Control of a natural gas liquid recovery plant in a GSP unit under feed and composition disturbances

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Abstract: Recent technological improvements have driven the rapid increase in natural gas production from unconventional reservoirs. The heaviest hydrocarbon fraction of this fossil fuel, the so-called natural gas liquids (NGL), have greater economic interest justifying the attention on its separation process from the raw gas. Various process schemes have been developed and studied for the NGL recovery, including the conventional, cold residue recycle (CRR), and the gas subcooled process (GSP). This study aims to assess different control strategies for a GSP unit and determine the most appropriate and effective process control scheme. For this, the dynamic responses for each control scheme are evaluated by changing feed flow rate and composition. The main targets are the achievement of 84% ethane recovery and low levels of methane impurity at the bottom of the demethanizer column. Due to the high cost of composition analyzers and the high delays introduced by composition controllers under the presence of flow disturbances, the control goals are reached by the knowledge of on-line temperature measurements. This is done by considering different temperature control structures such as the direct temperature control and cascade control, plus a pressure compensator. The results are compared, in presence of composition disturbances, with the action of a hybrid cascade control that uses in-line delayed concentration measurements to update the controller reference at each sampling period. Here, the hybrid and the simple cascade controls show the best control performance.

Keywords: Natural gas liquids recovery, Dynamic process simulation, Pressure compensated temperature, Hybrid cascade control, Discrete composition control.

1. INTRODUCTION

Among the different fossil energy sources, the natural gas represents one of the cheapest and cleanest. This energy source is mainly used as an alternative to more traditional fossil fuels such as oil and coal. The use of natural gas to produce energy is not a recent practice (Kidnay et al., 2011), but in the last decades its utilization has increased due to improvements in extraction techniques including drilling and hydraulic fracturing. This has led to an upsurge in natural gas availability and a decrease in its selling price, dropped dramatically to less than 30% of its previous highs (Luyben, 2013). Natural gas is a naturally occurring hydrocarbon gas mixture consisting mainly of methane, but commonly including varying amounts of heavier hydrocarbons. The presence of these hydrocarbons, known as NGL (Natural Gas Liquids), represents an additional source of profit: these compounds are separated and used as feedstock for different industrial processes. Consequently, the NGL have significantly higher value as separate marketable products than as part of the raw gas (Mokhatab et al., 2015). For this reason, numerous studies have focused on the development of novel process configurations to separate these products.

Different process schemes have been studied and developed for the NGL recovery. One of the first proposed industrial schemes was the conventional turboexpander process. In this unit, refrigeration is supplied through the implementation of pressure jump between the inlet stream and the nominal conditions inside the column by means of a turbo-expander (Campbell and Wilkinson, 1981). Starting from this process scheme, plant modifications and energy integrations have been realized to improve the separation performance and reduce operating costs. Among the new process schemes, the Gas Subcooled Process (GSP) is one of the most widely employed units in industry. GSP utilizes the split-vapour concept to obtain a column reflux that contacts and rectifies the vapor leaving the turboexpander and entering the column in the lower stages (Pitman et al., 1998). The Cold Residue Recycling (CRR) process represents an extension of the GSP design. CRR retains all the main advantages of the GSP while adding a column reflux stream of nearly pure methane (Wilkinson and Hudson, 1992). This enhances the levels of ethane recovery in the separation process. The NGL recovery process is normally influenced by disturbances in operating variables such as pressure, temperature (Chebbi et al., 2010) and feed conditions. The most common disturbances are those related to inlet flow rate and composition variations. The flow rate may be subject to changes dictated by the need to treat or produce a smaller or larger amount of product, while the composition can be subject to fluctuations due to the characteristics of the natural gas extraction basin. To maintain

the product specifications of NGL recovery plants, it is necessary to analyse the process dynamics and design a control system capable of mitigating or eliminating the effect of these disturbances.

Several NGL recovery processes have been studied in the literature by Manning and Thompson (1991), Kidnay et al. (2011), Chebbi et al. (2010), Park et al. (2015) and Kherbeck and Chebbi (2015). However, these studies do not include the inherent complexities involved in the dynamics of the process and their control. The direct control of concentrations in the presence of disturbances in feed flow rate and inlet composition has been analysed by Luyben (2013) and Chebeir et al. (2019). In terms of realization, this type of control strategy is simple, but it possesses disadvantages related to the long delay times inside the control loops resulting from the measurement times of instruments and the high purchase and maintenance costs of equipment. An alternative to direct control is the indirect control of composition through temperature measurements (Hori and Skogestad, 2007). With this type of control, the choice of the tray with the highest temperature sensitivity and the consideration of non-key components impacting on the column temperature profile are essential in cases of multi-component columns.

In this work, the dynamic behaviour of the system resulting from the separation of NGL in a GSP unit is investigated under the effect of disturbances in flow rate and input composition, by comparing different control strategies. Particularly, an improvement of the bottom methane impurity control in the demethanizer column is achieved through the implementation of a ratio flow controller in cascade with temperature controller, with and without remote setpoint selection by delayed composition measurements. Additionally, a pressure compensator has been added in the separator to enhance the ethane recovery control. The purpose of this study is to evaluate the use of alternative control schemes concerning the maintenance of product specifications in the demethanizer column.

2. BACKGROUND

2.1 Flowsheet

The NGL separation process that takes place in a GSP unit, depicted in Figure 1, has been analysed by means of a dynamic simulation using Aspen HYSYS® and based on realistic operating conditions (Chebeir et al., 2019). The raw gas is fed to the plant with a flow rate of 4980 kmol/h, at the pressure condition of 5818 kPa and a temperature of 35 °C. The composition with a low content of liquids is reported in Table 1. The specifications required by the GSP separation unit are to achieve an ethane recovery of 84% and a methane composition of 1 mol% at the bottom of the demethanizer column (T-100).

The main part of the separation unit consists of a 30 stages demethanizer column with a reboiler (E-103) and no condenser. A separator (TK-100) is placed upstream of the column to remove the liquid fraction of the feed gas stream. Then, this condensed vapor is introduced into stage 26. Part of the vapor is sent to a turboexpander (TE-100) and fed on stage 8. The remaining vapor is cooled, expanded and fed to the top of the column.

An important variable for ethane recovery is given by the separator temperature TI-100, which depends on the conditions achieved in the chiller (E-101) used to precool the feed gas. The remaining refrigeration needed to meet the column nominal temperature is provided by expanding the column feeds through the turboexpander and Joule-Thompson valves (JT-100 and JT-101).

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

2.2 Temperature sensor placement

For implementing an indirect composition controller to maintain the level of methane impurity in the column bottom product below a specific value, the selection of the temperature sensor position is of most importance. In a multicomponent column, the best position to place a temperature sensor is in the section where the sensitivity to temperature changes is higher. Consequently, an analysis of the column temperature and composition gradients has been carried out. To relate the influence of components variation on the temperature gradient inside the column, the analysis of the temperature gradient with a per-component contribution diagram (Porru et al., 2013) has been performed. In this manner, the best position for the temperature sensor is determined to be in tray 28.

3. SYSTEM DYNAMICS AND CONTROL STRUCTURE

Under the effect of feed and composition disturbances, the achievement of the control objectives has been evaluated considering different control configurations. This work proposes a modification of the control structure realized by Chebeir et al. (2019) for the same plant, as it is depicted in Figure 1. First, basic indirect controls of concentrations through temperature control have been applied in order to achieve both ethane recovery and methane concentration targets in the column bottom product. Subsequently, several improvements have been examined to eliminate or mitigate, as far as possible, the steady states offsets.

3.1 Indirect temperature controllers

To reach the target values of 84% for ethane recovery and 1 mol% for methane concentration in the demethanizer bottom, two indirect composition controllers (hereafter indicated as C1 configuration) have been implemented using on-line

temperature measurements in the separator and column. In the first control loop, the controlled variable is the separator temperature TIC-100, which is directly connected with the ethane recovery. The manipulated variable is the chiller duty. In the second control loop, the controlled variable is represented by the tray temperature TIC-103 at stage 28, related to the concentration of methane, by manipulating the reboiler duty.

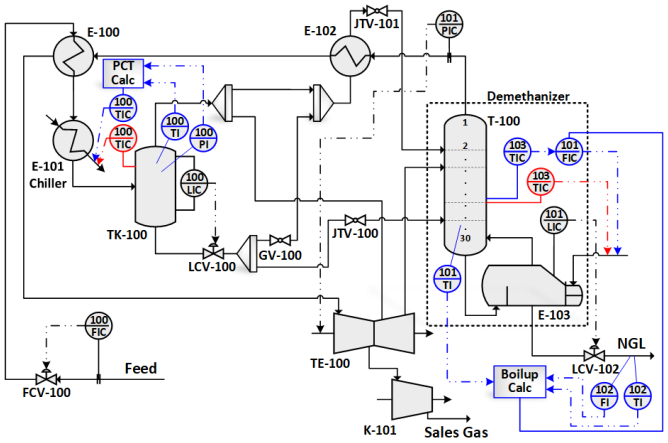

Figure 1: Schematic of the GSP unit and control structures: in red the conventional configuration (C1); in blue the C3 configuration.

3.2 Pressure compensate temperature

The feed flow variations are induced by manipulating the lines pressure through the valve upstream the separator (valve FCV-100 in Figure 1). This implies that the separator pressure PI-100 also undergoes variations. Since the separator temperature TIC-100 is maintained constant by the control action, pressure variations result in changes in the separator outflow composition that influences the ethane recovery. With the purpose of improving the separator direct temperature control and, consequently, the ethane recovery, a pressure compensator is placed in this unit. A new control variable has been considered within the loop instead of the separator temperature. This variable is represented by a fictitious temperature, calculated on basis of Antoine's law. This is a Pressure Compensated Temperature (PCT) estimated by using (1) (Brambilla, 2014):

$$PCT = T_m - C \ln \frac{P}{P_{rif}} \tag{1}$$

where T_m is the measured separator temperature TI-100; P is the separator pressure PI-100; P_{rif} is the separator pressure nominal value; and C is the compensation coefficient obtained from temperature data recorded during the imposition of pressure variations at constant composition with Aspen HYSYS®. By controlling the PCT (control configuration indicated with C2), it is possible to link the temperature variations to the pressure changes in the separator and mitigate the resulting concentration disturbances.

3.3 Indirect methane cascade control

Since the flow dynamics are faster than the ones of temperature and composition, a ratio flow controller FIC-101 is introduced in the control strategy. The resulting control is a cascade control system (this combined with PCT control is hereafter indicated with C3) used to speed up the responses obtained by the direct column tray temperature controller. The external loop is represented by a temperature controller TIC-103 that gives the setpoint to the internal loop. The internal loop is composed by a flow controller, the purpose of which is to maintain constant the ratio between the boilup and the column bottom product, by manipulating the reboiler duty. It is important to note that the ratio controller could guarantee the required methane composition at the bottom of the demethanizer column in case of feed flow variations, but it fails when feed composition changes. In this case, a temperature controller is required (Shinskey, 1996).

3.3.1 Boilup approximation

Considering that a boilup measurement is not generally available in real plants, an estimation of this variable is needed to implement the cascade control structure. A boilup approximation is obtained by the process information available and based on a balance around the reboiler expressed by (2):

$$\lambda V + cp(T_B - T_L)(V + B) = Q \tag{2}$$

Here, B is the bottom product stream FI-102; V is the actual boilup; Q is the reboiler duty; λ is the latent heat of vaporization of the mixture; cp is the specific heat capacity at constant pressure; T_B and T_L are the temperature of the stream B (TI-102) and the temperature of the liquid stream entering the reboiler (TI-101), respectively. Considering different values of inlet flow rate, the open-loop response of V, B, T_L, and T_B are registered by using the model testing function of Aspen HYSYS® on the reboiler duty with an amplitude of 2%. These data are used to perform a multi-linear regression considering the regression model (3):

$$\hat{V} = p_0 + p_1 B + p_2 T_B + p_3 T_L \tag{3}$$

where $\hat{V}$ is the boilup estimation. The regressors considered are the registered data of B, T_L, and T_B while p_0, p_1, p_2, p_3 are the model parameters.

3.3.2 In-line methane measurements

To compare this contribution with the studies developed by Luyben (2013) and Chebeir et al. (2019), the presence of a gas chromatograph is assumed in the plant. In general, only delayed and discontinuous composition measurements are available due to the typical delay in the in-line analysers, not considered in the mentioned contributions. To approach a real situation, it is assumed that the gas chromatograph is used to analyse the product composition of all the distillation units present in the distillation train. This is a common practice in

industrial plants, where only discrete and delayed composition measurements are available for the controller. Since further NGL separation requires other two distillation columns downstream the demethanizer and considering an analysis time of 10 minutes, reasonable for this type of mixtures, a sampling time of 30 minutes has been assumed. This takes into account the time needed by the instrument to complete a measuring cycle in the entire plant. Thanks to the knowledge of this measurement, another controller is added to the above-mentioned cascade. In particular, the discrete methane composition controller only acts in each delayed sampling time to adjust the shot of the temperature controller setpoint, leading to the implementation of a hybrid cascade controller (hereafter indicated with C4). To compare the hybrid control results with the conventional control ones, the methane concentration controller is also used in cascade with the conventional temperature controller (control scheme C5). The tuning parameters, obtained using the improved internal model control (IMC) approximate model controller tuning rules (Ogunnaike and Ray, 1994), are reported in Table 2 for the different control structures.

Table 2. PID tuning parameters

Controllers	Configuration	Kc	τ_I
Temperature	C1; C5	1.222	19.45
Concentration	C5	0.18	9
Temperature	C3; C4	0.802	11.5
Ratio	C3; C4	10.992	8.6
Concentration	C4	0.3	9

4. RESULTS

With the purpose of meeting the control objectives, the different control structures have been evaluated under the presence, in the feed gas to the plant, of 10% variation in the flow rate and 40% variation in the ethane concentration. These variations are chosen in accordance with the feed changes used by Chebeir et al. (2019) and the composition changes used by Luyben (2013). In this section, a comparison between the results obtained through the conventional control and the proposed control structures is performed. These results are relative to the worst-case variations corresponding to a decrease in the feed flow rate and the ethane concentration.

4.1 Ethane Recovery under feed disturbances

The ability to reach the ethane recovery target has been evaluated by analyzing the dynamic responses of the GSP unit under feed disturbances. A comparison between the conventional control (C1: direct temperature control in the separator) and the proposed control (C2: direct PCT control) is performed. Particularly, the use of a PCT control instead of temperature control is addressed to overcome the pressure changes induced in the separator by the inlet flow rate variations to the plant. This is an important aspect when feed flow upsets are present. The ethane recovery responses, performed without concentration control, for a decreasing of 10% in the feed flow rate are depicted in Figure 2.

By observing the top graph in Figure 2, it is possible to notice that the two temperature controllers are able to bring the process variables to the setpoint value after approximately 80 min with the proposed control and 95 min with the conventional one. In the bottom graph, we can observe that none of the two control structures is able to meet the ethane recovery specification.

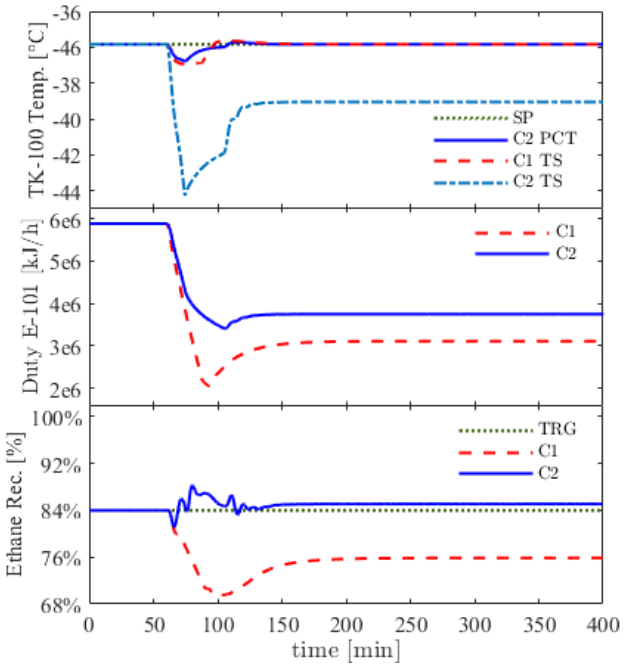

Figure 2: Responses obtained by a decrease of 10% in the feed nominal value of 4980 kmol/h for the ethane recovery.

The new steady state values for the ethane recovery are 76% with the conventional control, and 85% by controlling the PCT with the C2 control. The ethane recovery profile obtained by means of PCT control shows a lower initial deviation and a lower offset at the new steady states, demonstrating that the concentration upsets due to the feed variations are reduced. In the middle graph, the behaviour of the manipulated variable is reported. Here, the control action obtained with the proposed C2 control shows a lower variability.

4.2 Methane bottom impurity level under feed disturbances

The action of the proposed C3 (cascade control while controlling the PCT variable in the separator) and conventional C1 (direct temperature control) control schemes to limit the methane impurity concentration have been evaluated by comparing the profiles obtained in response to feed disturbances. The results are depicted in Figure 3.

By looking at the tray temperature profiles depicted in the top graph of Figure 3, it can be noticed that the C3 control structure has a faster response (it reaches the setpoint value of -90°C after around 98 min while the C1 control after around 190 min) and a lower initial variation. Observing the methane concentration profiles, in the bottom graph, it is noted that the conventional configuration cannot bring the methane impurity back to the target value of 1 mol%, even if the controller brings the controlled variable to the setpoint value. On the contrary,

the value reached with the cascade control is very close to the target. The methane concentration achieves a new value of 2% with the C1 control and a value of 1.03% with the C3 control, which also exhibits a lower initial deviation.

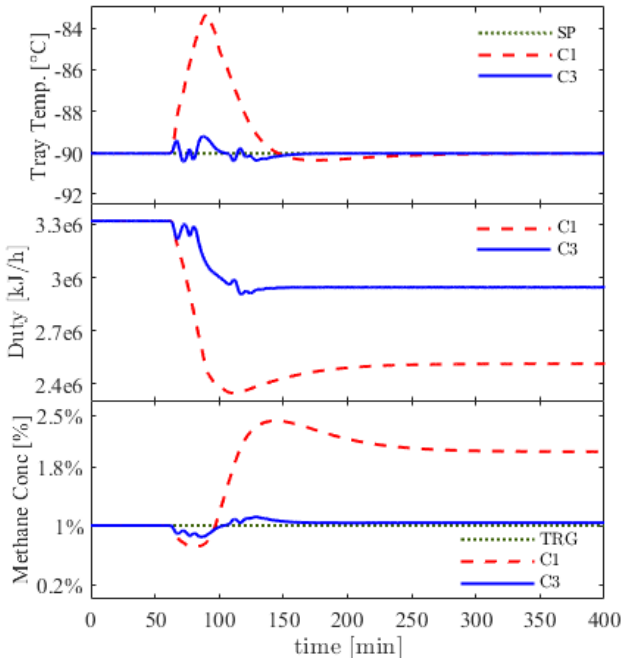

Figure 3 : Responses obtained by a decrease of 10% on the feed nominal value of 4980 kmol/h for the methane impurity level.

Under inlet flow disturbances, as displayed on Figure 2 and 3, the use of the PCT as a controlled variable has a positive influence not only on the ethane recovery but also on the methane impurity concentration. Overall, the offset values of ethane recovery and methane concentration are drastically reduced. With the pressure compensator, we can maintain almost constant the inlet column compositions, leading to an improvement of the control performances. The responses obtained by utilizing the cascade control are generally faster than those using direct temperature control on the plate. This is due to the implementation of the ratio controller. As the dynamics of the flows in the column are faster than those of temperature and concentration, a more aggressive control is attained. Thus, a cascade control combined with pressure compensator can reduce the deviation from the target.

4.3 Cascade control under composition disturbances

To consider the ability to meet the specifications of ethane recovery and methane impurity level in the presence of inlet composition disturbances, the conventional C1 and the proposed C3 control schemes are compared under variations of ethane inlet concentration. The profiles of ethane recovery and methane concentration obtained with the conventional control of the separator and column tray temperatures (C1) and with the proposed PCT control and column tray temperature controller in cascade with the ratio controller (C3) are shown in Figure 4.

Observing the profiles displayed in Figure 4, we can notice that none of the two control strategies can reach the selected ethane recovery and methane impurity level values. In the top graph, looking at the ethane recovery profiles, it can be noted that improvement is not registered in the case of ethane concentration upsets. Nevertheless, the distance between the value obtained with the PCT control is not excessively large, with a value of 1.8%, and a final offset of 6.6%. In the middle graph, the methane concentration offset at the new steady-state obtained with the C3 control is higher than the one obtained by the C1 control. A value of 0.70% is obtained with the cascade control while a value of 0.82% is achieved by the column tray direct temperature control.

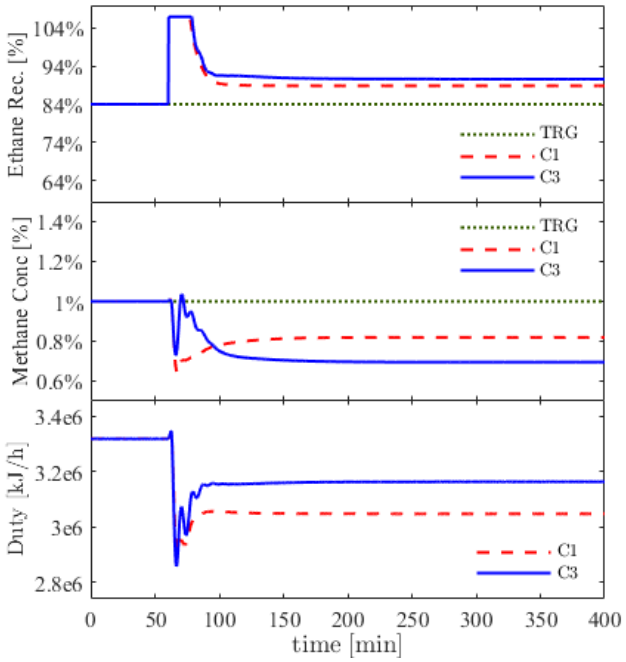

Figure 4: Responses obtained by a decrease of 40% in the ethane inlet concentration nominal value of 0.03 for ethane recovery and methane composition with C1 and C3 control.

The cascade control results under composition disturbances can be explained by considering that the composition dependent coefficient C in (1) has been evaluated at nominal feed compositions. When changing inlet composition with respect to the nominal one, the compensated temperature could not be properly estimated.

4.4 Hybrid cascade control under composition disturbances

Exploring the possibility of improving the profiles obtained for the methane concentration by the column tray temperature control in cascade with the ratio controller, a hybrid cascade (C4) control is proposed by introducing discrete methane measurements. The comparisons between the conventional control in cascade with a methane composition controller (C5 control) and the proposed C4 control for an increase of 40% in the inlet ethane concentration are shown in Figure 5.

Observing the top graph in Figure 5, it can be noted that the ethane recovery profile exhibits the same behaviour in both control structures. By considering the methane impurity

profiles reported in the middle graph, it is possible to observe that the C5 control depicts a higher initial variation and a higher speed of convergence, arriving at the steady-state around 30 min earlier. Both configurations achieve the methane control target at expense of speed of response. The higher response speed achieved with C5 control is probably due to the initial error of the composition controller. Since the conventional initial error is higher than the proposed cascade control error, this leads to a more aggressive initial control action. Besides, by looking at the methane concentration offsets depicted in Figure 4, it is possible to observe that the conventional control initially had to overcome a lower offset, leading to a faster response.

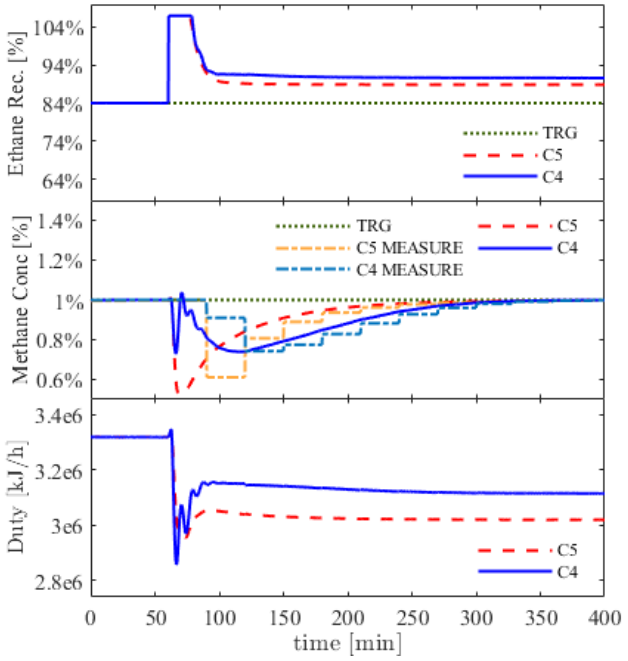

Figure 5: Responses obtained by a decrease of 40% on the ethane inlet concentration nominal value of 0.03 for ethane recovery and methane composition with the C5 control and proposed C4 control.

5. CONCLUSIONS

In the presence of input flow rate and composition variations, various control structures were compared with the main objective of achieving a methane composition of 1 mol% while maintaining an ethane recovery of 84% in a GSP recovery unit. In the presence of variations in the inlet flow rate, it was shown that the use of PCT in the separator as a controlled variable reduced the collateral composition disturbance in the column feed. Due to the implementation of the PCT control, the compositions coming out of the separator and then entering the column are kept almost constant. From the comparison of the indirect methane impurity level control obtained with the cascade control and the conventional direct temperature control in presence of inlet flow rate disturbances, it was possible to assert that the cascade configuration had the best control performance. This controller had the fastest response and maximum speed of convergence for controlling the temperature and composition of the methane in the bottom column stream. Under inlet composition disturbances, the implementation of the in-line delayed methane concentration controller in the cascade control structure, eliminated the steady-state offsets at the cost of a slower response while obtaining a better performance on the manipulated variable.

REFERENCES

Brambilla, A. (2014). Distillation Control and Optimization. McGraw-Hill Education, New York, NY

Campbell, R.E., Wilkinson, J.D. (1981). Hydrocarbon gas processing. Patent 4278457, USA.

Chebbi, R., Al-Amoodi, N.S., Abdel Jabbar, N.M., Husseini, G.A., Al Mazroui, K.A. (2010). Optimum ethane recovery in conventional turboexpander process. *Chem. Eng. Res. Des.* 88, pp. 779–787.

Chebeir, J., Salas, S.D., Romagnoli, J.A. (2019). Operability assessment on alternative natural gas liquids recovery schemes. *J. Nat. Gas Sci. Eng.* 71, pp. 102974.

Hori, E.S., Skogestad, S. (2007). Selection of control structure and temperature location for two-product distillation columns. *Chem. Eng. Res. Des.* 85, pp. 293–306.

Kherbeck, L., Chebbi, R. (2015). Optimizing ethane recovery in turboexpander processes. *J. Ind. Eng. Chem.* 21, pp. 292–297.

Kidnay, A.J., Parrish, W.R., McCartney, D.G. (2011). Fundamentals of Natural Gas Processing. CRC Press, Boca Raton, FL.

Luyben, W.L. (2013). NGL demethanizer control. *Ind. Eng. Chem. Res.* 52, pp. 11626–11638.

Manning, F., Thompson, R. (1991). Oilfield Processing: Natural Gas. PennWell Publishing Co. Tulsa, OK .

Mokhatab, S., Poe, W.A., Mak, J.Y. (2015). Handbook of Natural Gas Transmission and Processing: Principles and Practices. Gulf Professional Publishing, New York, NY.

Ogunnaike, B.A., Ray, W.H. (1994). Process dynamics, modeling, and control. Oxford University Press, New York, NY.

Park, J.H., Khan, M.S., Andika, R., Getu, M., Bahadori, A., Lee, M. (2015). Techno-economic evaluation of a novel NGL recovery scheme with nine patented schemes for offshore applications. *J. Nat. Gas Sci. Eng.* 27, pp. 2–17.

Pitman, R., Hudson, H., Wilkinson, J., Cuellar, K. (1998). Next generation processes for NGL/LPG recovery. *Gas Processors Association*. Dallas, TX.

Porru, M., Alvarez, J., Baratti, R. (2013). Composition estimator design for industrial multicomponent distillation column. *Chemical engineering transactions* 32, pp. 1975–1980.

Shinskey, F.G. (1996). Process Control Systems: Application, Design, and Tuning, fourth ed. ed. Mcgraw-hill Professional, New York, NY.

Wilkinson, J.D., Hudson, H.M. (1992). Improved NGL Recovery Designs: Maximise Operating Flexibility and Product Recoveries. Tulsa, OK.