BULETINUL Universităţii Petrol – Gaze din Ploieşti	Vol. LXVII No. 3/2015	94 – 100	Seria Tehnică

General Aspects Regarding Hierarchical Control of a Complex Industrial Process

Alin-Constantin Băieşu, Alina-Simona Băieşu

Honeywell România, Str. George Constantinescu 3, Bucureşti,
e-mail: alin.baiesu@honeywell.com
Universitatea Petrol – Gaze din Ploieşti, Bd. Bucureşti 39, 100680, Ploieşti,
e-mail: agutu@upg-ploiesti.ro

Abstract

In case of complex industrial processes, such petrochemical plants the hierarchical viewing of the control problem for simplifies the modeling effort. In order to do this, for each level of the hierarchical structure a model of the process is needed. The paper presents a method that is based on a hierarchy of models describing the complex industrial process at various levels of abstraction and a decomposition of the overall specification. The complexity of the resulting control scheme is considerably reduced in comparison to an unstructured (and therefore non-hierarchical) approach. The procedure is applied for a crude oil distillation unit.

Key words: *complex industrial process control, hierarchical control, crude oil distillation unit.*

Introduction

In order to properly analyze the steady state and dynamic control of any complex plant, adequate models are required. Using a hierarchical approach method, different types of models with different complexity would be required at different stages of plant details. Static models capture the long-horizon characteristics of the process and are useful for analysis at the higher levels of the plant. At a more detailed level where dynamics characterize the representations, the dynamic models must be used in order to synthesize the control structures.

Starting from an abstract viewpoint of the plant, the representation can be progressively refined into new viewpoints, each containing increasing levels of detail about the plant. Each viewpoint captures a particular range of characteristics of the process.

At lower levels, step-by-step the necessary details are introduced.

The viewing resolution is manifested in a number of process units that are observable from the specific viewpoint and the viewing span is measured by the details contained in the corresponding representation. In this paper the levels of hierarchical viewing have been adopted so that each process representation corresponds to a different time-scale of operations. The hierarchical framework also defines a temporal resolution in terms of planning horizons. Simpler process representation captures the plant behavior, which is observable over long time-horizons, while more detailed process representation brings out the faster dynamics in the process. Hence, within the hierarchical framework, the effects of process phenomena have been grouped according to their associated time-horizons, i.e. time scales. Each of these horizons characterizes the behaviour of the process at a particular level in the hierarchy of the plant's stratification.

A Hierarchical View of a Control System for a Complex Industrial Process

Controlling a complex system is difficult because of its features: dimension, geographical spreading and the great number of objectives.

Hierarchical control is an attempt to handle complex problems by decomposing them into smaller sub-problems, resolving them and reassembling their solutions in a hierarchical manner [3, 8].

Generally, the following features characterize any hierarchical control system [4]:

- the overall goal is decomposed into a number of high-level and low-level specifications;
- the high-level specification are usually concerned with long-term developments;
- low-level specifications are usually concerned with short-term developments.
- the higher level has the right to interfere and to impose performance criteria over lower levels;
- there is a permanent interaction between the units of the same level and between the units of different levels so that the system goal to be achieved.

Using a decentralizing approach, a complex system will be decomposed in sub-systems depending on their natural interactions.

Control systems for continuous plants are often built in a hierarchical manner, with regulatory control at the lowest level, a supervisory level above, and an optimization level on top [5].

A three-level hierarchical control system is shown in Figure 1.

For process characterization (controlled sub-systems and the two levels associated to them) the variables can be grouped in the following sets [7]:

- Ω – the coordinating variable set, setpoints for the third level (e.g. profit functions + product quality);
- Γ – the 3rd level information set;
- I – the coordinating variable set, setpoints for the second level (e.g. heat exchangers duties, product specifications, temperatures, etc);
- U – the command set (e.g. products flow setpoints);
- R – the 1st level feedback set (the process values, viewed from the second level);
- X – the 2nd level feedback set (the process values, viewed from the third level);
- D – the process disturbance set;
- Y – the process output set.

Usually the information flow in such a control hierarchy refers to the higher level sending commands to the layer below, and the lower layer reporting back with a measure of performance. To give the commands, every higher level must know the behaviour of the controlled lower level, represented by its model.

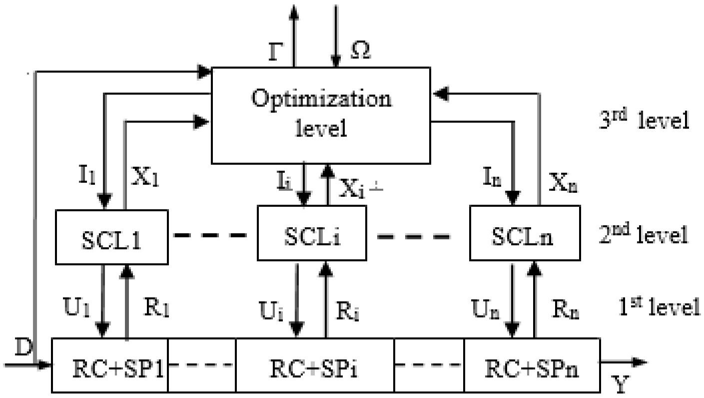

Fig. 1. A three-level hierarchical control system for a complex system: SCL_i - Supervisory Control Level (for sub-process i), $RC+SP_i$ - Regulatory Control + Sub-Process i.

Using the coordination variables and knowing how the controlled sub-system will behave the higher level computes the commands values.

The lower level decisions can be affected by the higher level by [4]:

- goal;
- imposing the criterion for the decision selection;
- choosing an alternative from many possibilities at a certain time.

On the other hand, the control layer is based also on the feedback information from the lower sub-systems.

The way that a sub-system from any level acts, is directly influenced by the higher level, most of the time by the direct higher level.

Although the important decision is taken by the higher level, the whole system performance depends on the behaviour of each sub-system, because the overall goal achievement implies the accomplishment of each local goal.

The higher level must take into account the interactions, which exists between the lower levels sub-systems. So it has to be able to assure a decoupling procedure between them.

A Hierarchical View of a Crude Oil Distillation Unit

A block diagram of a hierarchical control system for a crude oil distillation unit it is presented in Figure 2.

The goal of the optimization level is to assure that the lower level will act so that the specification will be achieved and the profit will be maximized. Therefore, at this level the setpoints for the lower levels are computes using some measures of profitability.

The profit P is defined as the difference between the revenues R and the total operating cost C_{total}, that is [2]:

$$P = R - C_{total} . \tag{1}$$

For any industrial process, the total operating cost is the sum of the raw materials and utility costs C_{RU} , the annualized capital cost C_{Cap}, the labour costs C_{Labor} , and the annualized cost for the control system hardware/software C_{CS}.

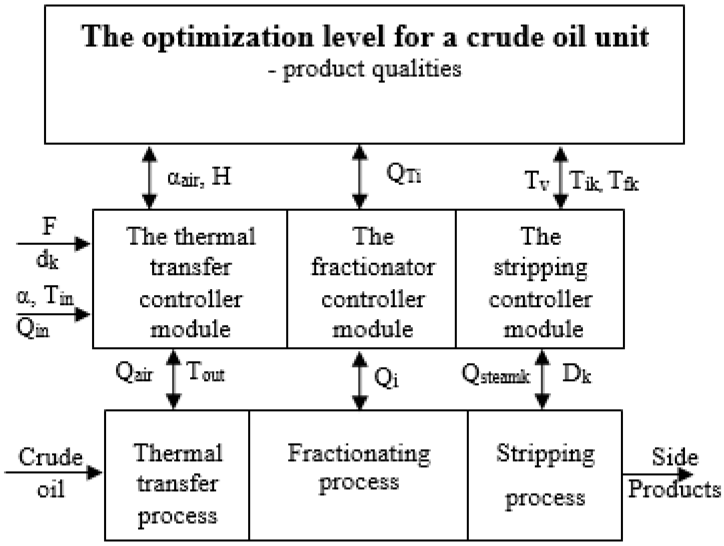

Fig. 5. A hierarchical control system for a crude oil distillation unit

Since a control system for an existing plant it was synthesized, the annualized capital and the control system costs are assumed to be constant. Also the labour costs can be assumed to be constant.

Thus, the following holds

$$\max P \Leftrightarrow \max[R - C_{RU}]. \tag{2}$$

Clearly, this optimization problem must be solved subject to many constraints such as production rate, equipment safety, etc. Also, the profit depends on the disturbances, D.

Thus

$$\max P \Leftrightarrow \max[R - C_{RU}](D), \tag{3}$$

subject to constraints.

This optimization problem is based on the model of the plant seen from the top level. Using this model the top level establishes the setpoints for the lower level so that this function to be maximized.

In order to accomplish this objective, the overall heating utility and the supply distributions must be minimized. Minimum utility calculations are performed by establishing several intervals of temperature.

The minimum heating utility, S_{min} is obtained by solving the following problem [1]:

$$S_{\min} = Min\, q_i, \tag{4}$$

where q_i is the heat surplus on interval i and is given by

$$q_i = \sum_{k \in I_i^H} F_k^H cp_k^H (T_{i-1} - T_i) - \sum_{j \in I_i^C} F_j^C cp_j^C (T_{i-1} - T_i). \tag{5}$$

Assuming that a set of n_u utilities are available at different temperature levels, if U_i is the amount of heat supplied by each utility, then the problem of minimizing the supply distribution become the following:

$$S_{\min} = Min \sum_{i=0}^{n_u} \omega_i U_i, \tag{6}$$

where ω_i is the utility costs and *i* is the interval that corresponds to the temperature interval obtained by minimizing the heat surplus.

This level also supplies the side-product quality specifications, which are initial and final STAS boiling points.

Because this plant can be decomposed in three main sub-processes: the thermal transfer process, the fractionator and the stripping sub-process, at the second level of the hierarchical control system we will have three control modules, one for each sub-process.

The thermal transfer controller module, using the setpoints from the higher level (the overheated percent and the air excess) and taking into account the disturbances (the air temperature, the inlet crude oil temperature and flow) computes the setpoints for the associated sub-system from the first level. These setpoints are the outlet temperature and the airflow.

From this level the process is viewed as

$$T_{out}, Q_{air} = f_1(\alpha, T_g, T_{air}, F, T_{in}, H), \tag{7}$$

where T_{out} is the crude oil outlet temperature, Q_{air} - air flow, α - air excess, T_g - the flue gases temperature, F - crude oil flow, T_{in} - the inlet crude oil temperature, H - the overheated percent.

The function f_1 can be a first order lag element for the channel airflow - air excess, disturbances and a second order lag element for the canal outlet temperature – overheated percent, disturbances.

The fractionator controller module using the thermal duties for the two pumparounds and the top temperature established by the higher level, computes the setpoints for the sub-system from the lower level. These setpoints are the two pumparounds flows.

From this level the process is viewed as

$$Q_i = f_2(T_v, F, Q_{T\,i}, T_i), \tag{8}$$

where Q_i is the setpoint for the pumparound's *i* flow, T_v - the top temperature, Q_{Ti} - the *i* pumparound duty, T_i - the difference between the inlet and outlet temperature of the pumparound *i*.

The stripping controller module computes the setpoints for the stripping process (the steam and side products flows) using the setpoints received from the higher level (the initial and final boiling temperatures) and taking into account the disturbances (the crude oil flow and its quality, which implies the potential of each side product). These setpoints are the side-draw flows and the stripping steam flow.

The models associated to this level are:

$$D_k, Q_{steamk} = f_3(T_{ik}, T_{fk}, d_k, F), \tag{9}$$

where D_k is the side product k flow, Q_{steamk} - the steam flow for side product *k*, T_{ik} - the initial boiling point of side product *k*, T_{fk} - the final boiling point of side product *k*, d_k - the potential of side product *k*.

These models can be first order lag elements with dead time.

Each level must also contain a decoupling block, to assure the compensation of interactions that exist between systems at the lower levels.

At the first level the process contains the models associated to the three sub-processes. At this point, every sub-process is characterized by more detailed models. Relations that capture the process intimacy such as mass and energy balances, vapour/liquid relationships, etc. give these models.

For example, for the thermal transfer process we have:

$$\begin{aligned}
&Q_f\,(q_f + h_f + h_{air} L_{\min} \alpha) = Q_T + Q_g h_g\,(T_g);\\
&Q_T = Q_{in}\,(H_{out}(T_{out}) - h_{in}(T_{in}));\\
&Q_T = \alpha_r A_{Tr}\,(T_g^4 - T_e^4) + \alpha_c A_{Tc}\,(T_g - T_e);\\
&T_e = \frac{T_{in} + T_{out}}{2}.
\end{aligned} \tag{10}$$

where Q_f is the fuel flow, q_f – heating value, H, h – the molar enthalpy, Q_g - the flue gases flow, α_r - the transfer thermal coefficient of the radiation zone, α_c - the transfer thermal coefficient of the convection zone, Q_{in} - the inlet crude oil flow, A_{Tr} – the radiation change area, A_{Tc} – the convection change area T_{in} - the inlet crude oil temperature, T_{out} - the crude oil outlet temperature, T_g - the flue gases temperature.

For the fractionator and stripping processes the model is given by mass and energy balances, and vapour/liquid relationships, as follows [6, 9]:

$$\frac{2}{3} bm_K \cdot LE_K^{-1/3} \cdot \frac{d\,LE_K}{dt} = FL_K + FV_K + L_{K-1} + V_{K+1} - VE_K - LE_K \tag{11}$$

where, $V_k = VE_k - DV_k$ and $L_k = LE_k - DL_k$ and m_K is the tray's molar holdup, FL_K, FV_K - inlet liquid and vapour molar flow rates, LE_K, VE_K - liquid and vapour molar flow rates, which come from the tray k, DL_K, DV_K - the side liquid and vapour molar flow rates, L_K, V_K - liquid and vapour molar flow rates, which arrive to the next tray, where

$$\begin{aligned}
&m_K \frac{d\,X_{K,i}}{dt} = FL_K\left(XFL_{K,i} - X_{K,i}\right) + FV_K\left(YFV_{K,i} - X_{K,i}\right) + L_{K-1}\left(X_{K-1,i} - X_{K,i}\right) +\\
&+ V_{K+1}\left(K_{K+1,i} \cdot X_{K+1,i} - X_{K,i}\right) - VE_K \cdot X_{K,i}\left(K_{K,i} - 1\right).\\
&0 = FL_K\left(HFL_K - HL_K\right) + FV_K\left(HFV_K - HL_K\right) + L_{K-1}\left(HL_{K-1} - HL_K\right) + V_{K+1}\left(HV_{K+1} - HL_K\right) -\\
&- VE_K\left(HV_K - HL_K\right) - m_K \sum_{i=1}^{NC}\left(HLP_{K,i} \cdot \frac{dX_{K,i}}{dt}\right) - m_K c_{pK} \frac{\sum_{i=1}^{NC}\left(K_{K,i} \frac{dX_{K,i}}{dt}\right)}{\sum_{i=1}^{NC}\left(\frac{\partial K_{K,i}}{\partial T} X_{K,i}\right)}.
\end{aligned} \tag{12}$$

XFL_{kj} is the i component liquid molar fraction in tray's k inlet, YFV_{kj} - the i component vapour molar fraction in tray's k inlet, $X_{K,i}$ - the i component liquid molar fraction on tray k, $Y_{K,i}$ - the i component vapour molar fraction on tray k.

Conclusions

Usually, complex systems have a complex behavior, which is difficult to analyze and comprehend. Using a hierarchical control approach one can decompose it into smaller sub-processes, solve their problems and then reassemble their solutions to obtain the overall solution. To use this method, there are needed different types of models with different complexity for different stages of system details. In this paper are presented some aspects regarding a hierarchical control system structure for a complex process, with particularization for a crude oil distillation unit, outlining the models associated with each level of the hierarchy.

References

1. Bagajewicz, M.J. – Energy savings horizons for the retrofit of chemical processes. Application to crude fraction units, *Computers & Chemical Engineering*, Vol. 23, 1998, p. 1.
2. Douglas, J.M. – *Conceptual design of chemical processes*, McGraw-Hill, New York, 1988.
3. Guran, M., Filip, F.G. – *Sisteme ierarhizate,în timp real, cu prelucrarea distribuită a datelor*, Bucureşti, Editura Tehnică, 1986.
4. Mesaroviç, M.D., Mako, D., Takahara, Y. – *Theory of hierarchical, multilevel systems*, Academic Press, New York, 1970.
5. Morari, M., Arkun, Y., Stephanopoulos, G. – Studies in the synthesis of control structures for chemical process, Part I: Formulation of the problem. Process decomposition and the classification of the control tasks. Analysis of the optimizing control structures, *AIChE Journal*, Vol. 26,1980, p. 220.
6. Olsen, I., Olav, G., Sira, T. – A rigorous and efficient distillation column model for engineering and training simulators, *Computers & Chemical Engineering*, Vol. 21, 1997, p. 193.
7. Paraschiv, N., Soare, D., Ceapă, Gh., Marinoiu, V. – A hierarchical control approach to a fluid catalytic cracking unit, *Control Engineering and Applied Informatics*, Vol. 3, 2001.
8. Raisch, J., Itigin, A., Moor, T. – Hierarchical control of hybrid systems, *ADPM2000, The International Symposium on Advanced Control of Chemical Processes*, 2000, pp. 67-72.
9. Rădulescu, G. – *Contribuţii privind reglarea după model a procesului de distilare atmosferică*, Ph.D. Thesis, Universitatea Petrol-Gaze Ploieşti, 2002.

Aspecte generale privind reglarea ierarhică a unui proces industrial complex

Rezumat

În cazul proceselor industriale complexe, cum ar fi instalaţiile petrochimice, abordarea ierarhică a problemei reglării simplifică efortul de modelare. Pentru a face acest lucru, pentru fiecare nivel al structurii ierarhice este nevoie de un model al procesului. Lucrarea prezintă o metodă care se bazează pe o ierarhie de modele care descriu procesul industrial complex la diferite niveluri de abstractizare și o descompunere a specificației de ansamblu. Complexitatea sistemului de reglare rezultat este redus considerabil în comparație cu o abordare nestructurată (și, prin urmare, neierarhică). Procedura este aplicată pentru o instalaţie de distilare a ţiţeiului.