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# CONTROL LOOP PERFORMANCE ASSESSMENT AND IMPROVEMENT OF AN INDUSTRIAL HYDROTREATING UNIT AND ITS ECONOMICAL BENEFITS

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## RESUMO

### **Auditoria e Melhoria de Desempenho de Malhas de Controle de uma Unidade Industrial de Hidrotratamento e seus Benefícios Econômicos**

Este artigo descreve a auditoria de desempenho de malhas de controle (CLPA) e sua melhoria aplicado a uma unidade industrial de hidrotratamento (HDT) localizada na Refinaria Alberto Pasqualini (REFAP S.A.). É realizada uma revisão sobre os métodos de avaliação do desempenho do controle regulatório e a ferramenta desenvolvida é descrita. Os passos básicos da metodologia de trabalho são também apresentados, com foco nas ações planejadas e implementadas para

melhorar o desempenho dinâmico da unidade. Finalizando o artigo, os principais resultados alcançados para uma aplicação industrial de longo prazo são apresentados, incluindo a pouco usual estimativa dos benefícios econômicos.

**PALAVRAS-CHAVE:** Auditoria de Desempenho de Malhas de Controle, Gerenciamento de Ativos.

## ABSTRACT

This paper describes the CLPA (Control Loop Performance Assessment) and its improvement applied to an industrial hydrotreating unit (HDT) located at Alberto Pasqualini Refinery (REFAP S.A.). It is presented a review on regulatory control performance evaluation methods and the developed tool is described. The basic steps of the work methodology are

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also presented, with focus on the planned and implanted actions to improve the unit dynamical performance. Finishing the paper, the main results for a long-term industrial application are presented, including the unusual economic gains evaluation.

**KEYWORDS:** Control Loop Performance Assessment, Asset Management.

## 1 INTRODUCTION

Since the publication of the seminal work of Harris (1989), the interest in control loop performance assessment (CLPA) has flourished in the control community. This has happened because, for the first time, this community clearly has seen it is possible to deal with the important problem of on-line evaluating the performance of control loops with a realistic framework. In subsequent years, several works concerning this subject arose in the open literature and some tools started to be developed by some control and automation enterprises. Instead of purchasing some of those tools, PETROBRAS, at the dawn of current millennium, decided to support Federal University of Rio Grande do Sul and the Trisolutions enterprise aimed at the development of a new tool for CLPA. This new tool has evolved since then and resulted in the BRPerfX software. This tool was applied to the REFAP's HDT last year. The presentation of this project and its results is the main concern of this paper.

In the sequence of this work, in section 2, it is presented a review on CLPA. Section 3 describes the developed tool, the HDT unit and the work methodology. In section 4, the main results are presented. These results encompass the baseline definition, the actions made to improve the unit dynamical performance, the evidences of that improvement and, in section 5, the estimation of the economic gains after the actions made by the technical team. Finally, some brief concerns about the overall work are presented in the conclusions section.

## 2 A REVIEW ON CONTROL LOOP PERFORMANCE ASSESSMENT

The performance evaluation of control loops is a subject that always was on focus since chemical processes started to be automatically controlled. Very simple and even rudimentary techniques of performance evaluation concerning their deterministic characteristics have always been well mastered. Visual analysis or comparative responses of the controllers for steps or pulses inputs allow little quantitative or subjective conclusions about the controller performance. The observation of the rise time, settling time, overshoot, integral square error among other features always allow the professional to

make findings about the action of the controller. However, these techniques are not of practical implementation for on-line monitoring of the process in routine operation because they require the introduction of disturbances in the process and/or open loop operation. Moreover, the large amount of control loops in a nowadays typical industry makes the task of monitoring the performance using these techniques a practically impossible task due to its huge requirement of human resources in full time. Furthermore, these techniques are not able to explicitly provide any information about the closed-loop performance under random changes, and in most cases not provide an evaluation index that is suitable for online implementation.

Towards automatic evaluation of control loop performance the first study dates back to 1967 (Åström, 1967). However, the basis for the formulation of methodologies for assessing the performance of control loops was defined in 1970 through the contribution of Karl J. Åström. In his book "Introduction to Stochastic Control Theory" (Åström, 1970) some formulations are presented for the prediction of optimal linear discrete stationary processes, from where the minimum variance controller (MVC) was derived. Few studies have been developed in the area in subsequent years.

In the late 1980's, there were several tools that would form a new field, called Statistical Process Control, aiming to monitor the processes variability and detect the presence and sources of disturbances. These techniques, however, do not judge the quality of outputs as a response from control actions.

It is natural that, when trying to quantify performance, to compare the assessment against some kind of reference. According with this logic, it came the proposal made by Harris et al. (1996), which used the minimum variance control as a reference to compare the performance of controllers. The idea was to create an index to quantify the controller performance compared to the theoretical minimum possible variance. Basically the index is the ratio of the variance in the process output generated by a controller that would produce minimum variance and the variance of real control loop output. This work was important because it showed how the techniques studied by Åström combined with simple time series analysis could be used to find an appropriate estimate for the variance of MVC from routine data for SISO (Single Input ? Single Output) systems. His contribution was significant in defining a new direction for the area of performance monitoring of control loops.

Continuing their work, Desborough and Harris (1992) proposed the use of simple linear regression to calculate an estimate of the minimum variance in an automated fashion, which had not been developed in previous work. A recur-

sive procedure, suitable for online calculation of the index, was also presented. An automatic methodology for monitoring the quality of control loops should be simple enough to not require human intervention and should be non-invasive. This means it should not make changes to the system where the monitoring is introduced and must use the least information possible, preferably only the information available in digital control systems. This is the great merit of the methodology proposed by Desborough and Harris, which only requires routine data from process in closed loop and previous knowledge of the process dead time.

Departing from these studies, several researchers began to explore various aspects of performance evaluation based controllers with minimum variance. Huang and Shah (1999) developed a method based on filtering and correlation (FCOR) to estimate the value of the minimum variance and thus calculate the performance index. In parallel, new tools or changes in assessment were introduced, such as procedures for detecting oscillations (Hägglund, 1995; Thornhill and Hägglund, 1997). From these developments, it was possible to develop applications that target real-time performance monitoring for process control systems.

Since then, several developments were made, and the academic productivity still is very active. Ko and Edgar (2000) proposed a MVC based method to estimate performance index of cascade controllers; feedforward controllers were explored by Desborough and Harris (1993) and Huang et al. (2000); MPC performance assessment was addressed by Patwardhan and Shah (2002), Ko and Edgar (2001) and Schäfer and Cinar (2004); among other proposals and approaches. Comprehensive reviews on the subject are available by Harris et al. (1999), Kempf (2003) and Qin (1998). The issue of knowing the a priori time delay was addressed by Huang et al. (2005). More recently, Farenzena (2008) decomposed the now labeled Harris Index in three parts: Nosi (Noise index), Deli (Delay index) and Tuni (Tuning index), according to the contribution of the noise, dead time and controller tuning, respectively, for the Harris index value.

Parallel to the MVC approach, other techniques have been developed or employed to complement the assessment of control loops. Oscillation detection procedures as proposed by Hägglund (2005), spectral analysis, autocorrelation functions, cross-correlation functions, simple statistics and metrics like valve travel and valve reversals are frequently used to monitor and diagnose control loops.

Practical applications are also reported in literature, as found in Harris et al. (1996), Huang and Shah (1998), Thornhill et al. (1999) and Paulonis and Cox (2003). Around 1997, the market started to produce software to perform the task of analyze and assess control loops.

### 3 MATERIALS AND METHODS

This section is divided in three parts. The first one describes the CLPA developed tool. The second part is a brief description of the HDT unit, where the developed tool was applied to. The last sub-section depicts the work methodology developed for improving the regulatory control of process units in general.

#### 3.1 The CLPA tool

Nowadays many companies have already undertaken international action to transform the technology and knowledge developed in software products and support and consulting services. Systems have been developed for evaluating the control loop performance, allowing the whole plant monitoring, and helping to diagnose problems.

Invariably, these applications must perform data collection, filtering of spurious data, execution of computing routine for performance indexes and metrics and storage of the available results for analysis. A good application should (1) be easy to install and configure, (2) have good connectivity with different process control systems, (3) perform accurate analysis, (4) provide friendly user interfaces for both configuration and analysis, and (5) generate detailed reports for control loops evaluation. In this work it was used a tool belonging to this class of software.

The BRPerfX is a real-time tool for performance monitoring and evaluation of control loops developed by Federal University of Rio Grande do Sul and Trisolutions enterprise and supported by PETROBRAS. It allows the identification of the main causes responsible for the bad control performance of a plant with large number of control loops. This is possible due to the state of art level of indexes and metrics available in the software. These includes: (1) oscillation detection indexes, (2) performance metrics based on minimum variance index, (3) traditional statistics, (4) Valve analysis, (5) controllers service factors, (6) PID tuning changing monitoring, (7) several graphical analysis and (8) customized weighted grades for equipments, units and plants.

In its user interface, the process is presented in a defined hierarchy, where the control loops belongs to equipment and the equipments are the contents of the units. Finally, the units belong to a plant and the set of plants to the factory or enterprise. A typical view of the web user interface is shown in figure 1.

The application works online without any interference in the plant operation. The essential information required for its correct work are the PV (Process Variable), SP (Set-Point), CO (Controller Output) and MO (Operation Mode). It is also

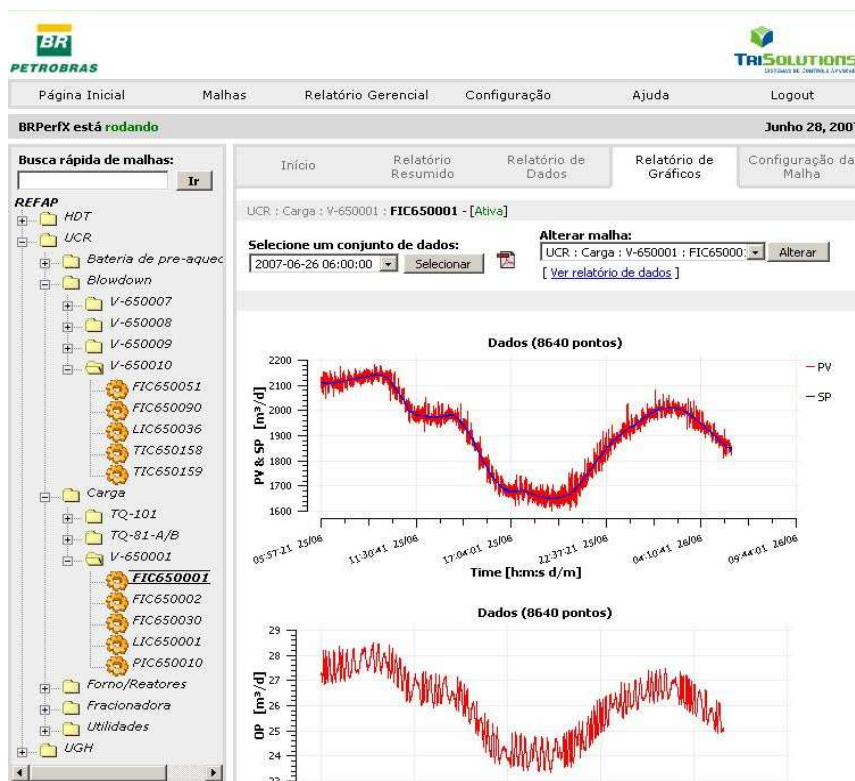


Figure 1: The BRPerfX user interface in web environment.

collected the control parameters (proportional gain, integral time and derivative time) in order to monitor tuning changes. To get these data, it uses an OPC (OLE for Process Control) DA (Data Access) interface, a standard to connectivity among systems largely adopted in industry. The software architecture is composed by three main parts: a Windows service, a database and a Web application. The Windows service is responsible for the OPC connection and data acquisition. The collected data are stored in the machine memory and, after a pre-defined time, the algorithms are run. These algorithms generate indexes and graphs that are saved in the database, being used afterward for analysis. The stored data, after the algorithms run, are discarded and the cycle starts again for a new report. This interval between two computations is labeled report generation period and its value is usually placed about 12 and 24 hours. The database is used to store the service computation results and the basic information of the registered control loops - necessary to the data acquisition and report generation - and, at last, to save and store the user configuration used in the Web application. The last part of the BRPerfX suite is the Web application. This was written in the PHP programming language. With the help of the application interface all the software management and configuration is done. The report visualization is also carried through this interface.

### 3.2 HDT unit

The REFAP HDT unit started up in December 2005 and its main objective is to remove sulfur from Diesel streams. This is done to improve the fuel quality, according to the current Brazilian legislation concerning the sulfur contents. In addition, the hydrotreating generates more favorable storage properties for the fuels and it allows the incorporation of new streams (originally inadmissible) to the diesel pool.

As diesel is the main refinery's product and it can only be sold after being hydro treated, it can be claimed that the continuous and stable operation of the HDT unit is fundamental for attending one of the most strategic enterprise objectives.

The HDT unit has a home made (PETROBRAS) design and is able to treat up to 4,400 cubic meters of diesel per day. A simplified scheme of that unit is shown in figure 2.

The crude petroleum presents some impurities that arise in its products, generating pollution, corrosion, odor and product instability.

The HDT process is based on sulfured and nitrogen compounds reactions with hydrogen gas in presence of a catalyst aiming to remove these pollutants. The hydrogen gas also

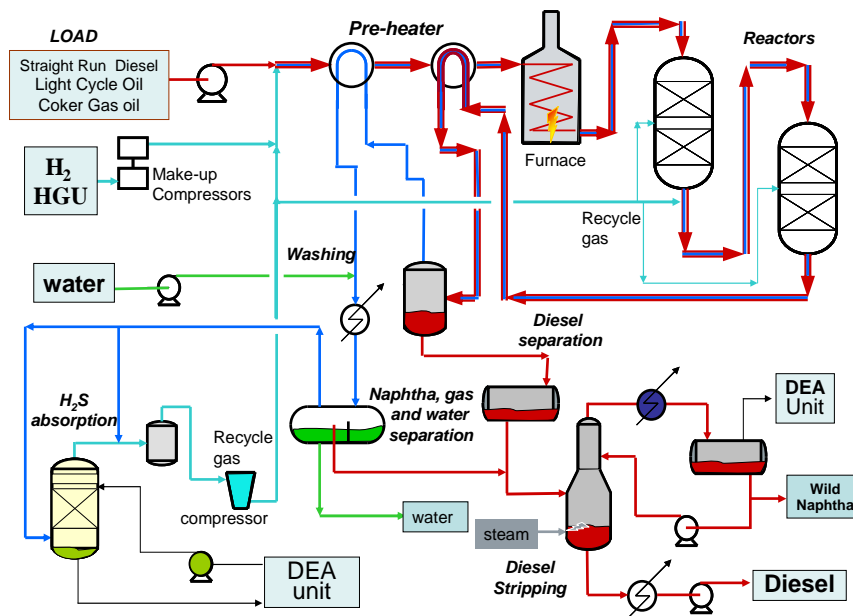


Figure 2: Scheme of REFAP HDT unit.

tackles the double bonds of olefins, improving the hydrocarbon saturation. These reactions, known as olefin saturation, denitrogenation and desulfurization, are highly exothermic and strongly influenced by the current operating pressure. The higher the pressure, the more severe is the process. In this specific case, the reactors operate at 81 kgf/cm<sup>2</sup>, featuring a medium to high severity hydrotreating process. The description of the process flux is given in the next paragraph.

The liquid load, as soon as it enters in the process, receives a hydrogen injection. Before going to the furnace, this stream is heated by a sequence of heat exchangers. After the heating in the furnace, the effluent goes to the reactors itself. The reactors have its temperature controlled by cooling gas (quench) injection. The reactor effluent follows to the cooling and separation sector. In this sector, the stream is divided in two parts: recycle gas and H<sub>2</sub>S enriched diesel. The diesel is sent to the rectification sector. In this sector the light fractions and H<sub>2</sub>S rich gas are removed from diesel. The light fractions form the wild naphtha (corrosive naphtha) that is sent to FCC (Fluid Catalytic Cracking) unit. The H<sub>2</sub>S enriched fuel gas follows to DEA treatment (other unit). The rectified diesel is sent to vacuum drying for removing solved water.

The hydrogen gas (H<sub>2</sub>) consumed in the whole process is supplied with 99.9% purity by an HGU (Hydrogen Generation Unit) that works in parallel with the HDT. The hydrogen admission is regulated by the HDT unit pressure. The HGU also suffered a control loop process assessment and improvement at same time (Longhi et al., 2009).

### 3.3 The work methodology

The objective of this section is to present the work methodology to improve the regulatory control of general units using a CLPA software. This methodology is the result of our practice in working for more than 2 years in this kind of project at the enterprise. It must be considered from now to the end of the paper that the CLPA tool is installed and working well at the chosen site.

#### 3.3.1 Minimum staff

The first concern is related to the necessary staff to successfully carry out the project. This team must be composed, at its minimum, by three people: 1 process control engineer, 1 process engineer and 1 process operator. Each one of these professionals has a fundamental contribution to the control loop analysis and, if one of them is lacking in the staff, the remaining people won't be able to adequately carry out the project.

#### 3.3.2 Control loops registering

The control loop registering is the next task to be done. The control engineer must download a list with all control loops in the DCS unit. After, he must upload these control loops to the CLPA tool. The BRPerfX allows the uploading using a CSV file constructed in Microsoft Excel. This is more convenient because the individual registering of each control loop can be very tedious.

It mustn't be forgotten that prior to diagnosing the control loops there must be an indication about the performance of the individuals control loops as well as the whole set of loops belonging to a unit. This may be attained by a correct definition of the control loop grades. After discussing this problem with the technicians involved in the control projects it was decided to use preferentially two grades. The first one should be used to evaluate the loops where the variability reduction is the main objective. This grade was labeled as Performance grade, as shown in equation 1 and illustrated in figure 3.

This is the most typical requirement for a control loop whose closeness to set-point is required. It must be emphasized that to have a reliable grade it is necessary to correctly inform the control loop time delay because the Harris index (see section 3.3.6) is highly dependent on this parameter. In addition to the Harris index, this grade also penalizes the percentage of time when the loop worked saturated or in manual mode.

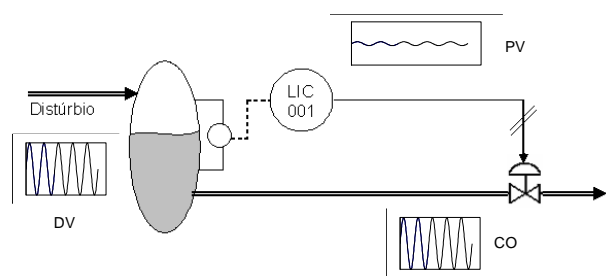


Figure 3: PV stabilization control loop.  
DV means Disturbance Variable.

$$GRADE_1 = \frac{(1 - Harris)}{100} \times \frac{(100 - \%SatTime)(100 - \%ManTime)}{100} \quad (1)$$

where:  $Harris = 1 - \sigma^2(MVCcontroller)/\sigma^2(PV)$ ,  $\%SatTime$  is the fraction of the chosen window time (12 hours in this particular case) where the control loop worked saturated and  $\%ManTime$  is the fraction of time where the controller worked in manual mode.

The other grade was developed for control loops where the main objective is to smooth the CO (Controller Output), typically pressure and level loops used to avoid variation in some material flows of the process, see figure 4.

For this case, it is suggested to use the grade from equation 2. It considers the amplification factor (see section 3.3.6) as the main performance metric. If the control loop has an amplification factor bigger than one, the oscillations are not being smoothed and the grade must be zero. On the contrary, if

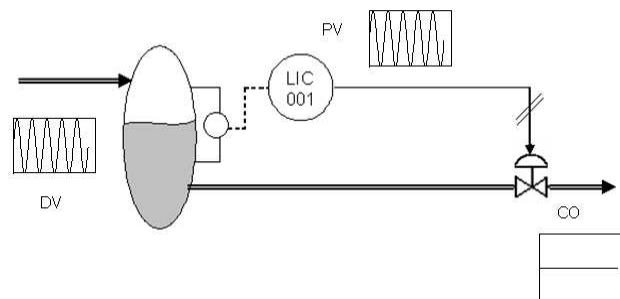


Figure 4: CO stabilization control loop.

the amplification factor is null, the control loop is working as a perfect smoother. Consequently, it deserves the biggest grade. Plus the amplification factor, this grade also penalizes the percentage of time of saturated operation. The manual mode is not considered because sometimes it is acceptable to work in manual for these loops.

$$GRADE_2 = \begin{cases} \text{IF} & AmpFactor < 1 \text{ THEN} \\ & (1 - AmpFactor) \times (100 - \%SatTime) \\ \text{ELSE} & 0 \\ \text{ENDIF} & \end{cases} \quad (2)$$

where:  $AmpFactor$  is the amplification factor, as it is explained in the item h (\*\*\*) in section 3.3.6. The other terms are the same as in equation 1.

### 3.3.3 Data acquisition

Being all control loops registered in the software it is necessary to wait some time before having a minimum amount of report that allow a representative analysis for the process control performance.

This time might vary from 2 weeks to 3 months, depending on the feed stream and operating conditions variations the unit are subject to. This doesn't hinder the execution of the next step.

### 3.3.4 Criticality analysis

This step is optional but it might be very useful to prioritize the future actions and also discard some loops performing bad but negligible for the process performance. In this step, the staff rank all control loops according to the following criteria: Impact on unit production, on product quality, on process safety and on the environment. Nowadays, it is used a

scale ranging from 0 (no impact) to 4 (maximum importance) to attribute grades. After, a weighting grade is composed for each loop using each criterion grade value. The higher is the composed grade, the more critical is the control loop. This step finishes with the ranking of the control loops according to their criticality.

### 3.3.5 Baseline definition

After being finished step 3.3.3 it can be defined the process control performance baseline. The baseline is a temporal portrait for the usual performance of the unit before the corrective actions. It will serve to compare and quantify the dynamical improvements and economic benefits gained after the project end. The baseline is composed by the average grade and indexes of the unit and the control loops considering a time interval that realistically represents, according the staff judgement, the usual plant operation. This period should be carefully chosen to avoid inclusion of maintenance operation periods, plant shutdown and uncommon instabilities, for example. This step can be done in conjunction with the next one.

### 3.3.6 Control loops diagnosis and corrective actions scheduling

Along the baseline definition, it starts to be felt the problems affecting each control loop. To make a realistic diagnosis, the grades and average indexes from the baseline are used as main tools, plus the graphical and information data generated by BRPerfX, as well as, the information contained in the enterprise database - in this particular case, the OSI Soft PI (Plant Information) software.

The diagnosis can be more properly made if each loop is classified in one or more of the options shown below:

**3.3.6.1 DCS (Distributed Control System):** DCS configuration problems (example: lacking of PV filter).

**3.3.6.2 BrperFX:** Incorrectly or inadequately registered control loop.

**3.3.6.3 Maintenance:** Final element or sensor problems.

**3.3.6.4 Tuning:** Performance limited by inadequate tuning.

**3.3.6.5 Operation:** Control loop being operated incorrectly.

**3.3.6.6 Design:** Inadequate control structure design or equipment sizing. It is necessary to review the control structure or to resize the equipments.

**3.3.6.7 OK:** Control loop operating as desired.

Every technician, when aiming to diagnose the control loop behavior, can identify or use a personal suite of indexes. In this present work, it was considered the metrics and graphics depicted from letters (a) to (k).

For a detailed explanation about the indexes (a) to (d) and its computation, it is strongly recommended to consult Farenzena (2008). For indexes (e) to (m), see Kempf (2003). Both references are easily available for download at LUME - UFRGS digital repository. The address is in the References section.

**a. Harris index.** It is an index that measures the potential of performance improvement of a loop in terms of variance reduction. The index is obtained basically by the complement from the ratio between the variance produced by a hypothetical MVC applied to the loop and the actual loop controller variance. The index is comprised in the range from 0 to 1, where zero indicates optimal performance (no variability reduction potential) and values near the unity indicate absence of control.

**b. Noise index (NOSI).** This index represents the percentage of the total variance that is caused by noise. If this index is high (and the Harris index is not adequate) it is recommended to implant some filter in the PV.

**c. Delay index (DELI).** This index is the percentage of the total variance due to the time delay (or dead time). This kind of variability mightn't be avoided by using feedback control because it is an intrinsic process performance limitation. The ideal situation is to have a good Harris index with a great value for the DELI index.

**d. Tuning index (TUNI).** This index is the percentage of the total variance accessible to feedback control. If this index is high (and the Harris index is not adequate) this is a clear indication of a need to tune the control loop parameters. To confirm tuning problems, it is necessary to analyze the Autocorrelation function.

**e. Autocorrelation Function (ACF).** The ACF furnishes extra information concerning the performance of control loops. As the main objective of a controller is to avoid correlation between the PV and its past values, its value must be close to null after passing the time delay (for a well tuned controller).

A natural occurring phenomena in sampled data systems is to get a data set in which a certain point is strongly dependent from previously collected points, that's to say, the state of the system is defined in great part by its past. The sampled signal of systems in free evolution results greatly correlated.

The role of a controller is to act over the system in a way to prevent the process from taking its natural tendency

and keeping at set-point. The signal over action of a well-tuned controller results not correlated. The ACF presents the correlation that a certain point of the sample has regarding to points previously sampled. As the time delay vanishes, period in which the controller cannot act, the ACF must be eliminated to characterize good control.

In some cases, the analysis of the ACF is more useful than the Harris index because the former can present nonlinear correlation between the control loop performance and the index value itself.

**f. Spectral/temporal analysis.** It is the analysis of the existence of oscillatory components in a signal. Peaks in a graphic of spectral/temporal analysis indicate the frequencies or periods in which oscillation occurs. The spectrum is obtained by the processing of the variable's signal by a Fast Fourier Transform (FFT). The spectrum produces results in frequency domain and can be converted to time domain (temporal analysis), where the results indicate oscillation periods in time units.

The resulting diagram shows the main oscillation frequencies in the PV signal. It can be very useful to corroborate the PV-CO analysis or to identify coupled variables (same peak frequencies).

**g. PV-CO diagram.** This diagram is very useful to qualitatively observe possible problems in the control final element. There are some typical shapes (ellipse and parallelogram) for this diagram that indicates the existence of control valve hysteresis or stiction. A person with a reasonable knowledge in control theory is able to recognize these patterns.

**h. Disturbance amplification factor (*AmpFactor*).** This is a technique applied only to surge level or pressure loops. This factor is given by the ratio between the summation of the controller output (in %) and the IAE of the process variable (converted to %, if necessary, using the PV range). A value greater than one indicate that the loop is not reducing disturbances at process input, while values smaller than one indicate that the process is smoothing disturbances.

**i. Traditional statistics.** Classic statistics like mean, variance, standard deviation, percentile error and operational measures like percentile of time in manual operation or percentile of time in saturated condition were also used in the worksheets and add value to the loop diagnosing.

**j. % Saturated time (*%SatTime*).** It indicates the percentage of time that the control loop works above or below the CO limits (high and low). It is useful to readily see the control loops that are not properly working.

It also is useful to understand why some grades are not good even in the evidence of other not so bad indexes values.

**k. % Manual time (*%MantTime*).** It indicates the percentage of time that the control loops work in manual (or not normal) mode. It is also useful to find control loops that are turned off. As the former index, it is also useful to understand why some loops have bad grades.

Furthermore, it is also noted some comments about the kind of necessary action to be made in case of control loop needing improvement. Based on the diagnosis, the average grades (baseline) and the criticality for the whole set of control loops, it is defined the priority for "attacking" the control loops. This choice is made using as the main criterion to improve the more process impacting control loops.

### 3.3.7 Corrective actions

After having clearly defined the more critical unit control loops, it is started up the fundamental step of the project. At this time, each loop is "attacked" to remove the limitations for a better control performance.

According to the limitation nature, there are different ways of solving:

**3.3.7.1 DCS:** In this case, there are DCS configuration problems and the refinery's Optimization or Instrumentation management staff easily solves them.

**3.3.7.2 BrperFX:** If the control loop is badly registered in the software, as in the former case, the Optimization management staff can solve it very easily.

**3.3.7.3 Maintenance:** This kind of problem is related to bad functioning of sensor or final elements. In this case, the process operators are advised to open an internal maintenance requirement. The Instrumentation and maintenance management are the responsible for its solving.

This kind of problem, depending on its origins, couldn't be solved before a plant shutdown. If the diagnosed problem was control valve stiction the Optimization management can insert a stiction compensator (an in-house development, under patenting process) meanwhile.

**3.3.7.4 Tuning:** In case of a bad performance caused by control tuning, the Optimization staff can solve it. The time to finish the task depends on the total number of control loops to be tuned. It can be estimated something



about 10 loops per week. So, for a typical unit, composed by 80 control loops and about 50% needing re-tuning, the forecasted time to execute the task is placed as one month of continuous work.

**3.3.7.5 Operation:** If the diagnosed problem is an incorrect control loop operation, the project staff contacts the process operators to elucidate the reasons for that, giving a new training if necessary or trying to solve the process or control problems responsible for operating in disagreement with the expectations.

**3.3.7.6 Design:** This kind of problem is related to an inadequate control project. It is not usual to find them. In that case, the control structure needs to be reviewed and the solving sequence is to send a new proposal for the Engineering, automation and process unit managers. As the changes involve many people and enterprise sectors, the problem solving can last more than the project duration. However, this time can vary, depending on the interest of the plant management.

**3.3.7.7 OK:** In this case, the control loop operates in accordance with the expectations. So, this control loop can be used as a replier for tuning similar loops.

**3.3.8 Evaluation of the implanted actions:** After having implanted the corrective actions, a sufficient time period (usually 1 or 2 months) must be waited before re-evaluate the grades and indexes of the changed control loops. The loops that don't present significant improvements are re-analyzed to confirm the correct classification and diagnosis of them.

**3.3.9 Economic benefits computation:** Having finished the first cycle of corrective actions, the economic gains of the unit are evaluated. It is used the difference between the profit and losses from the current situation - for a representative period of operation after the made actions - and the ones from the baseline. This will be explained better in subsequent sections;

Intangible benefits, as the higher plant reliability should only be cited in case of clear evidences of them.

**3.3.10 New improvement cycles:** When the project is finished, the enterprise units remain registered and generating reports in the software. Even the process analyst as the control engineer might continue to monitor the plant performance. Then, improvement actions for the whole unit or specific loops can continue to be done periodically or by demand.

## 4 MAIN RESULTS

In this section, it is presented the actions made in the HDT unit (section 4.1), and the dynamical improvements resulting from those actions (section 4.2).

### 4.1 Actions made in HDT unit

In order to help the work, the HDT unit was divided in six sectors, as shown in the tree directory structure of figure 5.

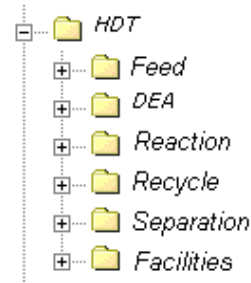


Figure 5: Tree structure for HDT process division.

It was followed the work methodology presented in section ???. The loops were classified according its problems. The summary of the baseline is presented in table 1.

Table 1: The baseline for the REFAP HDT unit.

Number of loops:	56	100%
Problem classification:	# loops	Percent
DCS configuration	9	16%
Maintenance	1	2%
Tuning	27	48%
Operation	8	14%
Design	4	7%
No problems detected - OK	3	6%

The project for the HDT unit started up in middle 2008 and finished at the end of the same year. At the time of writing this work only the tuning and DCS actions were finished. The comparison between the control loops classification before and after these actions is shown in table 2. For the managements involved in the HDT unit, the regulatory control is working well and the unit profitability can only be increased by higher level control and optimization strategies. So the remaining actions possibly won't be concluded.

In the sequence of this section, the evidences of the dynamical improvement for some illustrative control loops are presented and, in section 5, the economic benefits linked to the

Table 2: The control loops before and after classification.

Problem classification:	BEFORE	AFTER
DCS configuration	16%	16%
Maintenance	2%	2%
Tuning	48%	4%
Operation	14%	14%
Design	7%	7%
No problems detected - OK	13%	57%

results from table 2. The figures of this section are presented in Portuguese. This is because they are high fidelity copies from the CLPA software web interface or the plant operation HMI (Human Machine Interface) itself.

## 4.2 Dynamic improvements

In this section, it is presented some illustrative results showing the performance improvement due to the corrective actions made on some control loops.

The starting point is the feed section - see figure 6 - showing an example of a good working control loop. The loop in question is labeled FIC700001, responsible for the control of the SRD (Straight Run Diesel - the diesel coming straight from the distillation units) flow. If it is seen the PV-SP diagram alone, shown in figure 7 for a 12 hour time range, one can suppose but cannot conclude that the loop is well tuned or working well. To validate this conjecture, the engineer can analyze the Harris index (table 3) or the autocorrelation diagram (figure 8), both available in BRPerfX, for the same data period. The two information sources show that the loop controller is very close to the estimated minimum variance controller as well as the control loop is working to eliminate the influence of the past measurements (non-correlated signal). The TUNI index also tells that there is not so much to do for improving the controller tuning. So, it can be easily concluded its good performance.

After the presentation of an example from the feed section, now, it can be considered the very important reactors sector, figure 9.

This sector can start by showing an example of bad behavior caused by final control element. This situation has occurred in FIC700012, responsible for the quench admission in the second reactor.

This situation was supposed only by looking the PV-SP and CO dynamical plots (figure 10). This figure shows a typical behavior of control loop with a “stictioned” control valve.

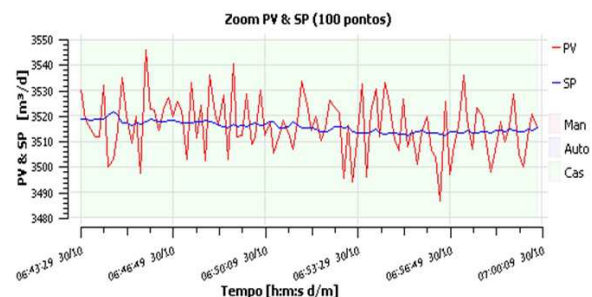


Figure 7: The FIC700001 PV-SP diagram for a twelve hours period.

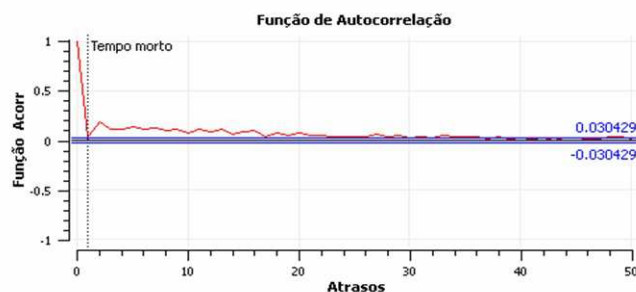


Figure 8: The FIC700001 PV-SP ACF diagram for the same time period of figure 3.

Table 3: Some indexes for FIC700001 related to the same period of figure 3.

Control loop	FIC700001
Control loop grade	92.029
Harris	0.08
NOSI	0.917
DELI	0
TUNI	0.083

Nowadays, it can be only confirmed this situation by analyzing the PV-CO diagram, shown for the same data period in figure 11. However, it is going to be released in a near future a new index for the BRPerfX suite which will help to quantify this kind of occurrence.

In fact, the stiction detected is about 0.4% of the valve moves. This has lead to a 400 Nm<sup>3</sup>/h PV variation (2.5% of range). These oscillations affect the hydrogen flow control and the grade of this control loop is placed about 20%, a very low grade for a flow control loop. Fortunately, the resulting high frequency of oscillation is not disturbing the reactors temper-

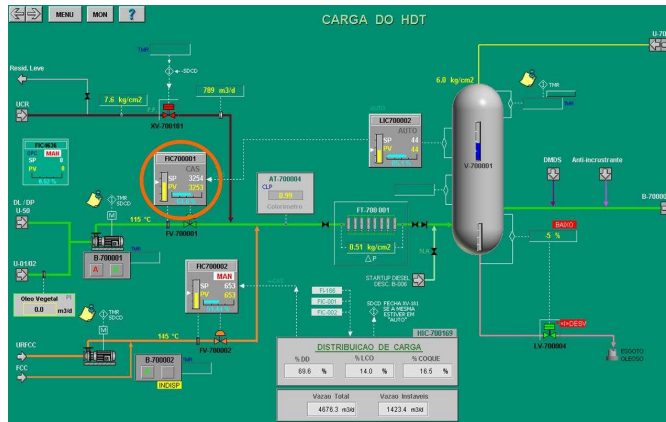


Figure 6: The DCS screen for the HDT unit load section (FIC700001 is circled).

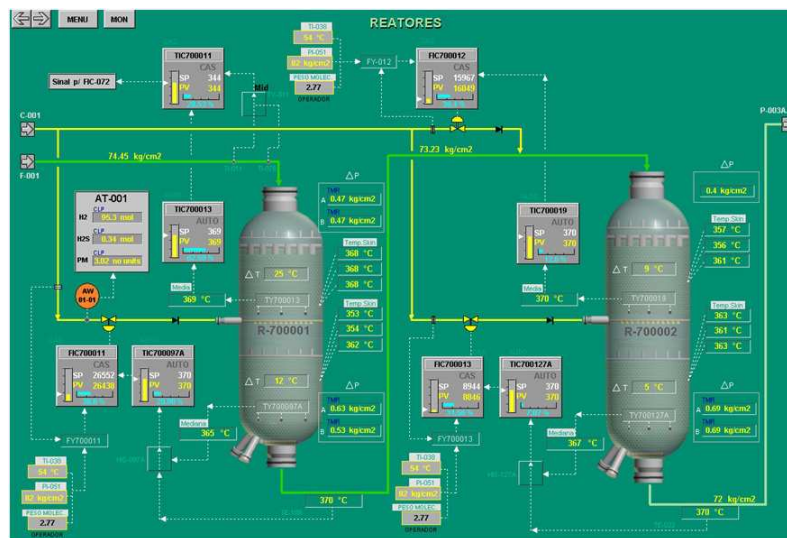


Figure 9: The DCS screen for the HDT unit reactor section.

ature control due to the filtering caused by the retention time of them.

Now let's consider the reactor temperature control. All temperature controllers from figure 4.5 were retuned. As an illustration it considers the second reactor effluent temperature control (TIC700127A).

As the other TICs from the reactor sector, this controller had a very slow action. Despite temperature controllers being slower than other PV controllers; the velocity wasn't the necessary one to adequately compensate the eventual disturbances of the unit. This can be seen in figure 12a. In the lower graph of this figure it can be seen the ACF diagram,

where it becomes clear that the controller velocity was really very slow.

So, it was performed a tuning in the PID, changing the gain more than 3 times its original value. The resulting performance is shown in figure 12b. The average error has diminished to half of the former value. Despite the ACF diagram (and the BRPerfX indexes) shows that the controller performance is still not the best achievable one, the grade of the loop has evolved from near to zero to something about 40%.

The results achieved in this controller tuning are similar to the others belonging to this section. This improvement also helped to turn on the temperature controller from the

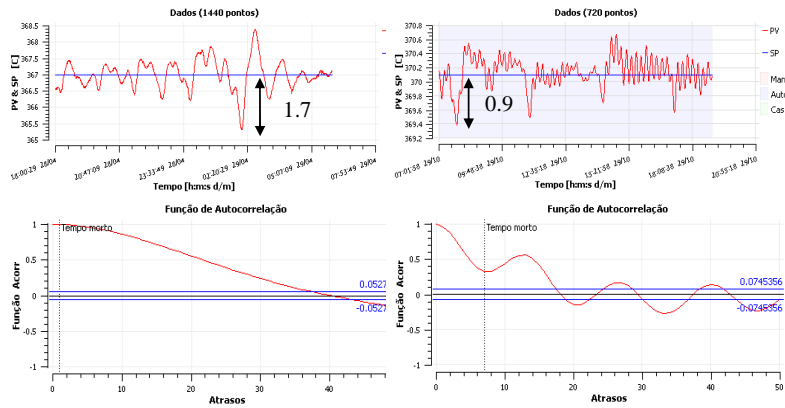


Figure 12: Illustrative PV-SP dynamical plot and ACF diagram before (a) and after (b) the corrective actions for TIC700127A.

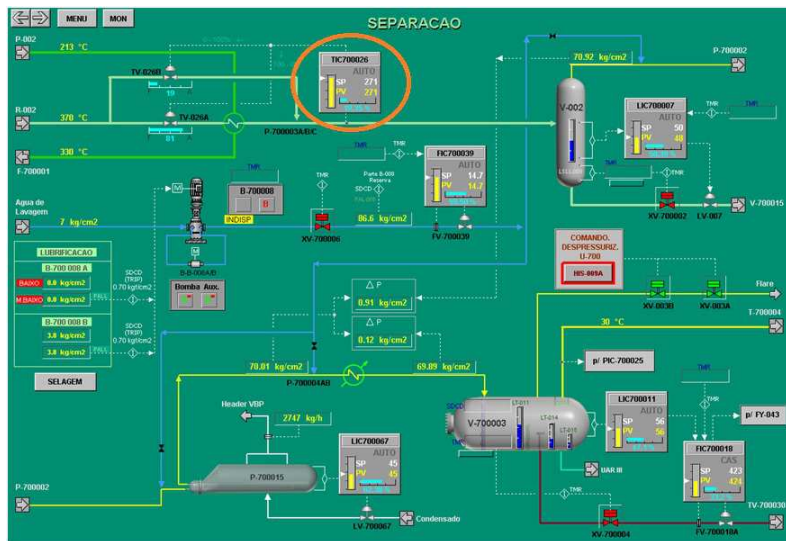


Figure 13: The DCS screen for the HDT separation section (TIC700026 is circled).

high pressure separation drum (figure 4.9) whose tag is TIC700026. This controller was turned off not only because of the affluent stream temperature but mainly due to the high and sudden disturbances on heat exchangers P700003 (three devices). This controller, instead of other temperature controllers, must be tuned very fast to compensate the disturbances from the heat exchangers. So, having the affluent temperature well controlled, the controller was retuned by reducing its integral time from 400 to 20 seconds and halving the gain. One illustrative figure showing the improvement in the control loop dynamical performance is presented in figure 14. In this figure it can be seen two disturbances occurrences (the graphs consider a 12 hour time range) and the corresponding controller response (lower plot) used to sustain the PV near to the set-point. After this new tuning, the

loop grade has changed from a null grade to values higher than 90%.

To end this section, it can also be said something about the rectifying tower T700001 (figure 15).

T700001 is responsible for the diesel to wild naphtha separation. It is highly desirable to maximize the production of diesel instead of naphtha. For doing this, it is necessary to reduce the top temperature until its safety limit. So, it is necessary to have a narrow control of its variation. The top temperature controller (TIC700029) was not retuned because, after the actions made in previous HDT sectors, the tower affluent had its variability reduced and the existing controller works well. This can be validated by looking figure 16 (ACF diagram) and 17 (PV histogram).



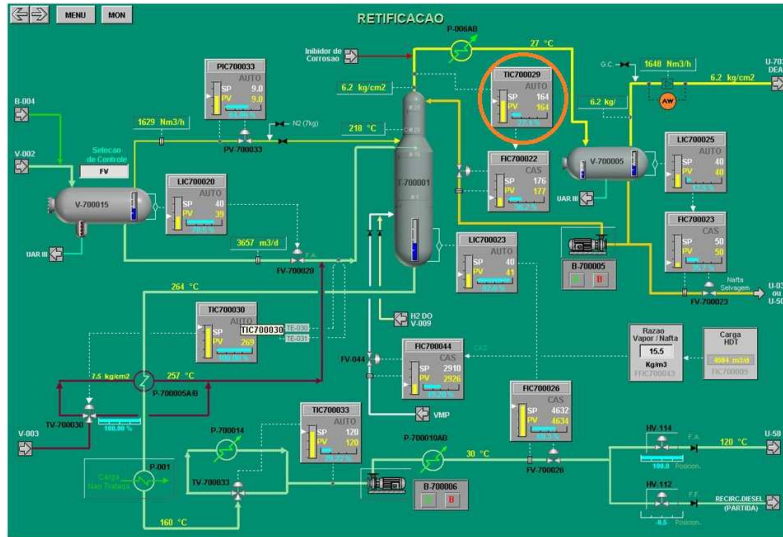


Figure 15: The DCS screen for the HDT rectification sector (TIC700029 is circled).

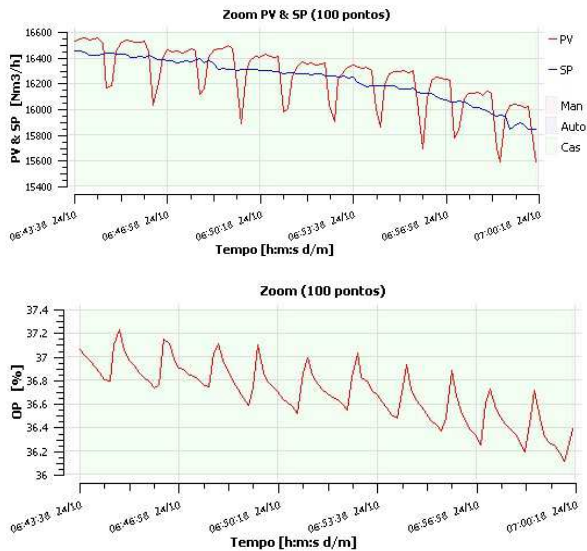


Figure 10: FIC700012 - PV-SP and CO dynamical plots for a 20 minutes time period.

In the next section this controller will be referred to show the financial benefits generated by diminishing the tower top controller temperature.

## 5 ECONOMIC GAINS EVALUATION

In this section, it is presented the economic benefits experienced by the unit after finishing the project. Starting up it can be claimed that in the process control context, it does not make sense to talk about control but a control (or manage-

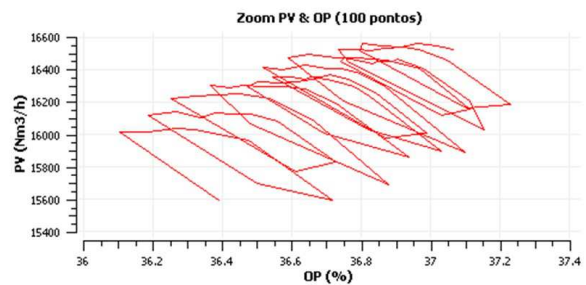


Figure 11: FIC700012 - PV-CO diagram for the same period as figure 10.

ment) hierarchy or pyramid. Despite the bigger (also riskier and longer term evaluated) gains being achieved in the higher level of that hierarchy (such as strategic planning and production scheduling) a well working regulatory control (the hierarchy lowest level) is necessary to guarantee the correct following of the orders from the highest levels. Then, the optimization of the regulatory control level (and also the communication and coordination among the levels) is fundamental to a safety and profitable plant operation. However, despite what the establishment says about the regulatory control layer economic relevance, there are certain objectives for this control level that never depend on the upper level decisions. These can generate high economical benefits (or losses) by itself.

Profit is the main objective of every plant. There are some exceptions, for instance, an effluent treatment plant. However, even for these plants, after attaining its fundamental purpose, the profit becomes the objective to be maximized (or mini-

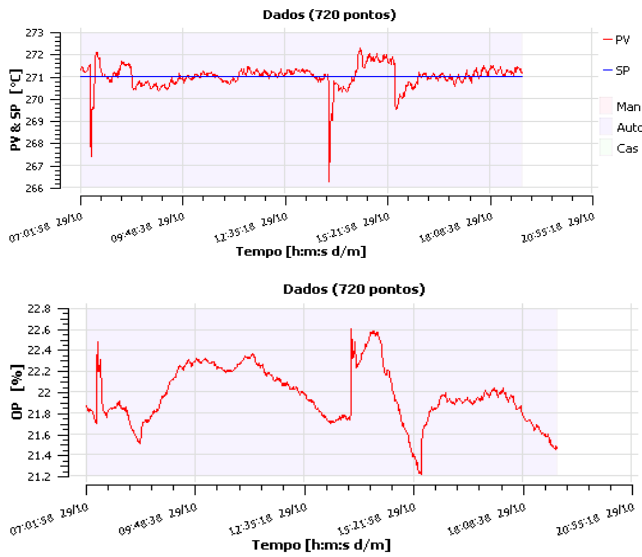


Figure 14: Illustrative PV-SP dynamical plot for TIC700026 after the corrective actions.

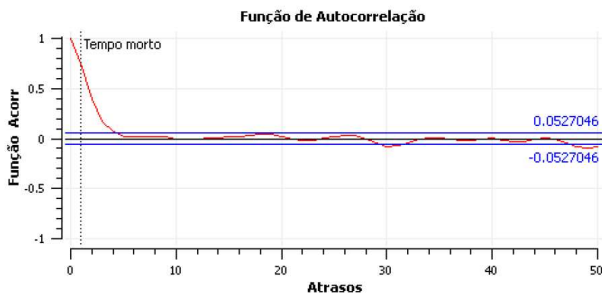


Figure 16: The ACF diagram for TIC700029 after corrective actions.

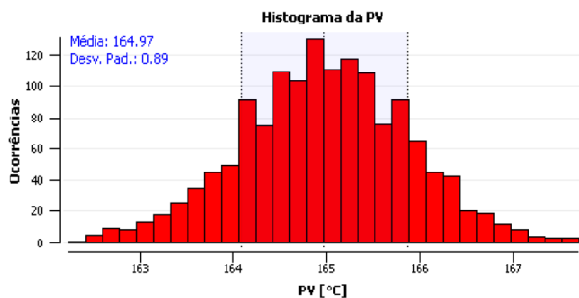


Figure 17: The PV histogram for TIC700029 after the HDT corrective actions.

mized, in case of losses). The plant profit is defined as the

difference between the sum of all revenues and the sum of all costs involved in the process (including the taxes):

$$\text{PROFIT} = \text{REVENUE} - \text{COSTS} \quad (3)$$

Finding the real profit or loss of a control action or improvement is a very hard task. However, if the intention is only to compute or estimate the gain, as in equation 4, the task becomes easier. This is because the reference for comparison is known (the baseline) and the same uncertain fixed costs are cut off. So, it is only needed to compute the net credits and debits for the difference from this baseline to the current situation (after the actions done):

$$\text{GAIN} = \begin{cases} \text{PROFIT}_{\text{after}} - \text{PROFIT}_{\text{before}} = \\ (\text{REVENUE}_{\text{after}} - \text{REVENUE}_{\text{before}}) + \\ (\text{COSTS}_{\text{before}} - \text{COSTS}_{\text{after}}) \end{cases} \quad (4)$$

In order to compute the economic benefits or gain, according to equation 4, it can be discarded the fixed credits and debits: depreciation, fixed overhead costs, subsidies, among others. Ignoring these factors, the process operation and operation may be represented as in figure 18, in which the main inputs to the process are shown on the left and the outputs on the right. Each is associated with a credit or debit term which influences the profit function, which must be kept as high as possible. The influences of constraints, and the process relationship which interrelate the inputs and the outputs, are also shown (Lowe and Hidden, 1971).

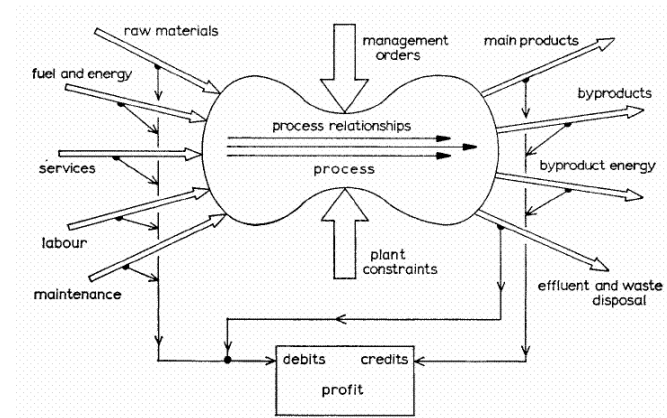


Figure 18: Relationships between process, profit and constraints, according to Lowe and Hidden (1971).

Some variable costs from figure 18 are usually not applied in case of regulatory control improvements, for example: The raw materials transportation and the variable overheads.

Other may not be tangibly computed, as the maintenance reduction and the labor spent to improve or manually correct the bad working control loops. Disregarding these revenues and costs, it can be depicted a table with the basic gains to be investigated after a CLPA and improvement project (table 4).

Table 4: Basic mechanisms for profit improvement in CLPA and improvement projects.

SOURCE	CLASSIFICATION
Raw material replacement for lower cost alternatives	CREDIT
Enlargement of the unit load (capacity)	CREDIT
Fuel consumption	DEBIT
Energy efficiency usage	DEBIT
Losses costs	DEBIT
Higher product selectivity	CREDIT
Product quality improvement	CREDIT <sup>a</sup>
Sale of new byproducts and/or energy	CREDIT
Effluent and emissions treatment and residuals disposal	DEBIT
Transportation costs	DEBIT - hard to compute <sup>b</sup>
Maintenance costs	DEBIT - hard to compute
Labour and services hired to solve control problems	DEBIT - hard to compute
Effluent and emissions reductions	CREDIT - Intangible <sup>c</sup>
Frequency reduction in the scheduled and/or not planned plant shutdown	CREDIT - hard to compute

<sup>a</sup> This is not true if the quality is not more valued after reaching a fixed specification.

<sup>b</sup> “hard to compute” due to the uncertainties involved in the computations.

<sup>c</sup> Exception is done if there is CER (Certified Emission Reduction) revenue or other similar emission reduction credit.

Considering the mechanisms organized in table 4, it was investigated where some of them could be applied to the RE-FAP HDT CLPA and improvement project. The result of the preliminary evaluation is presented in table 5. These results were used to select the point needing to be more carefully focused on.

Now, only the main opportunities are focused on and the estimated gain for each one is presented.

## 5.1 Reduction in the Hydrogen gas burning to control its production at HGU

The overall stabilization of the HDT, associated with the HGU stabilization, has lead to much lesser openings to atmosphere in the HGU control valves (Longhi et al., 2009). This gain was computed comparing the baseline openings and the current situation. The prices won't be presented here but they were computed using the calorific power of the relieved gas and the corresponding need to compensate them with natural gas usage (these gases would be used to generate hydrogen if not opened to flare). This gain is not a so high one and was estimated about R\$ 6,500.00<sup>1</sup> per month.

## 5.2 Augmentation of LCO load content and furnace fuel reduction

After the corrective actions, the temperature of the separation vessel (TIC700026) had its variability, characterized as the standard deviation, reduce from 2.9 C (baseline condition) to less than 1.0 C. This can be viewed in figure 19. This leads to a move in the vessel operation point of about 6 C. According to the process analysis calculations, this can lead to an augmentation in the LCO load content about 150 m3 per day. This is possible because there is an excess of LCO that has to be sold as less valued products (fuel oil) if it doesn't pass through the HDT unit. The benefit for this opportunity is evaluated using the price difference of the two alternatives revenues from selling LCO plus the extra usage of hydrogen gas caused by a bigger LCO processing. Using the prices from early 2009 it was arrived to a gain computed as R\$ 1,128,000.00 per month. This gain was completely accomplished at the end of 2009.

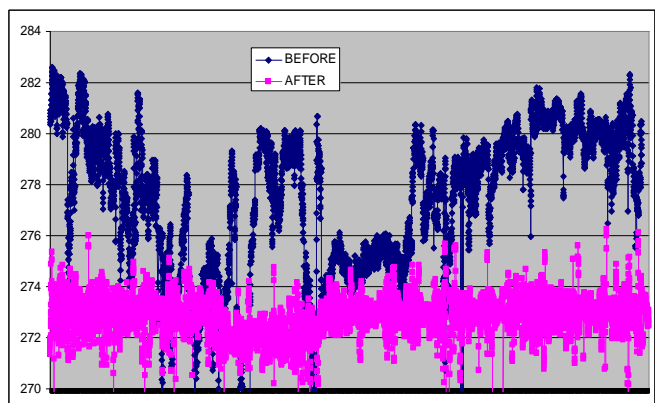


Figure 19: Five minutes sampled data for TIC700026 before and after the corrective actions.

<sup>1</sup>R\$ means BRD (Real, the Brazilian currency). Its average exchange rate for 2009 was 1 USD = 1.998 BRD (source: BOVESPA, the main Brazilian stock market).

Table 5: Preliminary analysis of the HDT unit gains.

SOURCE	STAFF SENTENCE
Raw material replacement for lower cost alternatives	→ After the temperature stabilization of V702002 it is possible to augment the LCO (Light Cycle Oil) content in the diesel pool. It should be considered in section 5.2.
Enlargement of the unit load (capacity)	→ The unit vibration reduction could lead to avoiding eventual load reductions used for control the unit stability. After analyzing the plant database, this opportunity was not detected.
Frequency reduction in the scheduled and/or not planned plant shutdown	→ It occurs, but it is very hard to compute. It will be presented only qualitatively in section 5.3.
Fuel Consumption	→ It is reduced due to the higher percentage of LCO in the load composition. This component is more exothermic than the others and leads to a lesser consumption in the furnace fuel. It will be considered at the end of section 5.2.
Losses reduction	→ The HDT and HGU stabilization reduces the disturbances in the hydrogen consumption and production, respectively. This leads to a smaller necessity to open relief valves to the atmosphere. It will be computed as presented in section 5.1.
Better selectivity	→ In the rectification sector, due to a better control in the sector tower, it can take place a better separation between wild naphtha and diesel. It will be computed, too (see section 5.4).
Product quality improvement	→ It is possible to happen for the diesel sulfur content. But the monitoring routine is inadequate for this. It should be discarded! The refinery was advised to improve the monitoring routine for the HDT unit.

The usage of bigger LCO percentage in the HDT load, due to its higher exothermicity, also allows a reduction in the load heating furnace fuel. This value was estimated, for the same temperature increment, using typical energy calculations, as R\$ 17,000.00 per month. Both gains from this section are based on the assumption that the only constraint for LCO augmentation in the load is the furnace minimum temperature difference. This is true most of the times, but not always.

### 5.3 Frequency reduction of passing corrosive water in V700003 and probable less plant shutdown due to staving in heat exchangers

The improvement in the unit control has been useful to diminish the occurrence frequency of passing corrosive water from the low temperature separating drum (V700003) to the double heat exchanger P700005. These heat exchangers had already been the reason for a plant shutdown in the past.

The gain related to this improvement is difficult to quantify. So, it'll be tried to illustrate it without any rigorous economic computation. To show the improvements, it is shown in fig-

ure 20, the dynamical evolution for instruments PI700034 (thin line) and TI700154 (bold line) for two similar time periods, before (upper) and after (lower) the corrective actions. When there is corrosive water passing through the heat exchanger, it can be noticed peaks in these instruments. So, as both graphs have the same scale it can be clearly seen that the passing has greatly diminished after the corrective period. To corroborate this qualitative observation, at 2008 December, the bundle of the exchanger was removed for evaluation and it was not reported any significant corrosion. As a consequence of that, the demand for its exchanging was suppressed. The material exchanging was rated as about R\$ 80,000.00

### 5.4 Diesel production augmentation by incorporating wild naphtha at T700001

The variability reduction in TIC700029, rectifying tower top temperature, allows reducing its set-point aimed at the diesel maximization and reducing the wild naphtha production. As the Standard deviation was diminished from 3.7 to 1.6 Celsius degrees, it is possible to safely reduce, according our



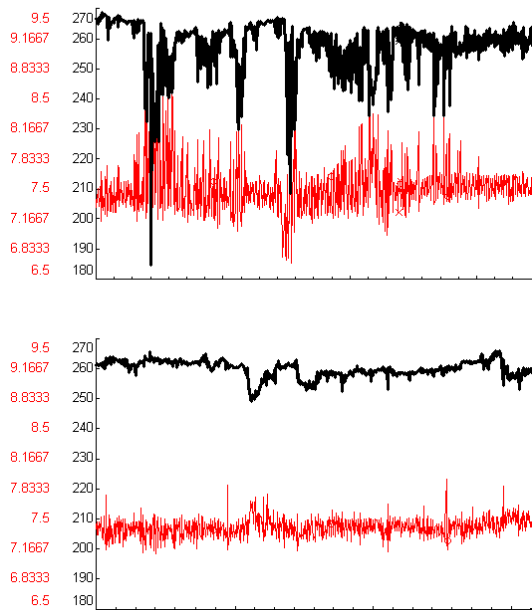


Figure 20: PI700027 (upper) and TI700154 (lower) for the baseline and current situation (early 2009).

process analysis calculations, 2 degrees in the top temperature set-point. Accounting of the prices of early 2009 for the exported naphtha and diesel in our internal reports, the benefit of this opportunity was estimated as R\$ 19.394,00 per month.

## 6 CONCLUSIONS

This work presents a project that applies new academic developments to solve practical industrial problems. The critical factors to its successful implementation are: IT infrastructure, reliable tool choice and, finally, staff and management commitment. It was shown the work methodology and the results achieved until now. It was also presented a non-conventional economic evaluation of the benefits generated by the project. This is a hard task to do, but it must be done to justify and motivate this kind of innovative work for the enterprise executive staff.

## ACKNOWLEDGEMENTS

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