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Design and Simulation of Hierarchical Control of Two Continuous Stirred Tank Heater in Series

Normah Abdullah[1*], Ramin Razmi[1], Tan Chung Chuan[1], Zulkifli Mohd Nopiah[2], Azah Mohamed[3] and Mohd Zaki Nuawi[4]

[1]Department of Chemical and Process Engineering, Faculty of Engineering and Built Environment, The National University of Malaysia, Bangi - 43600, Malaysia; normah.abdullah@ukm.edu.my, razmi.ramin@yahoo.com, tcc_90@hotmail.com
[2]Fundamental Studies of Engineering Unit, Faculty of Engineering and Built Environment, The National University of Malaysia, Bangi - 43600, Malaysia; zmn@ukm.edu.my
[3]Department of Electrical, Electronic and Systems Engineering, Faculty of Engineering and Built Environment, The National University of Malaysia, Bangi - 43600, Malaysia; profazah@gmail.com
[4]Department of Mechanical and Materials Engineering, Faculty of Engineering and Built Environment, The National University of Malaysia, Bangi - 43600, Malaysia; mzn@ukm.edu.my

Abstract

Background/Objectives: The present investigation was carried out to study the application of process control hierarchy to design an optimal control system for a two continuous stirred tank heaters in series (2-CSTHs). **Methods/Statistical Analysis:** The optimal control strategy of the 2-CSTHs is designed in a hierarchy of three level structures which employs optimization in the real-time optimization (RTO) control level, feedback control scheme in the regulating control level and sensors, pumps and control valves in the measurement and actuation level. All these levels should be coordinated and require information transfer. The proposed optimal control system of the 2-CSTHs was simulated using MATLAB/SIMULINK and HYSYS. **Findings:** The simulation results show that the PID feedback closed-loop of the 2-CSTHs for level and temperature controllers are able to achieve satisfactory performance for servo and regulating control. Simulation on the RTO algorithm also yields results which indicate that the non-derivative technique is an efficient algorithm in minimizing the square of tracking error of temperature in both tanks of the 2-CSTHs. **Application/Improvements:** The non-derivative algorithm has the advantage over other algorithms in the family of ISOPE algorithm for not requiring process derivatives which make it attractive for industrial applications.

Keywords: Continuous Stirred Tank Heater, Feedback Control, Hierarchical Control, Regulating Control, Real-Time Optimization (RTO), Supervisory Control

1. Introduction

A chemical plant is an arrangement of processing units integrated with each other in a systematic and rational manner. The plant's overall objective is to convert certain raw materials into desired products using available sources of energy, in most economical way which can contribute to significant decrease in operating cost. Heaters are processing units which are commonly used in chemical plants for heating purposes. The heating of tanks is a common industrial practice in many applications[1]. Utilization of energy in economical way can be accomplished in heaters by applying advanced process control schemes which may contribute to minimization of energy usage in chemical plants. The development of modern process industries and the increasing fierce competition of the world market have inevitably led to new demand on process control from various industrial sectors. Not only the outputs of the controlled plant are required to best follow their set-points, but also the

* *Author for correspondence*

operation of the whole industrial plant is required to be well controlled so that the operational indices (i.e. the production quality, efficiency and consumptions during production phase) are well controlled into their targeted ranges. Moreover, the quality and the efficiency indices can be enhanced as much as possible whilst the consumption indices are reduced to their lowest possible level so that optimal operation control for industrial processes can be realized. The fast development of computer and communication technologies has provided an implementation platform for optimal operational control in industrial processes[2].

Increasing need for engineers to lower production cost to withstand global competition has prompted engineers to look for rigorous methods in decision making such as optimization methods[3]. Optimization is one of the most important engineering tools that have been used to address the above issues[4] for the purpose of increasing plant efficiency especially for large scale industrial processes. To accomplish the optimization and control of a large-scale system, hierarchical scheme has been applied on the basis of "decomposition and coordination" strategy[5]. Theory of hierarchical systems which is useful for applications has been studied by a group of researchers in the Technical University of Warsaw and the study spurs for much further work in both theoretical and practical aspects[6]. Over the years, research works have shown that hierarchical structures provide stable and robust control in large-scale complex systems[7-11]. The successful implementation of process control hierarchy in the design of control system plays a critical role in making the plant operation profitable. Some of the recent research works involving the integration process design and control using hierarchical control structure for chemical processes can be found in Zhou et al.[12], Mazaeda et al.[13], and Prada et al[14]. In the hierarchical control system, the optimal operation for industrial processes is implemented through on-line calculation of optimal set point, also known as RTO, which allows the profits from the process to be maximized while satisfying the constraints. Recent research works in

This paper is focused on the application of hierarchical control structure in the design of optimal control system for 2-CSTHs. The simulation studies on the 2-CSTHs were performed by firstly considering, the base case design using HYSYS to determine the optimal operating conditions and equipment sizing. Secondly, the simulation consider the PID feedback control system with the objective of determining the performance of the PID feedback controller at the regulating control level and the RTO designed with the objective of evaluating the performance of the RTO control algorithm at the supervisory control level.

The remainder of the paper is organized such that section 2 presents the general idea of process hierarchical control technique. Section 3 describes the process description, modelling and based case designed values of 2-CSTHs. Section 4 elaborates on the process hierarchical control design of the 2-CSTHs. Section 5 presents and describes simulation results of the hierarchical control system of the 2-CSTHs and Section 6 concludes the paper.

2. Process Control Hierarchy

The control task of an industrial plant is to maximize the economics of the process control by considering objectives such as: i. maintaining the plant in a safe operation mode, ii. satisfying demands on product quality and economical usage of technological apparatus, and iii. maximizing the current production profit. Undesirable and uncontrolled process behavior usually leads to serious losses while failure with meeting demands on product quality parameters lead usually to decreased profits. After ensuring safety and quality of products, on-line optimization or RTO can be used in calculating optimal values of the state variables of the processes in the plant[15]. A hierarchical control structure is widely applied in designing an optimal control system for complex chemical plants, which consists of process control activities organized in the form of hierarchy with the required functions at the desired lowest levels, but optional functions at the higher levels. The successful implementation of these process control activities is a critical factor in making plant operation as profitable as possible[16]. Figure 1 shows the four levels in the process hierarchical, which is normally employed in controller design, in which various control activities such as plant management, optimization, control, monitoring and data acquisition are employed.

2.1 Measurement and Actuation

As shown in Figure 1, the lowest level consists of the measurement devices (sensors and transmitters) and actuation equipment (pumps and control valves) which are used to measure process variables and implement the calculated control action. These devices are interfaced to the control system[16].

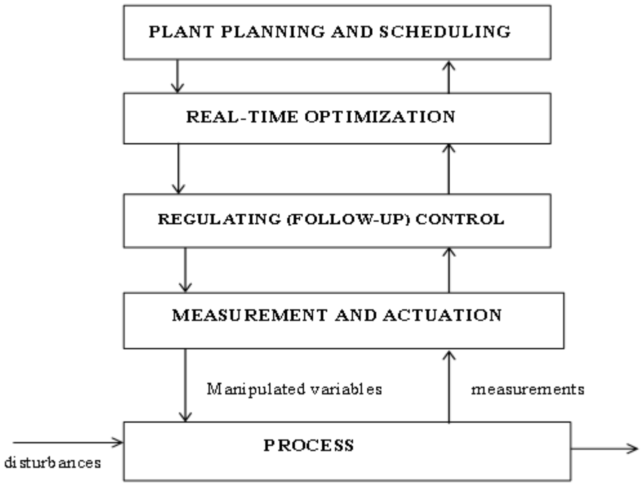

Figure 1. Hierarchical control structure for chemical processes.

2.1.1 Pump Sizing

Energy equation is used to size pumps that are employed in the optimal controller designed of 2-CSTHs[17] and it is given by,

$$g\Delta z+\frac{\Delta P}{\rho}-\frac{\Delta P_f}{\rho}-W=0 \quad (1)$$

where:
W: Work done by fluid, J/kg
Δz: Difference in elevations (z_1-z_2), m
ΔP: Difference in system pressures (P_1-P_2), N/m^2
ΔP_f: Pressure drop due to friction, including miscellaneous losses, and equipment losses, N/m^2
P: Liquid density, kg/m^3
g: Acceleration due to gravity, m/s^2

2.1.2 Valve Sizing

It is common that a valve should be designed to use 10-15 % of the total pressure drop or 10 psi, whichever is greater. The control valves are used as final elements (actuator) in feedback control loops, where the device enables a process variable to be manipulated so that the controlled variable will be at the desired set points. Eq. (2) is a design equation used for sizing control valves which relates the actual flow rate, $\bar{f}$ to the valve coefficient C_v[18].

$$C_v=\bar{f}\sqrt{\frac{G_f}{\Delta P_v}} \quad (2)$$

where
$\bar{f}$ Liquid flow, u.s.gpm
ΔP_v: Pressure drop across the valve, psi
G_f :Specific gravity of liquid at flowing condition

2.2 Regulating Control

At the regulatory level as shown in Figure 1, the controlled variables such as level and temperature are maintained at or close to their set points. Standard feedback control technique is used for maintaining the plants at the desired controlled set points. Proportional-Integral-derivative (PID) control algorithm which is the most popular controller used in industrial control, is applied in the feedback control of the plant due to its simplicity and also satisfactory performance. The PID controller can be represented as[18]:

$$m(t)=\bar{m}+\frac{K_c}{\tau_I}\int e(t)dt+K_c\tau_D\frac{de(t)}{dt} \quad (3)$$

where
m(t): controller output
$\bar{m}$: bias
K_c: controller gain
τ_I: integral (or reset) time
τ_D: derivative (or rate) time
e(t): error signal

The three parameters, K_c, τ_I, and τ_D must be tuned to obtain satisfactory control.

2.3 Real-Time Optimization

For achieving optimal process operation in MPC/classical PID, the steady-state set-point optimizing control (SSOC) algorithm is employed, with the purpose of calculating online optimal set-points for MPC/classical PID[15], by maximizing economic profit. Research in SSOC has become increasingly important since the last three decades. Numerous types of SSOC algorithms are available in the literature but most of these algorithms are suboptimal. One class of algorithm well known under the name of Integrated System Optimization and Parameter Estimation (ISOPE) has attractive features compared to other SSOC algorithms in which it is able to generate a series of set-points converging to the real plant optimal solution in spite of uncertainty in the process model and disturbance estimates[19].

3. Process Description

The process consists of two continuous stirred tank heaters in series (2-CSTHs) with recycle as shown in Figure 2. In the first tank, the input feed stream of F_1 and the recycle stream of F_2 are mixed and heated. The outlet flow of F_4 from the first tank flows into the second tank where it is again heated. Both tanks are heated by electrical heaters.

The variables used in the modeling and controlling of 2-CSTHs are summarized in Table 1.

Table 1. Variables used in modeling and control of 2-CSTHs

Variables	Units
F_1	Water feed flow rate, $m^3\ min^{-1}$
F_2	Output flow rate, $m^3\ min^{-1}$
F_3	Inlet flow rate to tank 2 of 2-CSTHs, $m^3\ min^{-1}$
F_4	Output flow rate from tank 2 of 2-CSTHs, $m^3\ min^{-1}$
F_5	Output flow rate from tank 2 of 2-CSTHs, $m^3\ min^{-1}$
T_o	Feed temperature to 2-CSTHs for tank 1, ^OC
T_1	Temperature in CSTH 1, ^OC
T_2	Temperature in 2-CSTHs in tank 2, ^OC
h_1	Liquid level in 2-CSTHs for tank 1, m
h_2	Liquid level in tank 2 of 2-CSTHs, m
ρ	Water density, $kg\ m^{-3}$
T_{ref}	Reference temperature, ^OC
C_p	The heat capacity of water, $kJ\ kg^{-1}\ K^{-1}$
A	Cross section area of both tanks in 2-CSTHs, m^2
Q_1	The amount of heat supplied by electrical heater 1, K_W
Q_2	The amount of heat supplied by electrical heater 2, K_W
m_1	Signal to electrical heater 1
m_2	Signal to electrical heater 2
V_{F2}	Valve 2 opening
V_{F3}	Valve 3 opening
T_m	Measured temperature, ^OC
T_1^*	Real process temperature of 2-CSTHs for tank 1, ^OC
T_2^*	Real process temperature of 2-CSTHs for tank 2, ^OC
T_{1sp}	Temperature set-point for 2-CSTHs for tank 1, ^OC
T_{2sp}	Real process temperature of 2-CSTHs for tank 2, ^OC

The process model of 2-CSTHs consists of two linear and two nonlinear ordinary differential equations which are given as follows:

$$A\frac{dh_1}{dt} = F_1 + F_2 + F_3 \tag{4}$$

$$A\frac{dh_2}{dt} = F_4 - F_5 - F_6 \tag{5}$$

$$Ah_1\frac{dT_2}{dt} = \left(F_1T_0 + F_2T_2\right) - F_3T_1 + \frac{Q_1}{\rho C_p} \tag{6}$$

$$Ah_2\frac{dT_3}{dt} = F_4T_1 - \left(F_5 + F_6\right)T_2 + \frac{Q_2}{\rho C_p} \tag{7}$$

where h_1 and h_2 are the heights of the liquid levels in CSTH 1 and 2, respectively. T_2 and T_3 are the temperature in the CSTH 1 and 2, respectively. The remaining variables and parameters and nominal operating conditions are shown in Table 2.

Table 2. Tanks sizes and steam properties

Variables	Symbol	Unit	Value
Diameter	D	m	0.5
Tank height	H	m	1.639
Cross-section area of both tanks	A	m^2	0.197
Feed flow rate to tank 1of 2-CSTHs	F_1	$m^3\ min^{-1}$	0.012
Output flow rate from tank 1 of 2-CSTHs	F_2	$m^3\ min^{-1}$	0.015
Inlet flow rate to tank 2 of 2-CSTHs	F_3	$m^3\ min^{-1}$	0.015
Output flow rate from tank 2 of 2-CSTHs	F_4	$m^3\ min^{-1}$	0.012
Output flow rate from tank 2 of 2-CSTHs	F_5	$m^3\ min^{-1}$	0.003
Feed temperature to tank 1 of 2-CSTHs	T_O	^OC	20
Temperature in tank 1 of 2-CSTHs	T_1	^OC	50
Temperature in tank 2 of 2-CSTHs	T_2	^OC	80

The base case design for 2-CSTHs has been carried out by the HYSYS software simulation[1].

4. Process Hierarchical Control Design

The optimal control strategy of the 2-CSTHs is organized in a three level hierarchical structure which employs optimization in the RTO control level, feedback control scheme in the direct control level and sensors, pumps and control valves in the measurement and actuation level.

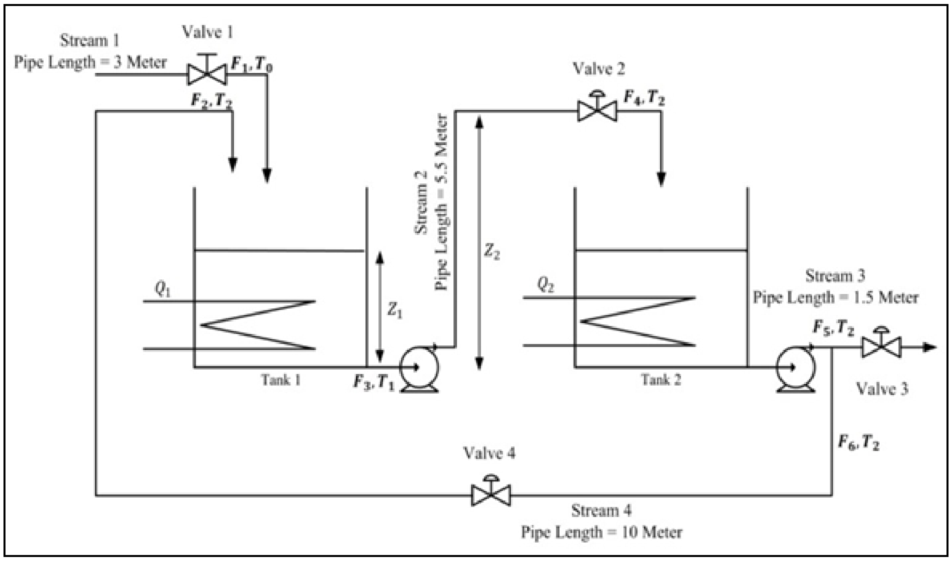

Figure 2. A schematic process flow diagram of 2-CSTHs.

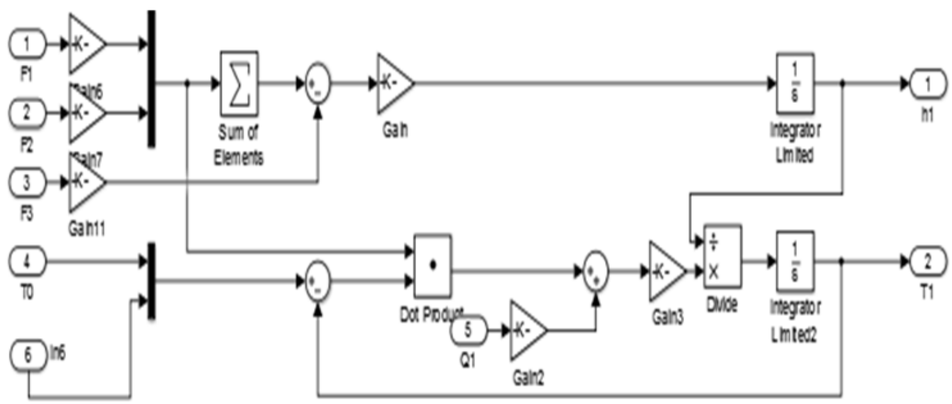

Figure 3. The Simulink model for tank 1 of 2-CSTHs.

4.1 Plant Model Simulator

The simulator for the plant model of 2-CSTHs was developed by considering two linear and two non-linear ordinary differential equations (Eqs. (4) to (7)). The simulator was programmed in the MATLAB/ SIMULINK environment. Figures 3 and 4 show the simulator component models of tank 1 and 2 of the 2-CSTHs, respectively.

The MATLAB-SIMULINK simulator consisting of tank 1 and tank 2 of the 2-CSTHs was developed as interconnected components as shown in Figure 5.

4.2 PID Feedback Controller

Standard PID feedback controller is used at the regulatory control level to control the height of the liquid levels and temperature in both tanks of the 2-CSTHs. Figure 6 shows the Simulink block diagram for the feedback controller. The control objectives are described as follows:

- Liquid level h_1 and h_2 are controlled by manipulating output flow streams of rate F_3 and F_5, respectively.
- Temperatures T_3 and T_4 are controlled by manipulating the current flow through the electrical coils in tank 1 and 2 of the 2-CSTHs, respectively.

The controller is tuned automatically using the built in Simulink PID block and the controller parameters of the 2-CSTHs are shown in Table 3.

Table 3. PID controller tuned parameters

	h_1 controller	h_2 controller	T_1 controller	T_2 controller
K_c	-7.20	-9.01	0.52	0.35
τ_I	-6.33	-7.91	0.62	0.48
τ_D	3.39	4.24	-0.20	-0.12

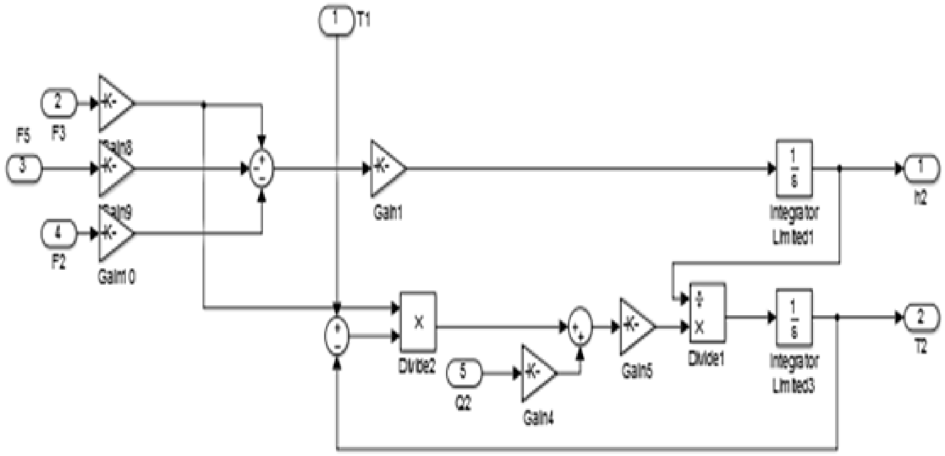

Figure 4. The Simulink model for tank 2 of 2-CSTHs.

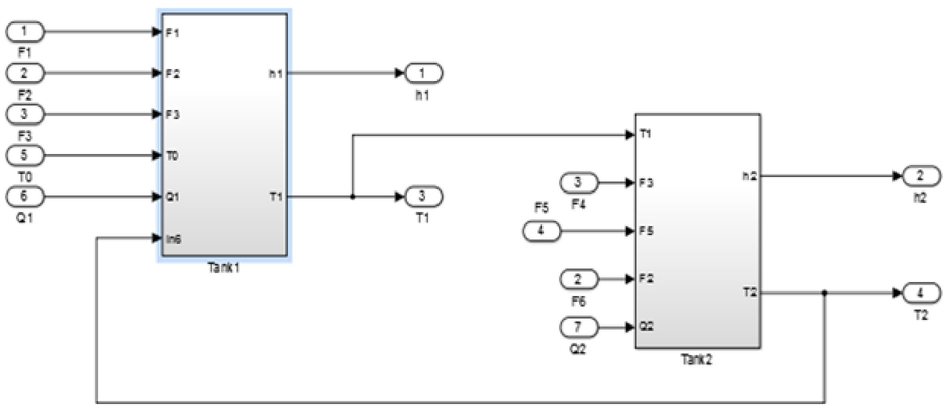

Figure 5. Model Simulink of 2-CSTHs.

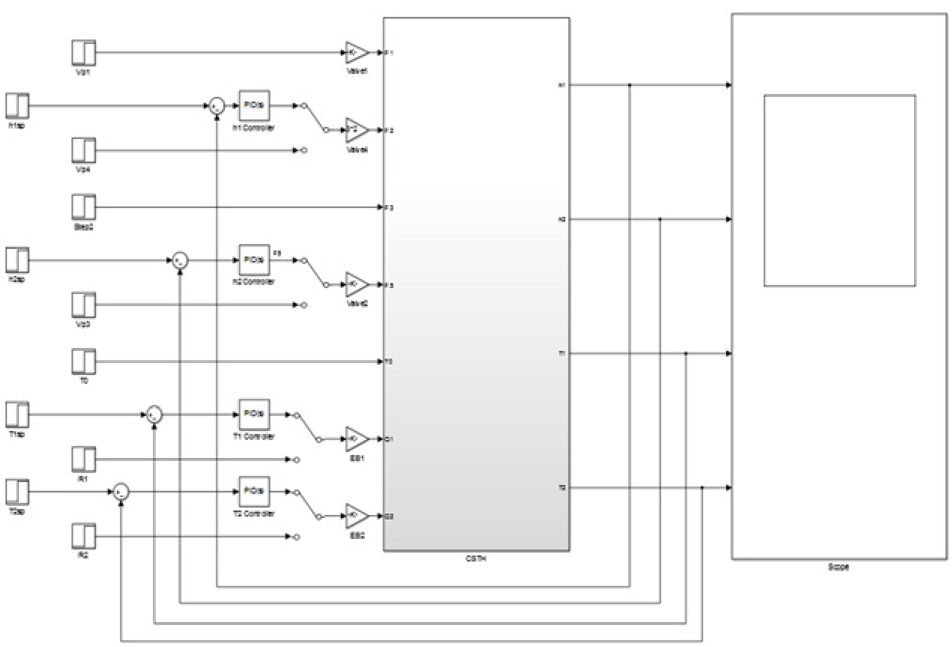

Figure 6. Feedback closed loop controller strategy for the 2-CSTHs.

4.3 RTO Controller Design

In the optimal control design of the 2-CSTHs, RTO also known as supervisory controller, is added into the hierarchical system without modifying the regulating level. The objective function of the optimization problem includes the plant profit, operational costs, energy consumption of the process or other criteria, including regulatory objectives such as the square of the tracking error. The optimal control strategy proposed is at the supervisory level that determines the optimal set-point for a given regulatory control as shown in Figure 7[20]. The supervisory level optimizes a general steady state objective function subject to equality and inequality constraints.

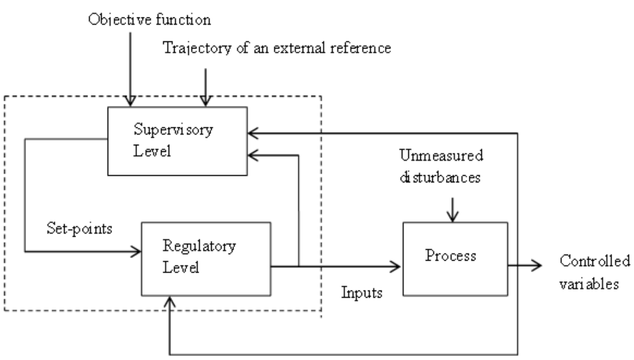

Figure 7. Control diagram of the supervisory and regulatory control levels structure.

The real-time optimization control configuration of the 2-CSTHs is shown in Figure 8. It is assumed that the water level in both tanks is fixed and the optimization unit does not change the control valve openings V_{p2}, V_{ps} instead the control signals to the electrical boxes.

4.3.1 RTO Control Algorithms

In this work, non-derivative algorithm from the family of ISOPE was applied in the RTO control level to recalculate the optimum set points for obtaining the current optimum operating conditions for the 2-CSTHs. The non-derivative algorithm was chosen because the algorithm converges to the plant real optimal steady-state points with least number of iterations and does not require process derivatives as compared to other centralized algorithms in the family of ISOPE algorithms[21].

The Optimization Control Problem (OCP) for the process plant can be mathematically formulated as

$$\min Q(c, y) \qquad (8)$$

$$y = F_*(C)$$

$$\in C = \{c : G(c) \leq 0\}$$

where $c = [c_1, c_2, ..., c_n]^T \in R^n$ and $y = [y_1, y_2, ..., y_n]^T \in R^m$ are the process control (set point) and output vectors, respectively. $Q : R^n \times R^m \rightarrow R$ is a given performance map, $F_* = R^n \rightarrow R^m$ is the real process input-output map and

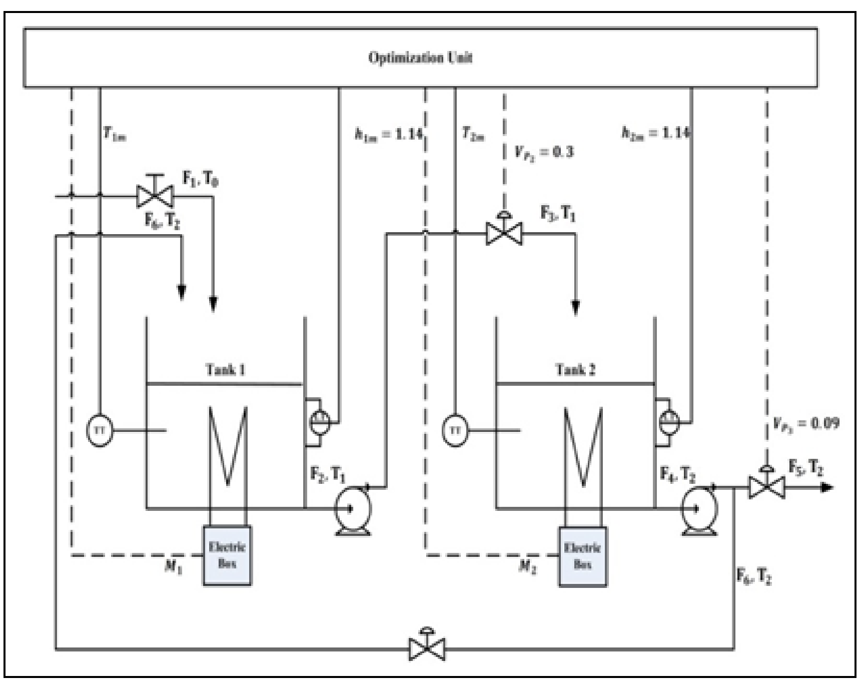

Figure 8. RTO control configuration of the 2-CSTHs.

$G : R^n \rightarrow R^p$ is the process inequality constraint map. The maps Q and G are assumed to be known exactly. The real process input-output map is given by,

$$Q : R^n \text{ x } R^m \rightarrow R \tag{9}$$

Where $a \in R^s$ are the process model parameters.

It is assumed that the map $F_*(.)$ is not known exactly, but that we can measure the process output $y = F_*(v)$ and output derivative $F'_*(v)$ at any $v \in C$

The non-derivative algorithm can be implemented as follows:

Step 1: Choose v^{-1}, $V^0 \in C$ and $\varepsilon > 0$. Set i=0. The output $y = F_*(v^{-1})$ has been achieved.

Step 2: Measure the reality output $y = F_*(v^i)$ at the point $v^i \in C$.

Step 3: Construct the linear function

$$F\left(c, v^{i-1}\right) = \left(K^i\right)^t c + B^I \tag{10}$$

To fit the function F_*I, pass through the points v^{i-1}, $v^i \in C$.

Step 4: Solve the sub problem

$$\min_{c \in C} Q\left(c, F\left(c, v^{i-1}, v^i\right)\right) + \frac{\rho}{2}\left(c - v^i\right)^2 \tag{11}$$

Where $\rho > 0$ is a penalty coefficient.

Let c^I be a solution. If $c^I = v^I$, then v^I is a solution, and the algorithm is terminated.

Otherwise continue to the next step.

Step 5: Update v^I by using the following equation:

$$v^{i+1} = c^I \tag{12}$$

Set i=i+1 and return to step 2.

The flow diagram for implementing the non-derivative algorithm is shown in Figure 9.

In formulating the OCP for the 2-CSTHs, several assumptions were made as follows:

$$Q_1 = \frac{q_{1\max}}{100} m_1 \tag{13}$$

$$Q_2 = \frac{q_{2\max}}{100} m_2 \tag{14}$$

$$F_5 = 0.8F_4 \tag{15}$$

$$F_6 = 0.2F_4 \tag{16}$$

The OCP of the 2-CSTHs is derived from the process model (Eqs. (4) to (7) and Eq. (13) to (16)) and it is described as follows:

Objective function,

$$\min\left\{\left\|T_{1m} - T_{1sp}\right\|^2 + \left\|T_{2m} - T_{2sp}\right\|^2\right\} \tag{17}$$

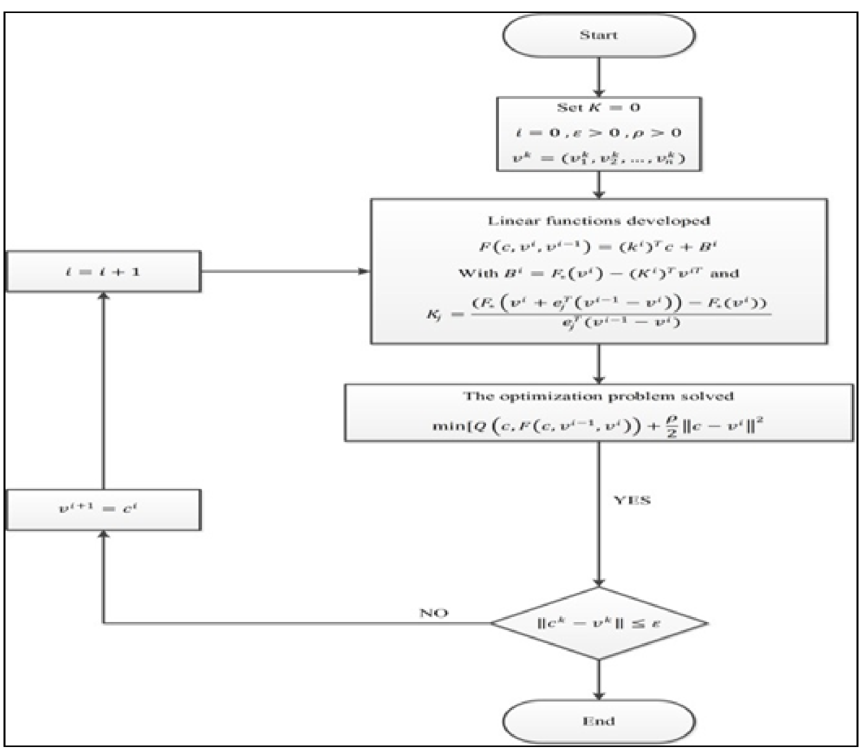

Figure 9. The non-derivative algorithm flow chart.

Real output,

$$\begin{bmatrix} T_1^* \\ T_2^* \end{bmatrix} = \begin{bmatrix} 63.4 + 696m_1 \\ 1169m_2 \end{bmatrix} \begin{bmatrix} 7.83 & -0.8 \\ -7.83 & 7.7 \end{bmatrix}^{-1} \quad (18)$$

Constraint equations,

$$0 \le m_1 \le 100 \quad (19)$$

$$0 \le m_2 \le 100 \quad (20)$$

5. Simulation studies

In this section, the optimal control system of the 2-CSTHs designed based on hierarchical control technique is evaluated. The control strategy scheme is structured in a hierarchy of three levels structure which employs optimization in the RTO control level, PID feedback control scheme in the regulating control level and sensors, pumps and control valves in the measurement and actuation level. The base case design has been performed using the HYSYS for valves, heaters, pumps, tanks and pipe sizing. The feedback closed loop for PID was developed and implemented using the MATLAB/SIMULINK software for the purpose of evaluating the performance of the PID controller. The non-derivative algorithm is applied at the RTO control level for evaluating online optimal set-points for PID controller, with the objective of minimizing the square of tracking error.

5.1 Based Case Design

The base case design has been carried out for both manual calculation and the HYSYS software simulation. Results from both methods are presented in Table 4 and these values are used in the hierarchical control design of the 2-CSTHs.

Table 4. HYSYS simulation and manual calculation design results of 2-CSTHs

	Valve sizing	
	Manual calculation (gpm psi$^{-1/2}$)	**HYSYS simulation (gpm psi$^{-1/2}$)**
Valve 1	35.90	38.31
Valve 2	10.58	10.65
Valve 3	25	33.65
Valve 4	2	2.038
	Heater sizing	
	Manual calculation	HYSYS simulation
Heater 1	-	19.35
Heater 2	-	32.46
	Tank sizing	
	Manual calculation	HYSYS simulation
Tank 1 and 2	Diameter = 0.5 m, Height = 1.639 m, Volume = 0.3218 m^3	Diameter = 0.5 m, Height = 1.639 m, Volume = 0.3218 m^3
	Pipe sizing for each stream	
	Manual calculation (m)	HYSYS simulation (m)
Stream 1	0.0254	0.0254
Stream 2	0.0254	0.0254
Stream 3	0.0254	0.0254
Stream 4	0.0127	0.0127
	Pressure loss in each stream	
	Manual calculation (kNm^{-2})	HYSYS simulation
Stream 1	1.3	0.4656
Stream 2	10.72	9.627
Stream 3	1.32	0.6328
Stream 4	11.66	10.78
	Pump sizing	
	Manual calculation (Watt)	HYSYS simulation (Watt)
Pump 1	7	4
Pump 2	8	5

5.2 Feedback Closed-loop Response

Feedback closed loop responses for ±5% step changes in T_1 and T_2 are shown in Figures 10a and 10b, respectively. As shown for these step changes, both controllers are able to achieve satisfactory servo performance in controlling the temperature of both tanks of the 2-CSTHs.

Feedback closed loop responses for ±0.14m step changes in h_1 and h_2 are shown in Figures 11a and 11b, respectively. As shown for these step changes, both controllers are able to achieve satisfactory servo performance in controlling the level of both 2-CSTHs.

Figure 12 shows the feedback close-loop response in the presence of ±10% step change in the feed flow rate, F_1 for level and temperature control of both CSTH. Both controllers in both CSTH are able to reject the disturbance change occurring in the feed flow rate and returned back to their set-points.

5.3 Real-Time Optimization Response

The RTO algorithm was programmed by using MATLAB to simulate the algorithm of non-derivative by following the procedure highlighted in the methodology (section 4.3.1). The results obtained from simulation for the Non-Derivative algorithm and the real process solutions are presented in Table 5. The real solution is obtained by performing optimization on OCP which involve Eqs. 17, 18, 20 and 21.

Table 5. Comparison between the results of non-derivative and real solution

	Non-derivative	Real solution
Objective function	0	0
LS	2	-
LT	6	-
P	0.1	-
ε_v	1.0	-
m_1	20.96	20.70
m_2	23.59	23.29

The non-derivative algorithm converges to real optimum within two iterations as shown in Figure 13

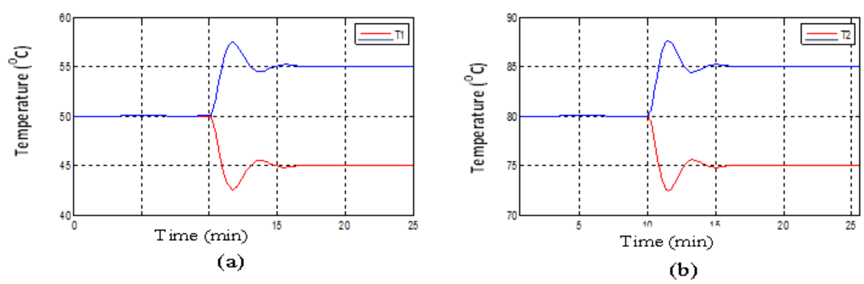

Figure 10. Feedback closed loop response with ± 5% set-point step changes in T_1 and T_2.

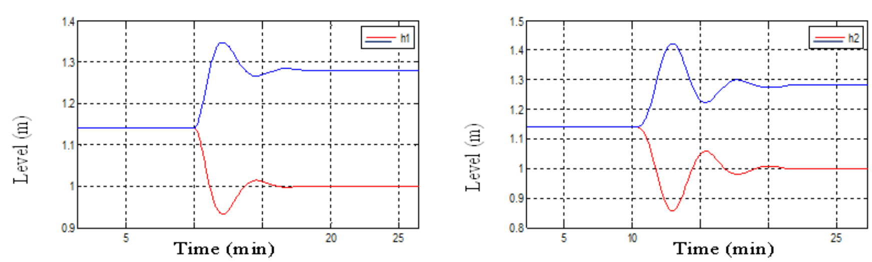

Figure 11. Feedback closed loop response with ± 5% set-point step changes in h_1 and h_2.

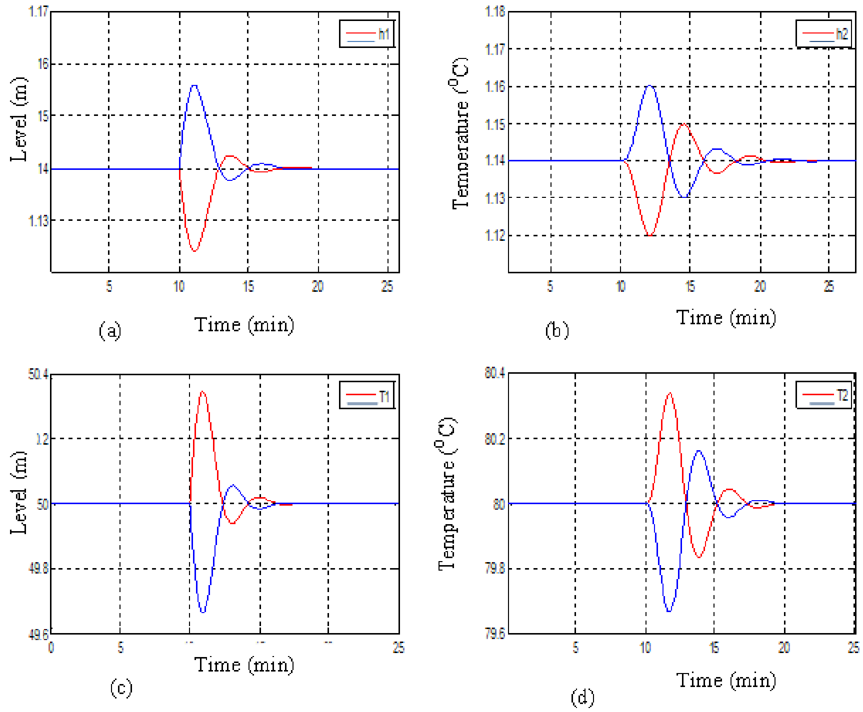

Figure 12. Feedback closed loop response with ±10 % disturbance step change in feed flow rate, F_1.

and it is found that the objective function value obtained from the real solution and the non-derivative algorithm is very close as shown in Table 5. Since the convergence of non-derivative algorithm is fast and it does not require any real process derivatives, therefore it will be attractive for industrial application. Figures 13 and 14 show the performance of the objective function and the set points changes during the optimization process for the applied algorithm, respectively.

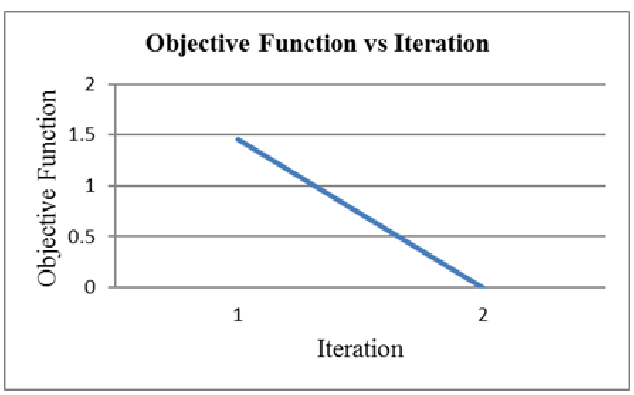

Figure 13. Objective function versus iteration during optimization process for the 2-CSTHs.

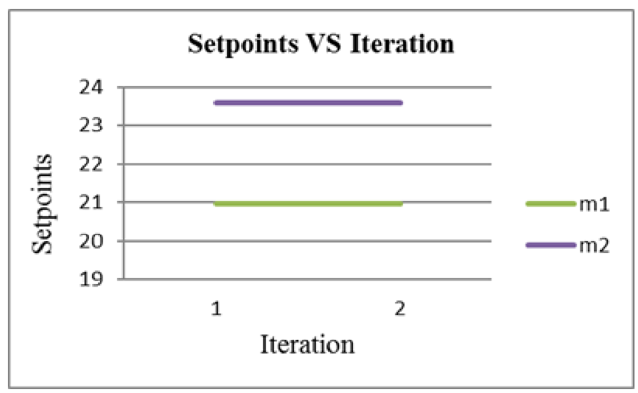

Figure 14. Set points versus iteration during optimization process of the 2-CSTHs.

6. Conclusion

A detailed study on the application of hierarchical control scheme applied to the 2-CSTHs for designing an optimal control system has been presented. The design conditions and equipment sizing for laboratory scale 2-CSTHs are obtained from based case design using manual calculation and HYSYS simulation. The simulation results show that the PID feedback closed-loop is also able to achieve satisfactory servo and regulating performance. Simulation on the RTO algorithm also yields results which indicate that the non-derivative technique is an efficient algorithm in minimizing the square of tracking error of temperature in both tanks of the 2-CSTHs for which only two iterations are required to converge to real optimum. The algorithm has the advantage over other algorithms in the family of ISOPE algorithm for not requiring process derivatives which make it attractive for industrial applications.

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