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RESEARCH ARTICLE

Design of centralized controller for multivariable process using MOPSO algorithm

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Abstract

Objective: To estimate centralized PID controller parameters for 4 outputs and 5 inputs crude distillation non-square system with RHP zeros process. **Methods/Analysis:** The Multi- Objective Particle Swam Optimization (MOPSO) algorithm is applied to determine the PID controller parameters for the considered distillation column process. **Findings:** The performance of the proposed controller is compared with two centralized controller schemes, Davison's and Tanttu and Lieslehto methods. The Integral Square Error (ISE), Integral Absolute Error (IAE) and Integral of Time Absolute Error (ITAE) are chosen as performance indices. The simulation results prove that MOPSO tuned centralized controller gives the best performance when compared to other analytical techniques for both set point tracking and in disturbance rejection environment. **Novelty:** In practice, conventional PID controllers are tuned using classical methods, which require complex numerical calculations. In this paper, an attempt is made to fine tune the PID controller for a MIMO process using Multi Objective optimization technique and obtained challenging results as compared to conventional methods.

Keywords: Nonsquare system; Centralized control; Multi Objective Particle Swam Optimization; PID controller

1 Introduction

Most of the industrial control processes are multivariable processes. Design of controller for a multivariable process is difficult because of the interaction effect among the loops. The interaction is mainly due to changes in one input effect with respect to several outputs. Distillation column is a major unit of operation in chemical, oil and gas processes. Design of controllers for the distillation column poses a tedious job in the process control instrumentation field. The interaction and location of transmission zeros are important in Multi-Input Multi-Output (MIMO) systems. A system is called non -minimum phase system when it has one or more right half plane (RHP) transmission zeros. These RHP zeros impose limitations on stability and controllability of the system [1,2]. Design of controller for the system with positive zeros is a major task because it will affect both the amplitude and phase angle. Adding extra phase lag by RHP zero causes instability to the process and is also difficult to control. The processes with an unequal number of inputs and outputs are called non-square systems and these systems have two control methods which are called centralized and decentralized controllers. The feedback will be used in the centralized controller and each input is manipulated from all measured outputs. In decentralized controller feedback will be implemented after pairing one output with one input.

Many heuristic algorithms such as Differential Evolution (DE), Simulated Annealing (SA), Particle Swarm Optimization (PSO), Genetic Algorithm (GA) and Bacterial Forage algorithm are powerful methods for solving many non-linear and tough optimization problems. The multi-objective particle swarm optimization (MOPSO) is one of the most promising stochastic search methods among these metaheuristics which originates from the simulation of behaviour of bird flocks, because of its easy implementation and high convergence speed.

Liu et al[3] proposed modified Internal Model Control (IMC) method and Smith delay compensator structure based on static decoupling for non-square processes with right half plant zeros and multiple time delays. The performances of the proposed controllers provided improved response when compared with other methods mentioned in the literature. Guoa et al[4] suggested Smith Decoupling compensation control method which provides good robustness and less interference performance over model mismatch on the system by model approximation using suboptimal reduction algorithm. The PSO based controller designed for spherical tank system shows improved performances when compared with various other optimization algorithms[5,6].

Madiouni[7] presented MOPSO algorithm based PID controller for various non-linear problems and proved its performances better than the non-dominated sorting genetic algorithm II in terms of better computation time. Ram and Chidambaram[8] proposed steady state gain matrix (SSGM) based centralized PI controller design for a multivariable process and validated its performance compared with controller design based on gain, time delay and time constant. Zhao et al[9] suggested controller design based on MOPSO for MIMO systems and also proved that the designed controller provided improved result in terms of ISE when compared to other optimal PID controllers. The multi-objective optimization algorithm[10,11] is applied to determine controller parameters to solve various difficult problems. Fu et al[12] employed multi objective optimization algorithm to estimate PI controller parameters to control superheated steam temperature. The designed controller improved the performance of steam temperature control for both set-point tracking and disturbance rejection. Monica et al[13] designed MOPSO based PID controller for ball and beam system and proved that the proposed controller provided improved performance compared to Skogestead's Internal Model Control in terms of time response analysis.

Perng et al[14] derived PID controller parameters using MOPSO, Genetic Algorithm, Bee colony optimization algorithm, Reinforcement Learning algorithm to water turbine governor based on the frequency domain sensitivity. The estimated PID controller performances are compared based on rise time (t_r), integral square-error, integral of time-multiplied squared-error, integral absolute error, and integral of time multiplied by absolute error. From the obtained results, they proved that MOPSO tuned PID controller provided improved performances when compared with other optimization algorithms. Gomez et al[15] suggested MOPSO based PID controller tuning for Unmanned aerial vehicles. The selected PID tuning parameters are applied to case study of quadrotor and also proved through the simulation results that MOPSO tuned PID controller provided good performance in terms of overshoot, rise time and root-mean-square error. Oliveira et al[16] proposed PSO based PID controller tuning method for Arduino-based Temperature Control Laboratory and compared its performance with the Grey Wolf Optimization algorithm in terms of integral absolute error and the total variation criteria. Through the obtained result it is observed that the PSO tuned controller provided improved control performance than the other optimization algorithm.

Gomez et al[17] introduced MOPSO based controller tuning procedure to control altitude for Px4- based Unmanned aerial vehicles. From the simulation results it is evident that the proposed controller gives good performance in terms of rise time, overshoot and root-mean-square error of step response of the P-PID controllers. Al-Khazraji et al[18] designed MOPSO algorithm to select the best system control parameter for production–inventory system with multivariable input and multivariable output. He tested this algorithm at automatic pipeline, inventory, and order based production control system (APIOBPCS) model and the newly modified two automatic pipeline inventory and order based production control system (2APIOBPCS) model for optimal control of production. The simulation results indicated that 2APIOBPCS model performed better than the APIOBPCS model to achieve an optimal performance in terms of balancing the order rate and stock level under different conditions.

In this work, the crude distillation column by Levein was considered for study. The MOPSO based optimization algorithm is used to determine the gain matrix. The controller performance is compared with Davison and Tanttu and Lieslehto control methods in terms of ISE, IAE and ITAE. From the simulation results it is identified that MOPSO optimization tuned controller gives improved response than the other methods.

2 Multi Objective Particle Swam Optimization

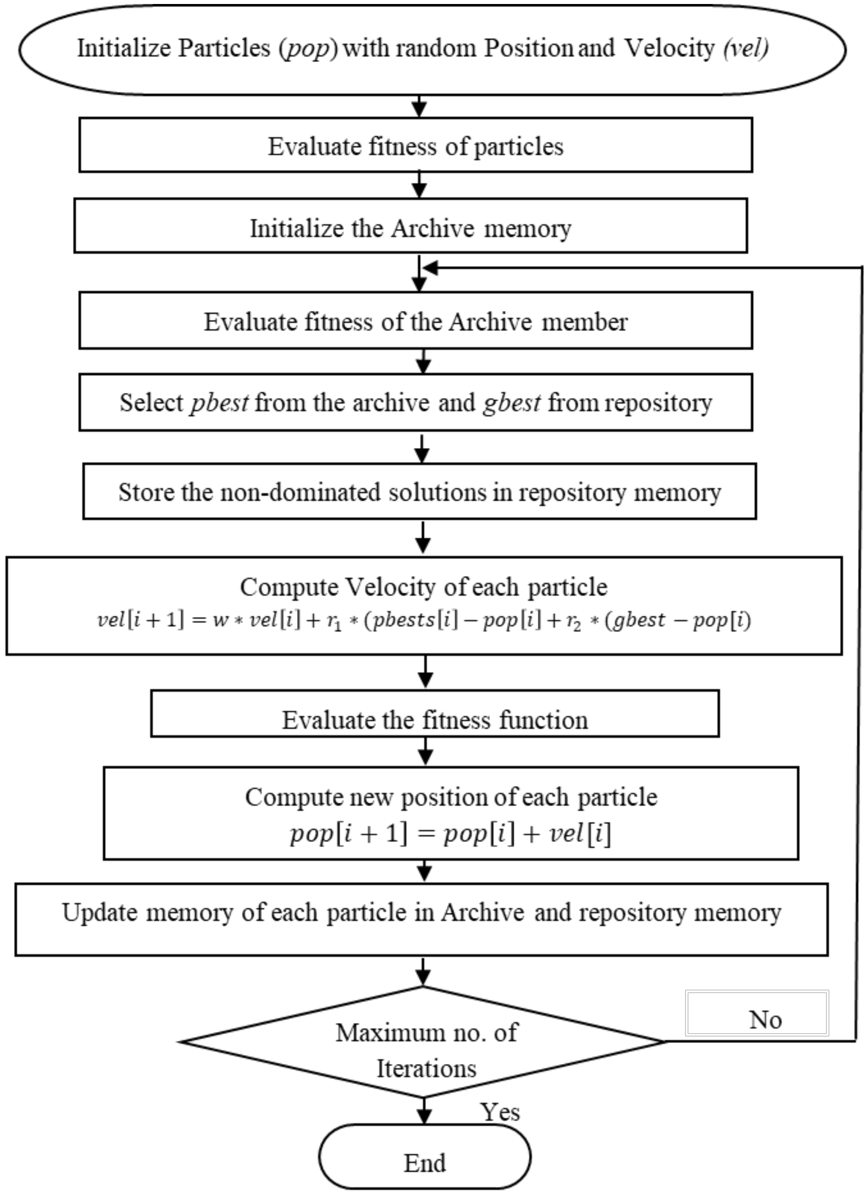

Fig 1. Flow chart of MOPSO

Particle Swarm Optimization (PSO) technique, is an evolutionary-type global optimization algorithm developed by Kennedy and Eberhart [19] based on social activities in flock of birds and school of fish and is widely applied to solve various engineering problems because of its simplicity and high computational efficiency problems that have more than one objective which is referred to as multi-objective optimization. The multi objective problem is found in various fields such as mathematics, engineering, agriculture, social studies, economics, aviation, etc.

Multi-objective optimization (MOO) problems consist of several objectives that are needed to be handled simultaneously. The most suitable MOO problems are Pareto based approaches, because of its ability to generate multiple solutions in less computation time [20]. Due to the high speed of convergence PSO based algorithms are suitable for multi-objective optimization problems. The flowchart shows Multi Objective Particle Swarm Optimization (MOPSO) algorithm and it uses the idea of a repository of particles and a mutation operator to get better solution. In detail, at each iteration, the objective value is calculated for each individual and then used to determine the relationship of non-dominance in the population in order to select a potentially better solution [21].

MOPSO is applied to estimate PID controller parameter for the crude distillation column. Distillation column is a major unit operation in chemical, oil and gas processes. Design of controllers for the distillation column poses a tedious job in the process control instrumentation field. The transfer function of crude distillation unit is shown in equation (4). The process has 4 outputs and five inputs. The controlled variables are naphtha/kerosene cut-point (y1), kerosene/light gas oil (LGO) cut-point (y2), LGO/ heavy gas oil (HGO) cut-point (y3) and measured over flash (y4). Manipulated variables are top temperature (u1), kerosene yield (u2), LGO yield (u3), HGO yield (u4) and heater outlet temperature (u5).

Figure 2 illustrates the basic block diagram of an optimization algorithm based PID controller tuning for a considered distillation process. The MOPSO algorithm based controller design is attempted to study and improve the performance of distillation unit. The controller design process is to select the suitable values for tuning parameters from the search space that minimizes the objective function. The objectives are minimization of ISE, IAE and ITAE values. The accuracy of the heuristic algorithm based controller tuning mainly depends on the objective function, which guides the optimization search to estimate suitable controller parameter values.

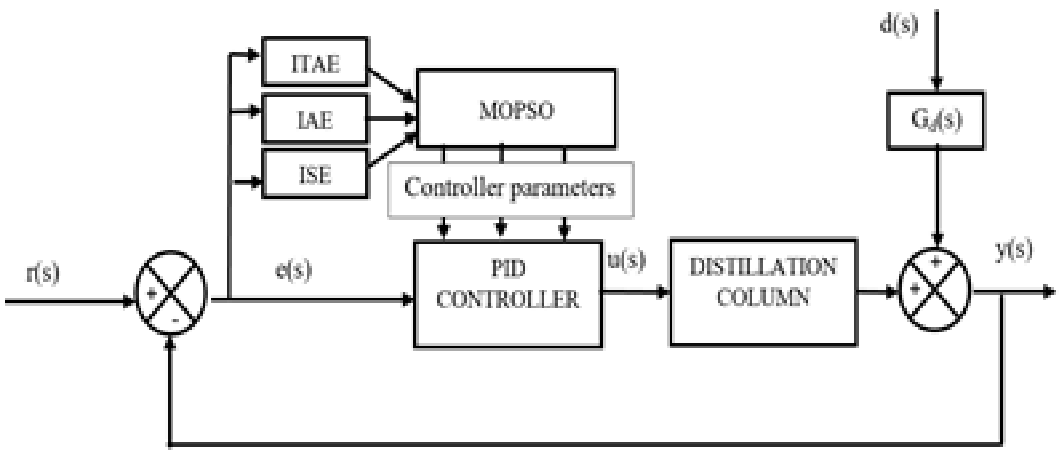

Fig 2. Block diagram of optimization scheme

3 Centralized Controller Design

3.1 Davison's Method

Centralized multivariable PID controller tuning method as suggested by Davison [1] for square systemsis discussed as follows. The proportional and integral gain matrices are given by

$$K_c = \delta[G(S=0)]^{-1} \tag{1}$$

$$K_I = \varepsilon[G(S=0)]^{-1} \tag{2}$$

Where $[G(S{=}0)]^{-1}$ is called the rough tuning matrix, and δ and ε are the fine-tuning parameters, generally its range from 0 to 1. This method is also applied to non-square system. There exists no inverse for non-square system. The Moore-Penrose pseudo inverse is used. For matrix A, it is given by

$$A^{\dagger} = A^{H}\left(A \times A^{H}\right)^{-1} \tag{3}$$

Where $A^{\dagger}$ is the inverse of non- square matrix and A^{H} is the Hermitian matrix of A. The PID controller gains for a non-square system are given by

$$K_c = \delta[G(S=0)]^{\dagger} \tag{4}$$

$$K_I = \varepsilon[G(S=0)]^{\dagger} \tag{5}$$

$$K_D = [G(S=0)]^{\dagger} \tag{6}$$

3.2 Tanttu and Lieslehto method

Based on IMC principles Tanttu and Lieslehto[1] have devised a multivariable PI controller tuning method. For a first order time delay process,

$$K_{cij} = \left(2_{\tau i\ j} + I_{ij}\right)/2\lambda k_{ij} \tag{7}$$

$$\tau_{Iij} = \tau_{ij} + 0.5\, L_{ij} \tag{8}$$

$$K_{Iij} = K_{cij}/\tau_{Iij} \tag{9}$$

where kij and Lij are the process gain and dead time of an element in model of the process for the i^{th} output and j^{th} input. The proportional gain and integral time constant are Kcij and $\tau_I$ij of the internal model controller of the ij^{th} loop. Then the multivariable PID controllers are designed by taking the pseudo-inverse for non-square system

$$K_c = \left[1/K_{cij}\right]^{\dagger} \tag{10}$$

$$K_I = [1/K_{Iij}^{\dagger}] \tag{11}$$

$$K_D = \left[1/K_{Dij}\right]^{\dagger} \tag{12}$$

where K_{Iij} is the integral gain constant of the ij^{th} loop

3.3 MOPSO Algorithm based Tuning method

The MOPSO is applied to estimate a centralized PID controller gain matrix with RHP zero. The Integral Squared Error (ISE), Integral Absolute Error (IAE), and Integral of the Time-Weighted Absolute Error (ITAE) are used as performance indices for minimization of error.

$$ISE = \int_0^t e^2(t)dt \tag{13}$$

$$IAE = \int_0^t |e(t)|dt\,| \tag{14}$$

$$ITAE = \int_0^t t|e(t)|dt \tag{15}$$

Where t is the simulation time =150 sec

4 Result and Discussion

4.1 Simulation Studies

To analyze the effectiveness of the centralized controller for a non-square process, we considered a crude distillation column with 5-input and 4-output process for simulation. The process transfer function [1] is given by,

$$G(s)=\begin{bmatrix} \begin{array}{c}\dfrac{3.8(16s+1)}{140s^2+14s+1}\\ \dfrac{-0.73(-16s+1)e^{-4s}}{150s^2+20s+1}\end{array} & \dfrac{2.9e^{-6s}}{10s+1} & 0 & 0 \\ \begin{array}{c}\dfrac{3.9(4.5s+1)}{96s^2+17s+1}\\ \dfrac{16se^{-2s}}{(5s+1)(14s+1)}\end{array} & \dfrac{6.3}{20s+1} & 0 & 0 \\ \begin{array}{c}\dfrac{3.8(0.8s+1)}{23s^2+13s+1}\\ \dfrac{22se^{-2s}}{(5s+1)(10s+1)}\end{array} & \dfrac{6.1(12s+1)e^{-s}}{337s^2+34s+1} & \dfrac{3.4e^{-2s}}{6.9s+1} & 0 \\ \begin{array}{c}\dfrac{-1.62(5.3s+1)e^{-s}}{13s^2+13s+1}\\ \dfrac{0.32(-9.1s+1)e^{-s}}{12s^2+15s+1}\end{array} & \dfrac{-1.53(3.1s+1)}{5.1s^2+7.1s+1} & \dfrac{-1.3(7.6s+1)}{4.7s^2+7.1s+1} & \dfrac{-0.6e^{-s}}{2s+1} \end{bmatrix} \tag{16}$$

The steady-state gain matrix of the above model and the centralized PID controller matrix is determined by using Davison's, Tanttu and Lieslehto methods as elaborated by Sarma and Chidambaram [1].

4.2 MOPSO based Centralized Controller method for crude distillation column

The entire operation of the crude distillation unit at Cosmo Oil's Sakai Refinery is discussed by Sarma and Chidambaram [1]. Simulation is carried out 50 iteration for the considered process and the tuning parameter values are obtained by assuming minimum and maximum values with 20 populations to find the global solution in a less competitive time. The PID controller matrix based on MOPSO algorithm is estimated as

$$G_C=\begin{bmatrix} 0.3519+\dfrac{0.2514}{s}+0.1661s & 0.0974+\dfrac{0.1142}{s}-0.1337s & 0.138+\dfrac{0.0115}{s}+0.117s & 0.0288+\dfrac{0.0261}{s}+0.0279s \\ -0.1380-\dfrac{0.1729}{s}-0.1459s & 0.1804+\dfrac{0.1357}{s}+0.1648s & 0.0099-\dfrac{0.0096}{s}-0.0110s & -0.0254-\dfrac{0.0238}{s}-0.213s \\ -0.0015-\dfrac{0.0019}{s}-0.0014s & -0.1411-\dfrac{0.0894}{s}-0.1859s & 0.1492+\dfrac{0.1356}{s}+0.1663s & -0.00029-\dfrac{0.00025}{s}-0.00022s \\ -0.2890-\dfrac{0.2932}{s}-0.3165s & 0.1965+\dfrac{0.3292}{s}+0.2615s & -0.4712-\dfrac{0.4215}{s}-0.2619s & -0.8225-\dfrac{1.2905}{s}-0.1769s \\ -0.0758-\dfrac{0.0665}{s}-0.0953s & 0.0386+\dfrac{0.0502}{s}+0.0352s & 0.0425+\dfrac{0.0299}{s}+0.0318s & 0.0959+\dfrac{0.0635}{s}+0.0819s \end{bmatrix}$$

(17)

5 Simulation Results

Crude distillation is a large-scale problem with 4 outputs and 5 inputs. Simulation study is carried out on the Matlab/Simulink platform for both servo and regulatory problems with step input. Simulation is carried out for 150 seconds, keeping a sampling period of 1 second. Results are compared in terms of performance indices such as IAE, ISE and ITAE values for all the three controller methods. The error values are shown in Tables 1, 2 and 3. From the tabulated values it is observed that MOPSO tuned controller provides the lowest error values (ISE, IAE, ITAE) than the other two analytical methods for applying step changes in Y1, Y2, Y4 for servo problems. For the given step change at Y3 the Davison controller gives lowest error values than the optimized Tanttu controllers. The Tanttu controller gives four times higher error values than the other two controllers.

A step change is applied in set point y1 and the corresponding output responses are compared which is shown inFigure 3. From Figure 3, it is noted that MOPSO controller quickly follows the changes in the set point with high overshoot but settled in less time period compared to other methods. Settling time of optimization tuned controller decreased around 15% as compared to Davison method; Tanttu and Lieslehto method takes more settling time than the other methods. Similarly set point change is also applied in Y2, Y3 and Y4 and its corresponding output responses are recorded in Figures 4, 5 and 6. From the Figures 4 and 5 it is noted that the optimization algorithm based controller response has greatly improved with regards to settling time and rise time with initial high overshoot. When we apply the step change at Y4 and its corresponding response as shown in Figure 6, it is observed that MOPSO controller provides less undershoot and quickly reaches the set point and also gives reduced settling time compared to Davison and Tanttu methods. The settling time and undershoot of Davison method is more compared to the proposed method and less by Tanttu method.

The designed controller's performance is also compared for change in load variables. Figures 7, 8, 9, 10 and 11 show comparison of the output variable for unit step change in load variables d1, d2, d3, d4, d5 respectively. From the Figure 7 , it is noted that the applied disturbances are eliminated quickly within 60 seconds by the MOPSO controller with less overshoots when compared to other two methods. The Davison method provides more overshoot than optimized controller method and less overshoot than Tanttu method. From all the regulatory responses (Figures 8, 9, 10 and 11) the Tanttu controller takes 2-3 times more to eliminate the applied disturbance. The error values for regulatory operations are shown in Tables 4, 5 and 6. From Table 4 , it clearly indicates that MOPSO algorithm tuned controller exhibits good response and also provides 2-3 times lower ISE values than other two controller techniques.

When we apply step change in load variable d3, the ISE values of the optimized and the Davison controllers are nearly equal as compared to Tanttu method. From Table 5, it is evident the MOPSO tuned controller for all unit step changes d1, d2, d4, d5 provide 2-3 times lesser IAE values than other analytical control methods. When we apply unit step change at d3, Davison based controller given less IAE values than the proposed and the Tanttu controllers. From Table 6, it is noted that the MOPSO controller provided two times lesser ITAE values than the Davison controller and five times lesser than the Tanttu controller.

Performance comparison of MOPSO, Davison's method and Tanttu and Lieslehto method for unit step changes

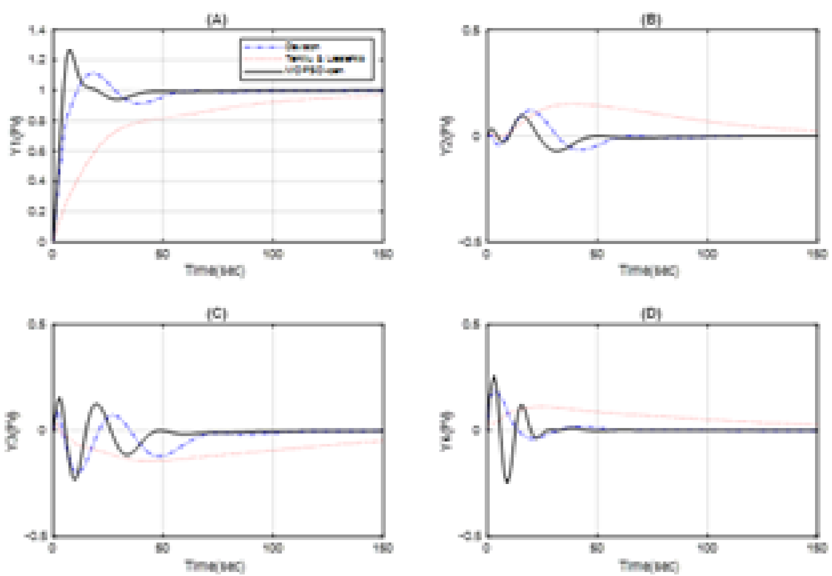

Fig 3. Servo problem for (A) y1, (B) y2, (C) y3 and (D) y4 for crude distillation process from y1 step input changes

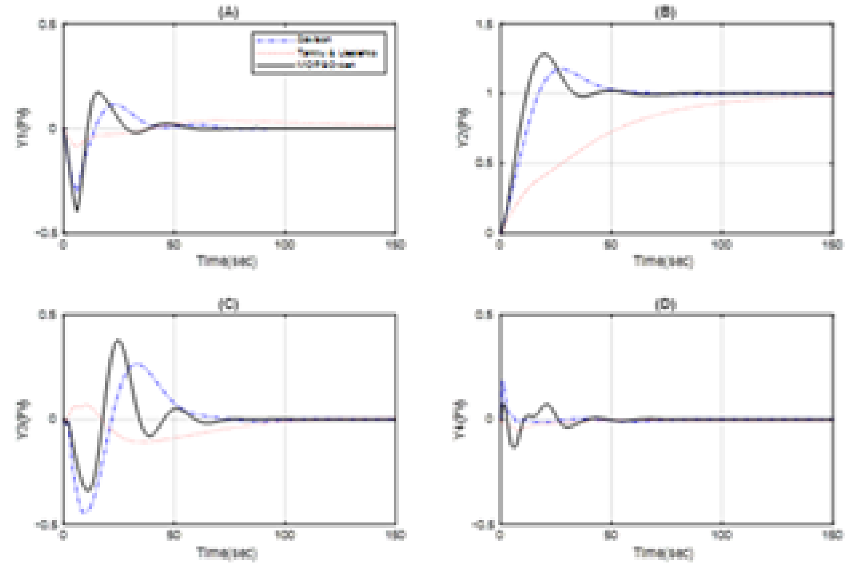

Fig 4. Servo problem for (A) y1, (B) y2, (C) y3 and (D) y4 for crude distillation process from y2 step input changes

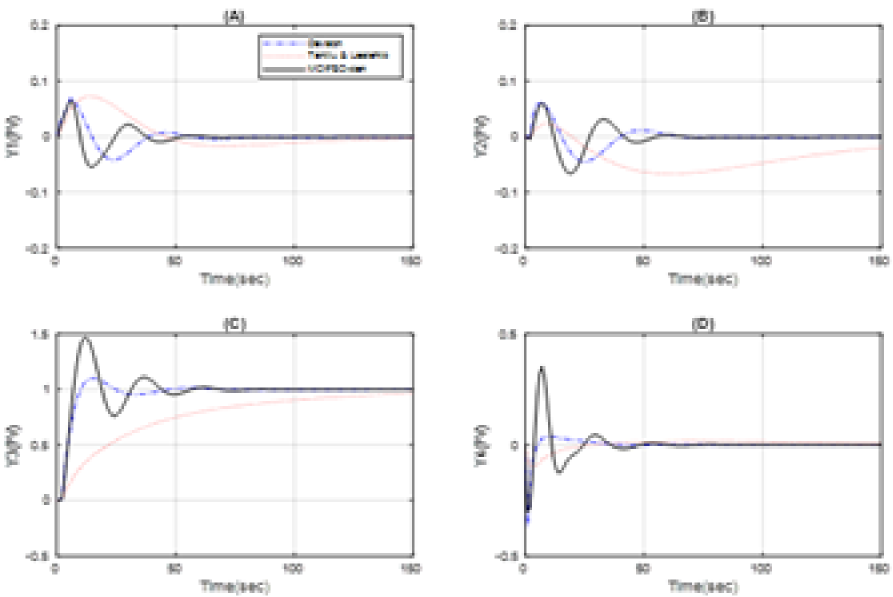

Fig 5. Servo problem for (A) y1, (B) y2, (C) y3 and (D) y4 for crude distillation process from y3 step input changes

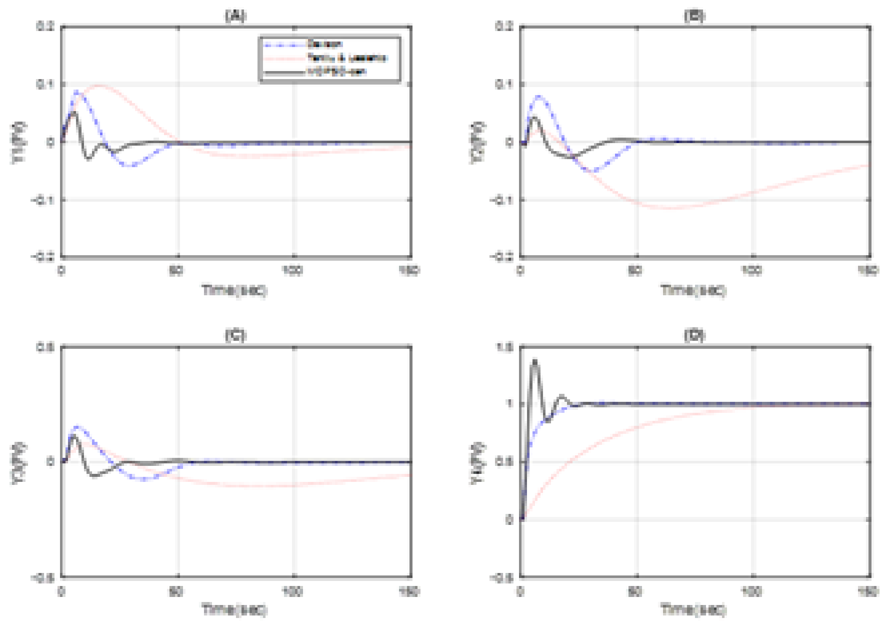

Fig 6. Servo problem for (A) y1, (B) y2, (C) y3 and (D) y4 for crude distillation process from y4 step input changes

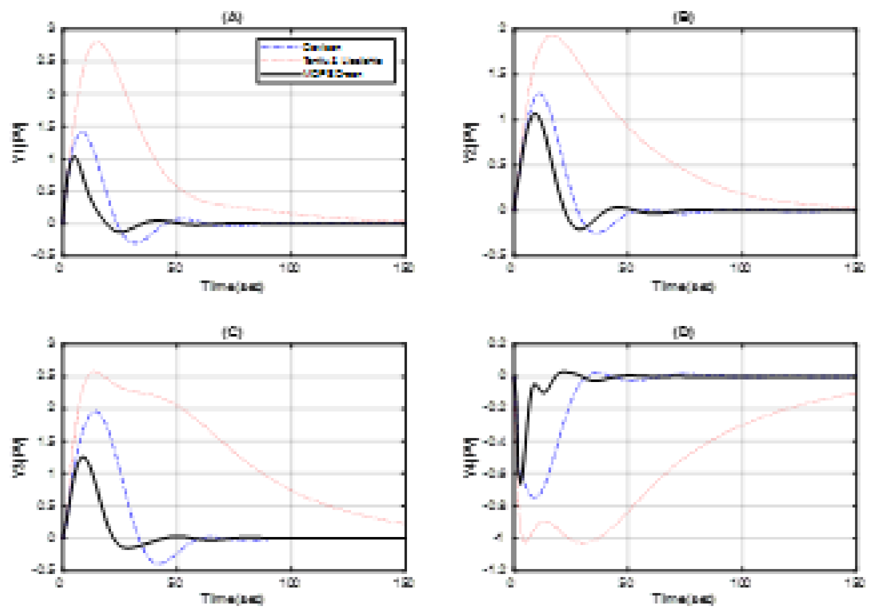

Fig 7. Regulatory problem for (A) y1, (B) y2, (C) y3 and (D) y4 for crude distillation process from y1 step input changes

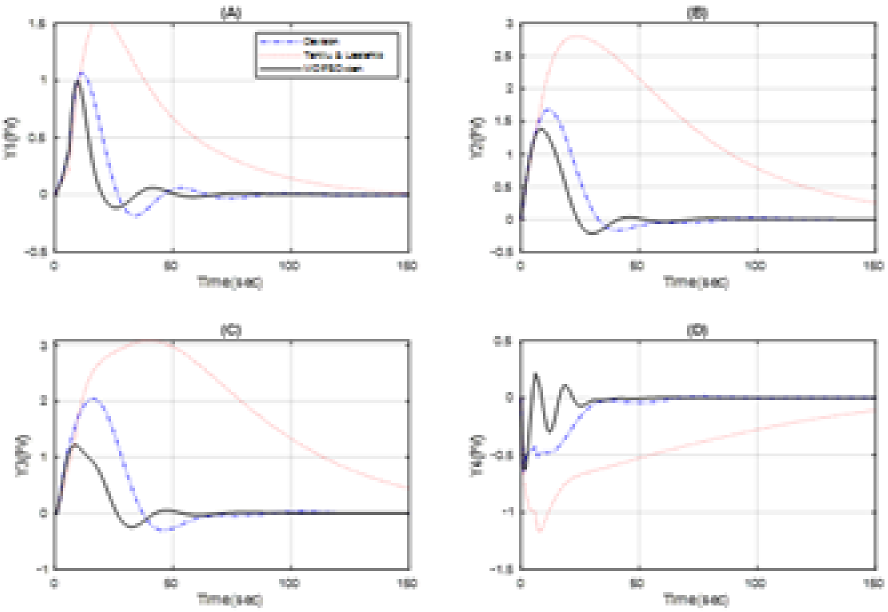

Fig 8. Regulatory problem for (A) y1, (B) y2, (C) y3 and (D) y4 for crude distillation process from y2 step input changes.

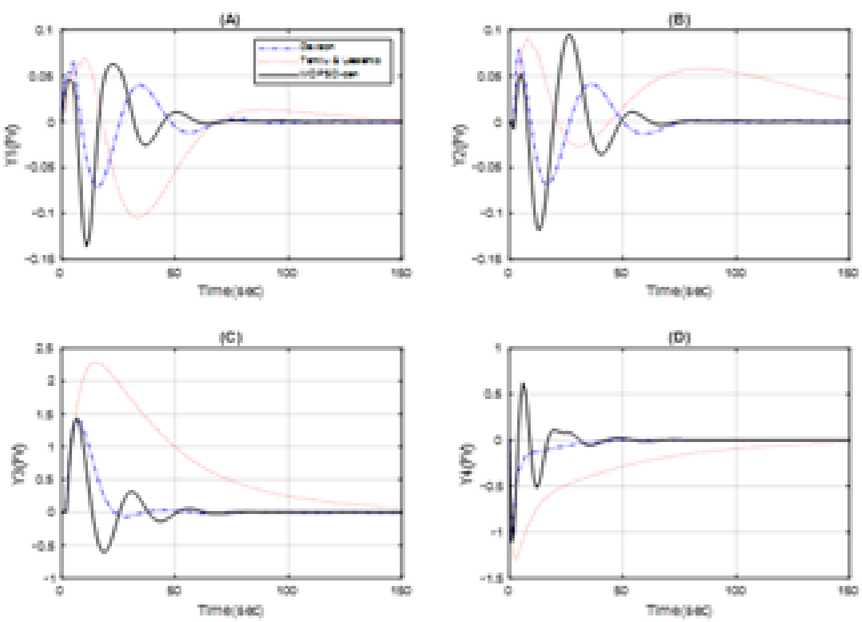

Fig 9. Regulatory problem for (A) y1, (B) y2, (C) y3 and (D) y4 for crude distillation process from y3 step input changes

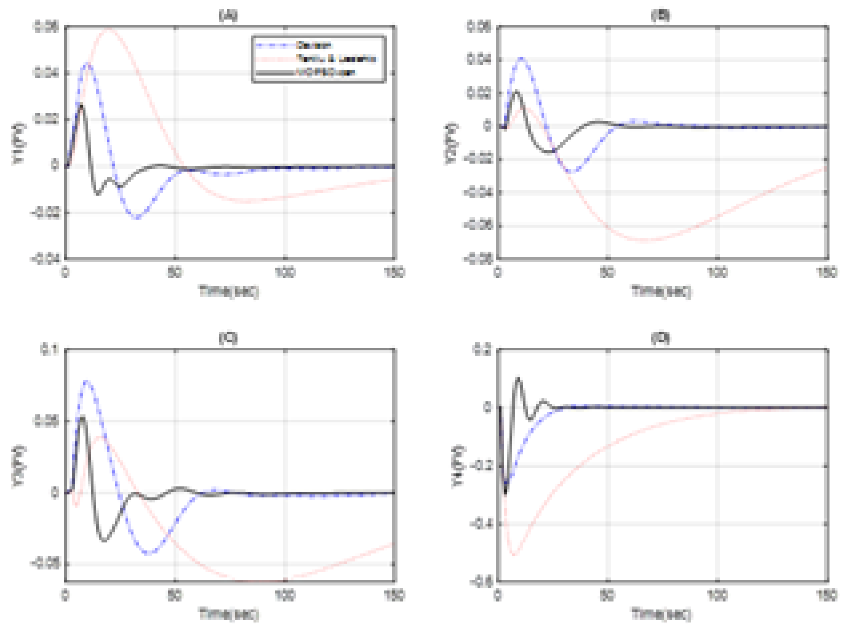

Fig 10. Regulatory problem for (A) y1, (B) y2, (C) y3 and (D) y4 for crude distillation process from y4 step input changes

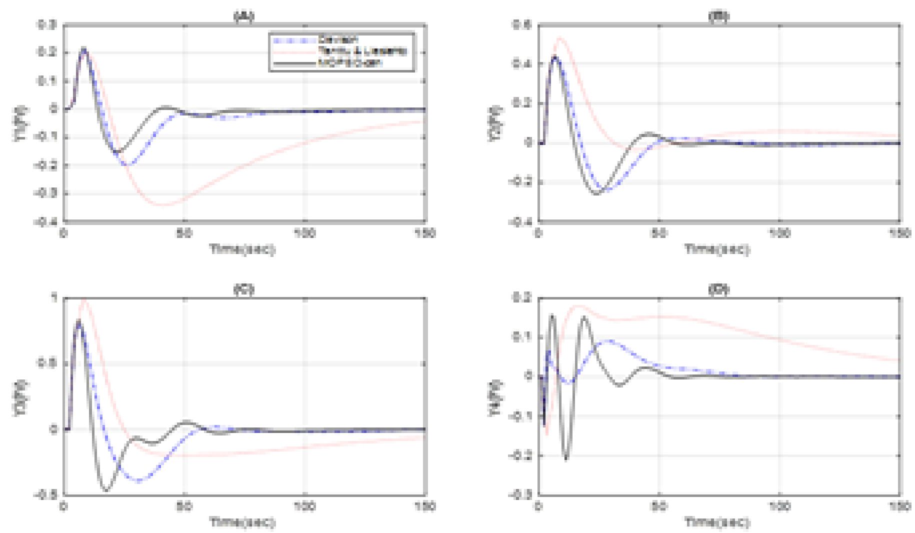

Fig 11. Regulatory problem for (A) y1, (B) y2, (C) y3 and (D) y4 for crude distillation process from y5 step input changes.

Table 1. ISE values of the servo problem for centralized controller

METHOD	STEP IN	ISE Values				SUM OF ISE
		Y1	Y2	Y3	Y4	
Davison's	Y1	2.468518	0.221825	0.604317	0.248185	3.542845
	Y2	0.620711	6.528954	3.040682	0.058512	10.24886
	Y3	0.049365	0.0514	4.320502	0.191621	4.612888
	Y4	0.084841	0.08828	0.295454	3.010368	3.478943
Tanttu & Lieslehto	Y1	11.71437	1.317035	1.515135	0.774036	15.32058
	Y2	0.149957	18.90483	0.439583	0.033135	19.5275
	Y3	0.250799	0.926327	0.911667	15.93994	18.02873
	Y4	15.71780	10.81561	2.490706	0.104345	29.12846
MOPSO	Y1	2.011713	0.116685	0.521512	0.461725	3.11163
	Y2	0.895941	5.687901	2.065506	0.114612	8.76396
	Y3	0.042204	0.05744	5.999197	0.715217	6.81405
	Y4	0.016358	0.015908	0.077131	2.951472	3.06087

Table 2. IAE values of the servo problem for centralized controller

METHOD	STEP IN	IAE Values				SUM OF IAE
		Y1	Y2	Y3	Y4	
Davison's	Y1	7.017111	3.317635	6.096304	2.448388	18.87944
	Y2	4.464637	12.75727	11.57743	0.914591	29.71393
	Y3	1.359765	1.419553	7.299364	1.357556	11.43624
	Y4	1.904284	1.926667	3.583132	5.768111	13.18219
Tanttu & Lieslehto	Y1	29.06583	12.23885	14.20777	9.984795	65.49724
	Y2	4.261004	37.13822	6.044405	1.698667	49.14229
	Y3	4.626874	10.59989	10.86113	30.37771	56.4656
	Y4	42.91012	38.79753	14.15686	3.330265	99.19478
MOPSO	Y1	5.201128	2.284543	4.622425	2.869113	14.97721
	Y2	4.55005	10.72069	8.394566	1.751351	25.41666
	Y3	1.159747	1.412987	11.91399	3.736781	18.22351
	Y4	12.24083	17.15978	18.52329	4.07642	7.937867

Table 3. ITAE values of the servo problem for centralized controller

METHOD	STEP IN	ITAE Values				SUM OF ITAE
		Y1	Y2	Y3	Y4	
Davison's	Y1	1052.567	497.6453	914.4456	367.2582	2831.916
	Y2	669.6956	1913.591	1736.614	137.1887	4457.09
	Y3	203.9648	212.9329	1094.905	203.6334	1715.436
	Y4	3912.632	3856.545	6969.586	2166.736	1977.329
Tanttu & Lieslehto	Y1	4359.875	1835.827	2131.165	1497.719	9824.586
	Y2	639.1506	5570.732	906.6608	254.8	7371.344
	Y3	694.031	1589.984	1629.169	4556.656	8469.84
	Y4	6436.518	5819.629	2123.529	499.5397	14879.22
MOPSO	Y1	780.1693	342.6814	693.3637	430.367	2246.581
	Y2	682.5074	1608.104	1259.185	262.7026	3812.499
	Y3	173.962	211.948	1787.099	560.5172	2733.526
	Y4	95.62492	109.0247	203.2698	782.7607	1190.68

Table 4. ISE values of the regulatory problem for centralized controller

METHOD	STEP IN	ISE Values				SUM OF ISE
		Y1	Y2	Y3	Y4	
Davison's	d1	23.6427778	21.90663	62.36149	8.100033	116.0109
	d2	11.9973427	41.79147	73.76714	4.622125	132.1781
	d3	0.07930666	0.081288	16.70909	2.568755	19.43844
	d4	0.02542171	0.026355	0.09004	0.513724	0.65554
	d5	0.79236529	2.166121	6.734299	0.175927	9.868713
Tanttu & Lieslehto	d1	182.595704	123.2726	351.5919	61.504	718.9642
	d2	68.3814179	396.1179	613.2996	38.62084	1116.42
	d3	0.27870924	0.282236	160.3206	22.8267	183.7082
	d4	0.0898157	0.331235	0.307326	5.201697	5.930075
	d5	5.4158826	3.312883	11.70408	2.11644	22.54928
MOPSO	d1	7.76270852	11.89699	15.32378	1.515756	36.49923
	d2	6.22646555	21.72669	18.63809	1.466382	48.05763
	d3	0.13085538	0.161931	14.38668	4.5622731	19.24174
	d4	0.00431723	0.005052	0.021837	0.2896073	0.320813
	d5	0.49058153	1.91373	5.091145	0.3617842	7.857241

Table 5. IAE values of the regulatory problem for centralized controller

METHOD	STEP IN	IAE Values				SUM OF IAE
		Y1	Y2	Y3	Y4	
Davison's	d1	26.08421	25.7103	46.46391	14.44491	112.7033
	d2	18.24468	36.30209	52.05609	12.16026	118.7631
	d3	1.889074	1.911063	17.08375	5.882622	26.76651
	d4	1.08813	1.087037	2.056617	3.086636	7.31842
	d5	6.317469	9.163132	16.7351	3.044152	35.25986
Tanttu & Lieslehto	d1	104.3735	98.88363	195.9327	82.20293	481.3927
	d2	71.03731	206.2506	268.9047	65.63882	611.8314
	d3	4.349445	5.810961	112.7541	40.44941	163.3639
	d4	2.761562	6.270148	6.188787	17.89926	33.11976
	d5	24.36947	12.68927	29.70691	16.5995	83.36515
MOPSO	d1	12.24083	17.15978	18.52329	4.07642	52.00033
	d2	10.97219	23.19806	22.97597	4.90838	62.0546
	d3	2.024289	2.404323	19.15299	8.770329	32.35193
	d4	0.35452	0.422865	0.764441	1.690606	3.232432
	d5	4.284821	8.001342	12.14946	3.069841	27.50546

Table 6. ITAE values of the regulatory problem for centralized controller

METHOD	STEP IN	ITAE Values				SUM OF ITAE
		Y1	Y2	Y3	Y4	
Davison's	d1	3912.632	3856.545	6969.586	2166.736	16905.5
	d2	2736.702	5445.313	7808.414	1824.039	17814.47
	d3	283.3612	286.6595	2562.563	6067.411	4014.977
	d4	163.2195	163.0556	308.4925	462.9954	1097.763
	d5	947.6203	1374.47	2510.265	456.6228	5288.978
Tanttu & Lieslehto	d1	15656.02	14832.55	29389.91	12330.44	72208.91
	d2	10655.6	30937.58	40335.7	9845.823	91774.71
	d3	652.4167	871.6442	16913.11	6067.411	24504.59
	d4	414.2343	940.5222	928.318	2684.889	4967.964
	d5	3655.42	1903.39	4456.037	2489.925	12504.77

Continued on next page

Table 6 continued

MOPSO	d1	1836.125	2573.968	2778.494	611.4629	7800.049
	d2	1645.829	3479.709	3446.396	736.257	9308.19
	d3	303.6434	360.6484	2872.948	1315.549	4852.789
	d4	53.17807	63.42971	114.6662	253.5909	484.8648
	d5	642.7232	1200.201	1822.419	460.4761	4125.82

6 Conclusion

The MOPSO algorithm based optimization technique is used to design centralized controller for non-square systems. The performances of the controller technique are compared with other simple centralized controller methods such as Davison's method, Tanttu and Lieslehto method. Simulations are carried out for both servo and regulatory operations. The IAE, ISE and ITAE values are tabulated. The MOPSO algorithm based controller method has shown the lowest error values for both servo and regulatory operations and better performance compared to other two methods in terms of settling time. The MOPSO controller reduced about 15% lesser settling time compared to other controllers with higher overshoot. Also, the proposed controller reduces the error two times lesser than Davison controller and 4-5 times lesser than Tanttu controller. A possible direction for future work is that, this algorithm may be combined with other optimization algorithms to improve the convergence. Convergence of MOPSO algorithm seems to be premature for complex large scale systems which limit the searching efficiency for global optimal solution.

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