Empirical Model Estimation and Model Comparison on a Temperature Pilot Plant

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***Abstract*—Model of a plant is used to get an insight of the physical system behavior and there always a scope exists to improve the model parameters by using proper estimation techniques. First principle model is more accurate if the physics of the plant is well known and less complex whereas empirical model estimation is used to estimate the model if the plant is complex and will give an accurate result over the operating region of estimation. In this work, first principle method is used to find the model of a temperature plant and a user interface is developed to provide flexibility to estimate the model for a range of physical parameter. The user interface provides an online view of the model and its response along with a future to simulate the model behavior. In the second half of the work, an empirical model is estimated using open loop experiment data. The models are compared and found to be closed matching. This work provides a simple user interface to estimate the model and to understand the model response with a wide variety of features like data storage, trend plot and with open loop and closed loop response analysis.**

***Keywords*— Empirical model, First principle model, FOPDT model, Open loop response, Temperature plant**

I. INTRODUCTION

A temperature process is usually considered as a slow process and exhibit one of its kind characteristics like if there is an increase in temperature it is very rapid until it reaches steady state whereas heat dissipation will be very slow and take a good amount of time. In this work, the model of a temperature pilot plant is estimated by two different methods namely, First Principle Method and Empirical Method. The procedures of both the methods are discussed in the subsequent section.

A. First Principle Method [1,2,3]:

Considering a thermal plant as shown in Fig. 1, ΔH be a small change in the heat input rate from its steady state value can be given as the sum of change in heat output rate with the change in heat storage rate as

$$\Delta H = \Delta H_1 + \Delta H_2 \qquad (1.1)$$

The change in outflow heat rate can be given as

$$\Delta H_1 = Q\ Cs\ \Delta\theta = \Delta\theta / R \qquad (1.2)$$

Where

ΔH_1 = Change in heat output rate
Q = Steady state liquid flow rate
C_s = Specific heat of liquid
Δθ = Change in temperature of outflow
R = $1/QC_s$ which is defined as the *Thermal Resistance.*

The change in heat storage rate is given by

$$\Delta H_2 = MC_s\, d\Delta\theta/dt = C\, d\theta\Delta/dt \qquad (1.3)$$

Where

M = mass of the liquid in the tank
dθΔ/dt = rate of rise of temperature in the tank
C = MC_s which is defined as *Thermal Capacitance*.

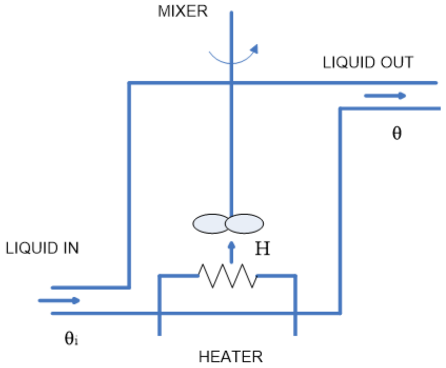


Fig 1. Typical thermal system

The mathematical model of a thermal system can be derived and given as [1]

$$\Delta H = \frac{\Delta\theta}{R} + C\frac{d\Delta\theta}{dt} \qquad (1.5)$$

$$\frac{\Delta\theta(s)}{\Delta H(s)} = \left[\frac{R}{1+RCs}\right] \qquad (1.6)$$

B. Empirical Method[4]:

Empirical model is based on experimentation about a nominal operating condition. An input excitation is induced and the resulting dynamic response will be captured to estimate the model. The procedure is a linearization of the process that is valid for some region. The method is called as 'Process Reaction Method'.

This method involves the following steps:

- Excite a small input and allow the process to reach steady state.
- Introduce a single step change.
- Collect output response data until the process again reaches steady state.
- Perform the graphical process reaction curve calculations.

A first-order-with-dead-time model will be estimated using this approach. With X(s) and Y(s) as input and output in Laplace form, the model in FOPDT form can be given as:

$$\frac{Y(s)}{X(s)} = \frac{K_p e^{-\theta s}}{\tau s + 1} \qquad (1.5)$$

This technique is suggested by Ziegler and Nichols (1942), uses the graphical calculations shown in Fig 2.

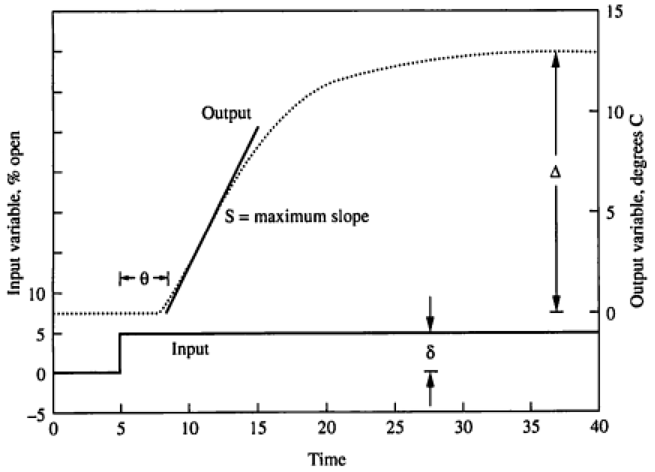


Fig 2. Process Reaction Curve

Where,
S - Maximum Slope
Δ - magnitude of the steady-state change in the output
δ - magnitude of the input change

The model parameters can be calculated as

$$K_p = \frac{\Delta}{\delta} \text{ and } \tau = \frac{\Delta}{S}$$

II. EXPERIMENTAL SETUP

The experimental setup and its P&I diagram are shown in Fig 3 and 4 [5]. It consists of a process tank fitted with a heater to heat the fluid. The tank has inflow and outflow pipelines to allow continuous water supply. The inflow can be viewed and controlled by a rotameter. The temperature of the outflow is measured by a RTD temperature transmitter. The heater supply is controlled by a digital indicating controller by means of a solid state relay (SSR). These units along with necessary piping are fitted on the support frame. The setup is designed for tabletop placement and access. The controller is connected to computer through USB for monitoring and controlling the process.

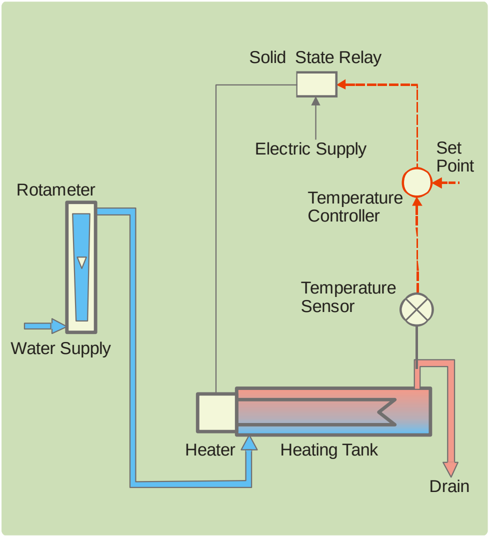


Fig 3. P&I diagram of temperature process

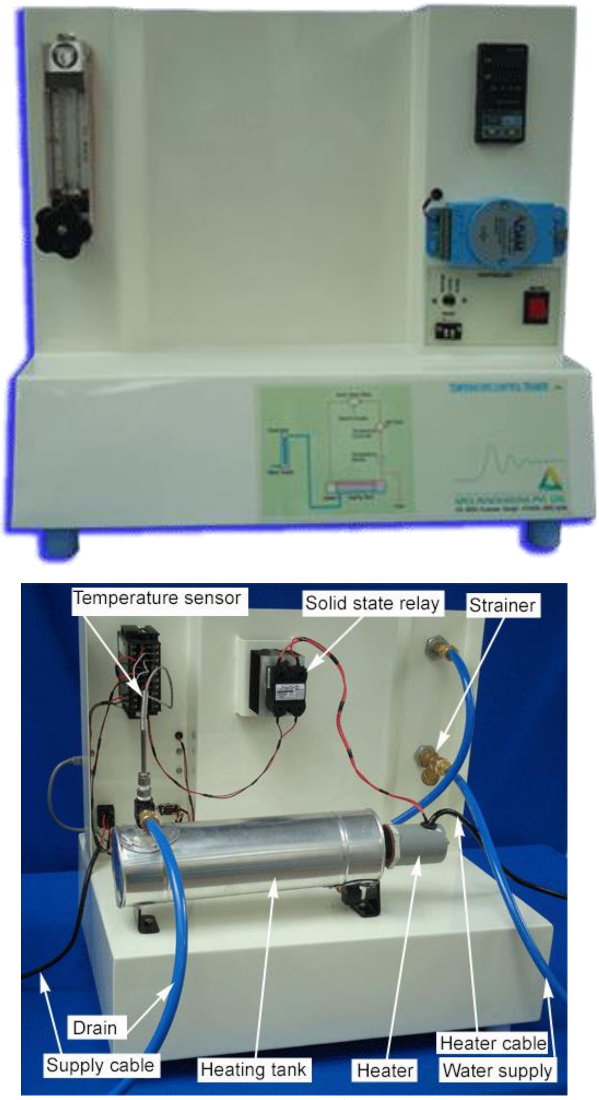


Fig 4. a) Front view b) Back view of the temperature plant

III. MODEL ESTIMATION

The model parameters are tabulated below:

TABLE I. Model parameters

Volume of heater tank	0.5 L
Heater Capacity	3 KW
Specific Heat of Fluid (water)	1
Steady state flow	50 LPH

Based on the standard equations given in Eqn. 1.1 to 1.3, the model is calculated. It can also be noted that the transfer function is normalized to 1% of heater power change and proper conversion is taken care in converting the engineering units to normalized units. The first principle model can be given as

$$G(s) = \left[e^{-13s} \frac{1.0368}{1+72s} \right] \tag{2.1}$$

The time delay provided in the above equation is directly observed from the plant and used. A set of experimental data is observed from the plant by applying a step change in the open loop mode and the response is given in Fig 5.

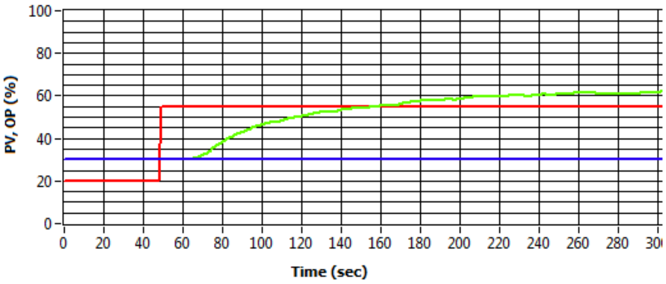


Fig 5. Open loop response for a step change of 20% to 55% controller output

To distinguish actual time lag and time lag because of the coldness of the fluid, a controller output of 20% is applied and the process is allowed to reach steady state slightly higher than the room temperature before applying the actual step input. This process will warm up the fluid and a better model can be estimated from the response.

The slope, lag and change in PV and controller output are calculated the empirical model is calculated as

$$G(s) = \left[e^{-14s}\,\frac{0.9}{1+70s} \right] \qquad (2.2)$$

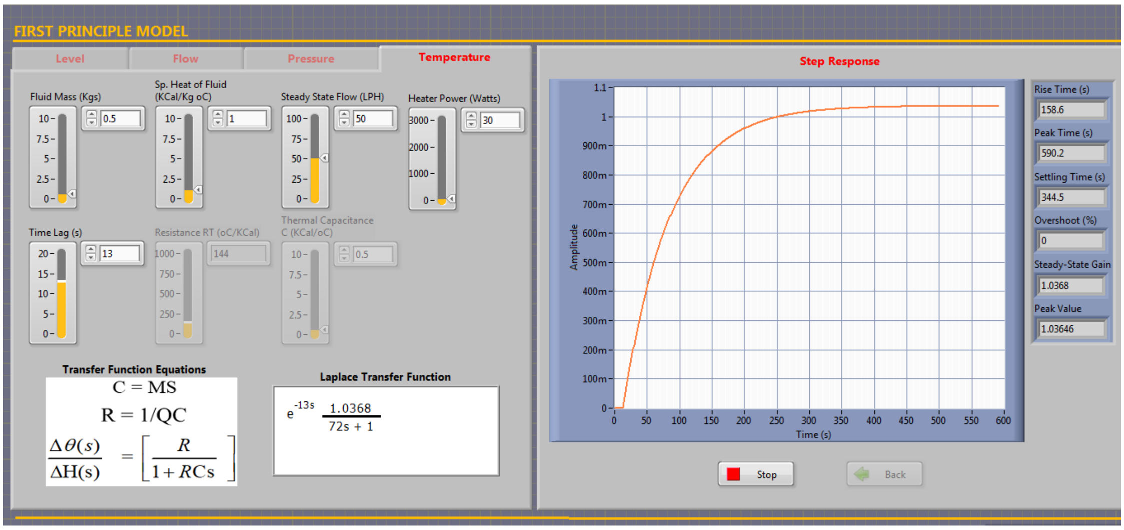
It can be observed that the model obtained from both the methods are close enough and based on the model a user interface is developed as discussed in the next section.

IV. GRAPHICAL USER INTERFACE (GUI)

Based on the first principle model a graphical user interface (GUI) is developed to estimate the model for a range of input parameters [6] as shown in Fig 6. The user interface is developed in LabVIEW [7] platform and accepts user input for a variety of parameters like fluid mass, specific head of fluid, steady state flow, heater power and time lag. The software automatically calculates heat resistance, heat capacitance and the mathematical model in first order plus dead time (FOPDT) form. It also displays the response of the plant model along with model characteristics such as rise time, peak time, settling time, overshoot, steady state gain and peak value.

The software is fully dynamic in the sense that any change in model parameter will give a display because of the change immediately. This software will be used to understand or estimate the model for any of similar temperature process plants.

As the second phase of the work, the model is used to simulate a temperature plant in closed loop with a PID controller [8-11] as shown in Fig.7.



Fluid Mass (Kgs): 0.5
Sp. Heat of Fluid (KCal/Kg oC): 1
Steady State Flow (LPH): 50
Heater Power (Watts): 30
Time Lag (s): 13
Resistance RT (oC/KCal): 144
Thermal Capacitance C (KCal/oC): 0.5

Transfer Function Equations

$$C = MS$$

$$R = 1/QC$$

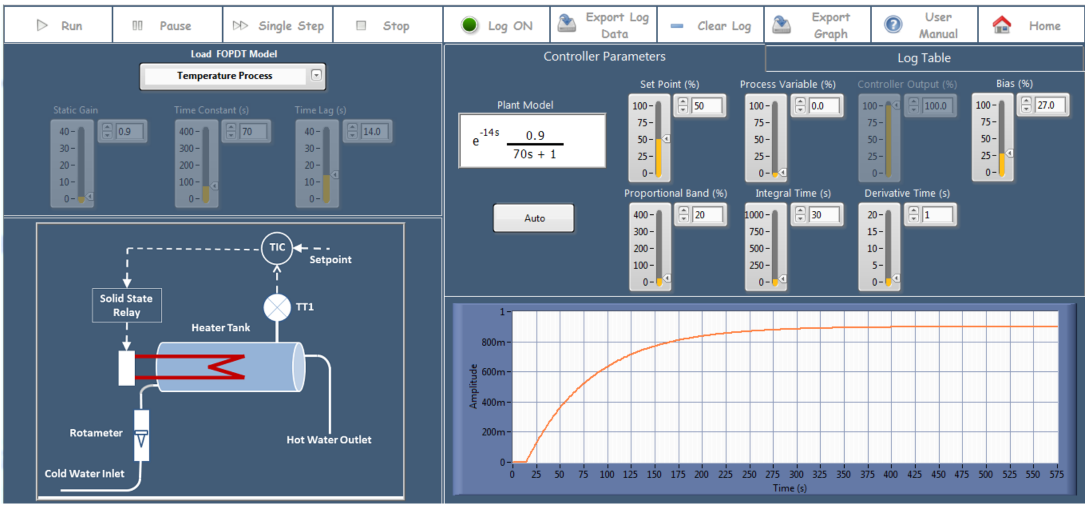
$$\frac{\Delta\theta(s)}{\Delta H(s)} = \left[\frac{R}{1+RCs}\right]$$

Laplace Transfer Function

$$e^{-13s}\,\frac{1.0368}{72s+1}$$

Rise Time (s): 158.6
Peak Time (s): 590.2
Settling Time (s): 344.5
Overshoot (%): 0
Steady-State Gain: 1.0368
Peak Value: 1.03646

Fig 6. Graphical User Interface (GUI) to estimate First Principle Model



Load FOPDT Model: Temperature Process
Static Gain: 0.9
Time Constant (s): 70
Time Lag (s): 14.0

Plant Model

$$e^{-14s}\,\frac{0.9}{70s+1}$$

Set Point (%): 50
Process Variable (%): 0.0
Controller Output (%): 100.0
Bias (%): 27.0
Proportional Band (%): 20
Integral Time (s): 30
Derivative Time (s): 1

Fig 7. GUI to simulate closed loop with PID control

V. RESULTS AND CONCLUSION

A set of real time experiments has been performed on the plant and the results are shown below. The response of a PID controller from the startup with the room temperature close to 26^{0}C (26%) is shown in figure 8. Tuning of PID is not considered as a part of this work. This work is focused on comparison between real time response to the model based simulator response for a particular value of Kp, Ki and Kd. For convenience, Proportional Band (%), Integral Time (s) and Derivative time (s) is considered with a value of PB=50%, IT=30s and DT=1s and the responses are shown in figure 9 and 10.

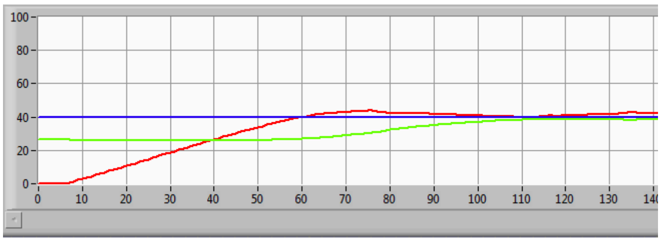


Fig 8. Plant response of a PID controller with 40% setpoint

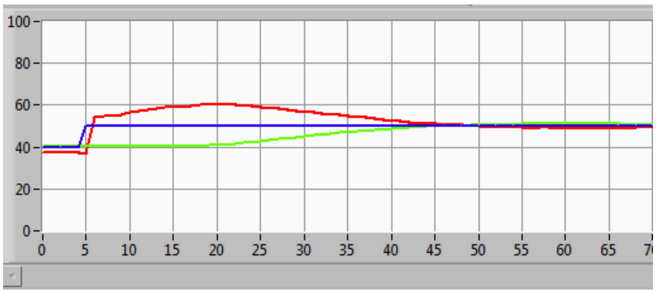


Fig 9. PID response for a setpoint change

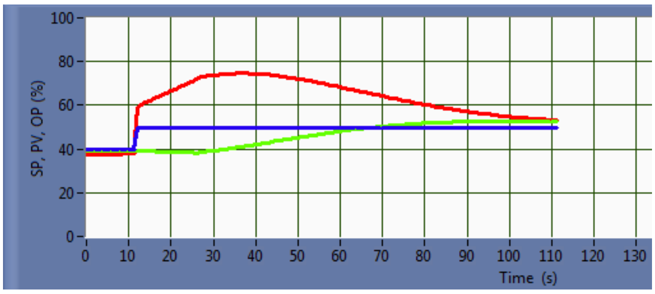
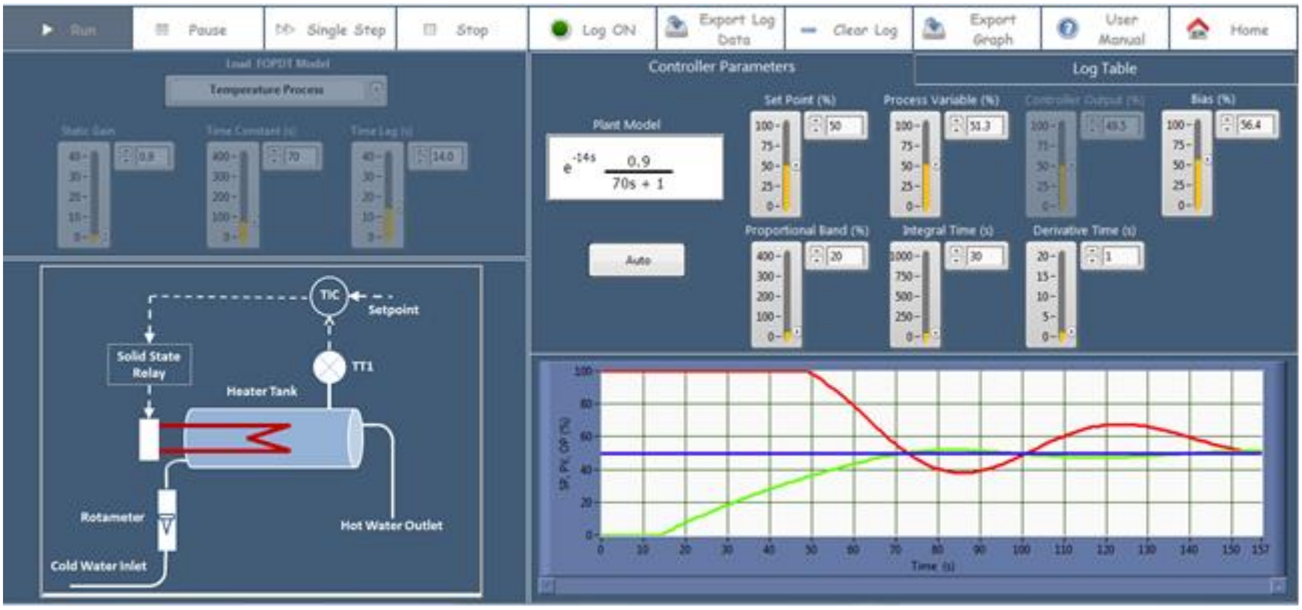


Fig 10. Simulator response for a setpoint change

Model based simulator response for a setpoint change of 40% to 50% is shown in figure 9. It can be observed that the response shown in figure 9 and figure 10 are closely matching. It can be concluded that the simulator can be used to understand the response of the actual plant and can also be used to study the impact of PID parameters. The simulator also has Auto / Manual provision to understand open loop and closed loop response of the plant. It also has a pause option to freeze the execution and a single step input to execute the simulation for one iteration. The simulation interval (sampling time) is internally set as 1 sec.

As a drawback of the simulator, the actual plant response starts from room temperature of the fluid, whereas the simulator response starts from 0%. It can be modified by allowing a user input of room temperature as an additional input. The complete view as a screenshot of the simulator in execution mode is shown in figure 11.



Static Gain: 0.9

Time Constant (s): 70

Time Lag (s): 14.0

Plant Model: $e^{-14s}\frac{0.9}{70s+1}$

Set Point (%): 50

Process Variable (%): 51.3

Controller Output (%): 49.5

Bias (%): 56.4

Proportional Band (%): 20

Integral Time (s): 30

Derivative Time (s): 1

Fig 11. Screenshot of temperature process simulator

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