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An Analytical Design Method of PI and PID Controllers for Industrial Series Cascade Processes With Time Delay Under Robustness Constraints

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Abstract—**In this paper, an analytical method is proposed to design Proportional-Integral (PI) and Proporional-Integral-Derivative (PID) controllers for two-stage industrial series cascade processes with transport delay under robustness constraints. The main rationale behind using series cascade control structure is that the disturbances in the inner loop are suppressed by the secondary controller before being transmitted to the outer loop. The presented design procedure consists of two steps: In the first step, the controller $C_b(s)$ in the inner loop is designed, while in the second step the controller $C_a(s)$ in the outer loop is obtained. The obtained controller is of PI or PID-type structure depending on the number of selected terms used in Maclaurin's approximation of the transfer function of the high-order controller. By specifying the robustness constraints within design procedure one define values of adjustable parameters to achieve compromise between robustness and performance indicators. The result is efficient suppression of load disturbances, evaluated by the integral of the absolute error (IAE). The step reference response can be additionally reshaped by using Two Degree of Freedom – 2DoF control structure via suitable selection of set-point weighting factor b, $0 \leq b \leq 1$, which acts on the control signal through displaced proportional action of the controller. The proposed design method is analyzed with simulations on wide class of typical representatives of industrial processes including stable, integral and unstable processes with time delay. A comparison with recent studies shows the effectiveness of the proposed tuning method for industrial cascade processes.**

Keywords- robustness; maximum sensitivity function; analytical controller design; cascade control;

I. INTRODUCTION

Cascade control structures are used in various industries with the aim of improving the efficiency of load disturbance rejection, reducing the sensitivity of the system to variations in process parameters, and generally improving the dynamic indicators of closed-loop system behavior [1]. Of particular importance is its application in the process industry, where the advantages of cascade control in the elimination of disturbances related to the control signal and when the control object (secondary process) has a nonlinear behavior are particularly emphasized [2]. Disturbance rejection in the process industry is of greater interest than set-point tracking in many process control applications. The reason for this is that set-point changes only occur when the production rate is changed.

In the standard two-stage series cascade control, there are two feedback loops with two controllers in each [3]. The controller in the inner loop is usually called the secondary controller (slave controller), while the primary controller (master controller) is located in the outer loop [4]. Some applications of series cascade control are: steam-fed water heaters [5], natural draft furnaces [6,7], polymerization reactors [8], etc. The basic concept of this configuration is that the disturbances in the inner control loop are suppressed by the secondary controller before being passed to the outer control loop. The benefits of using series control configurations are particularly emphasized under the following circumstances: when the inner control loop is faster than the outer control loop, when the inner control loop has influence on the outer control loop, and when the disturbances in the inner control loop are less severe than the disturbances in the outer control loop. As for the disadvantages of cascade control, industry consultants point to the additional cost of the extra sensor and controller that adds complexity to the control system. As a result, double tuning of the controllers is required. If the inner loop is too aggressive and the two processes operate on narrow time scales, two controllers can compete with each other, which can lead to instability of the control loop [9]. Performance improvement by applying cascade control over a conventional control loop is achieved when both controllers are adequately tuned.

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Most of the developed design techniques are related to conventional proportional-integral-differential (PID) controllers. Research studies have shown that about 94% of feedbacks in industry have a PI/PID structure [10], while in petroleum, chemical, pulp and paper industries, their percentage is over 97% [11,12]. Most of the developed design techniques refer to conventional proportional-integral-differential (PID) type controllers. Research studies have shown that about 94% of feedback in the industry have a PI/PID structure [10], while their percentage in the petroleum, chemical and pulp and paper industries is over 97% [11,12]. In a recent study [13], PID control was ranked first in terms of importance for industry, with the conclusion that efficiency and ease of implementation give PID algorithms preference over advanced control algorithms (predictive model control, intelligent nonlinear control, adaptive control, etc.). In this sense, different techniques have been developed for the design of the parameters of the primary and secondary controller, which can be generally divided into analytical and numerical (optimization) methods. Although numerical algorithms are more efficient in adjusting the parameters of controllers [14-19], they are computationally expensive compared to analytical rules and often cannot be easily adjusted by field engineers [20-21]. Therefore, various analytical design methods are available in the literature based on: configuration of the internal model control structure (Internal Model Control - IMC rules and their modifications) [22-25], the specification of the desired form of complementary sensitivity function of the closed-loop system [26-28], the application of the pole spectrum techniques [29,30], the use of D-decomposition techniques [31,32] and many other principles. Analogously, methodologies for adjusting the parameters of PI/PID controllers for the purpose of cascade regulation of industrial processes have been specifically developed.

Efficient analytical tuning methods for industrial cascade processes based on the IMC principle, among others, have been the subject of numerous studies [33-42], including the derivation of tuning rules considering robustness constraints in some of these researches. In [33, 34], both primary and secondary control loops controlled by 2DOF controllers are designed according to the IMC paradigm and tuning rules are presented. In [35], an automatic tuning method for a cascade control system is proposed, which allows simultaneous tuning of two PID controllers after estimating the process parameters by evaluating a set-point step response. Similarly, a single closed-loop step test is performed in [36] to identify the required process information using B-spline series representation for the step responses. The desired robustness levels guide the selection of IMC tuning parameters and allow full automation of controller tuning. In [37], an IMC-based method for tuning PID controllers while maintaining gain and phase margin specifications for cascade control systems is proposed.

The design technique using the IMC paradigm in [38] leads to a complete set of tuning parameters for the inner (2-DoF PI) controller and the outer (2-DoF PID) controller. The robustness of the cascade control system was analyzed in [39] using the structured singular value concept. In [40, 41], modified cascade structures are proposed to achieve improvements with both the IMC-based controllers and the process disturbances. Most of the aforementioned analytical design methods are related to a specific, narrower class of processes, while the method of direct synthesis of frequency-domain controllers for a broader class of processes has been recently proposed in [43,44]. The processes under considerations in this study include stable, integral and unstable time-delayed industrial processes. In [43], a proportional-integral controller is used in both the inner and outer control loops. The secondary controller is designed based on the desired transfer function of the closed loop system using the direct synthesis approach to achieve efficient suppression of load disturbances. A primary controller is then designed to provide good set-point tracking by considering the secondary process and the secondary controller as part of the primary process. The prefilter is added to remove unnecessary overshoot within the set-point response. A similar approach, defined as a combination of the direct synthesis approach and the pole placement method, was elaborated in [44] to achieve improved control performance. It should be noted that there are special control structures developed exclusively for systems with dominant transport delay [45], which are not the subject of discussion in the present work.

The subject of this work is the design of PI and PID controllers for industrial cascade processes with time delay, taking into account robustness requirements. The parameters of PI/PID controllers in the inner and outer loops of a series cascade control system are determined to achieve an adequate compromise between performance and robustness measures [46]. It is assumed that the pair of process models is known and described by the transfer function. The design procedure of PI/PID controllers, in particular for the inner and outer loop, is based on the determination of the high-order controller $C(s)$ [47], which is then approximated by low-order PI or PID controllers. In this paper, the controller $C(s)$ is defined for general process transfer function $G_p(s)$ given in rational form with time delay for specified complementary sensitivity transfer function $T_d(s)$ (where the subscript d denotes the "desired" form). The obtained results are tested by a series of numerical simulations with corresponding analysis of robustness and performance indicators for a wide class of industrial processes: stable, integral and unstable processes with time delay.

The merits of the present work are listed below:

i. The presented methodology is simple and straightforward, it is characterized by the flexibility to shape load disturbance and reference responses accordingly. It requires only two tuning parameters (one for each controller, in the primary and secondary control loops). The values of these parameters are adjusted to meet desired trade-off between performance and robustness. Moreover, the operator may have at its disposal the calculated value of the adjustable parameters under previously specified robustness constraint.

ii. The presented approach is applicable to a wide class of industrial processes including stable, integral and unstable processes with time delay. It does not require any reduction in the order of the process model and no loss of process dynamics which affects the quality of regulation. On the other side, for the derivation of tuning formulas in explicit form, it is desirable to have low-order models.

iii. The use of the proposed methodology to industrial series cascade processes leads to an improvement in the robust performance of the closed-loop system.

iv. The proposed methodology provides an efficient way for fine-tuning of the standard series cascade control systems, and the same methodology can be applied to multistage cascade control of industrial processes without loss of generality.

The rest of the paper is structured as follows. The general approach for designing controllers based on the specified complementary sensitivity function is given in Section 2. Then, in the remainder of the section, the presented methodology is demonstrated specifically for the series cascade control structure. In Section 3, the comparative simulation analysis is performed on the test batch consisting of ten pairs of representatives of the dynamical characteristics of industrial processes. Concluding remarks are given in the Section 4.

II. Analytical Design Methodology for PI/PID Controllers for Cascade Industrial Processes with Time Delay under Robustness Constraints

A. *The fundamentals of the general approach for controller design*

The simplified control structure with controller $C(s)$ is presented in Fig. 1. The following notation is used: r – reference signal, d – load disturbance at the input of process, n – measurement noise, y – output of the system, $G_{ff}(s)$ is the prefilter, and $G_p(s)$ is the transfer function of the controlled process.

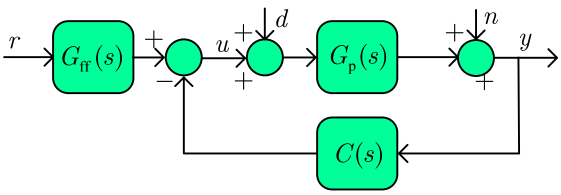

Figure 1. The simplified control structure

Among others, the approach of designing conventional controllers based on selected complementary sensitivity functions is one of the most common, characterized by its simplicity and flexibility in setting the desired trade-off between performance and robustness indicators. Various forms of complementary sensitivity functions have been considered in the literature [27-29, 33-38, 47]. In choosing the desired form of the transfer function two competing requirements must be met. First, the number of tunable parameters should be as small as possible while it is desirable to tune different closed-loop characteristics independently [21].

The complementary sensitivity function of the control system in Fig. 1 is given by the relation:

$$T(s)=\frac{L(s)}{1+L(s)}, \tag{1}$$

where $L(s)=C(s)G_p(s)$ is the loop transfer function. For the design of the PI/PID controller, we choose the desired complementary sensitivity function T_d as follows:

$$T_d(s)=\frac{N(s)e^{-\tau s}}{P(s)},\quad N(s)=1+\sum_{j=1}^{n}\eta_j s^j,\; P(s)=1+\sum_{k=1}^{p}\lambda_k s^k, \tag{2}$$

where $p \ge 2n$, $p,n \in N$ and adjustable parameters $\lambda_k > 0$, $k=\overline{1,p}$, $\eta_j \in R$, $j=\overline{1,n}$. The parameters η_j and λ_k are determined based on the desired behavior of the closed-loop system. Note that polynomial $P(s)$ is taken in wide scientific and professional literature in the form $P(s)=(\lambda s+1)^p$.

The controller $C(s)$ is defined from relations (2), for the process transfer function $G_p(s)$ with time delay τ, in order to achieve more efficient load disturbance suppression d, as well as, attenuation of measurement noise n [48]. From expressions (1) and (2), the transfer function of controller $C(s)$ is defined as:

$$C(s)=\frac{1}{G_p(s)}\frac{T_d(s)}{1-T_d(s)}=\frac{1}{G_p(s)}\frac{N(s)e^{-\tau s}}{F(s)}, \tag{3}$$

where $F(s)=P(s)-e^{-\tau s}N(s)$. In the general case, the parameters $\overline{\eta_1,\eta_n}$ are determined such that the poles of the process $G_p(s)$ for the nominal operating mode are eliminated with zeros of the function $F(s)$ [27, 47]. In this way, if the denominator of the complementary sensitivity function is chosen as follows: $P(s)=(\lambda s+1)^p$, (or in a similar manner), the parameters $\overline{\eta_1,\eta_n}$ are uniquely defined as a function of the single adjustable parameter λ, that should be selected to meet the desired design requirements.

For the purpose of obtaining low-order controllers from (3), the common approach is to use Maclaurin approximation [33,34]. In this way, the parameters of the PI controller defined by

$$C_{PI}(s)=k+\frac{k_i}{s} \tag{4}$$

are obtained by expanding the function $f(s)=sC(s)$ into Maclaurin series in s as in [15]. By using only the first two terms, the parameters of the approximation, i.e. $f(s)\approx f(0)+f'(0)s$, the integral and proportional gains are, respectively:

$$\begin{aligned} k_i &= f(0),\\ k &= f'(0) \end{aligned} \tag{5}$$

By following the described procedure, the parameters of the PID controller given by

$$C_{PID}(s)=\frac{k_d s^2+ks+k_i}{s(T_f s+1)} \tag{6}$$

are obtained by the Maclaurin approximation of the function $f(s)=s(T_f s+1)C(s)$. By preserving only three terms in the expansion, i.e. $f(s)\approx f(0)+f'(0)s+0.5f''(0)s^2$, parameters of the PID controllers are determined by relations:

$$\begin{aligned} k_i &= f(0),\\ k &= f'(0),\\ k_d &= \frac{f''(0)}{2}. \end{aligned} \tag{7}$$

The filter time constant can be defined as $T_f = \lambda / N$, where the value of parameter N is chosen taking into account the sensitivity to measurement noise. One possibility is to consider the maximum sensitivity to measurement noise $M_n = \max_{\omega} |S_n(i\omega)|$ [18], where $S_n(s)$ denotes sensitivity function with respect to the measurement noise:

$$S_n(s) = \frac{C_{PID}(s)}{1 + C_{PID}(s)G_p(s)} \tag{8}$$

In general, the parameter N can be chosen such that a change in the control signal caused by the measurement noise is acceptable [49].

It should be particularly noted that the previously described approximations of the complex (high-order) controller are suitable for processes where the transport (time) delay is not dominant with respect to the basic dynamics of the regulated process. Otherwise, controllers of more complex structure and appropriate control structures should be used to compensate for the time delays included in the characteristic equation of the control system to achieve the desired compromise between performance and robustness [37].

Hence, the parameters of the PI/PID controllers are expressed as a function of the free parameter λ, on the basis of which the desired indicators of robustness or closed-loop feedback system behavior can be achieved under previously defined conditions for processes where the transport delay is not dominant. The trade-off between the suppression of load disturbances and robustness can be expressed by the quantitative measure $M_s = \max_{\omega} |S(i\omega)|$, where

$$S(s) = \frac{1}{1 + C_x(s)G_p(s)} \tag{9}$$

is the sensitivity function for x=PI or x=PID. Following this procedure, the time constant λ should meet the condition:

$$\max_{\omega,\lambda} |1 / (1 + C_x(i\omega)G_p(i\omega))| = M_s, \tag{10}$$

so for stable control processes M_s should be within acceptable limits $1{,}2 \le M_s \le 2$ [42]. The evaluation of the robustness of the system to different types of uncertainties in the process such as modeling errors, can be estimated by the maximum value of complementary sensitivity function, i.e. $M_p = \max_{\omega} |T(i\omega)|$, where $T(s)=C_x(s)G_p(s)/(1+C_x(s)G_p(s))$, where x=PI or x=PID.

The efficiency of load disturbance rejection is evaluated on tha basis of Integral of Absolute Error – IAE, which is defined as follows:

$$\mathrm{IAE}_d = \int_0^{\infty} |e_d(t)|\, dt, \tag{11}$$

where $e_d(t) = \mathcal{L}^{-1}\{G_p(s) / (s(1 + L(s))\}$ is the response of the system to the unit step disturbance [39].

B. The proposed design approach for series cascade control structure

The idea behind using series cascade control structure is to supress disturbances before they affect the output signal (controlled variable This is not possible in the conventional single loop configuration, where disturbance suppression occurs when the output of the system deviates from the desired set-point. The effectiveness of cascade control is particularly evident when larger disturbances enter the secondary loop and the dynamics of the inner secondary loop are faster than the dynamics of the outer loop.

The structural block diagram of the series cascade control is shown in Fig. 2. The two control loops are nested, with the secondary (inner) loop inside the primary (outer) loop. The primary and secondary processes are described by the transfer functions $G_a(s)$ and $G_b(s)$, respectively. The corresponding controllers are denoted by $C_a(s)$ – primary (master) and $C_b(s)$ – secondary (slave) controller. The outputs of the controllers, i.e. the control signals, are denoted by u_a and u_b, respectively. The output of the primary control loop is denoted by y_1, while the output of the secondary control loop is denoted by y_2. The disturbances d_1 and d_2 are modeled to act on the input of the primary and secondary process models. The reference signal (set-point) of the primary control loop is represented by r_1, while the output signal of the primary controller y_2 serves as the set-point for the secondary controller. The measurement noise of the primary control loop is denoted by n_1, that of the secondary control loop by n_2.

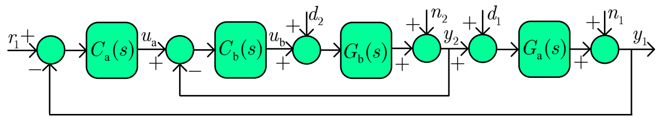

Figure 2. The series cascade control structure

The analytical approach described previously is used to design both PI/PID controllers in the internal and external control loops. The standard procedure for designing the primary controller $C_a(s)$ and the secondary controller $C_b(s)$ in the cascade control structure from Fig. 2, consists of two steps.

Step 1. First, the controller $C_b(s)$ in inner control loop is designed (if possible) for the process $G_b(s)$ [1]. The desired form of the complementary sensitivity function $T_{db}(s)$ is specified and the controller $C_{xb}(s)$ is designed (x=PI or x=PID at the designer's choice) under the desired robustness constraint M_{sb} or/and M_{pb} (or some other specifications for internal control loop).

$$M_{sb} = \max_{\omega} \left| \frac{1}{1 + C_b(s)G_b(s)} \right|, \quad M_{pb} = \max_{\omega} \left| \frac{C_b(s)G_b(s)}{1 + C_b(s)G_b(s)} \right|, \tag{12}$$

Step 2. In the second step, the controller $C_a(s)$ in the outer control loop is designed for the equivalent process described by the transfer function

$$G_{ae}(s) = \frac{G_a(s)G_b(s)C_b(s)}{1 + C_b(s)G_b(s)} \tag{13}$$

In this step, the designer assumes the form of the desired complementary sensitivity function $T_{da}(s)$ for the outer control loop. As a result, the controller $C_{xa}(s)$ (x=PI or x=PID) is designed for specified robustness sensitivities M_{sa} or/and M_{pa}, defined by relations:

$$M_{\mathrm{sa}} = \max_{\omega}\left|\frac{1}{1+C_{\mathrm{a}}(s)G_{\mathrm{ae}}(s)}\right|,\quad M_{\mathrm{pa}} = \max_{\omega}\left|\frac{C_{\mathrm{a}}(s)G_{\mathrm{ae}}(s)}{1+C_{\mathrm{a}}(s)G_{\mathrm{ae}}(s)}\right|. \tag{14}$$

Remark 1. The robustness constraints defined in (12) and (14) are used to design controllers in inner and outer loops, respectively. Due to the loop interaction in a cascade structure, the concept of structured singular values can be used for robustness analysis, as described in [39]. Therefore, the maximum of the sensitivity function of the overall system from Fig. 2 is defined as in [39]:

$$M_{\mathrm{s}} = \max_{\omega}\left|\frac{1}{1+C_{\mathrm{b}}(s)G_{\mathrm{b}}(s)+C_{\mathrm{a}}(s)G_{\mathrm{a}}(s)C_{\mathrm{b}}(s)G_{\mathrm{b}}(s)}\right| \tag{15}$$

Note that both the u_a and u_b control signals can be implemented independently such that the proportional action for the PI controller and the proportional and differential actions for the PID controller are shifted to the feedback path, while only the integral action of the k_i/s controller remains in the forward path of the reference signal, to which partially proportional action bk, $0\leq b\leq 1$ is added [49, 50]. In this way, this control structure prevents large spikes in the control signal due to changes in the reference signal. In addition, set-point weighting factor may be used to increase the response speed. It should also be noted that the described controller design procedure can be directly applied to multi-stage series cascade regulation of industrial processes without loss of generality.

III. COMPARATIVE ANALYSIS AND SIMULATIONS

Parameters of obtained C_{xa}/C_{xb} controllers (x=PI ili x=PID) are given in Table I for ten pairs of representatives of dynamical characteristics of industrial processes:

$$G_{a1}=\frac{2e^{-2s}}{5s+1},\quad G_{b1}=\frac{e^{-s}}{(2s+1)(3s+1)},\quad G_{a2}=\frac{e^{-s}}{(s+1)^2},$$
$$G_{b2}=\frac{e^{-0,1s}}{0,1s+1},\quad G_{a3}=\frac{e^{-0,25s}}{s},\quad G_{b3}=\frac{e^{-0,5s}}{(s+1)(2s+1)},$$
$$G_{a4}=\frac{2e^{-s}}{10s+1},\quad G_{b4}=\frac{e^{-0,5s}}{s},\quad G_{a5}=\frac{e^{-3s}}{10s-1},\quad G_{b5}=\frac{2e^{-s}}{s+1}$$
$$G_{a6}=\frac{2e^{-2s}}{20s+1},\quad G_{b6}=\frac{e^{-s}}{20s-1},\quad G_{a7}=\frac{e^{-0,7s}}{2s},\quad G_{b7}=\frac{2e^{-s}}{20s-1},$$
$$G_{a8}=\frac{e^{-4s}}{20s-1},\quad G_{b8}=\frac{2e^{-2s}}{s},\quad G_{a9}=\frac{e^{-4s}}{20s-1},\quad G_{b9}=\frac{2e^{-0,1s}}{10s-1}$$
$$G_{a10}=\frac{e^{-0,7s}}{s^2},\quad G_{b10}=\frac{e^{-0,3s}}{s+1}$$

In order to reduce the peak in response $y_1(t)$ to the step reference signal $r_1(t)$, the controller $C_a(s)$ is realized in all simulations so that only the integral action k_{ia}/s acts on the reference signal in the forward path, to which a partially proportional gain bk_a, $0\leq b\leq 1$, is added, while the other actions of the controller $C_a(s)$ are shifted to the feedback path. The controller $C_b(s)$ is placed in the forward path in all simulations. From the point of view of of practical realization it should be noted that it is necessary to prevent integral action from winding up, for which there are well-known (anti-windup) structures in practice, whether it is a PI or PID controller [50]. The parameters of the designed primary and secondary controller are listed in Table I, including the obtained quantitative performance and robustness measures [51].

The presented controller design method is explained in more details for the process $G_{p7}(s)$. The transfer function of the inner loop process model is $G_{b7}(s)=\frac{K_2e^{-L_2s}}{T_2s-1}$, where $K_2=2,\ L_2=1,\ T_2=20$. The process in the outer loop is described by the model $G_{a7}(s)=\frac{K_1e^{-L_1s}}{s}$, where $K_1=0.5$, $L_1=0.7$. First, the desired complementary sensitivity function for the inner loop is selected in the form $T_{db}(s)=\frac{(\eta_b s+1)e^{-L_2s}}{(\lambda_b s+1)^2}$. The unknown parameter η_b is then determined from the condition of cancellation of the process pole s=1/T_2 with a zero of the function $F_b(s)=(\lambda_b s+1)^2-(\eta_b s+1)e^{-L_2s}$ (see Eq. 3). As a result, η_b can be analytically calculated as $\eta_b=\left(T_2^2(1-e^{-L_2/T_2})+2\lambda_b T_2+\lambda_b^2\right)/\left(T_2e^{-L_2/T_2}\right)$. The filter time constant of the PID controller for the inner loop is assumed to be as follows $T_{fb}=\lambda_b/20$. The resulting high-order controller $C_b(s)=T_{db}(s)/((1-T_{db}(s)G_{7b}(s))$ depends only on an adjustable parameter for inner loop, i.e. λ_b. Based on equations (6) and (7), a PID controller in terms of an adjustable parameter λ_b is obtained. Finally, for the specified value M_{sb}=2, one obtains $\lambda_b\approx 1.53$ (parameters of controllers are given in Table I). Alternatively, λ_b may be changed until the inner loop reaches desired robustness levels expressed by M_{sb} and/or M_{pb}. In the second step, we analyze the equivalent process $G_{ae}=G_{a7}(s)G_{b7}(s)C_b(s)/(1+C_b(s)G_{b7}(s))$ described with Eq. (13), where $C_b(s)$ is previously obtained PID controller. The complementary sensitivity function for the outer control loop is chosen in the form $T_{da}(s)=\frac{(\eta_a s+1)e^{-(L_1+L_2)s}}{(\lambda_a s+1)^2(T_{fa}s+1)}$. For simplicity, let us adopt $T_{fa}\approx\eta_b$. Thus, the next step is to perform cancellation of the pole s=0 of the process $G_{a7}(s)$ with zero of the auxiliary function $F_a(s)=(\lambda_a s+1)^2(T_{fb}s+1)-(\eta_a s+1)e^{-(L_1+L_2)s}$. To achieve this, the parameter $\eta_a=2\lambda_a+T_{fa}+L_1+L_2$ is obtained from the condition $\left.\frac{d}{dt}\left(F_a(s)\right)\right|_{s=0}=0$. The resulting high-order controller $C_a(s)=T_{da}(s)/((1-T_{da}(s)G_{ae}(s))$ depends only on an adjustable outer loop parameter λ_a. By the Maclaurin's approximation mentioned above, a PID controller is obtained as a function of the parameter λ_a. The specification of M_{sa}=2 leads to $\lambda_a \approx 2.89$ and parameters of PID controller given in Table I. Here, the set-point weighting factor is chosen to be b=0.3, which leads to an additional shaping of the reference response. The analogous procedure is repeated for all processes considered.

The results of the simulation analysis of the proposed design methodology of controllers for series cascade processes including the comparison with two techniques recently elaborated in [43,44], are presented in Fig. 3 - Fig. 13. The comparison of responses to reference and load disturbances shown in Fig. 3 for process $G_1(s)$ in the cases where the designed secondary controllers are of PID-type, while the primary controllers have the structure PI or PID. As a result, IAE_{d2} is slightly smaller (slightly more efficient suppression of load disturbance d_2), while rejection of load disturbance d_1 is more efficient with PIDa compared to PIa controller, which is to be expected. Under the same constraints to modeling uncertainties M_{pa}=1, the PIDa and PIDb controllers are designed for process $G_2(s)$. Fig. 4 presents control signals $u_a(t)$ and $u_b(t)$ for process $G_1(s)$. Evidently, speeding up the reference response leads to more excessive control activity. Generally, preventing the integral action is done through known anti-windup structures used in practice, whether it is a PI or PID controller [50,52]. From Fig. 5 can be concluded that the system efficiently suppresses external disturbances compared to the controllers designed in [44]. Fig. 6 and Fig. 7 show the responses to reference and load disturbances for processes $G_3(s)$ and $G_4(s)$ for scenarios with different sensitivity constraints. It is evident that for smaller values of robustness constraints one obtains poorer load disturbance rejection. The use of PIDb instead of PIb controllers in the inner loop is justified for smaller values of IAE_{d1} and IAE_{d2}, as illustrated for process $G_5(s)$ in Fig. 8. Here, the controllers in the inner loop (PIDb and PIb) are designed under the constraint M_{sb}=2, and then the controllers PIDa in the outer loop are obtained for the specified M_{sa}=3.5. Fig. 9. Shows the system responses to the reference signal and disturbance for a fixed value M_{sa} = 2 and two values to constrain the robustness level of the inner loop (M_{sb}=2 and M_{sb}=3) for process $G_6(s)$. As expected, it results that for lower values of M_{sb}, a worse suppression of load disturbances is achieved, but the system is then less sensitive to modeling errors of the considered process. The same conclusion can be derived for process $G_7(s)$ for which the results are shown in Fig. 10. For process $G_8(s)$, larger values of IAE_{d1} and IAE_{d2} are obtained for the same constraint M_{sa}, but significantly lower values for M_{sb} and M_{pb} compared to the controller from [43], as can be seen in Fig. 11. The comparison of the performance/robustness of the system with the proposed PID controller in the inner and outer loops with [44] for processes $G_9(s)$ and $G_{10}(s)$ is shown in Fig. 12 and Fig. 13. Under the smaller constraint on robustness M_{sa}=2.46 (for proposed controller) compared to M_{sa}=2.54 (Ref. [44]), more efficient disturbance suppression is achieved for process $G_9(s)$ with the proposed PIDa, PIDb controllers (IAE_{d1}=6.32, IAE_{d2}=0.02) compared to the PID1, PID2 controllers from [44] (IAE_{d1}=9.57, IAE_{d2}=0.18). The same conclusion is drawn for the process $G_{10}(s)$ where the controllers are designed under the constraint M_{sa}=3.05. For all the considered processes, the reaction speed of the system is simply adjusted (increased) by a suitable choice of set-point weighting parameter b.

TABLE I. PARAMETERS OF C_{XA}/C_{XB}, (X=PI OR X=PID) FOR PROCESSES $G_{AJ}(S)/G_{BJ}$, J=1,2,3...,10 AND FOR SPECIFIED VALUES OF M_{SA} AND M_{SB} AS WELL AS PARAMETERS OF CONTROLLERS FROM RECENT STUDIES

Process pairs	k_a	k_{ia}	k_{da}	T_{fa}	b	M_{sa}	M_{pa}	k_b	k_{ib}	k_{db}	T_{fb}	M_{sb}	M_{pb}	IAE_{d1}	IAE_{d2}	M_s
$G_1(s)$	0.5524	0.1061	0.4884	0.1025	0.3	2.0	1.10	3.3928	0.8250	3.7773	0.0521	2.0	1.34	9.47	2.31	2.12
	0.3788	0.0763	-	-	0.8	1.8	1.00	3.3928	0.8250	3.7773	0.0521	2.0	1.34	13.1	2.69	1.76
$G_2(s)$	0.9598	0.4681	0.4712	0.0300	0.35	1.68	1.00	1.1397	8.5496	0.0346	0.0027	2.0	1.09	2.14	0.14	1.99
Ref. [44]	0.2449	0.1870	0.0170	0	1.00	1.34	1.00	0.5920	5.1380	0.0090	0	1.48	1.00	5.35	0.31	1.48
$G_3(s)$	0.5594	0.0742	0.6919	1.7550	0.41	2.0	1.43	4.1129	1.8308	2.4420	0.0278	2.0	1.38	13.49	2.06	2.40
	0.4547	0.0517	0.5835	1.7966	0.45	1.8	1.41	3.5430	1.5174	2.0988	0.0306	1.8	1.24	19.36	2.95	1.96
$G_4(s)$	1.4678	0.3127	-	-	0	2.0	1.21	1.5962	0.7599	0.2921	0.0350	2.0	1.46	3.26	1.15	1.99
	1.3645	0.2117	-	-	0.4	1.8	1.07	1.5962	0.7599	0.2921	0.0350	2.0	1.46	4.72	1.19	1.98
	1.2792	0.1903	-	-	0.4	1.8	1.05	1.3771	0.5617	0.2443	0.0440	1.8	1.39	5.26	1.67	1.76
$G_5(s)$	2.4432	0.0968	3.5501	0.1560	0.185	3.5	3.08	0.5698	0.4275	0.1731	0.0273	2.0	1.09	10.33	3.80	3.18
	2.3298	0.0809	4.0372	0.1765	0.20	3.5	3.07	0.4111	0.3183	-	-	2.0	1.17	12.37	5.46	2.62
$G_6(s)$	1.3709	0.1479	-	-	0	2.0	1.19	15.614	3.2769	5.8002	0.0765	2.0	1.52	6.93	0.27	1.97
	1.6680	0.1962	-	-	0	2.0	1.24	22.079	6.7223	9.007	0.0420	3.0	2.12	5.25	0.12	3.26
$G_7(s)$	0.5517	0.0383	1.2332	4.3637	0.30	2.0	1.49	7.8069	1.6384	2.9001	0.0765	2.0	1.52	26.16	2.36	1.97
	0.5814	0.0439	1.1225	3.7947	0.3	2.0	1.49	8.7008	2.0666	3.3109	0.0638	2.2	1.59	22.79	1.78	2.06
$G_8(s)$	2.6181	0.0551	7.9475	4.1722	0.22	3.0	2.64	0.3428	0.0665	0.2912	0.0543	5.0	4.10	18.15	21.1	7.02
Ref. [43]	2.4300	0.0730	-	-	0	3.0	2.50	0.2800	0.0700	-	-	37.6	38.1	15.42	18.78	6.47
$G_9(s)$	3.6822	0.1581	2.8892	0.1735	0.2	2.46	2.25	21.9546	35.2929	0.0588	0.0168	1.81	1.53	6.32	0.02	2.02
Ref. [44]	3.0000	0.1050	0.1300	0	0.2	2.54	2.38	10.960	5.5000	0.5300	0	1.17	1.22	9.57	0.18	1.17
$G_{10}(s)$	0.2086	0.0305	0.6076	0.0585	0.28	3.05	2.78	2.9959	4.0580	0.3078	0.0141	2.0	1.24	32.8	2.60	2.19
Ref. [44]	0.2000	0.0410	0.7500	0	0	3.05	2.40	3.1100	3.7300	0.3620	0	2.0	1.10	34.7	2.77	2.27

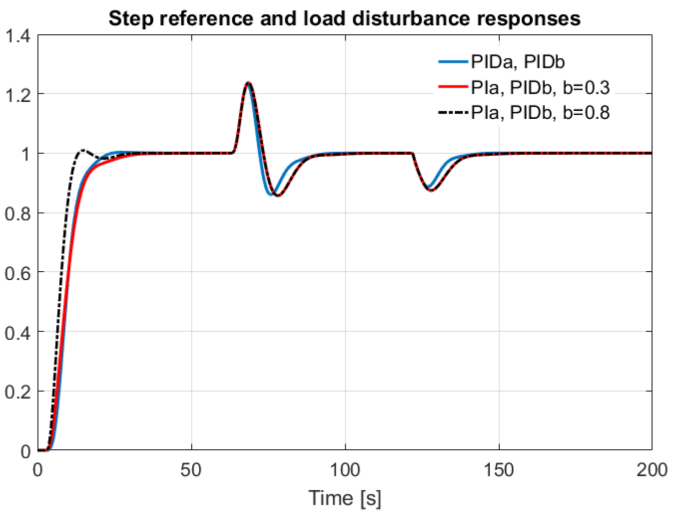

Figure 3. Responses of the system to step reference $R_1(s)=1/s$, and step load disturbances $D_1(s)=-0.1\exp(-120s)/s$ and $D_2(s)=\exp(-60s)/s$ for pair of processes $G_1(s)$ with proposed controllers PIDa,PIDb and PIa,PIDb

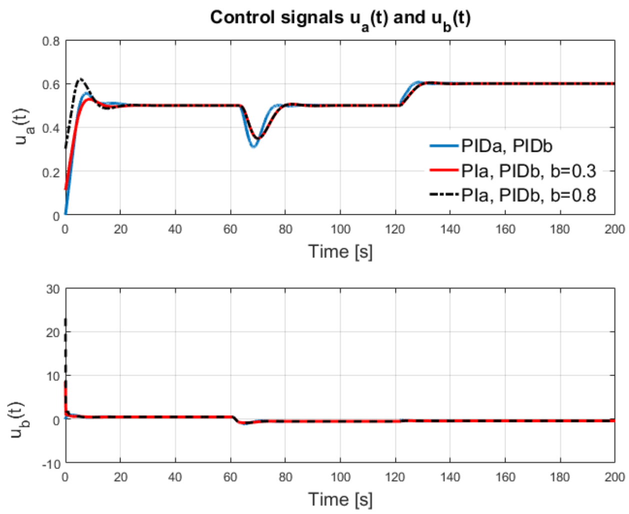

Figure 4. Corresponding control signal $u_a(t)$ and $u_b(t)$ of the system for pair of processes $G_1(s)$ with proposed controllers PIDa,PIDb and PIa,PIDb

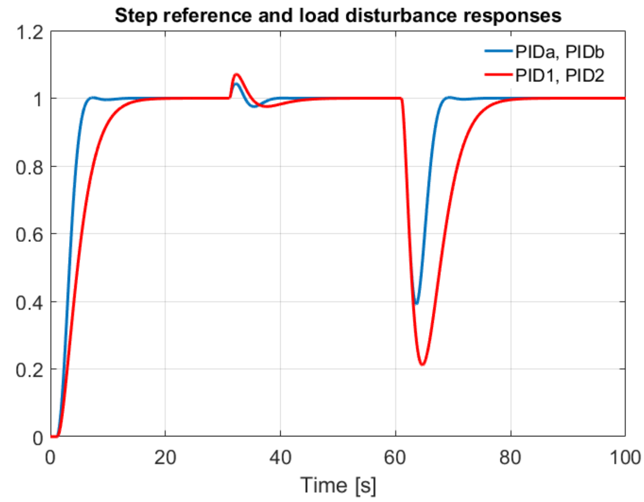

Figure 5. Responses of the system to step reference $R_1(s)=1/s$, and step load disturbances $D_1(s)=-\exp(-60s)/s$ and $D_2(s)=\exp(-60s)/s$ for pair of processes $G_2(s)$ with proposed controllers PIDa,PIDb (blue line) and controllers PID1,PID2 (red line) from reference [44] under same constraints $M_{pa}=M_{p1}$.

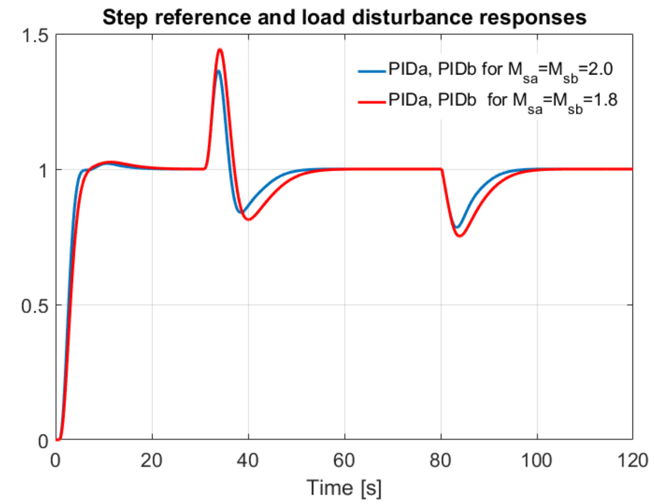

Figure 6. Responses of the system to step reference $R_1(s)=1/s$, and step load disturbances $D_1(s)=-0.1\exp(-80s)/s$ and $D_2(s)=\exp(-30s)/s$ for pair of processes $G_3(s)$ with proposed controllers PIDa,PIDb for different robustness constraints.

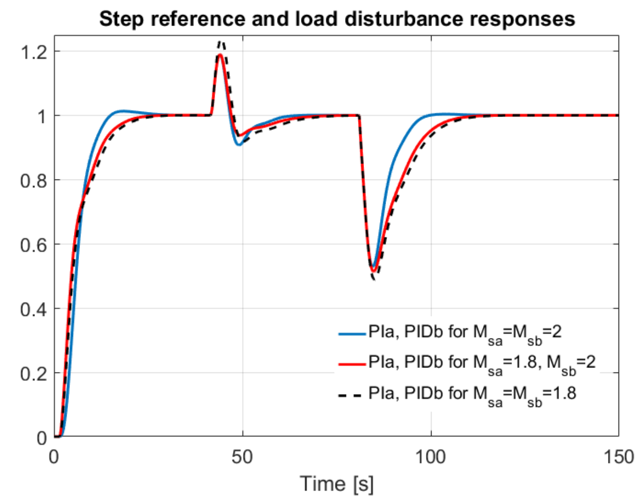

Figure 7. Responses of the system to step reference $R_1(s)=1/s$, and step load disturbances $D_1(s)=-1\exp(-80s)/s$ and $D_2(s)=\exp(-40s)/s$ for pair of processes $G_4(s)$ with proposed controllers PIa,PIDb under different robustness constraints.

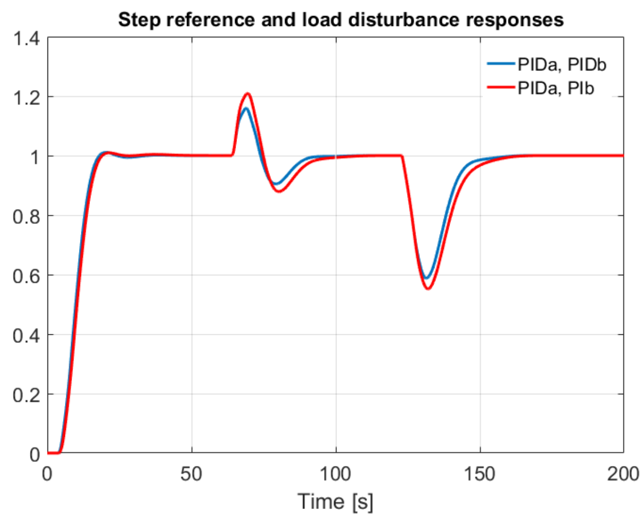

Figure 8. Responses of the system to step reference $R_1(s)=1/s$, and step load disturbances $D_1(s)=-0.5\exp(-120s)/s$ and $D_2(s)=0.5\exp(-60s)/s$ for pair of processes $G_5(s)$ with proposed controllers PIDa,PIDb and PIDa, PIb.

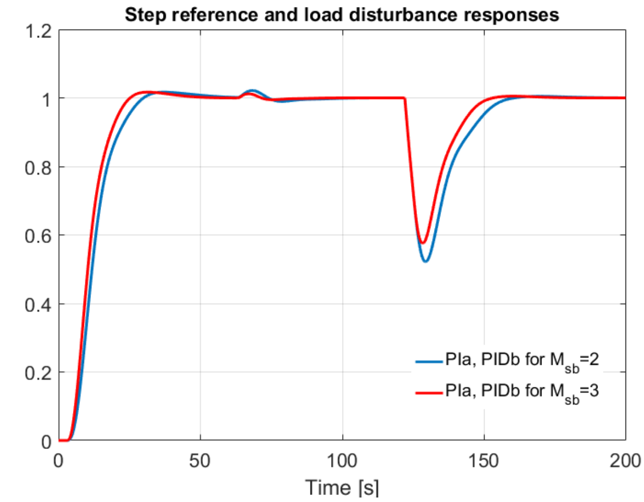

Figure 9. Responses of the system to step reference $R_1(s)=1/s$, and step load disturbances $D_1(s)=-\exp(-120s)/s$ and $D_2(s)=\exp(-60s)/s$ for pair of processes $G_6(s)$ with proposed controllers PIa,PIDb for $M_{sb}=2$ and PIa, PIDb for $M_{sb}=3$.

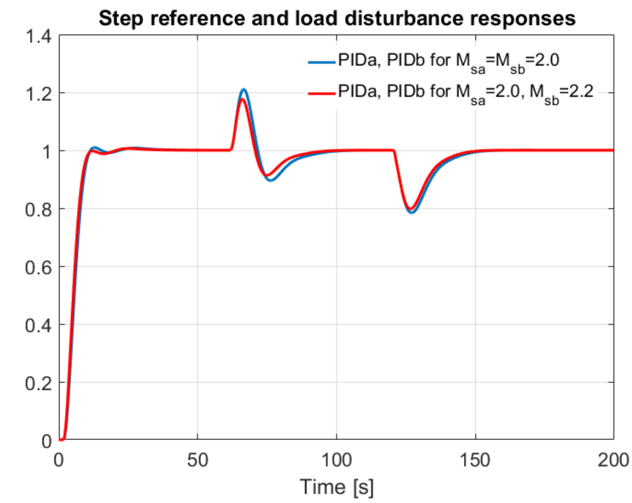

Figure 10. Responses of the system to step reference $R_1(s)=1/s$, and step load disturbances $D_1(s)=-0.1\exp(-120s)/s$ and $D_2(s)=\exp(-60s)/s$ for pair of processes $G_7(s)$ with proposed controllers PIDa,PIDb designed for different robustness constraints.

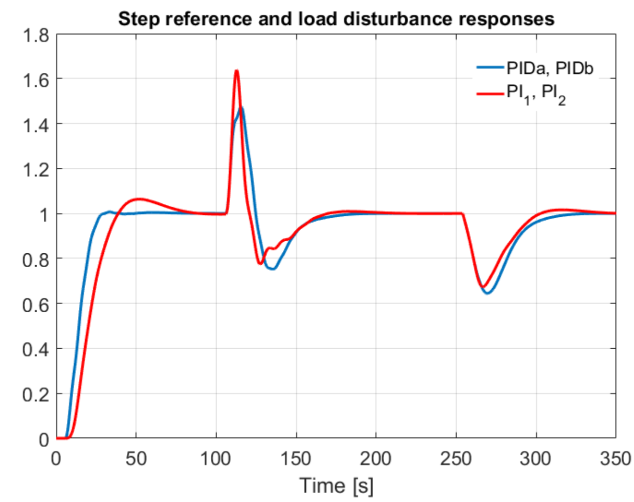

Figure 11. Responses of the system to step reference $R_1(s)=1/s$, and step load disturbances $D_1(s)=-0.5\exp(-250s)/s$ and $D_2(s)=0.5\exp(-100s)/s$ for pair of processes $G_8(s)$ with proposed controllers PIDa,PIDb and controllers PI_1,PI_2 from [43] designed under same robustness constraint $M_{sa}=M_{s1}$.

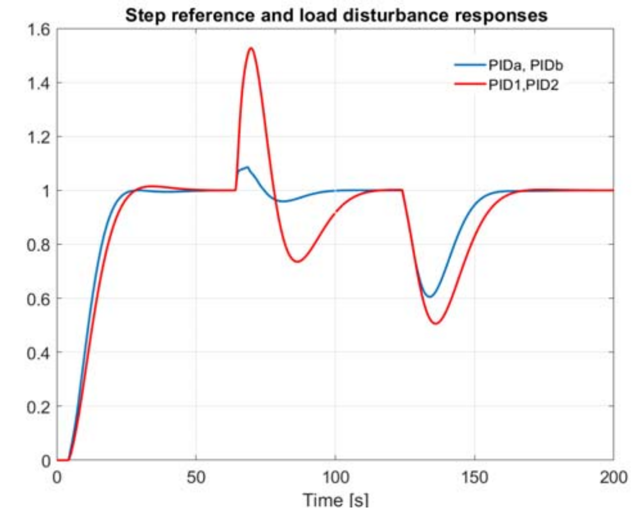

Figure 12. Responses of the system to step reference $R_1(s)=1/s$, and step load disturbances $D_1(s)=-\exp(-120s)/s$ and $D_2(s)=50\exp(-60s)/s$ for pair of processes $G_9(s)$ with proposed controllers PIDa,PIDb and controllers PID1,PID2 from [44] designed under same robustness constraint $M_{sa}=M_{s1}$.

Figure 13. Responses of the system to step reference $R_1(s)=1/s$, and step load disturbances $D_1(s)=-0.1\exp(-120s)/s$ and $D_2(s)=\exp(-60s)/s$ for pair of processes $G_{10}(s)$ with proposed controllers PIDa,PIDb and controllers PID1,PID2 from [44] designed under same robustness constraint $M_{sa}=M_{s1}$.

IV. CONCLUSIONS

The paper presents an analytical method for determining the parameters of PI and PID controllers for cascaded industrial processes with time delay. The controllers in the inner and outer loops are designed considering robustness constraints by specifying the maximum sensitivity or/and the complementary sensitivity function. Simulation results show that the proposed technique is robust and provides satisfactory improvements in system performance indicators compared to recently published methods for tuning the parameters of conventional controllers for industrial cascade processes with time delay. Alternatively, the operator can also modify the desired trade-off between robustness and performance measures by selecting the adjustable parameters. This shows that the proposed method is characterized by flexibility and provides an efficient way to fine-tune standard cascade control systems, and that the same methodology can be applied, to multi-stage cascade control of industrial processes, without loss the generality.

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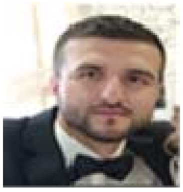

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