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Design PI Controller for Tank Level in Industrial Process

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Abstract
In today's chemical, refinery, and petrochemical sectors, separation tanks are one of the most significant separating processes. One or more separation tanks must operate consistently and reliably for multiple facilities' safe and efficient operation. Therefore, in this paper, a PI controller unit has been designed to improve the performance of the tank level controller of the industrial process in Basrah Refinery Station. The overall system mathematical model has been derived and simulated by MATLAB to evaluate the performance. Further, to improve the performance of the tank level controller, optimal PI parameters should be calculated, which Closed-Loop PID Autotuner has been used for this task. Several experiments have been conducted to evaluate the performance of the proposed system. The results indicated that the PI controller based on the Autotuner Method is superior to the conventional PI controller in terms of ease to implement and configuration also less time to get optimal PI gains.

KEYWORDS: Tank level controller, Industrial process, Refinery Unit, PID controller, PID Autotuner.

I. INTRODUCTION

Tank level control is one of the most common applications in the industry [1-3]. In industry, several approaches and techniques for level control are used. It's necessary to keep track of liquid levels in process tanks [4]. Mass flow rate regulation is utilized in a range of technological applications, including steam generators in power plants, reactors in several chemical industries, and storage tanks in the oil and gas industry [5]. The tanks were sized based on the fluid flow characteristics and rates that would be observed.

The liquid-phase hydrocarbon was the subject of research, and it was split into three stages. A two-phase separator, which separates fluid into gas and liquid phases, was used in the first step [6]. A two-phase vapor separator is a device that separates vapor from a vapor-liquid combination using a density gradient. One of the most important input facilities for oil refineries is phase separators. The combined-head streams' combined gaseous and liquid components are separated as part of the initial petroleum refining [7]. Such separations are generally caused by physical property changes in the fluid stream. There are two or more phases with different compositions in heterogeneous mixtures. Chemical reactions do not occur between the components of these mixtures, and the stages' borders are well defined. One of the drawbacks of the flash or separation process within production units is that it is extremely sensitive to input disturbances: temperature, concentration, level, and mass flow in the feed current all have a significant impact on flash separation efficiency. In this case, a thorough control system must be established to decline the effects of stream vapor or liquid disruption. Level (liquid phase) and pressure (vapor phase) are directly affected by changes in separation variation (how much liquid and vapor flow from intake valve to flash tank) [8].

In refining and gas processing, the separation of liquids and particles from a gas stream is critical. The two controllers can keep the level and temperature (and indirectly the pressure) around the intended levels during normal operation, assuming that process equipment such as pressure sensors and valves are working properly [9]. However, several fault conditions may result in an unsafe situation in which excessively high pressure in the drum is obtained (possible causes of such unsafe conditions could include faults in the top vapor effluent valve and the bottom liquid effluent valve that cause them to close) [10].

Recontacting Drum D5204(RD) is a separator vessel used in Iraq's Basrah Refinery to boost gasoline output. Because the liquid leaving the tank towards the bottom comprises a substantial amount of (reformate, LPG, and off-gas), liquid control is critical for this drum to retain these components available in the drum for other operations.

Before entering RD, the Recontacted combination process (vapor and liquid) is chilled with cooling water in the RD Cooler.

Two PI controllers control the liquid level and pressure in the flash drum, and this control system also includes a pressure relief valve. The three PID controllers have distinct actions in the process. The proportional controller (Kp) reduced rise time reduces but does not eliminate steady-state error. Integral control (KI) eliminates steady-state error but may exacerbate abrupt response. A derivative control (KD) improves system stability, reduces overshoot, and improves transient responsiveness. When the valve regulating the outlet vapor stream from the drum fails, modifying the tuning parameters of one of the other PI controllers while the safety system is activated improves closed-loop performance compared to the case where the tuning parameters of that PI controller remain the same.

Despite the evolution of more advanced control techniques, proportional-integral-derivative (PID) controllers are still widely used in the process industries [11]. PI controllers are often used to control industrial process variables (e.g., position, speed, current, temperature, pressure, humidity, and level); in reality, the derivative section is usually turned off due to measurement noise. For pressure and level control in gas tanks, a PI controller is sufficient. Adding the derivative element, on the other hand, improves closed-loop stability and shortens the time it takes for the step response to climb [12]. Proportional, integral, and derivative control are the three components of PID control, and determining the ideal value for these parameters without knowing anything about the plant is difficult. Throughout the previous many decades, various open and closed-loop-based tuning procedures have been submitted. All tuning methods demonstrate how to find the optimal PID parameter value [13,14].

Tuning processes have the same goal which is to find PID settings that allow the plant to overshoot less, settle faster, and be more robust to disturbances. Various strategies for tuning PID controllers for integrating systems with time delay have been presented in the literature. S. K. Pandey et al. were offered an auto-tuning algorithm for PID control parameters[13]. A simple PID auto-tuning algorithm is developed to implement and be applicable for the heating and cooling process. N.-S. Pai et al. have been suggested a calculation method of a practical PI/PID controller tuning for integrating processes with dead time and inverse response based on a model [15]. Analytical expressions for PI/PID controller settings based on the model using a direct synthesis method for disturbance rejection (DS-d). when F. Padula et al were presented tuning rules for integer-order and fractional-order PID controllers [16]. The tuning rules have been devised to minimize the integrated absolute error with a constraint on the maximum sensitivity. In 2016, K. Amoura et al. offered an experiential method for tuning a new type of fractional controller known as PID-Fractional-Order-Filter (FOF-PID) [17]. Furthermore, D. Castellanos-Cárdenas et al. have been submitted an IMC-based PID tuning method for inverse-response second-order plus, dead time systems [18]. The tuning rules are based on the optimization of an objective function that combines performance and robustness. Similarly, L. has been suggested a new proportional—integral—derivative (PID) controller auto-tuners using frequency sampling filters (FSFs) for the estimation of plant frequency response information under relay feedback control [19]. Additionally, K. Sinthipsomboon et al. have formulated a hybrid of fuzzy and fuzzy self-tuning PID controller for motor speed control of a SEHS. The described control technique consists of two components: a fuzzy controller and a fuzzy self-tuning PID controller [20]. Despite various approaches available to develop PID controllers for integrated systems, a review of the literature suggests that there is still room to enhance the performance and durability of these controllers. For the regulation of integrating processes, many authors presented a complicated structure with a large number of controllers.

In this paper, a part of the operation process represented by the "Enhancing Gasoline Production" unit has been studied that is located in Basrah Refinery, South of Iraq to improve the performance of this system that already has a PI controller but with manual tuning. This operation process is performed by a separator drum, in which a mathematical model of this drum is derived and then the performance of the process operation is evaluated by MATLAB. PID controller has been designed to improve the performance of the system, in which optimal PID parameters have been evaluated to achieve the best performance. Two methods are used to evaluate the PID parameters, the first one by Ziegler and Nichols method [21] and the second one by Autotuning method with new philosophy obtain. The proffered system is simulated by MATLAB, and the results show the response has less overshoot, faster settle time, and is more resistant to disturbances.

This paper is organized as follows: In section II, there is a description Enhancing Gasoline Production unit in Basrah Refinery. Section III presents the modeling of the separator drum system and PID controller auto-tuning. Section IV will show experiments and results for several scenarios in the case of manual and auto PID tuning. Finally, in Section V, conclusions have been presented.

II. Background

South Refineries Company was founded in 1969 through the establishment of Basrah Refinery, which began production in 1974 by establishing (Refining Unit No. 1), which is mentioned in Fig. 1 which is one of the major manufacturing units in the country where the production of oil derivatives using the latest scientific methods and advanced technology in production, which resemble their high-quality products, including foreign and meets the requirements of consumers. And continue to modernize and expand the company diversify their products and improve quality, it has been done by the establishment of a refinery and the second unit to improve gasoline and oil refinery [22]. Tasks of the company are:

1) Product the following: (Gasoline, Kerosene, Light gas oil, diesel, Fuel oil, marine fuel oil, LPG, jet fuel)
2) Product oils such as (Base lube oil grade 30, Pale 600k, Spindle oil, Bright Stock)
3) Plastic cans (capacity 1 liter, 5 liters), iron drums (200 liters).

The Naphtha HDS with Stabilization and Splitting Unit (U7501) is designed for feedstock preparation for newly built Catalytic Reforming and Isomerization Units within Basrah Refinery.

Naphtha feed from U7501 to a Catalytic Reforming unit typically contains C6 to C11 paraffin, naphthenic, and aromatics. The purpose of this reforming process is to produce high-octane aromatics from naphthenic and paraffin for use as a high-octane gasoline blending component.

The D5204 vapor phase constitutes hydrogen production. Part of the hydrogen production upstream of the pressure control valve is used in the naphtha hydrotreatment unit as make-up that is all noticed in Fig. 2. The remaining hydrogen-rich gas is routed to the hydrogen-rich gas network at 24 bar g. The separated liquid from RD D5204 is sent under level control to the stabilization section. The work preview descriptor for RD D5204 emphasizes the importance of strong performance management for levels and pressures to provide optimal functioning.

Fig. 1: UNIT 7501 in Basrah Refinery.

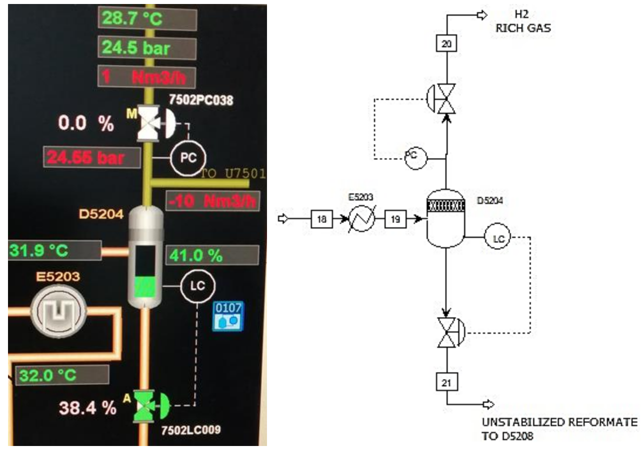


Fig. 2: DRUM D5204 schematic diagram.

III. Mathematical Model

Process modeling is a useful tool for doing process synthesis, analysis, or operational optimization. There are three types of models: black box, white box, and gray box models. In this essay, the first and second classes are not discussed. It is because experimental models are dependent on data that they lack the specialized process knowledge that they are called " experimental ". Physical models and white boxes are utilized because they demand thorough comprehension of the laws and theories guiding all of the actions involved. Semi physical models based on mass and energy balances are used here to simulate the flash process.

A. Mathematical model for RD D5204 (separator or flash drum).

By using the flash drum (separator drum) expressed in Fig. 3 that has the following specification data:

Item Tag: 7502-D5204. The mixture is available at (25) bar, (45) ^oC. At steady state, the molar inlet is (1159.01) kmole/h or 48141 kg/h and gravity 759 kg/m^3. This stream will be in two-phase the liquid by 85% and vapor by 15% with pressure (25). The physical separator occurs in a flash tank of 5.2 m in height and 1.6 m in diameter. Liquid-vapor equilibrium consolidates in the flash drum. Because there is enough difference between the relative volatility, the current liquid that leaves the tank for the bottom contains a large percentage of (reformate + LPG + off-gas) at pressure 25.5 bar and gravity 834.9 kg/m^3 also from top leaves rich gas (65-70 %) H2 + C1, C2, C3 … at pressure 25 bar.

To apply the conservation principle to all determined Process systems. Over each PS, it is advised to do the following balances: one total mass balance, n component mass balance, and total energy balance. When there are significant pressure or density variations in the process, momentum balancing is utilized. The Dynamics Balance Equations (DBE) are balance equations that assume that all balances are originally dynamic. If the process contains certain static characteristics, some of these can be converted to static equations. The mass and energy balance is established in the flash process being discussed here [8].

Fig. 3: D5204 in Basrah Refinery.

Balance over Total Process System (PS total):

- Total mass balance:

$$\frac{dM}{dt} = \dot{m_{2V}} + \dot{m_{2L}} - \dot{m}_3 - \dot{m}_4 \quad (1)$$

M: the total mass, $\dot{m_i}$: the mass flow of current *i*.
With subscript V for vapor and L for liquid in multiphase currents.

- Energy balance:

$$\frac{dQ}{dt} - \frac{dW}{dt} + \dot{m_{2L}} * \widehat{H_{2L}} + \dot{m_{2V}} * \widehat{H_{2V}} - \dot{m}_3 * \widehat{H_3} - \dot{m}_4 * \widehat{H_4} = \frac{dE_T}{dt} \quad (2)$$

With $\frac{dQ}{dt}$ the heat flow and $\frac{dW}{dt}$ the workflow of the system, $\widehat{H_i}$ the specific enthalpy of current ET the total energy of total PS.

- Balance over Liquid Process System (PS liquid):
  - Total mass balance:

$$\frac{dL_L}{dt} = \frac{1}{\rho L*\pi*r^2} * (\dot{m_{2L}} - \dot{m}_4) \quad (3)$$

With L_L the liquid level in the tank, ρ_i density of phase *i* (liquid or vapor), and r radius of the tank.

- Balance over Vapor Process System (PS vapor):
  - Total mass balance:

$$\frac{dP}{dt} = \frac{1}{V_V} * \left(\frac{R*T}{M_{mV}} * \dot{m_{2V}} - \frac{R*T}{M_{mV}} * \dot{m}_3 - P * \frac{dV_V}{dt}\right) \quad (4)$$

With P the pressure in the vapor phase inside the tank, R the universal gas constant the operating temperature, M_{mV} the molecular mass of vapor, and V_V the volume of vapor phase inside the flash tank.

Liquid or vapor flow through output flash tank valve, required to calculate $\dot{m}_3$ and $\dot{m}_4$, vapor and liquid flows exiting the flash tank:

$$\dot{m_i} = Cv_i * \frac{\gamma_i}{100} * \sqrt{\Delta P_i * \rho_i} \quad (5)$$

With Cv_i the coefficient of valve for phase i, γ_i the valve opining percentage, and ΔP_i the pressure drop through the valve acting over phase i.In valve sizing w_i is taken equal to 50% for nominal or design flow.

B. Mathematical model for Autotuning PID Controller.

A PID controller consists of three terms: the proportional (P) term, the integral (I) term, and the derivative (D) term. In an ideal form, the output u(t) of a PID controller is the sum of the three terms,

$$u(t) = K_c e(t) + \frac{K_c}{\tau_1}\int_0^t e(\tau)d\tau + K_c\tau_D\frac{de(t)}{dt} \quad (6)$$

where e(t) = r(t) − y(t) is the feedback error signal between the reference signal r(t) and the output y(t), and τD is the derivative control gain. The Laplace transfer function of the PID controller is

$$C(s) = \frac{U(s)}{E(S)} = K_c\left(1 + \frac{1}{\tau_I s} + \tau_D S\right) \quad (7)$$

The P action (mode) adjusts the controller output dependent on the error magnitude. An I action (mode) may eliminate the steady-state offset, but the D action anticipates the future trend (mode) (mode). These helpful functions are adequate for a broad variety of process applications, and the features' transparency leads to widespread user acceptance [23].

To design the PID controller, can assume that two estimated frequency response points from the relay testing data are the fundamental frequency Gp(jω1), $\omega_1 = 2\pi/N\Delta t$ and Gp(jω_2). If there is no disturbance in the system, then ω2 is selected as $3\omega_1$.Otherwise, ω_2 is selected as $2\omega_1$.

Converting the ideal PID structure into another polynomial form:

$$c(s) = \frac{c_2 s^2 + c_1 s + c_0}{s} \quad (8)$$

where Kc = c1, $\tau_I = c_1/c_0$ and $\tau_D = c_2/c_1$, the PID controller design using the estimated frequency response is to find the parameters c_2, c_1 and c_0.

Automatically adjust PID gains in real time depending on the expected plant frequency response from a closed-loop experiment. Utilize the Frequency Response Estimator to do a real-time experiment-based estimate on a physical plant. To acquire a frequency response estimate.

- Injects sinusoidal test signals into the plant at the nominal operating point.
- Collects response data from the plant output.
- Computes the estimated frequency response.

A recursive least squares (RLS) algorithm [24-26] to compute the estimated frequency response. Assume that the plant frequency response is $G(j\omega) = \gamma\angle j\theta$. When a signal $u(t) = A\sin(\omega t)$ excites the plant, the steady-state plant is $y(t) = A\gamma\sin(\omega t + \theta)$, which is equivalent to:
$y(t) = (\gamma cos\theta)Asin(\omega t) + (\gamma sin\theta)Acos(\omega t)$
At any given time, $Asin(\omega t)$ and $Acos(\omega t)$ are known, therefore, they can be used as regresses in an RLS algorithm to estimate $\gamma\cos(\theta)$ and $\gamma\sin(\theta)$ from the measured plant output y(t) at run time.

When the excitation signal contains a superposition of multiple signals, then:
$u(t) = A_1\sin(\omega_1 t) + A_2\sin(\omega_2 t) + \cdots$

In this case, the plant output becomes:
$y(t) = (\gamma_1 cos\theta_1)A_1\sin(\omega_1 t) + (\gamma_1\sin\theta_1)A_1\cos(\omega_1 t) + \cdots$

$(\gamma_2 cos\theta_2)A_2\sin(\omega_2 t) + (\gamma_2 sin\theta_2)A_2\cos(\omega_2 t) + \cdots$

The estimation algorithm uses $A_i \sin(\omega_i t)$ and $A_i \cos(\omega_i t)$ as regressors to estimate $\gamma_i cos\theta_i$ and $\gamma_i sin\theta_i$. For N frequencies, the algorithm uses 2N regresses. The computation assumes that the perturbation signal u(t) is applied to a plant with zero nominal input and output. To achieve this condition, the block subtracts from the measured plant input and output signals their values measured at the start of the experiment.

IV. EXPERIMENTS AND RESULTS

This section will concentrate on building systems in MATLAB Simlink, and work will be based on simulations of real industrial systems and the results of numerous experiments, so we will first evaluate the system's closed-loop performance without any control, then change configuration with a new PI control and finally offered an auto-tuner method by adding new blocks to the system to opportunistically optimize the system.
initiated many experiments that tested control on tank D5204 illustrated in Fig. 4 with its below specification:

Height:	5,200 mm tang to tang.
Dimeter:	1,600 mm
Total volume:	11.5 m^3
Operation pressure:	25 bar
Operation temperature:	45 ℃
Liquid density at 45 ℃:	759 kg/m^3
Liquid operation density for inlet stream:	650,52 kg/m^3
Liquid operation density for output stream:	722,8 kg/m^3
The Coefficient of level control valve:	cvi=44.6
The pressure drop through the valve acting:	10 bar
Inlet liquid flow rate to tank m^2:	600 Kmole/h
Valve opening in % yi:	50%
Desired level set point:	40%

From equations (3 and 5), the design of the system model will be implemented by MATLAB software, in which experiments will have divided into several scenarios:

1) Study the response of system without controller under factors effect (inlet feed and opening output valve)
2) Study the response of the system with PI control (manual tuning PI parameters) under factors effect (inlet feed and opening valve).
3) Study the system with PI control (auto-tuning PI parameters) under factors effect (inlet feed and opening valve).

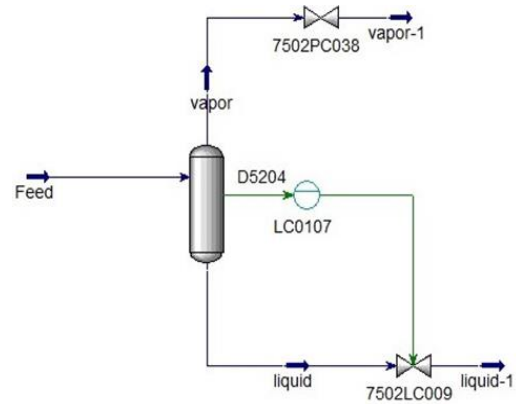


Fig. 4: graphical presented for model.

First Scenario: Without Controller

Simulink models of the tank are illustrated in Fig. 5. The system has been provided with a constant reference level with a predefined peak value, as well as an input feed for the separator tank, which we specified with a variable value (m2). The performance of the system is presented in Fig. 6 (*the curves illustrates both the desired set point level (blue trace) and the simulated actual level (red trace) as functions of time*) and its characteristics are listed in Table 1. Clearly, the results (rise time and overshoot values) show that the system is not controlled or acceptable to manage the plant. Due to the system's primary attribute being non-linearity, it was important to analyze the change in level using both the integrator and the other system components. All of these features have the potential to cause disruptions if they are lost or diminished for whatever reason.

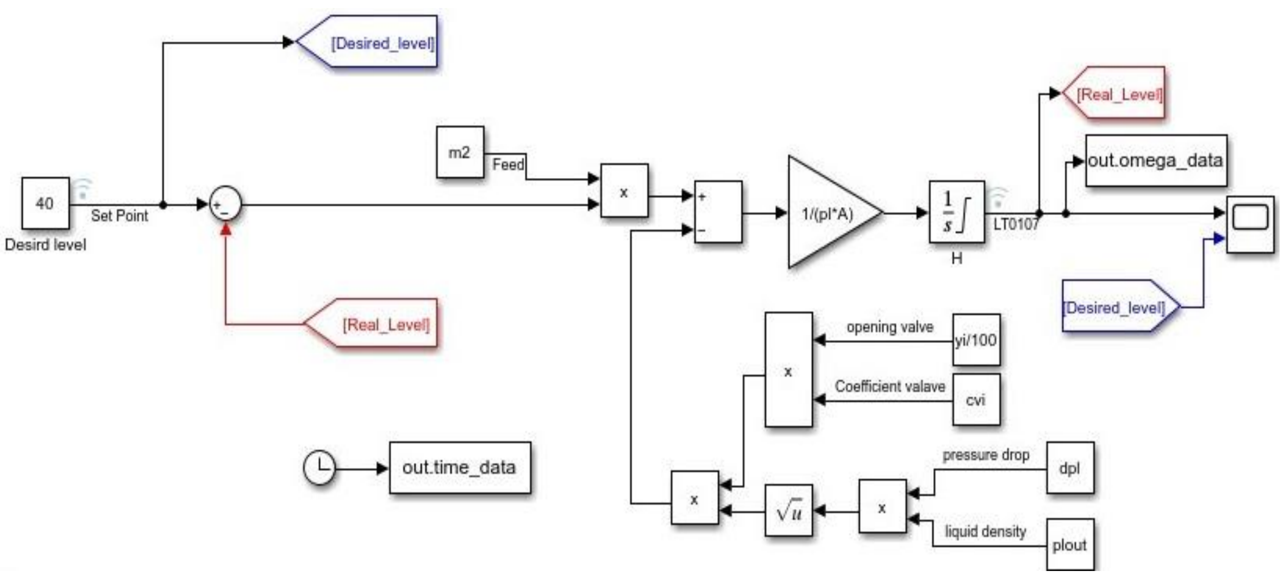


Fig. 5: System diagram without a controller.

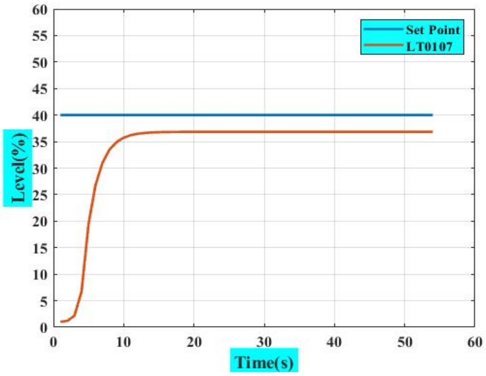


Fig. 6: System response of the plant without controller.

TABLE 1

STEP-RESPONSE CHARACTERISTICS FOR SYSTEM IN FIG. 5

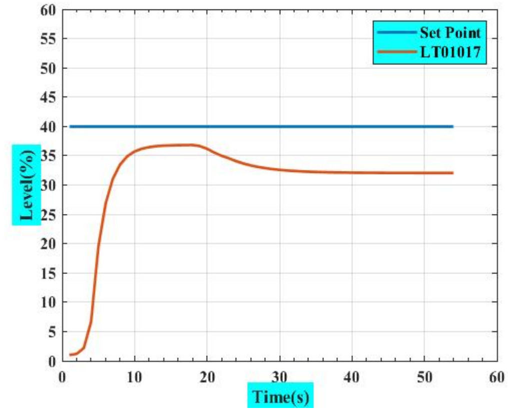
Rise Time	3.5862
Transient Time	8.5391
Settling Time	8.5149
SettlingMin	33.5413
SettlingMax	36.8369
Overshoot	3.7767e-05 %
Undershoot	0
Peak	36.8369
Peak Time	12

Certain problems occur often in refineries for a number of reasons such as lower input supply, valve failures, and so on. Due to this failure occurring immediately, reducing the situational feed rate to 250 kg/h while the system is operating causes the desired level to deviate from the real level, which will prevent the system from meeting the required performance that ultimately affect things all product specifications. Another aspect affecting system performance is the output control valve's precision; hence, any change or failure in this component would impair system responsiveness as illustrated in Fig. 7.

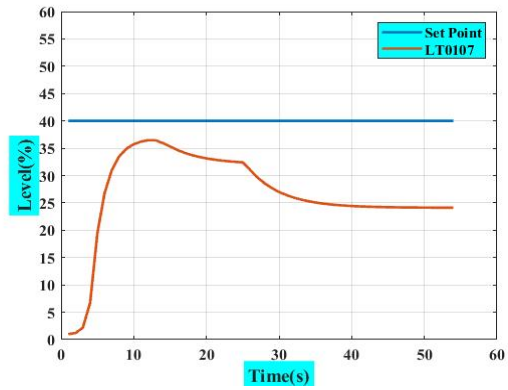
Second Scenario: With PI Controller (Manual Tuning)

A PI controller has been used to improve the performance of the system as recognized in Fig. 8. A trial and error approach will be used to determine the optimal parameters for the PIs. This technique is frequently employed in almost every industrial application, and it is quite successful.

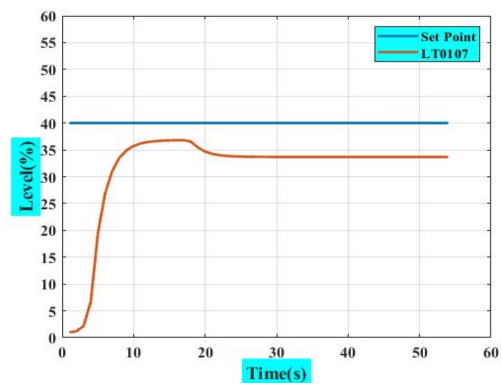
Due to several challenges, including the requirement for an instrumentation and control engineer with extensive experience working with online operating controllers and knowledge of reading and observing system responses when PI parameter values are changed, this operation in the real process takes between 4 and 6 hours to complete after adding the PI controller and simulating the system in the reallocation (Basrah Refinery) to obtain the PI parameters. It required several simulations with various P and I value to produce the system response displayed below. The results are illustrated in Fig. 9.



(a) Decrease m2 to 250 kmole/h



(b) Decrease m2 to 250 kmole and yi to 100%



(c) Increase yi to 100%

Fig. 7: system response after some disturbance.

After adjusting the PI controller, the system's response characteristics are depicted in Table 2, along with its reaction to disturbances such as altering the input feed m2 or opening valve value and recording the system's response to these changes.

Overshoot and rising time are the most important components to consider when examining the system's response and can evaluate the overall performance of the system based on the results, which suggest a high value for overshoot and other parameters in the system. Figure 9 shows the response of the system simulation with the PI controller, while Fig. 10 shows the performance of the system with the PI controller for several values of feed inlet (m2) and outlet valve control. Table 3 clearly expresses changes in values especially overshoot and transient time, in which using PI controller has improved the performance of the system.

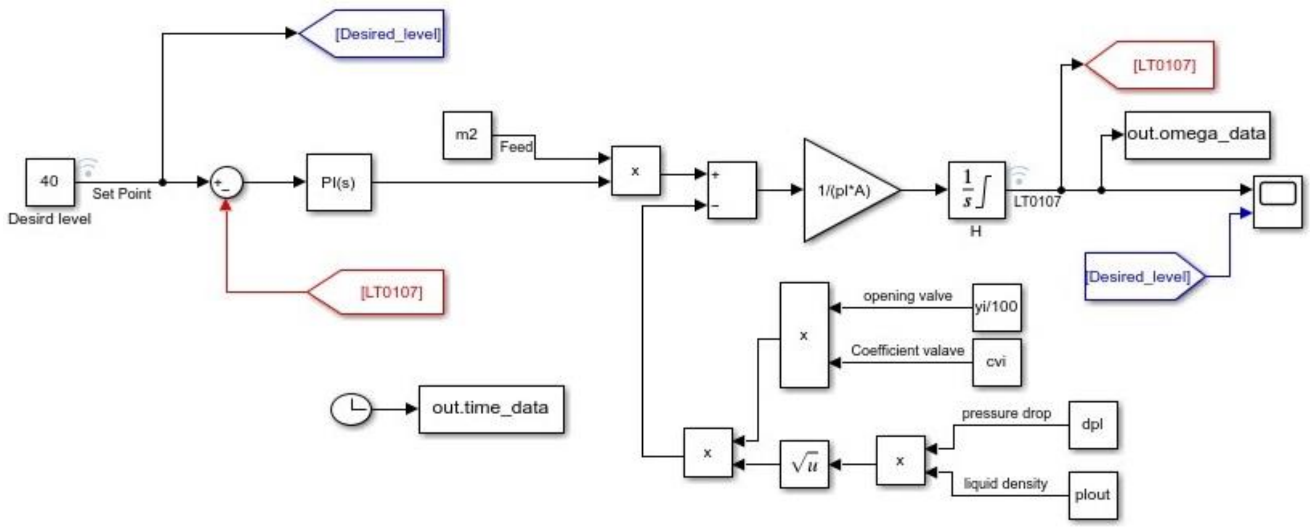


Fig. 8: System diagram with PI controller.

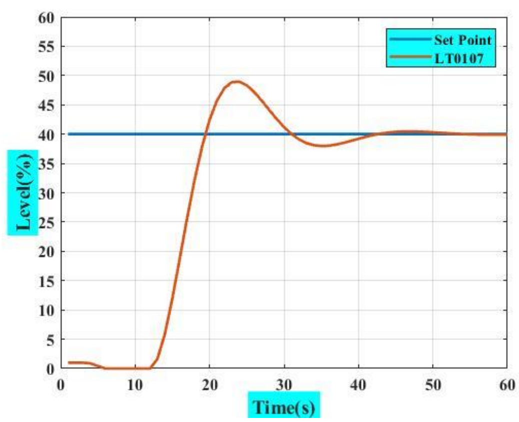
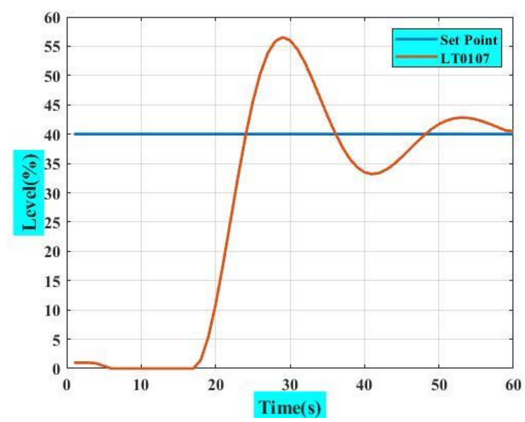
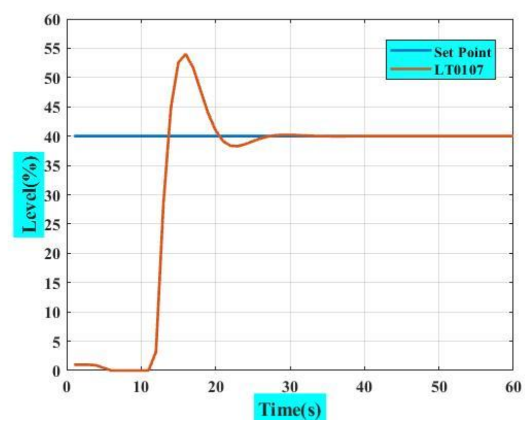


Fig. 9: System response of the plant with PI controller (manual tuning).



(a) Decrease m2 to 250 kmole/h



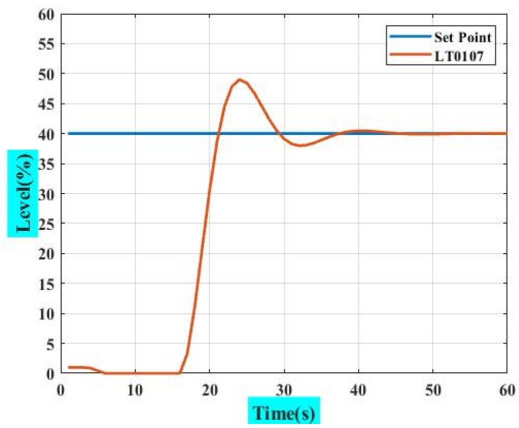
(b) Increase m2 to 1000 kmole/h

Fig. 10: System response in case of several values of m2 and yi.

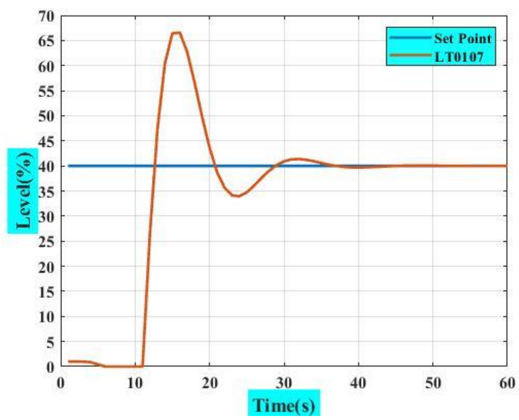
TABLE 2

STEP-RESPONSE CHARACTERISTICS OF SYSTEM IN Fig. 9

Rise Time	3.5769
Transient Time	30.5807
Settling Time	30.5807
SettlingMin	37.9758
SettlingMax	49.0099
Overshoot	22.5402%
Undershoot	0
Peak	49.0099
Peak Time	19



(c) Increase yi to 100%.



(d) Decrease yi to 0%.
Fig. 10: Continued.

TABLE 3
STEP-RESPONSE CHARACTERISTICS FOR SYSTEM IN FIG. 10.

Parameter	m2 = 250 kmole/h	m2 = 1000 kmole/h	yi = 100%	yi = 0%
Rise Time	4.9	1.42	3.57	1.03
Transient Time	57.46	25.15	35.61	35.49
Settling Time	57.46	25.15	35.61	35.49
SettlingMin	33.11	38.28	37.97	32.23
SettlingMax	56.61	53.98	49.1	74.56
Overshoot	38.98%	34.96%	22.5%	86.46%
Undershoot	0	0	0	0
Peak	56.61	53.98	49.1	74.56
Peak Time	29	16	24	16

Third Scenario: PI Controller (Autotuning)

The PI Autotuner controller has been intended to improve the performance of the system. The modified system has been depicted in Fig. 9, in which the Closed-Loop PID Autotuner block has been used for this purpose, therefore, get the system shown in Fig. 11. The parameters of this block should be set as follow:

- **Target bandwidth:** Determines how fast you want the controller to respond. The target bandwidth is roughly 2/desired rise time. For a desired rise time of 4 seconds, set target bandwidth = 2/4 = 0.5 rad/s.
- **Target phase margin:** Determines how robust you want the controller to be. In this example, start with the default value of 75 degrees.
- **Experiment sample time:** Sample time for the experiment performed by the Autotuner block. Use the recommended 0.02/bandwidth for sample time = 0.02/0.5 = 0.04s.

The Experiment tab has three main experiment settings:

- **Plant Type:** Specifies whether the plant is asymptotically stable or integrating. In this example, the separator drum D5204 System plant is integrating.
- **Plant Sign:** Specifies whether the plant has a positive or negative sign. The plant sign is positive if a positive change in the plant input at the nominal operating point results in a positive change in the plant output when the plant reaches a new steady state. In this example a positive plant sign.
- **Sine Amplitudes:** Specifies amplitudes of the injected sine wave perturbations. In this example, specify a sine amplitude of 0.3.

Start the experiment at 140 secs to ensure that the martial level has reached steady-state H=40. The recommended experiment duration is 200, bandwidth = 200/0.4 = 500 sec. With a start time of 140 secs, the stop time is 640 sec. The simulation stop time is further increased to capture the full experiment.

During simulation, the Closed-Loop PID Autotuner block performed the following:

1) Injects a test signal into the plant to collect plant input-output data and estimate frequency response in real-time. The test signal is a combination of sinusoidal perturbation signals added on top of the plant input.
2) At the end of the experiment tunes PID controller parameters are based on estimated plant frequency responses near the target bandwidth.
3) Updates a PID Controller block or a custom PID controller with the tuned parameters, allowing to validate closed-loop performance in real-time.

The Autotuning parameters strategy has the significant advantage of taking a short time to calculate optimal PI parameters and improving other characteristics like rising time and overshoot that are presented in Fig. 12. The testing control system with the same disturbance that is taken in the previous scenarios and behavior system against it clear appear in Flappers and Tables 5. From the results can be seen that the performance in the case of using the PID Autotuning is superior to the other methods.

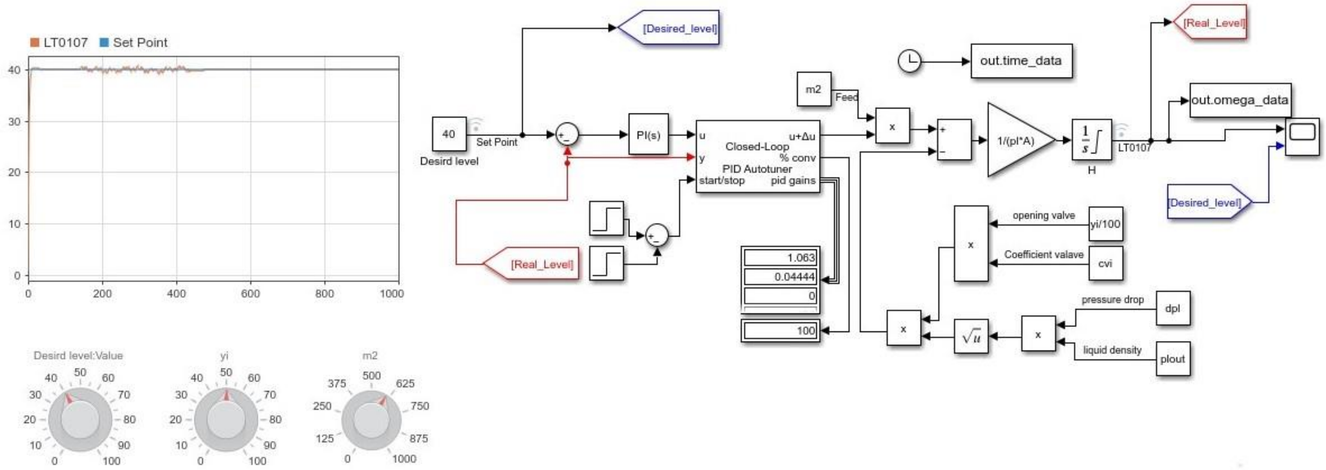


Fig. 11: System diagram with Closed-Loop Autotuning PID controller.

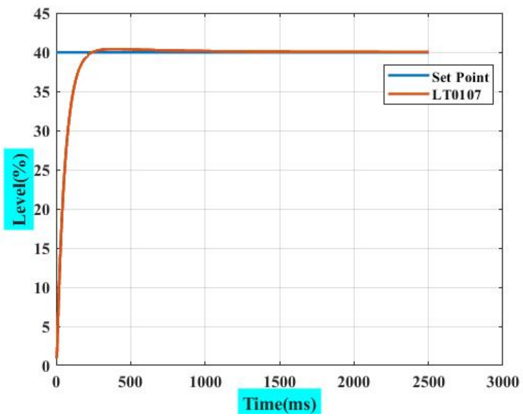
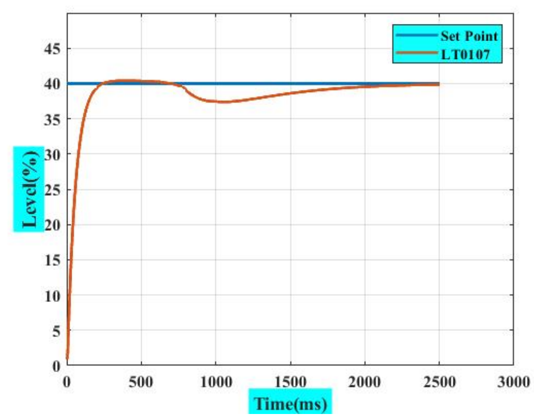


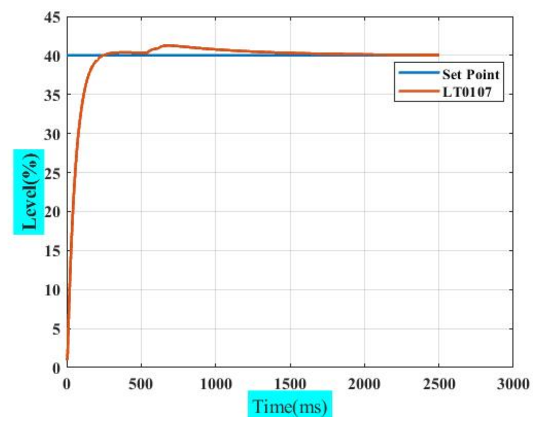
Fig. 12: System response of the plant with PI controller (auto tuning)

TABLE 4
STEP-RESPONSE CHARACTERISTICS FOR SYSTEM IN FIG. 11.

Rise Time	116.5438
Transient Time	194.4419
Settling Time	193.6038
SettlingMin	36.0012
SettlingMax	40.3922
Overshoot	0.9806 %
Undershoot	0
Peak	40.3922
Peak Time	388

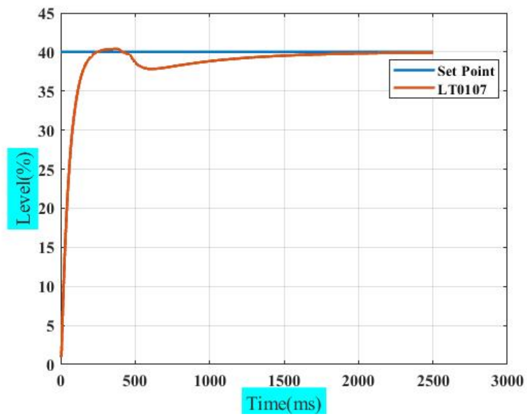


(a) Decrease m2 to 250 kmole/h

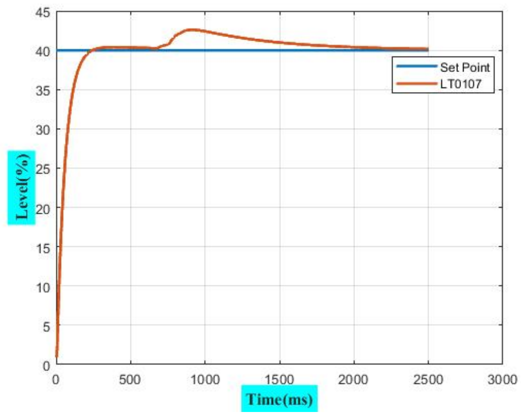


(b) Increase m2 to 1000 kmole/h.

Fig. 13: Response of system with Closed-Loop Autotuning PID controller in case of several values of m2 and yi.



(c) Increase yi to 100%



(d) Decrease yi to zero%

Fig. 13: Continued.

TABLE 5

STEP-RESPONSE CHARACTERISTICS FOR SYSTEM IN FIG. 13

Parameter	m2 = 250 kmole/h	m2=1000 kmole/h	yi=100%	yi=0%
Rise Time	116.54	116.54	116.54	116.54
Transient Time	2637.5	194.44	1642.7	5780
Settling Time	2626.3	193.60	1629	5766.1
SettlingMin	36.00	36.00	36.00	36.00
SettlingMax	41.04	40.72	40.39	41.79
Overshoot	0.98%	1.82%	0.98%	4.47%
Undershoot	0	0	0	0
Peak	40.39	40.72	40.39	41.79
Peak Time	388	6193	388	4923

V. CONCLUSION

The level controller has been designed for nonlinear Separator Drum D5204 based on PI controller, in which the PI parameters have been optimized by manual tuning and Autotuning method. The offered system has been modeled and then simulated by MATLAB R2021b based on real system characteristics. Several scenarios have been implemented for the experiments to validate the recommended system in the presence of some common disturbances or changes in industrial applications, such as reducing feed inlet to the drum or failure with either the inlet or outlet control valve, and observed how the control system responded to these disturbances. The results show, the performance has improved of the system in the case of using a PI controller, further, the system resists sudden disturbances in case of using the PI controller. In the case of using PID controller as PI controller, the performance of the system is superior to the other methods in respect to more resistance to disturbances cases and others such as easy to implementation and configuration as well less time to get best values for PI parameters.

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CONFLICT OF INTEREST

The authors have no conflict of relevant interest to this article.

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