Some Aspects Regarding Feedforward Control of Fractionating Processes

Nicolae I. Paraschiv and Gabriel N. Radulescu

***Abstract*—Feedforward control structures, applied to fractionating processes, prove their performances by taking into account the special dynamic behavior of these industrial plants, as well as specific products quality requirements. Real-time control by using feedforward algorithms imposes low-order simplified models for implemented software controllers. These control models are mainly related to some process behavior specifications; this is the reason why they have to be permanently tuned.**

This paper presents a point of view on how to build up, validate and tune such a control model for the fractionating processes.

***Index Terms*—Feedforward control model, fractionating process, simulation model, validation procedure.**

I. Introduction

At present, the well-known limitations of the standard PID feedback control structures impose as an alternative different concepts and algorithms (for example Feedforward Control, Model Predictive Control, Internal Model Control). This constitutes the large field of "advanced process control". All these "intelligent" structures are based on (or even include in a specific manner) the controlled process mathematical model.

As to fractionating processes, at present the advanced control schemes built on feedforward principle has the most important place within the advanced control structures mentioned above. For an industrial implementation, such a control system first requires an input-output characterization of the controlled process. The next step is the process modeling, followed by the control algorithm design and implementation. Finally, once the process and the control structure have been modeled, the aggregated system behavior must be tested and validated via (dynamic) simulation.

The authors present in this paper some relevant and problematic aspects related to the feedforward control principle, fractionating process dynamics modeling and simulation, as well as control algorithm synthesis. In the end, an integrated hardware/software platform for testing the aggregated system (fractionating plant connected to the control structure) is also presented.

N. I. Paraschiv is with the "Petroleum-Gas" University of Ploiesti, Department of Control Engineering and Computers, 39 Bucuresti Blvd, Ploiesti, ROMANIA (phone: +40-244-573171 ext.270, fax: +40-244-575847, e-mail: nparaschiv@mail.upg-ploiesti.ro).

G. N. Radulescu is with the "Petroleum-Gas" University of Ploiesti, Department of Control Engineering and Computers, 39 Bucuresti Blvd, Ploiesti, ROMANIA (e-mail: gabriel.radulescu@mail.upg-ploiesti.ro).

II. Feedforward Versus Feedback Control

There are two important possibilities of controller output determination within control systems, namely feedback and feedforward principles.

The feedback control principle, illustrated in Fig. 1, imposes controller output modification when there is a difference between the set point value and the process controlled variable current value.

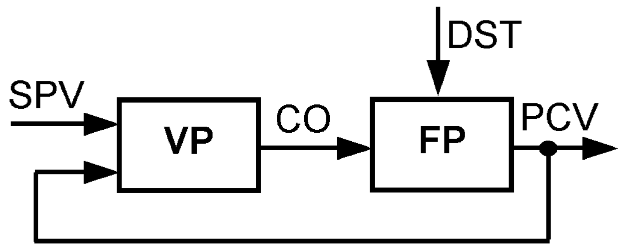

Fig. 1. Structure of a feedback control system: VP – variable part; FP – fixed part; SPV – Set Point Variable; PCV – Process Controlled Variable; CO – controller output; DST – Disturbances.

In the structure in Fig. 1, VP represents the controller and FP includes the process, the transducer and the actuator. The objective of this control system is to maintain PCV and SPV values equal when DST modify.

An advantage of feedback control is the use of universal control algorithms, such as PID. But, because controller output modification is made only after a PCV deviation from SPV occurred, there are periods of time when the system fails to achieve its objective. This aspect represents a disadvantage of feedback control systems, which is important if FP is characterized by a slow dynamic behavior and the set point state can hardly be re-established. A significant aspect of fractionating processes is that, until the set point state is re-established, there are quality nonconformities that have important consequences regarding controlled process efficiency.

The feedforward control principle, illustrated in Fig. 2, imposes controller output modification when there appear variations of the considered disturbances.

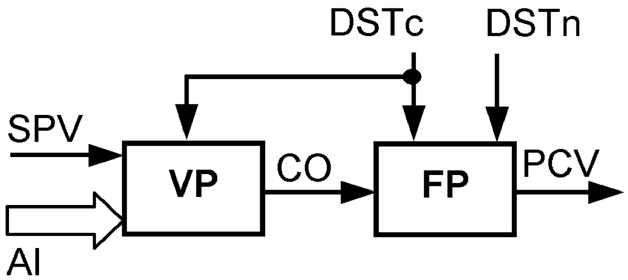

Fig. 2. Structure of a feedforward control system: VP – variable part; FP – fixed part; SPV – Set Point Variable; AI – Auxiliary Inputs; PCV –

Process Controlled Variable; CO – controller output; DSTc – Considered Disturbances; DSTn – Neglected Disturbances .

The controller action must compensate for that of the considered disturbances so that there is no difference between PCV and SPV. Under these circumstances, when only changes in DSTc appear, feedforward control systems entirely meet their objectives.

For these systems VP must contain a process control mathematical model whose dimensions can diminish the control system's real time performances.

The most important disadvantage of feedforward control systems is the impossibility of re-establishing the set point state when modifying DSTn.

Control systems with combined action (feedback and feedforward), having the structure shown in Fig. 3, combine the advantages of feedback and feedforward control and significantly reject their disadvantages.

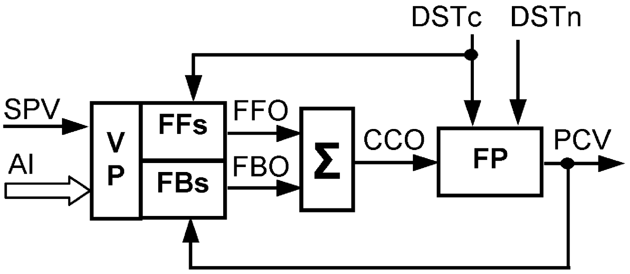

Fig. 3. Structure of combined actions control system: VP – variable part; FFs – FeedForward section of VP; FBs – FeedBack section of VP; Σ – summation element; FP – fixed part; SPV – Set Point Variable; AI – Auxiliary Inputs; PCV – Process Controlled Variable; CCO – combined controller output; FFO – FeedForward Output; FBO – FeedBack Output; DSTc – Considered Disturbances; DSTn – Neglected Disturbances.

Upon DSTc modification there runs the feedforward section (and PCV does not deviate from SPV), and when modifying DSTn it is the feedback section which is active, the set point state being re-established after transitory regime stoppage.

As a conclusion, feedforward control is necessary when there are disturbances with significant variations and when it takes a long time to eliminate the offset, which affects process efficiency. This is the case of fractionating processes whose feedforward control will be dealt with below.

III. Fractionating Process Variables Structuring

This process aims at obtaining high-value products by fractionating some mixtures whose composition is known.

The separated products' value is determined by their quality. An important indicator of quality is represented by composition. For example, given a binary mixture fractionation, the quality indicators are represented by top and bottom quality products from a fractionating column.

The fractionated products' quality is determined by:

- feed flow rate;
- feed quality;
- fractionating equipment state;
- other utilities (e.g. steam) properties.

In this context, the qualities of the column-extracted products represent output variables, and the variables that influence these qualities are considered to be input variables.

This classification depends on the purpose of the model in which these variables make part and therefore it is not unique. For example, for design models qualities are input variables, while for control models these are output variables.

Adopting an evolved control structure triggers the structuring of the inputs set in subsets, as shown in Fig. 4.

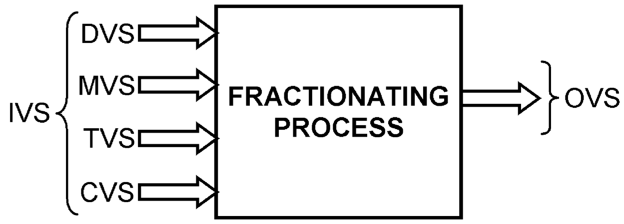

Fig. 4. Fractionating process' variables sets and subsets: IVS – Input Variables Set; OVS – Output Variables Set; DVS – Disturbance Variables Subset; MVS – Manipulated Variables Subset; TVS – Tuning Variables Subset; CVS – Constraints' associated Variables Subset.

The fractionating process takes place in fractionating columns and consists in mass transfer between liquid and vapor phases. A fractionating column may have one or several feed flows, respectively two or more fractionating products flows. Components of sets and subsets variables will be established for a binary mixture fractionating column presented in Fig. 5.

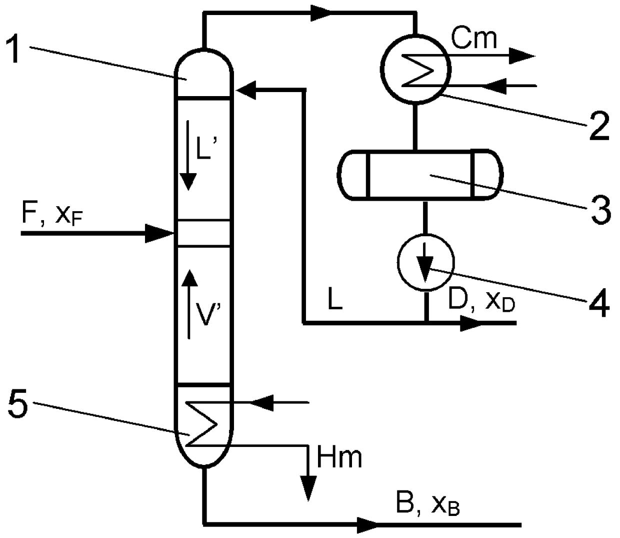

Fig. 5. Fractionating system: 1 – Fractionating column; 2 – Vapor condenser; 3 – Reflux drum (accumulator); 4 - Pressure source (pump); 5 – Incorporated reboiler; F, D, B, L – feed, distillate, bottom and external reflux flowrates; x_F, x_D, x_B – Concentrations of the more volatile component in liquid phase for feed, distillate, bottom flows; $L^{'}$, $V^{'}$ – Internal liquid and vapor flows; Cm, Hm – Cooling and heating agent medium flowrates.

The keeping of quality conditions for fractionating products requires the following variables to be controlled [1]:

- top and bottom streams composition;
- pressure inside the column;
- liquid levels at the base of the column and in reflux drum.

MVS components are resources for fractionating process control systems. In this context, MVS contains the following variables:

$$MVS = \{L,\ D,\ B,\ V',\ Cm\}\,, \quad (1)$$

where for *MVS* components there keep valid the significances from Fig. 5.

The distribution of manipulated variables (1) according to process control requirements is vital for the fractionating process' efficiency [1]. A method used by the authors to establish this distribution is Relative Gains Matrix Method [2].

As to *DVS*, the two main disturbances that affect the fractionating process are associated to feed channel, *DVS* being structured as follows:

$$DVS = \{F,\ x_F\}\,. \quad (2)$$

This *DVS* structure is justified by frequent changes of flow and composition feed and by their important influence upon fractionating products' quality.

The control models tuning necessity impose the presence of TVS within IVS. The authors' experience [3] has proved that the modification of a control model's validity conditions is determined by factors related to the fractionating equipment's state and to feed mixture characteristics.

Changes in the fractionating column state influence its efficiency, so they also influence *the number of theoretical trays N* [4]. As regards fractionated mixture characteristics, they are very well reflected in equilibrium constants values (for multi-component mixtures) and in *relative volatility α* (for binary mixtures). Consequently, for a binary mixture, *TVS* has the following structure:

$$TVS = \{N,\ \alpha\}\,. \quad (3)$$

CVS contains elements related to process operating restrictions. It is worth mentioning the ones related to imposed values (set points) for x_D and x_B, operating costs and environmental protection.

As regards *OVS*, this contains quality indicators represented by the light component concentration in top and bottom products, respectively:

$$OVS = \{x_D,\ x_B\}\,. \quad (4)$$

IV. Issues In the Fractionating Process Dynamic Simulation

To dynamically simulate a given physical system means in fact to use a representation for this one, which can offer a qualitative and/or quantitative image of the real system behavior in a time interval, when a set of inputs changes. This representation must have at least two functional sections: a model for the physical system and a "simulation engine" (see Fig. 6). There must also be a correlation between the physical system inputs/outputs (real inputs/outputs) and the simulator inputs/outputs (modeled inputs/outputs).

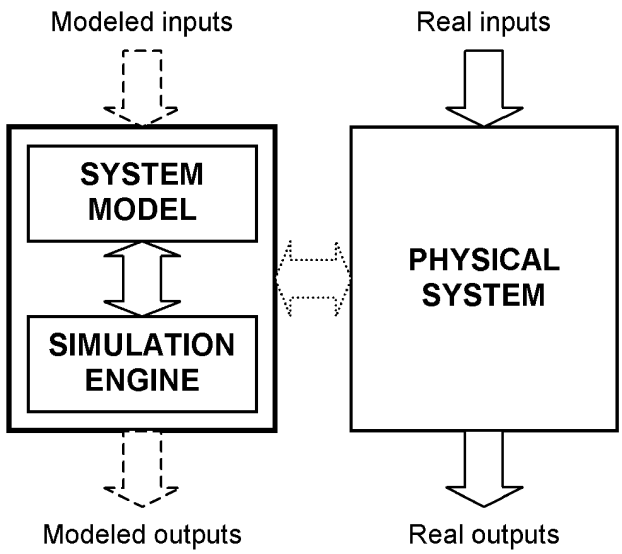

Fig. 6. The general simulator architecture.

When a mathematical model characterizes the system and the "engine" is a software environment, this is the case of a typical software simulator.

As regards the fractionating process, the authors' experience in this field allows them to emphasize three major topics requiring the process dynamic simulation. First, a dynamic software simulator, based on a "complete", rigorous and accurate mathematical model, may be a very useful tool in process intimacy research, offering a pertinent overview on system behavior. Second, using the dynamic simulation technique, there are several ways to obtain from it both dynamic and steady state reduced-scale models, as a specific application requires (typically in the fractionating process advanced control area). Third, using an appropriate software/hardware environment, a particular control structure may be tested via simulation. In this case (where this paper is focused) the real process is represented by its "complete" rigorous mathematical model as a "heart" of an independent simulator, while the software controller may run on a separate hardware host, both machines being interconnected via standard signals.

As to the proposed general simulator architecture, some problematic aspects must also be revealed. The main problem when modeling the fractionating process is to find the best model for the column, usually obtained by coupling the independent models for each tray, which generally consists in mass balance, energy balance and equilibrium equations. A good compromise between the results accuracy and a reasonable model dimension in order to require a non-prohibitive execution time for the integration routine may be made by using some classic simplifying assumptions [5]:

- the plant feeds can be divided in pure components or pseudo-components with known physical properties;
- the (pseudo) components are always perfectly mixed on column trays;
- the trays are ideal (theoretical trays), between the vapor and liquid phase there being an equilibrium state;
- the vapor holdup on each tray is negligible;
- the pressure profile is constant;

- the condenser with the accumulator is regarded as the first equilibrium stage (theoretical tray) in the column, while the column bottom, together with the reboiler) may be considered as the last tray.

The model for an entire plant, which may consist in several columns, sidestripers, pumparounds, condensers and accumulation tanks, is usually obtained from the models for each particular element, this technique being called the "structural approach" in the industrial plant modeling [5]. A general formulation of the mathematical model for the binary mixtures fractionating column (presented in Fig. 5) may have the following form:

$$A\frac{dS}{dt} = F(S, MVS, DVS, TVS, CVS), \tag{5}$$

where S is the generalized state variable, A represents the process time constants and F a generalized nonlinear function of the process/model inputs and the process/model state variables. Usually, S, A and F are in a matrix form.

Thus, regarding the generalized state S, a particular s_{ij} element represents the "i" state variable on the "j" tray. In their work, the authors use four independent state variables for each equilibrium stage: x_j (the light component mole fraction in the liquid phase), L_j (the liquid leaving the tray flow rate), V_j (the vapor leaving the tray flow rate) and T_j (the temperature) [6]. At the same time, the matrix function F represents in fact the model formulation, as particular mass, energy and equilibrium equations. Also, as a remark, A is usually a rare matrix (most of its elements being zero) and has a diagonal form, the model excluding any implicit correlations between $d(s_{ij})/dt$ elements.

In the authors' opinion, the model formulation, far from being a facile task, is only a small problem; many others must be revealed when analyzing the resulting model characteristics. First, the system is non-linear and stiff, because of the different time scales in the model, imposing serious limitations for the integration step in order to have a stable numerical solution. Thus, the best solution possible is to use a routine with variable step size, which permits a significant number of step reductions [6].

Then, the model has to be validated, in order to make sure that it can give practically relevant solutions. At the same, time this is a way to evaluate how the simplifying assumptions affect the modeled system response. In a "normal" situation, having reasonable dimensions and complexity, there are available a few analytical methods to study the inherent model properties: its solutions existence, uniqueness, continuity depending on the input data and especially their behavior in correlation with the physical sense. But unfortunately it is not our case and the only method to validate the model is to study very carefully its behavior during a significant number of simulations. The authors' opinion is that such a model has a good accuracy when there is an obvious concordance between the obtained results and the data directly taken from the real plant, taking into account the good sense remarks from the plant operators, too.

Another problem could be how to choose an appropriate integration method. Obviously, an explicit "recipe" on "how to generally choose" cannot be formulated, but some criteria may be taken into account: the system order, sensitivity, maximum error tolerance, the simulation hypothesis (with/without known consistent initial values), the inputs dynamic changes amplitude (requiring fixed/variable integration step size), inherent stability and so on. Routines as Runge-Kutta and Euler implicit type (like SDASSL, LIMEXS, RADAUS, EULERB, SDIRK4) are mentioned as best suitable for usual applications by the literature, in particular for the fractionating process simulation [7].

Then, the model equations being a differential algebraic system, the main problem is to determine consistent initial values for the integration (in fact to compute values of the algebraic variables which are consistent with the given initial values of the dynamic variables). A fractionating plant's "true" steady state, with practical relevance, is a special case where this is satisfied. But how this state can be obtained? One possible answer should be: by an arbitrary initialization of the dynamic variables followed by a long time horizon simulation, supposing it leads to a "true" plant steady state. This manner of work is suitable for simple systems, but in the case of the mentioned plant it systematically may fail to provide relevant results, this being a typical non-minimum phase system. For these complex plants, the authors suggest a hybrid methodology, by using a derived steady-state model to get initial values for a sub-set of the state variables, then initializing the dynamic simulator with these values and simulating the system until a real steady state is obtained [7].

V. FRACTIONATING PROCESSES FEEDFORWARD CONTROL

Feedback control loops performances are not satisfactory when we deal with very slow processes, as to the ones that take place in fractionating columns. These unsatisfactory performances are concerned mainly with the long duration of the transitory regime when quality conditions are not observed anymore. This leads to unconformities of the products obtained by fractionating, which results in financial losses.

Feedforward control represents a challenging way of maintaining the specified quality and increasing process efficiency.

Feedforward control structure needs to maintain x_D and x_B at the desired values when there are changes in F and/or x_F disturbances. From MVS there will be selected a pair from the following four variables B, D, V', L. Variables V' and L influence mass process transfer into column, while B and D influence the overall material balance.

For a feedforward control law implementation, the adequate mathematical relationships between the set points values x_D^*, x_B^*, the disturbances F, x_F and the pair of manipulated variables are needed.

The values for B and D (manipulated variables) can be calculated from fundamental mass balance relationships valid for the desired values x_D^* and x_B^*:

$$F = B + D, \tag{6}$$

$$Fx_F = Bx_B^* + Dx_D^*. \tag{7}$$

By eliminating D or B, the desired relationships are obtained, respectively:

$$B = F\frac{x_D^* - x_F}{x_D^* - x_B^*} \ , \qquad (8)$$

when B is the manipulated variable, or

$$D = F\frac{x_F - x_B^*}{x_D^* - x_B^*} \ , \qquad (9)$$

when D is the manipulated variable.

In order to determine the manipulated variables value corresponding to fractioning, simplified models such as those based on Fenske-Underwood-Gilliland (FUG) [4] or Douglas-Jafarey-Mc. Avoy (DJM) relationships [8] can be used.

FUG model implies determining reflux ratio R[1] from Gilliland correlation [4], expressed by the following function:

$$\frac{N - N_{\min}}{N+1} = f\left[\frac{R - R_{\min}}{R+1}\right], \qquad (10)$$

where: N, N_{min} are number and minimum number of theoretical trays;

R, R_{min} – reflux ratio and minimum reflux ratio.

The parameter R_{min} is determined from Fenske-Underwood relationship

$$N_{\min} = \left[\ln\frac{x_D^*(1-x_D^*)}{x_B^*(1-x_B^*)}\right](\ln\alpha)^{-1}, \qquad (11)$$

and N_{min} from Underwood relationships:

$$R_{\min} = \frac{\alpha x_D}{\alpha - \theta} + \frac{1 - x_D}{1-\theta} - 1, \qquad (12)$$

$$\theta = \frac{\alpha}{1 + x_F(\alpha - 1)} \ . \qquad (13)$$

As regards f function, the following form is proposed in the paper [9]:

$$Y = 9.025\cdot 10^{-2} X^2 - 0.648X + 5.708\cdot 10^{-3} X^{-1} - \\ - 7.606\cdot 10^{-5} X^{-2} + 0.541 \qquad (14)$$

$$\text{where } X = \frac{R - R_{\min}}{R+1} \text{ and } Y = \frac{N - N_{\min}}{N+1} \ . \qquad (15)$$

The determination of reflux ratio R makes it possible to determine the respective L and V' values:

$$V' = D\cdot(R+1) \, , \qquad (16)$$

$$L = (F - B)\cdot R \, . \qquad (17)$$

By using (8), (9), (16) and (17) we can determine the steady state values of B, D, V', L manipulated variables. However, feedforward control needs dynamics values for these variables. This is imposed by the necessity of a synchronism between the effect of disturbances and of controllers' outputs on the controlled variables. The dynamics of transmitting the two effects is considered to have remarkable practical results. It is described by the by 1st order with dead time transfer function

$$H(s) = \frac{e^{-\tau s}}{as+1} \, , \qquad (18)$$

or, in time domain,

$$a_{MV}\frac{dMV(t)}{dt} + MV(t) = MV_{st}(t - \tau_{MV}) \, , \qquad (19)$$

where: $MV(t)$, MV_{st} are the current and, respectively, the steady state values, for the selected MV;

a_{MV}, τ_{MV} – time constant and dead time for MV channel.

Current dynamic values of MV are obtained by solving the differential equation (19), which represents the dynamic section of feedforward control model.

The described model is parametrized, the modification of α and N parameters allowing model tuning when there are differences between current and desired values for the controlled variables.

A key problem dealing with simplified models for feedferward control is that of their validation. In the work example at the end of this paper we will describe a method of testing simplified control models.

VI. WORK EXAMPLE

For practically testing the theoretical results presented above, after years of experiments in this field, the authors propose a complex hardware/software platform, reproducing in a very close manner the real situation from the petro-chemical processing plants. The system architecture is detailed in Fig. 7.

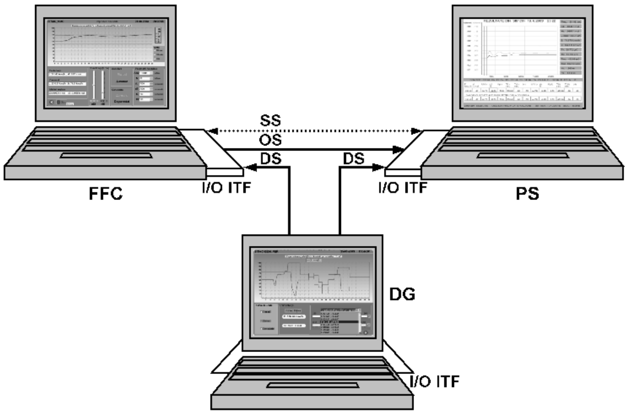

Fig. 7. The testing platform architecture: PS – Process Simulator; FFC – Feed Forward Controller; DG – Disturbances Generator; I/O ITF – Input/Output Interface; DS – Disturbance Signals; OS – Output Signals; SS – Synchronization Signals.

The system core is the binary fractionating process simulator PS, which makes use of a complex process dynamic model, in accordance with the modeling principles presented in chapter IV. For better performances (especially a faster integration routine), PS is entirely written in FORTRAN, having a modular structure. It includes dedicated routines for:

- simulator initialization (here all specific information must be specified: the plant geometrical details, the mixture components physical properties, initial operating parameters and so on);
- system mathematical model translation (a FORTRAN-coded representation for the generic equation (5))
- initial steady state identification (as consistent initial conditions for the integration routine);
- process model integration (implicit-type Euler routine, with constrained variable stepsize);
- user interface implementation (which permits online operating parameters changes, during simulation).

The authors first released the simulator in 1997 [6] as an

[1] Reflux ratio R can be determined as R=L/D

independent application, but it was continuously improved. At present, it permits not only the standard interaction by computer keyboard and display, but also by analogical (voltage) signals, as the simulated process accepts now inputs from external controllers and offers connections with its integrated transducers [10]. This way, taking into account another simulator feature (infinite time horizon simulation), it can substitute the real plant for safe operating experiments using classical control equipment. At the same time, if the simulator is well tuned, connected and synchronized with the real process, it can offer a way for inferential measurements, making possible the substitution of some expensive transducers (like quality analyzers).

The advanced feedforward controller FFC is implemented on another hardware host, in a flexible manner that permits the user not only to interact with and to tune the control algorithm, but also to choose it from an algorithms open library (FUG, DJM). The FFC workstation is equipped, as the PS host, with a multi-purpose input/output interface I/O ITF, making possible the external communication via analogical and digital signals with the controlled process actuators and transducers.

As shown in Fig. 7, between PS and FFC there are established two communication ways. First, the controller's outputs *L* and *B* are applied to the column simulator by the OS lines. Then, to make possible a faster simulation (in compressed time) and taking into account the variable stepsize for the PS's integration routine, a synchronization bi-directional bus has to be established, PS and FFC communicating one with each other the real simulation time.

The third system's host is the disturbances generator DG. It is, in fact, an analogical signal generator, which gives both to PS and FFC the current values for column feed flowrate *F* and propylene concentration x_F. DG resides on a PC workstation equipped with a similar I/O ITF like PS and FFC.

Owing to the lack of space, the authors will present only two screen captions from the PS workstation while testing the FGU control model (Fig. 8) versus a classical (PID) feedback control scheme (Fig. 9) on a propylene-propane fractionating column, on which our work is focused since many years [3], [6], [9], [10]. This stands for an example of the platform's capacities. The column has 72 theoretical trays and the nominal operating point is at *F* between 205kmol/h and 270kmol/h, x_F between 0.5mole frac. and 0.7mole frac., $x_D^* = 0.92$ mole frac., $x_B^* = 0.7$ mole frac., $\alpha = 1.1238$.

For this column, Relative Gains Matrix Method shows that L should be allocated to control x_D and B should be allocated to control x_D (usually this is called the "L-B structure").

By decreasing the feed flow rate from 242.5kmol/h to 220kmol/h, both control structures try to keep the products quality as specified, but the FGU feedforward algorithm proves to assure a better system behavior (as stability and dynamic response).

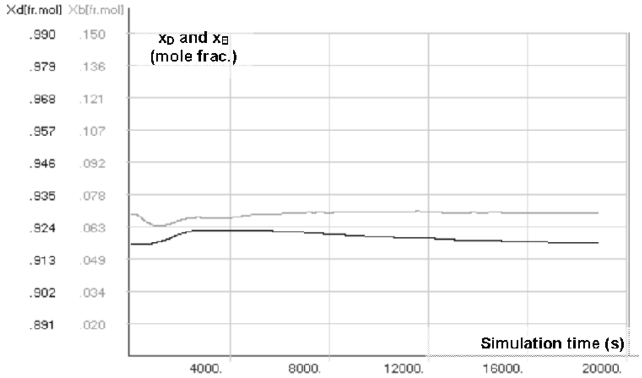

Fig. 8. Response when using the FGU feedforward control scheme.

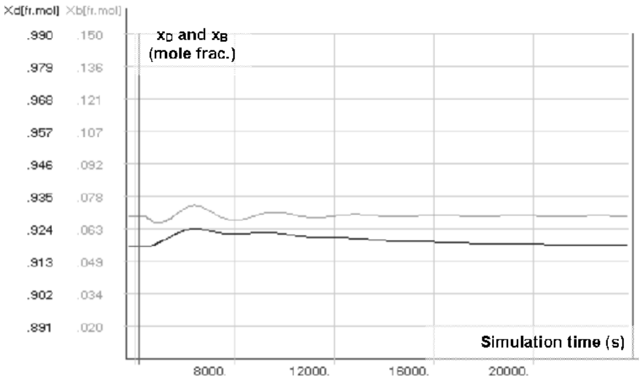

Fig. 9. Response when using a PID feedback control scheme.

VII. Conclusions

At present, for fractionating processes the advanced control schemes built on feedforward principle has the most important place within the advanced control structures. A feedforward control system implementation requires a few steps: first, an extensive (input-output) characterization of the controlled process, next, the process modeling, then the control algorithm design and implementation. Finally, the system behavior must be tested and validated via simulation.

This paper presented some aspects related to the feedforward control principle, fractionating process modeling and simulation, as well as feedforward control algorithm synthesis. For testing purposes, an integrated hardware/software platform originally developed and implemented by authors was also presented.

References

[1] N. Paraschiv and V. Marinoiu, *Automatizarea proceselor chimice*, vol. 2, Editura Tehnica, Bucuresti, 1992.

[2] N. Paraschiv, C. Marinescu and O. Cangea, "Stabilirea structurii buclelor de reglare asociate coloanelor de fractionare prin utilizarea tehnicii amplificarilor relative," *Romanian Journal of Informatics and Automatic Control*, vol. 7, no. 1, 1997.

[3] N. Paraschiv, *Echipamente si programe de conducere optimala a proceselor de fractionare a produselor petroliere*, PhD Thesis, Petroleum-Gas Institute of Ploiesti, 1987.

[4] C. Stratula, *Fractionarea. Principii si metode de calcul*, Editura Tehnica, Bucuresti, 1986.

[5] G. Radulescu, N. Paraschiv and V. Marinoiu, "A model for the dynamic simulation of a crude oil unit," *Control Engineering and Applied Informatics*, vol. 2, no. 1, 2000.

[6] G. Radulescu, V. Marinoiu and C. Marinescu, "Un simulator performant pentru investigarea reglarii automate a coloanelor de fractionare," *Romanian Journal of Informatics and Automatic Control*, vol. 8, no. 4, 1998.

[7] G. Radulescu, *Contributii la reglarea dupa model a procesului de distilare atmosferica*, PhD Thesis, Petroleum-Gas University of Ploiesti, 2002.
[8] A. Jafarey, M. J. Douglas, J. T. McAvoy, “Short-cut techniques for distillation columns design and control. 1. Column design,” *Ind. Eng. Chem. Process Des. Dev.*, vol 18, nr. 2, 1979.
[9] N. Paraschiv, “O forma analitica a corelatiei grafice Gilliland destinata modelelor de conducere automata evoluata a proceselor de fractionare,” *Romanian Journal of Chemistry*, no. 7-8, 1990.
[10] N. Paraschiv, G. Radulescu, D. Ionita, “Sistem integrat pentru testarea modelelor de conducere automata a proceselor de fractionare,” *Proceedings of the 27th Annual Congress of the American Romanian Academy of Arts and Science*, Oradea, 2002.