AN IMPROVED PRODUCTS QUALITY CONTROL SCHEME FOR THE CRUDE OIL UNIT

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Abstract: This work presents an improved control structure for side-products of a classical crude oil unit. Taking its basic principles from a Shinskey idea, it offers two original solutions by implementing a control loops dynamic decoupler and changing from PID to robust feed-forward controllers (for level control in the main column and also in sidestripers). The proposed control scheme was extensively tested by simulation, very good results being obtained by following many operating scenarios, as one can see from the selected examples in this paper. Taking this fact into account, such an improved products quality structure may be subject for an industrial implementation.

Keywords: crude oil distillation, quality control, dynamic behavior.

1. INTRODUCTION

The crude oil unit, as part of the Atmospheric and Vacuum Distillation plant, is the first processing unit in a refinery. Due to its position (its products becoming either final products or feedstock for other processing units), its complexity and taking into account the highly energy-consuming process characteristic, it is very important having robust and efficient control schemes, in order to assure the high quality for products and reduce the energy consumption.

Obviously, the old-classical PID control loops cannot solve such a complex task, while the very efficient computer-based distributed systems may have a prohibitive price. Despite this deep dichotomy, excellent results could be obtained by using clever hybrid control structures at normal investment costs, as this work tries to prove.

2. RECONSIDERING THE CONTROL SCHEME STRUCTURE

As depicted in fig. 1, the crude oil plant considered as example in this paper consists in one main column with two pumparounds and having four sidedraws to the sidestripers; the top vapor is totally condensed and stored in a tank where the water is decanted; a part of the top product turns back into column as external reflux.

As shown in literature (Hsie, 1989; Marinoiu and Paraschiv, 1992; Pătrăşcioiu, *et al*., 1998; Muske, *et al*., 1991; Shinskey, 1988), a classical control scheme for the crude oil plant provides independent control loops for upper and lower pumparounds flowrates, bottom levels in the main column and sidestripers, steam to column, sidestripers and also products flowrates. But, taking into account the products quality specifications (as stiff or relaxed restrictions), such a structure works only if well-trained operating personnel, having good knowledge on the process behavior, assist it.

Related to the direct product quality control (concerning the *Start boiling Point* – SP and *End boiling Point* – EP control), it is revealed that any action on pumparounds flowrates leads to strong changes in the main column internal reflux profile; it has a global influence on products quality and cannot be used for independent SP and EP control.

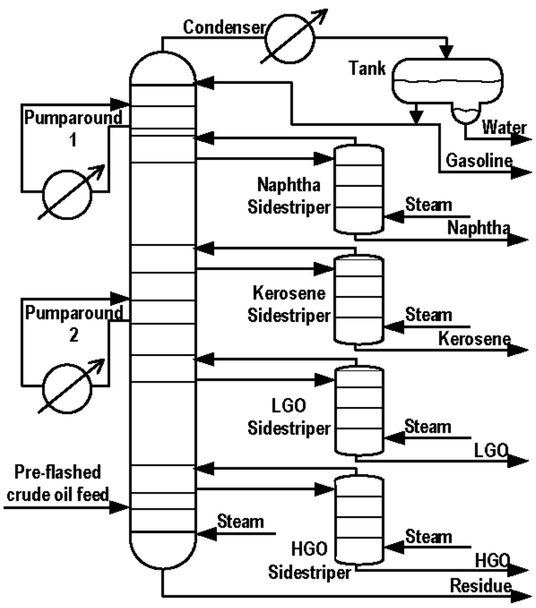

Fig. 1. The crude oil distillation plant.

To achieve this, a standard quality control scheme uses the (side) products flowrates to control each product EP and the steam (to sidestripers) flowrates to control each product SP, as known from literature (Hsie, 1989; Marinoiu and Paraschiv, 1992; Chung, *et al.* 1995; Pătrăşcioiu, *et al.*, 1998; Marinoiu, 2000) and practice.

These classical schemes bring not only a way to control the products quality, but also an undesirable bad dynamic behavior to the crude oil plant, due to the strong interactions among control loops, which in many cases lead to a very unstable system (Rădulescu, 2002).

Reconsidering these structures, two objectives were identified, speaking here about decoupling the control loops and finding improved solutions for level control (a major issue both for the main column and sidestrippers).

In order to answer to these announced objectives, the authors of this paper propose an EP control structure for side-products, based on a Shinskey idea, which overrides these difficulties by using an unidirectional decoupler (from column top to bottom) between EP control loops (Shinskey, 1984; Marinoiu, *et al.*, 1992; Pătrăşcioiu, *et al.*, 1998). As shown in fig. 2, the decoupler purpose is to eliminate the influence of a change in "p" product flowrate on "p+1", "p+2", ... products EP, its aim being to keep constant the column internal liquid reflux flowrate from the "p+1" product extraction tray to the column bottom. As shown in fig. 2, the analyzer-controller AC (PID-type EP controller for "p" product) output is the desired sum of lighter products 1,…, "p-1" flowrates and the "p" product itself flowrate ($\sum_{j=1}^{p} DL^j$). The DL^p flowrate is obtained as subtraction between the EP controller output (for "p" product) and the time-delayed sum of lighter products flowrates $\sum_{j=1}^{p-1}\left[f_{j,p}\left(t, DL_j\right)\right]$.

An original structure of "delaying channels" (1→p),(2→p)…(p-1→p) represent the difference between this proposed control scheme and those suggested by Shinskey (1984) and improved by Pătrăşcioiu, *et al.* (1998). The first one makes use of a global (unique) delaying function f (which in our opinion may be not reflecting with a high accuracy the real process behavior under influence of its inner hydraulic regime). The second one, apart from improving some weak aspects in the Shinskey's scheme, does not take into account any process dynamics.

This proposed structure, using 2nd order delaying elements for $f_{j,p}(t)$, implements a dynamic decoupler if correct time constants are determined through identification techniques for a particular application (Rădulescu, 2002) and may be regarded as a very basic component of a model-based controller.

Another original system feature is the solution adopted by authors for level control.

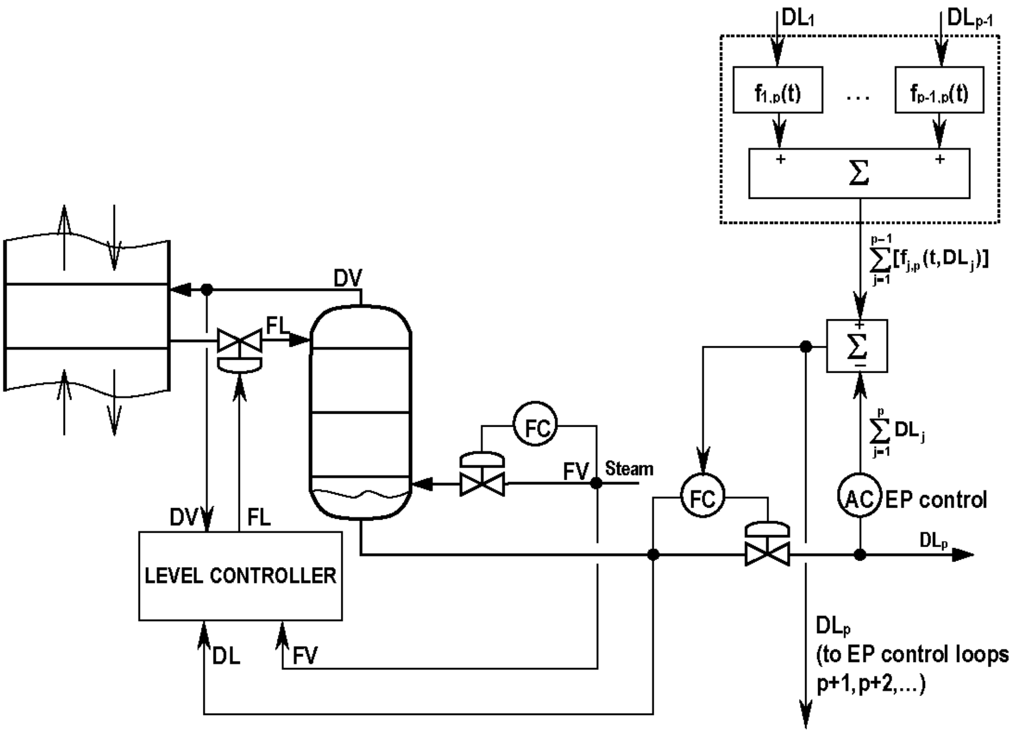

Fig. 2. The proposed control structure for the side-product "p".

Due to the process dynamics, especially the big delays (typical characteristic of the hydraulic regime in the columns with trays), in their studies the authors met difficulties in using standard feedback control schemes with PID controllers (a fact also revealed from the practical experience too), which may lead to a very sensitive and stiff system behavior (Hsie, 1989; Marinoiu, *et al.*, 1992; Marinoiu, 2000).

Instead of this classical solution, a very robust and efficient feed-forward control scheme was tested. The control algorithm makes a total material balance equation to be satisfied at each particular moment. As example, in the specific case of the sidestripers (fig. 3), where the bottom level is controlled by the liquid sidedraw from the main column (FL), the mathematical model for the level controller is:

$$FL = DL + DV - FV. \quad (1)$$

In practice, FV, DL and DV would be the measured variables and FL the manipulated variable (controller output), as shown in fig. 3. One can observe that equation (1) represents in fact the total material balance equation for the sidestriper.

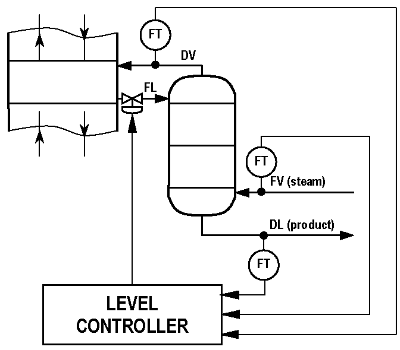

Fig. 3. The original level control scheme.

This solution may be applied even in the case of the main column, where the bottom level is controlled by the residue flowrate. The authors experimented there a slightly different equation, representing the total material balance for the last tray instead of the balance for the whole column. In both cases, this represents a "perfect" level controller (Rădulescu, 2002).

As remark, due to the very different dynamics in the column, the authors did not observe any negative influence when simulating the system equipped with these non-standard level controllers; furthermore, it brings more robustness for the plant.

3. SIMULATION TESTS

In order to evaluate the proposed structure behavior and performances, the authors performed some extensive simulation scenarios, using DIVA software environment (Mangold, *et al*., 2000), but – due to the lack of space – only a significant set of results is included in this paper. With this aim, a crude oil unit with the structure as in fig. 1 was considered; some other relevant data are presented in table1.

Table 1. Some data about the unit.

Number of trays in the main column	30
Number of trays in the main sidestriper	3
Feed type	Pre-flashed
Feed flowrate	0.57 kmol/s
Reflux ratio	0.23
Molar holdup on column trays	3.0 kmol
Molar holdup on sidestriper trays	3.0 kmol

3.1. The column open-loop response

In order to obtain the open-loop responses (as image on the intrinsic process characteristics), only the level controllers were left in service; that means any other system inputs (except the flowrates of the bottom residue for the main column and sidedraws to sidestripers) may be subject to changes. Also, no decoupling structure is present.

As example, fig. 4 shows the products EP's evolution when naphtha flowrate increases with 10%. To avoid decreasing the bottom level in naphtha sidestriper, the corresponding controller corrects its output – this way it increases the column liquid sidedraw flowrate. But in the same time it leads to an internal liquid reflux rate decreasing from the naphtha tray to the column bottom, which permits to heavier components to increase their mole fractions in the liquid phase.

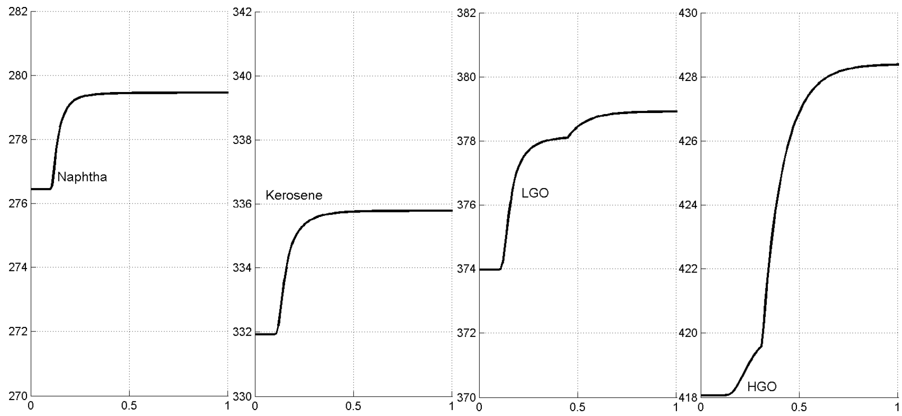

Fig. 4. Plant open-loop response for a 10% increase in naphtha flowrate. Product EP's, in [^{o}C], are depicted. The simulation time is expressed in [seconds × 10^4].

This affects the naphtha, kerosene, LGO (Light Gas Oil) and HGO (Heavy Gas Oil) EP's, as seen in fig. 4, by significantly increasing it (as known from practice too). As remark, the authors preferred to use absolute values instead deviations for EP's, to take into account the physical sense of the simulation results. They are in good agreement (qualitatively and quantitatively) with practical experience from the real plant.

A change in the liquid internal reflux flowrate affects more the heavier products and less the lighter ones, due to the different proportion of heavier pseudo-components in each product. In our case, by increasing naphtha flowrate with

10% it leads to a deviation with approx. 3°C in naphtha EP, 4°C in kerosene EP, 5°C in LGO EP and 11°C in HGO EP.

3.2. The column with decoupler response

Fig. 5 shows the products EP's when the naphtha flowrate increases with 10%, having the decoupler connected (EP control loops are not in service yet). Two aspects have to be emphasized, in comparison with the EP's evolution in fig. 4 (the same 10% naphtha flowrate increase, but without decoupler). First, the system sensitivity to inputs (side-products flowrates) significantly increases: while fig. 4 shows a naphtha EP increase with about 3°C, in fig. 5 it increases with 17°C. Second, the heavier products EP's are not seriously affected anymore (only a 2.5°C deviation in kerosene EP, LGO and HGO EP's having almost constant values); the remaining deviations could be easily now corrected by the EP controllers.

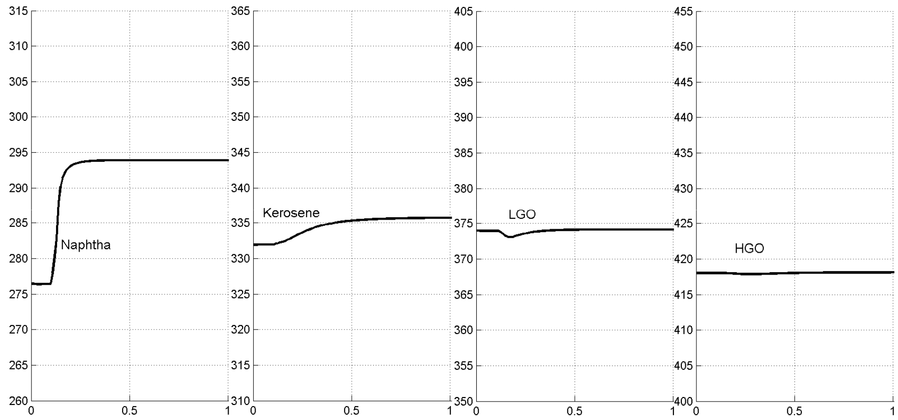

Fig. 5. Plant response for a 10% increase in naphtha flowrate; the decoupler is connected. Product EP's, in [°C], are depicted. The simulation time is expressed in [seconds × 10^4].

3.3. Closed-loop behavior

The efficiency of the proposed control structure can be observed now in fig. 6, where all EP control loops were left in service. A very good dynamic behavior of the modeled plant can be revealed, when the set point for naphtha EP controller increases with 2°C. Practically, the transient time is about 1.5 hours and kerosene, LGO and HGO EP's are not affected.

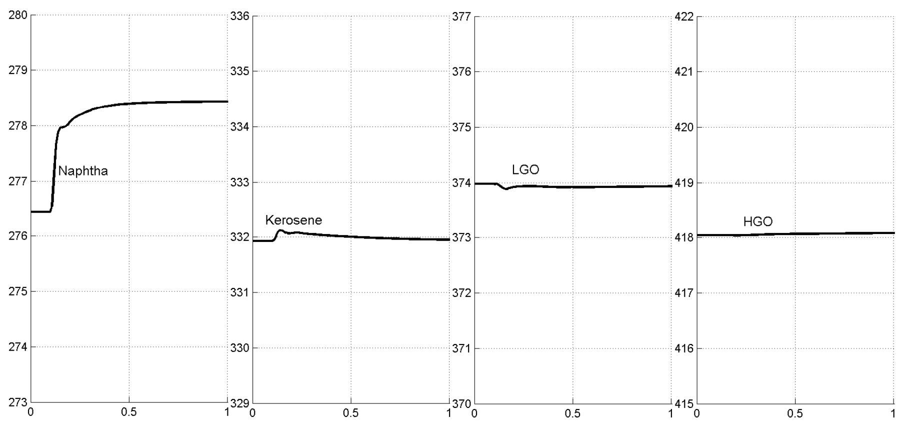

Fig. 6. Plant response when the set point for naphtha EP controller increases with 2°C. Product EP's, in [°C], are depicted. The simulation time is expressed in [seconds × 10^4].

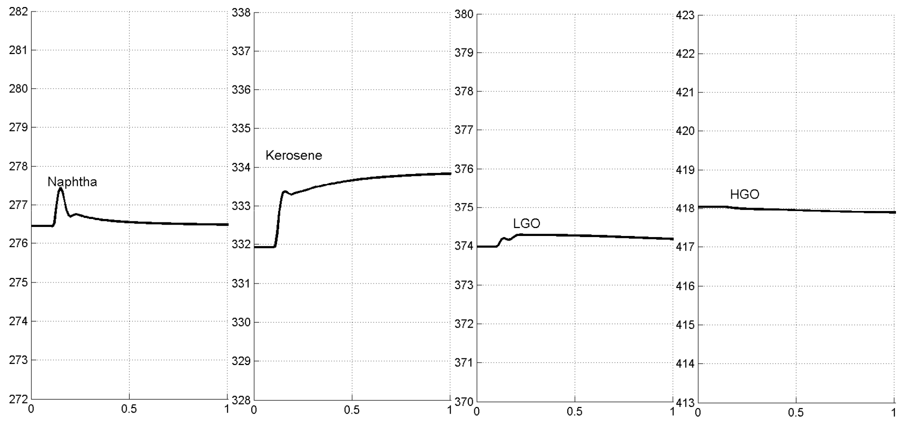

Fig. 7. Plant response when the set point for kerosene EP controller increases with 2°C. Product EP's, in [°C], are depicted. The simulation time is expressed in [seconds × 10^4].

The same good behavior can be observed in fig. 7, where product EP's evolution is presented, considering the set point for kerosene EP controller increases with 2°C. During the transient time (about 3 hours), the naphtha EP is slightly affected but its controller corrects the deviation in 1 hour; in the same time LGO and HGO are not significantly affected, due to the decoupler effect and an appropriate EP controllers tuning.

4. CONCLUSIONS

An original product quality control scheme for the crude oil unit was proposed in this paper. It is based on a Shinskey idea, but offers two original solutions, one for implementing a control loops decoupler and the other one for level control. Improved performances were revealed after extensive simulation tests. All good results may recommend this structure as subject for industrial implementation.

NOMENCLATURE

DL	liquid sidedraw from tray;
DV	vapor sidedraw from tray;
FL	liquid feed on tray;
FV	vapor feed on tray;
f	delaying function;
j, p	subscripts for product order number.

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