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Direct Synthesis Based Controller Design for Interacting Process Using ETF

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Abstract: A scheme for designing controller for multi-input multi-output systems with associating interactions from other loops using equivalent transfer function (ETF) is presented in this paper. Differing from customary existing techniques for deriving Equivalent transfer function here the Equivalent transfer function is directly assessed by exploiting the relationship between the equivalent closed-loop transfer function and inverse of open-loop transfer function. Based on the derived transfer function centralized and decentralized controllers are composed utilizing the existing tuning rules and simple Direct synthesis control method. Simulation examples are considered to illustrate the better control performance of a system with centralized controllers over decentralized controllers.

Index Terms: Interactions, ETF, centralized and decentralized controller.

I.INTRODUCTION

Due to the demand of high product quality and energy integration, most modern industrial processes take the form of multi-input multi-output systems. The PID control is still widely used method for controlling the MIMO systems because of its adequacy and moderately basic structure contrasted with other cofounded strategies. Be that as it may, because of interactions associated among control loops, MIMO systems are generously harder to control compared to single input single output systems.

Therefore, a lot of research has been made to successfully consider loop interactions into account while deriving equivalent SISO systems from MIMO systems [1]. Numerous control design methods have been outlined, which includes the detuning method [2, 3], sequential loop closing method [4, 5], independent design method [6, 7] ,and so on. Another way to deal with this problem and for utilization of single input single output system Huang et al. proposed equivalent transfer function strategy by building closely related process transfer function for each loop.

In detuning strategy, at first every individual controller is composed using Z-N tuning rules without managing process interactions from different loops. At that point, the interactions are considered by detuning each controller until the point when the moment that multivariable Nyquist stability is satisfied. Yet, burden of this strategy is that the controller settings are impervious to change.

In Sequential loop closing system, every controller is tuned successively. At first the controller of speediest loop is tuned by considering a picked input-output combination; this loop is then shut and after that the controller of the slowest loop is tuned for a next match while the main loop staying shut and so on. In independent loop technique, every controller is independently composed based closed and open loop transfer functions, fulfilling the disparity constraints on process interactions. A noteworthy inconvenience of independent loop technique is that it doesn't exploit the data with respect to controllers in different loops. The potential bother engaged with above procedures is that controllers of each loop communicate with different loops, the execution of one loop can't be accessed without data of controllers in different loop.

To solve this issue, a few approaches by developing equivalent individual loop have been proposed. The successful open loop transfer function is formulated without earlier data of controller progression of different loops. Additionally, the compelling open-loop transfer function work is inferred to separate multi-loop control system into an arrangement of proportionate independent single loop systems, at this point decentralized controllers are composed independently in view of the comparable open-loop transfer function model. The ETF is acquired by increasing original transfer function by multiplicative model factor inside the neighborhood of individual control loop critical frequency.

In this paper, a novel unified PI controller outline strategy for multivariable processes is proposed. By utilization of the relationship in between the inverse of open-loop transfer function and equivalent closed loop transfer function the articulation of ETF model is made. On the basis of designed ETF, multi-loop PI controllers are composed utilizing DS control technique. Especially, this ETF strategy is more advantageous for higher dimensional processes with entangled interactions nodes compared with other already available methodologies. The proposed technique is checked utilizing cases deliberately to exhibit the suitability and viability of proposed strategy.

II. EQUIVALENT TRANSFER FUNCTION

Consider an n * n open-loop stable multivariable system as shown in Fig. 1, where r_i, u_i, e_i and y_i , i □ 1, 2,..., n are the reference inputs, manipulated variables, error and outputs of system, respectively; $G_c(s)$ is the controller to be found and $G(s)$ is the process transfer function matrix, both of which are given by

$$G(s) = \begin{bmatrix} g_{11}(s) & g_{12}(s) & \cdots & g_{1n}(s) \\ g_{21}(s) & g_{22}(s) & \cdots & g_{2n}(s) \\ \vdots & \vdots & \ddots & \vdots \\ g_{n1}(s) & g_{n2}(s) & \cdots & g_{nn}(s) \end{bmatrix}, \quad (1)$$

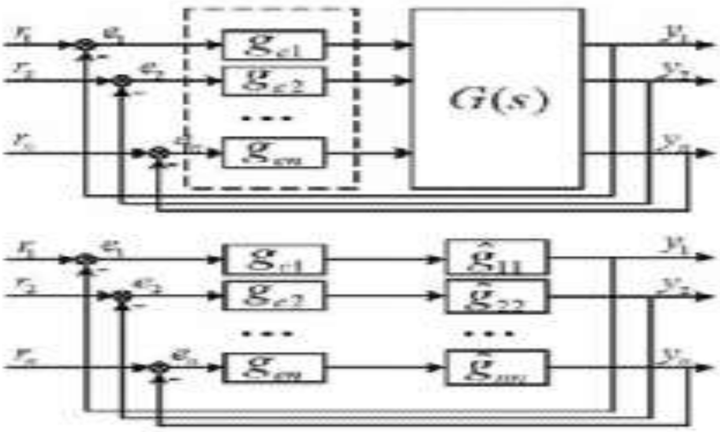

Fig. 1. Multi-loop control system and independent SISO system

As portrayed in Fig. 1, designing a multi-loop controller is breakdown to design of controller for equivalent single input and single output system. When entire system is closed, there are interactions among control loops due to the presence of elements other than diagonal elements in the transfer function. Considering the influence of other loops, the equivalent transfer functions can be compactly given in terms of Dynamic relative gain array (DRGA) , as follows[11]

$$\hat{G}(s) = [G(s)]^{-T} \quad (2)$$

And $\hat{G}(s)$ is

$$\hat{G}(s) = \begin{bmatrix} \frac{1}{\hat{g}_{11}(s)} & \frac{1}{\hat{g}_{12}(s)} & \cdots & \frac{1}{\hat{g}_{1n}(s)} \\ \frac{1}{\hat{g}_{21}(s)} & \frac{1}{\hat{g}_{22}(s)} & \cdots & \frac{1}{\hat{g}_{2n}(s)} \\ \vdots & \vdots & \ddots & \vdots \\ \frac{1}{\hat{g}_{n1}(s)} & \frac{1}{\hat{g}_{n2}(s)} & \cdots & \frac{1}{\hat{g}_{nn}(s)} \end{bmatrix}. \quad (3)$$

Note that the Equivalent transfer function (ETF) of each loop comprise of process dynamics terms only and they do not have knowledge of controller parameters under the assumption of perfect control estimate.

DERIVATION OF ETF:

As seen from equation (2), we got fundamental relationship between ETF and $G^{-T}(s)$. The equivalent transfer function can be derived as follows by utilizing the steady and transient information of given processes,

Considering the first order plus time delay model.

$$g(s) = \frac{ke^{-\theta s}}{\tau s+1} \quad (4)$$

From (2) and (4) , The equivalent transfer function for each loop is derived as

$$\frac{\tau s+1}{k} e^{-\theta i*s} = \frac{adj\ G_{ii}}{|G|} \quad (5)$$

Where $|G|$the determinant of G, and adj is$G_{ii}(s)$ is the adjugate matrix corresponding to $G_{ii}(s)$

Taking double derivative on both sides of equation (5) and letting s = 0 gives the parameters of the first order plus time delay model and they are given as:

$$k_i = \frac{1}{a_{ii}}$$

$$\tau_i = \sqrt{(\frac{b_{ii}}{a_{ii}})^2 - (\frac{c_{ii}}{a_{ii}})}$$

$\theta_{ii}=\frac{b_{ii}}{a_{ii}}-\sqrt{(\frac{b_{ii}}{a_{ii}})^2-(\frac{c_{ii}}{a_{ii}})}=\frac{b_{ii}}{a_{ii}}-\tau_i$

$a_{ii}=\frac{adj\ k_{ii}}{|k|}$ at s=0 (6)

$b_{ii}=-\frac{1}{|k|^2}\sum_{p=1}^{n}[\sum_{q=1}^{n}(adj\ K_{iq})\grave{g}_{qp}]\ adj\ k_{pi}$ at s=0 (7)

(7)

$c_{ii}=\frac{1}{|k|^3}\sum_{m=1}^{n}[\sum_{l=1}^{n}[\sum_{p=1}^{n}[\sum_{q=1}^{n}(adj\ K_{iq})\grave{g}_{qp}]]\grave{g}_{lm}]\ adj\ K_{mi}-\frac{1}{|k|^2}\sum_{p=1}^{n}[\sum_{q=1}^{n}(adj\ K_{iq})\ddot{g}_{qp}]\ adj\ k_{pi}$ at s=0 (8)

In order for the resulting FOPTD ETF model to be realizable, τ_i and θ_i should be real and positive.
Therefore from equations (6) – (8) following conditions must be satisfied

$\left(\frac{b_{ii}}{a_{ii}}\right)^2>\frac{c_{ii}}{a_{ii}}>0$ (9)

III. DESIGN OF DS CONTROLLER FOR MULTIVARIABLE PROCESS

As equivalent transfer functions now contains information of loop interactions, the complex multivariable process can be down converted to set of single input single output (SISO) process and then controllers can be designed using Direct synthesis method [15] to stabilize these SISO loops independently.

Direct synthesis method is based on a desired form system response and then finding controller strategy and parameters to give that response.

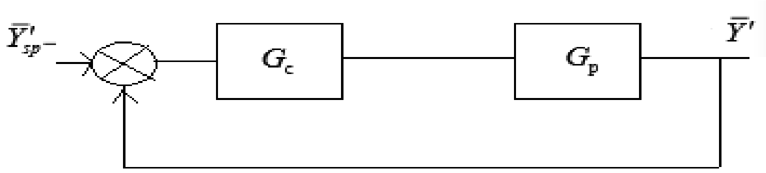

Fig.2: Block diagram for unity feedback system

A unity feedback system is illustrated in Fig .2
For the feedback control above the closed loop transfer functions between the output and set point are

$\frac{\grave{Y}}{Ysp}=\frac{G_c*G_p}{1+G_c*G_p}$ (10)

Where G_c*G_p the loop gain one. A more viable response would be:

$\frac{\grave{Y}}{Ysp}=\frac{e^{-ds}}{\lambda s+1}$ (11)

Where λ and d are adjustable parameters to satisfy performance and implementation requirement, respectively.
By substituting Eq. (11) in (10) we get

$G_c=\left[\frac{\grave{Y}}{Ysp}\right]*C$ (12)

Where $C=\frac{1}{1-\frac{e^{-ds}}{\lambda s+1}}$ (13)

Which can be implemented as by the unit shown in Fig.3

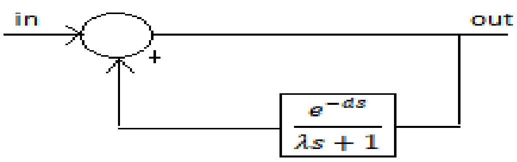

Fig.3. Feedback control unit

to convert the DS controller to PI controller we have to expand the G_c in Maclaurine series as follows

G_c =f(s)/s=$\frac{1}{s}[f(0)+\grave{f}(0)s+hod\ terms]$ (14)

Expanding G_c in frequency domain and neglecting higher order terms gives

$g_c=k_c(1+(1/\tau_{Ii}s))$ (15)

Comparing (15) and (16) we get controller parameters as

$k_c=\grave{f}(0)$. (16)

$\tau_{Ii}=\frac{f(0)}{\grave{f}(0)}$. (17)

If the equivalent transfer function has a strong lead term, the derivative and integral constants calculated using (16, 17) may have negative values. Under the said conditions, a controller with first order lag filter structure is recommended.

IV. CONTROL PERFORMANCE AND ROBUST STABILITY

Generally for detailed analysis and comparison with other existing methods integral absolute error and integral time absolute error criterions are considered.
IAE criterion is considered for evaluation of closed loop systems, which is defined as:

$$IAE=\int_0^{\infty}|e(t)|\,dt \qquad (18)$$

Where e (t) is r (t)-y (t)
Secondly, ITAE criterion to evaluate performance over long period of time.

$$ITAE = \int_0^{\infty}|e(t)|t\,dt \qquad (19)$$

At last, to investigate the robustness [16, 17] of came about control system, a robust technique is utilized for a reasonable examination with other existing strategies. The robustness can be researched under output multiplicative uncertainty as it is regularly less prohibitive than input uncertainty regarding execution. For a framework with yield vulnerability as [I+Δ(s)] G(s), the closed loop is steady if

$$\Upsilon < 1/\max(\text{sigma}(Y(s)/U(s))) \qquad (20)$$

ϒ represents degree of robust stability which is double norm of $\Delta(jw)$, where $\Delta(jw)$represents multiplicative output uncertainty. For a reasonable correlation ϒ esteem ought to be the same as or bigger than that of other reenactment strategies.

V. Simulation studies

Example 2. The distillation column of ISP reactor was quoted to verify performance of proposed equivalent transfer function method.

$$G(s)=\begin{bmatrix}\frac{22.89*e^{-0.2s}}{4.572s+1} & \frac{-11.64*e^{-0.4s}}{1.807s+1}\\ \frac{4.689*e^{-0.2s}}{2.174s+1} & \frac{5.8*e^{-0.4s}}{1.801s+1}\end{bmatrix}.$$

The SOPTD model with a lead term is considered.
We get equivalent transfer functions as below

$$g11eff = \frac{32.3*(2.878*s+1)e^{-0.2s}}{9.972s^2+6.753*s+1}.,\quad g12eff = -\frac{39.9535*(2.8757*s+1)e^{-0.4s}}{8.2562s^2+6.3779s+1}.$$

$$g21eff = \frac{16.09437*(2.8755*s+1)e^{-0.2s}}{9.9313s^2+6.744s+1}.,\quad g22eff = \frac{8.18*(2.876*s+1)e^{-0.4s}}{3.925s^2+3.98s+1}.$$

Loop pairing technique for decentralized controller is 1-1/2-2 [8].Robustness index ϒ is 0.59.

Implementation of Decentralized Control system:

The decentralized two input two output system can be implemented as follows.

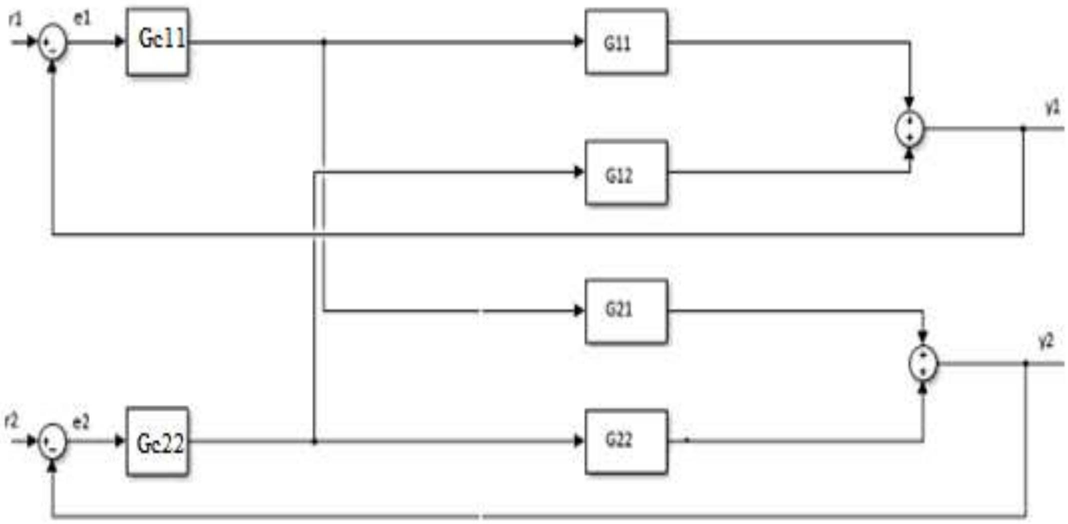

Fig.4. Implementation of Decentralized controller for TITO process

$$G_{eff}(s) = \begin{bmatrix}\frac{32.3*(2.878s+1)e^{-0.2s}}{9.972s^2+6.753*s+1} & 0\\ 0 & \frac{8.18*(2.876s+1)e^{-0.4s}}{3.925s^2+3.98s+1}\end{bmatrix}$$

Direct synthesis based decentralized controller are derived using equations (12) and (13) are as follows:

$G_{c11} = \left[\frac{(9.972s^2+6.753s+1)(1-0.2s)}{s(15.80s^2+98.47s+32.3003)}\right]$. , $G_{c22} = \left[\frac{(3.925s^2+3.98s+1)(1-0.2s)}{s(15.80s^2+98.47s+32.3003)}\right]$.

Implementation of centralized Control system:

The centralized two input two output system can be implemented as shown in Fig.5.

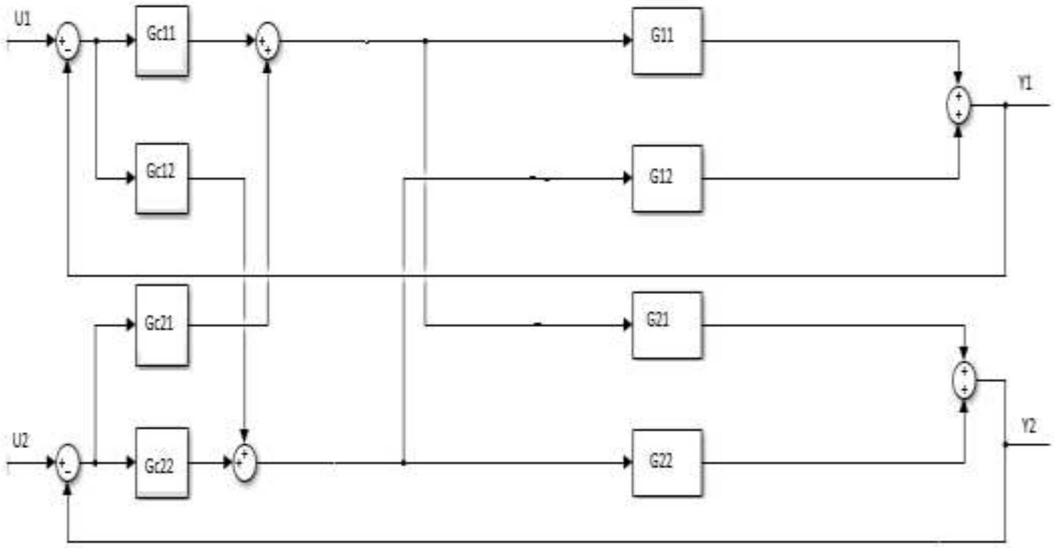

Fig.5. Implementation of centralized controller for TITO process

$G_{eff}(s) =$

$$\begin{bmatrix} \frac{32.3*(2.878s+1)e^{-0.2s}}{9.972s^2+6.753*s+1} & -\frac{39.953*(2.8757s+1)e^{-0.4s}}{8.2562s^2+6.3779s+1} \\ \frac{16.0947*(2.875s+1)e^{-0.2s}}{9.9313s^2+6.744s+1} & \frac{8.18*(2.876s+1)e^{-0.4s}}{3.925s^2+3.98s+1} \end{bmatrix}.$$

Direct synthesis based centralized controllers are derived using equations (12) and (13) as follows

Gc=

$$\begin{bmatrix} \frac{0.031*(9.972s^2+6.753s+1)e^{-0.2s}*c1}{2.446s^2+3.728s+1} & \frac{0.062*(9.9313s^2+6.744s+1)e^{-0.2s}*c2}{3.45s^2+4.075s+1} \\ \frac{-0.025(8.2562s^2+6.3779s+1)*c1}{2.44*s^2+3.725s+1} & \frac{0.122*(3.925s^2+3.98s+1)*c2}{3.45*s^2+4.076s+1} \end{bmatrix}$$

Where c1 = $\frac{1}{1-\frac{e^{-0.4s}}{0.85s+1}}$, c2=$\frac{1}{1-\frac{e^{-0.4s}}{1.2s+1}}$.

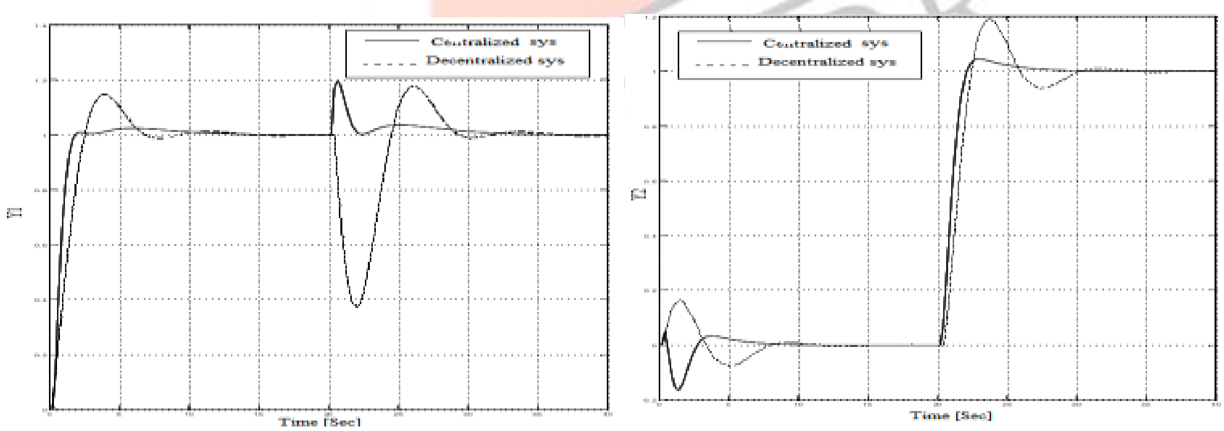

Fig.6. Outputs of ISP reactor for loop 1 and loop2 with decentralized controller and centralized controllers.

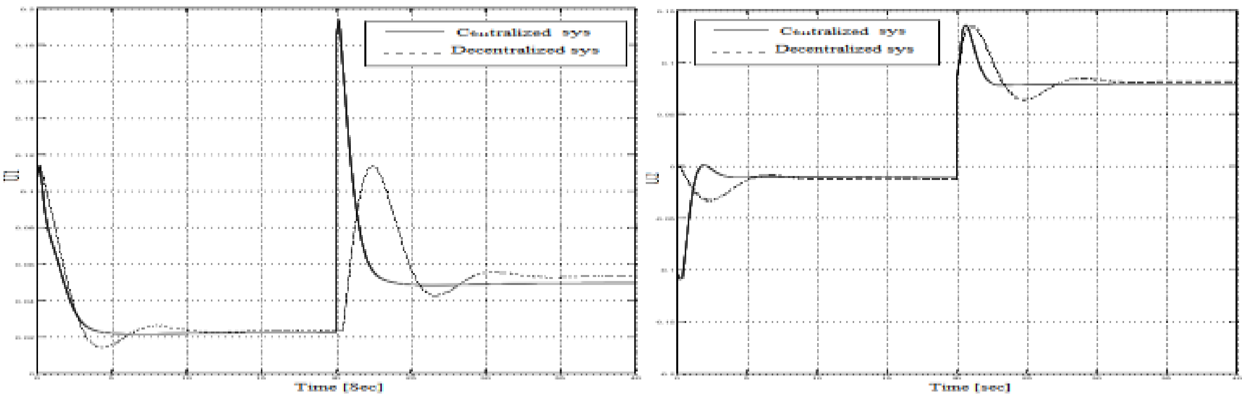

Fig.7. Controller outputs of ISP reactor for loop1 and loop2 with decentralized and centralized controllers

Tuning Method	Loop	K_{ci}	τ_{li}	IAE		ITAE	
				Nominal	ISP (+50%)	Nominal	ISP (+50%)
Decentrali-zed control	1	0.108	0.268	1.353	1.422	11.3	13.01
	2	0.096	0.906	25.08	1.478	1.443	23.04

Table 1. Controller parameters for decentralized controllers and performance index for ISP reactor with robustness index λ_i =0.85. +50 % indicate plant model uncertainty.

Tuning Method	Loop	J=1		J=2		IAE		ITAE	
		K_{ci}	τ_{li}	K_{ci}	τ_{li}	Nominal	ISP (+50%)	Nominal	ISP (+50%)
Centralized control	1	0.108	0.268	0.162	0.271	3.748	2.457	52.55	29.4
	2	-0.16	0.125	0.096	0.906	2.522	1.878	45.82	34.09

Table 2. Controller parameters for decentralized controllers and performance index for ISP reactor λ_{11} *is* 0.85, λ_{22} *is* 1.2 , +50 % indicate plant model uncertainty

Example2: Consider a 3*3 transfer function matrix [14].

$$G(s)=\begin{bmatrix} \frac{e^{-0.1s}}{s^2+2s+1} & \frac{0.3e^{-0.1s}}{2s^2+3s+1} & \frac{0.3e^{-0.1s}}{2s^2+3s+1} \\ \frac{0.3e^{-0.1s}}{2s^2+3s+1} & \frac{e^{-0.1s}}{s^2+2s+1} & \frac{0.3e^{-0.1s}}{2s^2+3s+1} \\ \frac{0.3e^{-0.1s}}{2s^2+3s+1} & \frac{0.3e^{-0.1s}}{2s^2+3s+1} & \frac{e^{-0.1s}}{s^2+2s+1} \end{bmatrix}$$

We get equivalent transfer functions as below

$g_{ij(i=j),eff} = \frac{0.86153*e^{-0.582s}}{1.258*s+1}$. , $g_{ij(i\neq j),eff} = \frac{-3.7335*(2.227*s+1)e^{-0.576s}}{2.155s^2+2.792s+1}$

Where i=1, 2, 3 j=1, 2, 3 ,Loop pairing is 1-1/2-2 [8],Robustness index ϒ is 0.85.

Implementation of Decentralized Control system:

$g_{c,ij(i=j)} = \left[\frac{(1.258s+1)}{s*0.8615*(1.08)}\right]$

Direct synthesis based decentralized controller
are derived using equations (12) and (13) are as follows:

$g_{c11} = gc22 = gc33 = \left[\frac{(1.258s+1)}{s*0.8615*(1.08)}\right]$.

The Decentralized three input three output System can be implemented as follows

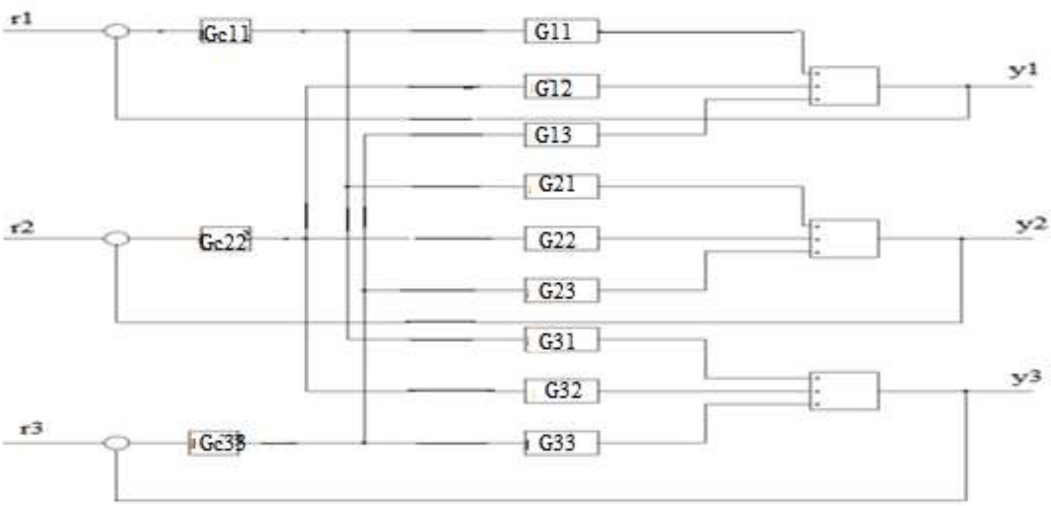

Fig: 8.Implementation of Decentralized controller for Three input three output process

Implementation of centralized Control system: The centralized two input two output systemCan be implemented as shown in Fig.9

$g_{c,ij(i=j)} = \left[\frac{(1.258s+1)}{s*0.8615*(1.08)}\right]$,$g_{c,ij(i\neq j)} = \left[\frac{(2.155s^2+2.792*s+1)e^{-0.526s}}{-3.733s(2.227s+1)(1.08s)}\right]$

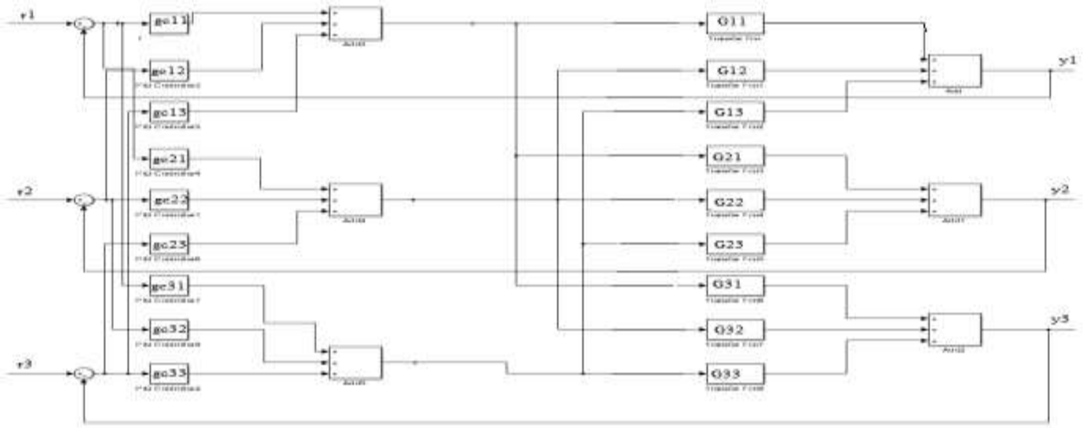

Fig.9: Implementation of centralized controller for three input three output process.

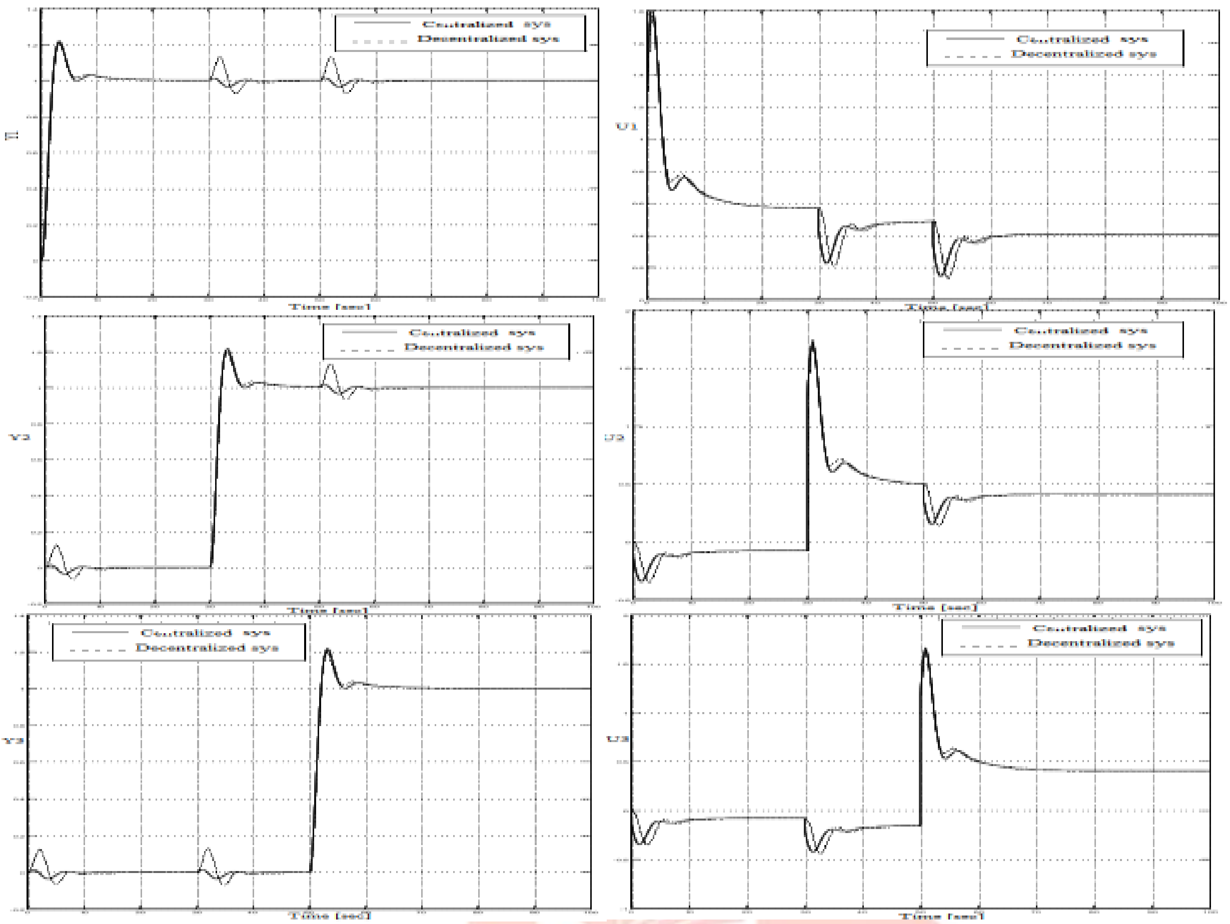

Fig.11: Output Y2, Y3 and their corresponding controller output for centralized and decentralized system

Tuning Method	Loop	K_{ci}	τ_{Ii}	IAE		ITAE	
				Nominal	+30%	Nominal	+30%
Decentralized	1	1.34	0.798	2.765	2.579	44.73	44.34
	2	1.34	0.798	2.685	2.519	84.52	79.04
	3	1.34	0.798	2.702	2.511	112.8	102

Table 3. Controller parameters for decentralized controllers and performance index with $\lambda_i = 0.5$ for example 2

Tuning Method	Loop	J=1		J=2		J=3		IAE		ITAE	
		K_{cij}	τ_{Ij}	K_{cij}	τ_{Ij}	K_{cij}	τ_{Ij}	Nominal	+30%	Nominal	+30%
Centralized	1	1.34	0.798	-0.140	1.77	-0.140	1.77	2.114	1.995	16.08	15.85
	2	-0.140	1.77	1.34	0.798	-0.140	1.77	2.081	1.978	66.93	63.25
	3	-0.14	1.77	-0.140	1.77	1.34	0.798	2.102	1.976	102.5	95.26

Table 4. Controller parameters for centralized controllers and performance index with $\lambda_{11} = \lambda_{22} = 0.5$ for example 2, +30 % indicate plant model uncertainty.

It is clear from the response of example 1 and example 2 that interactions of one loop over other loops can be minimized by using centralized controllers than decentralized controllers.

VI. Conclusion

A ETF based controller design has been developed in this project. The analytical expression of equivalent transfer function is derived based on relationship between closed-loop and inverse of open-loop transfer function. Simulation results for typical industrial processes with decentralized and centralized controllers shows that the centralized controllers provides better or compatible performance compared to decentralized controller. A conceivable heading for future work is the expansion of the ETF technique to non – square multivariable systems.

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