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Simulation and Control Configuration of Integrated Three-Product (Petlyuk) Distillation Column

R. Solar-Gonzalez[#], R. Monroy-Loperena [*], A. Velasco-Perez[#]

[#]*División Estudios de Postgrado, Universidad del Istmo*
Ciudad Universitaria, Ba. Sta. Cruz Tagolaba, Sto Domingo Tehuantepec, Oaxaca, 70760, México
[*]*ROMON S.A.*
Paseo de los Pirules 124, México, 04250, México
[#]*Departamento de Ingeniería Química, Universidad Veracruzana*
Prolongación de oriente 6 No. 1009 Col. Rafael Alvarado, Orizaba, Veracruz, 94340, México

***Abstract*— In a Petlyuk distillation column, two extra degrees of freedom can be used for optimization purposes. It has been reported that a typical energy saving of 30% is achievable with a Petlyuk distillation column, compared to conventional distillation arrangements. However, the optimal steady-state operation point can be difficult to maintain in practice. In this work we propose and studied the performance of a parallel control configuration, using habituating parameters, for the Petlyuk distillation column in presence of disturbances. The results show that parallel control configuration can be used to improve the robustness of operation by extending traditional decentralized product composition control loops for two and three points by using the two degrees of freedom as manipulated variable in a parallel feedback control loop.**

***Keywords*— Continuous distillation, Thermal coupled distillation, Petlyuk, composition control, Parallel control.**

I. Introduction

Distillation is a unit operation widely used to separate, multicomponent mixtures in spite of its high energy consumption and low thermodynamic. As a result, engineers and researchers are interested in developing new configurations capable of reducing and improving the use of energy. It has been demonstrated that thermal linking can reduce energy demands between 30 and 50% depending on the composition of the mixture to be separated. The three thermally coupled distillation sequences that have been explored in detail are the columns that using a side rectifier, the columns with a side stripper and the fully thermally coupled distillation sequence or Petlyuk system[1]. Because of the reduction in the oil reserves of the reduction in the oil reserves and policies of reduction in carbon dioxide emissions [2], significant research has focused on the design, optimization and control of the thermally coupled distillation sequences]. The Petlyuk column has also been described as the most efficient arrangement for three-component mixtures. It has been written [3] that, "For most separations, the fully thermally coupled distillation column is thermodynamically more efficient than the conventional arrangements, and as a consequence, has lower energy requirements." The lower energy requirements of the Petluyk column configuration lead to smaller column diameters and lower overall heat exchanger area in the reboiler and condenser. The reduced number of reboilers and condensers also results in a lower capital investment. However, in spite of all these apparently attractive features, the PC configuration has not found a wide industrial use. Even for sub ambient temperature distillations, where energy consumption is of great importance, the Petlyuk column configuration is yet to be used. This is rather surprising because the PC configuration has been known for nearly 50 years[4-5].

The lack of widespread use of the PC configuration has been attributed to its difficult design and control. However, considerable advancement has been made recently in both the design and control aspects of the PC configuration. The expectance that the dynamic properties of the PC configuration may cause more operational problems than conventional sequences is one of the factors that have contributed to their lack of industrial implementation. This conflict of an energy-efficient system leads to tight designs, which in turn are more difficult to control.

A two-column implementation of the PC design consists of a prefractionator with reflux and boilup from the downstream three-product column, a setup with only one reboiler and one condenser. As proposed by Wright [6], practical implementation of such a column can be accomplished in a single shell by inserting a vertical wall through the middle section of the column, thus separating the feed and side product draw. Petlyuk's main reason for this design was to avoid thermodynamic losses from mixing different streams at the feed tray location.

Respect to the degrees of freedom (see for instance, Dixon [7]), we have that in a conventional given column with fixed stages, feed locations, etc. Starting with binary distillation and considering a steady-state, where it is assumed that the holdups (condenser level, reboiler level, and pressure) are already controlled, we find two independent (manipulated) variables remain, for example reflux rate (L) and vapor boilup rate (V). On the other hand, in a PC we get at steady-state three additional degrees of freedom [8-9], one for each of the three additional streams leaving the main column. Referring to Figure 1, these are the side-stream S plus the streams SL and SV sent back to the prefractionator. Note that in this analysis the prefractionator itself does not have any degrees of freedom at the steady-state. In a traditional point of view of distillation control, the five degrees of freedom for the PC design may be used to specify (control) the top and bottom composition and one or two compositions in the side-stream. This leaves one or two degrees of freedom for optimization purposes.

We want to report in this work a two and three points composition control for the PC configuration, following a habituating approach [10-12], due to the PC configuration can be seen as a plant with more manipulated inputs than controlled outputs, then we resort in the premise that the control objectives can be satisfied more easily by utilizing additional input variables. In this way, we propose a simple procedure to expand traditional decentralized composition control structures applied to a PC configuration. Rigorous nonlinear numerical simulations are used to illustrate the performance of the proposed control scheme under typical operating conditions in a PC configuration.

II. Statement of the Control Problem by a Habituating Approach

A Let us assume a plant description of the form

$$\mathbf{y}(s) = \mathbf{G}(s)\,\mathbf{u}(s)$$

where **G** denote the process plant model, **y** the measurements and **u** the manipulated inputs. If the plant is an input/output linear process with more manipulated inputs than controlled output, the underlying premise is that the control objectives can be satisfied more easily by utilizing additional input variables. For the case of 4 input/ 2 outputs we have that

$$\mathbf{G}(s)=\begin{bmatrix} \frac{K_{1,1}}{\tau_{1,1}s+1} & \frac{K_{1,2}}{\tau_{1,2}s+1} & \frac{K_{1,3}}{\tau_{1,3}s+1} & \frac{K_{1,4}}{\tau_{1,4}s+1} \\ \frac{K_{2,1}}{\tau_{2,1}s+1} & \frac{K_{2,2}}{\tau_{2,2}s+1} & \frac{K_{2,3}}{\tau_{2,3}s+1} & \frac{K_{2,4}}{\tau_{2,4}s+1} \end{bmatrix} e^{-\theta s}$$

$$\mathbf{y}(s) = [y_1(s), y_2(s)]$$

$$\mathbf{u}(s) = [u_1(s), u_2(s), u_3(s), u_4(s)]$$

where $K_{i,j}$ is the steady-state gain, $\tau_{i,j}$ is the open-loop time constant, both for output i with input j and θ is the time delay of the proposed process.

We propose to solve the control problem by pairing the variables proceeding along with the following map

$$(u_1, u_2) \rightarrow y_1$$
$$(u_3, u_4) \rightarrow y_2$$

then we arrive to a diagonal block structure, that is

$$\mathbf{G}(s) = \begin{bmatrix} \begin{bmatrix} \frac{K_{1,1}}{\tau_{1,1}s+1} & \frac{K_{1,2}}{\tau_{1,2}s+1} \end{bmatrix} & \\ & \begin{bmatrix} \frac{K_{2,3}}{\tau_{2,3}s+1} & \frac{K_{2,4}}{\tau_{2,4}s+1} \end{bmatrix} \end{bmatrix} e^{-\theta s}$$

The above structure can be distinguished as a Diagonal Parallel Control Structure.

In this way the original problem was split into two 2-inputs/1-output problems, whose solution can be arrived as follows. Let us take the first problem,

$$y_1 = \left[\frac{K_{1,1}}{\tau_{1,1}s+1} u_1 + \frac{K_{1,2}}{\tau_{1,2}s+1} u_2 \right] e^{-\theta s}$$

In this non-square case, one has to specify a strategy for the additional control input u_2.

One alternative is to use a sort of regularization technique to square the rectangular control system. We propose to split the input/output model by introducing an habituating parameter, , as follows:

$$y_{1,1} = \beta_1 \frac{K_{1,1}}{\tau_{1,1}s+1} u_1$$
$$y_{1,2} = (1-\beta_1) \frac{K_{1,2}}{\tau_{1,2}s+1} u_2$$

where

$$y_{1,1} = \beta_1 y_1$$
$$y_{1,2} = (1-\beta_1) y_1$$

so that $y_1 = y_{1,1} + y_{1,2}$. Notice that the control output signal has been partitioned into two components: $y_{1,1}$ and $y_{1,2}$, corresponding to fractions of the regulated output affected respectively by the control inputs u_1 and u_2. Physically, this splitting of the controlled output can be seen as a distribution of the control effort between the two control inputs. Proposing the use of PI controls, , then the inputs can be manipulated as

$$u_1(s) = \beta_1 C_{PI,1}(s)$$
$$u_2(s) = (1-\beta_1) C_{PI,2}(s)$$

where

$$C_{PI,j} = K_{C_j} + K_{I_j} s^{-1}$$

for which well-known control techniques (e.g., Internal Model Control (IMC)) can be used for control design, that is

$$K_{C_j} = \frac{1}{K_{i,j}} \frac{\tau_{i,j}}{\tau_{C_j} + \theta}$$
$$\tau_{I_j} = \tau_{i,j}$$
$$\tau_{C_j} \cong 0.75\tau_{i,j}$$

By introducing the variables

$$z_1 = y_{1,1}$$
$$z_2 = y_{1,2}$$
$$z_3 = y_{2,1}$$
$$z_4 = y_{2,2}$$

a square 4x4 PI diagonal controller is obtained, that is:

$$\begin{bmatrix} z_1 \\ z_2 \\ z_3 \\ z_4 \end{bmatrix} = \begin{bmatrix} \beta_1 \frac{K_{C_1}s+K_{I_1}}{s} & & & \\ & (1-\beta_1)\frac{K_{C_2}s+K_{I_2}}{s} & & \\ & & \beta_2 \frac{K_{C_3}s+K_{I_3}}{s} & \\ & & & (1-\beta_2)\frac{K_{C_4}s+K_{I_4}}{s} \end{bmatrix}$$

Note that if β_i is unity then we recover a traditional 2x2 diagonal PI configuration.

III. Numerical Simulations

The case study is similar to case IV of Weyburn and Seader [13]. The feed is liquid at its bubble point with a flow rate of 1000 mol/min: 200 mol/min of benzene, 400 mol/min of toluene, and 400 mol/min of o-xylene. At the bubble point, the temperature and pressure of the mixture are 383.4 K and 101.33 kPa, respectively. The following general assumptions are made:

(1) The reflux divider and all stages, except the condenser and reboilers, are considered to be adiabatic.

(2) Heat-transfer rates for condensers and reboilers are never specified.

(3) The system operates at atmospheric pressure (101.33 kPa) with zero pressure drop throughout the system.

(4) The configuration of the system, that is, number of stages and locations of feeds, side streams, and interlinks, is known.

(5) All product streams leave the system as saturated liquids.

(6) The model description considers only the material balance of the system, assuming constant molar flows and ideal behavior, as the mixture studied (benzene, toluene, and o-xylene) is nearly ideal. Accordingly, the components have the following relative volatilities: [6.01, 2.61, 1.0]

The system consists of a 20-stage prefactionator interlinked with a 32-stage fractionator. The latter includes a total condenser and a partial reboiler. The stage locations for the feed, products, and interlinks are all included in Figure 1. All model equations were simultaneously solved by means of an implicit Euler approximation with time-step equal to 0.1min. At each time-step, the resulting nonlinear algebraically equations were solved by means of a Newton-Raphson method endowed with a sparse matrix solver to exploit the sparsity of the incidence Jacobian matrix. Because the trajectories of the PC are continuous in time, the Newton-Raphson method was implemented as a continuation method where the state of the preceding time step was taken as initial guess for the actual step. In this way, the Newton-Raphson required 2 or 3 iteration to achieve a relative error lower than 2%. The numerical procedure was implemented as a FORTRAN program in a Pentium IV computer.

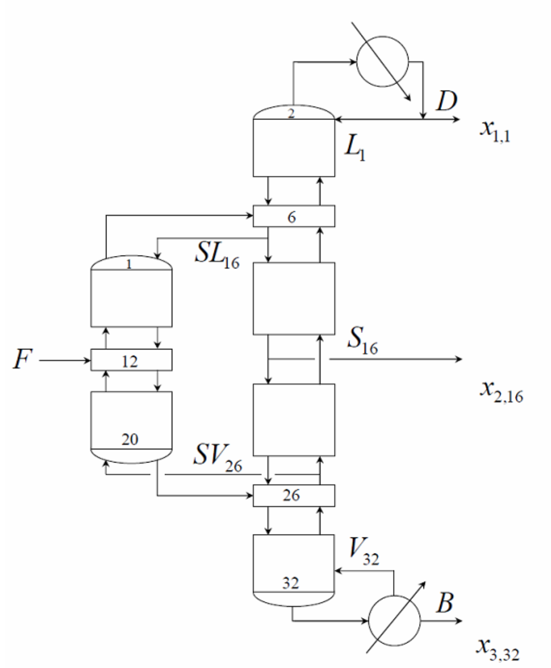

Fig. 1 Schematic diagram of the Petlyuk distillation column studied.

The nominal operating point studied has L_1 = 1050.0 mol/min, V_{32} = 2060 mol/min, SL_6 = 500 mol/min, SV_{26} = 1500 mol/min, and SL_{16} = 410 mol/min, with $x_{1,1}$ = 0.9701, $x_{2,16}$ = 0.9486 and $x_{3,32}$ = 0.9358.

For a PC configuration one must at least adjust two product flow rates correctly (e.g., L_1 to match the distillate composition and V_{32} to match the bottom composition), thus at least two-point control is required. However, three product flow rates are preferred to be adjusted (e.g., L_1 to match the distillate composition, L_{16} to match the intermediate product composition and V_{32} to match the bottom composition), to avoid that the composition of the intermediate product be lowered considerably under external disturbances. Hereunder these two cases will be analysed. The PI parameters for the necessary cases were obtained from step responses with IMC rulers. Under this set of control parameters, the following numerical simulations were carried out.

In this case a LV dual compositioncontrol configuration is used (where the reflux flow rate, L_1, is used to control the top composition, $x_{1,1}$, and the vapour boilup flow rate, V_{32} is used control the bottom composition, $x_{3,32}$). However, it should be stressed that other control configurations can be studied along the same lines. The column was started from the nominal conditions and then regulated at the setpoint ($x_{1,1}, x_{3,32}$) = (0.95, 0.99). At $t = 300$ min, the setpoint was changed to ($x_{1,1}, x_{3,32}$) = (0.99, 0.99), and at $t = 1800$ min, the feed composition was changed from (0.2, 0.4, 0.4) to (0.25, 0.35, 0.4).

Para use the diagonal PI model, it is necessary to mate the original manipulated variables with those identified as control degrees of freedom that are not used for control purposes. In our case we want to pair off the reflux rate, L_1, with the liquid side stream flow rate, SL_6, or the vapour side-stream flow rate, SV_{26}, and either to pair off the vapour boil-up flow rate, V_{32}, with any of these two variables. Referring to open-loop steady-state gains, we can notice that if we select to pair off the reflux flow rate, L_1 , with the vapour side-stream flow rate, SV_{26}, and the vapour boilup flow rate, V_{32}, with the liquid side-stream flow rate, SL_6, then we have the pairing that most affect over the output variables (distillate and bottom compositions). Then, the following assignment to the input variables will be used: $u_1 = L_1, u_2 = SV_{26}, u_3 = V_{32}, u_4 = SL_6$, with the following output variables assignment: $y_1 = x_{1,1}, y_2 = x_{3,32}$. Figure 2, presents the behaviour of the PI controller for four values of the habituating parameters, β_i . Notice that in principle as we decrease that the habituating parameter a better performance is obtained, this performance is better that the diagonal LV-PI configuration (that is, when $\beta_i = 1.0$).

Also notice that for the values of 0.95 and 0.90 of the habituating parameters the internal flow rate are decreased, this implies that lower demands of energy are necessary. In turn, this habituating control operation of the PC will resound in economic saving because less refrigerant and heat agents will be used when the habituating are well selected.

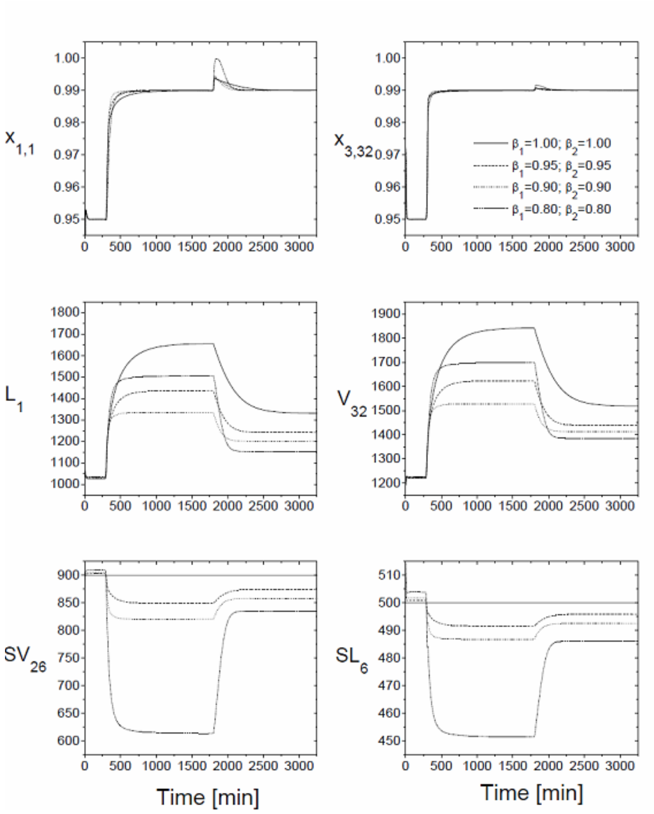

Fig. 2 Dynamic behaviour of the Petlyuk distillation column for various values of the habituating parameters in two point composition control, when the input variables are selected in accordance with its effect over the output variables.

To analyse the effect of each pairing input variable, we turn off one of the two pairing variables by setting its habituating parameter to unity. As a first case consider the case where $\beta_1 = 1.0$, then the top composition, $x_{1,1}$, will only be regulated by means of manipulating the reflux rate, L_1. In the Figure 3, we note a better performance when β_2 is decreased. However the energy requirements are larger. On the other hand, consider the case when $\beta_2 = 1.0$, then the bottom composition, $x_{3,32}$, will only be regulated by means of manipulating the vapour boil-up flow rate, V_{32}.

In the Figure 4 is presented the case when the setpoint change, note as β_1 is decreased a better performance is obtained until it reach a critical value where this behaviour is opposed. To avoid excessive control actions we satured the input signal, SV_{26}, to 600 mol/min. The same observation can be made for β_1, when this parameter is decreased the process has less energy requirements until it reach a critical value, too.

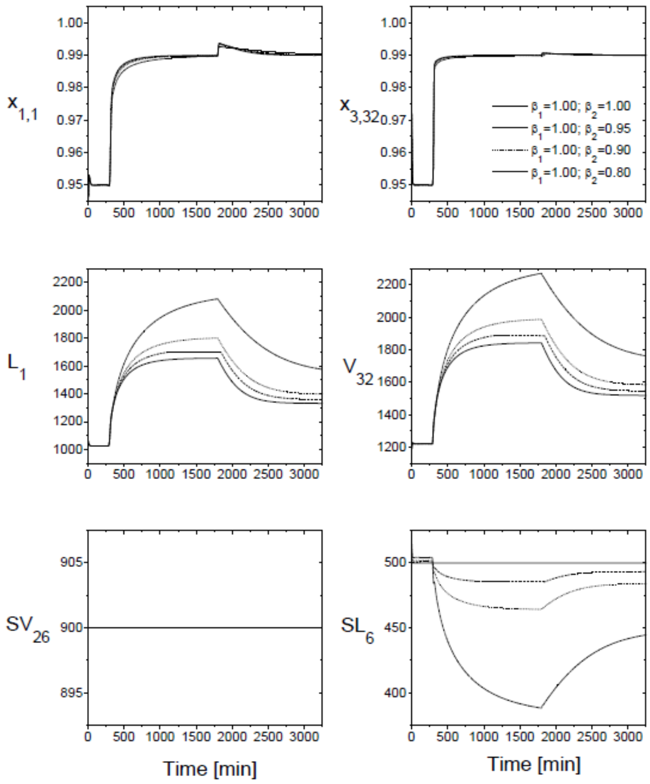

Fig. 3 Dynamic behavior of the Petlyuk distillation column when β_2 is fixed to one and various values of β_1, in a two point composition control, when the input variables are selected in accordance with its effect over the output variables.

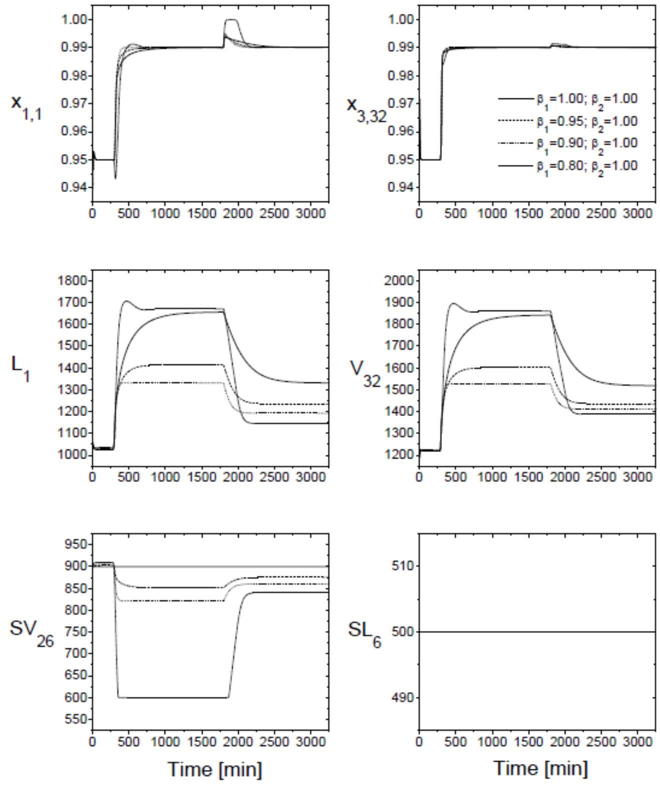

Fig. 4 Dynamic behaviour of the Petlyuk distillation column for various values of the habituating parameters in two point composition control, when the input variables are selected in accordance with its effect over the output variables.

Regarding then merits of the numerical experimentation presented above, the following comments must be done: (i) when a candidate of input variable that can be habituated is found, it must be selected by preference in which have the greater effect over the output variable. (ii) The input variable can be habituated must not provoke a signal conflict. For example, considering the two point case the worst behaviour was obtained when SL_6, was paired off with L_1, to control the distillate composition, because both are near and began a competition. (iii) The value of the habituating parameters has a direct impact on the performance of the controller. This observation suggests the existence of an optimal value of the habituating parameters. (iv) When the pair off the variables and the values of the habituating parameters are appropriate selected this resounds in economic saving, because small internal flows are obtained.

IV. Conclusions

A habituating control approach was used to balance the separation tasks provided by the distillation column and the prefractionator of a Petlyuk distillation configuration. It was observed that an appropriate selection of variables to be habituated resounds in economic saving, because small internal flows are obtained and then less refrigerant and heat agents will be used.

The main idea was to include the liquid side stream, SL_6, and the vapor side-stream SV_{26}, that usually are not employed for control purposes, in traditional two and three points PI diagonal structures. As a result, with an appropriate selection of the pairing variables, under expected operating conditions (that is, change in the charge composition to the Petlyuk column and change in the set point) the rejection times can be reduced and even more less energy requirements, can be found, giving as a consequence a better performance with respect to traditional diagonal PI composition control configurations.

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