Modified approach to distillation column control

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Abstract
This paper contains methodology research for forming the control algorithm for a distillation column, modeled as TITO (two-input two-output) process. Its modified form was obtained by connecting the two parts, and this combination hasn't been applied for such a industrial plant, until now. These parts include: a simplified decoupler which was first designed and decentralized PID controller obtained using D-decomposition method for such decoupled process. The decoupler was designed in order to make process become diagonal, and parameters of PID controllers are defined for the two SISO (single-input single-output) processes starting from relation between IE (integral error) criterion and integrator gain, taking into account desired response characteristics deriving from technological requirements of controlled plant. Their connecting provides centralized control. Analysis of the processes responses, obtained by the proposed algorithm and their comparison with the results from the literature, were performed after the completion of the simulations. The proposed approach to the centralized controller design, beside its simplicity of usage and flexibility in achieving diversity of process dynamic behavior, gives better response characteristics, in comparison with existing control algorithms for distillation column in the literature.

Keywords: distillation column, decentralized controller design, decoupling control, PID control, D-decomposition, simulation.

Available online at the Journal website: http://www.ache.org.rs/HI/

SCIENTIFIC PAPER

UDC 66.048.011:519.1

Hem. Ind. **71** (3) 183–193 (2017)

Process industry abounds in plants like a distillation columns. Therefore, their control attracts great attention of researchers. Directions of improvement, beside always accented system stability, robustness and performance, go towards the simplicity of control system implementation and its handling by operators. Nowadays, it is mainly trying to achieve using the PID controllers, so their tuning is subject of many researches. There are a lot of methods for controller tuning taking into account process multi-variability. Accordingly, the distillation column has been modelled as a two-input two-output (TITO) process. Some methods are used to design a controller that simultaneously performs process decoupling, *i.e.*, interaction compensation, while the other methods involve diagonal controller and separately designed decouplers. Because of the great possibilities for comparison with the results presented in the literature, which are listed below, as well as the complexity of the model in terms of loop interaction and time delays, binary distillation column (water–methanol) modeled by Wood and Berry [1] has been taken into consideration in this paper.

Tuning method for finite number of SISO (single-input single-output) PI controllers is derived in [2] and based on Ziegler–Nichols method, in the case that input–output variable pairing problem has been already solved.

The same number of decentralized relay feedback tests as the order of process transfer function matrix has been used, for the estimation of its frequency response matrix [3]. Based on this matrix, decoupling control is made, *i.e.*, controller with PI diagonal elements and PID off-diagonal elements is formed. Consideration of the time delays and eventual non-minimum phase zeros of the open-loops, was precondition for decoupling control design based on internal model (IMC) and model reduction, which simplifies theoretical controller for easier usage in the practice [4]. A strategy that is seen in the socio-political frameworks in imperial struggle for colonies has been applied to achieve the convergence of PID controllers' parameters towards their best values regarding integral absolute error (IAE). This approach, called "Colonial Competitive Algorithm" was used for multivariable PID controller design, which decouples controlled process [5]. The inverted decoupling within centralized control for TITO process is another approach with PI diagonal controller and

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Paper received: 26 March, 2016
Paper accepted: 22 June, 2016
https://doi.org/10.2298/HEMIND160326028P

filtered derivative compensators plus time delay as its off-diagonal elements [6].

Besides the foregoing references relating to the controller tuning methods, that contain process decoupling action, the following gives an overview of studies with emphasis on methods for separate decoupler design. Iterative tuning method for decentralized PID controller and decoupler realized under condition that the transfer function matrix of a generalized process is diagonal is shown in [7]. Often utilized strategy is static decoupling, with decentralized controllers designed through the integral gain maximization, by making a compromise between the performance of the individual control loops and their interaction [8]. A decoupler introduced in [9] enables independent control of loops while decentralized PI or PID controller is formed using an iterative numeric algorithm. An ideal PI decoupler, designed with respecting condition that open loop transfer matrix become diagonal, was approximated by the controller with four PI elements without time delay, and desired system performance have been obtained by tuning proportional gain [10]. One more control algorithm with inverted decoupling is derived in [11], where PI controllers are designed for separate control loops. It is noteworthy that, besides aforementioned strategies for distillation column control, whose variations were briefly presented in the above listed papers, the research in this field has various directions. For example, a neuro-fuzzy model utilization for distillation plant control, where ethanol is separated after corn starch fermentation [12]. Further, a multivariable PID controller based on Lyapunov quadratic index variation and characteristic matrix eigenvalues has been formed for distillation column control [13]. In recent times, the robustness of the time-delay systems is very often emphasized, either they are considered as a multivariable processes like in [14–16] or as a SISO processes in [17].

Proposed algorithm involves combination of simplified decoupler [11,18], and decentralized PID controller design using D-decomposition method [19–25] with introduced possibility to emphasize, *i.e.*, to constrain, some indicators of process behavior. Namely, two independent PID controllers for time-delay system are designed in this paper. Due to complexity of the transfer function of process with delay time, approximation of delay time is performed, *i.e.*, that process is described by high-order process of finite dimension [26]. One of the contributions of the present paper is solution how the PID controller parameters estimation for high-order linear systems of finite dimension can be effectively applied to the system with a time-delay which is of infinite dimension. Another contribution, especially in relation to the reference [22], is reflected in the introduction of derivative term and therefore complete PID controller design, unlike [22], where PI controller was designed in the K_p–K_i parameter plane. In this research K_p–K_d parametric plane is formed for the PID controller parameters estimation, while integral gain K_i is observed as one of the parameters that meets the requirement of minimum IE (integral error) criterion.

The rest of the paper has been arranged as follows. In section Decoupler design method for decoupling of distillation column process has been applied, while section Controller design shows an adapted PID controller design method that uses criteria defined by desired characteristics of given technological process. Section Results and discussion contains the proposed algorithm for the distillation column control, with analysis and discussion of the results obtained by numerical simulation of this multivariable system. Conclusions and possibilities for extension of proposed method application are given in the homonymous section.

DECOUPLER DESIGN

Well-known decoupling control structure is shown in Figure 1.

In this approach, the goal of 2×2 process decoupling is to obtain two independent SISO systems and it is necessary to design PID controllers for them.

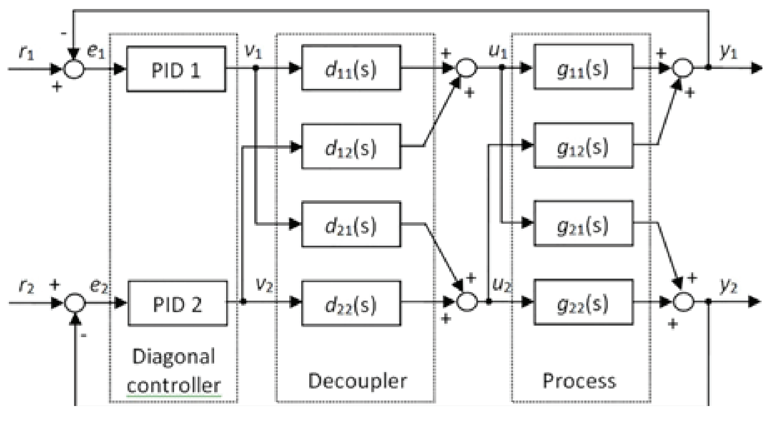

Figure 1. Centralized control as a combination of diagonal decentralized PID controller and decoupler [10].

The general form of 2×2 process transfer function matrix is:

$$G(s)=\begin{bmatrix} g_{11}(s) & g_{12}(s) \\ g_{21}(s) & g_{22}(s) \end{bmatrix} \tag{1}$$

while general expression for its decoupler is given by Eq. (2):

$$D(s)=\begin{bmatrix} d_{11}(s) & d_{12}(s) \\ d_{21}(s) & d_{22}(s) \end{bmatrix} \tag{2}$$

Independent SISO systems that should be determined are, in fact, the diagonal elements of the product $G(s)\cdot D(s)$ as a diagonal matrix.

Therefore, configuration in Figure 1 has to take the form like in Figure 2.

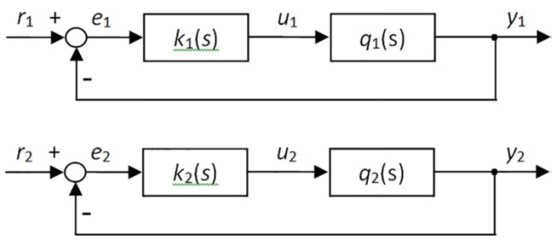

Figure 2. Decoupled system presented as two separated closed loops.

Diagonal matrix (apparent process) follows from Eq. (3) [11]:

$$\begin{aligned} Q&=\begin{bmatrix} g_{11} & g_{12} \\ g_{21} & g_{22} \end{bmatrix}\cdot\begin{bmatrix} d_{11} & d_{12} \\ d_{21} & d_{22} \end{bmatrix}= \\ &=\begin{bmatrix} g_{11}d_{11}+g_{12}d_{21} & g_{11}d_{12}+g_{12}d_{22} \\ g_{21}d_{11}+g_{22}d_{21} & g_{21}d_{12}+g_{22}d_{22} \end{bmatrix}=\begin{bmatrix} q_1 & 0 \\ 0 & q_2 \end{bmatrix} \end{aligned} \tag{3}$$

Laplace operator s was omitted in the Eq. (3) to make it shorter.

Simplified decoupling, of course, enables simpler expression for decoupler, but more complicated expression for the mentioned diagonal matrix (3), is given by:

$$\begin{aligned} Q(s)&=\begin{bmatrix} g_{11}(s) & g_{12}(s) \\ g_{21}(s) & g_{22}(s) \end{bmatrix}\cdot\begin{bmatrix} 1 & d_{12}(s) \\ d_{21}(s) & 1 \end{bmatrix}= \\ &=\begin{bmatrix} g_{11}(s)-\dfrac{g_{21}(s)\,g_{12}(s)}{g_{22}(s)} & 0 \\ 0 & g_{22}(s)-\dfrac{g_{21}(s)g_{12}(s)}{g_{11}(s)} \end{bmatrix} \end{aligned} \tag{4}$$

Eqation (4) presents TITO process as two independent SISO processes. They are used for controllers design, that need to control separate outputs.

CONTROLLER DESIGN

Controller, that should be design for the 2×2 process is given by Eq. (5):

$$\begin{aligned} K(s)&=\begin{bmatrix} k_1(s) & 0 \\ 0 & k_2(s) \end{bmatrix}= \\ &=\begin{bmatrix} K_{p1}+\dfrac{K_{i1}}{s}+K_{d1}s & 0 \\ 0 & K_{p2}+\dfrac{K_{i2}}{s}+K_{d2}s \end{bmatrix} \end{aligned} \tag{5}$$

Where are K_{p1} and K_{p2} are proportional gains, K_{i1} and K_{i2} are integral gains, K_{d1} and K_{d2} are derivative gains for the first and second closed loops, respectively. Elements $q_1(s)$ and $q_2(s)$ of apparent process (3) can often be of high order. Therefore, procedure proposed in [22], where tuning begins from the condition which establishes a direct relation between the IE criterion and integrator gain (higher integrator gain gives smaller value of IE criterion), has been adapted and applied in this research. The result has been extended by introducing the engineering specifications (relative stability and the settling time). In contrast to the [22], the main contribution of this paper is the introduction of derivative term, *i.e.*, member $K_d s$, in the controller transfer function. In accordance to that, the estimation of the controller parameters is performed in the K_p–K_d parametric plane under the condition that the integral gain K_i satisfies the requirement of minimum IE criterion ($K_{imax}= 1/IE_{min}$). So, while in [22] K_p–K_i parametric plane was in the focus, here, K_p–K_d parametric plane is formed for the different damping coefficients ξ, where, beside the natural frequency ω_n, integral gain K_i is parameter that is beforehand determined for the desired damping coefficient. This approach enables simple and efficient procedure for diagonal decentralized PID controller design for high-order systems, which can be applied without difficulties to the time-delay systems using Padé approximation of order n.

The transfer function of a SISO process obtained after decoupling is represented by:

$$q(s)=\frac{N(s)}{M(s)}=\frac{\sum_{k=0}^{m} b_k s^k}{\sum_{k=0}^{n} a_k s^k},\quad m\le n \tag{6}$$

In Eq. (6) a_k and b_k are numerator and denominator coefficients of the transfer function, respectively.

The characteristic equation of one closed loop system in Figure 2 is determined by Eq. (7):

$$f(s)=1+k(s)q(s)=0 \tag{7}$$

Joining Eqs. (5)–(7) gives Eq. (8):

$$f(s)=1+\left(K_\mathrm{p}+\frac{K_\mathrm{i}}{s}+K_\mathrm{d}s\right)\frac{N(s)}{M(s)}=0 \quad (8)$$

It follows that:

$$f(s)=sM(s)+\left(K_\mathrm{d}s^2+K_\mathrm{p}s+K_i\right)N(s)=0 \quad (9)$$

After invocation of:

$$f_1(s)=sM(s)=\sum_{k=0}^{n}a_k s^{k+1} \quad (10)$$

into Eq. (9), the final expression for the system characteristic equation in the complex domain is obtained.

$$f(s)=f_1(s)+\left(K_\mathrm{d}s^2+K_\mathrm{p}s+K_\mathrm{i}\right)N(s)=0 \quad (11)$$

Consequently, it is necessary to express the Laplace operator *s* in a suitable form and through it, establish a relation between the damping coefficient ξ and variable controller parameters K_d, K_p and K_i contained in the characteristic Eq. (11). So, region from the "*s*" plane under straight line ξ = const. (Figure 3), is mapped into the region of the corresponding damping coefficient, represented by curve for the ξ = const. in the parameter plane of the adjustable controller parameters (K_d,K_p).

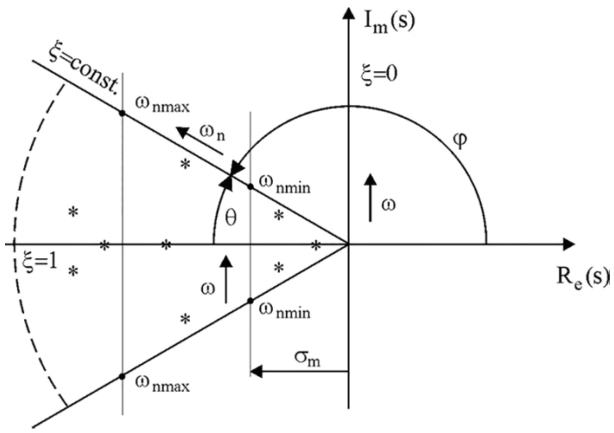

Figure 3. Region with required settling time and relative stability [22].

Knowing that:

$$s=-\omega_\mathrm{n}\xi+j\omega_\mathrm{n}\sqrt{1-\xi^2},\quad \xi=\cos\theta \quad (12)$$

Where ω_n is natural (undamped) frequency, and σ_m (Figure 3) is real part of the transfer function poles. Introducing of Eq. (12) into Eq. (11), gives system characteristic equation as:

$$\begin{aligned}&f_1(\xi,\omega_\mathrm{n})+[K_\mathrm{d}\omega_\mathrm{n}^2\left(\left(2\xi^2-1\right)-j\left(2\xi\sqrt{1-\xi^2}\right)\right)+\\&+K_\mathrm{p}\left(-\xi\omega_\mathrm{n}+j\omega_\mathrm{n}\sqrt{1-\xi^2}\right)+K_i]N(\xi,\omega_\mathrm{n})=0\end{aligned} \quad (13)$$

where

$$f_1(\xi,\omega_\mathrm{n})=\alpha(\xi,\omega_\mathrm{n})+j\beta(\xi,\omega_\mathrm{n}) \quad (14)$$

In Eq. (14) $\alpha(\xi,\omega_\mathrm{n})$ and $\beta(\xi,\omega_\mathrm{n})$ are real and imaginary part of polynomial $f_1(\xi,\omega_\mathrm{n})$, respectively. These are expressed through the Eqs. (15), (16) and (18):

$$\alpha(\xi,\omega_\mathrm{n})=\sum_{k=1}^{n}a_{k-1}(-1)^k\omega_\mathrm{n}^kT_k(\xi) \quad (15)$$

$$\beta(\xi,\omega_\mathrm{n})=\sqrt{1-\xi^2}\sum_{k=1}^{n}a_{k-1}(-1)^{k+1}\omega_\mathrm{n}^kU_k(\xi) \quad (16)$$

Expression:

$$\beta(\xi,\omega_\mathrm{n})=\sum_{k=1}^{n}a_{k-1}(-1)^{k+1}\omega_\mathrm{n}^kU_k(\xi) \quad (17)$$

is introduced because of the shorter writing. Now Eq. (16) becomes:

$$\beta(\xi,\omega_\mathrm{n})=\sqrt{1-\xi^2}B(\xi,\omega_\mathrm{n}) \quad (18)$$

T_k and U_k are Chebyshev functions of the first and second kinds. Following recurrent equations hold for them:

$$T_{k+1}=2\xi T_k-T_{k-1},\quad U_{k+1}=2\xi U_k-U_{k-1} \quad (19)$$

$$T_0=1,\quad T_1=\xi,\quad U_0=0,\quad U_1=1 \quad (20)$$

$$N(\xi,\omega_\mathrm{n})=\gamma(\xi,\omega_\mathrm{n})+j\delta(\xi,\omega_\mathrm{n}) \quad (21)$$

$\gamma(\xi,\omega_\mathrm{n})$ is real, and $\delta(\xi,\omega_\mathrm{n})$ is imaginary part of polynomial $N(\xi,\omega_\mathrm{n})$ and they are determined based on Eq. (22), (23) and (25):

$$\gamma(\xi,\omega_\mathrm{n})=\sum_{k=0}^{m}b_k(-1)^k\omega_\mathrm{n}^kT_k(\xi) \quad (22)$$

$$\delta(\xi,\omega_\mathrm{n})=\sqrt{1-\xi^2}\sum_{k=0}^{m}b_k(-1)^{k+1}\omega_\mathrm{n}^kU_k(\xi) \quad (23)$$

Introducing of:

$$D(\xi,\omega_\mathrm{n})=\sum_{k=0}^{m}b_k(-1)^{k+1}\omega_\mathrm{n}^kU_k(\xi) \quad (24)$$

into Eq. (23) gives Eq. (25):

$$\delta(\xi,\omega_n) = \sqrt{1-\xi^2}D(\xi,\omega_n) \tag{25}$$

Connecting Eq. (14)–(25) with Eq. (13), and thereafter suitable mathematical transformations and separating the real and imaginary parts, enables the following system of equations:

$$\begin{aligned} &K_d(\xi,\omega_n)\omega_n^2(2\xi^2-1) - K_p(\xi,\omega_n)\xi\omega_n = \\ &= -\frac{\alpha(\xi,\omega_n)\gamma(\xi,\omega_n)+\beta(\xi,\omega_n)\delta(\xi,\omega_n)}{\gamma^2(\xi,\omega_n)+\delta^2(\xi,\omega_n)} - \\ &-K_i(\xi,\omega_n) \\ &2K_d(\xi,\omega_n)\omega_n^2\xi - K_p(\xi,\omega_n)\omega_n = \\ &= -\frac{\alpha(\xi,\omega_n)D(\xi,\omega_n)+\gamma(\xi,\omega_n)B(\xi,\omega_n)}{\gamma^2(\xi,\omega_n)+\delta^2(\xi,\omega_n)} \end{aligned} \tag{26}$$

Solving a system of Eqs. (26) at $\omega_n \neq 0$ and $0 \leq \xi < 1$ provides expressions for the parameters of the PID controller K_d, K_p and K_i respecting the requirement that the integral gain meets the minimum of IE criteria, with a limit on the required damping coefficient of the closed loop. Aiming simplicity, this process is supported by MATLAB software. From the system of equations (26), for $\omega_n = 0$, the singular straight lines described by Eq. (27) are defined:

$$K_i(\xi,0) = 0 \tag{27}$$

Graphic interpretation of Eq. (26) is curve in (K_p–K_d) parametric plane for the selected damping coefficient under the condition that the integral gain K_i meets minimum of IE criterion. This curve and singular straight lines given by Eq. (27) represent enclosed area of possible solutions for the PID controller parameters.

$$G(s) = \begin{bmatrix} \frac{12.8}{16.7s+1}e^{-s} & \frac{-18.9}{21s+1}e^{-3s} \\ \frac{6.6}{10.9s+1}e^{-7s} & \frac{-19.4}{14.4s+1}e^{-3s} \end{bmatrix} \tag{29}$$

Here are outputs (controlled variables): $X_D(s)$ – percentage of methanol in the distillate, $X_B(s)$ – percentage of methanol in the bottom products, while manipulated variables are: $R(s)$ – reflux flow rate and $S(s)$ – steam flow rate in the reboiler. Therefore, the concrete example was taken for testing and analysis of the centralized control as a combination of simplified decoupling and diagonal decentralized controller, that consists of two PID controllers tuned by D-decomposition method. Accordingly [11], decoupler (30) has been calculated using simplified decoupling procedure. Its elements for the considered process (29) are given by Eq. (31):

$$D(s) = \begin{bmatrix} 1 & d_{12}(s) \\ d_{21}(s) & 1 \end{bmatrix} = \begin{bmatrix} 1 & \frac{-g_{12}(s)}{g_{11}(s)} \\ \frac{-g_{21}(s)}{g_{22}(s)} & 1 \end{bmatrix} \tag{30}$$

$$\begin{aligned} d_{12}(s) &= 1.47\frac{16.7s+1}{21s+1}e^{-2s}, \\ d_{21}(s) &= 0.34\frac{14.4s+1}{10.9s+1}e^{-4s} \end{aligned} \tag{31}$$

For the explored distillation column (29) a diagonal matrix (4), that represents system without interactions, i.e. whose diagonal elements are transfer functions of two independent SISO systems, is given by:

$$Q(s) = \begin{bmatrix} \frac{12.8e^{-s}}{16.7s+1} - \frac{6.4(14.4s+1)e^{-7s}}{228.9s^2+31.9s+1} & 0 \\ 0 & -\frac{19.4e^{-3s}}{14.4s+1} + \frac{9.7(16.7s+1)e^{-9s}}{228.9s^2+31.9s+1} \end{bmatrix} \tag{32}$$

RESULTS AND DISCUSSION

Mathematical model of binary distillation column (water–methanol) is formed by Wood and Berry [1]. Since the model was proven as a good, it is extensively used, during recent decades in many investigations, as an object for checking the effectiveness of various control algorithms. It was experimentally obtained and expressed by Eqs. (28) and (29):

$$\begin{bmatrix} X_D(s) \\ X_B(s) \end{bmatrix} = G(s)\begin{bmatrix} R(s) \\ S(s) \end{bmatrix} \tag{28}$$

Taking into account the structure in Figure 2 PID controllers are designed, i.e. more precisely PID 1 for $q_1(s)$ and PID 2 for $q_2(s)$. In order to enable the design of K_p–K_d parametric plane and taking into account the structure and complexity of the transfer functions given by Eq. (32), approximation of time-delay system with the system of finite dimension was necessary. Approximation of delay was carried out using fourth-order Padé approximation, for the frequency range $\omega_n \in$ (0–1.2) rad/s, which is necessary for the design of K_p–K_d parametric plane. The results of this approximation are presented in the frequency domain in Figures 4 and 5 using MATLAB software. These figures show excellent agreement of frequency characteristics for the selected

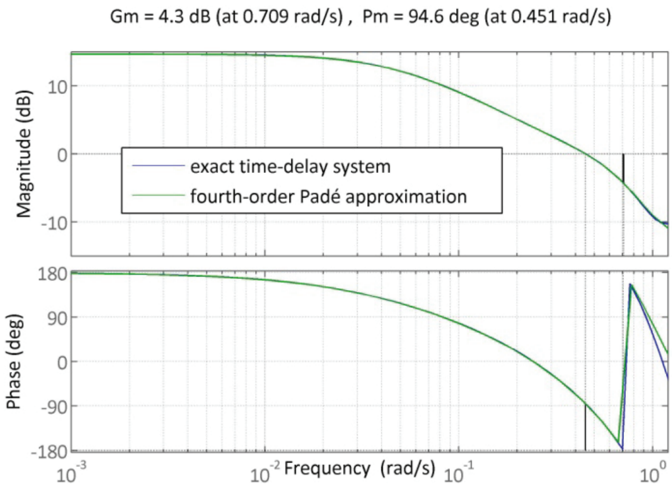

Figure 4. Comparative Bode diagrams of the exact time-delay system and approximated system.

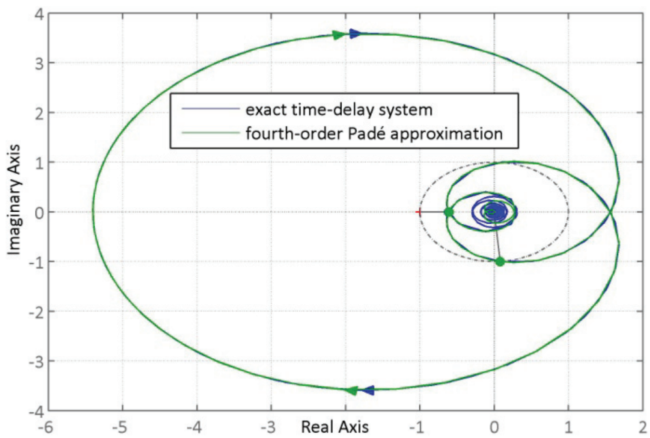

Figure 5. Comparative Nyquist diagrams of the exact time-delay system and approximated system.

frequency range, of the time-delay system approximated with fourth-order Padé approximation and the exact system. Accordingly, the possible disadvantages of the designed controller on the system of finite dimension should not show the shortcomings in the application to the system of infinite dimension (time-delay system).

In order to depict possibilities of this approach regarding satisfying a wide range of desired process performance, PID controllers are designed for two cases. In the first case, minimization of overshoot is required, while in the other the response speed is emphasized. Namely, borderline cases are presented, but the trade-offs between them are also obtainable, of course, based on technological requirements defined by the operators of this chemical plants. Simulations in the MATLAB software were carried out to obtain process responses.

Mentioned K_p–K_d parametric planes are given in Figures 6 and 7 for the first (PID1) and second (PID 2) controller, respectively, for the two values of damping coefficient (ξ = 0.5 and ξ = 1) and changing of natural frequency $\omega_n \in$ (0–1,2) rad/s.

Based on these planes for both controllers, K_p and K_d controller gains were determined from the curves that refer to the two different damping coefficients ξ = = 0.5 and ξ = 1, for a given value of the integrated gain K_i that meets minimum IE criterion for each value of damping coefficient. In terms of borderline cases of the system dynamic behavior in this area (Figures 6 and 7), the value of the damping coefficient ξ closer to 1 leads to the elimination of overshoot and reduction in the response speed, while a value closer to 0.5 makes system faster but with greater overshoot. Values between the above mentioned are used to accomplish compromise between these boundary cases to meet the technological requirements of the process.

First case (minimizing overshoot – damping coefficient ξ = 1)

In this case following controller parameters were calculated PID 1 (green line in Figure 6): K_{p1} = 0.2302; K_{i1} = 0.0575; K_{d1} = 0.1004 and PID 2 (green line in Figure 7): K_{p2} = –0.0675; K_{i2} = –0.01912; K_{d2} = –0.01. The resulting controller and decoupler (31) were applied in the centralized control strategy shown in Figure 1. Obtained responses were compared with four cases from the literature realized using different decoupling and controller tuning and shown together in Figure 8.

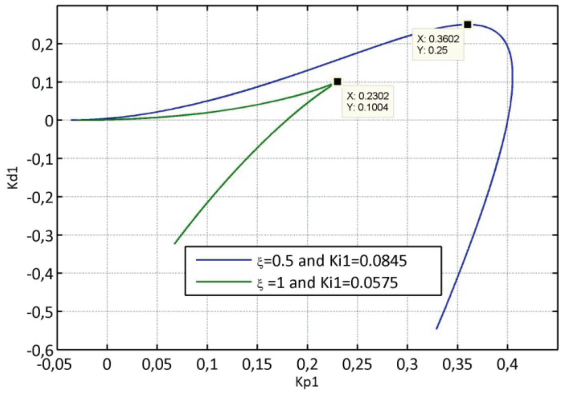

Figure 6. K_{p1}–K_{d1} parametric plane for the two values of damping coefficient (ξ = 0.5 and ξ = 1).

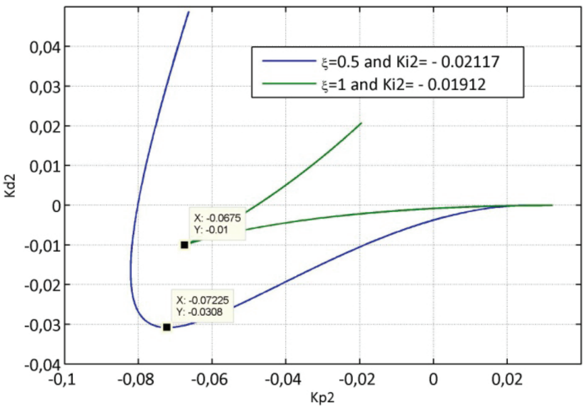

Figure 7. K_{p2}–K_{d2} parametric plane for the two values of damping coefficient (ξ = 0.5 and ξ = 1).

Figure 8 shows that proposed method is more efficient than the others, regarding overshoot reduction in both outputs. It also provides less oscillatory responses. In Figures 9 and 10, the second output is postponed by 50 s in order to system interaction become more evident. Now, it is noticeable that the proposed method provides satisfactory decoupling and taking into account the responses in Figure 8, it can be concluded that it is applicable in the practice.

Second case (faster responses - damping coefficient ξ = 0.5)

Controller parameters for this case are PID 1 (blue line in Figure 6): K_{p1} = 0.3602; K_{i1} = 0.0845; K_{d1} = 0.25 and PID 2 (blue line in Figure 7): K_{p2} = –0.07225; K_{i2} = = –0.02117; K_{d2} = –0.0308. Even under such requests for system performance, proposed method is more successful than the others in literature. By comparing the achieved results with the other control algorithms (Figure 11), it is obvious that the speed of both responses is improved, without significant deterioration of their overshoot and monotony. The feasibility of such a controller is confirmed by responses in Figure 10, where it can be seen that system interaction is reduced to a minimum.

More accurate comparison of the proposed control algorithm with methods from literature applied for distillation column control is enabled by Table 1. Namely, it contains a comparative view of rise time, settling time and overshoot. Even though some of the methods give better certain quality indicators than proposed one, nevertheless, taking into account overall quality of both responses, *i.e.*, all displayed parameters and especially the defined requirement for controller design, it is evident that the proposed method provides responses with better characteristics.

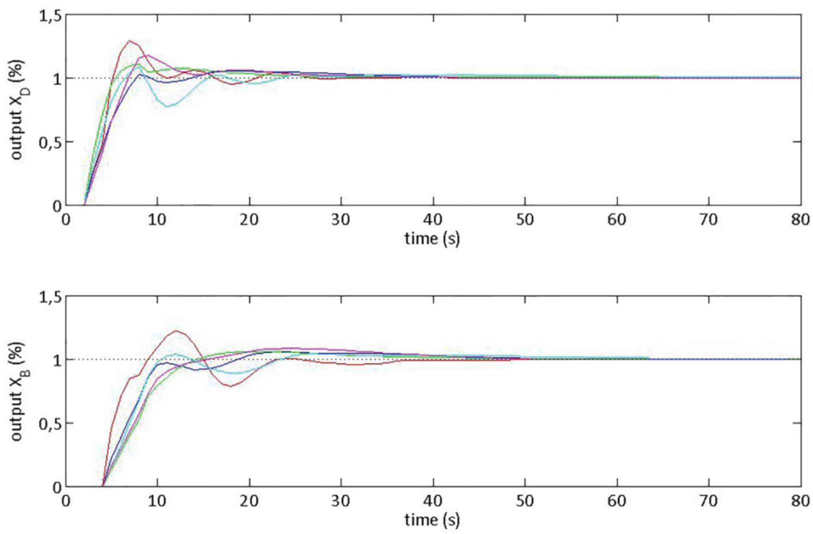

Figure 8. Responses of the Wood/Berry distillation column (minimized overshoot): ____ Atashpaz-Gargari et al. 2008, ____ proposed method, ____ Garrido et al. 2010, ____ Morilla et al. 2008, ____ Wang et al. 2000, _ _ _ Reference value.

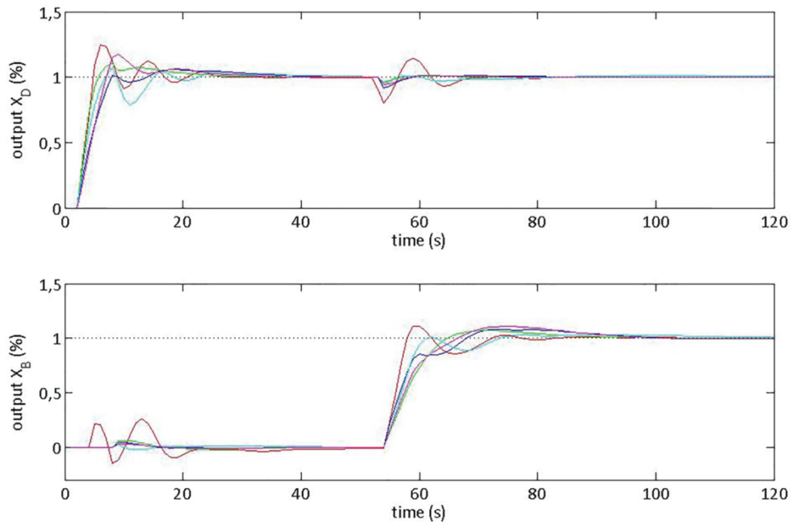

Figure 9. Display of interaction in the Wood/Berry distillation column (minimized overshoot): ____ Atashpaz-Gargari et al. 2008, ____ proposed method, ____ Garrido et al. 2010, ____ Morilla et al. 2008, ____ Wang et al. 2000, _ _ _ Reference value.

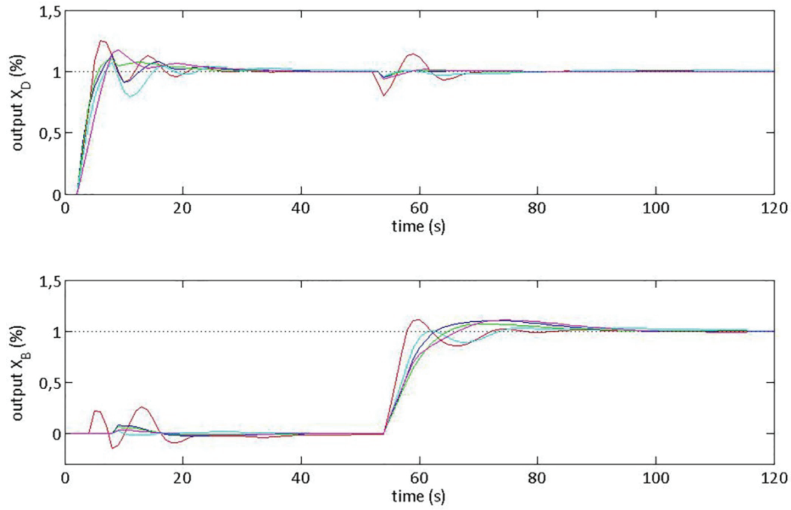

Figure 10. Display of interaction in the Wood/Berry distillation column (faster responses): ____ Atashpaz-Gargari et al. 2008, ____ proposed method, ____ Garrido et al. 2010, ____ Morilla et al. 2008, ____ Wang et al. 2000, _ _ _ Reference value.

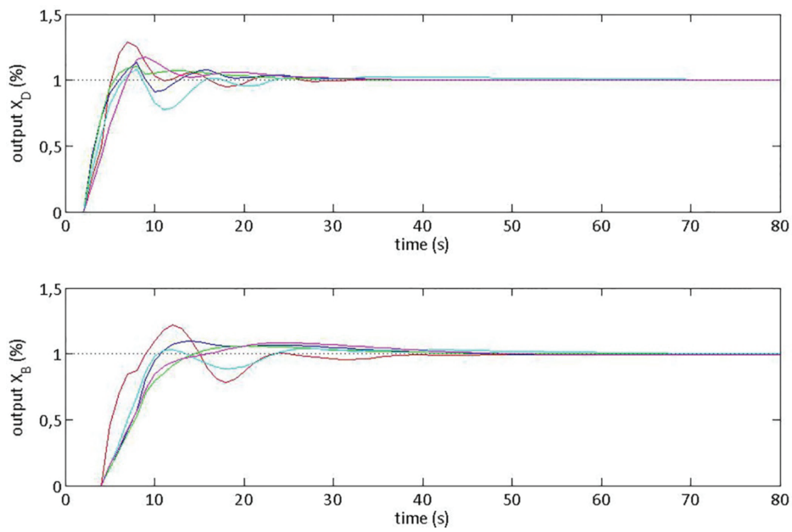

Figure 11. Responses of the Wood/Berry distillation column (faster responses): ____ Atashpaz-Gargari et al. 2008, ____ proposed method, ____ Garrido et al. 2010, ____ Morilla et al. 2008, ____ Wang et al. 2000, _ _ _ Reference value.

Table 1. The responses quality indicators of the explored distillation column under different control algorithms

Method	First output, X_D			Second output, X_B		
	Rise time T_r / s	Settling time T_s / s	Overshoot O / %	Rise time T_r / s	Settling time T_s / s	Overshoot O / %
Proposed (minimized overshoot ξ = 1)	4.381	33.639	5.339	5.002	42.322	5.684
Proposed (faster responses ξ = 0.5)	2.795	26.433	13.832	4.921	37.376	9.725
Atashpaz-Gargari *et al*., 2008	2.498	25.052	28.827	4.012	36.051	22.202
Garrido *et al*., 2010	2.544	22.885	10.354	7.006	34.759	5.864
Morilla *et al*., 2008	3.292	44.776	8.519	4.675	50.519	4.137
Wang *et al*., 2000	3.796	27.881	17.7	6.31	41.983	8.377

Table 1 shows that for the value of damping coefficient ξ = 1, overshoot is in the range of 5,339––5,684% at response speeds from 4,381 to 5,002 s. For the value of the damping coefficient ξ = 0.5 overshoot is in the range of 9,725–13,832% at response speeds from 2,795 s to 4,921 s. Having this in mind, change of the damping coefficient ξ, can enable satisfying of the different project requirements in terms of achieving adequate transition process quality of the observed system.

CONCLUSIONS

Approach to distillation column control, characterized by simplified decoupler synthesis and diagonal decentralized PID controller tuning based on D-decomposition method, is proved as good. The proposed method gives better process responses compared to the others applied for control of this process, and it does not complicate the designing procedure. It also provides good flexibility with regard to achieving of different system performances, defined in accordance with the technological requirements of industrial plants. These requirements can cause the system dynamic behavior between these two boundary cases, *i.e*., reduction or eventual elimination of overshoot on the one hand and the maximum achievable response speed on the other. Thus, performed research has introduced a way of making trade-off between these cases. Moreover, by implementing several iterations it is possible to design PID controllers, which provide the best value of one response characteristic under defined limitation on the any other one. This research, on the well-known and already verified model of distillation column, serves as proof that this algorithm can be used for other similar types of distillation columns. Foregoing facts support the acceptance of the proposed method by the operators of these plants, i.e. its introduction in the practice.

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IZVOD

MODIFIKOVANI PRISTUP UPRAVLJANJU DESTILACIONE KOLONE

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Postrojenja u procesnoj industriji su veoma važan i neiscrpan skup objekata za istraživanje sa stanovišta automatskog upravljanja. Osim stabilnosti, robusnosti i performansi sistema, teži se ka jednostavnosti implementacije upravljačkog sistema kao i njegovog rukovanja. U ovom radu je istražena metodologija formiranja upravljačkog algoritma za destilacionu kolonu, modelovanu kao proces sa dva ulaza i dva izlaza (TITO proces). Istraživani pristup upravljanju je provjeren koristeći već više puta u literaturi upotrijebljen matematički model pomenutog procesa. Prilagođavanjem i spajanjem dva metoda projektovanja njegovih komponenti, čija kombinacija do sada nije primjenjena za upravljanje ovog industrijskog postrojenja, dobijen je modifikovani upravljački sistem. To su pojednostavljeni rasprezivač koji se prvi projektuje i decentralizovani PID regulator dobijen pomoću metode D-dekopozicije za prethodno raspregnut proces. Rasprezivač je projektovan iz uslova dijagonalnosti procesa, a parametri PID regulatora su definisani za dva odvojena procesa sa jednim ulazom i jednim izlazom (SISO procesa) polazeći od veze između IE (*integrala greške*) kriterijuma i pojačanja integratora i uzimajući u obzir definisane željene karakteristike odziva, koje se mogu mijenjati u zavisnosti od tehnoloških zahtjeva upravljanog postrojenja. Njihovim sprezanjem ostvareno je centralizovano upravljanje. Naime, na proces sa kašnjenjem primijenjena je procedura za projektovanje dijagonalnog decentralizovanog PID regulatora za sisteme visokog reda. U okviru te procedure izražavajući Laplasov operator u odgovarajućoj formi preko stepena prigušenja i prirodne učestanosti uvedena je zavisnost između stepena prigušenja i promjenljivih parametara PID regulatora. Analiza odziva procesa dobijenih primjenom predstavljenog algoritma i njihovo poređenje sa rezultatima iz literature su sprovedeni nakon izvršenih simulacija. Ilustrovani su granični slučajevi dinamičkog ponašanja sistema: minimalni preskok i maksimalna brzina odziva. Zbog njihove dobro poznate kontradiktornosti predložen je metod za ostvarivanje kompromisa između ova dva slučaja izborom stepena prigušenja, s tim da se vodilo računa o granici stabilnosti sistema automatskog regulisanja. Dobijeni pristup projektovanju centralizovanog regulatora, pored svoje jednostavnosti primjene i fleksibilnosti u ostvarivanju različitih dinamičkih ponašanja procesa, daje bolje karakteristike odziva od postojećih algoritama upravljanja destilacionom kolonom u literaturi.

Ključne reči: Destilaciona kolona • Decentralizovano projektovanje regulatora • Rasprezanje • PID upravljanje • D-dekompozicija • Simulacija