International Journal of Applied Engineering Research
ISSN 0973-4562 Volume 10, Number 4 (2015) pp. 8935-8952
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Particle Swarm Optimization Tuned Fractional Order PID Controller For A Non-Linear MIMO System

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Abstract

Conventional Proportional Integral (PI) controller is a well known controller used in almost all process Industries for controlling the process parameters at desired set value. The tuning of these controllers is done by a classical Zeigler Nichols (ZN) tuning. As the process tanks are connected in an interacting Multi Input Multi Output (MIMO) mode, there exhibits a highly nonlinear dynamic behavior and time delay between tanks input and output. The Classical ZN tuned PID controller parameters does not cope with all operating points as it exhibits different non linear characteristics. One of the prime applications of fractional calculus is fractional order Proportional Integral Derivative (FOPID) controller. The key challenge in designing fraction order PID controller is to determine the Integral order (α),the derivative order β .This paper proposes the design and modeling of Fractional order PID controller for a nonlinear MIMO spherical tank process. The aim of the paper is to enhance the controller performance of an nonlinear MIMO System by obtaining the optimal set of tuning values at all operating points using Particle Swarm Optimization (PSO) technique and to compare their control performances. The superiority of swarm intelligence in tuning FOPID proves that better enhanced controller performance was obtained for a PSO tuned FOPID controller than that of conventional PID controller in terms of robustness and stability of the closed loop system.

Keywords: PID controller, PSO, Fractional Order Controller, MIMO systems.

Introduction

Proportional Integral Derivative Controller has been using in Industrial control applications for a long time. The reasons for their wide popularity lies in the simplicity of design and good performance which includes low percent over shoot and small settling time for slow process [1,2].According to the survey in 1989, 90 percentage of process industries uses the conventional PID controllers [3].The wide spread use of the PID controller in the Industry is due to their simplicity and ease of retuning online [4].One of the most successful and oldest classical techniques is Zeigler Nichols method. It was put forward by John Ziegler and Nathaniel Nichols [3] in 1942 and is still a simple, fairly effective PID tuning method. He proposed two methods, one is open loop step response method and the second is closed loop frequency response method. The Ziegler and Nichols first PID tuning method is the techniques made based on certain controller assumptions. Hence, there is always a requirement of further tuning; because the controller settings derived are rather aggressive and thus result in excessive overshoot and oscillatory response. Also the controller parameters are rather difficult to estimate in noisy environment. The second method is based on knowledge of the response to specific frequencies. The idea is that the controller settings can be based on the most critical frequency points for stability. This method is based on experimentally determining the point of marginal stability. This frequency can be found by increasing the proportional gain of the controller, until the process becomes marginally stable. These two parameters define one point in the Nyquist plot. The gain is called ultimate gain Ku and the time period Tp. The PID parameter setting is given in [3-6].

Fractional order calculus has gained acceptance in the last couple of decades.J Liouville made the study of fractional calculus in 1832. In 1867 A.K. Grunwaled worked on the fractional operations. G.F.B. Riemann developed the theory of fractional integration in 1892. Fractional order mathematical phenomena allow us to describe and model a real object more accurately than the classical integer methods. Fractional order PID (FOPID), which was first proposed by Podlubny, is the extended version of conventional integer order PID (IOPID) [34]. FOPID possesses unique characteristics of infinite dimensions, memory effects, low sensitiveness to external disturbances, and so forth, when comparing with IOPID. Moreover, abundant dynamics, high robustness and fine tracking accuracy of control systems can be obtained when FOPID is applied [35–37]. One of the challenging problems for practical applications of FOPID is the determination of the controller parameters, which highly influences the stability and tracking performances of servo systems. However, there are no universal methods for optimally determining these parameters due to the complexity of fractional order operations [38-40]. Fractional order PID controllers are used in many control applications [18]. Schlegel Milos et.al [10], performed a comparison between classical controller and fractional controller and summarized that the fractional order controller outperforms the classical controller. D Xue et.al [11],implemented fractional order PID control in DC motor. Chuang Zhao, Xiangde Zhang, et.al [14], implemented fractional order controller in position servo mechanism. Varsha Bhambhani et.al [16] implemented fractional order controller in water level control. Hyo-Sung Ahn et.al [19] implemented fractional order integral derivative controller for temperature profile control. Concepcion A. Monje et.al [20], and Fabrizio Padula et.al [21], devised methods for tuning and auto tuning of fractional order controller for industry level control. Venu Kishore Kadiyala et.al [22] implemented fractional order controller for aerofin control. Fabrizio Padula, Antonio Visioli et.al [23]devised theTuning rules for optimal PID and fractional order PID controllers. A state space design method based on feedback pole placement can be viewed in [14] The control of liquid level is a basic problem in Process Industries such as Petro chemical plants, Chemical recovery of Fertilizer industry etc. It is essential for control system engineers to understand the behavior of non linear Spherical tank level control system and how the level control problems are solved. Several real time works are being carried out to control the level of the spherical tank such as Design of self tuning Fuzzy controllers for non linear systems, by Rajni Jain,N.Sivakumaran,et.al [25]Model based Controller Design for a Spherical tank process by S.Nithya, N.Sivakumaran, etal [26], Soft Computing Based Controllers Implementation for Non Linear Process in Real Time by S.Nithya, N.Sivakumaran et.al[27], Design of Controller using Simulated Annealing for a real time process'by S.M.Girirajkunar , N.Anantharaman et.al[28], Experimental study of Fractional Order Proportional Integral Controller for water level Control carried out by Varsha Bhambani and Y.Q.Chen[29-40], State Feedback with Intergal Controller for a Non-Linear spherical tank Process was done by PradeepKannan.D, S.Sathiyamoorthy [32] etc. Almost all the above works are focused on design of a controller for a nonlinear SISO system. The problem of level control in a nonlinear coupled spherical tank MIMO process is quite cumbersome. PradeepKannan.D, S.Sathiyamoorthy has worked on gain scheduled controller Implementation on a coupled spherical tank process [41]. The PID controllers have been used for several decades in industries for process control applications..The performance of PID controllers can be further improved by appropriate settings of fractional-I and fractional-D actions. This paper attempts to study the behavior of fractional PID controllers. Finding $[\alpha, \beta, kp,\tau_i,\tau_d]$ as an optimal solution to a given process thus calls for optimization on a five dimensional space. The performance of the FOPID controllers is better than its conventional PID both tuned by particle swarm optimization technique. Thus fractional order PID controller can be applied for coupled nonlinear spherical tank process to control the level of the nonlinear tank by manipulating the inflow rate of the process tank by providing optimal values of parameters such as k_p,τ_I, τ_d, α and β. so as to obtain a desired performance characteristics. The proposed design will find extensive application not only in non linear spherical tank process but also for other nonlinear process with simple fine tuning in the values of α and β.

This paper is organized as follows. Section II explains about the mathematical modeling of the non-linear coupled spherical tank MIMO process. Section III describes the Experimental set up. Section IV illustrates the implementation of PSO tuned conventional PID controller and Section V illustrates the implementation of PSO tuned FOPID controller. Section VI describes the Results obtained for servo, regulatory, robustness and closed loop stability of the system operating at various operating points of the tank for both the controllers. Finally the conclusion is given in section VII.

Mathematical Modelling

Consider the coupled Spherical tank process shown in Fig 1. The objective of the process is to control the level of the two tanks namely $h_a(t)$ and $h_b(t)$.

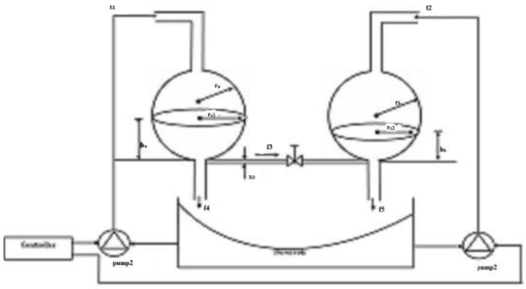

Figure 1: Schematic diagram of Coupled Conical Tank Process

The inlet water flow from the two pumps f_1 (t) and f_2 (t) are used as manipulated variables so as to keep the control variable at the desired set point. The level of the tank at any instant is measured by the combination of orifice and Differential pressure transmitters which are mounted on the respective tanks discharge line whose output is 4 -20mA. This output are compared with the desired set point value of level one and level two which will be configured as 4 -20mA. The error signal is amplified based on the controller specification. The controller outputs is used to vary the inflow rates $f_1(t)$ and $f_2(t)$ of the spherical tank so as to maintain the set point at desired level h_a and h_b of the tanks. Electro pneumatic converters are used to convert the controller outputs of 4 -20mA in to a pneumatic signal of 3-15 psi so that the final control element will be able to throttle the inflow rates f_1 and f_2. Using the law of Conservation of mass, the non linear plant equations are obtained.

For tank 1, $f_1 - f_3 – f_4 = (Ah)\, dh_a/dt$ [1]

For tank 2, $f_3 + f_2 – f_5 = (Ah)\, dh_b/dt$ [2]

Where, f_1 and f_2 are in flow rates to tank_1 and tank_2 respectively in (m^3/sec).
f_4 and f_5 are outflow rate of tank 1 and tank 2 m^3/s.
f_3 flow rate between the tanks m^3/s.

Radius on the surface of the fluid varies according to the surface level of fluid in the tank. Let this radius be as r_s. Therefore $r_{s1}^2 = \surd 2r_ah_a – h_a^2$ and $r_{s2}^2 = \surd 2r_bh_b – h_b^2$

Where r_a= radius of tank_1 in metres.
r_b= radius of tank_2 in metres.
$r_a = r_b = r = 0.5$metres
h_a = fluid level in tank_1 in metres.
h_b = fluid level in tank_2 in metres.
x_0 = thickness of pipe 0.04 in metres.
Also

$f_3 = a \surd 2g(h_a – h_b)$ [3]

$$f_4 = a\sqrt{2g(h_a - x_0)} \qquad [4]$$

$$f_5 = a\sqrt{2g(h_b - x_0)} \qquad [5]$$

Where $a=\pi[x_0/2]^2$

h_{as} - steady state water level of tank 1 = 0.22m

h_{bs} - steady state water level of tank 2 = 0.21m.

The linearised plant transfer function in S domain are given as

$$G_{11}(s) = \frac{1.855s + 0.0881}{s^2 + 0.0794s + 0.0008} \qquad [6]$$

$$G_{22}(s) = \frac{1.919s + 0.0612}{s^2 + 0.0794s + 0.0008} \qquad [7]$$

$$G_{21}(s) = \frac{0.0489}{s^2 + 0.0794s + 0.0008} \qquad [8]$$

$$G_{12}(s) = \frac{0.0492}{s^2 + .0794s + .0008} \qquad [9]$$

A. *Black box modeling*

Consider the first order system with dead time represented by the following transfer function

$$G(s) = \frac{k_p e^{-t_d s}}{\tau_p s + 1} \qquad [10]$$

The output response to a step change in input is given by,

$$y(t) = k_p \Delta u(1 - exp(-(t-\theta)/\tau_p)) \quad t \geq \theta \qquad [11]$$

The measured output is in deviation variable form. The three process parameters K_p, τ_p, θ can be estimated by performing a single step test on process input. The process gain is found as simply the long term change in process output to the change in process input. The time delay is the amount of time, after the input change, before a significant output response is observed.

B. *Estimation of time constant by two point method*

Smith has obtained the parameters of FOPDT transverse function model by letting the response of actual system and that of the model to meet the two points which describe the parameter τ_p and t_d. Here, times required for the process output to make 28.3% and 63.2% respectively are measured. The time constant and time delay can be estimated from the equation given below for both the transfer function,

$$\tau_p = \frac{(2t/3) - (t/3)}{0.7}, and t_d = t/3 - 0.4\tau_p \qquad [12]$$

Thus the obtained FOPDT models are given by $G_{11}(s) = \frac{110e^{-4.492s}}{81.58s + 1}$ [13]

Similarly $$G_{22}(s)=\frac{76.425e^{-8.79s}}{70.72s+1} \qquad [14]$$

The parameters of PID and PI controllers as per Ziegler Nichols tuning method are determined.

C. *Decoupler design*

The effect of interactions in a MIMO system is minimized with the introduction of decoupler. Thus for the coupled non linear interacting spherical tank system, the decoupler are designed as given below.

$$D_{12}(s)=\frac{-0.0492}{1.8550s+0.0881} \qquad [15]$$

$$D_{21}(s)=\frac{-0.0489}{1.919s+0.0612} \qquad [16]$$

D. *Model validation*

The FOPDT model obtained from two point method is given by equations 13 and 14 and the derived transfer function on substituting the dimensions of the spherical tank is given by equations 6 to 9. The open loop response for both the cases for a given step change in set point is obtained as follows.

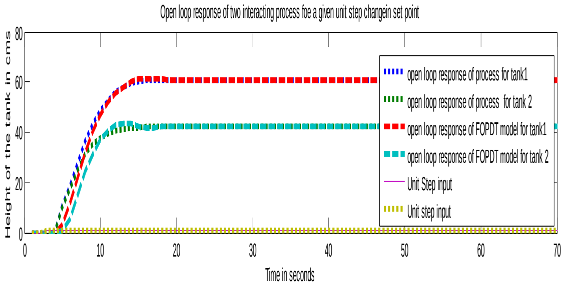

Figure 2: Open loop response of coupled Spherical tank system

It is evident from the figure 2 that both the responses coincides and have similar time domain response. Hence FOPDT model approximation can be used for tuning the PID controller to obtain optimal values of controller parameters.

Experimental Set Up of Coupled Interacting Spherical Tank Process

Figure 3: Experimental set up of coupled spherical tank system.

The experimental set up of the Coupled spherical tank system is shown in figure 3. The spherical tanks are made up of SS316 material which has high corrosion resistance property and has a maximum height, radius of 0.5 meter. Water from the reservoir tank whose size of 1250 mm x 450 mm x 450 mm are pumped through a Kirloskar make 0.5 HP pumps having a discharge of 1500 litres per hour that flows through a Teleline make rotameter having a range 40 - 440 LPH to the spherical tank. The levels of the $tank_1$ and $tank_2$ at any instant are measured by the Rose mount make Differential pressure transmitters having a measurement range of 0-400 mm of water column corresponds to the output range of 4 -20mA.

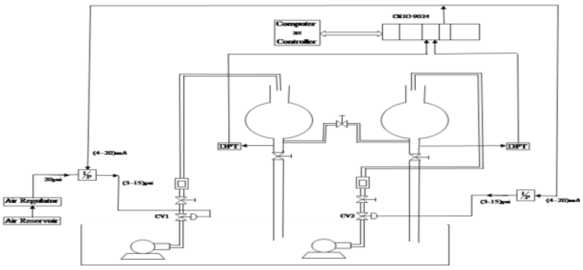

Figure 4: Schematic block diagram of coupled spherical tank system

These outputs are compared with the desired set value of levels in $tank_1$ and $tank_2$ respectively which will be scaled in the control law as 4 -20mA. The error signals are also in the range of 4-20mA which is then amplified based on the controller specifications. These output are used to vary the inflow rate $f_1(t)$ and $f_2(t)$ of the spherical tank with the help of an ABB make Electro pneumatic converters which converts the output of 4 -20mA in to a pneumatic signal of 3-15 psi so that the Equal percentage control valve will be able to throttle the inflow rates $f_1$and f_2 respectively. The controller parameters are determined for both ZN tuned PI Controller as well as FOPI controllers. The coupled spherical tank set up is interface to the real time controller NI Compact RIO 9024which is an 8 module chassis having an inbuilt accuracy of 200ppm.These values are applied to the PID controller block of LabVIEW program for a given step change in inflow rates f_1 and f_2, and the corresponding steady state response are obtained.

PSO Tuned Conventional PID Controller

Particle swarm optimization (PSO) is a population based stochastic optimization technique developed by Dr. Eberhart and Dr. Kennedy in 1995, inspired by social behavior of bird flocking or fish schooling.PSO shares many similarities with evolutionary computation techniques such as Genetic Algorithms (GA). The system is initialized with a population of random solutions and searches for optima by updating generations. In PSO, the potential solutions, called particles, fly through the problem space by following the current optimum particles. Each particle keeps track of its coordinates in the problem space which are associated with the best fitness solution, it has achieved so far is stored. This value is called *pbest*. Another "best" value that is tracked by the particle swarm optimizer is the best value, obtained so far by any particle in the neighbors of the particle. This location is called *lbest*. When a particle takes all the population as its topological neighbors, the best value is a global best and is called *gbest*. The particle swarm optimization concept consists of, at each time step, changing the velocity of each particle toward its p*best* and *lbest* locations. Acceleration is weighted by a random term, with separate random numbers being generated for acceleration toward *pbest* and *lbest* locations. The fitness function evaluates the performance of particles to determine whether the best fitting solution is achieved. During the run, the fitness of the best individual improves over time and typically tends to stagnate towards the end of the run. Ideally, the stagnation of the process coincides with the successful discovery of the global optimum.

E. PSO Algorithm

Let D be the dimension of the search space,

$$x_i = [x_{i1}, x_{i2}, \ldots x_{iD}]^T \quad [17]$$

denote the current position of ith particle of the swarm.

Then, $x_i pbest = [x_{i1} pbest, x_{i2} pbest, \ldots x_{iD} pbest]^T$ [18]

denote the best position ever visited by the particle.

$$Xgbest = [x_{i1} gbest, x_{i2} gbest, \ldots x_{iD} gbest]^T \quad [19]$$

represents ‘gbest’ the best position obtained by any particle in the population.

$$Vi = [v_{i1}, v_{i2}, \ldots v_{iD}]^T \quad [20]$$

represents the velocity of ith particle.

$$Vi_{max} = [v_{i1max}, v_{i2max}, \ldots v_{iDmax}]^T \quad [21]$$

denotes the upper bound on the absolute value of the velocity with which the particle can move at each step.

The position and velocity of the particles is adjusted as per the following equation

$$V_{id} = w * v_{id} + c_1.r_1.(x_{iD}\ pbest - x_{iD}) + c_2.r_2\ (x_{iD}\ gbest - x_{iD}) \qquad [22]$$

where c_1 and c_2 are positive constants, represent the cognitive and social parameter respectively; r_1 and r_2 are random numbers uniformly distributed in the range [0,1], w is inertia weight to balance the global and local search ability.

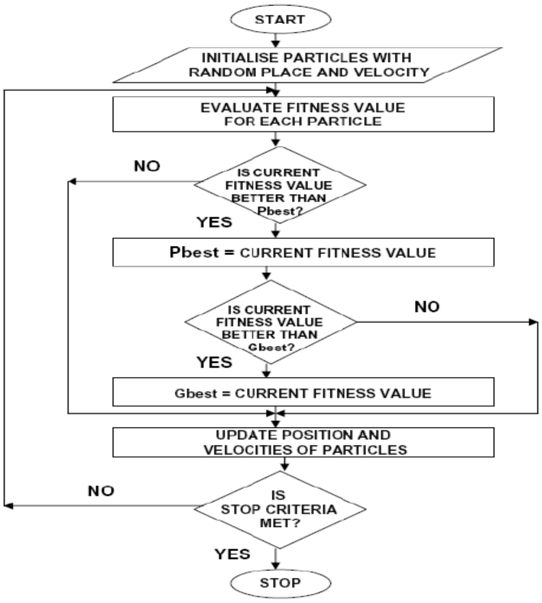

Figure 5: Flow chart for PSO based PID tuning Process

F. *Implementation of PSO tuned PID Controller.*

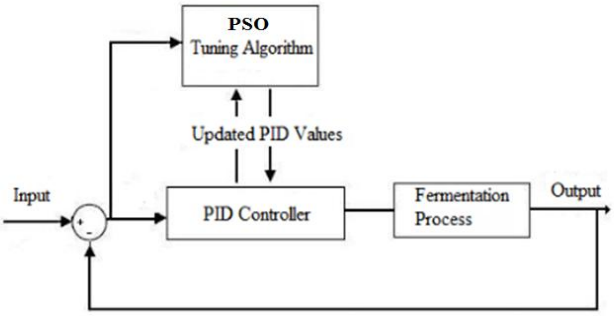

Figure 6: Schematic diagram of PSO based PID tuning Process

The figure 6 shows the block diagram of PSO based PID controller. Optimized PID values are obtained using the Optimization tool available in Matlab. Optimization tool in Matlab is initialized with the parameters such as population size and number of iterations that are randomly chosen appropriately for both the tanks. Further fine tuning the values of population size and number of iteration until the fitness function is achieved provides the optimal response of PID controller for coupled spherical tank process as shown in the responses respectively. The optimum values of kp, ki, kd are found to be 0.6, 4.29,0.1 for tank1 and 1.96, 2.17, 0.1 for tank2 respectively.

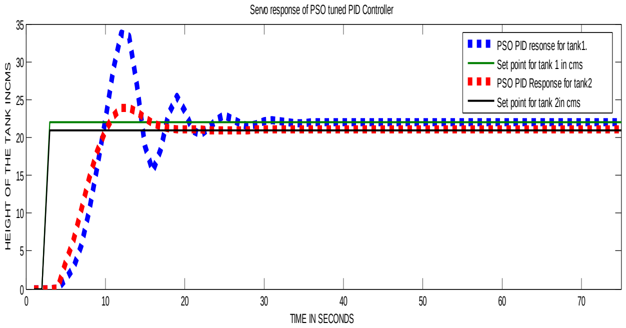

Figure 7: Servo response of PSO Tuned PID Controller.

The figure 7 shows from the response that the PSO tuned PID controller output could track the set point at a different operating point such as 22cm and 21cm of water column for both the tanks with a nominal settling time and with slightly moderate overshoots.

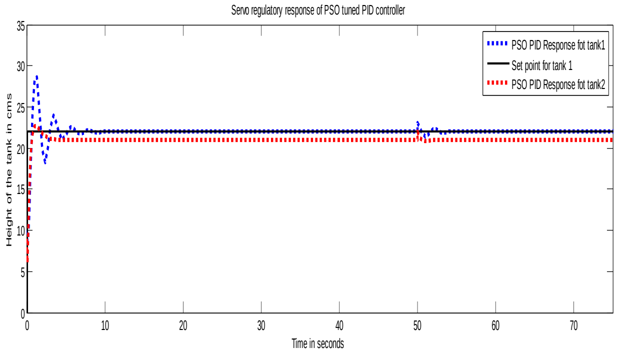

Figure 8: Servo Regulatory response of PSO Tuned PID Controller

The figure 8 shows from the servo regulatory response that the PSO tuned PID controller output could track the set point and rejects the disturbances at an operating point of 22cm and 21cm of water column for both the tanks where a disturbance is applied at 50th second.

PSO Tuned Fractional Order PID Controller

The fractional-order controller will be represented by fractional-order $PI^{\alpha}{}_{D}{}^{\beta}$ controller with transfer function given by the following expression.

$$Gfopi(s) = \frac{u(s)}{e(s)} = K_c\,[\,1 + \frac{1}{\tau_i S^{\alpha}} + \tau_d S\,] \qquad [23]$$

Where α and β are arbitrary real numbers, K_p is controller gain, τ_i is integral time and $\tau_{d,}$ is derivative time. Taking α =1 and β=1,a classical PID controller is obtained. The $PI^{\alpha}{}_{D}{}^{\beta}$ controller is more flexible and gives an opportunity to better adjust the dynamics of control system. It is compact and simple but the analog realization of fractional order system is very difficult and challenging.

The fraction order PI, PID controller can be realized with FAMCON tool box, Ninteger tool box in matlab. Determination of optimal values of $k_p,\tau_{I,}$ $\tau_{d,}$ α and β for a FOPID controller is done by a minimum search PSO algorithm Keeping the objective function as minimization of integral square error. The values of $k_p,\tau_{I,}$ $\tau_{d,}$ α and β FOPID controller G_{c1} are found to be 4,4.0,4.0,0.74,0.2 for tank1 and similarly the values of $k_p,\tau_{I,}$ $\tau_{d,}$ α and β for FOPID Controller G_{c2} are found to be 10.0,1.0,2.0,0.89,.68 respectively.

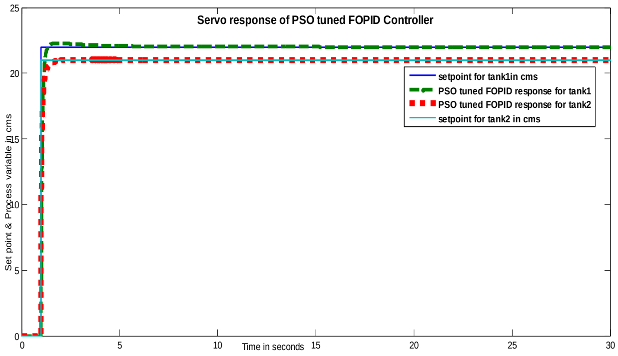

Figure 9: Servo response of PSO Tuned FOPID Controller

The figure 9 shows from the response that the PSO tuned FOPID controller output could track the set point at an operating point such as 22cm and 21cm of water column for both the tanks with a lesser settling time and with very minimal overshoots.

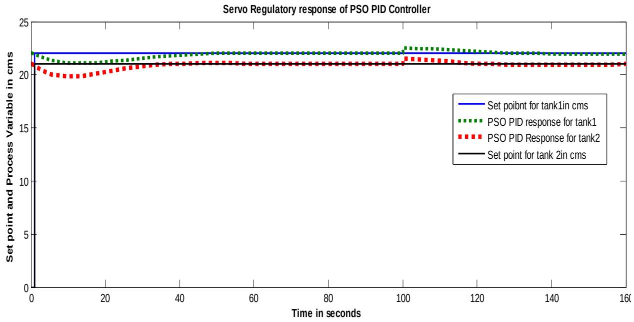

Figure 10: Servo Regulatory response of PSO Tuned PID Controller

The figure 8 shows from the servo regulatory response that the PSO tuned FOPID controller output could track the set point and rejects the disturbances at an operating point of 22cm and 21cm of water column for both the tanks where a disturbance is applied at 400th second.

Table 1

Type of controller	**SP h_a in cm**	**SP h_b in cm**	**Rise time in sec**	**Settling time in sec**	**MP over shoot in %**	**Yss**
PSO PID	22	-	22.9	75.70	7.05	22.01
PSOPID	-	21	16.60	59.20	9.33	21.01
PSO FOPID	22		0.25	3.13	1.28	22.0
PSO FOPID		21	0.39	2.13	0.12	21.0

Table 1 shows the servo response at an operating point of 22and 21 cms height of the tanks with time domain specification of conventional PSO tuned PID controller against PSO tuned FOPID controller.

G. *Robustness Investigation*

Robustness analysis has been done for both the PSO tuned PID controller as well as PSO tuned FOPID Controller. A 10 percent change in process parameters such as k_p t_d and t_i are applied and the corresponding FOPDT model are determined as

$$G_{11}(s)=\frac{121e^{-4.043s}}{89.738s+1} \qquad [24]$$

$$G_{22}(s)=\frac{84.0675e^{-7.911s}}{77.79s+1} \qquad [25]$$

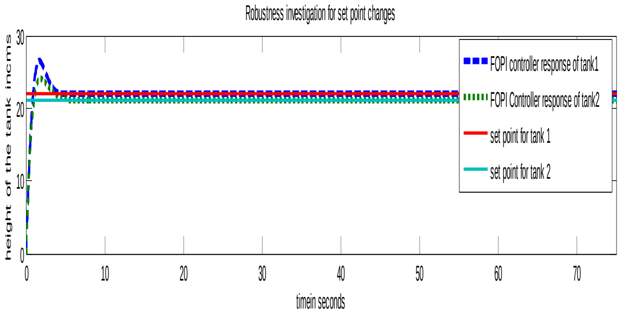

Figure 11: Robust Response of PSO FOPID controller for a set point of,22, 21cm of water column of tank1 and tank2

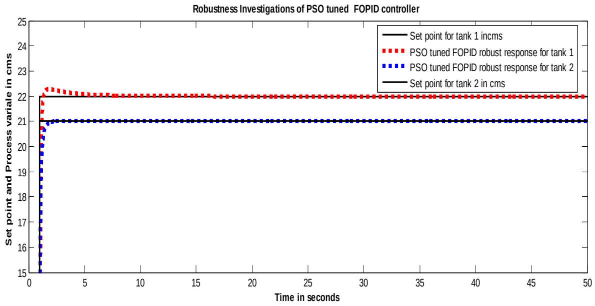

Figure 12: Robust Response of PSO PID controller for a set point of,22, 21cm of water column of tank1 and tank2.

The figures 11 and 12 show that the robustness response of PSO PID controller as well as PSO tuned FOPID controller for a step change in set point of both the tank 1and 2 respectively. From the above graphs, that the PSO tuned FOPID controller is robust enough to track the set point variation in spite of process parameters variation.

H. Stability Investigations:

Stability of the closed loop system has been analyzed for both the PSO PID and PSO FOPID Controller.

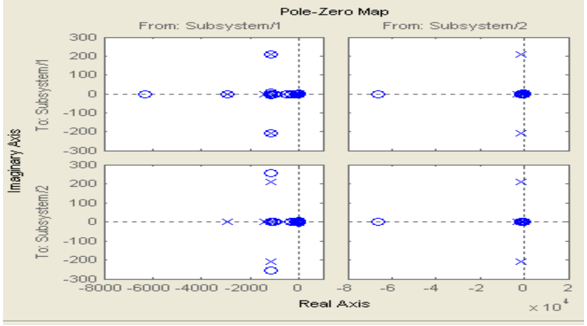

Figure 13: closed loop pole zero map for PSO PID Controller

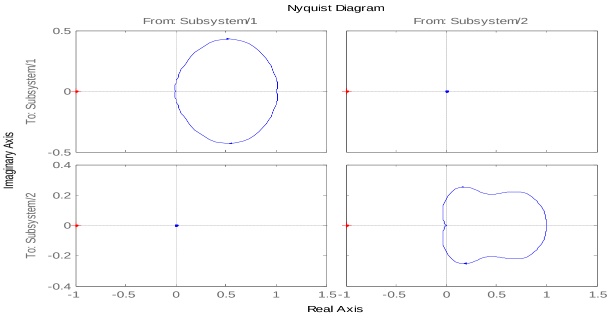

Figure 14: closed loop Nyquist plot for PSO PID Controller

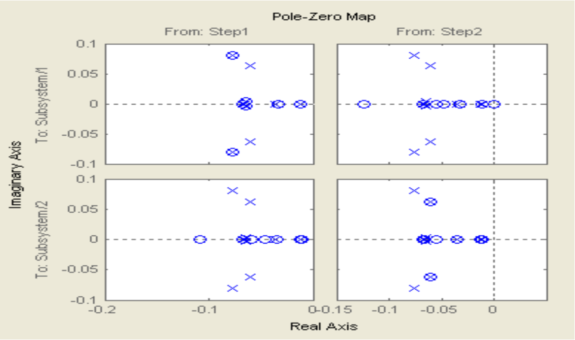

Figure 15: closed loop pole zero map for PSO FOPID Controller

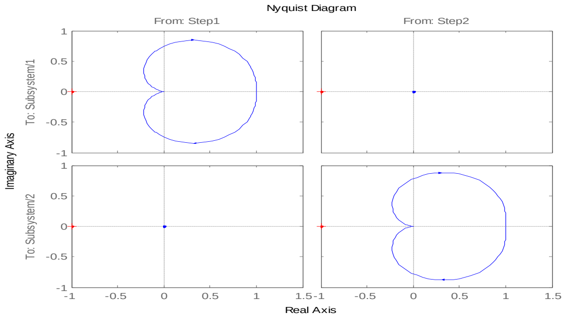

Figure 16: closed loop Nyquist plot for PSO FOPID Controller

It is evident from the above figures 13,14,15 and 16 that the closed loop poles lie on the left half of the S plane and the contour of the Nyquist plots does not include the coordinates -1,j0. Hence the closed loop transfer function of the MIMO System is stable.

Results and Discussion

An step response of PSO tuned PID controller and FOPID controller are obtained for servo and servo regulatory problems and their performance characteristics are compared in terms of time domain specifications. The table 1 shows the time domain specifications for the proposed and conventional controller. It is evident from the table 1 that the PSO tuned PID controller output could track the set point at an operating point of 22cm,21cm of water column for both tank 1 and tank2 with a settling time of 75.7 and 59.2s and with % peak overshoots of 7.05% and 9.33 %. From the results, it is observed that the PSO tuned FOPID controller output could track the set point at the same operating points such as 22cm for tank1 and 21cm of water column for tank2 with settling time of 3.25 and 2.38s. Further there exist a drastic reduction in maximum peak overshoot from 7.05% to 1.28 % for tank1 and 9.33% to 0.117% for tank2.

Conclusion

FOPID controller always cope up with tank non-linear characteristics at a given operating points such as 22, 21cm of water column in the coupled spherical tank. PSO tuned FOPID controller outperforms the response of conventional PID controller with less overshoot, faster settling time and peak time, when subjected to servo and regulatory operations. Further the FOPID controllers proves itself that it is robust enough for changes in process uncertainties and keep tracks the set point and closed loop stability of the system is guaranteed. Tuning FOPID controller can also be applied for other non linear processes such as continuous stirred tank reactor, fermentation process, water treatment process etc to obtain desired controller performance characteristics.

Acknowledgment

This work was supported by department of electronics and Instrumentation engineering and research and development unit of K.L.N. College of Engineering, Pottapalayam , Sivagangai district, Tamilnadu, India .

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