Head Pressure minimization of a Visbreaking column through an advanced PID controllers architecture

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Abstract: In this paper an advanced PID control architecture developed to minimize the head pressure of a visbreaking column is presented. The implemented control strategy has been designed through an advanced interconnection of *PID* controllers typically used in industrial processes: cascade control, feedforward control and override control. In order to correctly identify the interactions of the key variables, a RGA analysis has been performed. The proposed control system has been previously tested on simulation, thus been able to evaluate its performances in terms of robustness, stability and rejection of the main disturbances of the process. The results on the plant of the column optimization confirmed its effectiveness in the minimization of the head pressure. In this way, the new controller guaranties an increase of the separation between the heavy and light components and an increase of the extraction of the valuable products like gasoline and Light GasOil. The application of the proposed control architecture has allowed reaching an important economic recovery.

Keywords: Distillation, PID, Model Identification, Relative Gain Array (RGA) analysis, cascade control, override control.

1. INTRODUCTION

Visbreaking is a non-catalytic thermal process that converts atmospheric or vacuum residues via thermal cracking to gas, naphtha, distillates, and visbroken residue. Atmospheric and vacuum residues are typically charged to a visbreaker to reduce fuel oil viscosity and increase distillate yield in the refiner (e.g. Maples R.E 2000, Speight J.G 2006).
The process name of "visbreaking" refers to the fact that the process reduces (i.e. breaks) the viscosity of the residual oil. The objectives of visbreaking are to:

- *Reduce the viscosity of the feed stream*: typically this is the residue from either the refinery's atmospheric or vacuum distillation of crude oil but can also be used to reduce the viscosity of oils produced in the processing of tar sands, certain high viscosity crude oils and other high viscosity oils.
- *Reduce the amount of endproduct residual fuel oil produced by a refinery*: Residual fuel oil is generally regarded as a low value product.
- *Increase the proportion of middle distillates in the refinery output*: middle distillates are used as a diluent with residual oils to bring their viscosity down to a marketable level. By reducing the viscosity of the residual stream in a visbreaker, a fuel oil can be made using less diluent and the middle distillate saved can be diverted to higher value diesel or heating oil manufacture.

Basically, a visbreaker thermally cracks large hydrocarbon molecules into smaller molecules by heating residuals oil thus reducing its viscosity. Typically the feed stream is the residue from either the refinery's atmospheric or vacuum distillation of crude oil but can also be oils produced in the processing of tar sands, certain high viscosity crude oils and other high viscosity oils. Visbreaker produces naphtha and fuel oils as well as refinery fuel gas (see Figure 1).

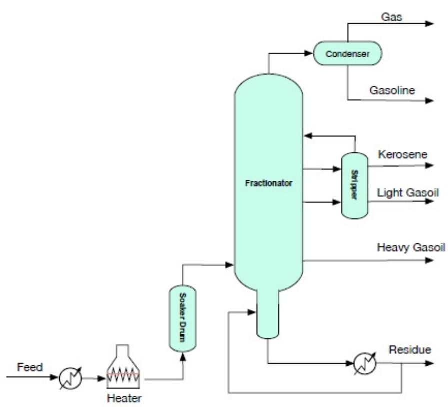

Fig. 1: Scheme of the visbreaking process

There are two types of visbreaking technology that are commercially available: the 'coil' or 'furnace' type and the 'soaker' process. In the coil process, conversion is achieved by high temperature cracking for a predetermined, relatively short period of time in the heater. In the soaker process, which is a low temperature/high residence time process, the majority of conversion occurs in a reaction vessel or soaker drum, where the two-phase heater effluent is held at a lower temperature for a longer period of time.
In this work a soaker visbreaking column of a refinement plant has been considered. The scope of this unit is the production of gas, gasoline, kerosene, Light GasOil (*LGO*)

and Heavy GasOil (*HGO*). For each product different draw-off trays are employed; in order to improve the product separation inside the column a portion of the total stream of the products extracted is returned to the column as reflux, whereas the rest leaves the column and is sent to the downstream units.

In the present paper an advanced PID architecture that guarantees head pressure minimization is presented. The minimization of the head pressure allows to increase the separation between the heavy and light components and to increase the extraction of the valuable products like gasoline and Light GasOil (*LGO*). A formal approach based on an initial identification phase and on a subsequent Relative Gain Array RGA analysis has been adopted thus been able to prove a substantial design error in variable coupling of the formerly implemented control logic.

2. PROBLEM DEFINITION

The first objective of the visbreaking column under study is the regulation of the final point of the gasoline distillation curve (see Figure 2).

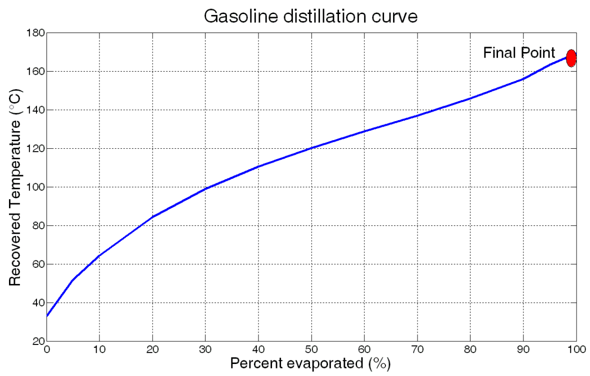

Fig. 2: Gasoline distillation curve

In order to achieve this objective the control system that was implemented at the time this research study was started was designed according to the architecture in Figure 3. The final point of the gasoline stream was controlled by a cascade control: the master controller been a temperature controller (*CTC18997* – which set-point is the final point) and the slave controller been a gasoline reflux controller (*FC18999*). The stream of the kerosene was controlled by a single flow controller (*FC18998*). The scope of the cooler is to decrease the temperature of the kerosene stream mixture with the gasoline stream at the entrance of the head of the column. In the old scheme the cooler was controlled by a temperature controller CTC1803 in a split range configuration on the two valves used in parallel. The slope of the air fin was controlled manually by a *hand controller HC1899*.

The main disadvantage of this control architecture was the high value of the head pressure; the direct effect was an increase of the temperature profile of the entire column and consequently a difficult separation between the light and heavy components. Furthermore, since the final points of each of the components extracted from the visbreaking column are affected by pressure and temperature profile, it resulted very hard to maintain the components distribution curves at their target. Consequently, some light components were extracted at the bottom of the column and impurities (heavy components) were extracted along with the top products. In particular, the analysis revealed that the impurities concerning the heavy components of the Heavy GasOil (*HGO*) were extracted at the same tray where the Light GasOil (*LGO*) (a high economic value product) was extracted this representing a not tolerable giveaway.

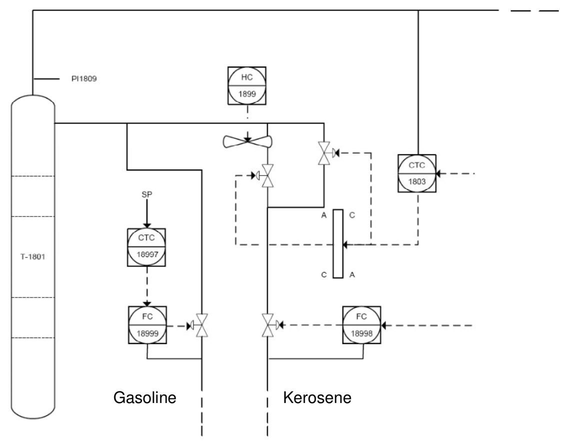

Fig. 3: Old Control Strategy: control of the gasoline final point

Basically the existing control strategies aimed at maintaining constant the quality of the products (output temperature of the column), avoiding excessive fluctuations in the quantities of products (even at the expense of constant quality), maintaining liquid levels within limits and maintaining the internal column load within constraints (draw off – reflux ratio) whereas no particular attention was given at the maximization of system efficiency. The last point has motivated the present work. The proposed architecture allows fulfilling the above specifications while attaining a greater control efficiency, that is, the maximization of valuable products extraction, with important improvements in terms of performances and economic returns.

The first step in the design of the new control architecture is the identification of the head pressure model and the gasoline final point model of the column; in this way it is possible to evaluate the effects of the two refluxes on the gasoline final point target and on the head pressure.

3. MODEL IDENTIFICATION

The target of the new control system strategy is to control the gasoline final point and to minimize the head pressure. The visbreaking column is a critical plant and any modification of its setting must be carefully investigated before its actual implementation. This has motivated the development of a multivariable model, capable to describe the most important dynamics of the process under study, so to evaluate the effects of the new control architecture and its possible improvements in an offline environment. As it is well presented in literature (e.g. Maciejowski J.M. 2002, Zhu Y.C., Backx T 1993) the dynamics of most of the industrial processes can be suitably described by a first order transfer function with delay. The generic transfer function has the following structure:

$$g_i(s) = \frac{K_i}{1+\tau_i s} e^{-T_{d_i} s}$$

where K_i is the process gain, τ_i is the process time constant and T_{d_i} is the process delay.

3.1 Head Pressure Identification

The head column pressure can be well predicted from the two streams which are recirculated in the head of the column (gasoline reflux -*FC18999*- and kerosene reflux -*FC18998*-). After a step test performed on the plant, a MISO model for the prediction of the head pressure has been developed. In Figure 4 measured and predicted values of the head pressure are shown; taking into account resolution and sensibility of the sensor the model can be stated to be in a good agreement with the system.

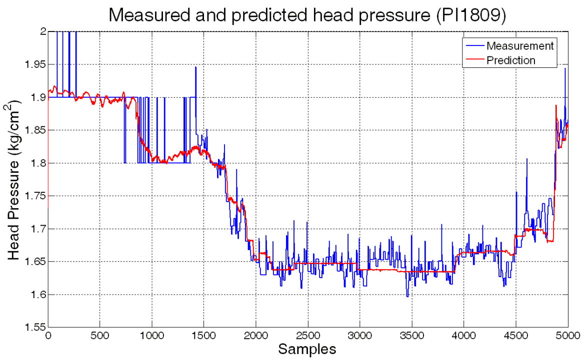

Fig. 4: Head Pressure identification (PI18809)

3.2 Gasoline Final Point Identification

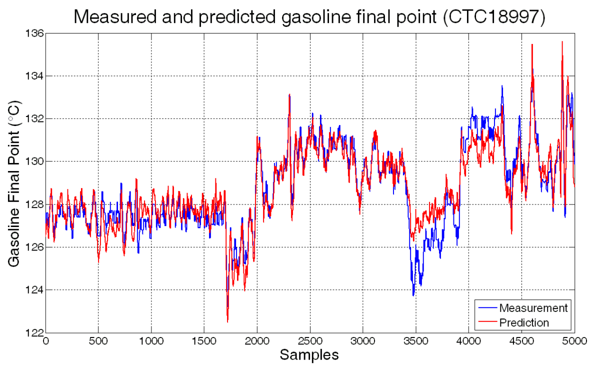

Fig. 5: Gasoline final point identification (CTC18997)

The gasoline final point is influenced by the flows of the recirculated streams and their temperatures. Thus to accurately estimate the final point of the gasoline distillation curve it is necessary to introduce in the model the contribution of the cooler used to cool the kerosene reflux stream (out of the temperature controller *CTC1803*). The results relative to the prediction of the gasoline final point are shown in Figure 5.

3.3 Pairing Manipulated and Controlled Variables

After process models identification, the problem of the paring between Manipulated and Controlled Variables (MV/CV) has been considered. To solve this problem a Relative Gain Array (RGA) approach has been adopted (e.g Shinskey F.G 1996, Luyben W.L., Luyben M.L 1997).
Relative gain analysis is a widely used technique in control system design for multivariable processes. The analysis is based on a Relative Gain Array which measures the interactions of all possible single-input single-output (SISO) pairings of the considered system variables. The RGA thus indicates the preferable variable pairings in a decentralized (multi-loop SISO) control system based on interaction considerations. The steady-state relative gain array Λ of the system G(s) is defined

$$\Lambda(G) = G \circ G^{-T}$$

where "$\circ$" denotes the element-by-element product. For the considered system, G and Λ are in the form:

$$G = \begin{bmatrix} g_{11} & g_{12} \\ g_{21} & g_{22} \end{bmatrix}, \quad \Lambda = \begin{bmatrix} \lambda_{11} & \lambda_{12} \\ \lambda_{21} & \lambda_{22} \end{bmatrix}.$$

The inspection of the λ_{ij} elements gives useful insights about interactions of the control loops. From literature (e.g. Skogestad S., Postlethwaite I. 2005) some RGA properties can be summarized as follows:

1. $\lambda_{ij} = 1$, no interaction between loops other than i,j;
2. $\lambda_{ij} = 0$, manipulated input i, has no effect on output j.
3. $\lambda_{ij} = 0.5$, high degree of interaction.
4. $0.5 < \lambda_{ij} < 1$, interaction between the control loops, preferable pairing.
5. $\lambda_{ij} < 0$, negative off-diagonal elements are critical: they indicate a change of the sign of the gain and related possible instability problems.

3.4 Case study

In the multivariable system under study it is necessary to establish what is the best paring between the two manipulated variables, gasoline reflux stream (*FC18999_PV*) and kerosene reflux stream (*FC18998_PV*), and the two controlled variable, head pressure (*PI1809*) and gasoline final point (*TI18997*). In Table 1 the values of the relative gain matrix (Λ) are shown.

Manipulated Variable (MV)	**Controlled Variable (CV)**	
	TI18997	PI1809
FC18998_PV	**0,74**	0,26
FC18999_PV	0,26	**0,74**

Table. 1: Relative Gain Array matrix

Considering RGA properties (see previous section) the optimal paring between Controlled Variables (*CV*) and Manipulated Variables (*MV*) is the one that relates the head pressure (*PI1809*) with the gasoline reflux and the gasoline final point (*CTC18997*) with the kerosene reflux. This result highlights a not optimal pairing of the previously implemented controller. In fact, as it can be observed from Figure 3 the temperature (TI18997) acts on the gasoline reflux (FC18999_PV) while the head pressure (PI1809) is controlled by kerosene reflux (FC18998_PV).

4. NEW CONTROL STRATEGY

The purpose of the control, as mentioned in section 2, is to stabilize the column's final point and to reduce the column head pressure at a standard target. The proposed control strategy has been developed by an advanced architecture of PID controllers (e.g. Astrom K.J., Hagglung T. 1995). The new control adopts the variables pairing as suggested by the RGA analysis. The first step has concerned the introduction of a cascade control (e.g. Morari M., Zafiriou E 1989; Seborg D.E., Edgar T.F., Mellichamp D.A. 1989) between the temperature controller relative to the gasoline final point (CTC18997) and the flow controller relative to the kerosene stream (FC18998). In order to minimize the head pressure of the column it is necessary to minimize the stream of the gasoline reflux. To achieve this objective a new valve position controller (VPC18999A) (e.g. Stephanopoulus G. 1995) has been introduced as master with a very slow tuning in a cascade control architecture and the existing flow controller (FC18999) as a slave. The controlled variable of the VPC is the position of valve used to regulate the kerosene stream; the output is the set point (SP) of the gasoline stream. In order to maintain a low flow in the gasoline stream it is necessary to set a high SP on the new VPC1999A (up to 75-80%). As a side effect of the minimization of the gasoline stream, an unfavourable increment of the temperature at the head of the column was expected. In order to early compensate this lack of the cooler stream, a feedforward controller (XC18997) has been introduced; the compensation used to compute the setpoint of the kerosene flow controller takes into account the gasoline flow and the heat exchanged by the gasoline stream and kerosene stream as in the following:

$$FC18998_{SP} = \left(\frac{C_{P\ \text{gasoline}} \cdot FC18999 \cdot (TI1824 - TI1879)}{C_{P\ \text{kerosene}} \cdot (TI18889 - TI18999)} \right)$$

where $C_{P\text{gasoline}}$ and $C_{P\text{kerosene}}$ are the specific heats at constant pressure and the TIs are the temperatures of the products along the transfer.
The new control system architecture also provides to exploit at maximum the use of the cooler employed along the kerosene pipe; to achieve this target a further valve position controller has been designed (VPC18999B). The CV of this controller is the position of the valve regulated by the new temperature controller (TC18999) which has been introduced to control the temperature of the kerosene stream. The output of this VPC is the SP of the temperature controller; in order to maximize the use of the cooler it is required to set a high setpoint. Finally to avoid a very low temperature (manipulated with TC18999) at the head of the column an override control strategy (e.g. Coughanowr D.R. 1991) has been introduced: the logic selects the maximum value between the output of the new valve position control (VP*C18999B*), which tends to reduce the SP of TC18999, and the output of the new temperature controller (T*C18998*) that controls the temperature of the kerosene and gasoline mix streams. The resulting final control system is reported in the following Figure 6.

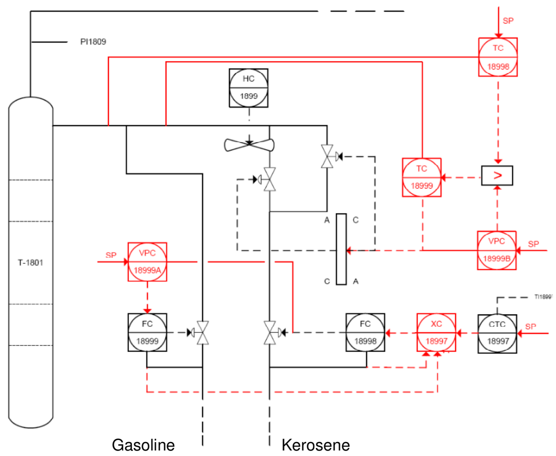

Fig. 6: New control strategy: control of the gasoline final point and minimization of the head pressure.

4.1 PIDs Tuning

PID tuning was realized off-line using models indentified in section 3. Different tuning methods (e.g. Anirudda D. 2000; Tan Kok Kiong 1999) have been considered which ensure different performance indexes. Controller tuning methods provide the controller parameters in the form of formulae or algorithms. The implemented methods are: Ziegler-Nichols, Cohen e Coon, Internal Model Control, ITAE (Integral of Time Absolute Error), gain and phase margin optimization, Pole Placement.

	Kp	Ti [sec]	Td [sec]
Ziegler Nichols	3,354	20,000	5,000
Cohen Coon	3,904	22,333	3,478
Int. Mod. Control	2,861	45,000	4,444
ITAE	2,208	52,684	3,399
Gain & Phase Margin	1,630	41,372	0,904
Pole Placement	0,632	19,550	0,000

Table. 2: Tuning method parameters

These methods require knowledge about the controlled process, in particular the process Gain (K_i), the lag time (τ_i) and the delay time (T_{di}).
Tuning results presented in this section are related to the flow controller of the kerosene reflux FC18999 (process parameters are $K_i = 1.43$, $T_{di} = 13$ s, $\tau_i = 41$ s). The computed tuning parameters are shown in Table 2 while the step responses of the controlled system as resulting from the different tunings are depicted in Figure 7. Cohen Coon and Ziegler Nichols settings give the greatest rise time, overshoot and oscillatory response than the other methods. In some cases the overshoot is greater than 50 percent.
The responses obtained with the ITAE, gain and phase margin and pole placement methods show rise times of about 15 seconds which are higher than the previous methods rise times but still acceptable. On the other hand overshoot is considerably decreased and robustness is guaranteed by the

adopted algorithms. Moreover lower control efforts are requested by these last three methods (see Figure 8).

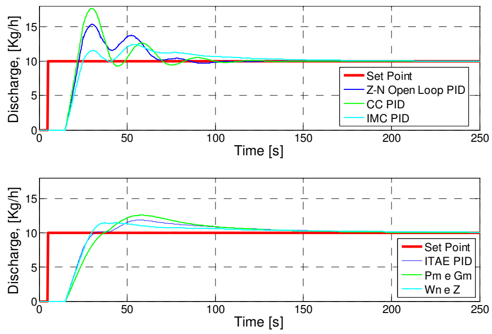

Fig. 7: Comparison between the proposed control methods

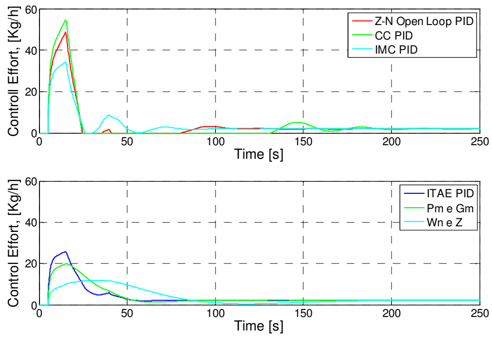

Fig. 8: Control effort.

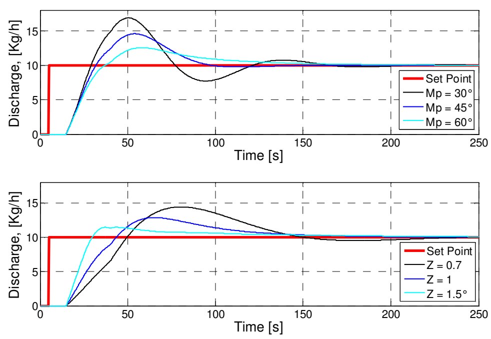

Fig. 9: Different tuning values for Gain & phase margin method (upper figure) and pole placement method (lower figure).

Figure 9 shows the system behavior at different tuning values for gain-phase margin and pole placement methods. The Bode diagram of the open loop transfer function is shown in Figure 10 for the *Gain & Phase Margins* (*GPM*) considered tuning: it's possible to note that the phase margin and gain margin are respectively 60° and 3 dB as it was required. Finally the performances of a PID architecture comprehensive of a setpoint weight (SPW) have been considered and the response behavior for all the considered tuning methods has been analyzed. As a general rule, this PID implementation allows an overshoot reduction without an evident decreasing of the rise time. As far as concerns the problem of the overshoot reduction, an implementation of a fuzzy PID architecture (e.g. Zanoli S.M., Conte G. 2003) is under development.

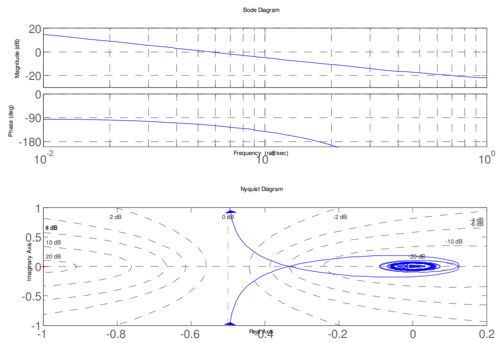

Fig. 10: Bode and Nyquist diagram.

5. RESULTS

The method used to tune the parameters in control system is the *Gain & Phase Margins* method characterized by a limited overshot and by an acceptable rise time while allowing the enforcing of desired robustness indexes.
The new control strategy was implemented in an Emerson DCS and the PID was realized in parallel configuration.
In the following pictures the results of the proposed control strategy are presented. Figure 11 shows the decreased values achieved in the head pressure of the column under study.

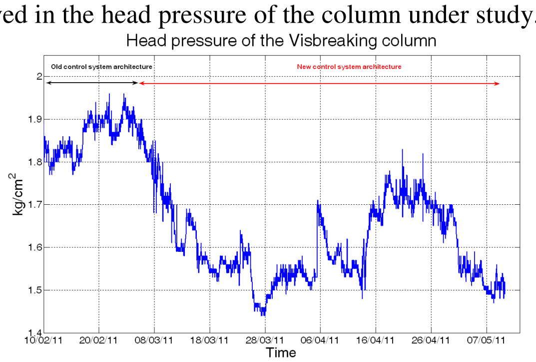

Fig. 11: Trend of the head pressure of the visbreaking column (PI1809) before and after the implementation of new control

As it can be noticed the head pressure has been decreased of about 0.3 kg/cm^2. This has been achieved decreasing the gasoline reflux (see Figure 12). The most valuable benefit, in term of economic recovery, concerns the improvement in the extraction of valuable products as Light GasOil (*LGO*). In fact, as it can be observed in Figure 13, owing to the implementation of the new control strategy, the percentage of the Heavy GasOil (*HGO*) recovered at 360 °C has been decreased of about four percentage points (from 17% to about 13%). The percentage recovered corresponds to an increment (of the same percentage) of the *LGO* flow extracted (with a high economic value). This result can also be assessed by the inspection of the *LGO* and *HGO* distillation curves. As it can be seen in the following figure (Figure 14), after the development of the new control strategy, the overlap between the *LGO* and *HGO* has been reduced of about 9 °C.

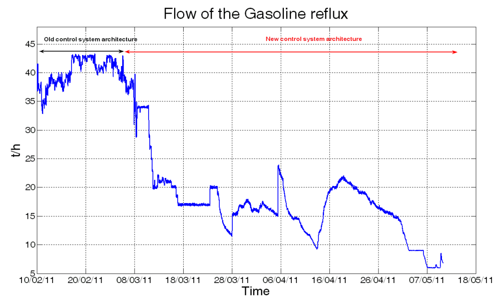

Fig. 11: Flow of the gasoline reflux (FC18999) before and after the implementation of the new control strategy

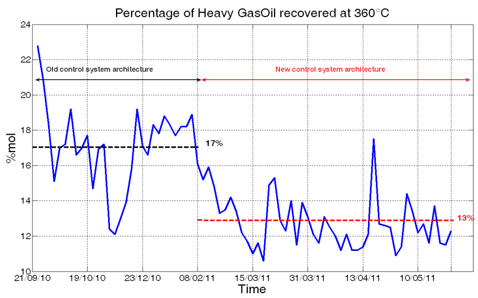

Fig. 13: Trend of the percentage of Heavy GasOil recovered at 360°C before and after the implementation of new control

Moreover a lower recirculation of heavy products in the upstream units has been achieved thus reducing the visbreaking process energy consumption. Taking into account different economic factors and process interactions, refinery managers have carried out a thorough analysis. From this analysis the economy return rate of the proposed control system has been estimated to be 2.5M $/year.

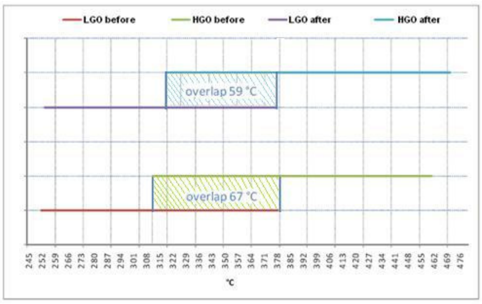

Fig.14: Overlap between the Light GasOil and Heavy GasOil before and after the implementation of the new control strategy

6. CONCLUSIONS

An advanced *PID* controller architecture design that guarantees the minimization of the head pressure of a visbreaking column is presented. Process models identification has been performed and the computation of a RGA analysis allowed detecting a not optimal variable pairing of former controller. New control solution has been proposed, analysed and tested in simulation. As a final step the proposed control system has been suitably translated for the implementation in the Distributed Control System used in the refinery plant and it is actually operative. The implementation of the proposed system confirmed the expected highly performing improvements: the pressure of the head column has been decreased of about 0.3 kg/cm^2 by means of the minimization of the stream of the gasoline reflux. The target of the gasoline final point has been achieved by maximization of the kerosene reflux and by maximization of the use of the cooler situated along the kerosene pipe. In fact, the major benefit of the minimization of the head column pressure concerns the significant decrease of the thermal profile. This allows to increase the separation between the heavy and light components and to increase the extraction of the gasoline and Light GasOil (*LGO*). The economic recovery introduced by the new control system has been evaluated of about 2.5M $/year.

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