OPTIMISATION OF NATURAL GAS PLANT - GAINS IN PROFITABILITY, STABILITY AND ENERGY EFFICIENCY

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Background

Due to the crescent demand for more efficient and sustainable processes, it becomes increasingly important to use new methodologies and tools to diagnose and optimize Natural Gas Processing Units (NGPUs) in order to improve operation and control performances. Process plants are designed based on many assumptions, like the availability of feed, feed properties, environmental conditions and market issues as product prices, quality specifications and the required production rate. However, these parameters can change dramatically, which usually leads to under-optimal operation if the original design or operational parameters are kept constant on the new situation. Therefore, it is of utmost importance to apply new control, optimisation and operating strategies, in order to keep NGPUs at their best operating points.

Advanced process control (APC) and optimisation systems are an industrial reality for oil & gas and petrochemical plants [Qin and Badgwell, 2003] [Campos et al., 2009[a]]. These systems are able to provide improved stability and safety, constraint management and higher profitability. Petrobras has been investing in the development of APC's and their related tools and systems for several years, during which they were implemented on many of its refinery process units [Zanin and Moro, 2004]. However, the application of these technologies to NGPUs is relatively recent in Petrobras [Campos et al., 2007] [Besch et al., 2009], despite of the significant potential of economical earnings.

In complex processes, a change in one variable may cause many effects. Also, on most situations the number of manipulated variables is smaller than the number of variables to be controlled. Therefore, it can be difficult to control all the desired variables at the same time. As a consequence, the operating point of a process unit without advanced control is usually defined with a great safety margin from any constraints, in order to provide enough time for the operators to recognize and respond to eventual disturbances (see figure 1 – Normal operation without APC). An optimisation system will try to push the unit to operate closer to its constraints, and as a consequence maximize production and energy efficiency, minimize losses, etc.

Therefore, the benefit of APC systems comprising optimisation modules is associated with the operation near of the process constraints, which is due to its ability to predict continuously, based on its model, the effects of disturbances on these constraints or controlled variables and make anticipated control actions that will keep the process always at its best operating point. Because it operates 24 hours a day, it makes the best of opportunities when there are changes in feed composition, flow rate and any other measurable external conditions.

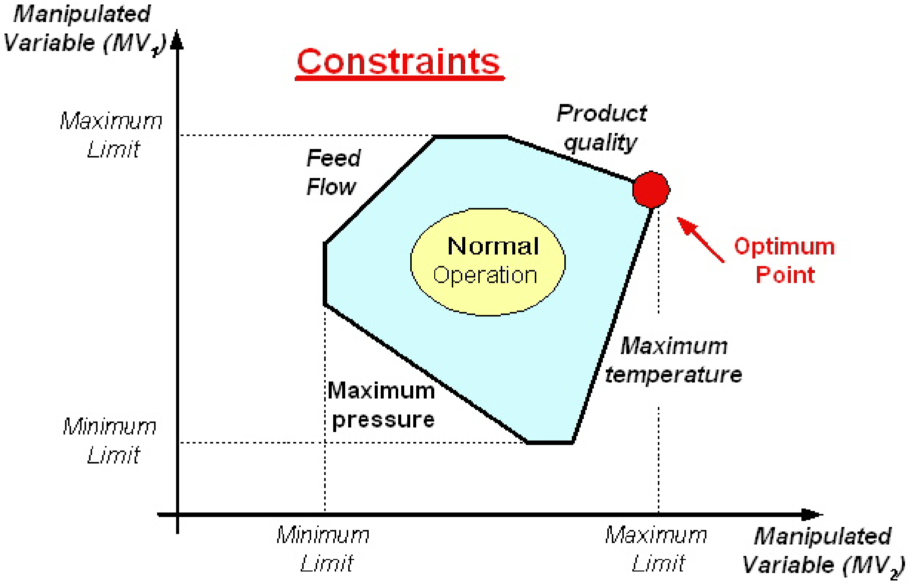


Figure 1 – APC benefits attained with operation closer to constraints.

This article will describe the results of a wide program for the deployment of APC systems including optimisation procedures and the improvement of the existing regulatory control systems in many Petrobras' NGPUs. A new methodology was conceived to maintain these tools, in order to avoid degradation of the obtained benefits related to improvements on stability, profitability, energy efficiency and sustainability.

The APC system discussed here is the Multivariate Predictive Control (MPC) [Camacho e Bordons, 1998] [Maciejowski, 2002] with its optimisation layer. This algorithm uses a dynamic model of the process to compute the optimal future control actions, minimizing the effect of disturbances. It tries to control several process variables simultaneously acting in many manipulated variables. The advanced control algorithm used in the NGPUs is called "CPM" and was developed by Petrobras. The CPM algorithm makes use of a multivariable and dynamic model of the process to increase the control system performance, especially when some of the following characteristics are present:

- Multivariable and coupled regulatory control loops;
- Processes with long dead times, long time constants or non-minimum phase behaviour.
- Processes whose operating point is continuously upset by disturbances.

Optimum operating set points are obtained by the CPM optimisation module, based on economical criteria and taking into account the process constraints.

Aims - Methodologies and Tools to Diagnose and Optimize Operation

An integrated vision of all automation levels is highly required in order to achieve the desired results. These levels are implemented as layers of the automation pyramid in figure 2. It is necessary to have a policy by which some key performance indicators are monitored, in order to diagnose and fix any arising problems, either in the process, instrumentation or control. This policy should involve actions and attention from operators, process engineers and control engineers.

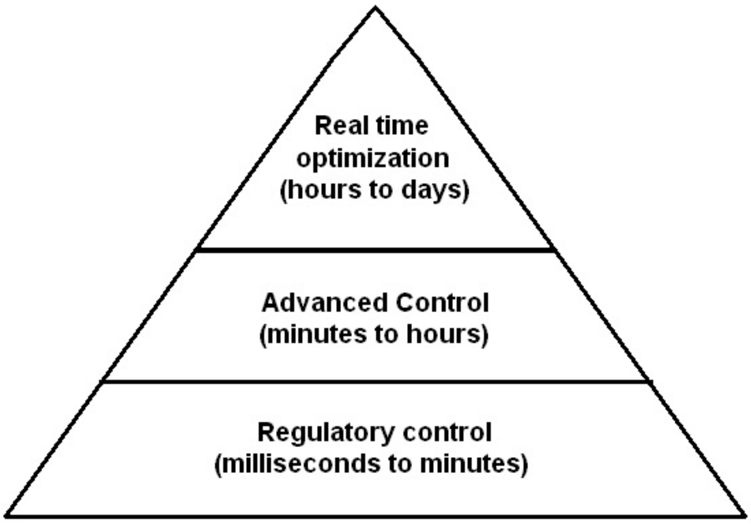


Figure 2 – Automation layers.

Regulatory Control Assessment

An advanced control system won't reach the expected economical benefit if turned off constantly by the operators. One of the main reasons that can cause this is the existence of problems on the pyramid base, mainly constituted by PID controllers [Campos and Teixeira, 2006] [Bequette, 2003] [Ogata, 1982] configured in the digital systems (DCS - Distributed control system or PLC - Programmable logical controllers). In many situations, a significant benefit can already be obtained with the improvement of the regulatory control layer, as will be shown in this paper. Therefore, the instruments, valves and the regulatory control (PIDs) should operate appropriately. Some of the common problems associated with the regulatory control are:

- Instrumentation problems - Valve stiction, miscalibration, measurement resolution, noisy sensors, bad sizing.
- Tuning of the PID controllers – oscillation and/or lack of stability.
- Control strategy – interaction and poor management of degrees of freedom.

An industrial plant usually has hundreds of control loops, and less and less engineers to maintain them. Therefore, the industries need tools to perform automatic analysis and diagnoses [Aströn, 1970][Harris, 1989][Kempf, 2003]. One of the most important features of this tool should be to provide automatic ways to prioritize the actions for each process that might result in a better performance. It should offer a standardized metric to compare different actions in different processes as well, even in different scales such as economical, environmental or safety.

Petrobras and Federal University of Rio Grande do Sul (UFRGS) have developed a tool for regulatory control assessment, the software called "BR-PerfX" [Farenzena et al., 2006] [Farenzena and Trierweiler, 2008]. Its main purpose is to compute some universal key performance indicators that reduce the subjectivity in the analysis and help engineers in their assessments and decisions about problems affecting the regulatory control. Figure 3 shows BR-PerfX interface.

One of the main problems found in the instrumentation is related to control valves. This equipment might present hysteresis, dead band and stiction, frequent causes of oscillations and increased variability in controlled variables [Garcia, 2008] [Munaro et al., 2008]. Figure 4 shows an example of control valves with and without dead band and stiction.

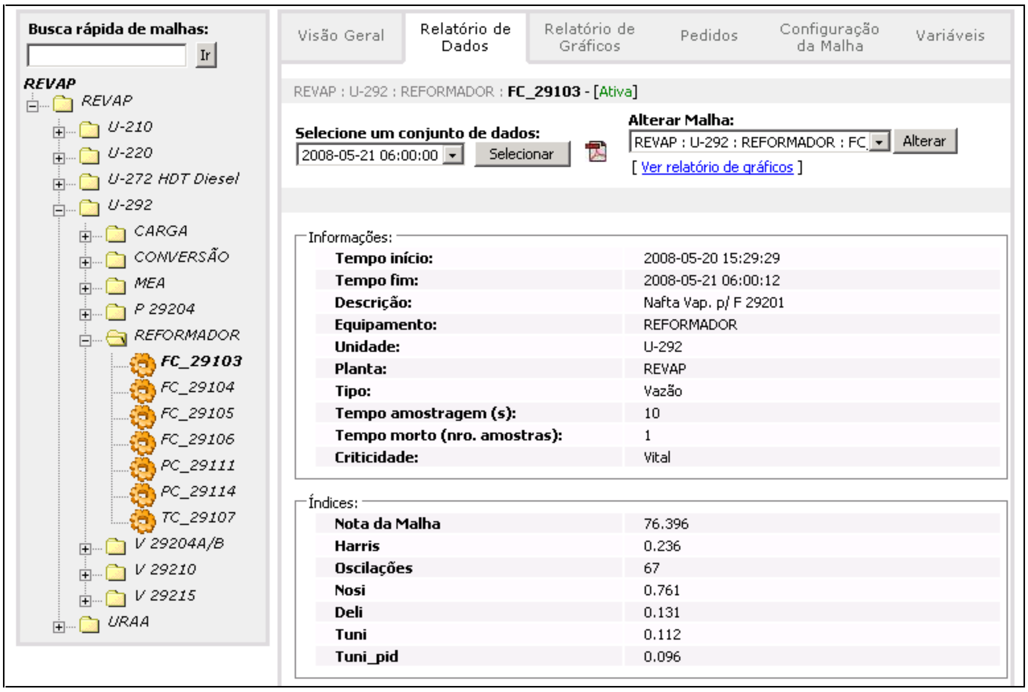


Figure 3 – BR-PerfX interface.

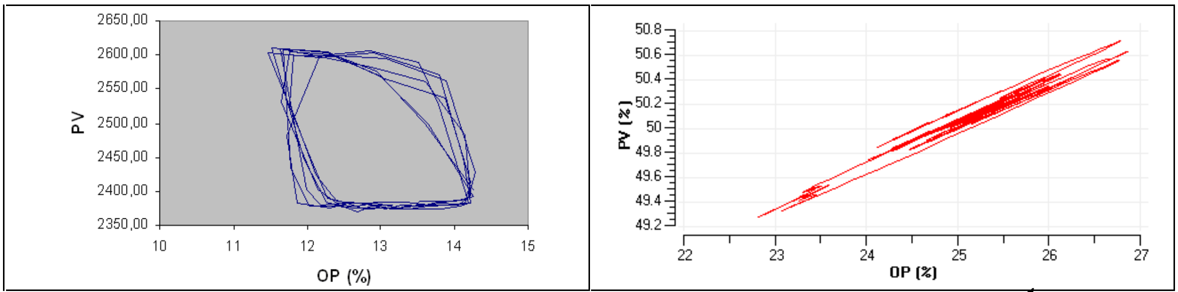


Figure 4 – example of control valves with and without dead band[1].

Figure 5 shows an example of oscillations due to dead band and stiction in control valves. Usually, when there isn't over sizing, maintenance procedures can solve these problems.

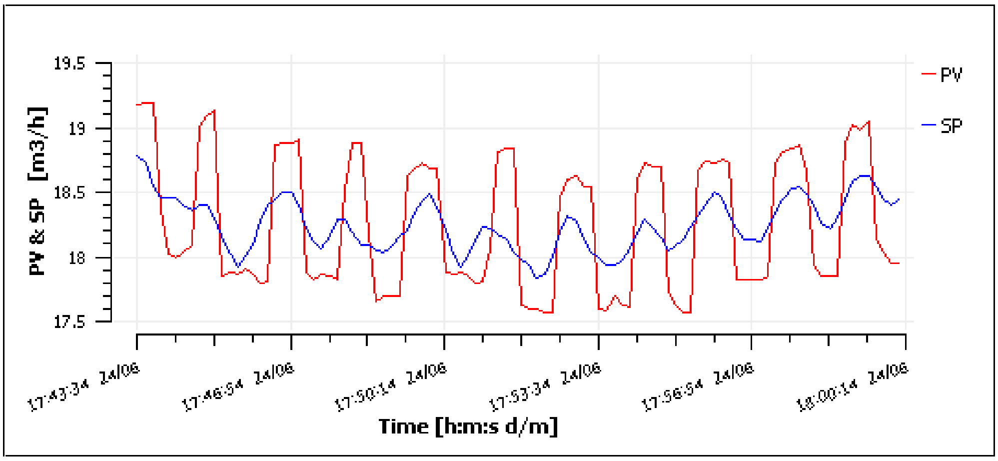


Figure 5 – Example of control oscillations due to dead band and stiction in control valves (PV is the controlled variable and SP the set point).

[1] PV is the controlled variable and OP is the controller output.

Regulatory Control - PID Tuning

If there are tuning problems on PID control loops the software "BR-Tuning" is used. This software was developed in partnership between Petrobras and the Federal University of Campina Grande (UFCG). BR-Tuning [Arruda and Barros, 2003] [Schmidt et al., 2008] [Acioli et al., 2009] implements a group of techniques for open and closed loop identification and proposes new tuning parameters. It communicates directly with the process automation system (DCS or PLC) using the OPC protocol [OPC, 2012]. Figure 6 shows an interface of this software.

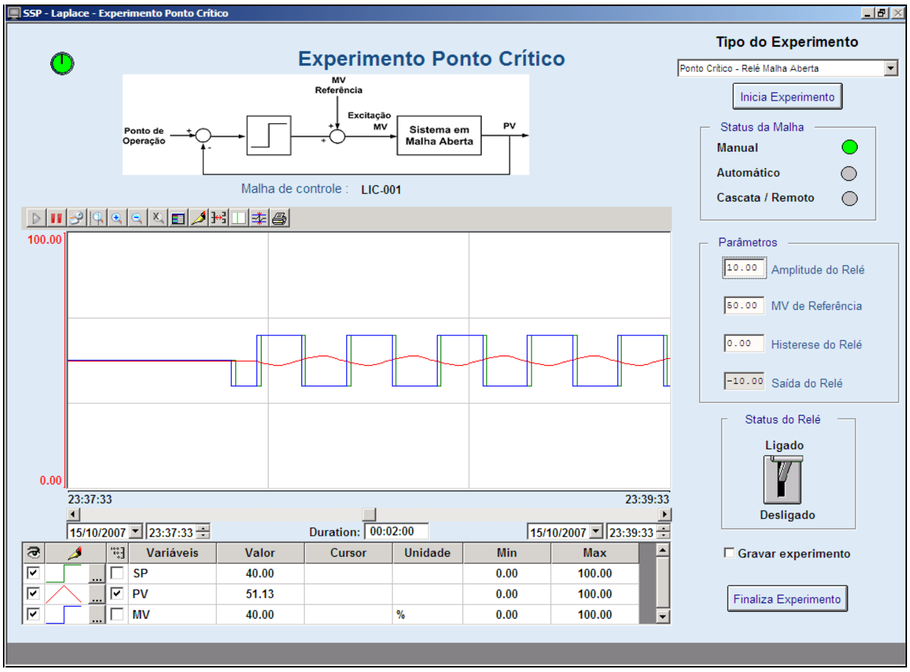


Figure 6 – BR-Tuning interface.

Advanced Control - Multivariable Predictive Control (MPC)

The advanced control systems are positioned in the second pyramid level. Model Predictive Control (MPC) [Cutler e Ramaker, 1980] [Richalet et al., 1978] [Clarke et al., 1987] [Camacho and Bordons, 1998] [Rawlings, 2000] [Maciejowski, 2002] is one of the most used advanced techniques. This algorithm considers the interaction between control loops, and makes use of an optimisation module or procedure. These algorithms are usually implemented in a process computer that communicates with the DCS or PLC using for example OPC protocol [OPC, 2012]. The outputs of this advanced control are usually the set points of the PID controllers of the regulatory control. So if there is a problem in the advanced control layer, the plant operation can go on with the last PID set points in the DCS. The MPC controller determines a set of control actions (Δu) that optimize the following objective function:

$$J = \underline{e}^T \underline{\underline{W}}^T \underline{\underline{W}}\, \underline{e} + \Delta \underline{u}^T \underline{\underline{R}} \Delta \underline{u} \qquad (1)$$

$$\underline{e} = \underline{\hat{y}} - \underline{y}^{TG} \qquad (2)$$

where **e** is a vector of future deviations predicted between the controlled variables (y) and their respective targets or set points ($\underline{y}^{TG}$) along a given future prediction horizon. These targets are defined by the MPC optimisation layer.

This dynamic optimisation is subject to the following constraints:

$\underline{u}_{\min} \leq \underline{u} \leq \underline{u}_{\max}$ (3); $\Delta\underline{u}_{\min} \leq \Delta\underline{u} \leq \Delta\underline{u}_{\max}$ (4); $\underline{y}_{\min} \leq \underline{y}^{C} \leq \underline{y}_{\max}$ (5)

The relationship between the controlled variables **y** and the manipulated variables **u** is defined by the dynamic process model $\underline{\underline{S}}$. The controller uses these models to predict future deviations. If these deviations are written in terms of the manipulated variables, one can obtain:

$$\underline{e} = \underline{\underline{S}} \cdot \begin{bmatrix} \Delta\underline{u} \\ \Delta\underline{D} \end{bmatrix} \quad (6)$$

Where ΔD contains the movements of disturbance variables that might upset the process. The solution of this optimisation problem, given by equations (1) - (6), will provide a trajectory of future control actions (to each manipulated variable) that may lead the process to the defined optimal operating point, while avoiding the eventual effect of the incoming disturbances. However, in order to obtain suitable control actions, a set of tuning parameters are introduced. When applying them, the system solution is given by:

$$\Delta\underline{u} = \left(\underline{\underline{S}}^{T}\underline{\underline{W}}^{T}\underline{\underline{W}}\,\underline{\underline{S}} + \underline{\underline{R}}\right)^{-1}\underline{\underline{S}}^{T}\underline{\underline{W}}^{T}\underline{\underline{W}}\,\underline{e} \quad (7)$$

Matrix ***R*** is composed of the moving suppression factors, tuning parameters that define the intensity or speed of control actions. This matrix defines a penalization on the movements of the manipulated variables. Matrix ***W*** is composed of the equal concern parameters. With their introduction, a prioritization of the controlled variables is established, which provide the MPC controller the ability to “decide” which deviations should receive more attention at each situation.

In practice, the predictive controller computes at each run (typical execution period spans from 30 seconds to 6 minutes) a set of future control actions that minimize the objective function "J" (equation 1), subject to constraints, considering a prediction horizon (N). However, only the first control action is implemented. At the next time "k +1" all the optimisation and control sequence is repeated. Thus, this controller belongs to a class called receding horizon control [Kwon and Han, 2005]. Figure 7 shows a schematic of this predictive controller.

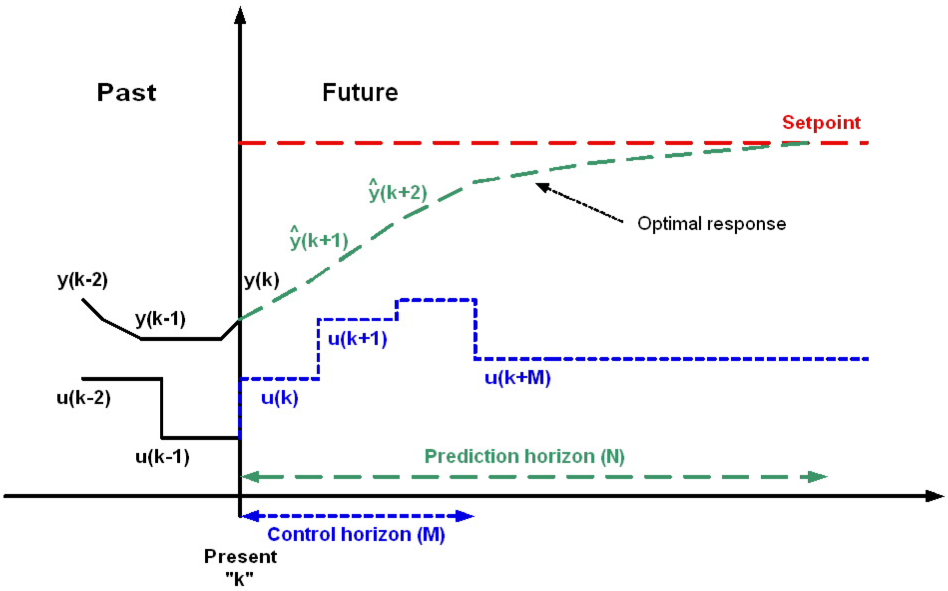


Figure 7 – Predictive controller actions.

Many industrial MPC controllers have an embedded optimisation procedure responsible for defining the optimal operating point to be sought, that takes into account the economic aspects of the process under analysis. The aim of this optimisation layer is usually the minimization of operational cost [Cutler and Perry, 1986] [Jing and Joseph, 1999] [Saffer and Doyle, 2004] [Tatjewski, 2008] [Nikandrov and Swartz, 2009]. This objective function comprises the cost of feeds and utilities (electricity, steam, cooling water and fuel gas), and the profit to be obtained with the product streams:

$$F = \sum_{i}^{nF} \$_i Feed_i + \sum_{l}^{nUtil} \$_l Utility_l - \sum_{k}^{nP} \$_k \mathrm{Pr}oduct_k \qquad (8)$$

Thus, the optimisation problem may be posed as the minimization of function F(u) subject to constraints. Examples of optimisation goals for a Natural Gas Processing Units (NGPUs) are:

- Minimize the loss of C_{3+} (NGL – Natural gas liquids) in the gas.
- Maximize the C_2 content in the NGL (up to a maximum specification limit).
- Increase the gas production or NGL production depending on the market.
- Minimization of operational cost.
- Increasing the operational stability.
- Increasing energy efficiency.
- Minimize gas losses in the flare.

The problem can be simplified by the linearization of the objective function around an operating point:

$$F \cong F(u_0) + \sum_{j=1}^{nU} \frac{\partial F}{\partial u_j}(u_0) \times \Delta u_j^{OT} = F(u_0) + \sum_{j=1}^{nU} P_j \times \Delta u_j^{OT} = F(u_0) + \Delta F\left(\mathbf{\Delta u^{OT}}\right) \qquad (9)$$

Where $\mathbf{\Delta u^{OT}} = \mathbf{u^{OT}} - \mathbf{u_0}$, while $\mathbf{u_0}$ is the vector of values of the manipulated variables at the initial instant and $\mathbf{u^{OT}}$ is the vector of optimal values to be achieved when the steady-state is established. Therefore, $\mathbf{\Delta u^{OT}}$ is the amount of change to be implemented to each manipulated variable in order that the optimal operating point can be reached. Using equation (9), the optimisation problem can be stated as the following linear programming problem:

$$\min_{\mathbf{\Delta u^{OT}}} \Delta F\left(\mathbf{\Delta u^{OT}}\right) \qquad \text{subject to :} \qquad \begin{aligned} & u_j^{\min} \le u_j^{OT} \le u_j^{\max} \quad j = 1,\dots,nU \\ & y_i^{\min} \le y_i^{OT} \le y_i^{\max} \quad i = 1,\dots,nY \\ & y_i^{OT} = y_i^{SS} + \sum_{j=1}^{nU} g_{ij} \Delta u_j^{OT} \end{aligned} \qquad (10)$$

Where **G** is a matrix with the static gains of the process $\mathbf{g_{ij}}$ that define the influence of each manipulated variable at each controlled variable, obtained from the dynamic model. The results of this optimisation problem (the sets of values for $\underline{\mathbf{u}}^{\mathbf{OT}}$ and $\underline{\mathbf{y}}^{\mathbf{OT}}$) are the targets or set points for the advanced control layer.

The relations between the manipulated variables and the objective function (P_j, on equation 9) must be obtained by the combination of prices for feeds, products and utilities and data from the process model. They should be defined carefully, in order to guarantee that consistent optimisation results will be obtained.

Real Time Optimisation (RTO)
The RTO layer is composed of optimization algorithms associated to first-principle, highly accurate models. They are used in association with MPC algorithms when an extra level of accuracy must be used [White, 1997], in order to determine the optimum operational targets that must be pursued by the MPC.

Methods

The proposed methodology for implementing the optimisation system is gradual in order to allow the absorption of this technology by operators and process engineers. There are three basic principles:

- The use of MPC Controllers as tools for control and optimisation requires, in general, changes in the way engineers and operators perform the monitoring and operation of the process unit.
- A successful implementation depends directly on the performance of the process, regulatory control and instrumentation.
- The implementation of MPC and its optimisation layer implies obtaining reliable process and economic models.

Thus, along all phases of the project special attention is paid to the training of the teams involved, with consolidation of concepts related to the process, tools and the controller. So, many hours were used in discussions with operators and engineers. This was crucial to the success of the project, ensuring the effective use of the system and assuring the economic benefit expected. Another relevant aspect was the commitment to analyze the process using simulators to define recommended operating regions of the units.
The projects were organized in the following steps:

- Definition of key performance indicators (KPIs);
- Conceptual design;
- Review of the regulatory control system and instrumentation;
- Process identification;
- Installation, configuration and tuning MPC;
- Training and Assisted Operation;
- Benefit estimates;
- Implementation of operational indicators in real time.

Definition of key performance indicators (KPI)
In this phase, indicators are conceived to measure the process unit situation regarding aspects as variability, profitability and energy efficiency. In the beginning of the project the KPIs are computed for a selected moment, defining a base case. These KPIs are reassessed at some key points of the project and compared, providing an estimate of the attained benefits.

Conceptual design
The conceptual design starts with a thorough analysis of the process, where the scope, economic and performance objectives of the controller are defined. From these objectives, a set of guidelines are proposed, establishing the ways to act on the process in order to

achieve the desired goals. Based on these guidelines, the set of process variables to be used by the APC system are selected, and defined as manipulated, controlled or disturbances. Manipulated variables are those used by the controller to act on the process, usually set points of PID control loops. Controlled variables are operational constraints, in general expressing product specifications, and safety or capacity limits. Disturbance variables represent any measurable external influences to the process unit, whose effect the APC controller is supposed to anticipate and reject. During the conceptual design it is also performed the definition of the hardware architecture (process computer, communication drivers and networks) and other software and hardware issues.

Review of the Regulatory Control System and instrumentation

As stated, the good performance of instrumentation and regulatory control is fundamental to the successful implementation of an advanced control system. This comes from the fact that these systems usually act in the set point PID controllers (Campos and Teixeira, 2006). So if there is a problem in regulatory control, advanced system performance will be compromised. With the multivariable predictive controller, set points associated with manipulated variables are changed frequently. So, regulatory controllers should be tuned slightly overdamped. The performance of each regulatory control loop must be evaluated for set point change.
Some control loops can be found unstable or unsuitable to work with the APC. In these cases, the control loop must be re-designed. This paper will show some examples where new regulatory control strategies provided significant benefits. Some instruments may be operating at full scale, or with inadequate resolution. Thus sometimes recalibration will be required for these instruments.

Process Identification

The goal of this phase is to identify the relationship between the manipulated and the controlled variables of a process, representing this relationship through a set of mathematical equations named the process dynamic model, already mentioned on previous sections. This model will be used throughout the MPC algorithm, on the computation of predictions, determination of the optimal operating point and on the computation of the control actions. The accuracy of this model prediction is crucial to the performance of the MPC controller. The model is obtained from plant tests, consisting of changing the manipulated variables of the controller in a safer and planned manner, and measuring the response of all controlled variables. Typically, several steps are performed using different amplitudes and duration. Process identification of complex processes is still a hard task, where a significant part of the effort on MPC implementation is spent.
The collected data is used along with an identification software as the “VIP” system (1999), a software developed by Petrobras that performs the computation of the dynamic models. Then, some meetings involving the implementation team, process engineers and plant operators are scheduled in order to validate the model, based on theoretical concepts and operational experience. After the necessary corrections are carried out, a consolidated model of the process is obtained. At this step, the economic model must also be determined.
Figure 8 shows an example of a test, where a manipulated variable (in blue) is varied in steps and its effect on a product composition (in yellow) can be measured. A new automatic system named “BR-Step” was devised to perform the plant test. By monitoring the process conditions during the plant test, this system saves time and avoids the loss of data. For instance, a step test will be performed only if certain conditions are verified, such us, the feed flow is in a certain range, or the outputs of PID controllers are not saturated.

The Process Identification phase is an excellent opportunity to expand the available knowledge about the unit, since it requires a behaviour analysis of the main process variables and the interactions between them, along with economic issues.

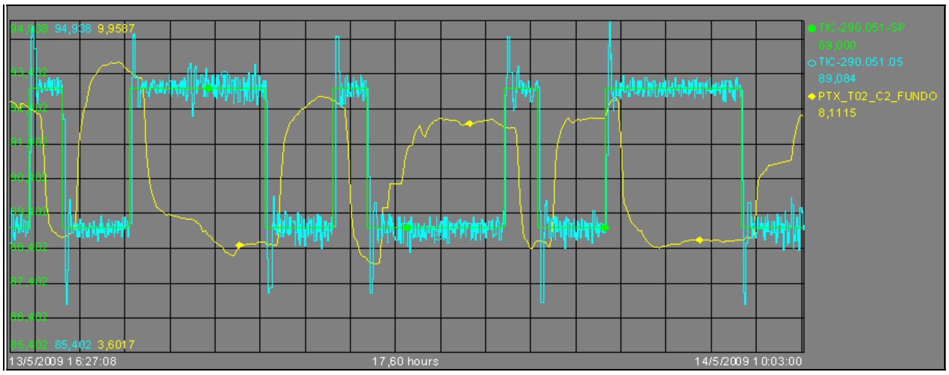


Figure 8 – Example of an identification test.

Figure 9 shows an example of the obtained models (the rows are manipulated variables or disturbances and the columns are the controlled variables).

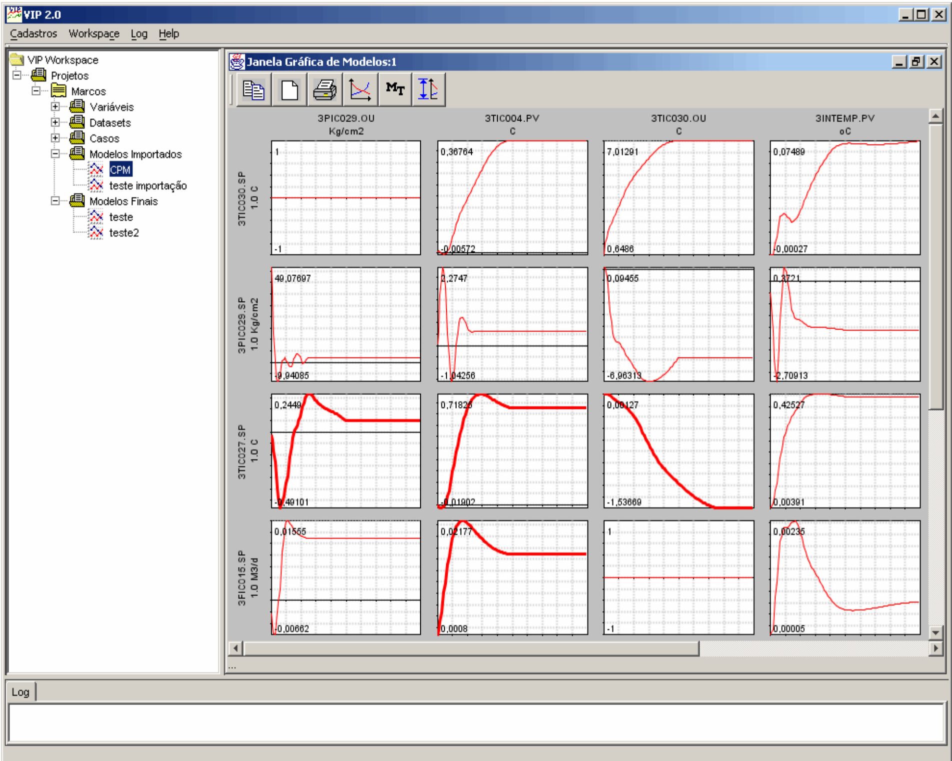


Figure 9 – Dynamic models identified.

Installation, configuration and tuning

The software configuration involves the OPC addressing of process variables, incorporation of the consolidated dynamic model and the creation of operator interfaces in the automation system (DCS - Distributed Control System or SCADA - Supervisory Control and Data Acquisition). Through the interface, the operator can start and stop the advanced control, enable or disable some variables, and change the operating ranges. Figure 10 shows the configuration screen of the controller.

Logic procedures that continuously verify the communication integrity between the APC system and the automation system are also implemented. If there is a failure, some actions are performed to ensure that the regulatory process control continues safely without the APC system.

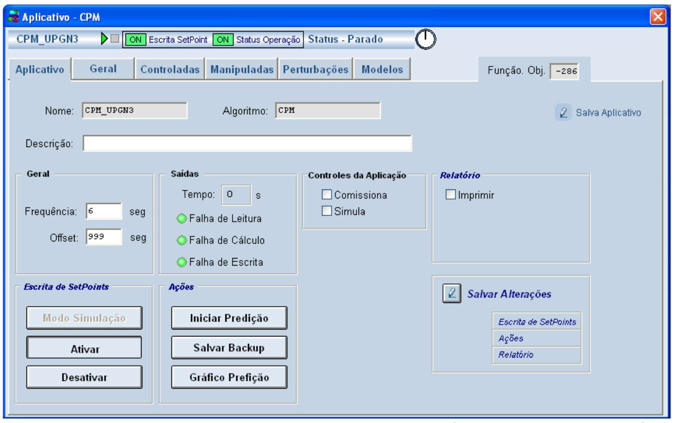


Figure 10 – Predictive controller: configuration interface.

A simplified dynamic simulation of the process is used to obta in an initial set of MPC tuning parameters. Then, the controller is put into operation in open loop, and the consistency of their control actions can be evaluated. With a first set of acceptable tuning parameters, the controller starts operating in closed loop, and some adjusts are done according to the desired performance.

Training

A key aspect for the success of the project is the training of the entire team. It is done continuously, with training sessions realized at the beginning of each phase of the project. The training goal is to provide information about the following phase and define tasks to be assigned to the plant personnel. However, every opportunity is used to consolidate knowledge about the process and about the use of the APC as a new optimization tool. Special attention must be given as well to the project documentation, which shall be used as a reference in the future. Once the APC system is fully implemented, there is an important period of assisted operation where a hands-on training is performed.

Benefit estimates

A careful estimation of benefits is essential in order to give credibility to the finished project, but also to encourage new investments. Both economical and non-economical benefits must be reported. However, the comparison of a process unit performance at different moments can be a challenging task. They must be consistent, in order to avoid either under or over estimates. Therefore, every comparison must take into account the feed flow rate and external aspects that can affect the process efficiency. In the case o NGPUs, any comparison must take into account the natural gas C_{3+} molar composition at different situations, as the process behaviour can change dramatically.

Implementation of operational indicators in real time

Advanced control performance can decay throughout time. There are many causes for this behaviour:

- Changes in the units operational objectives;
- Equipments efficiency losses (fouling);
- Changes in the feed quality;
- Problems in instruments and in the inferences (soft-sensors);
- Lacks of qualified personnel for the controller's maintenance.

If no maintenance work is done, the operators end up turning them off. So it is necessary to implement online operational indicators that measure the performance of the controller and can be used to identify the need for corrective actions. A second point is to have reliable tools to diagnose problems and point out the causes of bad performance, such as unreliable dynamic models, bad controller tuning, inference problems, non-linear behaviour or new process constraints.

Results

The major objectives of NGPUs are to fractionate the Natural Gas on desired products such as NGL (Natural Gas Liquid), LPG (Liquefied Petroleum Gas) and Gasoline while keeping them at the required specification. Some of the results of the application of the proposed methodology to some NGPUs are discussed here.

Example I - Natural Gas Processing Unit in Bahia, Brazil
The first unit where the new methodology was applied is located in Bahia, Brazil. In this unit, NGLs (C_{2+} fraction) are extracted from the Natural Gas by cryogenic expansion with turbo-expansion (see figure 11). It is desirable to maximize the NGLs recovery, while keeping a specification of maximum percentage of C_2.

Figure 11 – Photo of the gas processing plant – Bahia, Brazil.

Example I - Regulatory Control - Assessment of Instrumentation and Control
After an evaluation of regulatory control with the tool "BR-PerfX", it was generated a list of corrective actions, primarily associated with the PID tuning. In this unit, there were no valves with stiction problems, neither valves with oversized.

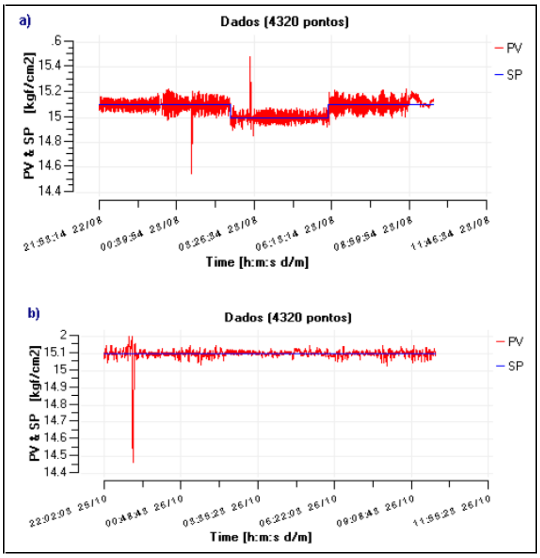


Figure 12 – Improvement of the performance of PIC-4002 after tuning (curve "b").

Figure 12 shows the results for the compressor pressure control loop (PIC-4002). The analysis indicates that the control loop had an aggressive tuning (curve "a"). It was noticed a real improvement of this controller with a reduction in the Harris and in valve travel index. This control loop changed the oscillations period from 170 to around 700 seconds, which are more typical of process disturbances. The reduction of the oscillatory behaviour of this control loop is also evidenced when comparing the graphics of the autocorrelation function, shown in Figure 13.

The pressure control of the deethanizer tower was also tuned. It was operating in manual with a high variability (≈1.5 kgf/cm^2), which is not desired for a distillation column.

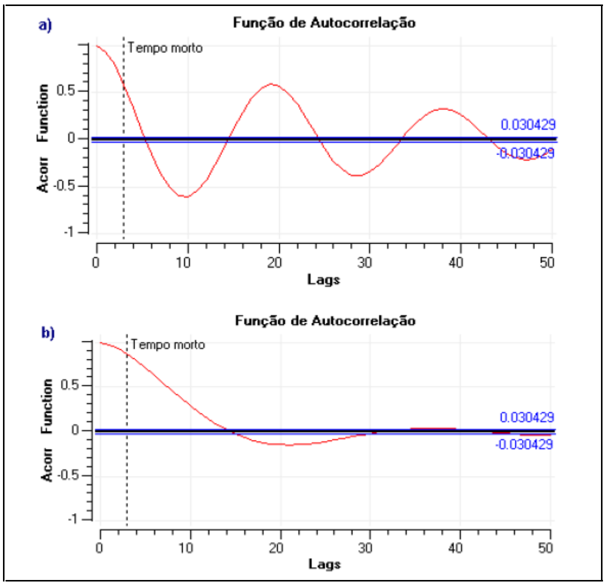


Figure 13 – Performance of the PIC-4002 after tuning (autocorrelation function).

Example I - Regulatory Control - Assessment of control strategy

After the tuning phase, the focus has turned to the analysis of regulatory control strategy. A well-designed control structure must be robust and flexible in order to avoid frequent and repetitive actions of the operators in the plant. Observing this unit, some problems were detected:

- Inappropriate pressure control of the deethaniser;
- Frequent operator's actions to adjust the compression according to the feed flow.

Figure 14 shows a simplified schematic of the structure that was used. The control strategy was designed to keep the deethaniser pressure constant acting on the admission valve of the gas to the turbo-expander. This control is also subject to two overrides: maximum expander suction pressure and maximum turbo-expander speed.

With this control strategy when the unit receives more natural gas, the pressure of the tower tends to rise causing the controller to close the inlet gas valve of the turbo-expander, thus resulting into higher pressure in the pipeline. So, the operator needs to change a "step" in the reciprocating compressor to force the system to admit more gas.

Figure 15 shows a simplified drawing of the proposed strategy. The first change was to control the inlet pressure of the unit manipulating the valve of the turbo-expander with override due to high pressure in the column and high speed in the machine. In this new philosophy, the tower controller is only to avoid high pressure in the tower. The pressure of the deethaniser is now maintained by the suction pressure controller of the reciprocating compressor. Thus, the manipulated variable is the compressor recirculation valve. With this strategy, the plant will operate as follows: when there is an increased in the natural gas production, the pressure between compressor and expander will tend to increase, making this controller to open automatically the inlet valve of the turbo-expander, so processing more gas. The pressure in the deethaniser was much more stable with this new strategy (≈ 0.10 kgf/cm^2 - reduction of 93% of the variability).

It was also implemented an automatic procedure to change the "step" of the sales gas compressor in order to maintain the recirculation valve around 10% opening to minimize energy cost. It should be noted that now the operators need only supervise the control, and they don't need to change the compressor's steps all the time.

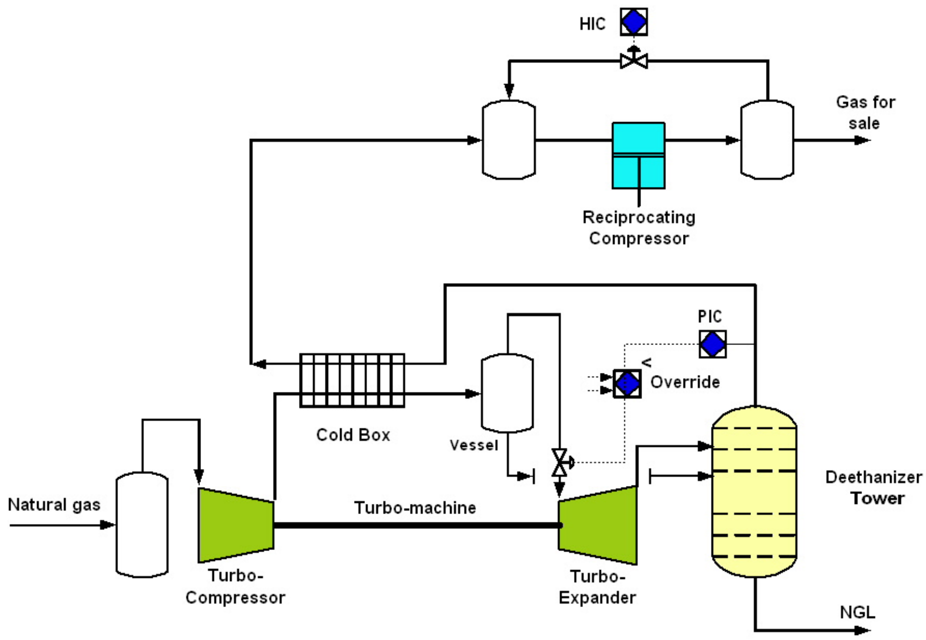


Figure 14 – Schematic of process plant and the old control strategy.

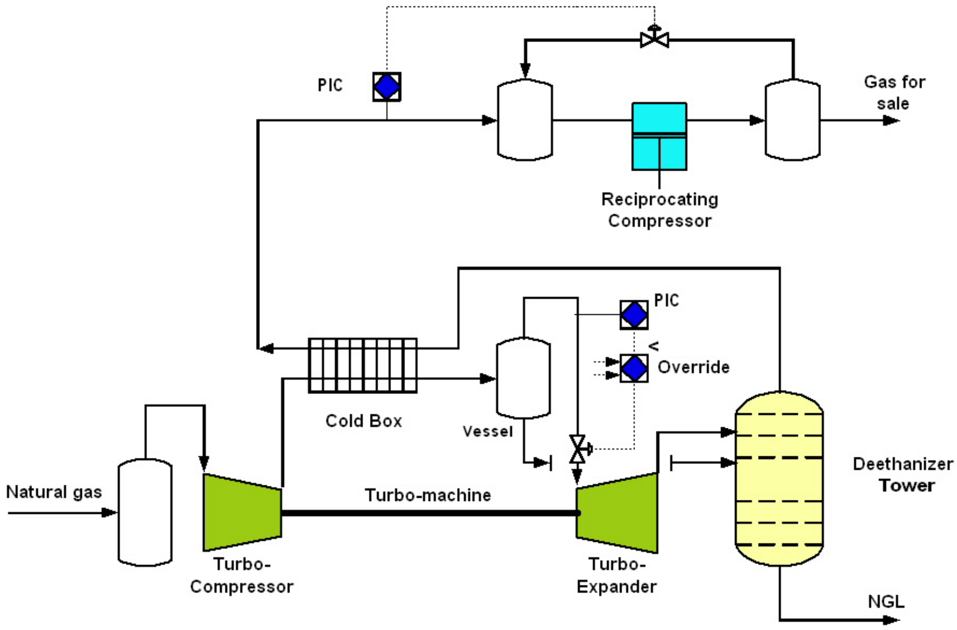


Figure 15 – Schematic of the new control strategy proposed.

Example I - Regulatory Control - Economic gains obtained with regulatory control

In order to calculate the increase in recovery of NGL (natural gas liquids) due to the implementation of this project, it was decided to establish an indicator of the plant performance. This indicator considers not only the absolute production of NGL, but also the feed flow of natural gas processed and its NGL content (percentage of heavy components C_{3+} in the feed).

So it was defined the following indicator of the recovery of NGL:

$$Indicator = \frac{\text{Production of NGL}}{\text{Feed Flow}} \times \frac{10^6}{\text{Percentage of } C_{3+} \text{ in the feed}}$$

Table 1 shows the evolution of this indicator during the project. After the startup, the turbo-compressor had some vibration problems that limited its maximum speed. Thus, the base case was considered after solving this problems with turbo-compressor (row "Before tuning" in Table 1).

All economic calculations were made using the following common base:

- Annual basis;
- Feed flow constant: 2.2 million Nm^3/d;
- Percentage of C_{3+} constant and equal to 5.7 mol%;
- Value of NGL equal to 45.4 US$/bbl.

Table 1 shows that only tuning the PID controllers, with the initial control strategy, resulted in an economic gain for the plant. Before tuning, the indicator was 34.3 and rose to 35.8 after this phase. The profitability increased around US$ 1800000 per year. Therefore, these tools of analysis and tuning of PID controllers can bring great economic gains to industrial gas plants.

The definition of a new control strategy for this plant resulted in another increase of the recovery of NGL. The recovery efficiency of NGL increased 4% after tuning and another 6% with this new control strategy. The total increase of profitability of this plant due to the regulatory control was around US$ 4400000 per year.

Table 1 - Indicator trends and economic gains.

	Indicator	NGL production	Economic gains
Before tuning	**34,3**	**430,9 m³/d**	X
After tuning	**35,8**	**449,1 m³/d**	**US$ 1.800.000 per year**
After new control strategy	**37,8**	**473,7 m³/d**	**US$ 4.400.000 per year**

Figure 16 shows how the loss of C_{3+} in gas for sales fell significantly after the implementation of this new control strategy (average of 0.3 mol% to 0.05% molar). This component is now being incorporated to NGL, and contributing to increased NGL production and profitability of the plant.

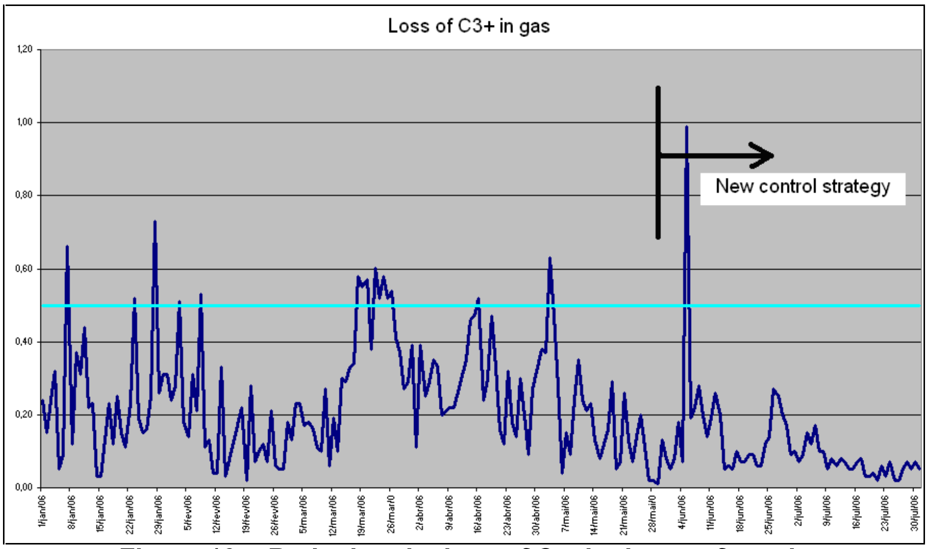


Figure 16 – Reducing the loss of C_{3+} in the gas for sale.

Example I – Advanced Control Implementation

The multivariable predictive controllers (MPCs) are powerful tools for the process optimisation of industrial plants. The benefits are more visible in complex processes like NPGUs where there are challenging dynamic responses due to disturbances (feed flow and composition, energy integration, etc.) that must be dealt with while taking into account process constraints.

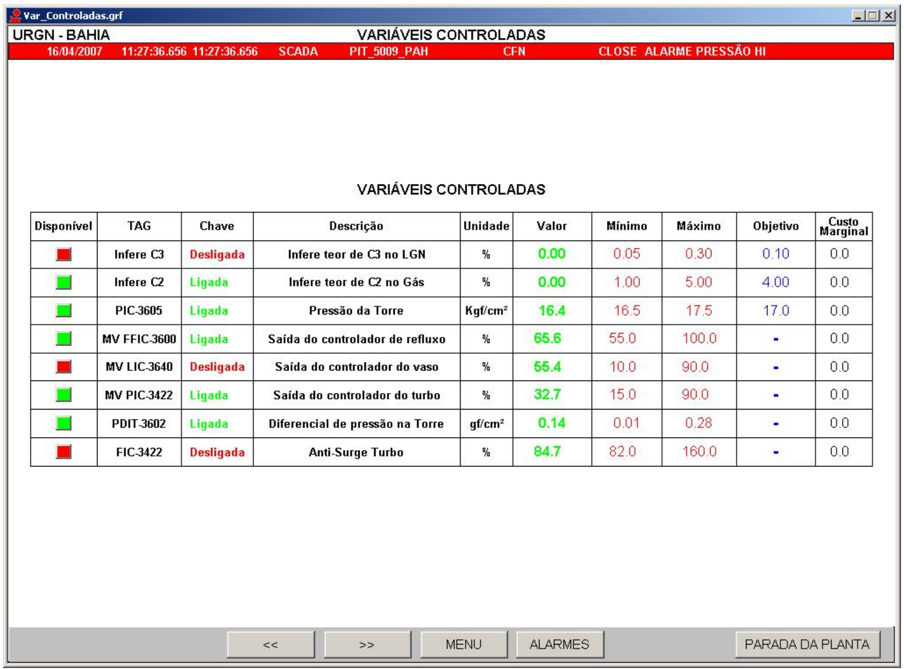
The first step to implement advanced control in this unit was to perform the functional design, where the main optimisation objectives and variables are defined.

For this unit, it was defined the following objectives for the advanced control system:

- Minimize the loss of C_3 in gas for sales.
- Maximize the C_2 contents in NGL (up to a maximum of 4.4%).
- Increase the production of gas or NGL depending on the market.
- Maximize profitability.
- Increase the operational stability.

The advanced control was designed to cover almost all areas of the unit. The controlled variables are the following: inference of C_3 in gas, inference of C_2 in NGL, pressure of the deethaniser, controller's output of the reflux of deethaniser, controller's output of the level of deethaniser, controller's output of the pressure of deethaniser, differential pressure of the tower (deethaniser), controller's output of the anti-surge and temperature of the vessel. It can be observed that only the first two variables (inferences) are associated with the specification

and production goals; all other are constraints to keep the plant in a desired operational range.
The manipulated variables are as follows: suction pressure of the reciprocating compressor, suction pressure of the turbo-expander, temperature at the bottom of the deethaniser, ratio between reflux flow and the feed, reflux of the deethaniser and ratio between two flows in the cold box. The feed flow is considered a disturbance. All manipulated variables are set points of PIDs controllers of the regulatory control.
After functional design, it was known all the dynamic models and inferences (virtual sensors) required for advanced control operation. Identification tests were performed to obtain the dynamic models of the controller. These models were validated with the operators and engineers of the unit, and then the predictive controller has been configured, tuned and implemented in the plant.
Figure 17 shows an example of the interface of the controller, where the operator can change the desired ranges for each variable and its activation (on / off). These screens have been configured in the supervisory system of the plant (SCADA).



Disponível	TAG	Chave	Descrição	Unidade	Valor	Mínimo	Máximo	Objetivo	Custo Marginal
	Infere C3	Desligada	Infere teor de C3 no LGN	%	0.00	0.05	0.30	0.10	0.0
	Infere C2	Ligada	Infere teor de C2 no Gás	%	0.00	1.00	5.00	4.00	0.0
	PIC-3605	Ligada	Pressão da Torre	Kgf/cm²	16.4	16.5	17.5	17.0	0.0
	MV FFIC-3600	Ligada	Saída do controlador de refluxo	%	65.6	55.0	100.0	-	0.0
	MV LIC-3640	Desligada	Saída do controlador do vaso	%	55.4	10.0	90.0	-	0.0
	MV PIC-3422	Ligada	Saída do controlador do turbo	%	32.7	15.0	90.0	-	0.0
	PDIT-3602	Ligada	Diferencial de pressão na Torre	gf/cm²	0.14	0.01	0.28	-	0.0
	FIC-3422	Desligada	Anti-Surge Turbo	%	84.7	82.0	160.0	-	0.0

Figure 17 – Operator's interface of the controller - Controlled variables.

Example I – Advanced Control Benefits

It was used the same indicator to evaluate the benefits of the advanced control:

$$Indicator = \frac{\text{Production of NGL}}{\text{Feed Flow}} \times \frac{10^6}{\text{Percentage of } C_{3+} \text{ in the feed}}$$

Table 2 shows the evolution of this indicator along the project. It is noted that the improvement of the regulatory control allowed a gain of around 10% and with the advanced control (MPC) the gain of the project increased to 31%. Therefore, the regulatory control was responsible for about 35% of the total gain obtained in this project. Figure 18 shows the trend of this performance indicator over the project (from 2005 to early 2009). It can be observed visually the improvement of performance indicator with the advanced control, which besides having a higher profitability (increased recovery of NGL), also showed a greater stability with less variability.

Table 2 - Evolution of performance indicator throughout the project.

	Indicator	NGL production	Economic gains
Before tuning	34,3	430,9 m³/d	X
After tuning	35,8	449,1 m³/d	US$ 1.800.000 per year
After new control strategy	37,8	473,7 m³/d	US$ 4.400.000 per year
After advanced control (MPC)	45	566 m³/d	US$ 13.000.000 per year

Figure 19 shows how the loss of C_{3+} in the gas fell significantly after the implementation of advanced control (average of 0.3% molar to values close to zero). This component is now being incorporated into NGL, and helping to increase plant's profitability. The peaks of losses in Figure 19 are associated with unscheduled stops or shutdowns.

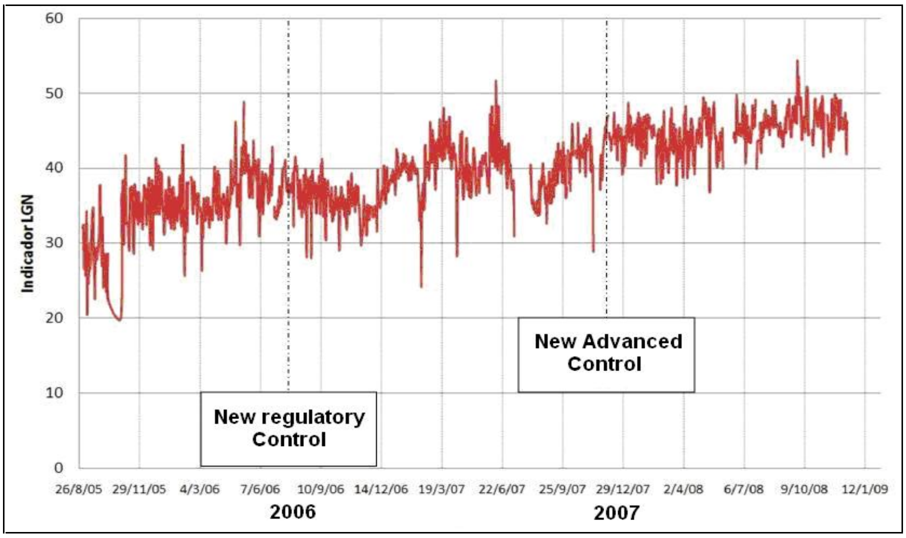


Figure 18 – Improving plant performance with Advanced Control (over three years).

This project provided an increase in profitability of around US$ 13 million per year, due to an increase in daily NGL production of 135 m^3/d. It was considered the same basis as described when assessing the gains from regulatory control.
The next part will described the optimisation of another natural gas plant.

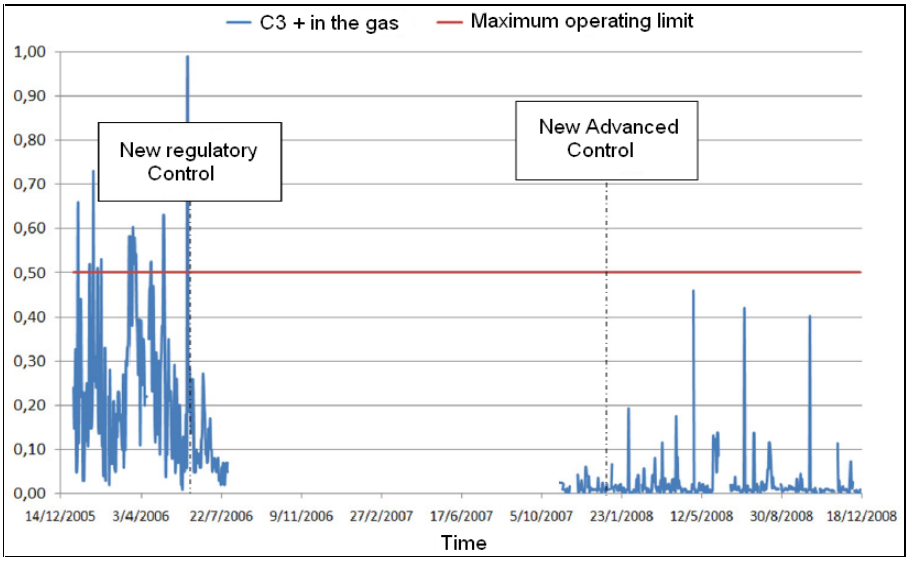


Figure 19 – Reduction of losses (C_{3+}) in gas with Advanced Control (over three years).

Example II - Natural Gas Processing Unit in Guamaré, Brazil

In this unit (see Figure 20), also based on cryogenic expansion with turbo-expander, the Natural Gas feed is fractionated into residual gas (mostly composed of C_1 and C_2), fuel gas (C_2 and a small mount of C_3), LPG and Gasoline. The economic goals here are to maximize the LPG yield, while minimizing C_3 losses [Campos et al., 2008] [Campos et al., 2009[b]]. The unit comprises three distillation columns (see Figure 21), a furnace and a cold box where most of the heat integration of the process is done (see Figure 22).

This NGPU makes use of three utilities: electricity (used in drivers of compressor and pumps), propane (used as refrigerant in the deethaniser's condenser) and fuel gas (used in heaters).

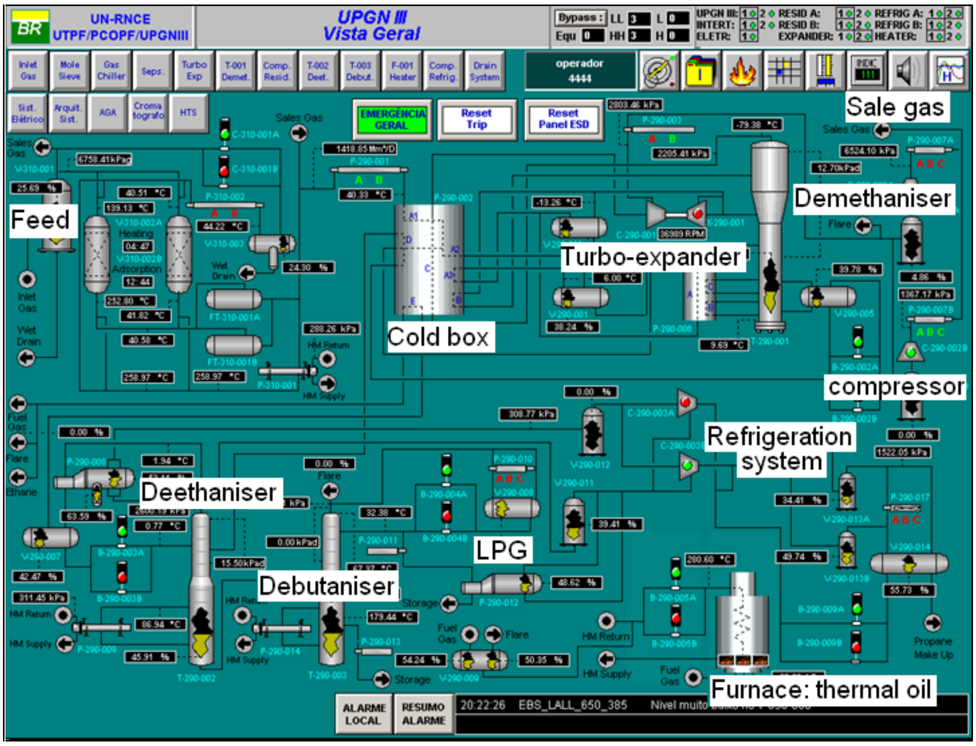


Figure 20 – Schematic of this Natural Gas Processing Unit (NGPU) – Guamaré - Brazil.

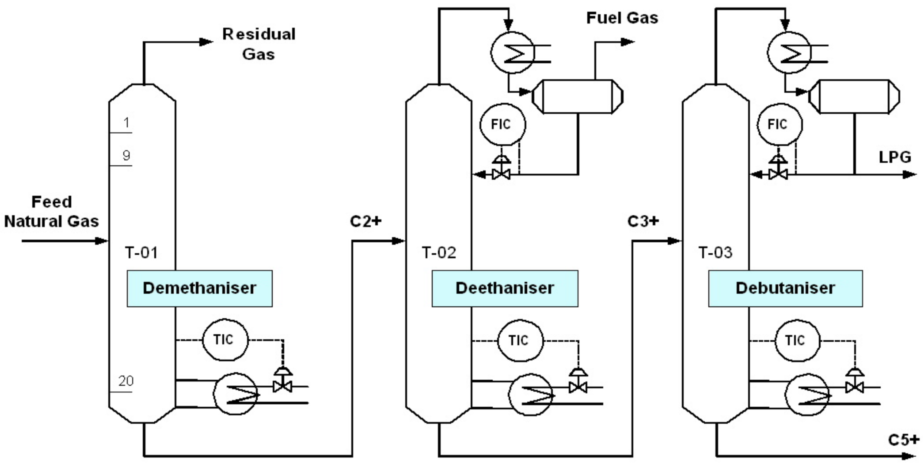


Figure 21 – Schematic of the distillation columns of this unit.

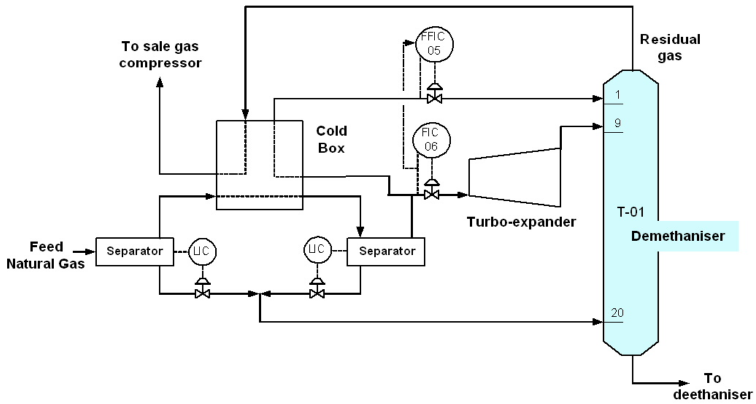


Figure 22 – Esquema de cargas da Torre Demetanizadora.

Example II - Regulatory Control - Assessment of Instrumentation and Control

This phase consisted of installing the software "BR-PerfX" for assessment of instrumentation and regulatory control. After, more than fifty control loops were analyzed and tuned, using the software "BR-tuning". Many levels controls were tuned in order to minimize the movement of the associated flow, improving the stability.

The main problem found in this plant was the turbo-expander flow control (FIC-01 in figure 23), which was operating continuously in manual mode. When operators tried to put this loop in automatic mode there were strong interactions with the pressure control (PIC-01), resulting in oscillations. Figure 23 displays the old control strategy of the unit.

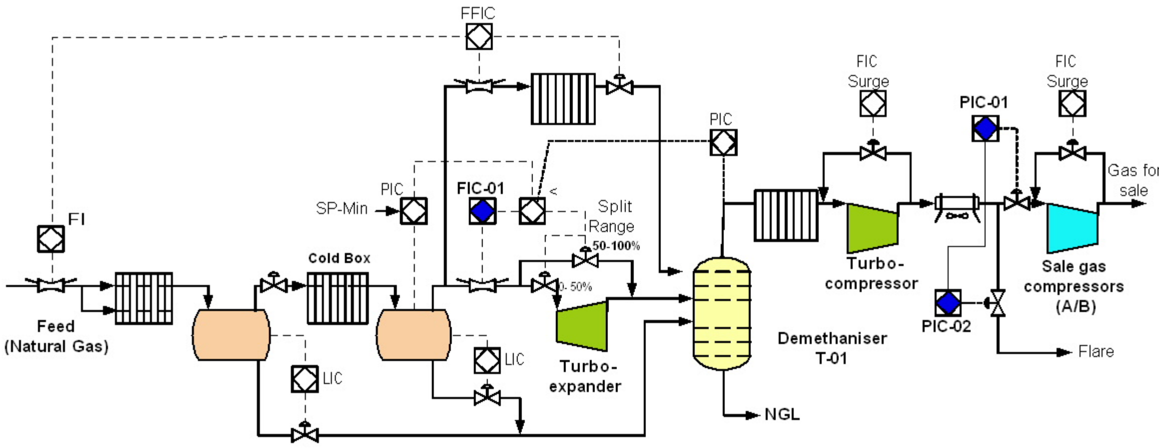


Figure 23 – Old regulatory control strategy to NGPU in Guamaré/Brazil.

The problems with this control strategy were the following ones:

- The set points of the pressure controls in the suction of the turbo-compressor associated with the turbo-expander (PIC-01 and PIC-02) were difficult to be adjusted, because they should be changed if the turbo-expander was operating or not. For example, if there were a shutdown in the turbo-expander, these set points should be changed quickly to lower values in order to avoid a trip in the whole unit due to high pressure in the distillation column (T-01). It was the cause of many trips of the unity, before the advanced control project, and was a great operators' concern.
- The degree of freedom was not well used, because the control strategy was fixing many variables of the turbo-machine (ex. flow, suction and discharged pressure). Therefore the control system was not stable in automatic mode.

- When the compressor associated with the turbo-expander opened its anti-surge recycle valve, the discharge pressure fell and the controller (PIC-01) closed the suction valve of the sales compressor. It is not a correct action, because it will put also this machine to surge condition. This event happened some times and the unit oscillated until a shutdown situation, generating losses.
- The pressure control of the demethaniser didn't operate in automatic mode, in spite of being the most important column of the plant. Abrupt variations in this pressure increase the losses of NGL.

A new control strategy was proposed, according to figure 24. Basically, it was changed the location of the process variables for the pressure controllers (PIC-01 and PIC-02). These controllers are now monitoring the pressure of the distillation column (T-01 - demethaniser), and the pressure between the discharged of the turbo-compressor and the sale gas compressors are free to vary. So, the turbo-expander flow control (FIC-01) in this new strategy will define indirectly the compressor speed, but the discharge pressure is free to accommodate changes in compressors efficiency (degree of freedom for the other control loops).

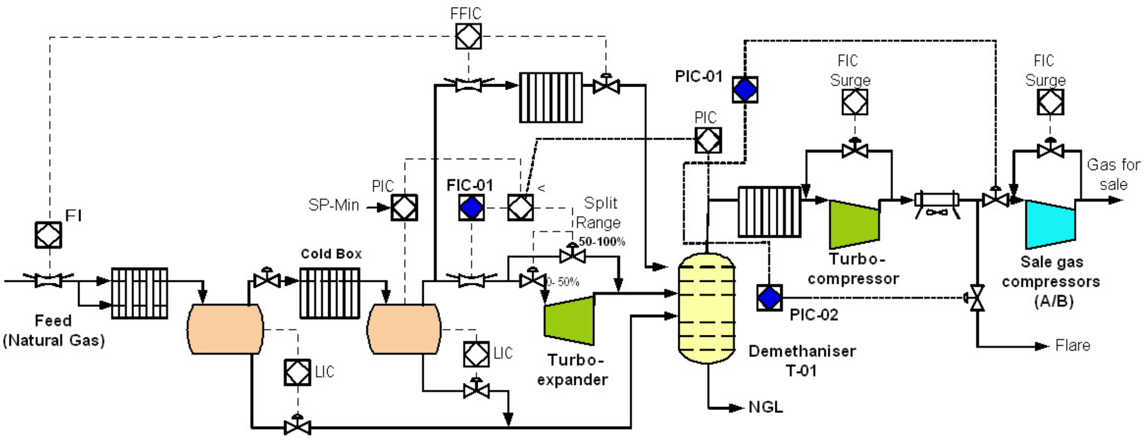


Figure 24 – New regulatory control strategy to NGPU in Guamaré/Brazil.

With this new control strategy, the plant flow control (FIC-01), that manipulates the gas to the turbo-expander, is now also operating in automatic mode continually (see figure 25 where set point is the green curve, flow is the blue curve, and the valve opening is the yellow curve). This new regulatory control strategy does not have the old problems that were associated with shutdown of the unit.

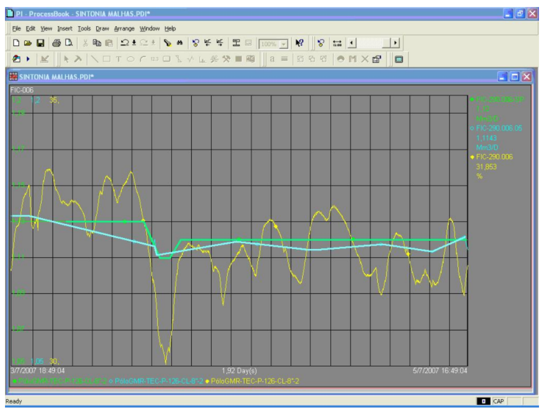


Figure 25 – Flow control in automatic, manipulating the gas to expander.

Another advantage of this new control strategy is that demethaniser pressure control is now in automatic with a good performance (see figure 26), manipulating the suction valve of the sales compressor (PIC-01). With this new regulatory control strategy, the unit is easier to be operated and it is easier to find a better operating point, allowing to increase the profitability of the plant.

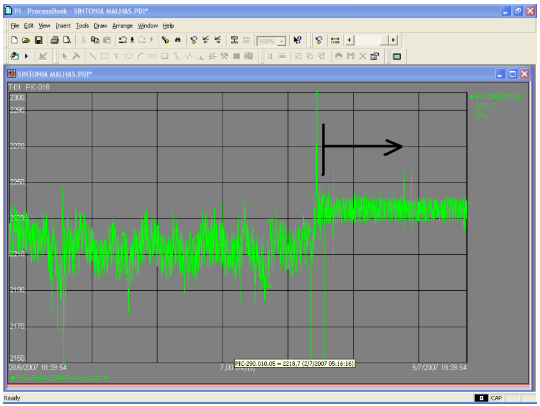


Figure 26 – Greater stability of the distillation column pressure.

Example II - Set point's change to increase the stability of the process

Another problem identified in this unit was associated with large pressure fluctuations in the propane refrigeration compressor. These oscillations were causing great disturbances in the pressures and compositions of the deethaniser, because propane is the refrigerant fluid of this distillation column. Figure 27 shows a schematic of the propane refrigeration compressor system.

In addition, it was not possible to identify dynamic models in this column due to these pressure variations. If it wasn't found a solution for this problem, the potential benefits of the advanced controller would be reduced due to this high variability.

When it was analyzed the causes of these oscillations, it was noticed that the air cooler system was at its limit (lack of heat exchanger area). So, it was not able to condense the propane at the desired discharge pressure of the compressor. Thus, during the night the pressure controllers of the refrigeration system worked well. But during the day, with a warmer air, they saturated. For example, during the day the suction pressure control of the compressor started accelerating until its maximum speed (100%). Beyond this point, the system loses control and the pressure rises in relation to its set point. Thus, all the pressures of the cooling system will vary according to the outdoor air temperature.

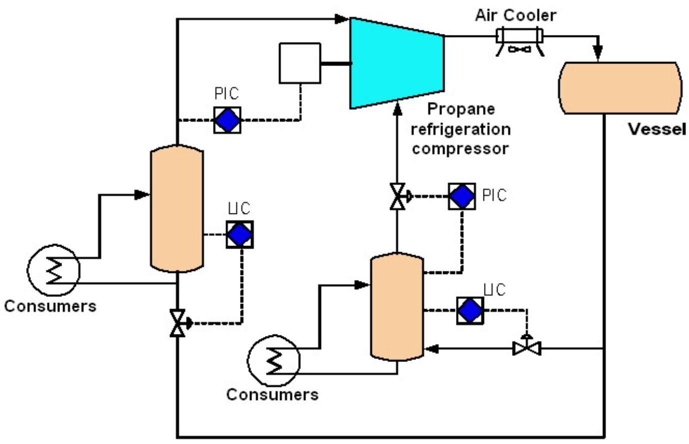


Figure 27 – Propane refrigeration system.

The following question was posed to the team: why not change the set points of the pressure controllers of the compression system to avoid saturation? Using a greater pressure it would be easier to condensate, so even at noon, depending on the set points, it was possible to keep the suction pressure under control. Through an analysis of the history data and process, it was defined new set points (higher values) for pressure controllers of the refrigeration system. Figure 28 shows the gain of stability achieved by this new operating point.

The advanced control can manipulate the set points of the pressure controllers and try to minimize them, as long as the outputs of these controllers are in the desired ranges (not saturated).

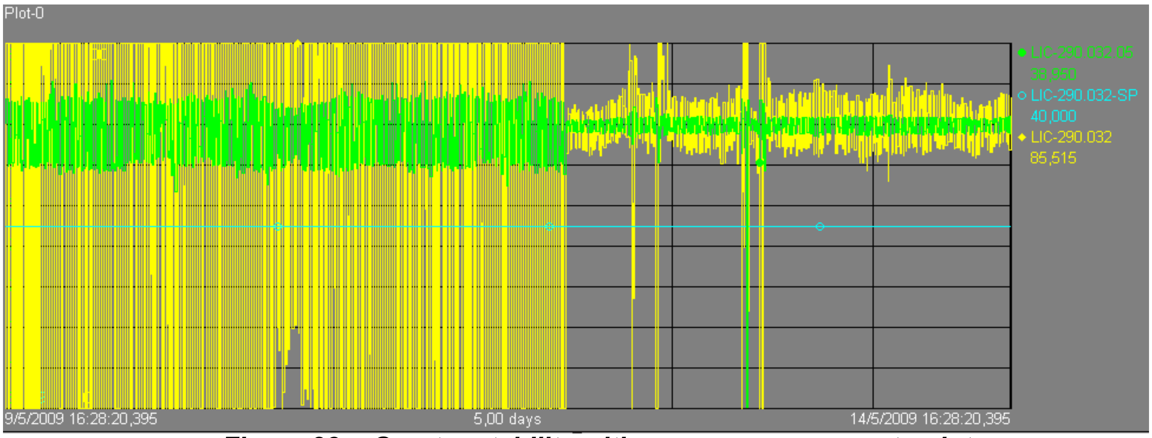


Figure 28 – Greater stability with new compressor set points.

Example II - Methodology for assessment of results

When the team started the task of estimating benefits of this project, two kinds of difficulties arose, that led to unreliable results:

- Computations with process data led to inconsistent results, as recoveries above 100%. These deviations are due to calibration problems in instruments.
- Comparing the performance of the unit at two different times was very difficult due to variations in quality and the flow of the natural gas feed.

To overcome the first problem, the process data were submitted to a procedure called data reconciliation. This method consists in solving an optimisation problem by which the raw data are checked against a reliable mathematical model. As a result of this procedure, it is obtained an equivalent set of data that is consistent with this mathematical model. Variables included in this optimisation process are the errors (bias) of the measurements. In this project, this model consisted of the balance equations for each component and the overall mass balance of the NGPU. Significant errors were detected and corrected. With this new data set, the results of computations became consistent.

To enable the performance comparison of the unit at different times a preliminary assessment was performed in order to determine the major external factors that could affect the overall separation efficiency of the unit. The analysis of several months of process data showed that two prominent effects have great impact on performance: feed flow rate and feed composition (characterized as the percentage of C_{3+}).

To compensate for these effects in calculations of the benefits, the values were corrected to standard values of feed flow and natural gas composition. Figure 29 illustrates the influence of the natural gas composition (blue curve) in the yield of LPG (yellow curve).

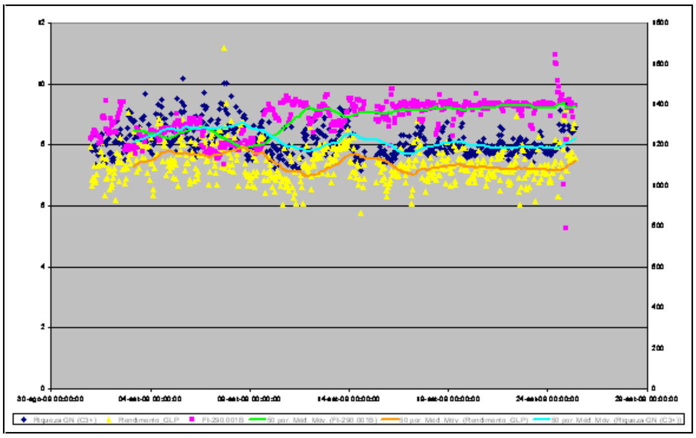


Figure 29 – Effect of the composition of natural gas on the yield of LPG.

Example II - Economic gains obtained with improvement of regulatory control
With the new control strategy implemented in this regulatory level, this NGPU was operating much more stable, minimizing the workload of operators, and minimizing the number of unplanned shutdowns, as shown in Figure 30 (a reduction of about 35%).

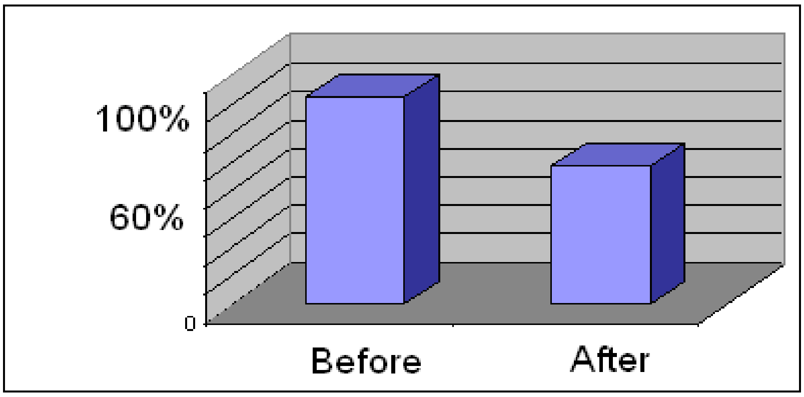


Figure 30 - Reduction in the number of unscheduled shutdowns after new regulatory control.

Table 3 contains the data used in the assessment of benefits. The yields computed were reconciled considering standard conditions (see above procedure). The parameters "a" and "b" correspond to the coefficients of correlations used in the correction of income.

Table 3 - Information for calculating the benefits of regulatory control.

Period	Composition	Feed	Yields reconciled				Adjusted yields for standard conditions			
	(%C3+)	Flow	%LPG	%Gas	%Fuel	%C5+	%LPG	%Gas	%Fuel	%C5+
jun/07	7.30	1435.2	6.01	85.02	7.57	1.4	6.7	84.7	7.2	1.4
nov/08	8.02	1519.6	7.18	83.06	8.4	1.36	7.3	83.1	8.1	1.4
mar/09	8.68	1373.4	7.67	82.89	8.1	1.34	7.3	83.0	8.4	1.3
apr/09	8.82	1394.3	7.4	83.05	8.28	1.28	7.0	83.2	8.5	1.3
may/09	8.37	1475.9	6.89	82.85	9.11	1.15	6.8	83.0	9.1	1.2
sep/09	8.14	1326.1	7.4	83.95	7.32	1.31	7.5	83.8	7.5	1.3
Average	8.22	1420.8					7.2	83.2	8.3	1.3
a (C3+)			0.7	-0.38	-0.37	0				
b (feed)			0	-0.00132	0.0018	0				

The yields were computed in six periods of time. The period from June 2007 corresponds to the period before the first regulatory control tuning, and therefore representative of past performance of the unit. The adjusted production in later periods was used for the calculation of average values, considered representative of the plant performance after June 2007. Thus, with the improvement of regulatory control, the recovery of LPG increased from 6.7% to 7.2%.

Considering the price difference between LPG and gas, and considering a standard natural gas feed flow of 3136 kmol/h (1414 Mm^3 / d), it was computed (Table 4) a benefit of about US$ 1800000 per year due to the improvements in the regulatory control (0.5% increase in recovery).

Table 4 - Benefits with better regulatory control.

Product	Price (US$/vol.)	Yield change (%molar/Feed)	Production change	Gains (US$/year)
LPG	**238.99**	**0,5238**	**1.59 m³/h @ 20C**	**2.735.964**
Gas	**207.37**	**– 1,463**	**– 0.89 Mm³/h Std**	**-1.320.575**
Fuel	**361.00**	**1,1124**	**0.58 Mm³/h Std**	**1.503.454**
C5+	**333.52**	**– 0,112**	**– 0.46 m³/h @ 20C**	**-1111929**
				1.806.913

There were also obtained gains with energy efficiency. The relationship between the fuel gas consumed in this NGPU and the natural gas feed flow was computed at four different times and corrected for the effect of disturbances (feed composition and flow) according to the procedure described above. Then, the average ratio was computed and compared to baseline in June/2007. There was a reduction in specific consumption of fuel gas from 0.0095 to 0.0078 (Mm^3/d fuel gas / Mm^3/d natural gas), which corresponds to a relative reduction of 18.6%. So, there was also a reduction in emissions by this unit, estimated at 1600 ton/year of CO_2 equivalent.

Figure 31 shows this specific consumption of fuel gas for this unit during this project. Note that there are two levels of gas consumption during normal operation of the plant, depending if the dryers are regenerating or not. It can be noted that there was a reduction in gas consumption for the two operation cases.

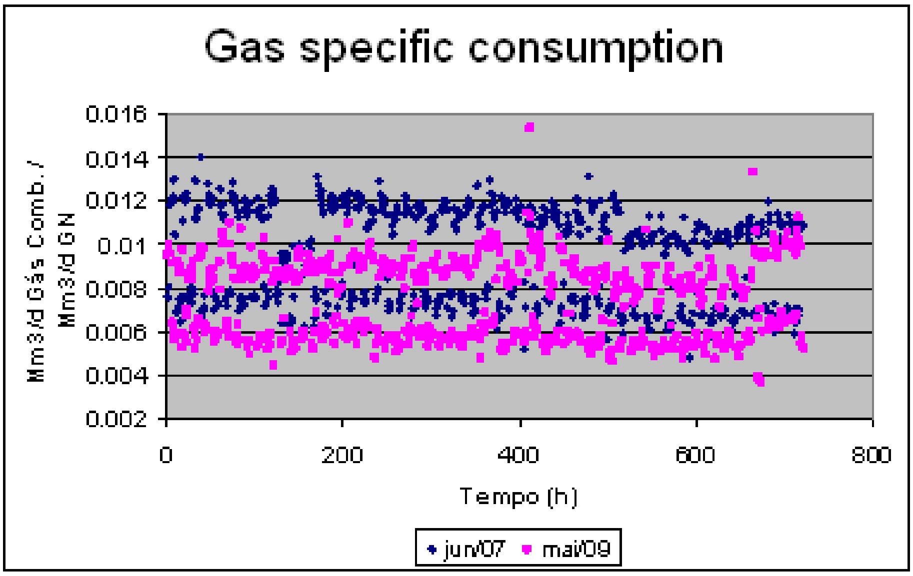


Figure 31 - Reduction in gas specific consumption after better regulatory control.

Example II - Advanced Control Implementation

The scope of this advanced control system covers all areas of this unit. The following objectives were defined for the controller:

- Maximize the natural gas feed flow.
- Minimize the loss of C_{3+}.
- Maximize the production of LPG.
- Guarantee supply fuel gas (ethane) to the consumers.

The functional design of this advanced controller chooses 10 manipulated variables (degrees of freedom) and 16 controlled variables. For this work have been developed inferences for six properties: mole fraction of propane at the top of demethaniser, mole fraction of propane at the top of deethaniser, mole fraction of ethane in the bottom of deethaniser, mole fraction of ethane in LPG (top of debutanizer), mole fraction of C_{5+} in LPG and Reid vapour pressure (RVP) in bottom of debutanizer. These inferences were based on a neural network algorithm and were configured into the advanced control environment. Identification tests were performed (5 batteries of 11 hours for each manipulated variables). The data was used to identify the dynamic model of the process (figure 32). Figure 33 shows the operator interface for manipulated variables, where he can changes the operation limits for each variable and its activation (on / off).

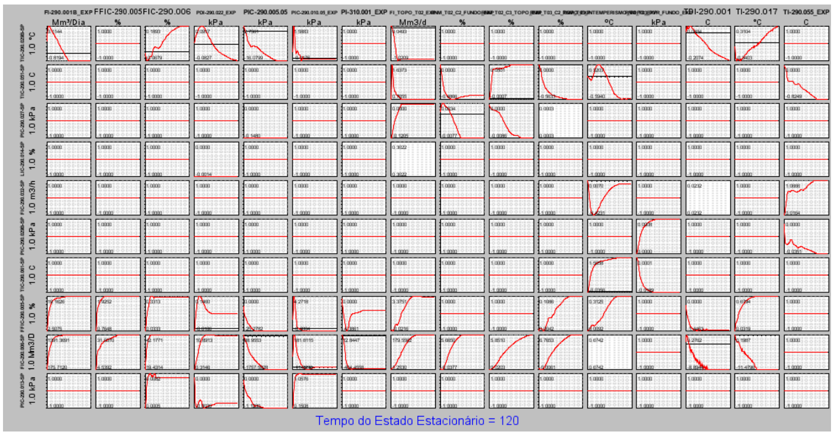
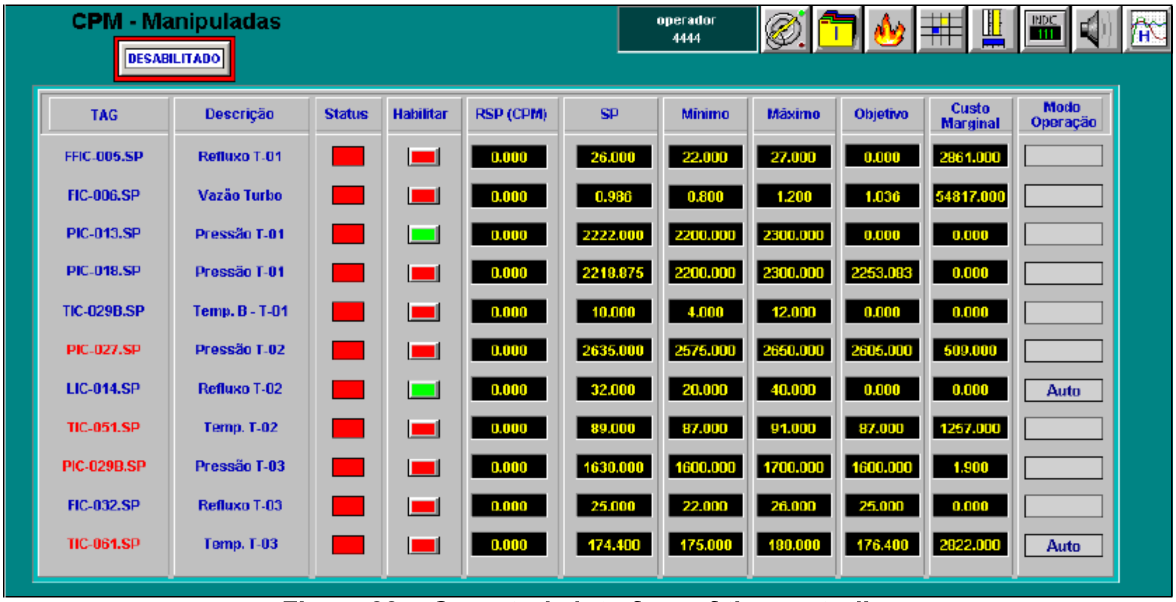


Figure 32 - Dynamic model of this process used in the advanced controller.



TAG	Descrição	Status	Habilitar	RSP (CPM)	SP	Mínimo	Máximo	Objetivo	Custo Marginal	Modo Operação
FFIC-005.SP	Refluxo T-01			0.000	26.000	22.000	27.000	0.000	2861.000	
FIC-006.SP	Vazão Turbo			0.000	0.986	0.800	1.200	1.036	54817.000	
PIC-013.SP	Pressão T-01			0.000	2222.000	2200.000	2300.000	0.000	0.000	
PIC-018.SP	Pressão T-01			0.000	2218.875	2200.000	2300.000	2253.083	0.000	
TIC-029B.SP	Temp. B - T-01			0.000	10.000	4.000	12.000	0.000	0.000	
PIC-027.SP	Pressão T-02			0.000	2635.000	2575.000	2650.000	2605.000	509.000	
LIC-014.SP	Refluxo T-02			0.000	32.000	20.000	40.000	0.000	0.000	Auto
TIC-051.SP	Temp. T-02			0.000	89.000	87.000	91.000	87.000	1257.000	
PIC-029B.SP	Pressão T-03			0.000	1630.000	1600.000	1700.000	1600.000	1.900	
FIC-032.SP	Refluxo T-03			0.000	25.000	22.000	26.000	25.000	0.000	
TIC-061.SP	Temp. T-03			0.000	174.400	175.000	180.000	176.400	2822.000	Auto

Figure 33 – Operator's interface of the controller.

Example II – Advanced Control Benefits

The performance of multivariable predictive control (MPC) has met the objectives set out in the functional design: maximize and stabilize the feed flow of natural gas, reduce variability and maximize LPG production.

The main gain related to the advanced control was to increase mole fraction of ethane in LPG, so increasing its recovery from 7.3% to 8%. It was also observed that variability in this concentration of ethane was reduced by 0.5% with this controller. Thus, the benefits of the advanced control for this unit were estimated in US$ 2500000 per year.

There were also others gains, such us, a better understanding about this unit by operators and engineers. Examples can be cited: the identification that there is a hydraulic limitation at demethanizer (T-01) that occasionally caused flooding and loss of C_{3+} and that condenser of debutanizer operated in its thermodynamic limit, making it heavily dependent on the outdoor air temperature.

Example III - Results in others NGPUs

This section will present further results of this project for other 8 (eight) NGPUs. As it was already said, the performance of instrumentation and regulatory control is fundamental to the successful implementation of an advanced control system. In this project it was evaluated 467 control loops, the results showed that only 29% were operating satisfactorily. The remainder, almost 70%, had some problem. The most common problem was PID tuning, which represents almost 50% of the problems. The maintenance problems (instrumentation) came in second place with 10% of the problems. Operation and configuration problems had similar rates of 8%. Design problems appeared with 5% and control strategy with 4%. It should be note that a control loop can have more than one problem.

After corrective actions, the percentage of control loops with a good performance rose from 29% to 68%. Figure 34 summarizes the overview performance of the regulatory control before and after this project.

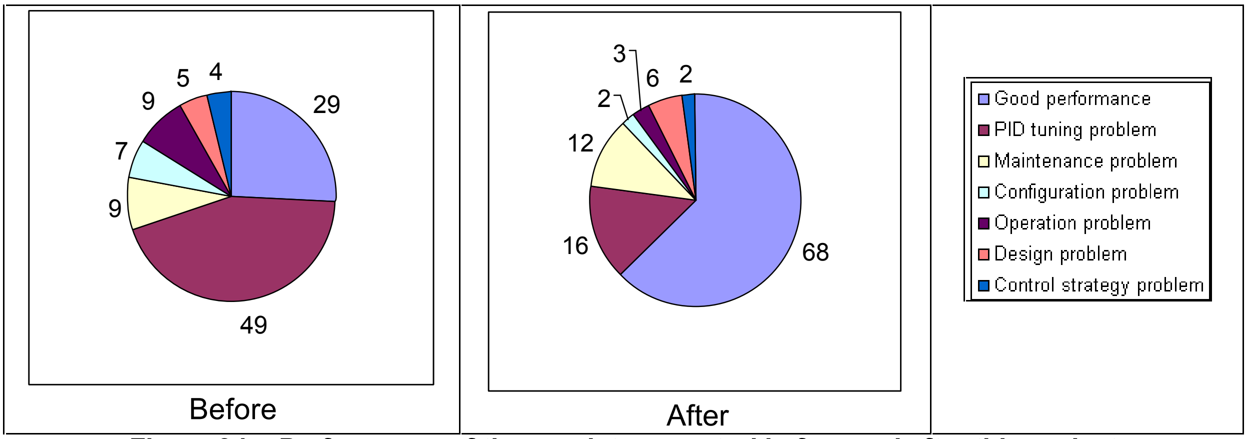
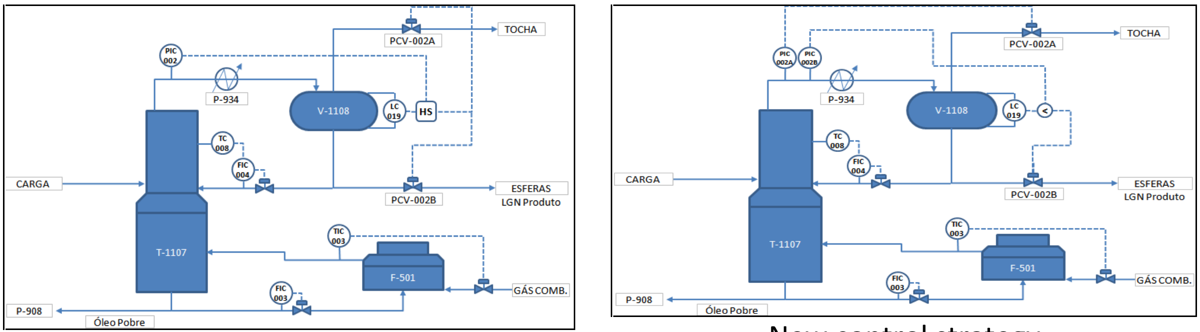


Figure 34 – Performance of the regulatory control before and after this project.

Besides PID tuning, a good control strategy is also an important point to improve the performance of the regulatory control. Another example is shown in Figure 35. The pressure control of this distillation column was accomplished by manipulating the top product. The top vessel (V-1108) operated completely flooded, and if the pressure was going up then the controller would increase its output (the valve position PCV-002B) to decrease the level in the condenser (P-934), exposing more area and so condensing more gas, thus reducing the pressure. But when all the condenser area is used, and the level in the vessel (V-1108) is too

low, then doesn't make sense to continue increasing the position of the top product valve (PCV-002B). So the original control strategy had level switch (HS) in the vessel that detected a low level and switched the pressure controller output to the flare valve (PCV-002A) and closed completely the top product valve (PCV-002B).
The problem with this old strategy is that it was sending to flare all the production when there was a condenser limitation or a disturbance.
The new control strategy doesn't use a switch. It uses a new PID level controller in override with the pressure controller. There is also a new pressure controller in the column with a higher set point. So, when there is a problem (limitation on the top condenser), and the level takes control of the product valve (PCV-002B - sending the product that has been condensed to storage), this new pressure controller will open the flare valve only to maintain pressure tower under control.
This new control strategy allowed to decrease the flow to flare by 40%, which means a reduction of CO_2 emissions by 4380 tonnes per year. Another advantage of this new control strategy was lower pressure variability and greater recovery of NGL (reduction of losses of NGL in the flare from 13 to 8 m^3/d).



Old control strategy New control strategy

Figure 35 – Definition of a better regulatory control strategy.

Advanced control system also allows increasing even more the performance of natural gas processing units. In one unite the recovery of ethane was increased from 40% to 59%, and the recovery of propane was increased from 72% to 82%. In another unit the NGL recovery increased 5%. Another advantage is the possibility to simulate and define better operating points when there are changes in feed flow and composition. For instance, for one unit of natural gas, simulations showed that it was better to change radically the distribution of absorption oil flow between the two towers, regarding to the design values (see figure 36). This was a huge operational paradigm and it took many multidisciplinary analyses to convince operators to adoptee new limits. However, this change allowed to increased recovery of Natural Gas Liquids (NGL) by 5%.

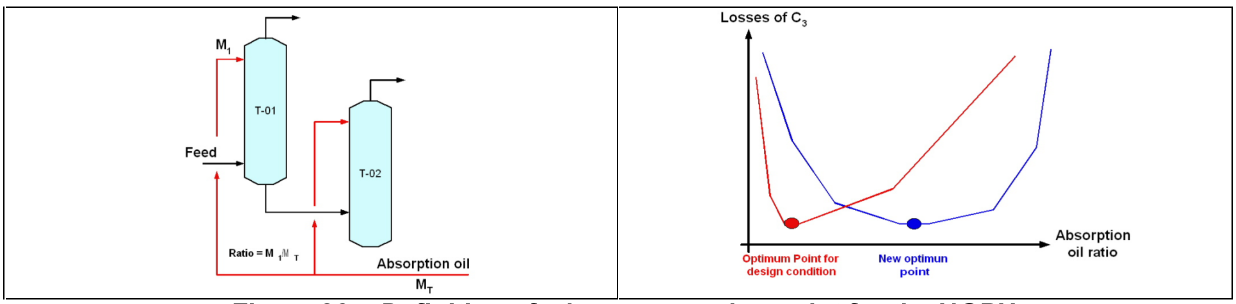


Figure 36 – Definition of a better operating point for the NGPU.

Conclusions

This paper presented a methodology to improve operation and control of several Natural Gas Processing Units (NGPUs), based on the use of advanced process control technologies. The process of APC implementation requires that a comprehensive view of the process unit under analysis is built, considering both the current status and the original design basis. As a consequence, a full revision of process operation procedures, and of the regulatory control system takes place. Therefore, improvements come not only by the action of optimization and advanced control algorithms, but also by the adjustments and corrections performed on other levels.

Between the key factors for successful implementations are the necessary availability of suitable tools that provide the implementation team the means to perform their tasks with good timing and accuracy. In PETROBRAS the proposition of new tools or the improvement of existing ones are a continuous process.

However, the human factor is the most essential. The implementation team is composed by external control and process consultants and engineers and operators from the plant staff that will combine their knowledge, experience and skills to perform the procedures considered by the proposed methodology. Therefore, one must make sure that the appointed plant personnel are available to take part of the project. It was observed that a continuous process of training and involvement of operation, maintenance and engineering teams (and also others like Process analysis, Production Planning and Programming, Automation, IT, Instrumentation and Quality Control) are critical to minimize the impact of "culture shock" caused by the introduction of new tools and operating philosophies, ensuring their continued use and the actual achievement of the benefits envisaged during the design phase.

The great advantage of the proposed methodology and tools is to operate UPGNs more efficiently, in order to increase energy efficiency, profitability and sustainability of these industrial processes.

The effectivity of the proposed methodology could be assessed by the evaluation of key performance indicators. Some of the results were discussed above and are synthesized here:

- Economic gains on NGL recovery of 30%, which generated an increase in profit of about US$ 13 million per year (see figures 18 and 19).
- Increases in energy efficiency by the reduction of 18% on fuel gas consumption, resulting in a reduction of CO_2 emission of 1600 tons/year) (figure 31).
- Minimization of emissions in flares due to higher process stability, providing a decrease of 4380 t/year on CO_2 emissions (figure 35).
- Minimization of the required equipment maintenance due to higher operational stability.
- Minimization of production losses was obtained in a NGPU where the number of shutdown events (unscheduled trips) was reduced on 33%, and the continuous reduction of NGL losses in the fuel gas (see figure 30).
- Significant increase (29% to 68%) on the percentage of regulatory control loops with good performance (figure 34).

An important qualitative gain of this methodology is the increased training of operators and engineers about the best operating practices and process understanding (figure 36).

The experiences of the implementation team along the program execution has led to the conception and proposition of new technologies that might aid implementation, assessment and maintenance of new and existing MPC applications.

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Nomenclature

APC - Advanced process control.
DCS - Distributed control system.
KPIs - key performance indicators.
LPG - Liquefied Petroleum Gas.
MPC - Model Predictive Control.
NGL – Natural gas liquids.
NGPUs - Natural Gas Processing Units.
PLC - Programmable logical controllers.
RVP - Reid vapour pressure.
RTO - Real Time Optimisation.
SCADA - Supervisory Control And Data Acquisition.

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