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Analysis of the Dynamic Characteristics of the Distillation Column of Plate Type

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Abstract

The relevance of the study related to the necessity to convert to "unmanned" production in a refinery and, as a consequence, we must derive a mathematical model of the high-octane gasoline process. This paper introduces a calculation procedure for modelling and dynamic analysis of a condensate distillation (rectification) column using by the mass balance structure. In this study, the reflux rate and the boil up rate are used as the inputs of the purity of the distillate overhead and the impurity of the bottom products. The dynamic analysis is accomplished over two phases: the basis nonlinear model of the plant and the full-order linearized model. This information will be used for modelling simulation to verify the applicable ability of the dynamic model and then to analyse and synthesize of the predictive model of the automatic control system of the rectification column.

Keywords: distillation column, mass balance, mathematical model, dynamics.

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1. Introduction

Automation of different types of production is an important area of scientific and technological development of the modern society, because it leads to eliminating the laborious and monotonous works and, finally, increases the quality of the product and capacity [1]. Today, one of the most relevant problems in oil and gas companies is an automation of technological processes and production due to the escalating competition in the hydrocarbon market and the necessity to reduce production expenses. Among the large variety of technological mass transfer processes in the petrochemical industry the most widely used the absorption, distillation, extraction, crystallization, ect. [3]. All of them are characterized the strict requirements for the mixture and quality of the product.

Main difficulties connected with developing of automatic control system of the distillation plant occur as a result of the peculiarity of a control object which is a large-sized device characterized by great values of time constant and lag [2]. The objective of the control of the considering plant is to achieve a given precision of separation of a multicomponent mixture with the highest possible intensity. The objective of the control system of distillation column is to maintain the product outputs concentrations despite the disturbance in the feed flow and the feed concentration. Changing of the feed flow rate, as a major factor had the negatively influence on the quality of a product, is not subject to stabilization [5]. Consumptions of the superheated stream, distillate, refrigerant and heating vapor are controlled inputs.

In practice, the multi-stage distillation process is carried out in the form of counter current distillation (rectification) in a column. The liquid mixture to be separated is fed to the bottom of the column, where it is brought to the boiling point. The vapor produced moves upwards inside the column, exists it at the top and is condensed. Part of the condensate is carried away as top product. The remainder flow back into the column and moves down-wards as liquid opposite phase.

The relevance of the study is connected with the transition to the concept of "unmanned" production at the plant JSC "Gaspromneft-Omsk Refinery" Omsk, Omsk region. Thus, in new conditions the control of rectification

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process has to include not only permanent monitoring of process parameters with aim to efficient operator intervention in case of exceeding the admissible fluctuations of control outputs from the given ones, but also be able to predict the current and future states in real time using appropriate mathematical models of the object.

The goal of this paper is to present a theoretical calculation procedure of a distillation column for simulation and analysis as an initial step of a project feasibility study.

2. Statement of the problem

The column under considering is design with $N = 49$ trays. A feed can be considered as a pseudo binary mixture of Ligas (iso-butane, n-butane and propane) and Naphtas (iso-pentane, n-pentane, and higher components).

The following assumptions were used for formulating the mathematical model of the distillation plant (JSC “Gazpromneft-Omsk Refinery”, Omsk, Omsk region) to analyze the dynamic characteristics [4]:

- The liquid holdups on each tray, the consider, and reboiler are constant and perfectly mixes;
- The flow of vapor, when it is in contact with liquid in the plates, can be described by using the hydrodynamic model of ideal displacement;
- The carryover of liquids from plates are absent by the vapor;
- The volumes of vapor and liquid streams do not change along the column height. This corresponds to separation of the components having similar boiling points and evaporation temperature;
- Reboiler works as a partial evaporator.

Using available information about features of technological process (fig. 1) we have to carry out the dynamic analyze of the distillation column, work out the mathematical model as a system of differential equations describing the dynamic properties through the different channels of controlled and disturbing impacts.

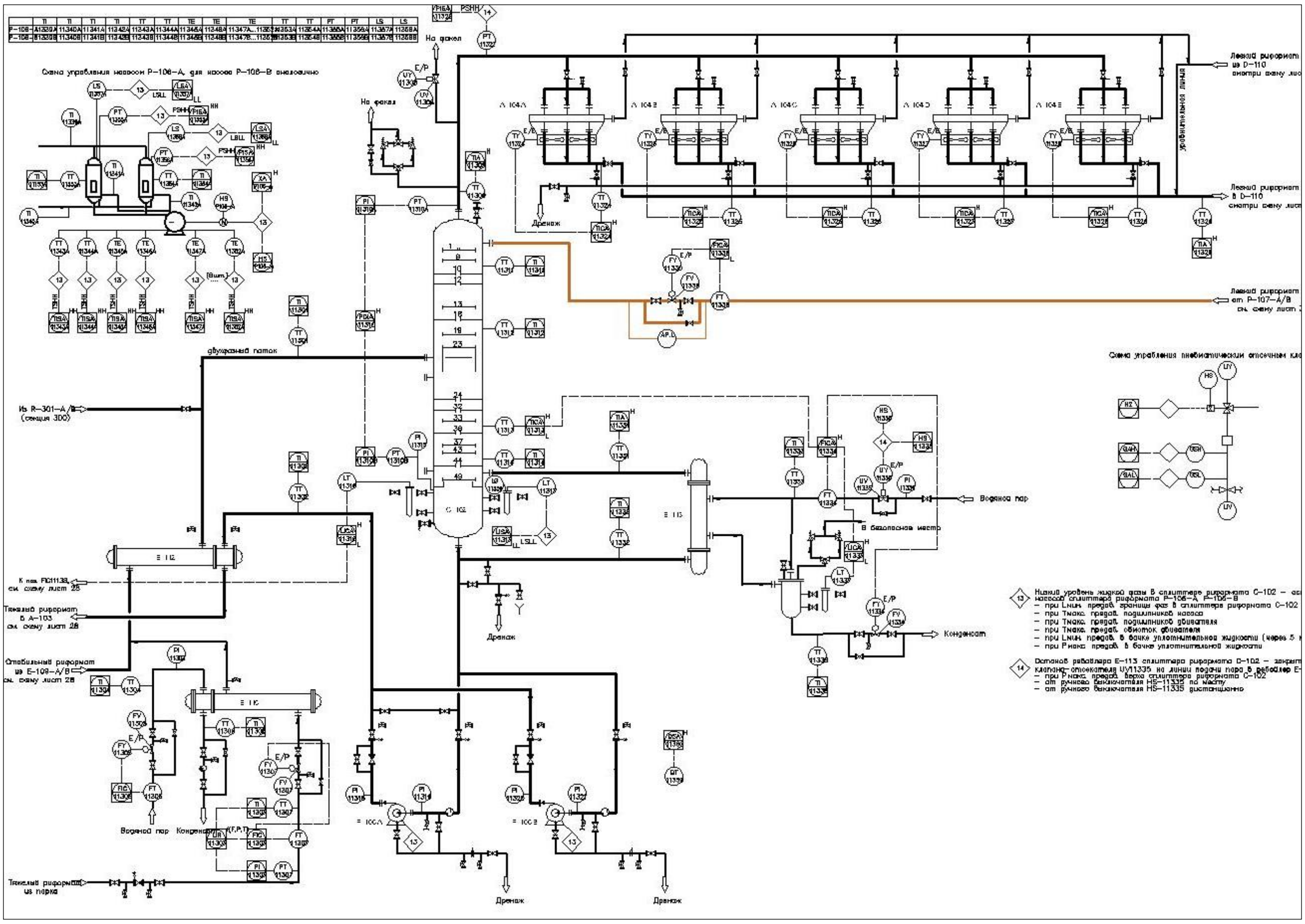

Figure 1. Functional scheme of automation of rectification process

3. Analysis of dynamic characteristic

For the rectifying section, the typical pressure drop per tray is 6.75kPa. Thus, the pressure at the top section is 4 atm.

For the stripping section, the base pressure is approximately equal to the pressure of the feed section therefore the pressure drop across this section is neglect.

If there is ideal mixing, changing of mixture of the liquid phase at a separate isolated tray will be depend on the variation of mixture of the vapor phase (or flow rate) and will be described using the differential equation of the first order [5, 7]. The value of the time constant is obtained from the empirical chart of phase equilibrium (fig. 2).

Using linear approximation the dependence between the volatile component concentrations in the liquid and vapor phases will be derived as:

$$w_V = a + bw_L \quad . \tag{1}$$

Where w_V, w_L (%) are volatile component concentrations in the liquid and vapor phases; a, b - unknown parameters of tangent passed through the point whose position depends on the selected mode of the plant.

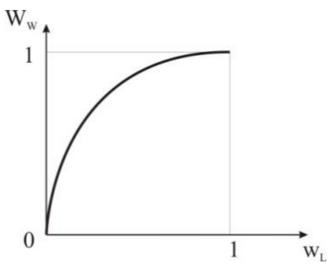

Figure 2. Equilibrium flash vaporization (EFV) curve

Mass balance equation for the volatile component in each tray is formed as follows:

- at the first tray:

$$m_{L1}\frac{dw_{L1}}{d\tau} = F_F w_F - F_B w_B + F_{L2} w_{L2} - F_{V1} w_{V1}; \tag{2}$$

- at the ith tray:

$$m_{Li}\frac{dw_{Li}}{d\tau} = F_{Vi-1} w_{Vi-1} - F_{Li+1} w_{Li+1}; \tag{3}$$

- at the last tray:

$$m_{Ln}\frac{dw_{Ln}}{d\tau} = F_{Vn-1} w_{Vn-1} - F_{Ln} w_{Ln} - F_D w_D. \tag{4}$$

Where $F_F, F_{Li}, F_{Bi}, F_B, F_D$ (kg/h) are masses of the flow rate in a feeding section (crude oil), vapors at the top of ith tray, liquid at the bottom of ith tray, liquids in a reboiler and a reflux drum, respectively. Masses in a fluid phase in each tray are denoted as $m_{L1},..,m_{Ln}$ (kg). Time and a number of trays in a distillation column are designated as τ (h) and n correspondently.

Under that $w_B = w_{L1}$; $w_{V1} = a_1 + b_1 w_{L1}$; $w_D = w_{Vn} = a_n + b_n w_{Ln}$, the dynamic model (2) – (4) can be expressed by following equations:

$$\begin{cases} m_{L1}\dfrac{dw_{L1}}{d\tau} = F_F w_F - F_B w_{L1} + F_{L2} w_{L2} - F_{V1}(a_1 + b_1 w_{L1}); \\ \dots \\ m_{Li}\dfrac{dw_{Li}}{d\tau} = F_{Vi-1}(a_{i-1} + b_{i-1} w_{Li-1}) - F_{Li} w_{Li}; \\ \dots \\ m_{Ln}\dfrac{dw_{Ln}}{d\tau} = F_{Vn-1}(a_{n-1} + b_{n-1} w_{Ln-1}) - F_{Ln} w_{Ln} - F_D(a_n + b_n w_{Ln}). \end{cases} \tag{5}$$

In order to obtain a linear control model for nonlinear system (5), we assume that the variables deviate only slightly from some operating conditions [1]. Then the nonlinear Eq. (5) can be expanded into Taylor's series. If the variation $x_n - \overline{x_n}$ is small, we neglect the higher-order terms. Rearranging variables $x = w_F - w_F^0$; $y_1 = w_{L1} - w_{L1}^0$; $y_i = w_{Li} - w_{Li}^0$; $y_n = w_{Ln} - w_{Ln}^0$, transform the system of equation (5). The linearization of the distillation leads to a 49th-order linear model in state space form:

$$\begin{cases} m_{L1}\dfrac{dy_1}{d\tau} = F_F x - F_B y_1 + F_{L2} y_2 - F_{V1} b_1 y_1; \\ \dots \\ m_{Li}\dfrac{dy_i}{d\tau} = F_{Vi-1} b_{i-1} y_{i-1} - F_{Li} y_i; \\ \dots \\ m_{Ln}\dfrac{dy_n}{d\tau} = F_{Vn-1} b_{n-1} y_{n-1} - F_{Ln} y_n - F_D b_n y_n. \end{cases} \tag{6}$$

By using Laplace transform techniques [1], system of differential Eq. (6) is rearranged into the system of algebraic equations to obtain the transfer function for each tray of the distillation apparatus:

$$\begin{cases} m_{L1} s\ell(y_1) = F_F \ell(x) - F_B \ell(y_1) + F_{L2}\ell(y_2) - F_{V1} b_1 \ell(y_1); \\ \dots \\ m_{Li} s\ell(y_i) = F_{Vi-1} b_{i-1} \ell(y_{i-1}) - F_{Li}\ell(y_i); \\ \dots \\ m_{Ln} s\ell(y_n) = F_{Vn-1} b_{n-1}\ell(y_{n-1}) - F_{Ln}\ell(y_n) - F_D b_n \ell(y_n). \end{cases} \tag{7}$$

Symbol ℓ is denoted the direct Laplace's transform. According with dependences (7), the transfer function of each tray can be express by zero-pole form:

$$W_1(s)=\frac{\ell(y_1)}{\ell(x)}=\frac{K_1(Ts+1)}{(T_1s+1)(T_2s+1)};$$
$$W_2(s)=\frac{\ell(y_2)}{\ell(x)}=\frac{K_1K_2}{(T_1s+1)(T_2s+1)};$$
$$W_3(s)=\frac{\ell(y_3)}{\ell(x)}=\frac{K_1K_2K_3}{(T_3s+1)(T_1s+1)(T_2s+1)};$$
...
$$W_i(s)=\frac{\ell(y_i)}{\ell(x)}=\prod_{j=1}^{i}\frac{K_j}{(T_js+1)}; \qquad (8)$$
$$W_n(s)=\frac{\ell(y_n)}{\ell(x)}=\prod_{j=1}^{n}\frac{K_j}{(T_js+1)}.$$

Empirically established that the largest numerical value of time constant in Eq. (8) corresponds to parameters T_1 of first tray. Parameter T_1 can be estimated from the following quadratic equation:

$$\left(\frac{m_{L1}m_{L2}}{F_BF_{L2}}\right)s^2+\left(\frac{m_{L1}(F_{L2}+F_B)+F_{V1}m_{L2}b_1}{F_BF_{L2}}\right)s+1=0 \qquad (9)$$

In order to approximately state the value of time constant, we assume that parameter b_1 in Eq. (9) is equal to one, so the approximation straight line will be passes through the point on the equilibrium curve (fig.2) and will be inclined at the angle 45°. In this case the time constant T_1 will be approximately equal to the average time taken τ_{aver} and can be estimated as a ratio between the volume of liquid at the first tray and flow rate in a feeding section (crude oil). Increasing the reflux ratio ($R=F_R/F_D$) almost equally effects on the process at all trays as well as reducing the value of time constants, which tend to zero in case of high degree of irrigation.

It was experimentally established that the transitional process for the mixtures are changed immediately at the trays of under considering rectification column which are located near to the feed flow section. Thus the obtained experimental curve can be approximated through a first-order linear differential equation. The mixture at the bottom and top of the column are varied with delay exceeding time taken at the tray in ($k\cdot n$) times, where n is quantity of trays between the feeding section and corresponding tray, factor k depends on the total number of tray and is normally selected at $k=(0.5\div 1)$.

4. Conclusion

In this study, the system identification has been employed using the distillation column of plate type at JSC “Gaspromneft-Omsk Refinary”. A mathematical model describing the dynamic behavior of the control object is made by using a mass balance equation.

It is found that the process in each plate of distillation column under considering is represented by a differential equation of the first order, that corresponds to typical inertial link with a time constant approximately equal τ_{aver}. (a ratio between the volume of liquid at the first tray and flow rate in a feeding section). Due to the high response rate of the pressure control loop, it does not influence as a dynamic disturbances on the associated temperature control loop. Because of the interplay between the latter one and level control loop the mass flow rate of the distillate can make significant fluctuations, so it is recommended to carry out all required corrections as slow as possible. Pressure fluctuations make the control more difficult and reduce the performance so assume that column operates at a constant pressure.

According to the experimental information about the distillation process at JSC “Gaspromneft-Omsk Refinary” the existed control system does not achieve the operational objective and nominal capacity of the plant will get less nominal value dealing with disturbance. Offering adaptive control system will be compensated for the variations in the characteristics of the process. Adaptive control is widely used in petroleum industries because of two reasons: firstly, most of processes are nonlinear: secondly, most of the processes are not stationary. Both of these facts lead to adapt the changing control parameters [6, 7]. Further, a reduce-order linear model will be derived such that it best reflects the dynamics of the distillation process and used as the reference model for a model-reference adaptive control system to verify the applicable ability of suggested model for a distillation column dealing with the disturbance and the model-plant mismatch as the influence of the plant feed disturbance.

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