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Study of Different Controller's Performance for a Real Time Non-Linear System

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Abstract:

The remarkable growth in the control mechanisms has been evidently seen in the last two decades. The controller Design has always been an important concern. In this paper we have chosen a real time Single Spherical Tank Liquid Level System (SSTLLS) for our investigation. The real time system is chosen to model the non-linear spherical system. This paper deals with the modeling of identified system in Simulink. System identification of this nonlinear process is done using black box model, which is identified to be nonlinear and approximated to be a First Order Plus Dead Time (FOPDT) model. A proportional and integral controller is designed in Simulink and various tuning methods including, Skogestad's, Ziegler Nicolas(ZN) , Cheng and Hung(CH), and SIMC PID(SPD) are implemented. The paper will provide details about the implementation of the controller, and compare the results of PI tuning methods used.

Keywords: Single Spherical Tank Liquid Level System (SSTLLS), PI Controller, Simulink, LabVIEW.

1. Introduction

In common terms, most of the industries have typical problems raised because of the dynamic non linear behavior. It is only because of the inherent non linearity, most of the chemical process industries, Hydrometallurgical industries, food process industries are in need of classical control techniques. The evidence of mushrooming interest in the nonlinear models and their controlling strategies [1, 2], which in turn described about the process dynamics around a larger operating region than the corresponding linear models have been gaining great popularity [3].The non linear models are obtained from first principles and further from the parameters which appear within such models that are procured from the data of the process. However , the search for the conventional methods is still on. Once the model has been developed, a controller design maintain the process under steady state. Proportional Integral Derivative (PID) controller is the name that is widely heard as a part of process control industry. Despite much advancement in control theory which has been recently seen, PID controllers are still extensively used in the process industry. Conventional PID controllers are simple, inexpensive in cost [4], easy to design and robust provided the system is linear. The PID controller operates with three parameters, which can be easily tuned by trial and error ,or by using different tuning strategies and rules available in literature such as Ziegler-Nichols [5], Z Cheng and Atherton [6],Sung et al. [7].These rules have their bases laid on open-loop stable first or second–order plus dead time process models. There are many other methods and approaches which have periodically evolved to improvise the performance of PID tuning, For instance the Aström–Hägglund phase margin method [8], the refined Ziegler-Nichols method by Cohen and Coon [9] as well as Hang et al. [10], the internal model control (IMC) design method [11,12],gain and phase margin design methods [13,14], and so on. The software and technology have been assisting the mankind by offering a potential to design and implement more sophisticated control algorithms. Despite all the effort, industries emphasize more on robust and transparent process control structure that uses simple

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controllers which makes PID controller the most widely implemented controller.

This paper endeavors to design a controller based on Multi Model Approach which focuses on dividing the complete nonlinear range into several linear parts which can be controlled and tuned separately using already defined classical methods. The transfer functions of various linear regions will be, then cascaded to give the overall transfer function of the nonlinear tank system. There on we implement various tuning techniques to design the PI controller so as to control the level parameter of the SSTLLS.

2. Experimental Process Description

The laboratory set up for this system basically comprises two spherical interacting tanks which are connected with a manually operable valve between them. Both the tanks have an inflow and outflow of water which is being pumped by the motor, which continuously sucks in the water from the water reservoir. The flow is regulated in to the tanks through the pneumatic control valves, whose position can be controlled by applying air to them. We employ a compressor so as to apply pressure to close and open the pneumatic valves. There is also provision given to manually measure the flow rate in both the tanks using rotameter. The level in the tanks are being measured by a differential pressure transmitter which has a typical output current range of 4mA-20mA.This differential pressure transmitter is interfaced to the computer connected to through the NI-DAQmx 6211 data acquisition module card which can support 16 analog inputs and 2 analog output channels with a voltage ranging between ±10 Volts. The sampling rate of the acquisition card module is 250Ks/S with 16 bit resolution. The graphical program written in LabVIEW is then linked to the set up through the acquisition module. Figure 1 shows the real time experimental setup of the process.

The process of operation starts when pneumatic control valve is closed by applying the air to adjust the flow of water pumped to the tank. This paper talks only about a single spherical tank liquid level system (SSTLLS), so we shall use only the Spherical Tank one for our usage throughout the experiment. The level of the water in tank is measured by the differential pressure transmitter and is transmitted in the form of current range of 4mA-20mA to the interfacing NI-DAQmx 6211 data acquisition module card to the personal computer (PC).After computing the control algorithm in the PC, control signal is transmitted to the I/ P converter which passes the pressure to the pneumatic valve proportional to the current provided to it. The pneumatic valve is actuated by the signal provided by I/P converter which in turn regulates the flow of water in to the tank. Figure 2 shows the interfaced NI-DAQmx 6211 data acquisition module card. Table 1 shows the technical specifications of the interacting two tank spherical tank liquid level system setup.

Figure 1: Real time experimental set up of the process.

Figure 2: Interfaced NI-DAQmx 6211 Data Acquisition Module Card.

3 .System identification and controller design

3.1. Mathematical Modeling of SSTLLS

The SSTLLS is a system with non linear nature in virtue of its varying diameter. The dynamics of non linearity for this system can be described by the first order differential equation.

$$\frac{dV}{dt} = q_1 - q_2 \quad (1)$$

Where,

V is the volume of the tank

q_1 is the Inlet flow rate and

q_2 is the Outlet flow rate.

The volume V of the spherical tank is given by,

$$V=\frac{4}{3}\pi h^3 \qquad (2)$$

Where h is the height of the tank in cms.

On application of the steady state values, and by solving the equations 1 and 2,the non linear spherical tank can be linearized by the following model,

$$\frac{H(S)}{Q1(S)}=\frac{Rt}{\tau s+1} \qquad (3)$$

Where, $\tau = 4\pi R_t h_S$ and $Rt = \frac{2hs}{Q2s}$

The system identification of SSTLLS is derived using the black box modeling. Under constant inflow and constant outflow rates of water, the tank is allowed to fill from (0-45) cm. Each sample is acquired by NI-DAQmx 6211 from the differential pressure transmitter through USB port in the range of (4-20) mA and the data is transferred to the Personal computer. This data is further scaled in terms of level (in cm).employing the open loop method, for a given change in the input variable; the output response of the system is recorded. Ziegler and Nichols [5] have obtained the time constant and time delay of a FOPTD model by constructing a tangent to the experimental open loop step response at its point of inflection. The intersection of the tangent with the time axis provides the estimate of time delay. The time constant is estimated by calculating the tangent intersection with the steady state output value divided by the model gain.

Cheng and Hung[15] have also proposed tangent and point of inflection methods for estimating FOPTD model parameters. The major disadvantage of all these methods is the difficulty in locating the point of inflection in practice and may not be accurate. Prabhu and Chidambaram [16] have obtained the parameters of the first order plus time delay model from the reaction curve obtained by solving the nonlinear differential equations model of a distillation column. Sundaresan and Krishnaswamy [17] have obtained the parameters of FOPTD transfer function model by collecting the open loop input-output response of the process and that of the model to meet at two points which describe the two parameters τ_p and θ. The proposed times t_1 and t_2, are estimated from a step response curve. The proposed times t1 and t2 , are estimated from a step response curve. This time corresponds to the 35.3% and 85.3% response times.

The time constant and time delay are calculated as follows.

$$\tau_p = 0.67(t_2 - t_1) \qquad (4)$$

$$\theta = 1.3t_1 - 0.29t_2 \qquad (5)$$

At a constant inlet and outlet flow rates, the system reaches the steady state. After that a step increment to the system is given by changing the flow rate and various values of the same are taken and recorded till the system becomes stable again. The experimental data are approximated to be a FOPDT model. The model parameters are designed for five different regions of operation in SSTLLS, The conventional FOPDT model is given by

$$G(s)=\frac{K.e^{-\theta s}}{\tau s+1} \qquad (7)$$

Using the FOPDT model, the transfer functions for various regions are determined. Table 2 shows the transfer functions for different regions.

Table 1: Technical Specifications of the Experimental Setup

PART NAME	DETAILS
Spherical Tank	Material: Stainless Steel Diameter:45 cm
Storage Tank	Material: Stainless Steel Volume:100 liters
Differential Pressure Transmitter	Type: Capacitance Range:(2.5 to 250)mBAR Output:(4 to 20)mA Make: ABB
Pump	Centrifugal 0.5 HP
Control Valve	Size: 1/4”,Pnematic actuated Type: Air to close Input(3-15)PSI 0.2-1 Kg/cm2
Rotometer	Range:(0-440)LPH
Air Regulator	Size 1/4” BSP Range:(0-2.2)BAR
I/P Converter	Input:4-20 mA Output: (3-15) PSI
Pressure Gauge	Range:(0-30) PSI Range:(0-100) PSI

Table 2:Transfer functions for different regions.

Height	Transfer Function
0-9	$G(s) = \frac{9 * e^{(-88.88*s)}}{1 + 91.12 * s}$
9-18	$G(s) = \frac{18 * e^{(-440.995*s)}}{1 + 142.04 * s}$
18-27	$G(s) = \frac{11.25 * e^{(-896.835*s)}}{1 + 122.61 * s}$
27-36	$G(s) = \frac{10 * e^{(-1224.16*s)}}{1 + 73.365 * s}$
36-45	$G(s) = \frac{11.25 * e^{(-1404.51*s)}}{1 + 27.805 * s}$

3.2. Design of PI Controller

The derivation of transfer function model will now pave the way to the controller design which shall be used to maintain the system to the optimal set point. This can be only obtained by properly selecting the tuning parameters K_p and τ_i for a PI controller. All the tuning methods are implemented for different regions. The controller is designed in Simulink and the time domain analysis is done for those tuning methods.

4. Results and discussions

Table 3 to 7 shows the time domain comparison and performance indices for each zone to find out the best suitable method for that particular zone.

4.1 Region 1 (0-9) cm:

For the first region, i.e. the level range from 0 to 9, the Rise time is comparatively less for Cheng and Hung and SIMC-PID (Γc=0),but there Peak Overshoot tends to be the highest. Therefore, in spite of low values of Rise time, these methods don't hold good enough. Peak Overshoot is an important parameter for tuning selection. Although ZN has no Peak Shoot but it has very high value of Rise time and settling time. Therefore, In the first region, SIMC-PID(Γc=Tm) proves to be the best with least settling time and Peak Overshoot and a quite good value of Rise time.

Table 3: Time Domain parameters for range 0-9

Method	Rise time(sec)	Settling time (sec)	Peak overshoot (%)
Skogested	169	538	4.07
CH	72.3	865	32.5
SPD(Γc=0)	71.1	1150	49.9
SPD(Γc=Tm)	174	536	3.55
ZN	77600	154000	0

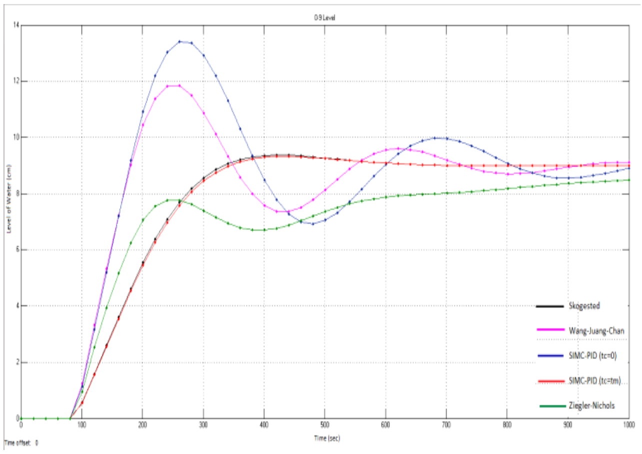

Figure 5:Time Domain comparison in Simulink for region 0-9

4.2 Region 2 (18-27) cm:

In the third region, Cheng and Hung is equally competitive with the Skogestad's method, Since Cheng and Hung has the lowest value of Settling time and Rise time, we cannot neglect the factor that Skogestad's has the least value of Peak Overshoot in this region but still Cheng and Hung proves to be the best method because of its least settling time and Rise time.

Table 4: Time Domain parameters for range 18-27

Method	Rise time(sec)	Settling time (sec)	Peak overshoot (%)
Skogested	1710	5438	4.02
CH	608	5000	16.1
SPD(Γc=0)	720	11500	49.6
SPD(Γc=Tm)	815	8210	38.1
ZN	51400	90800	0

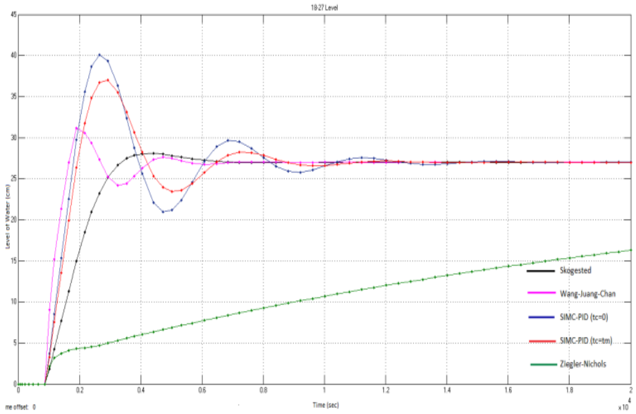

Figure 6: Time Domain comparison for region 18-27

4.3 Region3 (36-45)cms:

The lowest values of Rise time and Settling time for Cheng and Hung suggest that it is the best method in this operating range. Although it has a higher value of Peak Overshoot than Skogestad's, it is still an optimal solution for this region**.**

Table 7: Time Domain parameters for range 36-45

Method	Rise time(sec)	Settling time (sec)	Peak overshoot (%)
Skogested	2680	8520	4.03
CH	981	5930	11.9
SPD(Γc=0)	1130	18100	50
SPD(Γc=Tm)	1150	17800	47.9
ZN	427000	936000	0

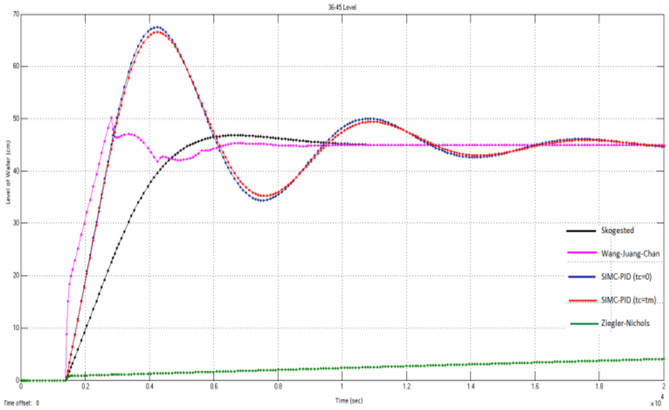

Figure 7: Time Domain comparison for region 36-45

5. Conclusions

Classical Controller might not be the best solution in the industry but is still the most used technique for controlling purposes and a better tuning is always helpful in maintaining the accuracy of the controller.From the discussions above, it is clearly seen that different tuning methods proved efficient than their various counterparts in different regions , taking into account the time domain analysis. We also tested the results for the remaining regions and tuning proved to be an important step of controller design.

References:

[1]. Biegler,L.T., and Rawlings,J.B.,1991,"Optimization approaches to nonlinear model predictive control",In chemical process control,CPCIV,pp.543-571,Austin,TX.
[2]. Kravaris,C.,and Arkun,Y.,1991,"Geometric nonlinear control-An Overview",In chemical process control,CPCIV,pp.477-515,Austin,TX,1991.
[3]. Raich,A.,Wu,X.,and Cinar,A.,1991,"Approximate dynamic models for chemical processes:A comparative study of neural networks and nonlinear time series modeling techniques".,AIChE Conference,Los Angeles,CA,
[4]. George K.I.,Boa-gang Hu.,and Raymand,1999,"Analysis of direct action fuzzy PID controller structures",IEEE Trans.on SMC,29(3),pp.371-388.
[5]. Ziegler J.G..,and Nicolas N.B.,1942,"Optimum settings for automatic controllers",Transctions of ASME,64,pp.759-768.
[6]. ZCheng.M.,and Atherton.D.P.,1993,Automatic tuning of optimum PID controllers,Pro. c.IEE 140(3),pp.216-224.
[7]. Sung S.W.,Lee.O. J.,Lee I..B.,Lee.J,and Yi I..B,1996,"Automatic tuning of PID controller using second order plus dead time delay model",J. of Chem.Eng.of Japan,29(6),pp991-999.
[8]. Astrom,K.J.,and Hagglund,T.,1984,"Automatic tuning of simple regulators with specifications on phase and amplitude margins,Automatica,20(5),pp.645-651.
[9]. Cohen G.H.,Coon G.A.,1953,"Theoritical consideration of retarted control".Transactions of ASME,75,pp 827-834.
[10]. Hang C.C.,and Astrom K.J.,Ho W.k.,1991,"Refinements of the Ziegler Nicolas tuning formula",Proc IEE Control Theory App,138,pp.111-118.
[11]. Garcia C.E.,and Morari M.,2000,"Internal model control:1.A unifying review and some new results,1 &EC Process Des.and Dev.,21(2),pp.308-323.
[12]. Rivera D.E.,Morari M,and Skogestad's S.,1986,Internal Model Control for PID controller Design,I.&E.C. Process Des.and Dev.,25(1),pp.22-265.
[13]. Astrom,K.J.,and Hagglund,1996,Automatic Tuning of PID Controllers",The Control Handbook,CRC Press,Florida,pp817-826.
[14]. De paor.A.M,and O'Malley M.,1989,:Controllers of Ziegler Nicolas type for unstable processes",Int.J.Control,49,pp.1273-1284.
[15]. G.S. Cheng and J.C.Hung.J.C."A least-squares based self tuning of PID Controller",Proc.IEEE South East Conference,North Carolina,1985,pp.325-332
[16]. E.S.Prabhu and M.Chidambram."Robust control of a distillation column by method of inequalities" Indian Chen.Engr.Vol 37,pp.181-187,1991.
[17] K.R. Sundaresan and R.R. Krishnaswamy,"Estimation of time delay,Time constant parameters in Time frequency and Laplace Domains",Can.J.Chem.Eng.Vol.9 No.4.