Optimization of Performance Attributes Using RTDA Controller for Dual CSTR

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Abstract

The abstract of this work is to design an alternative control scheme – RTD-A, that combines the simplicity of PID controller with technical – brilliance of MPC controller by avoiding the tuning problems associated with both, for a highly used industrial process, Dual CSTR. Continuous stirred-tank reactor (CSTR), is a standard process used in chemical industries/engineering and environmental engineering. Cascading two CSTRs will lead to decrease in cost and volume when compared to single CSTR. In the proposed work, the temperature control of coupled CSTR is attempted by implementing PID, adaptive control, MPC, and the new generation RTD-A controllers. The performance of the proposed control schemes is compared and it is proved that the RTDA controller outperforms the other control schemes in terms of settling time and ISE.

Keywords: DUAL CSTR; PID; MRAC; MPC; RTD-A

1. Introduction

A CSTR is a highly non-linear system and this non-linearity makes the chemical process unstable. When two CSTRs are in series configured as 'DUAL CSTR'', the first operates at a higher concentration, and with higher reaction rate. The second reactor in series builds on the conversion carried out in the first reactor. Assuming that the reactants are perfectly mixed, the calculations are performed with the two CSTRs. In a perfect CSTR, the composition of the feed will be identical to the yield and is a function of the reaction rate and the total residential time. By using dynamic and steady state characteristics, suitable control schemes are designed for the DUAL CSTR process. Due to the highly exothermic reaction, the temperature has to be maintained within the threshold value by varying flow rate of coolant supplied to the jacket.

In this work, the first principle model of the dual reactor is developed using MATLAB Simulink toolbox by using the differential equations that describes the process.

The widely used controller in industrial scenario is PID Controller because of its easy tuning. But designing a PID controller for the desired performance like set point tracking, disturbance rejection and robustness is difficult since the gains of the PID controller are not directly related to them.

Model Predictive Controller (MPC) is a control algorithm that uses a model of the process to predict the future response of a plant. Proportional Integral Derivative (PID) Controller cannot provide good enough performance in controlling highly complex, non-linear and uncertain processes. MPC algorithm requires the knowledge of input/output constraints and weights along with prediction Horizon and Control Horizons to be provided by the user for the control action.

MRAC is a direct adaptive strategy with adjustable controller parameters and an adjusting mechanism to adjust them. The adaptation law gives to find a set of parameters that minimize the error between the plant and the model outputs. Hence the parameters of the controller are adjusted until the error becomes zero but stability is not treated meticulously. It is in the need of high gain observes.

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RTDA controller processes the characteristics of MPC and PID controller and its structure is as simple as that of a PID controller. The tuning parameters θ_R, θ_T, θ_D and θ_A are directly related to performance parameters robustness, set point tracking, disturbance rejection and an additional parameter aggressiveness respectively. The main purpose of this paper is to design RTDA controller for controlling the temperature profile of dual CSTR and the performance of the proposed controller is compared with MPC, MRAC and PID control schemes.

This paper is organized as follows: Section 2 describes the Dual CSTR process and the steps involved in obtaining the first principle model of the same. In section 3 the procedure for obtaining the linearized model of Dual CSTR from the non-linear first principle model is discussed. Section 4 describes the design procedure of conventional and advanced control schemes for the simulated Dual CSTR. Section 5 depicts the comparison of the proposed controllers and Section 6 portrays the conclusion for the proposed work.

2. Process Description And Mathematical Modelling Of Dual Cstr

The schematic diagram of the dual tank CSTR is shown in Fig 1. By using energy balance and mass balance equations the dynamic model of Dual tank CSTR is developed. The component A with molar concentration of C_{A1} and flow rate of q is admitted as a feed material to reactor 1, which is maintained at a temperature of T_1. The concentration of tank 2 is denoted by C_{A2} and it is operated at a temperature of T_2. Stirrer is used for perfect mixing of components inside the reactor.

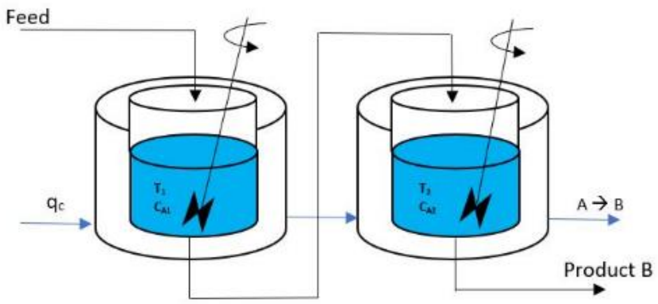

Figure 1: Dual Cstr

Keeping concentration C, constant flow rate q and temperature T, both product B and reactant A are taken out from the reactor. The coolant is circulated in the outer jacket of both the reactors to control the heat generator due to the exothermic chemical reaction. To maintain the temperature profile of both the reactors, the flow rate q_c of the coolant is to be manipulated. The coolant enters the jacket at the temperature of T_{cf}.

The objective of proposed controller is to maintain the temperature of reactor 2 (T_2), within the limit by adjusting the coolant flow rate q_c.

The dynamic model of dual CSTR is derived by the 4 differential equations i.e., 2 energy balance and 2 mass balance equations for the two reactors are depicted in equation (1) to (4) [2][14]:

$$\frac{dC_{A_1}}{dt} = \frac{q}{V_1}.\left(C_{A_f} - C_{A_1}\right) - C_{A_1}\alpha e^{\left(\frac{E}{-RT_1}\right)} \quad (1)$$

$$\frac{dC_{A_2}}{dt} = \frac{q}{V_2}.\left(C_{A_1} - C_{A_2}\right) - C_{A_2}\alpha e^{\left(\frac{E}{-RT_2}\right)} \quad (2)$$

$$\frac{dT_1}{dt} = \frac{q}{V_1}.\left(T_f - T_1\right) - \frac{C_{pc}\rho_c q_c}{V_1 C_p \rho}\left\{1 - e^{\left(\frac{U_{A_1}}{C_{pc}\rho_c q_c}\right)}\right\} * \left(T_{C_f} - T_1\right) - \frac{\Delta H \alpha C_{A_1}}{C_p \rho}.e^{\left(\frac{E}{-RT_1}\right)} \quad (3)$$

$$\frac{dT_2}{dt} = \frac{q}{V_1}.\left(T_1 - T_2\right) - \frac{C_{pc}\rho_c q_c}{V_2 C_p \rho}\left\{1 - e^{\left(\frac{U_{A_2}}{C_{pc}\rho_c q_c}\right)}\right\} * \left(T_1 - T_2\right) - \frac{\Delta H \alpha C_{A_2}}{C_p \rho}.e^{\left(\frac{E}{-RT_2}\right)} \quad (4)$$

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Where,

C_{A1},C_{A2}	Concentrations of A in reactor 1 & 2 in mol
T_1, T_2	Temperatures of reactor 1 and 2 in K

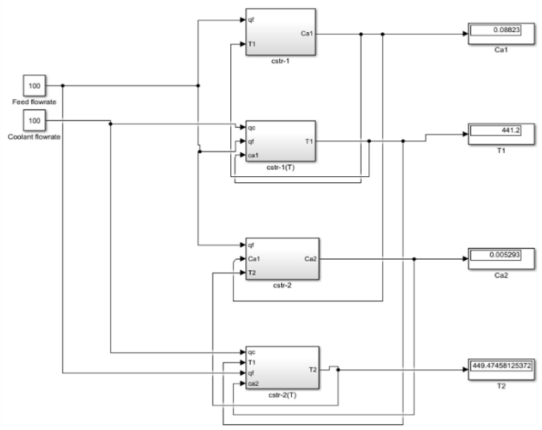

Figure 2: Simulink Model Of Dual CSTR

The Simulink model of the dual CSTR using the above differential equations is shown in Fig 2.

The nominal operating parameters used in the mathematical modeling of the dual CSTR is mentioned in TABLE 1.

Table 1: Specific Values Of Process Parameters

Process Parameters	**Values**
Flow rate (q)	100L/min
Feed Concentration (C_{Af})	1.0 mol/L
Feed Temperature (T_f)	350.0K
Coolant Temperature (T_{cf})	350.0K
Volume of Reactors (V_1, V_2)	100L
Overall Heat Transfer coefficient (U_{A1}, U_{A2})	1.67e5J/min-K
Pre-exponential Factor for A →B Arrhenius Equation (α)	7.2 e^{10}
Activation Energy / Gas Constant (E/R)	1e4K
Heat of Reaction, (ΔH)	4.784e4J/mol
Density of Fluid, (ρ)	1000g/L
Density of Coolant Fluid(ρ_c)	1000g/L
Heat Capacity of Fluid (C_p)	0.239J/g-K
Heat Capacity of Coolant Fluid (C_{pc})	0.239J/g-K
Coolant flowrate(q_c)	100L/min

Steady state operating points for the dual CSTR are taken as,

$$C_{A1_{ss}} = 0.088\ mol/L$$

$$T_{1_{ss}} = 441.219\ K$$

$$C_{A2_{ss}} = 0.00529\ mol/L$$

$$T_{2_{ss}} = 449.474\ K$$

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3. Linearization Of Dual Cstr Process

The non-linear dynamic model [1] relating two process variables u and y is

$$f(u,y) = \frac{dy}{dt} \tag{5}$$

Here nonlinear, dynamic model relating three process variables, T_1, T_2 and q_c:

$$\frac{dT_1}{dt} = f(T_1, q_c) \tag{6}$$

$$\frac{dT_2}{dt} = f(T_1, T_2, q_c) \tag{7}$$

Applying Taylor Series Expansion with $u = \bar{u}$ and $y = \bar{y}$ and truncate after the first order terms,

$$f(u,x) = f(\bar{u},\bar{y}) + \left|\frac{\partial f}{\partial y}\right|_{u'} + \left|\frac{\partial f}{\partial y}\right|_{y'} \tag{8}$$

Where $u' = u - \bar{u}$ and $y' = y - \bar{y}$.

The A,B,C and D Matrices of the state space model are obtained by the linearization using Taylor's series.

- $A = \begin{bmatrix} -7.36 & 0 \\ 1.998 & -1.178 \end{bmatrix}$
- $B = \begin{bmatrix} -0.905 \\ -0.0827 \end{bmatrix}$
- $C = \begin{bmatrix} 0 & 1 \end{bmatrix}$
- $D = 0$

The poles obtained from the state space model lie on the left half of s-plane (since the eigen values are -1.178 and -7.36) which denotes the operating points are stable. The process transfer function of the linearized model is,

$$G_p(S) = \frac{-0.0827s - 2.419}{s^2 + 8.546s + 8.68} \tag{9}$$

4. Controller Design

A. PID Controller Design

The coolant flow rate to the jacket is adjusted to maintain the temperature of reactor 2 at a desired value. The heat transfer involved in both the reactors makes the overall response sluggish and hence a PID controller would be the most appropriate controller [22]. The PID controller parameters for the above transfer function is tuned using the Ziegler–Nichols settings and are listed in TABLE 2.

Table 2: Parameters of PID Controller

Controller Parameter	Gain Value
Proportional	2.6789
Integral	1.6458
Derivative	0.136

The servo and regulatory response are obtained for a set point of 451K with disturbance in the feed flowrate which is introduced at t = 50 min (by changing from 100 L/min to 95 L/min) is shown in Fig 3.

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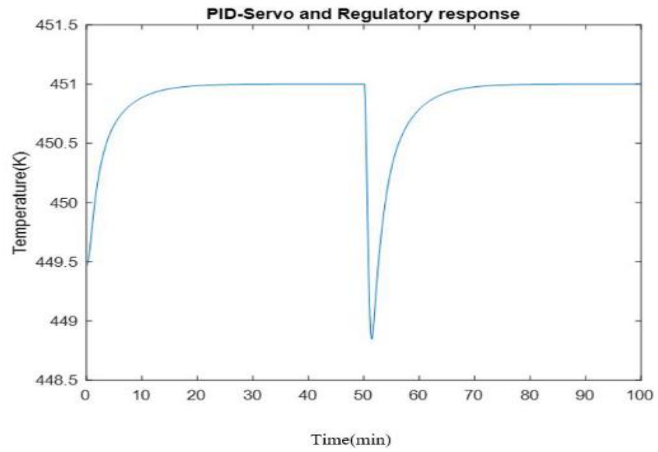

Figure 3: Response Of Pid Controller

B. Adaptive Controller

An adaptive controller has to tune its controller parameters or it has to modify the controller by accommodating the basic changes in the process behavior [3][18][19]. Hence the adaptive controller has to adjust the controller parameters continuously to

- Accommodate variations in process dynamics and disturbances
- Ensure best performance of the control system

The MIT rule based Adaptive controller [20] is used to take the control action for the Dual CSTR process [4][5][6].

The constraints to be satisfied are

- Maximum Overshoot (Mp) of 5%
- Settling time (Ts) of less than 3 seconds (In our work, 2.5 seconds is fixed as settling time).

To meet out the constraints, the damping ratio (ξ) and natural frequency (ω_n) are determined as 0.71 and 2.1834 rad/sec respectively.

Hence the transfer function of the reference model is determined as:

$$G_m(s) = \frac{4.76}{s^2+3.1s+4.76} \qquad (10)$$

The Simulink structure of proposed MRAC with MIT rule for the linearized model of the dual-CSTR[16] is shown in Fig 4.

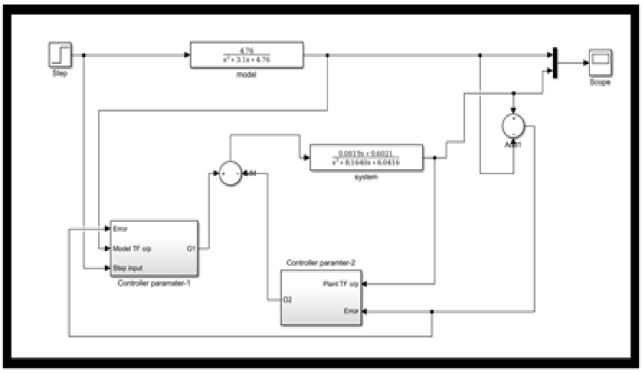

Figure 4: Mrac-Simulink Model For Linearized Model Of Dual-Cstr

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The performance of the proposed MRAC controller is validated by using different possible adaption gain values with a unit step change as a reference input and the response is shown in Fig 5.

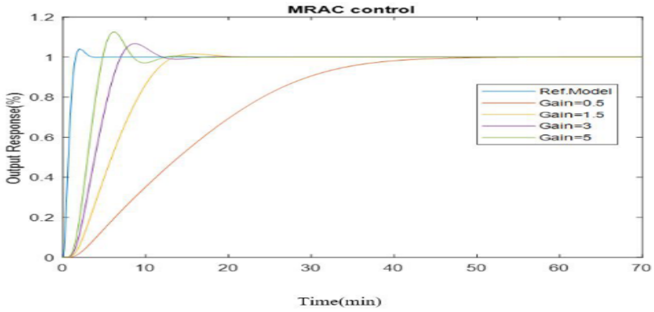

Figure 5: Response Of Mrac For Various Gains

Based on the performance of the MRAC controller, it is to be noted that when the adaptive gain is increased the settling time and rise time are decreased significantly. However adaptive gain was not increased beyond 5 since the system becomes unstable producing sustained oscillations.

The desired adaptive gain value is chosen as γ =0.8 for which the maximum peak overshoot is very negligible less than 1% (maximum overshoot of more than 5% is not desirable for a temperature control process). The servo response is plotted for temperature of reactor 2 for a set point of 451K with adaption gain of 0.8 which is shown in Fig 6. The regulatory response is obtained for a set point of 451K with disturbance in the feed flowrate which is introduced at t = 20 min (by changing it from 100 L/min to 95 L/min) is shown in Fig 7.

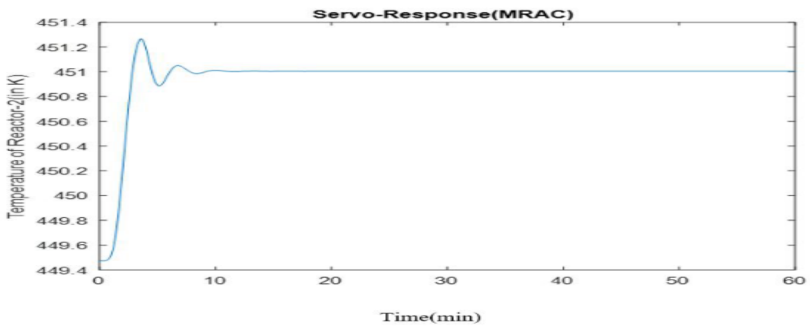

Figure 6: Servo Response Of Mrac

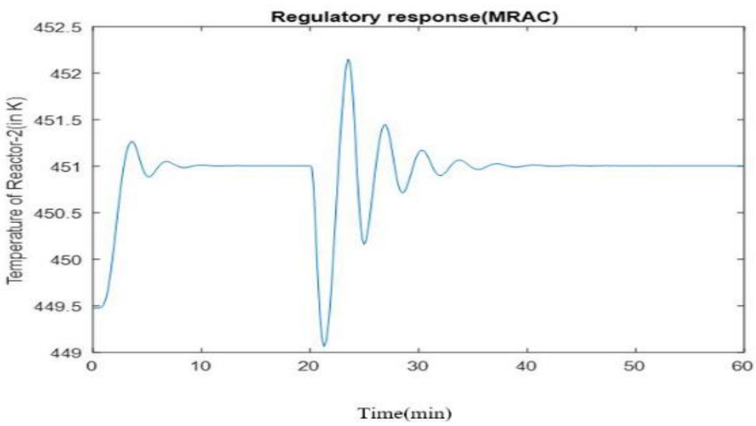

Figure 7: Regulatory Response Of Mrac

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C. DMC Based Model Predictive Control

Dynamic Matrix Control (DMC) is a control algorithm designed explicitly to predict the future response of a plant. The model predictive control [8][15] for achieving the temperature control of dual CSTR is carried out by using the linearized model as prediction model and the non-linear model as the actual plant. The MPC scheme is implemented by taking Prediction Horizon as 10 and Control Horizon as 2. The servo response of the MPC for the set point of 451K is plotted and shown in the Fig 8. The corresponding change in manipulated variable is shown in Fig 9.

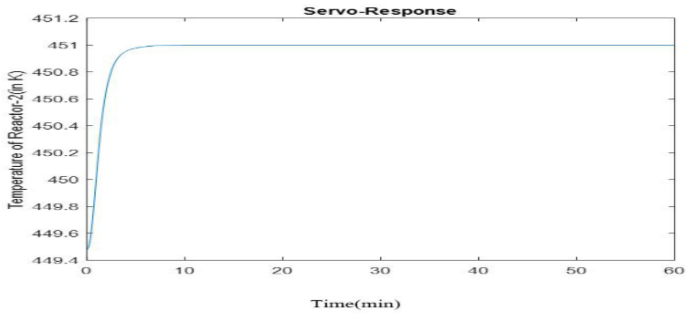

Figure 8: Servo Response Of Mpc

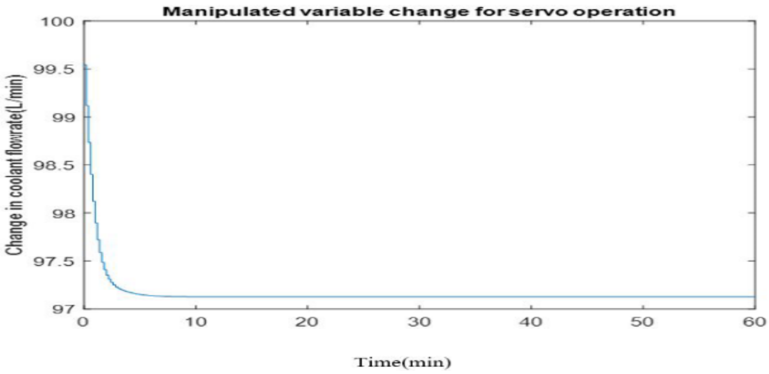

Figure 9: Controller Output Of Mpc (Coolant Flowrate)

The regulatory response of the designed MPC is obtained for a set point of 451K with disturbance in the feed flowrate which is introduced at t = 20 min (by changing it from 100 L/min to 95 L/min) is shown in Fig 10.

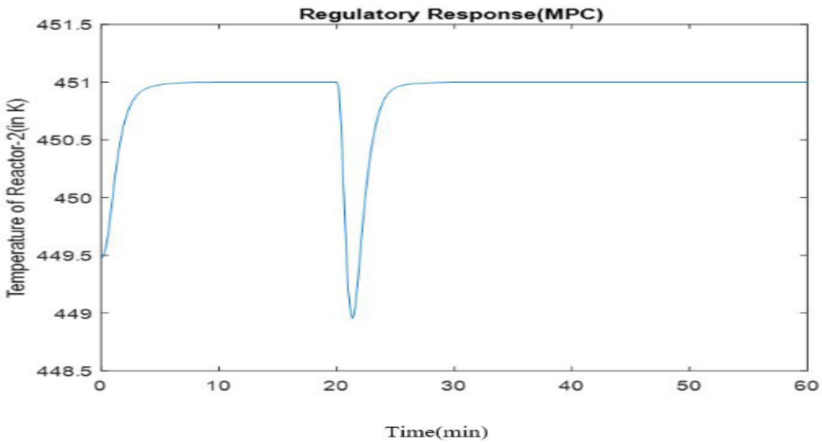

Figure 10: Regulatory Response Of Mpc

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D. RTD-A Controller

The RTD-A controller, comprises of a simplified linear model predictive control scheme which uses the process reaction curve information as like determining the parameters of PID controller. [9][21][25][26] The additional three tuning parameters, θ_R (robustness), θ_T (set point tracking) and θ_D (disturbance rejection) are normalized between 0 and 1 makes the tuning of the RTD-A controller easier. The fourth auxiliary parameter θ_A (also normalized between 0 and 1), provides the control over the controller aggressiveness which is independent of the three main tuning parameters.[10][11]

Bayesian estimation principle is used to estimate disturbance effect as,

$$\hat{e}_d(k) = \theta_R \hat{e}_d(k-1) + (1-\theta_R)e(k) \tag{11}$$

With the current disturbance estimate, the future error is then estimated to update model prediction,

$$\Delta\hat{e}_d(k) = \hat{e}_d(k) - \hat{e}_d(k-1) \tag{12}$$

The P-step ahead prediction horizon is given by,

$$\tilde{y}(k+i) = \hat{y}(k+i) + \hat{e}_d(k+i) \text{ for } 1 \le i \le P \tag{13}$$

The desired trajectory $y_t(k)$ for a given setpoint Sp(k) is represented as,

$$y_t(k) = \theta_T y_t(k-1) + (1-\theta_T)S_P(k) \tag{14}$$

$$y_t(k+i) = \theta_T^i y_t(k) + \left(1-\theta_T^i\right)S_P(k) \tag{15}$$

Equ 15 is for $1 \le i \le P$

The objective function of the controller for the proposed system is,

$$\min_{u(k)} = \sum_{i=1}^{P}\left(y_t(k+i) - \tilde{y}(k+i)\right)^2 \tag{16}$$

The control action u(k) is the minimization of the difference between model predicted output and reference trajectory for P-step,

$$u(k) = \left(\frac{1}{b}\right)\left(\frac{\sum_{i=1}^{P} \eta_i \psi_i(k)}{\sum_{i=1}^{P} \eta_i^2}\right) \tag{17}$$

$$\psi_i = y_t(k+i) - a^i\hat{y}(k) - \hat{e}_d(k+i) \tag{18}$$

Where,

N	Prediction horizon determined by θ_A
B	Process parameter
η_i	Process parameter
$\Psi i(k)$	Stipulated error

The aggressiveness parameter θ_A depends on prediction horizon P and is given by

$$P = 1 - \left(\frac{\tau}{t_s}\right)\ln(1-\theta_A) \tag{19}$$

Where, t_s is the sampling time.

Stipulated error is time varying with 3 contributing components:

1. Projected effect of past control action (uses parameters a,b,c)
2. Reference trajectory (uses θ_T)
3. Projected effect of unmeasured disturbances.

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Tuning parameters for each performance parameter of the controller is,

1. Robustness: $0 < \theta_R < 1$; depends on the current disturbance effect $\hat{e}_d(k)$
2. Setpoint Tracking: $0 < \theta_T < 1$; depends on desired output trajectory $y_t(k + i)$
3. Disturbance Rejection: $0 < \theta_D < 1$; depends on future disturbance effect prediction $\hat{e}_d(k + i)$
4. Overall Aggressiveness: $0 < \theta_A < 1$; depends on prediction horizon P

Usually, the overall aggressiveness will be specified by the control engineer. Small values of $\theta_A \sim 0$ corresponds to aggressive performance and large values of $\theta_A \sim 1$ corresponds to conservative performance.

Key Characteristics of the RTD-A

The RTD-A controller for the temperature control of non-linear dual CSTR is implemented and the individual attribute variations are analyzed [12] by keeping other attributes to a constant value of 0.3.

For different values of θ_R the response of the controller is obtained and it is observed that the error in the system decreases as the value of θ_R increases [13]. The graph is plotted for θ_R = 0.1, 0.5, 0.9 and is shown in Fig 11.

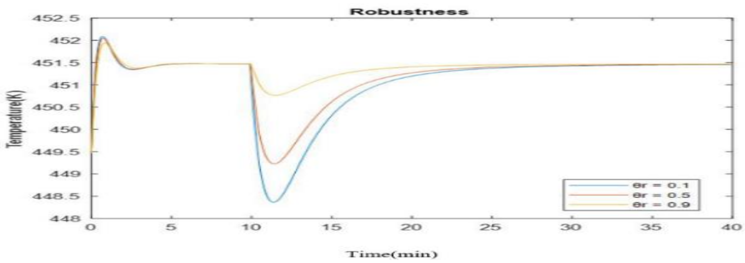

Figure 11: Robustness Using Rtd-A Controller

For the set point of 451.47K the response of the process is plotted for different values of θ_T = 0.1, 0.5, 0.9 and it is observed that for lesser values of θ_T the response is aggressive which is shown in Fig 12.

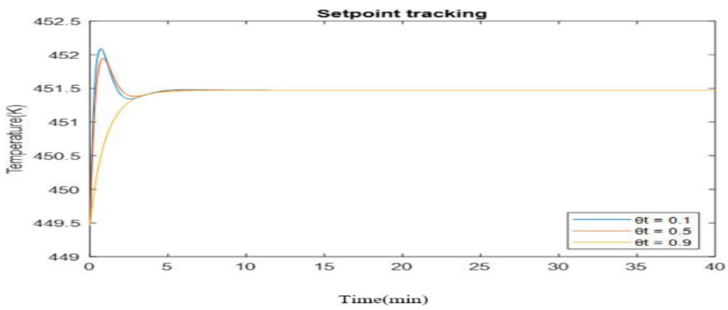

Figure 12: Set Point Tracking Using Rtd-A Controller

For validating the disturbance rejection capacity of the proposed RTDA-controller, disturbance is introduced in the feed flowrate at time instant t=10s by changing it from 100L/min to 95L/min and the responses are plotted for different values of θ_D =0.1, 0.5, 0.9 which is shown in Fig 13. It is observed that, the introduced disturbance is rejected quickly for the minimum value of parameter θ_D.

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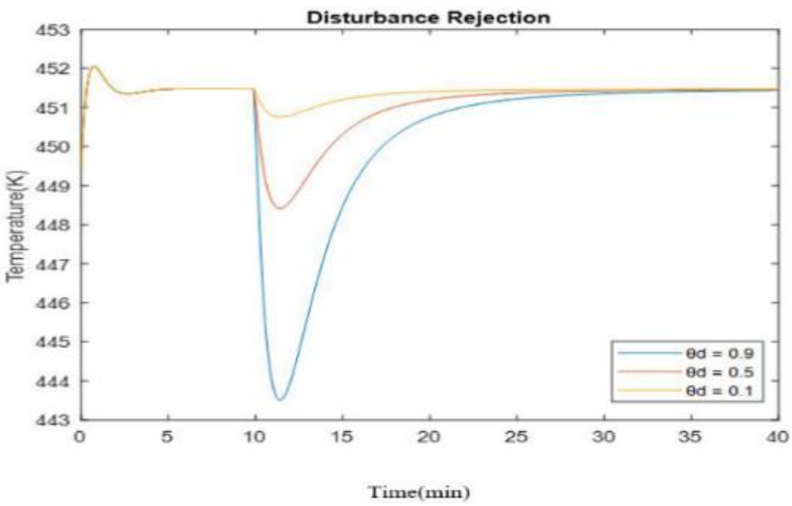

Figure 13: Disturbance Rejection Using Rtd-A Controller

For the set point of 451.47K the response of the process is plotted for different values of θ_A = 0.1, 0.3, 0.6 and it is observed that for lesser values of θ_A the response is conservative and larger values of the parameter the response is aggressive which is shown in Fig 14.

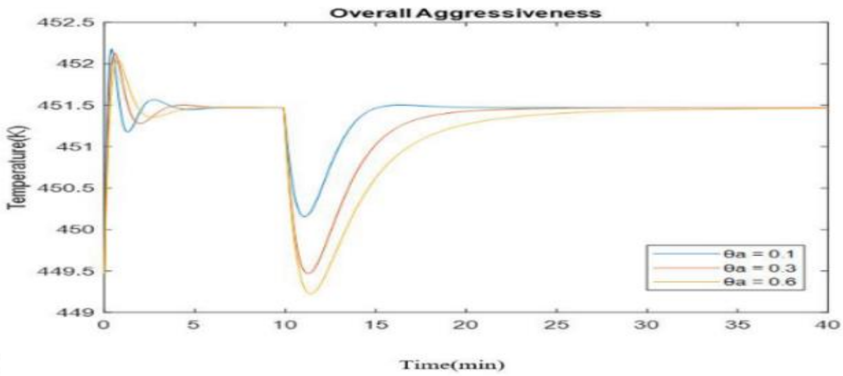

Figure 14: Overall Aggressiveness Using Rtd-A Controller

The servo and regulatory response of the designed RTD-A Controller is obtained for a set point of 451K with disturbance in the feed flowrate which is introduced at t = 10 min is shown in Fig 15.

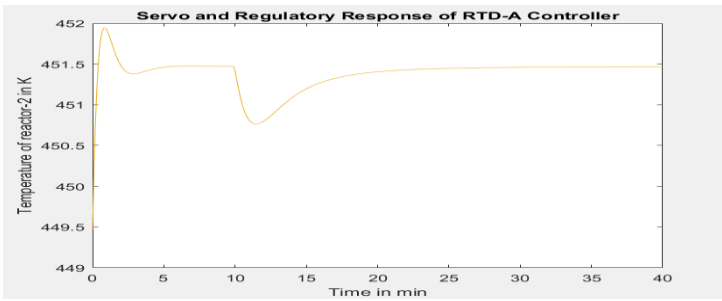

Figure 15: Servo And Regulatory Response Of Rtd-A Controller

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5.Comparison Of Proposed Controllers

The performance of the proposed controllers – PID, MRAC, MPC and RTD-A designed for Dual CSTR are compared by evaluating settling time and ISE as performance measures for both servo and regulatory operation is depicted in TABLE 3.

Table 3: Performance comparison of the four proposed control schemes

BASIS	PID	MRAC	MPC	RTD-A
Settling Time (in minutes)-Servo response	14.96	7.21	4.57	2.33
Settling Time (in minutes)- Regulatory response	14.26	14.13	5.12	3.92
Integral Squared Error (servo response)	4.04	3.46	2.175	0.593
Integral Squared Error (regulatory response)	8.15	6.27	5.51	0.451

6.Conclusion

In this work, the advanced control structure RTD-A, MPC, MRAC and PID are designed for the complex industrial process, Dual CSTR. Based on the performance of all the controllers, it is validated that RTD-A outperforms the other controllers in terms of settling time and ISE. The effectiveness of RTD-A controller is analysed by giving variation to the three tuning parameters, θ_R (robustness), θ_T (set point tracking) and θ_D (disturbance rejection) and the fourth auxiliary parameter θ_A, which provides the control over the controller aggressiveness which is independent of the three main tuning parameters. All the tuning parameters of RTD-A are normalized between 0 and 1 i.e., when the values are closes to 0 the performance will be aggressive where as they are closer to 1 the performance will be conservative. The optimization problem upon which the control action computation can be solved analytically which makes RTD-A controller a computationally simple and easily implementable in both software and hardware.

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