PREDICTIVE CONTROL DESIGN IN DISTILLATION COLUMN OF LPG - PLANT ULE

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1. ABSTRACT

PDVSA Gas has facilities in various areas of the Venezuela, dedicated to profitable exploitation of hydrocarbon reserves. Ule fractionation plant is located in western of Venezuela, in Zulia state, near to the east coast of Lake Maracaibo.

This plant is dedicated to the extraction and fractionation of natural gas liquids (NGL) and is formed by fractionation plants: LPG-1, LPG-2 and LPG-3, and extraction plant LPG-5, nowadays the only operating plant is LPG-2 with an average feed flow rate of 16,000 barrels per day (BPD) of NGL arising from Tia Juana 2 and Tia Juana 3 Extraction Plants, and which has a design processing capacity of 46,000 BPD of NGL, yielding: propane, mixed butanes and natural gasoline. Plant LPG-2 is composed of a depropanizer tower, a debutanizer tower, propane condensers, butane condenser, two feed drums, two reflux drums and two reboilers.

The basic control in the LPG-2 depropanizer tower is done through feedback controllers to maintain some of the variables in stable condition, such as, temperature, pressure and reflux rate which are fundamentals in order to obtain a determined product specification. This control strategy works properly once the steady state is reached; but is ineffective to respond to disturbances or variations that may occur unexpectedly, producing temperature levels that may turn the product out of specification in a given time period. Therefore, a predictive control strategy is proposed for the control of the depropanizer tower.

This research consists in the design of a predictive control in a LPG distillation tower, based on real data to generate mathematical models of the process, then the data was pre-processed and validated in order to have models that more closely resembles reality and expected behavior of the variables that interact in the LPG distillation process, then were developed control laws and predictive compensators for the mathemathical models and, finally, the verification of the control strategy. The methodology used underlies the proposed by Lewis and Yang (1999) to design control systems. This methodology consists of four phases: Phase I modeling or system identification, Phase II development of the control strategy, Phase III evaluation of the performance, Phase IV construction and verification. Each of the first three phases were successfully developed to obtain the expected result: the design of a predictive control strategy, feed-forward type, whose function is to anticipate disturbances that occur in the process where many variables are involved and strongly coupled which produces competition between control loops operating in the column. In this way we have obtained a dynamic model of a feed-forward compensator that inhibits the effect of the feed flow rate disturbances, in the top and bottom temperatures, having in consideration the behavior of the distillation column of LPG.

2. TABLE OF CONTENT

3. BODY OF PAPER

3.1. PROCESS DESCRIPTION AND CONTROL PHILOSOPHY

The LPG fractionation plant ULE of PDVSA Gas, located in the Tia Juana city, municipality Simon Bolívar, Zulia state, Venezuela, is responsible for fractionation of natural gas liquids (NGL) from two extraction plants located in Maracaibo Lake: Tia Juana 2 and Tia Juana 3.

In LPG-2 fractionation plant, the NGL feed, arising from Tia Juana 2 and Tia Juana 3 Extraction Plants, goes through a flow control valve towards the depropanizer column D8-504, where is fractionated into propane (top product) and a mixture of butane and heavier components (bottom product), as shown in Figure 1.
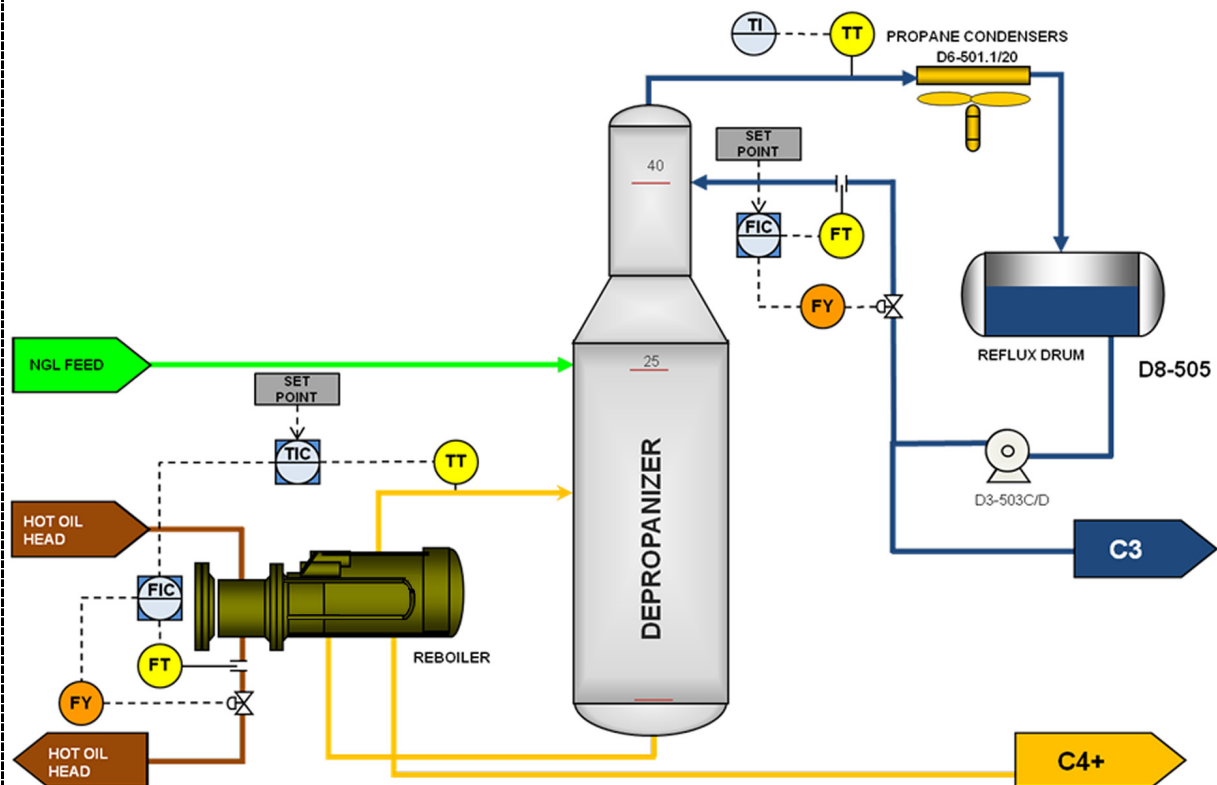

Figure 1: Depropanizer Tower D8-504

Vapors from the top (propane) are sent to the condensers D6-501 and D6-503, connected in parallel. In them, the condensation temperature and the production of liquids is regulated with start and stop of their respective fans, which are operated manually, giving the stop signal from the Distributed Control System (DCS), then the operator must switch them on locally.

Condensation of both coolers is discharged to the reflux drum D8-505. The operating pressure in the reflux drum D8-505 is maintained by a pressure control valve through the deviation of hot gas vapors from the tower top to the drum. Not condensed vapors in the reflux drum are discharged to the flow station TJ-16 or the vent through a pressure control valve.

Reflux pumps D3-503 C/D (one in operation and one spare) are feed from the drum D8-505, returning a portion of the flow through a flow control valve to the depropanizer column D8-504, as reflux. The operation consists in the regulation of the internal temperature in the distillation column and

the dragging of butanes and heavier (C4+) to the bottom again, where the difference in boiling points between the various components of the mixture allows lighter products to flow towards the enrichment zone (top of the tower) and the heavier products go to the depletion region (bottom of the tower) optimizing the extraction of propane.

The other part of propane is controlled by a control level which exerts an action on a level control valve which allows flow of remaining propane in the D8-505 to storage.

The liquid in the bottom of the column D8-504 is sent to the reboiler D2-504, where it is heated through a heat exchanger with hot oil (reboiler), the temperature is controlled by a cascade PID loop, where the temperature master control acts on the secondary flow controller, which regulates the passage of hot oil to the reboiler, according to the temperature requirements set by the operator on the master controller.

Vapors produced in the heat exchanger are returned to the depropanizer column D8-504 and the reboiler effluent liquid fraction is sent as feed to the debutanizer column D8-506, by a level control valve.

The flow from the bottom of the reboiler (composed by butanes and heavier products), which comes from the depropanizer column, flows as the feed to the debutanizer column (D8-506).

3.2. PROBLEM APPROACH

The basic control in LPG-2 depropanizer tower, is performed by controllers with feedback, this type of controller always tries to keep constant some of the variables such as temperature, pressure and reflux rate, which are essential in designing the column for achieving a particular specification. The system works properly when it reaches steady state, but it is inefficient to absorb the variations that may occur in any process, including changes in the composition or the feed rate of the column.

Generally a temperature control is placed in the stripping area to adjust the reflux rate to the reboiler and thus the bottom product composition. The top product quality is usually obtained as a function of reflux rate, either directly or modified by other variables. It is clear that when there is a variation in the quality or quantity of feed to the column, there will be a variation in the compositions of the top and bottom products until the rate of reflux and heating fluid are adjusted.

This operation is usually done manually by adjusting the set-point of the controllers involved at the discretion of the operator when visualizing the variables involved in the process, resulting in a process that tends to errors, which leads to operational problems. The operator cannot reasonably accommodate all the variables that need to be monitor and reset them every minute. The result is that the process will have its operating point to a certain distance from the natural constraints of the distillation process in order to give the operator time to recognize and respond to the input disturbances to the process, this can take time, causing loss of energy, out of specification product and difficulties to the operator when trying to obtain desired values of temperature, pressure and reflux in the distillation column of LPG.

3.3. OBJECTIVE OF THE RESEARCH

To design the Predictive Control for the distillation column installed in LPG-2 plant located in Fractionation Plant Ule, with the aim of improving product quality and reducing unnecessary use of energy in the process of propane fractionation. In order to achieve this objective is necessary the completion of the following tasks:

- To diagnose the current functioning of the LPG-2 depropanizer column.

- To obtain the transfer functions of the LPG-2 depropanizer column.

- To determine the predictive controller design.

- To evaluate through simulation the proper operation of the proposed design for the depropanizer column.

3.4. ASSUMPTIONS AND PREMISES

The bases and assumptions used in the study are:

- The temperature operating condition of the top and the bottom of the distillation column are 127 °F and 270 °F respectively.

- The process data was obtained in manual control (open-loop control).

- The perturbations performed in the control valves were consecutive unit steps.

- The dynamic state data was obtained from the Plant Information (PI) Process Book server connected to the DCS via OPC.

- The sample time, for the recording of the process variables, was one second.

- The transfer functions were obtained through the system identification toolbox in the MATLAB Software.

3.5. RESULTS

Analysis of open-loop results in the Distributed Control System (DCS).

When disturbances were made in the tower with open-loop controllers, it was obtained the data for the further design of the mathematical models. The following is the description of the operational procedure to obtain the process data.

Consecutive steps in the feeding flow control valve were performed starting from 18,000 BPD to 23,000 BPD approximately, and thus, the behavior of the top and bottom temperatures was recorded as shown below:
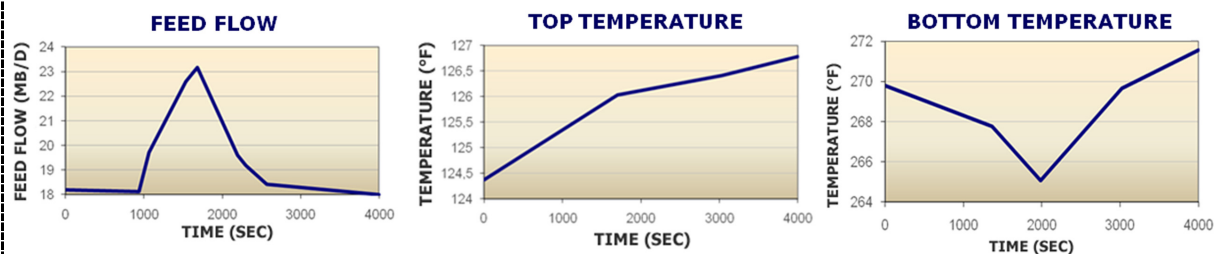

Figure 2: Incidence of Feed Flow Rate in Top and Bottom Temperatures

Then, a perturbation was made in the reflux flow control valve to record top temperature behavior. In the same way for the bottom of the distillation column; a perturbation was made in the hot oil flow control valve to record bottom temperature behavior.
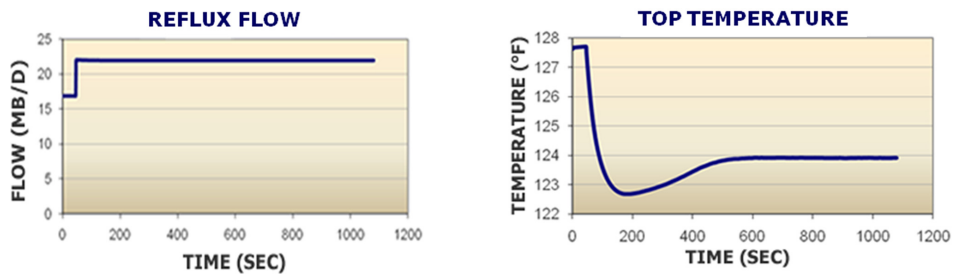

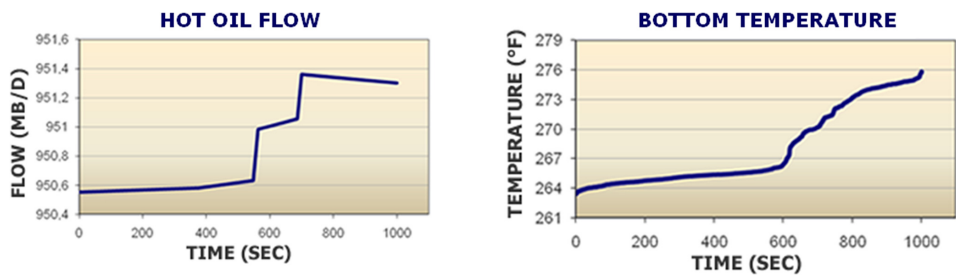Figure 3: Incidence of Reflux Rate in Top Temperature

Figure 4: Incidence of Hot Oil Flow Rate in Bottom Temperature

Obtaining process transfer functions

With the experimental data obtained by making perturbations as above, using the MATLAB as a tool for calculation, it was determined a mathematical model provided by the system identification toolbox. When testing with different types of algorithms, it was determined that the parametric models were most appropriate. The transfer functions were obtained to assemble the desired predictive control law.

The following figures show the transfer functions obtained and the block diagram for the predictive control strategy proposed:
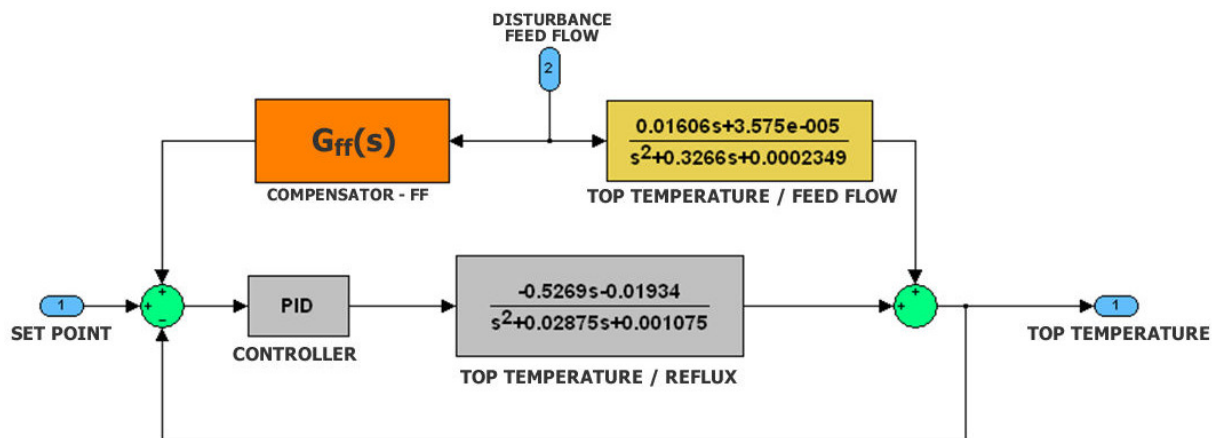

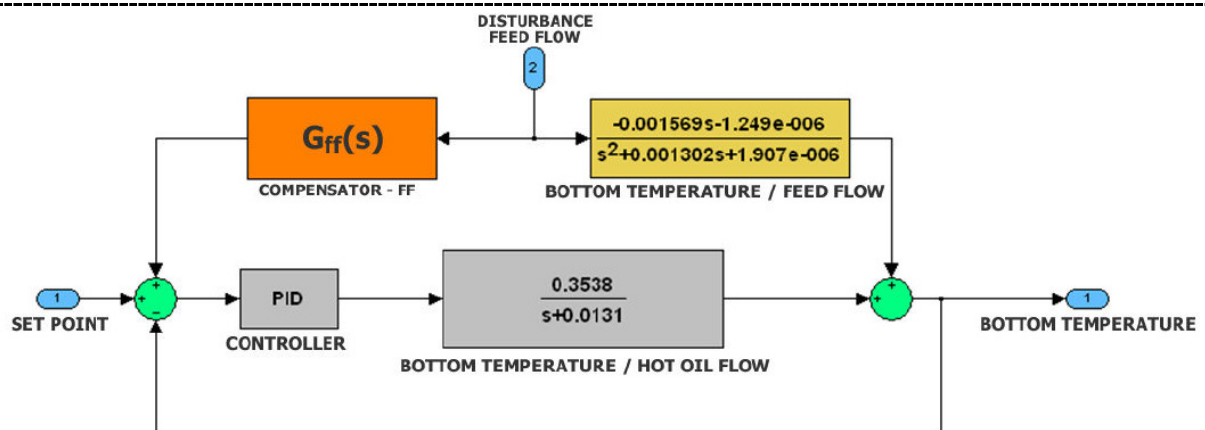Figure 5: Predictive Control Strategy for the Top Temperature

Figure 6: Predictive Control Strategy for the Bottom Temperature

Once obtained the transfer functions, proper analysis (root locus, bode diagram) were performed to each, which would allow to identify the design of the controller by satisfying the features needed for the process, such as, settling time and maximum percent overshot.

Tuning the controller

To tune the two PID controllers was used the sisotool program in the MATLAB Software by the locating of the closed-loop poles in the desired position depending on the needs of the process and thus, to get the transfer functions of the controllers, that stabilize the system at the shortest time possible. The application of the sisotool utility is shown below, where the poles and zeros thrown by the equations were moved and recalculated to find the gain and the dynamic parameters of the controller by selecting most suitable values, in the other hand, the frequency response allows to determine the gain margin for the controller that turns close-loop system unstable.
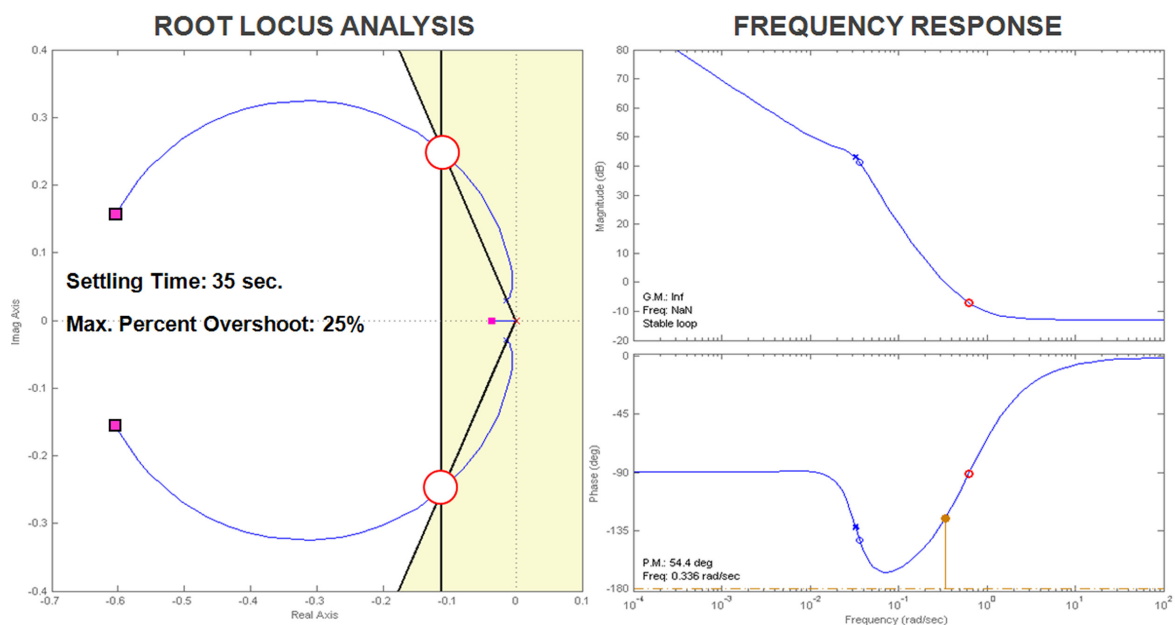

Figure 7: Root locus analysis and frequency response for the Top Temperature PID controller
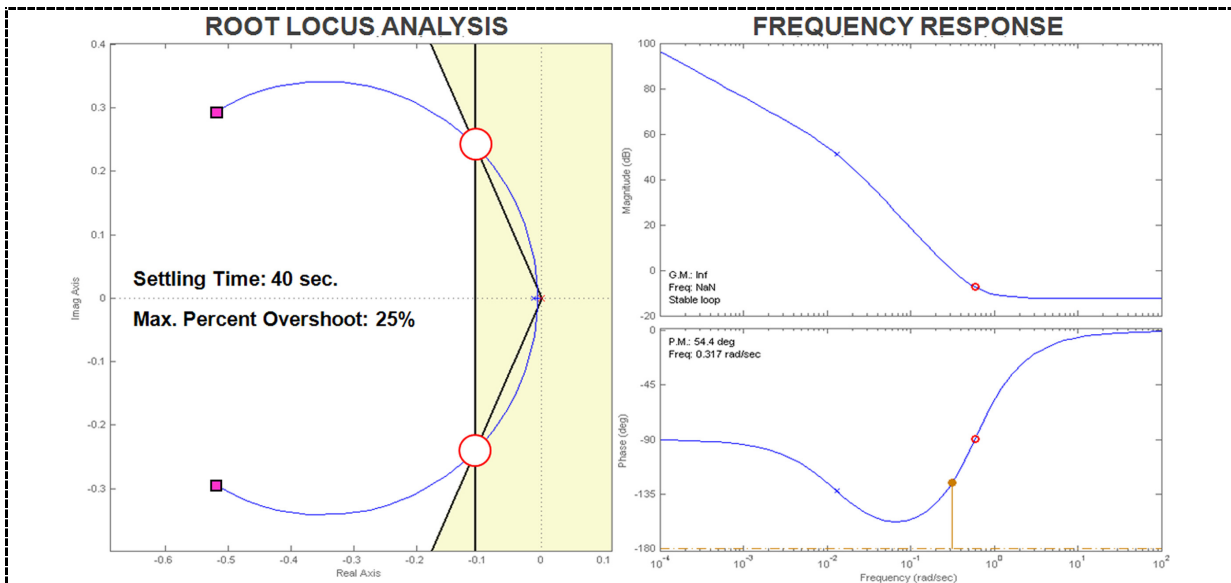
Figure 8: Root locus analysis and frequency response for the Bottom Temperature PID controller

The transfer functions, obtained for the PID controllers are shown in the following figures:

$$C(s) = -0.5201 - \frac{0.16776}{S} - 0.4295S$$

Figure 9: Top Temperature PID controller

$$C(s) = 0.7049 + \frac{0.2431}{S} + 0.7025S$$

Figure 10: Bottom Temperature PID controller

Obtaining the Feed-forward compensators

The calculation of the compensators, for each one of the controllers, was performed with the expression depicted in the figure below, where G(s) is the perturbation transfer function, C(s) is the controller and P(s) is the process transfer function. Afterwards, is shown the transfer functions calculated for each control loop.

$$G_{FF} = \frac{G(s)}{C(s)P(s)}$$

Figure 11: Expression for calculation of Feed-forward compensators

$$G_{FF(s)} = \frac{-0.01606S^4 - 0.000497475S^3 - 1.8292315 * 10^{-5}S^2 - 3.843125 * 10^{-8}S}{0.2263S^5 + 0.3562^4 + 0.1907S^3 + 0.03546S^2 + 0.001083S + 7.621 * 10^{-7}}$$

Figure 12: Feed-forward compensator for Top Temperature control loop

$$G_{FF(s)} = \frac{0.001569S^3 + 2.18 * 10^{-5}S^2 + 1.636 * 10^{-8}S}{0.2485S^4 + 0.2497S^3 + 0.08633S^2 + 0.0001125S + 1.67 * 10^{-7}}$$

Figure 13: Feed-forward compensator for Bottom Temperature control loop

Evaluation of Compensators

The following figure shows the behavior of top and bottom temperatures with and without compensation. It depicts how the compensated loop maintains the process inside the operational parameters:
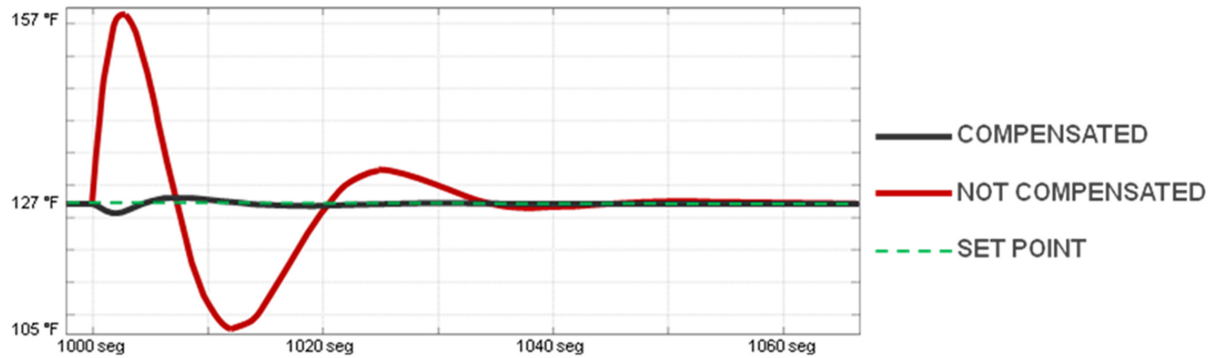
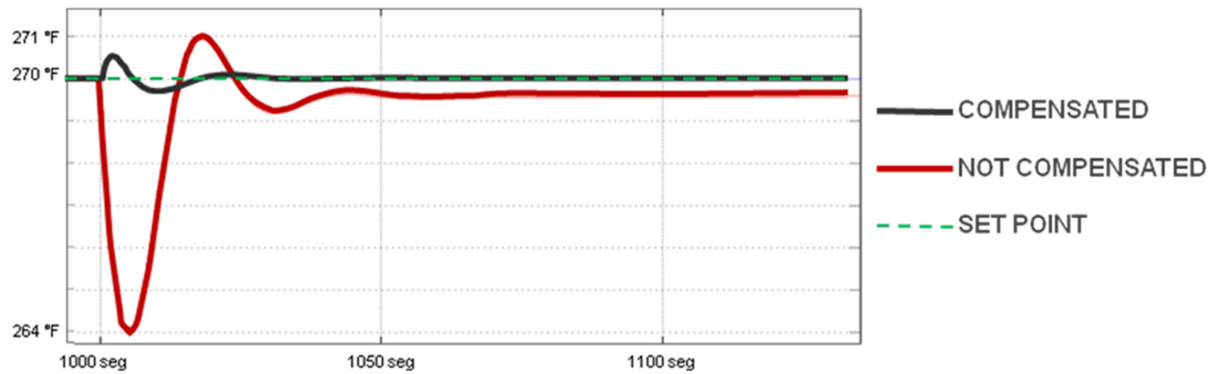Figure 14: Closed-loop response for Top Temperature control loop

Figure 15: Closed-loop response for Bottom Temperature control loop

Without compensator the system stabilizes normally in 36 seconds for the top and 40 seconds for the bottom.

With the compensator with a perturbation of feed the control of the top stabilizes normally in 20 seconds and shows a large decrease in the amplitude of the signal while avoiding sudden changes in the output of the controller.

With the compensator with a perturbation of feed the control of the bottom stabilizes normally in 7 seconds and immediately corrects this disturbance in the feed.

The response time of the signal is improved by 45.5% in the top temperature and by 82.5% in the bottom temperature.

The Feed-Forward Control designed acts on the desired value of the temperature control from the top and bottom of the tower, according to the flow of the feed stream and its corresponding reflux and hot oil flows. This ensures that products fulfill the quality plan to assure economic and energy savings and increase the system reliability and thus reduce human errors associated with the handling of operational parameters within the distributed control system.

3.6. CONCLUSIONS

The process of analysis presents a very particular response since all variables affect each other and depend on each other, the control system for such processes must be robust because they are industrial processes that have quite interaction between variables to control, so they cannot be subject to human errors that could alter the final product quality and hence the importance of this work, a new control strategy was designed for the plant under study, it became clear that the feed flow had a direct relation on allowing temperatures to generate a control strategy on top and bottom temperatures of the column, by controlling the flow of hot oil and reflux in the tower, assuming the feeding stream as a disturbance.

3.7. RECOMMENDATIONS

The implementation of an (advanced) control strategy for the handling of LPG distillation column is recommended, this process may be susceptible to both human error or disturbances that may occur in feedback control strategies, try to optimize the process by using alternative control schemes such as the Feed-forward, implement strategies to control the pitch of the existing PID loops in plant, thereby allowing the stabilization of the process, minimizing the response time to ensure the quality and reliability of the production.

4. REFERENCES

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5. LIST OF FIGURES