Chemical Technology, Control and Management

Volume 2019 | Issue 5 Article 5

12-28-2019

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Recommended Citation
Yusupbekov, Nodirbek Rustambekovich; Mukhitdinov, Djalol Patxiddinovich; and Kadirov, Yorkin Bakhodirovich (2019) "Modeling Of Multicomponent Rectification Under Uncertainty," *Chemical Technology, Control and Management*: Vol. 2019 : Iss. 5 , Article 5.
Available at: https://uzjournals.edu.uz/ijctcm/vol2019/iss5/5

ISSN 1815-4840
CHEMICAL TECHNOLOGY.
CONTROL AND MANAGEMENT
2019, №6 (90) pp.37-42
International scientific and technical journal
journal homepage: https://uzjournals.edu.uz/ijctcm/
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MODELING OF MULTICOMPONENT RECTIFICATION UNDER UNCERTAINTY

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Abstract: *This paper presents a complete computer model for an industrial distillation column, which considers the input, output and model (nonlinearity and high-frequency dynamics of the process) uncertainties. The proposed model has a great advantage of covering the entire working range of the distillation column. This model can be used as a main tool for the analysis and synthesis of distillation process control systems.*
Keywords: *mathematical model, rectification, uncertainty, nonlinearity, dynamic mode, input and output uncertainty, model uncertainty, robust control, multiplicative control.*

Аннотация: *Ишда кириш, чиқиш ва моделлаштириш (жараённинг ночизиқлилиги ва юқори частотали динамикаси) ноаниқликларини ҳисобга олган ҳолда саноат ректификациялаш колоннаси учун афзаллиги ректификациялаш колоннасининг тўлиқ иш диапазонини қамраб олганидан иборат тўлиқ компьютерли модель ишлаб чиқилган. Ректификациялаш жараёнини бошқариш системасини таҳлил қилиш ва синтезлаш учун асос сифатида фойдаланилиши мумкин.*
Таянч сўзлар*: математик модель, ректификация, ноаниқлик, ночизиқлилик, динамик режим, кириш ва чиқиш ноаниқлиги, моделли ноаниқлик, робастли бошқариш, мультипликатив бошқариш.*

Аннотация: *В работе разрабатывается полная компьютерная модель промышленной ректификационной колонны с учетом входной, выходной и модельной (учитывающей нелинейность и высокочастотную динамику процесса) неопределенностей, имеющая то преимущество, что она охватывает весь рабочий диапазон ректификационной колонны. Она может быть использована в качестве основы для анализа и синтеза систем управления процессом ректификации.*
Ключевые слова: *математическая модель, ректификация, неопределённость, нелинейность, динамический режим, входная и выходная неопределенность, модельная неопределенность, робастное управление, мультипликативное управление.*

The mathematical model considering the dynamic mode of the rectification process (whether it is linear or non-linear) can reflect only an approximate behavior of a real distillation column. The non-linear model can be used to control the process in a fairly wide range of operating conditions, and the use of the linear model in such conditions is limited by its error, which significantly increases with the deviation of the process parameters from its designed technological regulations, resulting form the process nonlinearity. Since the process is stochastically affected by the external environment, it is impossible to determine the error of its linear model [1, 2].

When developing mathematical models of distillation columns, the typical uncertain sources and events are as follows: the unknown structures and parameters of phase equilibrium models under production conditions; the impact of unmeasured disturbances in the composition of the raw material; variation of delay time and hydrodynamic regimes of the process; nonlinear characteristics of the interaction between the temperature profile control loops in the column-type systems, measurement errors, neglecting high-frequency dynamics in the process model, and non-linearity of the process [3,4,5]. All these uncertain sources exist simultaneously and can be classified as input uncertainty, model uncertainty, which should consider non-linearity and high-frequency dynamics of the process, and output uncertainty, i.e. measurement errors of output parameters.

The robust control systems [1] can be employed in order to save the output variables and error signals of the system within the specified acceptable intervals (despite the presence of uncertainties in the controlling process) [1].

In multiplicative uncertainty [2, 6], system “input-output” model can be represented in the following way:

$$\begin{pmatrix} y_1(s) \\ y_2(s) \\ \dots \\ y_N(s) \end{pmatrix} = \begin{pmatrix} g_{11}(s)g_{12}(s)\dots g_{1M}(s) \\ g_{21}(s)g_{22}(s)\dots g_{2M}(s) \\ \dots \\ g_{N1}(s)g_{N2}(s)\dots g_{NM}(s) \end{pmatrix}.$$

$$\left(\begin{pmatrix} 10\quad 0 \\ 01\dots 0 \\ \dots \\ 00\dots 1 \end{pmatrix} + \begin{pmatrix} \delta g_{11}(q)\ \delta g_{12}(q)\dots \delta g_{1M}(q) \\ \delta g_{21}(q)\ \delta g_{22}(q)\dots \delta g_{2M}(q) \\ \dots \\ \delta g_{N1}(q)\delta g_{N2}(q)\dots \delta g_{NM}(q) \end{pmatrix} \cdot \begin{pmatrix} x_1(s) \\ x_2(s) \\ \dots \\ x_M(s) \end{pmatrix}\right), \tag{1}$$

Or for the input-output model:

$$y(s) = G(s)(E + \delta G(s,q))x(s), \tag{2}$$

where, $y(s), x(s),$ - are the output and input signal; G(s) - transfer matrix of the nominal system; $\delta G(s,q)$ - transfer matrix of not simulated dynamics; E - unit matrix of the corresponding dimension.

Usually the robust systems are used to control processes with an unknown or incomplete mathematical model containing uncertainties. In this case, a little change of the parameters of the controlling process causes a small variation of the output of the closed control system, Therefore, the system control must be “rigid”, not resistant to the changes in the controlling process.

In the rectification process, the actual values of irrigation and steam boiler load never exactly correspond to the values which the controlling system requires. In statics, in the calculations of irrigation flows and steam rising along a distillation column, the result of calculating the rectification process is taken as a first approximation, taking into account the mutual influence of the components at constant values of the vapor and liquid flows over the height of the column. The flow rates of steam G_j and liquid flow L_j at the (k)th iteration are derived using the heat balance equations. In this case, enthalpies is used which is calculated through the components and temperatures.

The main causes of errors between given values and true values of steam and irrigation flows are: static and dynamic errors of measurement of indicators of a reflux condenser and a boiler; a change in the heat of vaporization due to instability of pressure and temperature; inertia of the boiler, delay and high-frequency oscillations of actuators.

The uncertain interval of input vectors can be modeled with the help of the multiplicative description of uncertainty with frequency-dependent interval of the error W_{ui} for *i*-th element of a vector connected with Vu_L for a dephlegmator L and Vu_L for the boiler [1]. These borders can be combined in a scalar matrix V_u. In this case the following equation is calculated:

$$\tilde{u}(j\omega) = \{I + \Delta_u(j\omega)W_u(j\omega)\}u(j\omega) \tag{3}$$

$$\left\|\Delta_u(j\omega)\right\|_\infty \le 1 \tag{4}$$

$$W_u(j\omega)=\begin{bmatrix} w_{u_L} & 0 \\ 0 & e_{u_V(j\omega)} \end{bmatrix} \tag{5}$$

The uncertainty model is presented in Figure 1.

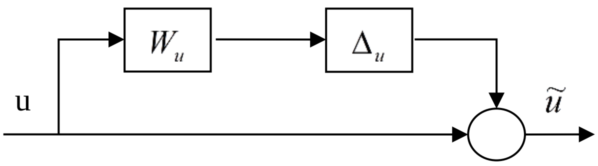

Figure 1. ***Multiplicative uncertainty describing the input vector.***

In practice, it is more likely that disturbances change over time, so the estimated and real error estimates are not equal [7]. The quality of any control system, ultimately, is determined by the magnitude of the error, equal to the difference between the required and actual values of the controlled quantity. To assess the accuracy of control in the synthesis of systems, error values are used in various typical modes, in which the perturbations are usually assumed to be constant and are considered to be applied to the output of the control object. In practice, it is more likely that disturbances change over time. Therefore, the estimated and real errors are not equal to each other [7].

In the paper [8] it is shown that the quality of control system for fine distillation depends significantly on the accuracy of measurement of controlled quantities. In order to design a control system, the error detection limits must be estimated with maximum accuracy.

In the low frequency range, the errors of the input controlled variables strongly depend on the measurement accuracy of technological parameters. Therefore, the steady-state error, together with the high-frequency error is proposed to describe the transfer function in the first order:

$$G(s)=K\frac{1+s/\omega_N}{1+s/\omega_D} \text{ или } \omega_N < \omega_D \tag{6}$$

where gain-coefficient K represents a steady-state error. Critical frequencies are usually selected according to $\omega_D > 10\omega_N$.

The mathematical description of the distillation column dynamics includes controlling variables (phlegm L, steam V) and several sources of disturbances, of which the most important are: changes of the input flow components and the input flow rate [4]. Instability of the input flow rate and its components leads to nonlinear behavior of distillation columns. When considering a simplified model of the dynamics of the liquid composition on the plate without taking into account the feed flow

$$\frac{dx_j}{dt}=\frac{1}{n_j}[L_{j-1}(x_{j-1}-x_j)+V_{j+1}(y_{j+1}-x_j)-V_j(y_j-x_j)] \tag{7}$$

it can be seen that the nonlinear behavior of the components depends on the actual flow rates of irrigation (L) and steam(V) (due to the high sensitivity of the distillation columns to changes in the speed of these flows; even though input flow change is very small, the response can go beyond the linear region) and the actual concentration profile of the components along the distillation column height.

When the operating range of the distillation column is limited to the maximum and minimum values of the flow rate and composition, maximum internal flow rates are observed for the minimum values of the mixture composition and reverse. The component concentration profiles for these two limited conditions restricts the region of stable state of controling system. Hence, it can be concluded that the distillation column in the low-frequency mode provides a high purity of the binary mixture distillation (at maximum and minimum load). The following three operating modes of binary rectification are proposed as a basis for further discussion:

Model 1 columns at rated load

Model 2 columns with maximum power and minimum concentration of power components (increased load)

Model 3 columns with minimum power and maximum concentration of power components (load reduction)

Data on different modes are given in table 1.

Table 1.

Considering Model	The value of the supply flow mol/min.	Concentration of supply components mol / mol
Model 1 (M-1)	*30*	0,6
Model 2 (M-2)	50	0.5
Model 3 (M-3)	20	0.7

One of the reasons for representing the column nonlinearity (involved with change of setting) is the multiplicative uncertainty of the product. Then the assumption of independence of uncertainty for each product from the actual value of the other ones can be displayed as follows (Fig. 2 and Equations (8)):

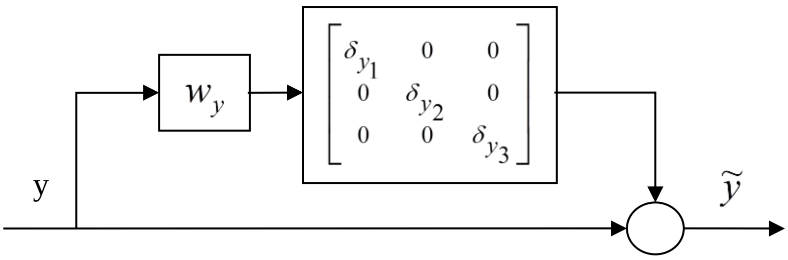

*Figure 2: **Multiplicative uncertainty at the output.***

$$\tilde{y}(j\omega)=\left\{I+\begin{bmatrix} \delta_{y_1}(j\omega) & 0 & 0 \\ 0 & \delta_{y_2}(j\omega) & 0 \\ 0 & 0 & \delta_{y_3}(j\omega) \end{bmatrix} W_y(j\omega)\right\} y(j\omega) \tag{8}$$

or

$$\left\|\delta_{y_1}\right\|_{\infty} \leq 1$$

and

$$y(j\omega)=G_N(j\omega)\begin{bmatrix} d(j\omega) \\ u(j\omega) \end{bmatrix}$$

The matrix W_y is a diagonal matrix with uncertainty limit for each product on the main diagonal. An upper limit for these uncertainties can be obtained by calculating the standard errors $\Delta G_2(j\omega)$ and $\Delta G_3(j\omega)$ (for each channel $u_i \to y_i$ or $d_i \to y_j$ of mode $\Delta G_2(j\omega)$ and $\Delta G_3(j\omega)$ respectively:

$$\Delta G_2(j\omega)=\left[G_2(j\omega)-G_N(j\omega)\right]G_N^{-1}(j\omega) \tag{9}$$

$$\Delta G_3(j\omega)=\left[G_3(j\omega)-G_N(j\omega)\right]G_N^{-1}(j\omega) \tag{10}$$

The upper limit of the weight error is the maximum of all normalized output errors .

In [9]. it is noted that the representation of the nonlinearity of the distillation column in the form of a simple multiplicative uncertainty limit at the output is not entirely justified.

Calculations show that the multiplicative uncertainty at the output exceeds 80% (for) in the low frequency range. This is significantly less than in the midrange, but increases dramatically at frequencies above 0.1 rad / min, where flow dynamics affect column behavior. The description of uncertainty with such a high multiplicative uncertainty in the low frequency region is very high for any control system structure.

At the same time, it should be noted that the errors are strongly correlated: a change in the stationary operating points causes a simultaneous increase or decrease in the singular values of the transfer functions from the control signals D (D and M) to the output model. Thus, we can assume that the dynamic behavior of the distillation column must be " somewhere between M-2 and M-3.", which can be represented as a linear combination of two models and (11) (figure 3)

$$G(j\omega) = \frac{G_2(j\omega)+G_3(j\omega)}{2} + \delta_G(j\omega)\frac{G_2(j\omega)-G_3(j\omega)}{2} \tag{7}$$

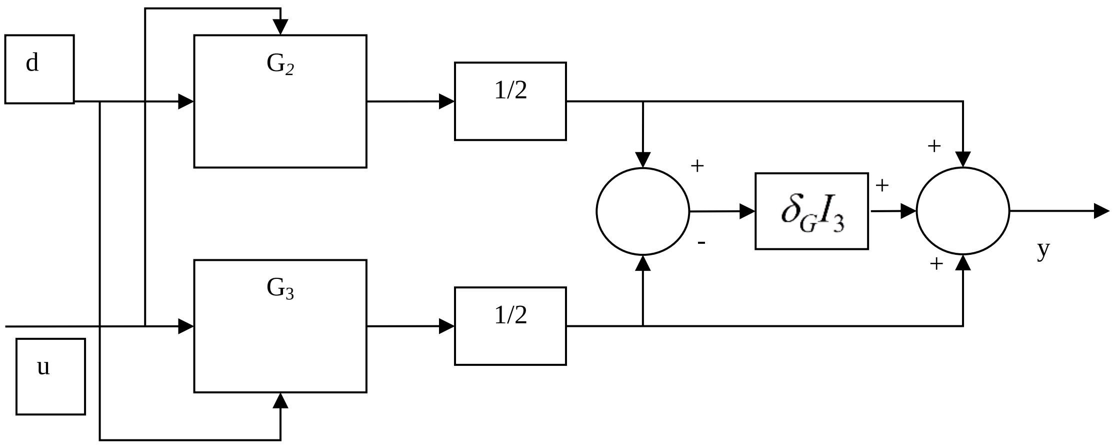

*Figure 3: **uncertainty Model due to nonlinearity in the low frequency range.***

The uncertainty parameter can be complex. In this case, it is appropriate to assume a possible phase shift for all models in the interval between G3 and G2. GRT means that the calculated values of all models in a certain set of Nyquist hodograph for a certain frequency may not lie on the same line.

On the basis of the given material the full model of uncertainty is made (Fig.4), consisting of input uncertainty (3) and model uncertainty (8,11), which has the advantage that it covers the entire operating range of the distillation column.

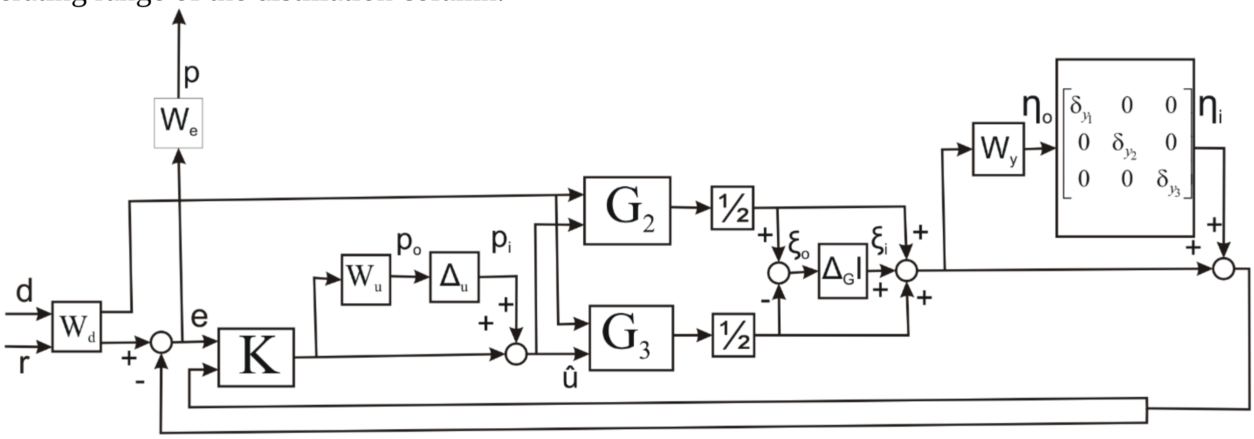

*Figure 4. **A complete model of uncertainty.***

The above suggests that a computer model has been developed for an industrial distillation column with account of certainties (input, model, which should take into account the nonlinearity and high-frequency dynamics of the process, and output, i.e. measurement errors of the output parameters). The developed model covers the entire operating range of the distillation column. It can be used as a basis for analysis and synthesis of rectification process control systems.

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