Effect of the demethanizer improved control strategy on the separation train for the NGL separation process

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Abstract: In recent years the attention on natural gas production and utilization is growing due to different fundamental aspects. First, the availability of natural gas has increased thanks to technological improvements in the extraction techniques that have made possible the production from unconventional reservoirs. Second, the interest in clean energy is growing, aiming to reduce CO_2 emission and thus global warming. Natural gas is a cleaner fossil fuel compared with other traditional energy sources such as oil and coal. Another reason that drives the attention on this fossil fuel is the increasing economic interest of recovering the heavier hydrocarbon fractions contained in it. The fractionation of natural gas liquids (NGL) is an energy-demanding process, often conducted with a separation train that includes cryogenic distillation columns. This work is intended to show how to achieve an energy-efficient recovery of NGLs with a proper control strategy and without composition analyzers. In particular, the effects of a combined cascade control with boilup approximation plus a pressure compensator control for the demethanizer section, on the desired NGL extraction product targets are investigated and compared under feed flowrate disturbances with conventional direct temperature controllers. Overall, it is shown that the cascade control plus pressure compensator provides the best performance and the lowest energy consumption.

Keywords: Natural gas liquids recovery, Dynamic process simulation, Pressure compensated temperature, Multicomponent distillation

1. INTRODUCTION

The consequence of the so-called "shale gas revolution" (Caporin and Fontini, 2017) derived from technological advances in extraction methods, including hydraulic fracturing and horizontal drilling, is an increase in the availability of natural gas and a decrease in the selling price dropped to less than 30% of its previous highs (Luyben, 2013a; Caporin and Fontini, 2017; Feng et al., 2019). A further drive to the production and use of natural gas is given by the growing interest in the climate issues caused by global warming, leading to the demand for cleaner and more environmentally friendly fuels. In this context, natural gas, mainly used as a fuel, is seen as a cheaper and cleaner alternative to traditional fossil fuels such as coal and oil. Natural gas is a hydrocarbon gas mixture constituted mainly of methane, and a variable fraction of heavier hydrocarbons, known as natural gas liquids (NGL). The increasing availability of natural gas opens the door to another economically profitable process, the separation of NGL. These liquefied hydrocarbons can be sold separately and used as feedstock for various industrial processes with a significantly higher market value than as part of the raw gas (Mokhatab et al., 2015). Numerous studies have focused on the development of novel process configurations for the NGL fractionation process. A widely used technology for the separation of NGL is cryogenic distillation due to its ability to recover the product at high purities (Olsen et al, 2012). The

fractionation unit consists of a train of distillation columns in which first a cryogenic distillation for methane separation takes place. Subsequently, ethane, propane and butanes are separated using several distillation columns in a direct sequence arrangement. The first distillation column of the train, the demethanizer, is the heart of the separation process. Various process modifications have been studied to enhance this separation, leading to the development of different process schemes (Newaz and Jobson, 2010). One of the first proposed industrial schemes was the conventional turboexpander process. In this unit, a turbo-expander and Joule-Thompson valves are used to implement the pressure jump between the inlet flows and the nominal conditions inside the column, providing the necessary refrigeration for cryogenic distillation (Campbell and Wilkinson, 1981). Starting from this process scheme, many improvements in the design of NGL recovery processes have been developed to improve separation performance and reduce operating costs. Among the process schemes developed, one of the most widely used in industry is the Gas Subcooled Process (GSP). The GSP unit utilises the split-vapour concept to improve recovery. In this process, part of the feed is used as column reflux to contact and rectify the vapour in the column (Pitman et al., 1998). A modification of the GSP scheme is the Cold Residue Recycling (CRR) process scheme. The CRR retains all the main advantages of GSP and further enhance the ethane recovery by providing a reflux stream of almost pure methane to the column (Wilkinson and Hudson, 1992). These separation processes are generally

influenced by disturbances in operating variables such as pressure and temperature (Chebbi et al., 2010) and feed conditions. The most common disturbances are variations in inlet flowrate and composition. The flowrate may be subject to variations dictated by changes in natural gas demand, while the feed composition may be subject to fluctuations due to the characteristics of the natural gas extraction basin. Therefore, to maintain the production targets, it is necessary to analyse the process dynamics and design a control system that can mitigate or eliminate the effect of these disturbances.

The NGL separation process has been extensively explored in the literature by numerous authors including Manning and Thompson (1991), Long and Lee (2012), Lee et al. (2012), Park et al. (2015), Mokhatab et al. (2015), Kherbeck and Chebbi (2015) and Uwitonze et al. (2020). However, in those studies, the complexity of the dynamics involved in the process and their control are not considered. Direct control of concentrations in the presence of disturbances in feed flowrate and inlet composition has been investigated by Luyben (2013a, 2013b), Chebeir et al. (2019) and Zhu et al. (2020). This type of control strategy is the simplest in terms of construction, but it has also several negative aspects, related to the long delay times within the control loops resulting from the measurement times of the instruments and the high costs of purchasing and maintaining the equipment. An alternative to direct control is the indirect control of composition through the knowledge of temperature measurements in the column (Hori and Skogestad, 2007). The use of indirect composition controls, employing temperature controllers, despite being widely used in industry, may not be the best control strategy to guarantee product specifications. Improvements to this strategy have been introduced for the NGL recovery process by Mandis et al. (2021). In this contribution, a control of the pressure compensated temperature (PCT) in the separator placed upstream of the separation train was implemented to improve the ethane recovery. The methane impurity level was maintained with a cascade control structure where the primary loop was given by the column tray temperature controller and the secondary loop was a ratio control between a boilup estimation and the column bottom product.

This work aims to evaluate the effects of the modifications proposed by Mandis et al. (2021) on the production targets of a CRR unit and the remaining columns of the fractionation train and to show how to achieve an energy-efficient recovery of NGL. The alternative control strategies are investigated and compared under feed flowrate disturbances.

2. BACKGROUND

2.1 Flowsheet

The NGL fractionation process based on realistic operating conditions (Chebeir et al., 2019) takes place in a CRR unit and subsequent distillation columns under feed flowrate disturbances. This process is simulated using the Aspen HYSYS® process simulator with the fluid package given by the Peng-Robinson equations of state (EOS). The feed to the plant is represented by a natural gas mixture with low content of liquids, whose feed composition is reported in Table 1. The

nominal feed conditions considered are a flowrate of 4980 kmol/h, a pressure of 5818 kPa and a temperature of 35 °C. Variations of feed flowrate are realized by manipulating the inlet line pressure of the plant, thus producing pressure variations in the separator located upstream to the separation train.

Table 1. Feed composition (Chebbi et al., 2010)

Components	Mole fractions
Nitrogen	0.01
Methane	0.93
Ethane	0.03
Propane	0.015
Butanes	0.009
Pentanes	0.003
Hexanes	0.003
%C2+	6

2.1.1 CRR Unit

The demethanizer located in the CRR unit consists of a 30 stages distillation column with a reboiler, no condenser and three feed streams. In this unit, the feed flow is precooled by using a heat exchanger and a chiller and sent to a separator where the liquid fraction formed is removed. Part of the vapour and liquid stream is fed to the 2nd tray of the column. The remaining vapour fraction is depressurized through a turbo-expander and entered in the column in the 8th tray, while the remaining liquid fraction is fed to the 26th column tray. To provide column reflux, the top product is split into two streams one of which is compressed using a cryogenic compressor, cooled, and sent to the top of the demethanizer as reflux of nearly pure methane. More details on the demethanizer column can be found in Mandis et al. (2021). The desired specifications of the demethanizer separation are an ethane recovery of 84% in the top product and a methane impurity level of 1 mol% in the bottom product.

2.1.2 Deethanizer column

The deethanizer is a 30 stages distillation column with a reboiler and a condenser. This unit is fed at the 12th tray with the liquid stream leaving the bottom of the demethanizer column. The feed conditions were a flowrate of 277 kmol/h, a temperature of -5.1°C and a pressure of 2459 kPa. The column removes the propane from the mixture, collecting an ethane product of high purity in the top product. The production targets required for the deethanizer separation are an impurity of propane of 1 mol% in the top product, and ethane concentration of 0.34 mol% in the bottom product.

2.1.3 Depropanizer column

The depropanizer is constituted by a 46 stages distillation column with a reboiler and a condenser. The feed, represented by the liquid stream leaving the bottom of the deethanizer, is introduced at the 19th tray fed with the bottom product of the deethanizer column. The feed conditions were an inlet flowrate of 148 kmol/h, a temperature of 83.5°C and a pressure of 1760 kPa. The specifications required for this unit are: column top product with a value of 0.6 mol% in the sum of the concentrations of iso-butane and n-butane; column bottom product with a propane concentration of 0.1 mol%.

2.1.4 Debutanizer column

The debutanizer is a 35 stages distillation column with a reboiler and a condenser. This column is responsible for the removal of the butane fraction from the remaining mixture. The debutanizer is fed at the 18th tray with the bottom product stream of the depropanizer column. This stream has a flowrate of 74 kmol/h, a temperature of 78.5°C and a pressure of 608 kPa. The product targets required for this separation are an impurity of iso-pentane of 0.2 mol% in the column top product and an n-butane impurity of 0.2 mol% in the bottom product.

2.2 Temperature sensor placement

The choice of the control system used to maintain the required product specifications for the deethanizer, depropanizer and debutanizer was based on the analysis of the composition and temperature profiles of the columns in steady state conditions. The possibility of using indirect composition controls employing temperature measurement was evaluated, considering the absence in the plant of expensive composition analysers. For the separation of multicomponent mixtures, the positioning of the temperature sensor and the effective presence in the column of detectable temperature variations are of most importance when temperature control is used instead of direct composition control. The thermocouple best position is expected in the section where the temperature is the most sensitive in response to key components variations. To highlight the variations of the temperature gradient in the column due to the variations in the key component concentrations and to consider the effects on this gradient of the non-key components, a temperature gradient analysis was carried out with a contribution diagram per component (Porru et al., 2013). The resulting diagrams of the temperature gradient with a per-component contribution diagram analysis and the total temperature gradient applied to the three columns are shown in Figures 1-3.
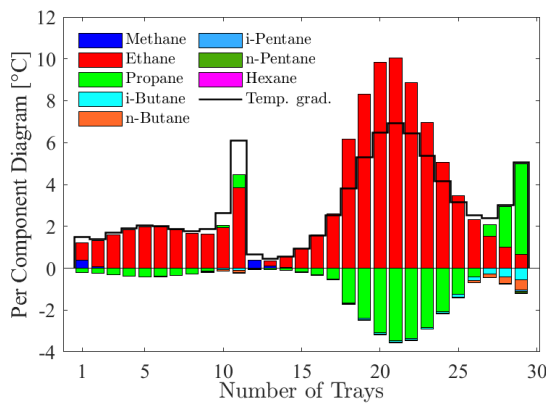

Figure 1: Deethanizer per component diagram; 1 is the top tray.

In the diagram in Figure 1, the results obtained for the deethanizer column are depicted. Except for the temperature variation around the feed plate, located in the 12th tray, the per component analysis can describe the influence of composition variations in the temperature gradient inside the column. As it is possible to visualize in the rectifying section, the variations of the heavy key component (propane) produce a negligible impact in the temperature profile. Thus, the implementation of a temperature controller in this section is not recommended.

On the other hand, light key component (ethane) variations in the stripping section have a high impact on the temperature profile, reaching a peak located in the 21st tray. Thus, the temperature sensor was located in this tray.
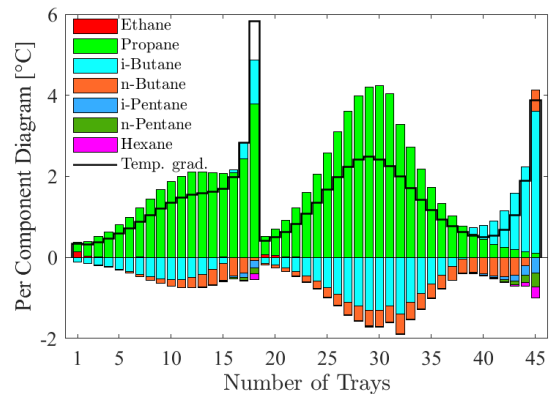

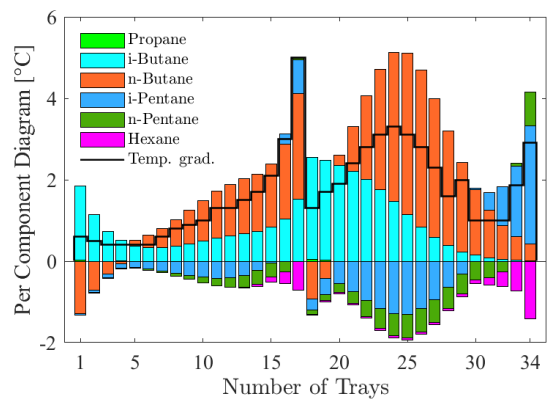Figure 2: Depropanizer per component diagram; 1 is the top tray.

Figure 3: Debutanizer per component diagram; 1 is the top tray

Analog results are obtained with the per component analysis performed on the depropanizer column (Figure 2) and debutanizer column (Figure 3). In the rectifying section of the depropanizer, the influence of the variation of the sum of iso-butane and n-butane leads to a region of trays (10-13) where a maximum in the temperature variations is not present. Additionally, the heavy key component (iso-pentane) variations produce neglectable variation in the temperature gradient in the same section of the debutanizer. Therefore, placing a temperature sensor in the enrichment section of depropanizer and debutanizer columns is not possible. As seen for the deethanizer, in the stripping section of depropanizer and debutanizer, the effect of the light key component (propane and n-butane, respectively), has a major influence in the temperature gradient of the columns. The trays where the influence on the tray temperature is the largest are the 30th in the depropanizer and the 26th in the debutanizer columns. Therefore, the temperature sensors for the stripping section of depropanizer and debutanizer columns were positioned in these trays.

3. CONTROL STRUCTURE

In this section, the control structures implemented to meet the production targets under the effect of feed disturbances are

described. Considering the high cost of composition analysers and the delays introduced by employing composition controllers, only indirect controls of concentrations has been applied in the section where temperature sensors were present. Reflux ratio controllers are used in the enriching section of the columns to mitigate the feed disturbances (Luyben, 2005). The implemented control loops for system stability, as pressure and level controllers, are not included in the discussion for the sake of brevity while the others are PI controllers tuned with the improved internal model control.

3.1 Demethanizer column control strategy

The demethanizer column has the purpose of removing the methane fraction from the feed gas mixture. The objective of the control strategy is the achievement of an ethane recovery of 84% and a level of methane impurity of 1 mol% in the bottom product of the column. Different control strategies were developed to optimize the maintenance of purity targets. The conventional approach, consisting of basic temperature controllers is described as follows:

1) The impurity composition in the bottom is maintained by controlling the temperature of the 28th tray by manipulating the reboiler duty;
2) The ethane recovery is maintained by controlling the separator temperature by manipulating the heat removed at the chiller.

The second control structure considered in this work is the one developed by Mandis et al. (2021) to improve the disturbances rejection and the control performances for the demethanizer operations. This control structure is explained as follows:

1) The maintenance of ethane recovery was achieved with a pressure compensator in the separator, by using the PCT as a controlled variable in the control loop;
2) The methane impurity at the bottom of the demethanizer column was maintained through the implementation of a ratio control between a boilup estimation and the column bottom product, in cascade with the column tray temperature controller.

3.2 Deethanizer column control structure

This column is responsible for separating the ethane from the stream leaving the demethanizer. The control strategy aims to maintain the propane concentration at 1 mol% in the top product and the impurity of ethane at the value of 0.34 mol% in the bottom product. To reach the bottom production target, thus, indirectly control the ethane composition in the column bottom product, the temperature of the 21st tray is controlled by manipulating the reboiler duty.

3.2 Depropanizer column control structure

The depropanizer column is used for the propane separation with the control objectives of maintaining impurity level values of 0.6 mol% for the sum of the concentrations of iso-butane and n-butane in the top product and 0.1 mol% for the propane concentration in the bottom product. To maintain the level of propane impurity in the column bottom product, this is indirectly controlled by controlling the temperature of the 30th tray by manipulating the reboiler duty.

3.3 Debutanizer column control structure

The purpose of this column is the separation of iso-butane and n-butane from the remaining heavier hydrocarbon mixture. The control goal consists in holding constant the impurity of iso-pentane of 0.2 mol% in the top product and the impurity of n-butane of 0.2 mol% in the bottom product. In order to maintain the concentration impurity target in the column bottom product, the n-butane composition is indirectly controlled by controlling the temperature of the 26th tray by manipulating the reboiler duty.

4. RESULTS

Under the presence of 10% variation in the plant feed flowrate (chosen following the feed changes used by Chebeir et al. 2019), we intend to verify the influence of the control strategy developed to optimize the achievement of the demethanizer column production targets on the other distillation columns of the train. The results of the implementation of the direct control of temperatures (indicated as conventional control structure) and the cascade arrangement with PCT control (indicated as proposed control structure) are compared and evaluated by considering the dynamics profiles obtained for impurity composition levels and reboiler duty of the deethanizer, depropanizer and debutanizer columns.

4.1 Deethanizer column results

The composition and the reboiler duty dynamic profiles obtained for the deethanizer column with the conventional and the proposed control structure, in presence of plant feed flow variation, are depicted in Figure 4. Here, it is possible to assert that, by increasing and decreasing the plant inlet flowrate with an amplitude of 10%, with the proposed configuration, the propane concentration profiles (Figure 4, top panel) show a lower initial variation and a higher speed of response. Additionally, for decreasing variation in the feed flowrate, the propane concentration profile exhibits, with the conventional control structure, a much higher offset when compared with the proposed one. In presence of both the considered plant feed variations, the ethane concentration profiles (Figure 4, middle panel) obtained with the proposed configuration depicted an improvement in the control performance. It is possible to observe a lower initial variation, higher speed of response and a lower offset at the new steady states. Under opposite variation of the same amplitude in the feed flowrate, with the proposed configuration, propane and ethane concentration profiles show symmetric behaviors. Thus, by using the cascade structure with PCT control, it is possible to linearize the behaviors of the deethanizer product targets. Moreover, by considering the reboiler duty (Figure 4, bottom panel) with the proposed control, it shows a lower variability, reductions of 22% and 28% are registered, respectively, for decreasing and increasing the plant feed. This allows to reach the product target with lower power consumption.

4.2 Depropanizer column results

Figure 5 shows the depropanizer dynamic profiles obtained for impurity concentration and reboiler duty with the conventional and the proposed control structures.
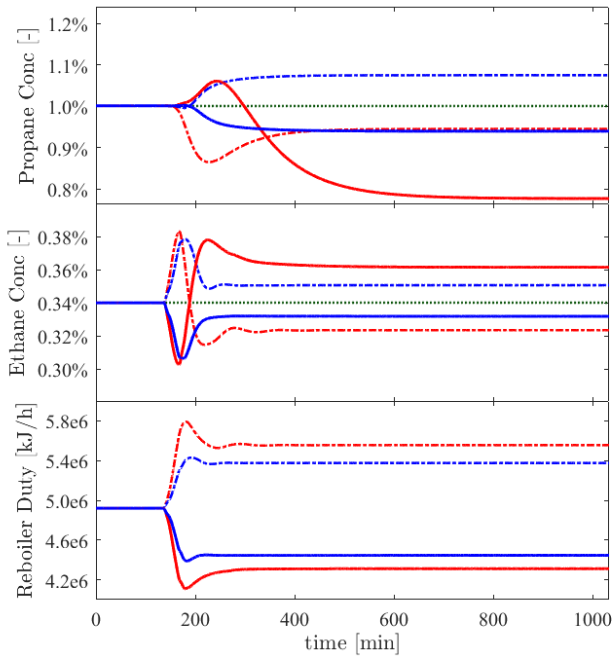
Figure 4: Responses obtained with the deethanizer column by the conventional control strategy (red lines) and the proposed control strategy (blue lines) for a 10% increase (dash-dotted lines) and a 10% decrease (continuous lines) in the plant feed flowrate. The green dotted lines depict the control targets.

By looking at the graphs obtained for a 10% increase in the nominal feed to the plant and a 10% decrease of the same variable, it is possible to assert that the impurity in the top product of the depropanizer (sum of the concentrations of iso-butane and n-butane depicted in the top panel of Figure 5) exhibits different behaviours. The impurity concentration profile shows, with the conventional control, a slightly higher initial variation, and a lower offset under the effect of an increase in the feed flowrate to the plant. In contrast, in the worst-case variation, represented by a decrease in the feed flowrate, and utilizing the proposed control strategy, a lower initial variation and a lower offset are obtained in the process.

Under the effect of both the applied disturbances, the same behaviors are observed for the propane concentration profiles (Figure 5, middle panel). Nevertheless, for meeting the control objective regarding the top and bottom impurities levels in the depropanizer products, with the proposed control strategy, it is possible to mitigate the effects of the worst variation, corresponding to the decreasing of the feed flowrate to the plant. Additionally, with the proposed configuration, the impurity concentration profiles in the top and bottom products depict symmetric behaviors, in response to symmetric variations in the disturbance applied. Thus, by this control structure, the dynamic behaviors of the depropanizer column product targets are linearized. By considering the bottom panel in the Figure 5, with the proposed control, the actions of the reboiler duty show a lower initial variation but in general, the actions obtained with the conventional and proposed control structure are comparable for both the variations applied.
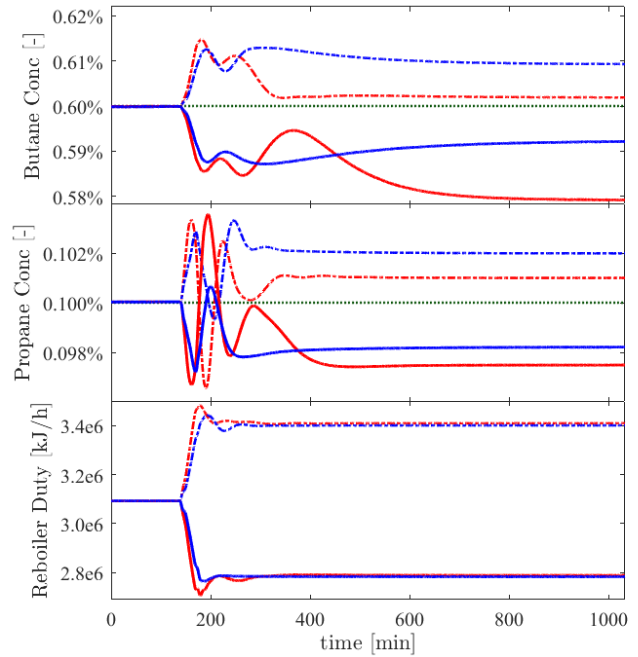

Figure 5: Responses obtained with the depropanizer column by the conventional control strategy (red lines) and the proposed control strategy (blue lines) for a 10% increase (dash-dotted lines) and a 10% decrease (continuous lines) in the plant feed flowrate. The green dotted lines depict the control targets.

4.3 Debutanizer column results

The composition profiles and the action of the reboiler duty, obtained in the debutanizer column with the conventional and the proposed control structure, are depicted in Figure 6, for a 10% increase and decrease in the plant feed flowrate.
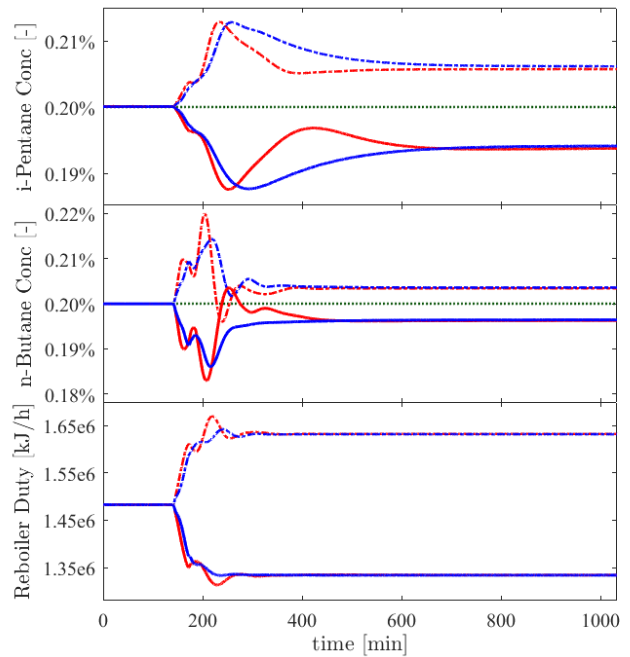

Figure 6: Responses obtained with the debutanizer column by the conventional control strategy (red lines) and the proposed control strategy (blue lines) for a 10% increase (dash-dotted lines) and a 10% decrease (continuous lines) in the plant feed flowrate. The green dotted lines depict the control targets.

By looking at the profiles, it is possible to assert that, considering the small composition variations, the iso-pentane impurity level profiles in the top product of the debutanizer (Figure 6, top panel) shows a similar behavior with both control configurations. However, the deviation from the target value appears to be lower during the transient time with the conventional structure. Under the effect of the two disturbances applied, by considering the n-butane profiles (Figure 6, middle panel) obtained with the proposed configuration, it is possible to observe a lower initial variation and a higher speed of response. By considering the actions of the reboiler duty (Figure 6, bottom panel), a lower initial variation is obtained with the proposed control. In general, the profiles obtained with the two control structures are comparable under both flowrate variations considered in the process.

5. CONCLUSIONS

It has been demonstrated that the use of the proposed control structure, given by the cascade and PCT control, can improve not only the control performance and the energy consumption achieved on the demethanizer but also has positive effects on the rest of the plant. The use of the proposed control configuration proves to be able to linearise the inherent non-linearities in the dynamic behaviour of the impurity concentrations at the top and bottom of the deethanizer and depropanizer. The comparison between the action of conventional and the proposed control structures on the impurity levels of the top and bottom product of deethanizer and depropanizer and for the bottom product of the debutanizer showed, with the proposed structure, best control performance or at least managed to improve the responses obtained in the worst variation. In addition, with this control scheme, less variability in the deethanizer reboiler duty was registered, resulting in the achievement of the product target with lower energy consumption.

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