# Modern Dynamic Simulation of a Crude Oil Plant

## Revising the dynamic decoupling between quality control loops

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*A careful analysis reveals the extreme complexity of crude oil unit control task, as much as the process itself is complex (as inner property), stiff, with strange and unexpected behavior. For that reason, everything that seems to be new and appropriate in this field may become a different way in approaching such a complicate plant control. After the dynamic modeling and crude oil open-loop representation validation this article proposes a new view on the additional structures which allows for an improved quality control to be performed.*

*Keywords: crude oil unit, dynamic decoupling, simulation*

At present, there are many refineries which treat the crude oil unit as one of the most important plants, but covering it with control structures reduced at a minimum – especially considering the high price of analyzers (usually implied by the product quality control) and continuous downstream processing (none of its products being a final one *as is).* Such a control structure for sideproduct *p* is depicted in figure 1 [1].

The sidestripper (SS) bottom level is controlled with main column sidedraw FL, also controlling the $DL_p$ sideproduct and steam flowrates. The last two setpoints are completely offered to plant operators, meaning they are fully entitled to manually correct the product(s) Start Boiling Point (SP) and End Boiling Point (EP), after consulting the (usually) 4 to 8 f analysis report. However, there are important drawbacks affecting the plant: it works based only on critical plant operators skills, the slow quality control loops make the controllers hard to be tuned, and the plant intricate behavior is already known to force interesting decisions to be made. In this respect, an alternate approach has to be found in order to efficiently operate the crude oil unit.

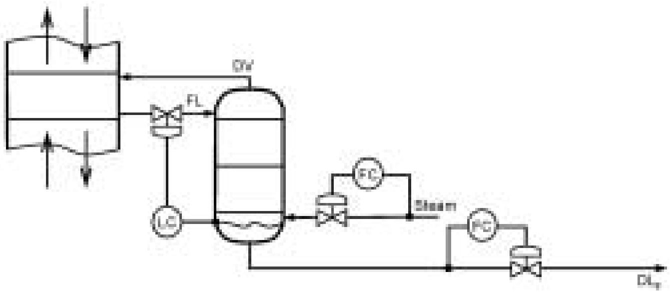


Fig. 1. Classical control structure for *p* sideproduct. Quality control loops are missing [1]

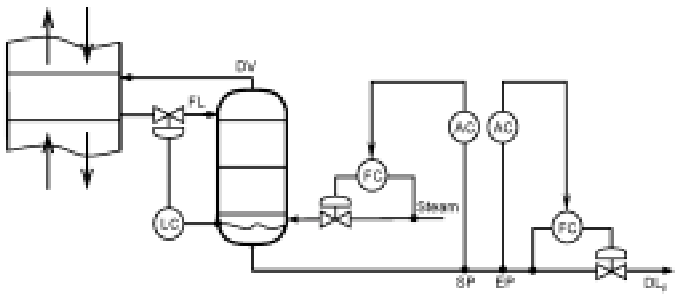


Fig. 2. Control structure for *p* sideproduct, with simple quality control loops [1]

*The classical quality control loops configuration*

A derived control structure is depicted in figure 2, where the human operator presence is simply replaced by using two quality control loops for *p* product SP and EP [1].

The only advantage now is the human operators' non-presence as part of the *controlling device*, because all other disadvantages are kept: long transient regimes, a system hard to be stabilized, strong interacting control quality loops. Also, the investment costs for the two analyzers are very high and their usage seems not to be required by the poor performances expected. For these reasons, the structure in picture 2 is not considered with practical relevance and is not implemented in industry [1].

*Decoupling the product quality loops*

As shown in our previous works [2-4], the sideproducts variation influences unidirectionally their EP's (to the column bottom) for a normally equipped column. Starting from this remark, different ways to approach the problem were presented, having one common purpose: the product quality loops' one-way decoupling – from column top to its bottom.

The first reference structure was proposed by Shinskey in [5]. As shown in figure 3, each sideproduct flow controller

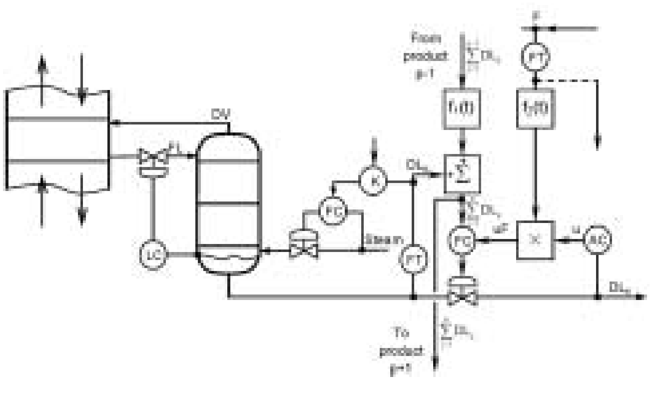


Fig. 3. Shinskey control structure for *p* sideproduct [1]

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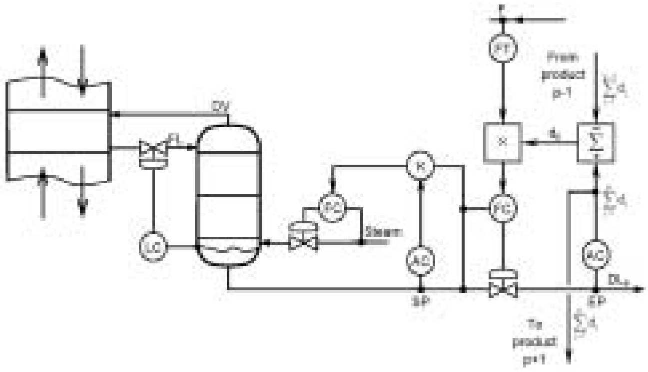


Fig.4. Alternative control structure for *p* sideproduct [1]

fixes the $\sum_{j=1}^{p} DL_j$ the sum of first *p* products flowrates, numbered from column top to its bottom), having as setpoint the column feed flowrate F multiplied with u (product quality controller output). If any of the *p*-1 flowrates modifies, the controller adjusts $DL_p$ in order to keep constant the sum $\sum_{j=1}^{p} DL_j$ (and the column internal reflux ratio, for loop decoupling) [1].

The structure is functional by adding the compensating elements $f_1(t)$ (to delay the upper products flow changes) and $f_2(t)$ (to delay the column feed disturbances), otherwise the system becomes highly unstable. The author of this paper considers that having only one $f_1(t)$ for all upper products changes is not enough (normally each product should have its own delaying element), but the quick response of the flow controller cancels all remaining unbalances. Another advantage of Shinskey's structure is the missing analyzers for products SP's, controlled only by a *K* report block [1]. But the main disadvantage is the absence of physical sense for the product quality controller output, and this is why (in 1998) Patrascioiu, Marinoiu, Paraschiv and Cîrtoaje proposed another structure [6]. Figure 4 shows this new approach, where two analyzers (for SP and EP) are already present.

The structure operates not with product flowrates, but with product potentials in feedstock ($d_j$). The EP controller has the output $\sum_{j=1}^{p} d_j$ and, whenever there is a change in $\sum_{j=1}^{p} d_j$, the correct value of $d_j$ multiplies the feed flowrate F and applies as sideproduct flow controller setpoint (maintaining constant the other p+1 EP's). Besides this unidirectional loop decoupling, this structure works by feedback (keeping a correct $d_j$ value) and by feedforward (including the F flowrate in the controlling algorithm) [1].

This alternative structure operates naturally with the plant, but still has two important drawbacks. It strongly depends on bottom product level control loops (for each

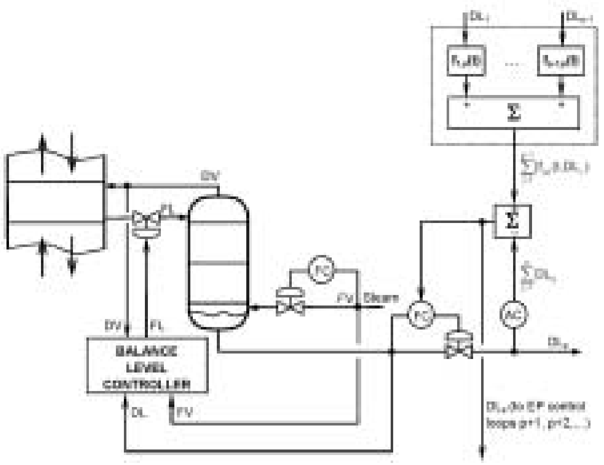


Fig. 5. Original control structure for *p* sideproduct, with quality loops dynamic decoupling [1]

sidestripper), so any local delays and oscillating behavior may become a characteristic for the whole system. Moreover, it totally omits all dynamic aspects, the authors testing it only in a (pseudo) steady state regime [1]. This is the main reason why the author of this paper developed an original controlling structure, trying to keep all good aspects from the Shinskey and this alternative regulating plant.

Figure 5 presents the new approach. The first feature to be mentioned is the "balance level controller" used to fix the sidestripper bottom level, which instantly reacts when there is a disturbance between inputs (DV – vapor back to column, DL – product flowrate and FV – steam flowrate) and output (FL – sidedraw to sidestripper). As the products SP are better controlled via column pumparounds [1], the steam flowrate setpoint is freely set. Also, the connection to column feedstock is missing, for all further experiments it being considered constant.

In order to decouple the quality loops, the controller for *p* product EP output is the desired sum $\sum_{j=1}^{p} DL_p$ between all 1, 2, ..., p products flowrates. It is subtracted from the sum $\sum_{j=1}^{p-1} [f_{j,p}(t, DL_j)]$ which individually delays all influences of j product flowrate in p product EP, finally the result being applied to product flowrate controller as remote setpoint. Here, the most important point is how these $f_{j,p}$ individual time functions are calculated. While Shinskey proposed aglobal influence approach, here all dependencies lighter product j – side product p, are individually treated. Moreover, in spite of Shinskey's approach, these dependencies are described through a second order delay system $f_{j,p}(DL_j, t) = a_{jp2}\frac{d^2DL_j}{dt^2} + a_{jp1}\frac{dDL_j}{dt} + DL_j$, $a_{jp2}$ and $a_{jp1}$ being the time constants. In this respect, we have a true, reduced dimensional model for the crude oil plant, specially adapted

**Table 1**
TIME CONSTANTS FOR THE DYNAMIC DECOUPLER

| | NAPHTHA (p=1) | KEROSENE (p=2) | LGO (p=3) | HGO (p=4) |
|---|---|---|---|---|
| GASOLINE (j=0) | $a_{012}$=120<br>$a_{011}$= 24 | $a_{022}$= 528<br>$a_{021}$= 48 | $a_{032}$= 1224<br>$a_{031}$= 72 | $a_{042}$= 1848<br>$a_{041}$= 88 |
| NAPHTHA (j=1) | - | $a_{122}$= 120<br>$a_{121}$= 24 | $a_{132}$= 528<br>$a_{131}$= 48 | $a_{142}$= 960<br>$a_{141}$= 64 |
| KEROSENE (j=2) | - | - | $a_{232}$= 120<br>$a_{231}$= 24 | $a_{242}$= 360<br>$a_{241}$= 40 |
| LGO (j=3) | - | - | - | $a_{342}$= 48<br>$a_{341}$= 16 |

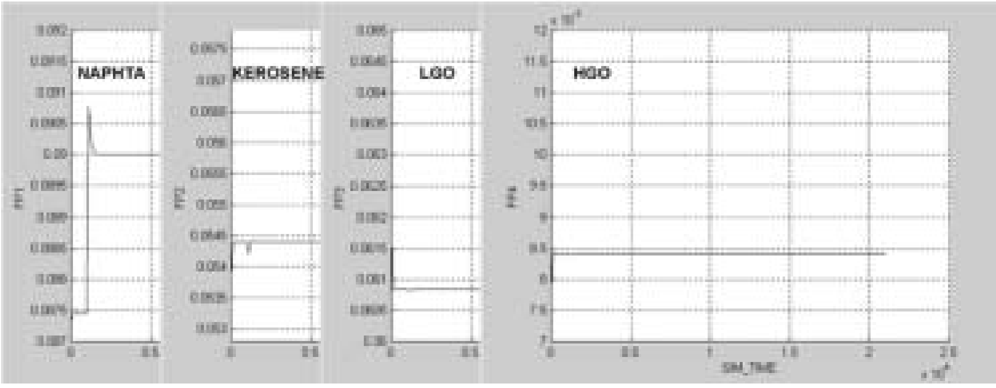


Fig. 6. Plant response for a 0.23 to 0.26 increase in reflux ratio, with the dynamic decoupler connected. Naphtha, kerosene, LGO and HGO flowrates are presented in [kmol/s]. The simulation time is expressed in [s× $10^4$].

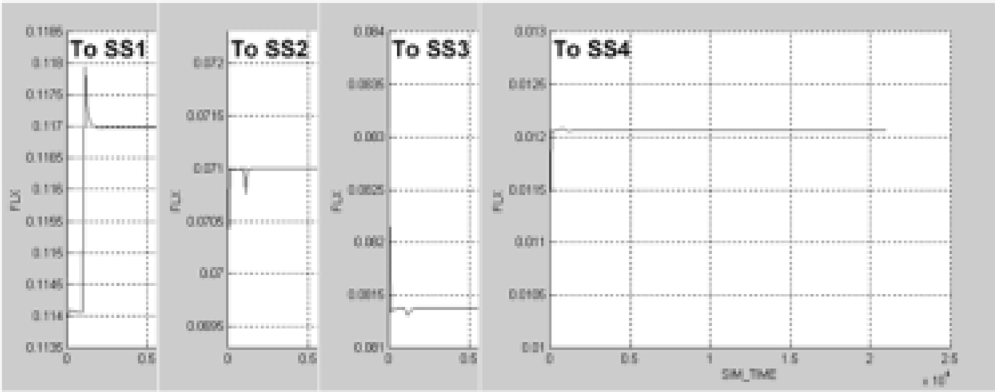


Fig. 7. Plant response for a 0.23 to 0.26 increase in reflux ratio, with the dynamic decoupler connected. Liquid sidedraws to SS1, SS2, SS3 and SS4 flowrates are presented in [kmol/s]. The simulation time is expressed in [s× $10^4$].

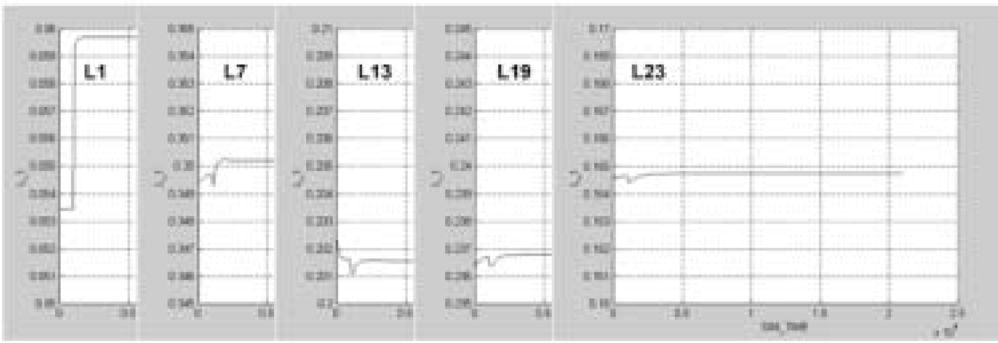


Fig. 8. Plant response for a 0.23 to 0.26 increase in reflux ratio, with the dynamic decoupler connected. Internal reflux from trays 1, 7, 13, 19 and 23, in [kmol/s], are depicted. The simulation time is expressed in [s× $10^4$].

to quality loops decoupling, $a_{jp2}$ and $a_{jp1}$ being its tuning variables. In table 1 these values are shown for the same crude oil unit presented in [1], taken as example in this work.

*Simulation results*

In order to test this dynamic decoupler behaviour, the simulation experiments propose an "open loop" configuration for the products' quality loops (gasoline and naphtha, kerosene, light gas oil – LGO, heavy gas oil – HGO), all end point temperature controllers being commuted in manual regime. Due to their very quick dynamics, the flow controllers are supposed to work "perfectly" in this environment.

All details regarding the crude oil unit are given in [2]. The simulations were performed in DIVA (Dynamische Simulation Verfahrenstechnischer Anlagen), developed at the Stuttgart University [7-12], a scientific software oriented on dynamic simulations for industrial environments.

The first test was a reflux ratio increase from 0.23 to 0.26 leading to gasoline flowrate decreasing.

As shown in figure 6, the decoupler leads to a naphtha flowrate increase by 0.003 kmol/s, simultaneously with the gasoline increasing flowrate keeping constant the sum $DL_0+DL_1$. All other side products flowrates remain constant.

The rapid action of the balance level controller is depicted in figure 7. Practically, the column sidedraw increase flowrate to SS1 (from 0.114 kmol/s to 0.117 kmol/s) makes the main column internal reflux to remain unchanged below tray 7 (extraction tray for non-stripped naphtha). The same efficient dynamic decoupler action may be seen in figure 8, where the column reflux ratio is shown (with its expected variation between tray 1 and 7).

The internal reflux restore below tray 7 leads to keeping constant the naphtha, kerosene, LGO and HGO EP's (only the gasoline EP being affected). This fact is depicted by figure 9, it proving an (almost) perfect unidirectional decoupling effect for quality control loops.

The second experiment is represented by increasing the naphtha EP controller manual output with 10%, the naphtha flowrate varying from 0.09 kmol/s to 0.14 kmol/s. In this case, the dynamic decoupler reset the sum $DL_0+DL_1+DL_2$ by lowering the kerosene flowrate ($DL_2$) with 0.05 kmol/s and preventing any LGO and HGO modifications (fig.10).

Like in the previous case, the step variations of naphtha and kerosene flowrates are quickly transformed (by the SS1 and SS2 balance level controller) in non-stripped product flowrate variations: tray 7 (to SS1) from 0.115 kmol/s to 0.175 kmol/s and tray 13 (to SS2) from 0.07 to 0.01 kmol/s (fig. 11). Obviously, the non-stripped fractions to SS3 and SS4 are not influenced at all.

The main column internal reflux flowrate is depicted in figure 12. As expected, the decoupler practically cancels any significant change on tray 13 (non-stripped kerosene), tray 19 (non-stripped LGO) and 23 (non-stripped HGO).

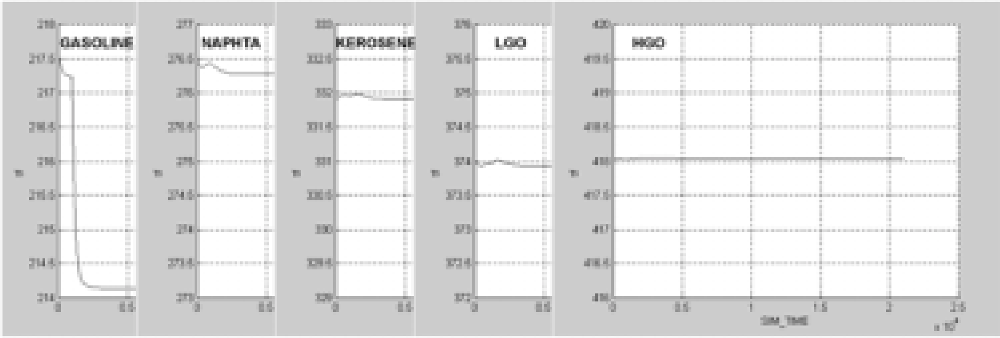


Fig. 9. Plant response for a 0.23 to 0.26 increase in reflux ratio, with the dynamic decoupler connected. The gasoline, naphtha, kerosene, LGO and HGO EP, in [$^0$C], are depicted. The simulation time is expressed in [s× $10^4$].

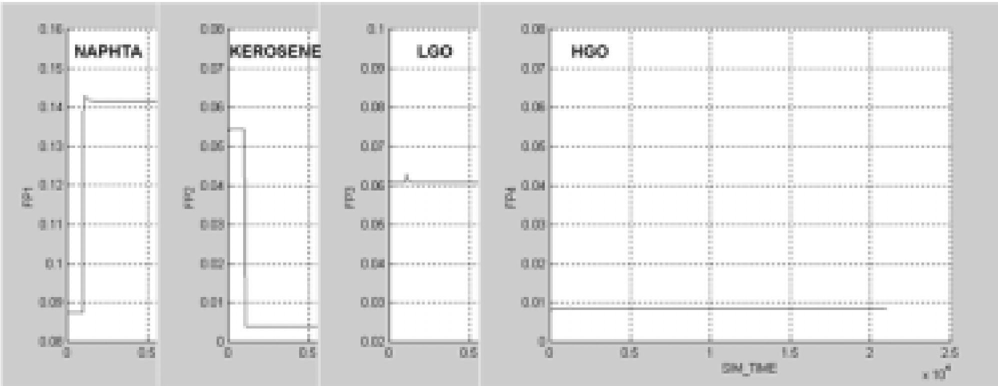


Fig. 10. Plant response for a 0.09 kmol/s to 0.14 kmol/s increase in naphtha flowrate, with the dynamic decoupler connected. Naphtha, kerosene, LGO and HGO flowrates are presented in [kmol/s]. The simulation time is expressed in [s× $10^4$].

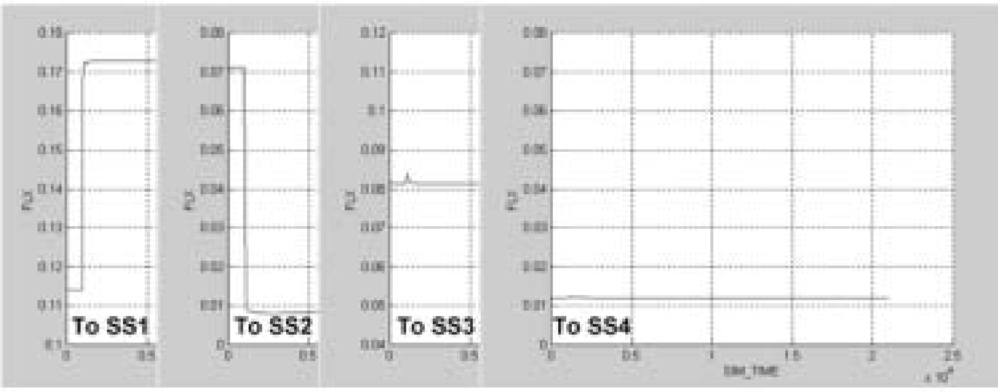


Fig. 11. Plant response for a 0.09 kmol/s to 0.14 kmol/s increase in naphtha flowrate, with the dynamic decoupler connected. Liquid sidedraws to SS1, SS2, SS3 and SS4 flowrates are presented in [kmol/s]. The simulation time is expressed in [seconds × $10^4$].

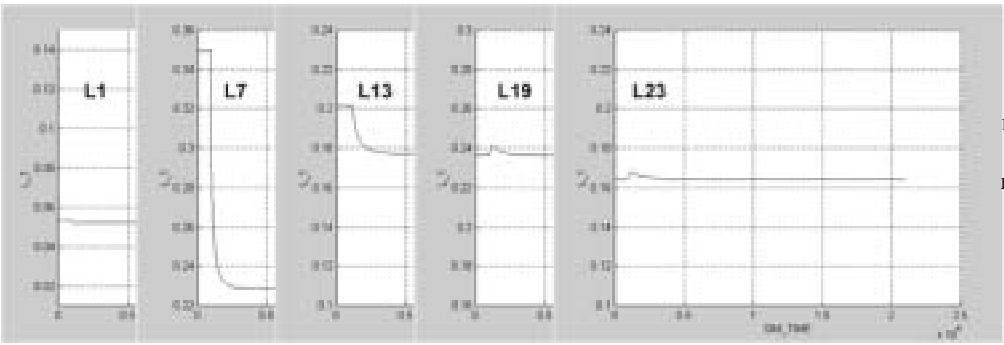


Fig. 12. Plant response for a 0.09 kmol/s to 0.14 kmol/s increase in naphtha flowrate, with the dynamic decoupler connected. Internal reflux from trays 1, 7, 13, 19 and 23, in [kmol/s], are depicted. The simulation time is expressed in [s× $10^4$].

In this case, a very light disturbance appears in the column, produced by the internal reflux flowrate variation (of 0.02 kmol/s) on tray 13 – although it is reduced by 85% by the decoupler presence. The explanation comes from the effective decoupler implementation which prohibits any final product flowrate to decrease below 0.005 kmol/s. This limit applies to kerosene flowrate (fig. 10), which will slightly differ from its calculated value. But nevertheless, on tray 19 this change disappeared, the reflux being the same as before (0.24 kmol/s). It is important to point out that, being subjected to big step variations (like this 10% change in naphtha flowrate), the whole system is very sensitive in the presence of the dynamic decoupler.

The internal reflux on tray 13 lowering with 0.02 kmol/s leads to an unwanted variation of +4$^0$C in kerosene EP, as shown in figure 13. Here are depicted a +17.5$^0$C increase in naphtha EP (expected) and a +2$^0$C increase in gasoline EP (undesirable, though), while LGO and HGO EP's are constantly kept – as it has to be.

This time we met a situation that cannot be explained only by normal reasoning, speaking here about this gasoline EP modification, which is in contradiction with the one-way disturbances propagation theory (from column top to its bottom). Nevertheless, we simply cannot say the internal reflux is the ONLY agent influencing the products' EP, this is why further investigation has to be made.

## Conclusions

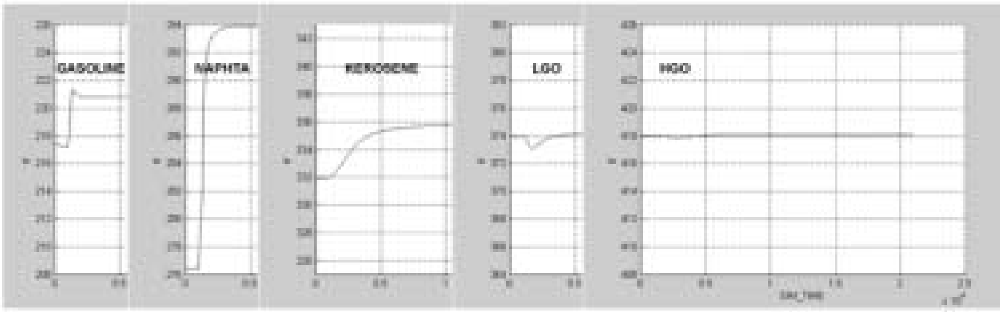


Fig. 13. Plant response for a 0.09 kmol/s to 0.14 kmol/s increase in naphtha flowrate, with the dynamic decoupler connected. The gasoline, naphtha, kerosene, LGO and HGO EP, in [$^0$C], are depicted. The simulation time is expressed in [s × $10^4$].

The crude oil plant is well-known for its complexity and stiffness, its strange and unexpected behavior being widely accepted. This is why everything new in this field may become a good way in approaching such a complicate plant control. In our case, after the dynamic modeling and crude oil "open-loop" representation validation, this paper presents a critical analysis on the additional structures which permit an improved quality control. An original decoupling structure for products' quality control loops is presented and tested, leaving the way open for a full control structure implementation.

## References

1.RADULESCU, G., Reglarea evoluatã a procesului de distilare atmosfericã, Editura Universitãii Petrol-Gaze din Ploiesti, Ploiesti, 2015.

2.RADULESCU, G., Rev. Chim. (Bucharest), **67**, no. 2, 2016, p. 314

3.RADULESCU, G., Rev. Chim. (Bucharest), **58**, no. 2, 2007, p. 239.

4.RADULESCU, G., PARASCHIV, N., KIENLE, Rev. Chim. (Bucharest), **58**, no. 3, 2007, p. 349

5.SHINSKEY, G. F., Distillation Control for Productivity and Energy Conservation, McGraw-Hill Book Company, New York, 1984.

6.PATRASCIOIU, C., MARINOIU, V., PARASCHIV, N., CIRTOAJE, V., Buletinul Universitatii Petrol-Gaze Ploieºti, **XLVII-L**, Nr. 12, 1998.

7.MARQUARDT, W., HOLL, P., GILLES, E. D., Proc. of the III$^{rd}$ World Congress of Chemical Engineering, Tokyo, 1986, p. 687.

8.GILLES, E. D., HOLL, P., MARQUARDT, W., MAHLER, R., SCHNEIDER, H., BRINKMANN, K., WILL, K. H., Automatisierungstechnische Praxis, **32**, 1990, p. 343.

9.MANGOLD, M., KIENLE, A., MOHL, K, D., GILLES, E. D., Chem. Engng. Sci., **55**, 2000, p. 441.

10.MARQUARDT, W., HOLL, P., BUTZ, D., GILLES, E. D., Chem. Engng. Tech., **10**, 1987, p. 164.

11.KRONER, A., HOLL, P., MARQUARDT, W., GILLES, E. D., Comput. Chem. Engng., **14**, 1990, p. 1289.

12.TRANKLE, F., GERSTLAUER, A., ZEITZ, M., GILLES, E. D., Comput. Chem. Engng. (Suppl.), **21**, 1997, p. 841

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