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USE OF A TOOL FOR THE SELECTION OF CONTROL STRUCTURES IN PETROCHEMICAL PROCESSES

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Abstract. *For an oil processing plant to become competitive and still meets the specifications of the products in the current market, structural rearrangements have been applied, resulting in more integrated plants with high recovery mass and energy. However, as a result of this integration during production disturbances propagate more easily by the process and there is therefore a need for a more robust control structure. These cases are ideal for applying plantwide control in which the control loop process in applied observing the behavior of the plant as a whole. However, depending on the size of the evaluated process, the use of tools can support and facilitate the resolution of steps required to implement the plantwide method. This paper discusses the application of a tool developed in VBA-EXCEL® which aims to achieve the best sets of controlled variables that can be used in control structures. The tool was applied to a deethanizer column simulation with the aid of PRO/II® software. The results showed that the tool has potential application in real engineering problems, allowing obtaining control loops for greater controllability of the process.*

Keywords: *Control Plantwide, Selection Tool, Control Structures, Deethanizer Column, PRO/II®.*

1. INTRODUCTION

Aiming a processes that meet safety, production rate and product quality and the great competitiveness between industries, it is common nowadays that the engineer use of auxiliary procedures, which when applied to processing plants control oil, provide satisfactory results. Considering that the process under supervision should remain in full operation, to get a robust control loop practice is essential in order to stabilize the plant efficiently when there are disturbances, or at least keep it under an operating condition in which the "losses" during production may be acceptable.

In this context, a technology that has stood out is the plantwide control, according Lyman and Georgakis (1995) aims to provide a global framework for coordinated control of many important variables of a multiunit process, including all the plant production and quality control of the product. In a chemical plant there are thousands of variables, among them such controlled variables, manipulated and measured, this causes a range of control loops possibilities existing for the control process. The choice of best variables, in the other words, variables that affect significantly the overall process, allowing controllability of the plant and hence product quality is performed in accordance with the control structure design, e.g Skogestad, Perkins and Luyben.

Until the 70s, some areas of process control theory presented important advances such as real-time optimization of multivariable predictive control and controllability analysis tools, however, only from 1980 began to be published work with theoretical developments on Global Process Control, developed by the group of Professor Stephanopoulos at MIT (Larsson and Skogestad, 2000; Andrade, 2008).

Systematic procedures for obtaining plantwide control structures are found in the literature, as described by Skogestad (2000) and Skogestad (2004), and despite the basic idea of applying global control structures seem simple to be developed, the existing procedures bring to light the difficulty and complexity of this application depending on how the control problem is observed. So that this methodology can be used more easily, not only in academia, but as an engineering tool, considering the resolution of the used mathematical model and variable selection manipulated and controlled to the process, the use of combined software becomes such a way facilitator.

Silva (2013) shows that it is possible to perform the steps of Skogestad method in which computational efforts are required, using a robust model (stationary) that represents the process with the aid of a tool developed in VBA-EXCEL®. With the aim to obtain the best set of controlled variables that can be used in control structures, the tool was applied in a simulation deethanizer column, built in the PRO / II® software, for an oil processing plant.

2. METHODOLOGY

The tool used to control deethanizer column is based on the application of the methodology presented by Skogestad(2004), which suggests a systematic procedure for the global control structures design of the plant as follows.

Step 1 - Definition of operational objectives, constraints and cost scalar function J

The operational objectives set are preferentially grouped in a scalar cost function J to be minimized taking into account the available degrees of freedom, i.e., manipulated variables (inputs).

The control optimization problem can be represented by Eq. (1) subject to some restrictions of equality or inequality process, as shown in Eq. (2) where d represents all involved disturbances, including external, as well as changes in the model in the specifications (restrictions) and the parameters (rates), the process restriction may be related to product specifications and limitations in operating conditions, for example.

$$\min_{u_0} J_0(x,u_0,d) \tag{1}$$

$$\begin{aligned} g_1(x,u_0,d) &= 0 \\ g_2(x,u_0,d) &\le 0 \end{aligned} \tag{2}$$

Finally, the problem is evaluated according to a function loss in the process, represented by Eq. (3). It is required a bit of difference between the values of the cost function when obtained by applying a control strategy, in this case by a set of controlled variables $c = c_s$ by manipulating u0, and the same function in your great value c_{opt}.

$$L(u,d) = J(u,d) - J_{opt}(d) \tag{3}$$

Step 2 - Identification of degrees of freedom and manipulated variables

This step consists in determining the degrees of freedom for the process, i.e., manipulated variables, u_0, which in turn define the controlled variables to be selected for the process, c.

Step 3 - Selection of the primary control variables

The selection of primary variables control is directly linked to a viable operation, in terms of keeping the variable in its setpoint, cs, to obtain an almost "optimal" operating, where for each set of controlled variables (CVs) selected there is an assessment of the loss equation L.

Regarding the selection of good controlled variables, c, four requirements must be meet as presented in Araújo's work.

i. The optimal value of c should be less sensitive to disturbances;
ii. Measurement and c control should be easy;
iii. The value of c must be sensitive to changes in the degrees of freedom in the steady state, i.e. the manipulated variables, u.
iv. For cases with more than one degree of unfettered freedom, the selected control variable should be independent.

Step 4 - Production Capacity

The production rate is sometimes given as fixed in the plant's entrance, while its outputs work for level control, however depending on what is required in the process, this rate can become maximized. Thus, given a degree of freedom for feed rate, as it maximizes some point the same process can be altered, because the idea of maximizing the production rate, this example brings up the problem called "bottleneck "and entail consumption degree of freedom in the system, as previously defined.

Step 5 - Regulatory and Supervisory Control Layer

This first layer refers to the "stabilization" of the plant, it generally contains control loops that need is running to allow the layer supervisory work in search for a satisfactory system operation. An important structural issue then is the choice of the controlled variables (CVs) secondary y_2 (extras assisting the primary controlled variables in control), so that the effects of the primary CVs disturbances are rejected.

The main function of the supervisory layer may be set as the primary control variable y_1 in the respective predetermined values as degrees of freedom using the optimal values y_2 and any other manipulated variables. While the regulatory layer is directly linked to operational objectives, the supervisory control layer is related to the economic objectives of the plant, namely the production rate and product quality.

Step 6 - Optimization Layer and Validation

The identification of the active constraints and the recalculation of optimal values for the controlled variables, y_1 are the optimization layer functions. Another controversial issue is also the real-time optimization (RTO) or manual for the process. In relation to validation of the control structure, it is possible by means of a dynamic model of the nonlinear plant.

2.1 Description of Deethanizer Column

A distillation column with a multicomponent vapor partial condenser, and the separation of the ethane component (C2) from propane (C3) as the top product and base, was studied by applying the plantwide control tool. The tab. 1 shows the values of the molar fraction of supply for each component present in the unit operation, for which it has a flow rate of 150 ton.hr-1 fed on the plate 11 (the top as a reference) at a temperature in the feed stream 65 °C pressure at the top equal to 19.2 kg.cm-2 and the number of stages equal to 37 (with condenser and reboiler).

The deethanizer column this is an example studied by Hori and Skogestad (2007) in search of a good control structure that came to keep the top and bottom compositions close to their desired values. In this study, plantwide control structures are obtained by applying the tool developed by Silva (2013).

Table 1. Configuration data for the nominal feed composition of deethanizer column

Ethane	0.021
Propane	0.598
Iso-Butane	0.125
N-Butane	0.006
Propene	0.25

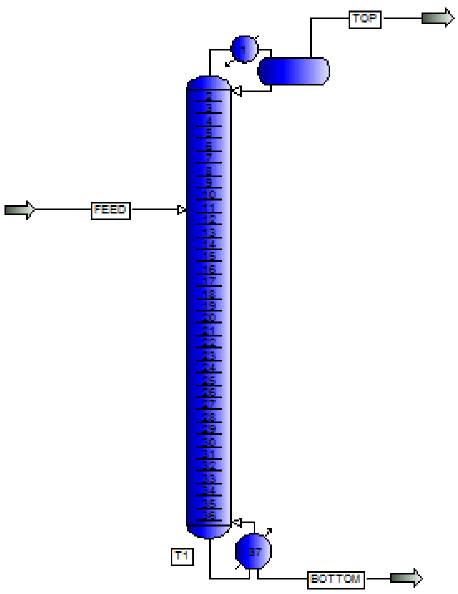


Figure 1. Deethanizer Column in the PRO / II® for application of the tool

2.2 Application of Methology in the Column

Analysis of the degree of freedom was performed to give a column Nu equal to 6, this being the number of degrees of freedom dynamic, that is, the number of manipulated variables for the process. However, considering the food as a given flow rate by a process upstream and having to be maintained at the column pressure in its pre-set value (P_{COL} = 19.2 kg.cm-2), two degrees of freedom for the system are extracted from counting performed.

With the number of degrees of freedom remaining (Nu = 4), two of them are required to stabilize the level of the condenser and referverdor, resulting in a system for the remaining number of degrees of freedom equal to 2. Consider the set of manipulated variables u_0 = [B, D, V, L], the question now is having two manipulated variables which must be controlled in the universe of variables involved in the process, so that we have a smaller economic loss.

To answer this question, one must know, the constraints and the objective function J to be minimized. From the set of MVs, the selection control structures for the column was performed considering the pair u_1 = [B, V] to the top and composition control based components of propane and ethane, respectively. The objective function is based on the deviation of ethane compositions at the bottom and propane (more valuable product for the process) at the top, the separation obtained with respect to the setpoint value of each chemical species, as shown in Eq. (4).

$$\min J = \left(\frac{x_{topo}^{P} - x_{topo,s}^{P}}{x_{topo,s}^{P}} \right)^2 + \left(\frac{x_{base}^{E} - x_{base,s}^{E}}{x_{base,s}^{E}} \right)^2 \quad (4)$$

where x is mole fraction of a given species. The indices P and E refer to species propane and ethane respectively. The index S applies to the value of the fraction of chemical species at face value.

As a next step, the optimization of the model was done by observing that the compositions of ethane at the bottom and propane on top kept on their maximum values, the presence of these two active constraints in the process entails the consumption of the remaining degree of freedom, (Nu = 0), restricted the problem of determining controlled variables. Thus, in order to keep the controlled variables at their primary values required, a study was conducted on the choice of secondary variables to control these temperatures related to the deethanizer column stages, selecting plates where occur the largest temperature variations as shown in Fig. 2.

The simulation was performed taking into consideration the study Skogestad and Hori (2007) and a previous analysis of the temperature profile, then in order to work with plates, which measuring temperature made possible in the actual column as described the work cited, the plates 19 and 25 were then added to the tool.

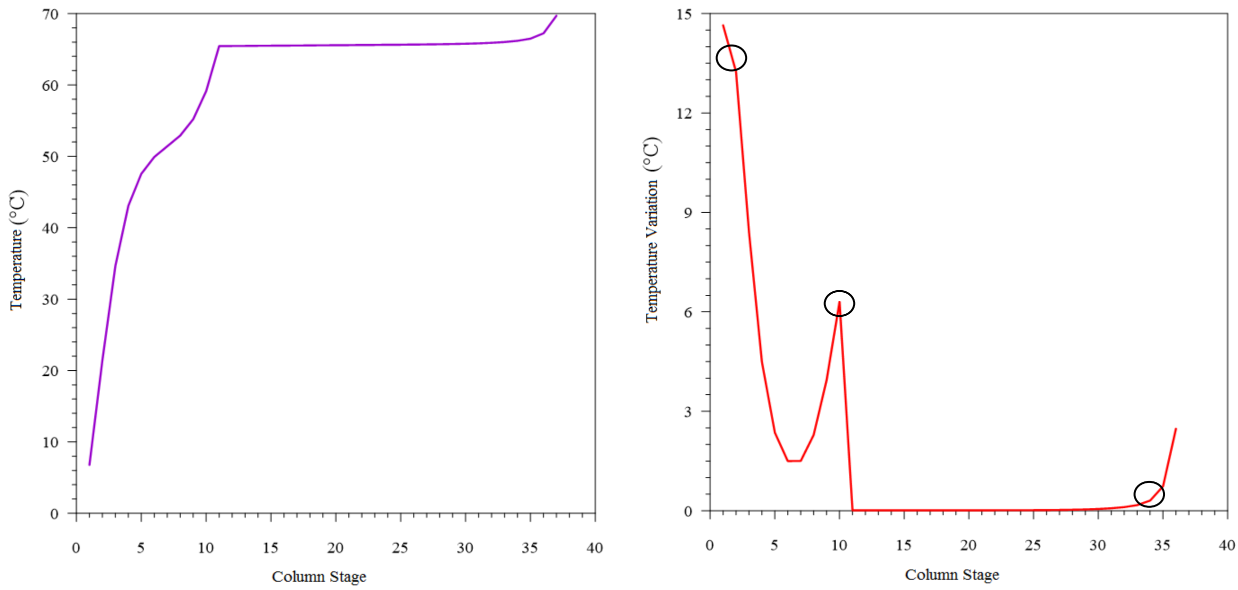


Figure 2. Temperature in the plates (top reference) obtained by simulating the PRO / II® software (a) Temperature Profile (b) Temperature variation profile

2.3 Developed Tool in VBA-EXCEL

Introducing a simple graphical interface, both input data as output data (Fig. 4) for handling the user tool is based on the simulation software communication PRO / II® (2012) owned by Invensys with Microsoft Office Excel® (2010) through the COM technology (Component Object Model). The Excel was chosen because it is an access package facilitated, and the implementation tool was accomplished in VBA Developer (Visual Basic for Applications) of the same package.

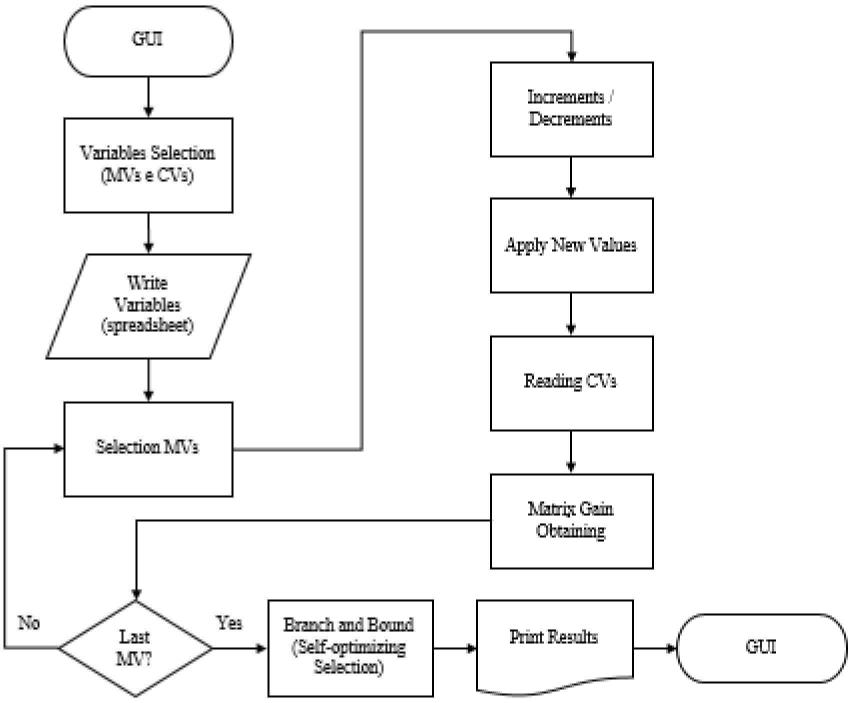
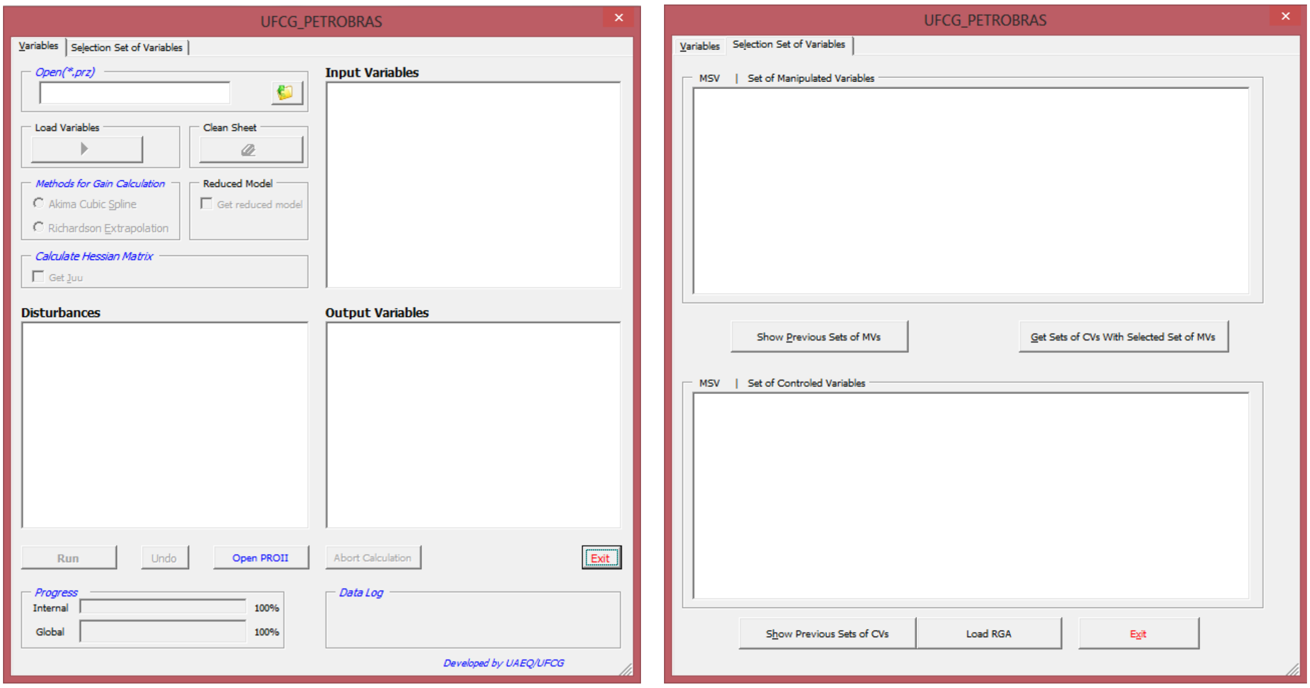


Figure 3. Flux diagram for selection tool sets of controlled variables to the plant.



(a) (b)

Figure 4. Interface selection tool (a) Graphical use interface selection tool control structures of the whole plant by Skogestad method (2000, 2004) (b) Response GUI selection tool sets of controlled variables to the plant by the Branch-and-Bound method.

3. RESULTS

3.1 Results Obtained by Tool

Through the degree of freedom for the analysis deethanizer column, as shown, a pair of MVs (manipulated variable) is previously selected based tool for the application of the methodology approached by Skogestad. The literature shows the set u_1 = [B, V] as a good set of VMs to plantwide control, so the flow of the base and the vapor flow for the process comprised the MV set considered.

This work is used the minimum singular value obtained by singular value decomposition (SVD), the selection of the controlled variables through available measurements. Tab. 2 shows eight of the best pairs of controlled variables, in descending order, obtained by the tool, ie the maximum for the minimum singular value and the position of the CV selected in the main Excel spreadsheet (BRPWC spreadsheet). You can see that for the first four subsets of controlled variables, the option to remain one of the manipulated variables of the set studied in its great value (constant) is considerable, to a smaller loss in the process.

Importantly, the method used by the tool (rule of maximum gain) works well, however in order to have greater certainty 'answer was a dynamic process analysis should be performed. Therefore in order to validate the result obtained was selected a couple of controlled variables to the manipulated variables and the results presented.

Table 2. Results generated by the first eight subsets CV tool for the set of MV = [B, V]

Controlled variables	Minimum Singular Value (α)
D + T2	0.79364
B + T2	0.79364
D + TD	0.61811
TD + B	0.61811
D + L/D	0.39558
B + L/D	0.39558
TD + T2	0.23206
L/D + T2	0.18353

3.2 Dynamic Model

The range of possibilities presented for controlled variables by applying the tool was observed above and in the case of even bigger problems, there is a large set of possible combinations to control the process. With the pair of manipulated variables being the base flow and vapor (u_1 = [B, V]), best candidates for CVs were listed in Tab. 2. In order to study the process dynamically the set of controlled variables with the highest minimum singular value after running tool, CV = [D, T2], that is, lower loss for the process is selected and then brought to Dynamics® simulation software Aspen, as shown in Fig. 5.

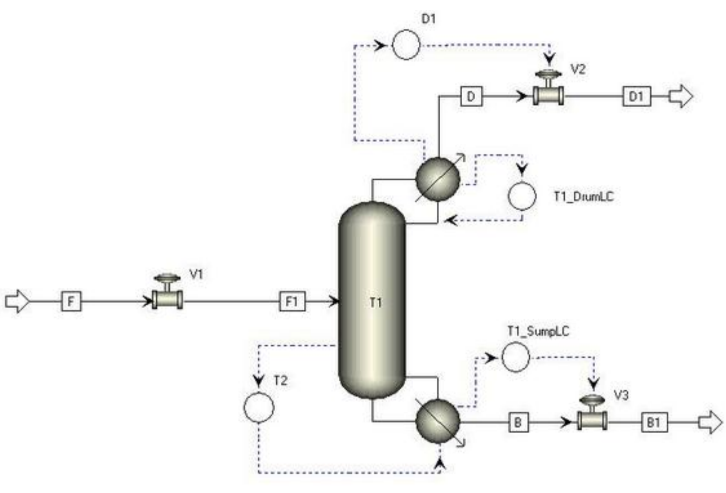
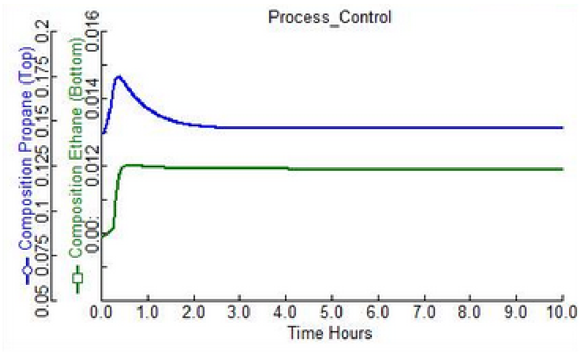
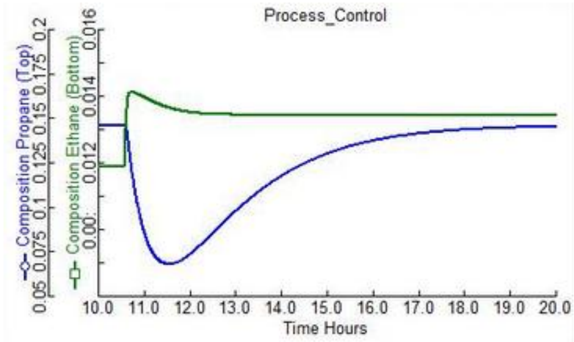


Figure 5. Deethanizer Column exported to Aspen Dynamics®

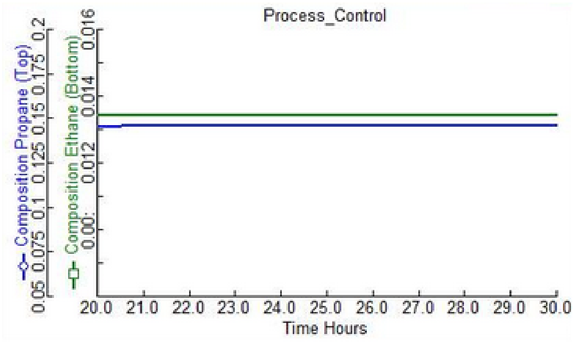
The controllers are required to interface the process variable (controlled variable) of the operating variable (manipulated variable), the first related to "indirect" control composition of ethane and propane at the bottom and top by means of top flow and the temperature T2 plate. For analysis of such compositions were built charts in itself Aspen Dynamics®, as shown in Fig. 6.



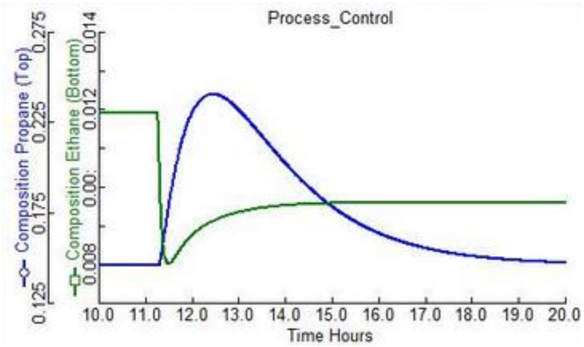
(a)



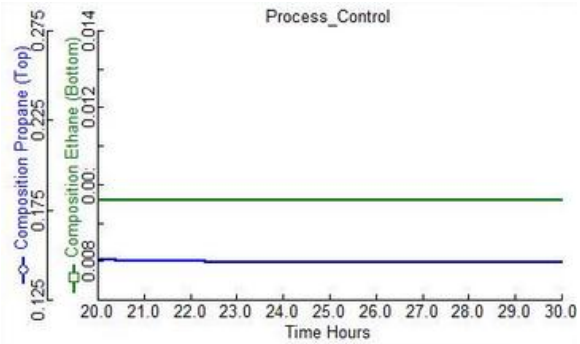
(b.1)



(b.2)



(c.1)



(c.2)

Figure 6. composition control through the temperature of the plate 2 (a) No disturbances (b.1) behavior disorder after application of + 10% (b.2) Reaction after 10 hours elapsed disorder + 10% (c.1) behavior after application of the disorder 10% (c.2) Reaction after 10 hours elapsed disturbance -10%

The behavior of composition control at the base and top to total chemical species shown in 6 (b) and 6 (c) refers to the application of a disturbance of approximately 10% in opening and / or closing the feeding valve of the column, with the power coming from an upstream process. Simulation times were selected so that each display change and stabilization process can be performed for each estimated disturbance range.

As restrictions on the process, the species of the compositions must be kept in their required values, which are $x(\text{BASE}) \leq 1\%$ ethane for the former $x(\text{TOP}) \leq 15\%$ for propane, these values were taken as basis of the analysis of cases studied. It can be seen from Fig. 6 that whereas manipulated variables as the base flow (B) and the top flow (V) of the spine, the secondary control variable selected responded well to disturbances (power system) in the process.

It is possible to observe in Fig. 6 that the case of propane at the top composition, satisfactory results were obtained, three hours elapsed since the CV subset related to flow rate and temperature of the top of the plate 2 (secondary variables) was able to return to composition of interest to within the allowable value when the performed increase or decrease of 10% in the process, there is a difference of about 0.01 compared to the setpoint for the disturbances applied.

Regarding the control of the ethane base composition, it was possible to obtain a satisfactory behavior in the same manner. Noting that if applied to a disturbance in the process 10% of the feed valve, through the secondary subset of variables adopted, was allowed to work under the ceiling (0.15) by a difference in value of 0.05 below the required limit. A small detour over the limit was observed for ethane composition control with 10% increase (disorder).

In short, it was observed that after a few hours of elapsed process, it has been possible to verify the efficiency of compositions for the control of propane and ethane, particularly for the main process component (propane), given the manipulated variables MV = [B, V] and the best set of controlled variables obtained by applying the control structures selection tool, ie, CV = [D, T2].

4. CONCLUSIONS

According to the results, you can see that the generation and application tools in the process control area of real engineering problems, greatly assists the process responsible for the selection of variables to achieve plantwide control structures that bring results economically satisfactory.

The methodologies described in the literature make an important foundation in the generation of these tools. The tool applied in this study, was based on the optimization of numerical methods and allowed to obtain results in margins required for the process in question. In addition, systematic procedures for control structures design combined with application engineering tools allow the study increasingly complex and complete, since depending on the size of the problem to the procedures described becomes of difficult execution.

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6. RESPONSIBILITY NOTICE