} - T_{pt11})/4;$$

$$T_{pt11} \ge T_{pt12} \Rightarrow T_{12} = 0,\ T_{22} = (T_{pt11} - T_{pt12})/4.$$

From these implications result mainly four variants of decoupler:

A. direct input-output channels are faster than the crossed ones, the case of A-type decoupler:

$$D_A = \begin{bmatrix} 1 & \dfrac{k_{12}}{T_{12}s+1} \\ \dfrac{k_{21}}{T_{21}s+1} & 1 \end{bmatrix}; \qquad (6)$$

B. direct input-output channels are slower than the crossed ones, the case of B-type decoupler:

$$D_B = \begin{bmatrix} \dfrac{1}{T_{11}s+1} & k_{12} \\ k_{21} & \dfrac{1}{T_{22}s+1} \end{bmatrix}; \qquad (7)$$

C. the direct input-output channel 1-1 is faster than the crossed channel 2-1, while the direct input-output channel 2-2 is slower than the crossed channel 1-2, the case of C-type decoupler:

$$D_C = \begin{bmatrix} \dfrac{1}{T_{11}s+1} & \dfrac{k_{12}}{T_{12}s+1} \\ k_{21} & 1 \end{bmatrix}; \qquad (8)$$

D. the direct input-output channel 1-1 is slower than the crossed channel 2-1, while the direct input-output channel 2-2 is faster than the crossed channel 1-2, the case of D-type decoupler:

$$D_D = \begin{bmatrix} 1 & k_{12} \\ \dfrac{k_{21}}{T_{21}s+1} & \dfrac{1}{T_{22}s+1} \end{bmatrix}. \qquad (9)$$

Compensating the effects of two parallel opposite input-output channels, the deadtimes are computed with:

$$\tau_{p22} \le \tau_{p21} \ \Rightarrow\ \tau_{11} = 0,\ \tau_{21} = \tau_{p21} - \tau_{p22};$$

$$\tau_{p22} \ge \tau_{p21} \ \Rightarrow\ \tau_{21} = 0,\ \tau_{11} = \tau_{p22} - \tau_{p21};$$

$$\tau_{p11} \le \tau_{p12} \ \Rightarrow\ \tau_{22} = 0,\ \tau_{12} = \tau_{p12} - \tau_{p11};$$

$$\tau_{p11} \ge \tau_{p12} \ \Rightarrow\ \tau_{12} = 0,\ \tau_{22} = \tau_{p11} - \tau_{p12}.$$

From these implications result four variants of decoupler:

$$\text{a. } \begin{cases} \tau_{p11} < \tau_{p12} \\ \tau_{p22} < \tau_{p21} \end{cases} \Rightarrow \begin{cases} \tau_{12} = \tau_{p12} - \tau_{p11} \\ \tau_{11} = \tau_{22} = 0 \\ \tau_{21} = \tau_{p21} - \tau_{p22} \end{cases};$$

$$\text{b. } \begin{cases} \tau_{p11} > \tau_{p12} \\ \tau_{p22} > \tau_{p21} \end{cases} \Rightarrow \begin{cases} \tau_{11} = \tau_{p22} - \tau_{p21} \\ \tau_{12} = \tau_{21} = 0 \\ \tau_{22} = \tau_{p11} - \tau_{p12} \end{cases};$$

$$\text{c. } \begin{cases} \tau_{p11} < \tau_{p12} \\ \tau_{p22} > \tau_{p21} \end{cases} \Rightarrow \begin{cases} \tau_{11} = \tau_{p22} - \tau_{p21} \\ \tau_{21} = \tau_{22} = 0 \\ \tau_{12} = \tau_{p12} - \tau_{p11} \end{cases};$$

$$\text{d. } \begin{cases} \tau_{p11} > \tau_{p12} \\ \tau_{p22} < \tau_{p21} \end{cases} \Rightarrow \begin{cases} \tau_{21} = \tau_{p21} - \tau_{p22} \\ \tau_{11} = \tau_{12} = 0 \\ \tau_{22} = \tau_{p11} - \tau_{p12} \end{cases}.$$

It can be stated that the proposed procedure is an experimental decoupling procedure for two input - two output processes, based on a standard structure decoupler that can be implemented in 16 distinct variants, according to the dynamics of direct and crossed channels process features.
The decoupler has 6 tuning parameters: two proportional gains, two deadtimes and two transient times. The initial values of the decoupler are determined by 3×4 process parameters, each of the 4 input-output process channels being characterized by three parameters: proportional gain, deadtime and transient time. The simulation results demonstrated the practical character of the proposed decoupling procedure, the decoupling procedure being tested on many multivariable processes [14].
Next, the hardware implementation requires the use of data acquisition modules. In this case *Test BCC* works with two KUSB 3100 serial interface from Keithley Instruments. One interface is used to acquire/generate the data from/to the real BDC (on line mode) and the other to do the same with the simulated column (off line mode).
The off line mode is very useful for testing the behavior of the BDC, because the results are obtained much faster than from real BDC. The resulted framework for Test BCC is presented in Fig.3. The PI monovariable controllers modules contains an user interface that allow tuning the K_P and T_i parameters. First the 6 decoupler parameters are tuned, after that the task of tuning the multivariable controller is much simpler, the 2 monovariable controllers being tuned as for two separate SISO control loops.
The proposed control technique is tested with a representative BDC, namely propane-propylene splitter.

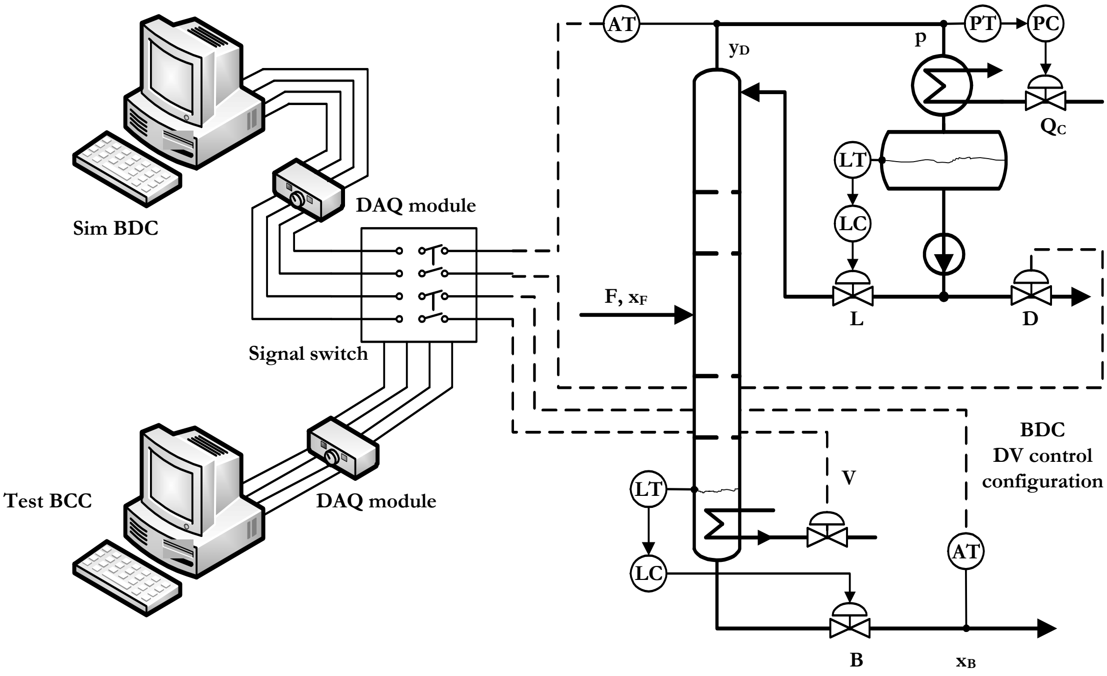

Fig.3. *Test BCC* framework.

3 Case Study

The propane-propylene splitter is a final distillation column from separation unit that process the most important products of FCCU. Consequently, the control structure is not imposed by plantwide criteria. From this point *Test BCC* offers the opportunity of studying the behavior of the column with several control configurations in offline mode (no composition control, one point control, two point control) and in online mode to the real distillation column with LB structure two point control. In Fig.4 it is presented the behavior of propane-propylene splitter in offline mode, one point control (top composition is controlled) with 8 potential control configurations, to a step decrease in feed flowrate (main disturbance).

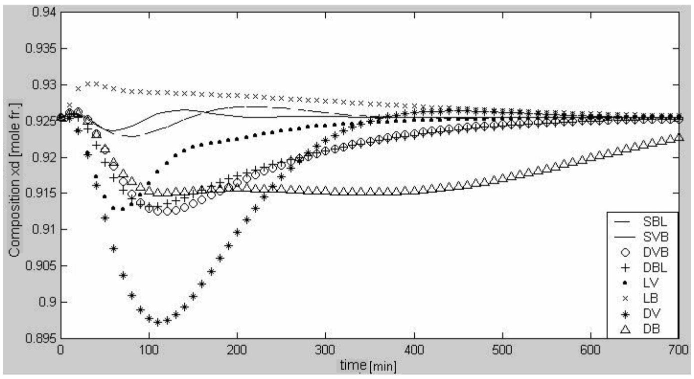

Fig.4. Top composition response to a step decrease in feed flowrate.

The PPDC that was studied is proven to be sensitive to any disturbance, especially to changes of the feed quality (x_F). The specifications for this column were stiff concerning the composition for the top product (90% vol. propylene) while the specification for bottom product is flexible.

The results concerning the RGA [6] proved that the best structure for the studied column is represented by ratio control schemes naming SV/B and DV/B configurations. The specifications for this column and the current x_B ranges (0.01…0.05% mole fr.) lead to RGA closest to 1. The most important parameter taken into account is the dynamic response of the structure. The SV/B structure has a faster dynamic response than DV/B structure, as a result of dynamic simulations. More important for SV/B structure is the feature that reject the effect of feed flowrate changes even in open loop mode. The SV/B structure associated to this column does not influence any downstream units.

Another important feature of *Test BCC* is the possibility to use a parametric decoupler in two point control mode (for the real or simulated distillation column).

For the studied column with SV/B structure the resulted decoupler is C-a type [14]

$$D_{C-a} = \begin{bmatrix} \dfrac{1}{T_{11}s+1} & \dfrac{k_{12}e^{-\tau_{12}s}}{T_{12}s+1} \\ k_{21}e^{-\tau_{21}s} & 1 \end{bmatrix}, \qquad (10)$$

with static parameters $k_{12} = 0.84$, $k_{21} = 0.24$, time constants $T_{11} = T_{12} = 50$ min and deadtimes $\tau_{12} = 1$ min and $\tau_{21} = 2$ min. The decoupling features of the process are improved about 20 times [14].

The limit of this approach is the need of two composition analyzers that usually are expensive and used only on very important products (the case of propane propylene splitter). Consequently the future work will consist in finding an adequate inferential composition measurement from available tray temperature transducers and extending the use area of *Test BCC* to BDC without online composition measurement.

4 Conclusions

The two point composition control problem for binary distillation columns has been addressed. The proposed application works with two modes (on line and off line) each of them with the possibility to choose no composition control, one point composition control, two point composition control. linear output-feedback two-way decoupling control scheme. For two point composition control, the combination of feedforward, feedback ideas led to a parametric decoupler with standard structure. The decoupler has 6 tuning parameters dependent on process features. The associated multivariable controller consisted of two PI feedback components and a parametric decoupler. The proposed application was tested with a representative BDC (propane-propylene splitter) matching or improving the behavior obtained with previous schemes.

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