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Design and development of modified RTD-A controller for unstable bioreactor

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Abstract: Control of open loop unstable process is a challenging one owing to the presence of right hand plane pole(s). The spotlight of this paper is to develop modified Robustness set point Tracking Disturbance rejection and Aggressiveness controller(RTD-A) for an unstable process. The RTD-A controller formulation is modified accordingly to suit the unstable process. The proposed control scheme has two control loops. The unstable process is initially stabilized by an inner feedback loop consisting of a proportional controller. The RTD-A control strategy is then implemented on the stabilized plant. The developed control scheme is validated on an nonlinear unstable bioreactor plant model in a simulation environment. It is found that the proposed controller has better servo and regulatory performance when compared to the conventional IMC-PID controller. Effect of variation in the tuning parameters of the controllers on the process output is analysed in detail.

Keywords: RTD-A controller, IMC-PID, unstable process, bioreactor.

1. INTRODUCTION

A class of controllers employing the model of the system for taking control action are termed as Model Predictive Controllers (MPC). Robust Tracking Disturbance - Overall Aggressiveness (RTDA) Controller is one such emerging controllers which is an alternative control strategy. RTD-A controller possess the predicting feature of conventional MPC which still retains the simplicity of the Proportional Integral Derivative (PID) controllers. Individual tuning parameters for robustness, set point tracking, disturbance rejection and overall aggressiveness of the controllers are directly related to the performance attributes of the controller [Ogunnaike and Mukati (2006) and Mukati et al. (2009)].

Ogunnaike and Mukati (2006) have developed the computational law for RTD-A controller for a First Order Plus Dead Time Process (FOPDT). The developed control strategy has been implemented on a simulation environment of a nonlinear polymerization reactor model. It is proven that the proposed controller outperforms the conventional IMC and PID controllers. As a continuation of the same, rules for the selection of the four tuning parameters based on stability criteria have been developed by Mukati et al. (2009). The tuning rules have been validated on temperature control in a physical vapor decomposition process. However the tuning rules developed mandatorily requires the selection of an uncertainty parameter which becomes slightly difficult in practice. Hence, block diagram representation and semi analytical tuning rules for RTD-A controllers was developed by Sendjaja et al. (2011).

As the design developed by Ogunnaike cannot handle second order process with dead time process (SOPDT) or SOPDT with minimum or non-minimum zero, Anbarasan and Srinivasan (2015), proposed a simplified RTD-A control algorithm suitable for SOPDT process with minimum or non-minimum zero. Closed loop block diagram and stability analysis of the proposed control algorithm has been effectively analyzed. Haseena and Srinivasan (2018) have developed a mixed constrained RTD-A controller capable of handling various linear inequality constraints in multi variable control framework.

Open loop unstable processes are predominant in some petroleum and chemical industries. Some of the industrial process may also have multiple steady states. Unstable systems are those which have a minimum of one pole in the right hand side of the s-plane. Control of such systems is a little tedious one. Jacob and Chidamabram (1996) have developed PID controller design formula for FOPDT unstable processes. From the simulation results it is evident that the proposed methods can handle perturbations in time delay, time constant and process gain.

MPC controller for open loop unstable processes have been proposed by Qi and Fisher (1993) using the state space formulation with an AR model in addition to the usual step response model . Nagrath et al. (2002), have formulated a state estimation based model predictive controllers for open loop unstable cascade systems. Kalman filter is incorporated for the estimation of states and the new augmented states of the original system is formed by extending the modeled disturbances as augmented states. Lee and Park (1991) have dealt with the design of MPC

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for multi variable unstable processes with constraints on manipulated variables. A quadratic dynamic matrix control scheme developed with state feedback for stabilization of the unstable process is found to handle constraints on the manipulated variable.

From the survey of literature of recent researches, it is evident that the RTD-A controller outperforms the other classical model predictive control schemes such as internal model control (IMC), dynamic matrix control (DMC) in terms of both set point tracking as well as disturbance rejection. Also researchers have so far developed RTD-A controllers for stable processes with or without constraints. This creates a motivation to formulate RTD-A control law for unstable processes. This paper focuses on development of RTD-A controller for Second order processes with one pole in the right hand plane creating instability.

The same existing RTDA-A control scheme cannot be directly implemented to the unstable problem as such. This paper proposes a RTD-A control strategy with modifications suitable to be used for unstable processes.

The paper is organized as follows:

Section 1 briefs about the history and recent researches carried out in RTD-A control algorithm. Section 2 proposes the modified design of RTD-A controller for Second order process with RHP pole. An illustrative numerical example is presented in Section 3. Design and Development of RTD-A controller for unstable bioreactor process is dealt in Section 4. Section 5 concludes the paper.

2. DESIGN OF MODIFIED RTD-A CONTROLLER FOR SECOND ORDER RHP SYSTEMS

The proposed control strategy consists of two control loops. The unstable process is first stabilized using a suitable proportional(P) only controller. Cascade control scheme of control law formulation is adopted where the inner loop consists of only the stabilizing controller. The RTD-A control scheme is implemented in the outer loop with the stabilized model of the unstable process. A simple proportional only controller suits well for stabilization of the open loop unstable process with one pole in the RHS. However, any other form of controller could be used in the inner loop for stabilization of more complicated unstable systems. The proposed control scheme is represented as shown in the block diagram.

An SOPDT system with one pole in the right hand side is considered as given below.

$$G(s) = \frac{Ke^{-\theta s}}{(s-a)(s+b)} \tag{1}$$

where K is the process gain, θ is the dead time, a and b are the unstable and stable poles respectively.

The RHP pole is first stabilized using a suitable Proportional (P) controller with gain K_c.

The proposed control strategy is of the cascade controller. The inner feedback loop is for the stabilization of the unstable process and the RTD-A controller in outer feed forward loop takes care of the performance attributes.

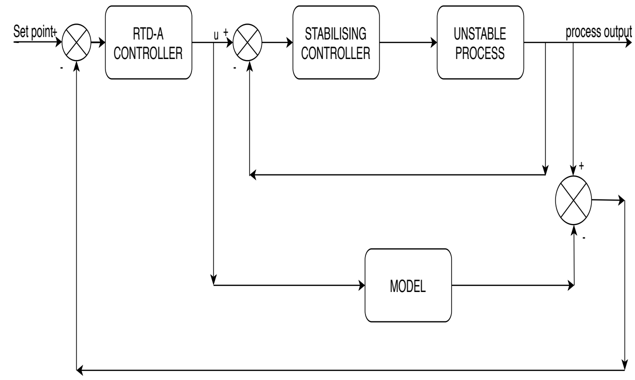

Fig. 1. Block diagram representation of modified RTD-A controller for unstable processes

The closed loop transfer function of the inner loop when the unstable system is stabilized using a proportional controller is derived in this section. The unstable transfer function is given as in equation (1).

$$G'(s) = \frac{K_cKe^{-\theta s}}{(s-a)(s+b)+K_cKe^{-\theta s}} \tag{2}$$

On proper approximation of the dead time as $(1-\theta s)$ using appropriate approximation equation (2) is rewritten as

$$G'(s) = \frac{K_cKe^{-\theta s}}{(s-a)(s+b)+K_cK(1-\theta s)} \tag{3}$$

On simplification

$$G'(s) = \frac{K_cKe^{-\theta s}}{s^2+(b-a-K_cK\theta)s+(K_cK-ab)} \tag{4}$$

Taking τ' as the time constant and ζ' as the damping coefficient of the stabilized system, equation (4) is rewritten as,

$$G'(s) = \frac{K'e^{-\theta s}}{\tau'^2s^2+2\zeta'\tau's+1} \tag{5}$$

One step ahead prediction in the discrete domain is given as,

$$\hat{y}(k+1) = c_1\hat{y}(k)+c_2\hat{y}(k-1)+d_1u(k-M)+d_2u(k-M-1) \tag{6}$$

Where, M stands for the delay period given by $round(\frac{\theta}{T_s})$ with T_s sampling time. c_1, c_2, d_1, d_2 are the discrete model parameters.

General 'L' step ahead future prediction of the discrete model output with M steps delay is written as,

$$\hat{y}(k+M+L) = \alpha\hat{y}(k)+\beta\hat{y}(k-1)+\gamma u(k-M-1)+ \sum_{j=0}^{M-1}\lambda(j)u(k-M+j)+\sum_{j=M}^{X-1}\lambda(j)u(k) \tag{7}$$

Where α, β, γ, λ are the respective coefficients and $X = M+1$.

The predicted model output is updated using the available process measurements. This forms the differentiating factor of the RTD-A control scheme from the rest of the model based control strategies.

The model update equation is given as,

$$\tilde{y}(k+M+L) = \alpha\hat{y}(k) + \beta\hat{y}(k-1) + \gamma u(k-M-1) + \sum_{j=0}^{M-1}\lambda(j)u(k-M+j) + \sum_{j=M}^{X-1}\lambda(j)u(k) + \hat{e}_d(k+M+L|k) \quad (8)$$

Where $\hat{e_d}(k+M+L|k)$ denotes the update of the future error prediction. It is dependent on the disturbance rejection parameter θ_d. $\hat{e}_d(k+M+L|k)$ is formulated as

$$\hat{e_d}(k+M+L|k) = \hat{e}_d(k) + \frac{(1-\theta_d)}{\theta_d}[1-(1-\theta_d)^{M+L}]\delta\hat{e}_d(k) \quad (9)$$

where, θ_d, the disturbance rejection parameter,can vary from 0 to 1.

$\delta\hat{e}_d(k)$ is the difference in the present and one-step previous value of the error. $\hat{e}_d(k)$ is the prediction of the current disturbance effect.

$$\hat{e}_d(k) = \theta_r\hat{e}_d(k-1) + (1-\theta_r)e(k) \quad (10)$$

where, e(k) is the modeling error. $\hat{e}_d(k)$, is dependent on the robustness parameter θ_r which lies between 0 and 1. When θ_r is 1, the closed loop system is highly stable.

The control law is formulated to bring the process output close to the set point and thus making error as zero. An error minimization based optimization problem is solved to compute the control law. The objective function of the optimization problem is posed as,

$$\min_{u(k)} \sum_{L=1}^{N}(y_{set}(k+L) - \tilde{y}(k+M+L))^2 \quad (11)$$

$y_{set}(k)$ denotes the set point trajectory for a defined set point y_{def} and $y_{set}(k+L)$ denotes the reference trajectory for L step ahead predictions. The set point trajectory, $y_{set}(k)$ is defined as,

$$y_{set}(k) = \theta_t y_{set}(k) + (1-\theta_t)y_{def} \quad (12)$$

θ_t indicates the set-point trajectory tracking parameter of the RTD-A controller. It lies between 0 and 1.

The final control law of the RTD-A controller is computed as,

$$u(k) = \frac{\sum_{L=1}^{N}\Phi_L(k)(\sum_{j=M}^{X-1}\lambda(j))}{\sum_{L=1}^{N}(\sum_{j=M}^{X-1}\lambda(j)^2} \quad (13)$$

Where $\Phi_L(k)$ is given as,

$$\Phi_L(k) = \sum_{L=1}^{M}(y_{set}(k+L) - \alpha\hat{y}(k) - \beta\hat{y}(k-1) - \gamma u(k-M-1) - \sum_{j=0}^{M-1}(\lambda(j)u(k-M+j) - \hat{e_d}(k+M+L|k)) \quad (14)$$

Prediction horizon, Ł defines the overall aggressiveness of the developed controller. Prediction horizon is related to the aggressiveness tuning parameter, θ_a by the following relation.

$$\theta_a = 1 - e^{((L-1)T_s)/\tau'} \quad (15)$$

The four tuning parameters of the RTD-A control scheme for robustness, set point tracking, disturbance rejection and aggressiveness are $\theta_r, \theta_t, \theta_d, \theta_a$ respectively.

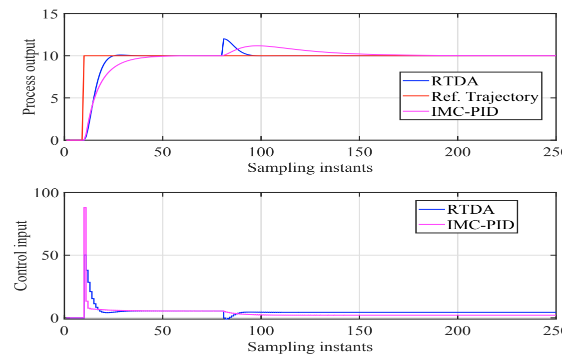

Fig. 2. Process output and control input of numerical example

3. ILLUSTRATIVE EXAMPLE

The above proposed RTD-A control scheme is demonstrated through an unstable transfer function model in this section. A transfer function with process gain 1, with one stable pole at -3.1086 and one unstable pole at 1.1086 is considered for simulation purpose.

The transfer function is as follows

$$G(s) = \frac{1}{s^2 + 2s - 3.446} \quad (16)$$

The stabilizing gain K_c for this transfer function is found to be 4. When the unstable transfer function model is stabilized with this stabilizing gain, the transfer function of the stabilized model is found to have stable poles at $(-1+1.56i)$ and $(-1-1.56i)$. This yields the stable transfer function model,

$$G'(s) = \frac{4}{s^2 + 2s + 0.554} \quad (17)$$

RTD-A controller is designed for the above transfer function model as given in Eq.(15). The tuning parameters $\theta_r, \theta_t, \theta_d, \theta_a$ are tuned as per the tuning rules given by Anbarasan and Srinivasan (2015). The tuning parameters for the system are set as $0.5, 0.9, 0.1$ and 0.4 with prediction horizon N as 8 respectively. The controller performance is compared with that of an IMC-based PID controller. The tuning parameter λ for IMC-PID is chosen as 0.5. The servo and regulatory response of the plant output for a set point of 10 is shown in Fig. 2 along with the control input. An output disturbance is introduced to the plant output and it is seen that the designed RTD-A controller is capable of rejecting the disturbance and track the set point in a better manner than the IMC-PID.

4. DESIGN OF MODIFIED RTD-A CONTROLLER FOR NON-LINEAR UNSTABLE BIOREACTOR PLANT

Biochemical reactor plants are widely used in various fields of applications such as treatment of waste water, fermentation and pasteurization processes. Substrate element is consumed by the biological mass to produce proliferated cells. D and Lim H C (1981) have given the governing mathematical modeling equations of a continuous stirred tank bioreactor for growth of the methanol utilizing microorganism Methylomonas under conditions where methanol is the rate limiting substrate.

The non-linear modeling equations of the continuous stirred tank bioreactor as stated by Srinivas and Chidambaram (1995), is given as,

$$\dot{x_1} = (\mu(x_2) - D)x_1 \tag{18}$$

$$\dot{x_2} = -\sigma(x_2)x_1 + D(x_{2f} - x_2) \tag{19}$$

where $\mu(x_2)$ and $\sigma(x_2)$, the specific growth rate and the specific consumption rate are given by the relation,

$$\mu(x_2) = \frac{0.504x_2(1 - 0.24x_2)}{0.00089 + x_2 + 0.406x_2^2} \tag{20}$$

$$\sigma(x_2) = \frac{x_2(1.32 + 3.86x_2 - 0.661x_2^2)}{0.00089 + x_2 + 0.406x_2^2} \tag{21}$$

Biomass concentration and reactor substrate concentration are two dimensionless quantities corresponding to the two states. D represents the dilution rate in h^{-1}. μ and σ represents the specific growth rate and the specific consumption rate respectively. x_{2f} represents the feed substrate concentration. The bioreactor plant is controlled for desired reactor substrate concentration x_2 with respect to the input dilution rate D.

Operating the above non-linear biochemical reactor model at a dilution rate of 0.4 and feed substrate concentration of 1.8 it results in three steady state operating points out of which one is unstable. $[0.24; 0.40]$ forms the unstable operating point.

The transfer function relating the substrate concentration (0.4) and the input dilution rate (0.4) having one RHP pole at 0.123 is given by Srinivas and Chidambaram (1995),

$$G_p(s) = \frac{1.46s + 0.5725}{s^2 + 0.8934s - 0.12571} \tag{22}$$

The above unstable RHP system is stabilized with a gain 0.65. The stabilized transfer function is given by,

$$G'_p(s) = \frac{0.91s + 0.3718}{s^2 + 1.80s + 0.24639} \tag{23}$$

The discrete model parameters c_1, c_2, d_1, d_2 for the RTD-A controller are $1.689, -0.6977, 0.1592$ and -0.1467 respectively. The controller is tuned effectively to achieve an aggressive closed loop characteristics to obtain faster response. The tuning parameters $\theta_r, \theta_t, \theta_d, \theta_a$ are chosen appropriately as $0.5, 0.01, 0.5$ and 0.3 respectively as per the tuning rules developed by Anbarasan and Srinivasan (2015). The proposed modified RTD-A control scheme is implemented on the non-linear model of the Bioreactor developed by Srinivas and Chidambaram (1995). The controller has been designed with the linearized model and implemented in the non-linear model. Comparison plot of servo and regulatory performance of RTD-A controller and IMC-PID controller on the unstable bioreactor plant is presented in Fig. 3. The controller is able to track the desired set point of 0.8 substrate concentration. An input disturbance of variation of the feed substrate concentration, x_{2f} from 1.28 to 3 is introduced and the proposed controller is capable of rejecting the input disturbance and track the desired reference trajectory. As the controller is developed using a linearised model and implemented on a non-linear model, the controller is automatically subjected to probability of plant-model mismatch. It is evident from the servo and regulatory response of the controller that the developed controller is robust enough to handle these plant model mismatch.

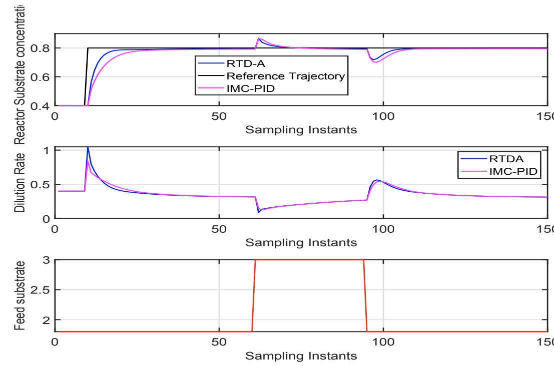

Fig. 3. Servo regulatory response of proposed modified RTD-A and conventional IMC-PID controllers on unstable bioreactor

5. ANALYSIS ON VARIATION IN TUNING PARAMETERS

Closer the value of the set point tracking parameter θ_t to 1, the response becomes more and more sluggish and when $\theta_t = 1$ the set point changes are fully ignored. When θ_t is close to 0 an instantaneous set point tracking performance is achieved. This variation in θ_t tuning parameter variation is shown in Fig. 4. Servo and regulatory responses when θ_t is $0.9, 0.01, 0.8$ while maintaining the other tuning parameters at a constant value is presented. It is inferred from the response that the set point tracking capability decreases as θ_t approaches 1 and the controller becomes more and more conservative in nature. The transient response of θ_t values $0.9, 0.8$ is found to be sluggish which makes the overall settling time delayed. It is also explicitly seen that variation of θ_t has its effect only on the servo problem while it has null effects in the disturbance rejection.

The process output response is shown in Fig. 5 for various values of θ_d ranging from 0 to 1 with other tuning parameters kept constant. When θ_d equals 1, from Eq.(9) it could be inferred that, at the estimated error at any current instant the prediction of disturbance effect in the future will be a constant non-changing value. Similarly, substitution of θ_d as 0 in Eq.(9) indicates that in the future

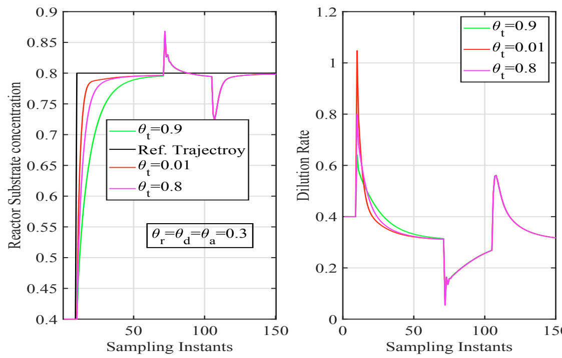

Fig. 4. Process output with respect to variation in set point tracking parameter

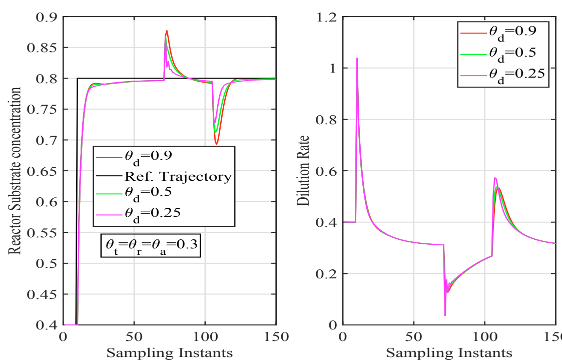

Fig. 5. Process output with respect to variation in disturbance rejection parameter

instants, the disturbance evolving as an increasing ramp. Thus, θ_d must be tuned to lie close to 0 to achieve an aggressive disturbance rejection capability and vice versa.

Fig. 6 showcases the plant response for fixed values of θ_t, θ_d and θ_a with various values of θ_r, the robustness tuning parameter. Plant model mismatch is handled by θ_r. From the plot in Fig. 6, it is evident that the controller has aggressive robustness capability when θ_r lies close to 0. θ_r is chosen to lie close to 1 when heavy effective filtering is required in case where there is significant model mismatch and the integrity of the model is not assured. It should be noted from Eq.(10) that selection of θ_r exactly equal to 0, there is nil model mismatch which is physically not realizable. Similarly, applying θ_r equal to 1, indicates that the model is developed with ignorance of process knowledge. This implies that selection of θ_r equal to 0 results in extreme problems in robust stability while θ_r equal to 0 results in steady state offset issues.

θ_a, the aggressiveness tuning parameter lies between 0 and 1. The prediction horizon is directly related to the aggressiveness tuning parameter, θ_a. When θ_a, is small and lies close to 0, it results in smaller prediction horizon assuring a better estimating capability. As θ_a, approaches 1, N also increases proportionally and when θ_a is made equal to 1, the prediction horizon becomes infinite resulting in conservative aggressiveness of the controller. This aggressiveness of the controller is evident from Fig. 7,

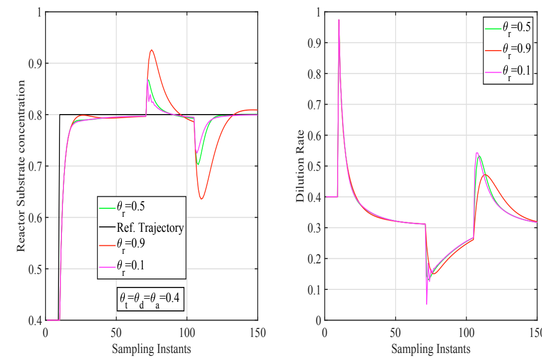

Fig. 6. Process output with respect to variation in robustness parameter

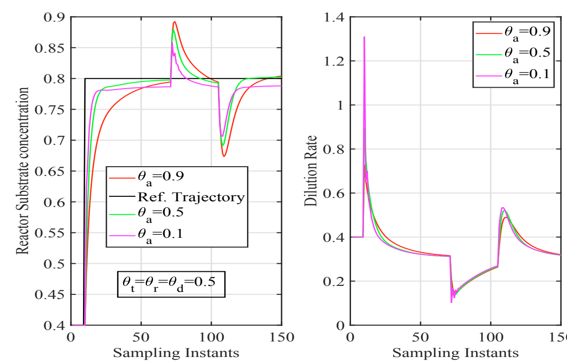

Fig. 7. Process output with respect to variation in aggressiveness parameter

where θ_a is varied while having constant values for the other tuning parameters.

6. CONCLUSION

Modified RTD-A controller suitable for unstable process has been designed and developed. Suitable proportional controller gain is chosen for the inner loop stabilization of the unstable process. The efficiency of the proposed modified RTD-A control scheme is exhibited via an illustrative numerical example and a non-linear unstable continuous stirred tank bioreactor plant. Comparison of the results with the conventional IMC-PID controller showcases the supremacy of the proposed control strategy in terms of servo and regulatory performances.

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REFERENCES

Anbarasan, K. and Srinivasan, K. (2015). Design of rtda controller for industrial process using sopdt model with minimum or non-minimum zero. *ISA Trans.*, 57, 1231–244.

D, D. and Lim H C, W.W.A. (1981). An experimental verification of stability and multiplicity of steady-states in a biological reactor. *AIChE J.*, 27, 284–292.

Haseena, B.A. and Srinivasan, K. (2018). Development of mixed constrained rtda controller for industrial applications. *ISA Trans.*, 81, 197–209.

Jacob, E.F. and Chidamabram, M. (1996). Design of controllers for unstable first-order plus dead time delay systems. *Computers chem. Engg.*, 20, 579–584.

Lee, J.K. and Park, S.W. (1991). Modelpredictive control for multivariable processes wkth constraints on manipulated ariables. *Korean Journalof Chem. Eng*, 8, 195–201.

Mukati, K., Rasch, M., and Ogunnaike, B.A. (2009). An alternative structure for next generation regulatory controllers. part ii: Stability analysis, tuning rules and experimental validation. *J. Process Contr.*, 19, 272–287.

Nagrath, D., Prasad, V., and Bequette, B.W. (2002). A model predictive formulation for control of open loop unstable cascade systems. *Chemical Engineering Science*, 57, 365–378.

Ogunnaike, B.A. and Mukati, K. (2006). An alternative structure for next generation regulatory controllers: Part i: Basic theory for design, development and implementation. *J. Process Contr.*, 16, 499–509.

Qi, K.Z. and Fisher, D.G. (1993). Model predictive control for open loop unstable process. *1993 American Control Conference*.

Sendjaja, A.Y., Ng, Z.F., How, S.S., and Kariwala, V. (2011). Analysis and tuning of rtd-a controllers. *Ind. Eng. Chem. Res.*, 50, 3415–3425.

Srinivas, M. and Chidambaram, M. (1995). Fuzzy logic control of an unstable bioreactor. *Bioprocess Engineering*, 12, 135–139.