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Liquid Level Control Performance Study of Conventional and Advanced Model-Based Controller for a Quadruple Tank System

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ABSTRACT

Liquid level control in a quadruple tank system (QTS) is crucial for various industrial applications due to its multivariable nature. Conventional controllers frequently encounter difficulties maintaining performance in complex applications, showing transient and steady-state response difficulties. This study evaluates and compares the effectiveness of Model Predictive Control (MPC) and Proportional-Integral-Derivative (PID) controller in regulating the liquid level of a quadruple tank system to improve control precision, stability and overall performance. The investigation utilized controller tuning to evaluate performance metrics across different operational scenarios and constraints, including rise time, settling time, and overshoot. The results demonstrate that MPC has superseded performance compared to PID controller, resulting in faster response time, shorter settling time, and decreased overshoot from the desired set point. As a result, constrained MPC achieved a 15% decrease in overshoot and a 20% decrease in settling time compared to the PID controller. To validate the robustness of MPC, set point changes were applied compared with PID controller performance metrics. MPC transitioned smoothly between 12 and 18 cm targets without overshoot and steady-state error. In contrast, PID incurred a persistent 0.5 cm error. This research contributes to the fact that MPC is a more effective method of controlling QTS. MPC demonstrates better performance in both transient and steady-state response.

Keywords: Liquid Level Control, Quadruple Tank System, PID controller, Model Predictive Control

1. INTRODUCTION

Coupled tank system serve as important process industry applications for investigating dynamic behaviour, such as in Abbas *et al.* [1]. The quadruple tank system (QTS) exhibits complex interactions between interconnected that require advanced control approaches to manage. Controller design must consider multivariable properties derived from the QTS linearized model in Johansson [2]. Tank level behaviour results from input-dependent relationships between elements in this multiple-input multiple-output (MIMO) controlled plant in Navratil *et al.* [3]. The QTS consists of four tanks with multiple pumps that simultaneously control liquid levels. Used widely in control education, such as in Numsomran et al. [4] and Pedroso et al. [5], the QTS configuration was created by combining two double-tank setups.

Implementing control strategies is important in directing the QTS towards its desired outcomes, guaranteeing stability, accuracy, and responsiveness. This study aims to examine conventional and advanced controller for their adaptability and efficiency in this dynamic environment. The insight gained can enhance control strategy comprehension. Conventional PID controllers have been employed in industry due to their well-established framework.

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Despite their success in several studies, Pedroso *et al.* [5] and Lakshmi *et al.* [6]. PID have inherent limitations, particularly in handling complex and dynamic systems with nonlinearities. This has led to the exploration and integration of advanced control methods, with MPC standing out as an alternative Özkan *et al.* [7]. Hence, the objective of this research is to develop advanced MPC controllers and conventional PID controllers to accurately monitor and control the liquid levels in the two bottom tanks of the QTS, meeting an optimal reference level profile while addressing the challenges posed by the system's inherent nonlinearity and multivariable nature.

2. METHODOLOGY

2.1 Development of System Equations for Each Tank

Figure 1 depicts the setup of the cylindrical tank used in this study. A first-principles modelling approach was undertaken to systematically derive the governing dynamical equations of the QTS. Non-linear differential equations describing the liquid level behavior in each tank volume were formulated based on fundamental conservation principles. Bernoulli's equation relating head, flow velocity, and pressure was leveraged to establish the inter-tank outflow relationship.

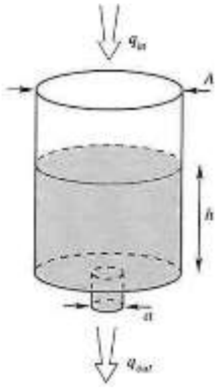


Figure 1. Schematic of a single tank with relevant geometric and hydraulic parameter represented

The nonlinear equations were simplified using the Taylor series approximation to create a model capturing system dynamics. Analysis of zero dynamics revealed coupling behaviors between volumes. This method provided a foundation for modal analysis and investigations into set point regulation capabilities for the quadruple tank plant. The resulting structure simulated benchmark process operations and validated control strategies for optimal liquid distribution. Mathematical representations for all tank subsystems were derived from Eqn. (1) to (4)

$$\frac{dh_1}{dt} = -\frac{a_1}{A_1}\sqrt{2gh_1} + \frac{a_3}{A_1}\sqrt{2gh_3} + \frac{\gamma_1 k_1}{A_1} u_1 \quad (1)$$

$$\frac{dh_2}{dt} = -\frac{a_2}{A_2}\sqrt{2gh_2} + \frac{a_4}{A_1}\sqrt{2gh_4} + \frac{\gamma_2 k_2}{A_2} u_2 \quad (2)$$

$$\frac{dh_3}{dt} = -\frac{a_3}{A_3}\sqrt{2gh_3} + \frac{(1-\gamma_2)k_2}{A_3} u_2 \quad (3)$$

$$\frac{dh_4}{dt} = -\frac{a_4}{A_4}\sqrt{2gh_4} + \frac{(1-\gamma_1)k_1}{A_4}u_1 \tag{4}$$

2.2 Generation of Simulation Data Using Transfer Function

A Simulink-based methodology was developed to mathematically represent the dynamic behavior of the QTS. Utilizing work by Johansson *et al.* [8] an original subsystem architecture was constructed in Simulink to simulate interconnected tank volumes. The design included in Figure 2 incorporated system parameters in MATLAB, establishing a virtual process plant for identification and validation studies. The developed numerical model provided a comprehensive understanding of the quadruple tank system's behaviors.

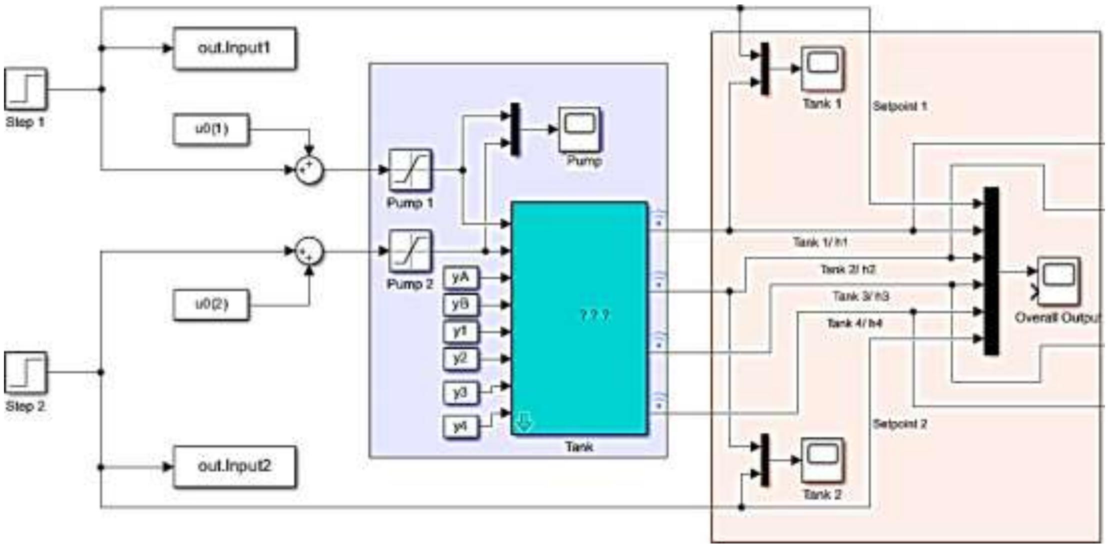


Figure 2. Simulink model for conducting open-loop system identification tests of quadruple tank system

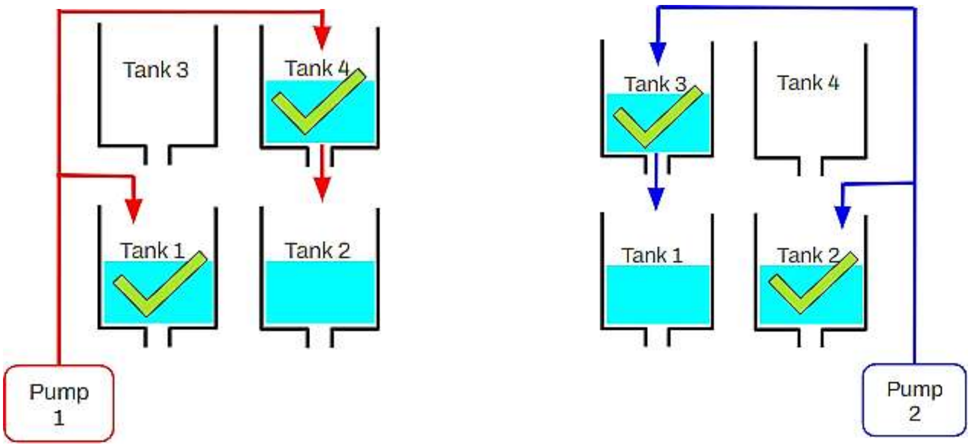
The system parameters used in this study were initially determined through prior experimental characterization work by Johansson [2]. Their study identified the key system attributes and dimensions that govern the dynamic response of the quadruple tank apparatus. These parameters are included in Table 1. The same parameter values were adopted here to maintain consistency with the established process model through replication experiments on an identical setup.

Table 1 Defined process parameters governing behaviour of the QTS

Parameter	Value
Tank cross-sectional area, A_1 and A_3	28 cm^2
Tank cross-sectional area, A_2 and A_4	32 cm^2
Nozzle cross-sectional area, a_1 and a_3	0.071 cm^2
Nozzle cross-sectional area, a_2 and a_4	0.057 cm^2
Acceleration of gravity, g	981 cm/s^2
Voltage applied to pump, V_1 and V_2	3,3 V

Pump constant, K_1 and K_2	3.33,3.35 cm^3/Vs
Flow distribution to lower and diagonal tank γ_1 and γ_2	0.70,0.80

The model structure was validated through open-loop step response tests under two operational configurations depicted in Figure 3. Test 1 involved exciting Pump 1 with a pseudo-random binary signal, isolating the transfer path between Tanks 1 and 4, while Test 2 characterized the transfer paths between Tanks 2 and 3 by activating Pump 2. The tests aimed to provide input-output data for system identification and established a virtual process plant for rigorous validation studies. The developed numerical model comprehensively characterized inherent QTS behaviors by estimating model parameters from test data and ensuring simulated responses reproduced empirical trends.



(a) Case 1: Pump 1 ON, Pump 2 OFF (b) Case 2: Pump 1 OFF, Pump 2 ON

Figure 3. QTS implementation for (a) Case 1 and (b) Case 2

2.3 Development and Validation of Transfer Function Model for the QTS

Johansson [2] illustrates the matrix configuration depicting flow relationships within the quadruple tank process, which is shown in Figure 4. Flow to tank 1, G_{11} is described as $\gamma_1 v_1 k_1$, while flow to tank 4, G_{22} is given as $(1-\gamma_1)k_1 v_1$. Similarly, flows to tank 2, G_{21} and tank 3, G_{12} are represented according to $\gamma_2 v_2 k_2$ and $(1-\gamma_2)k_2 v_2$ respectively. This arrangement can be captured via a flow matrix incorporating valve settings and distribution factors to precisely manage and analyze the multivariable nature of the system. The flow matrix formulation elucidates key dynamical interactions and dependencies between interconnected tanks.

Tank 1 (1st order) **Tank 3** (2nd order)

$$G(s)=\begin{bmatrix} \dfrac{\gamma_1 c_1}{1+sT_1} & \dfrac{(1-\gamma_2)c_1}{(1+sT_3)(1+sT_1)} \\ \dfrac{(1-\gamma_1)c_2}{(1+sT_4)(1+sT_2)} & \dfrac{\gamma_2 c_2}{1+sT_2} \end{bmatrix} = \begin{bmatrix} G_{11} & G_{12} \\ G_{21} & G_{22} \end{bmatrix}$$

Tank 4 (2nd order) **Tank 2** (1st order)

Figure 4. Flow matrix configuration from [2] depicting transfer function interactions between interconnected tanks in the QTS

A Simulink model was developed to simulate the transfer behavior of individual tanks via a multivariable control scheme. As depicted in Figure 5, the QTS was constructed such that G_{11}, G_{21}, G_{12}, and G_{22} represent tank 1 through 4, respectively. An assumption was made considering flow equality from upper tanks 3 and 4 to lower tanks, tanks 1 and 2, and from pumps to upper tanks. Although direct summation was used to combine upper tank transfer functions, mathematical modeling differences were deemed negligible for experimental and simulation accuracy. Overall, characterization of intrinsic flow properties and development of a Simulink representation incorporating the flow matrix formulation establishes and analytically defined virtual process for rigorous controller assessment and performance evaluation.

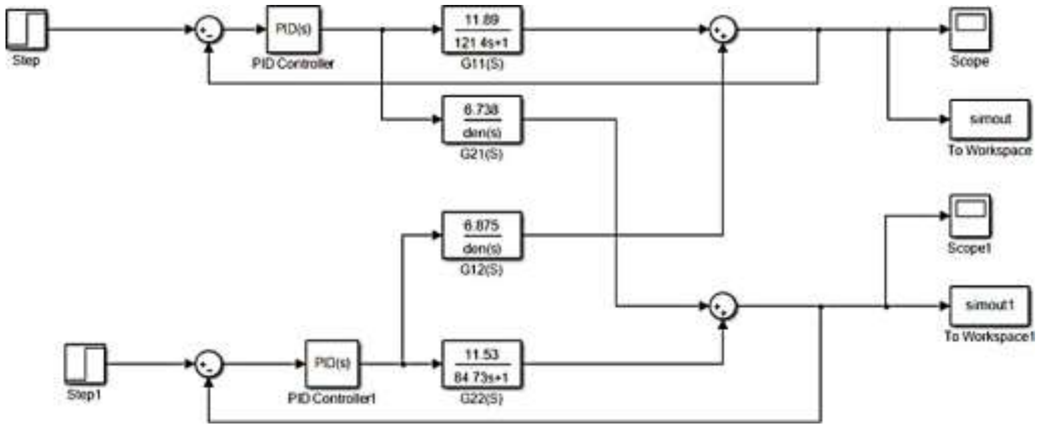
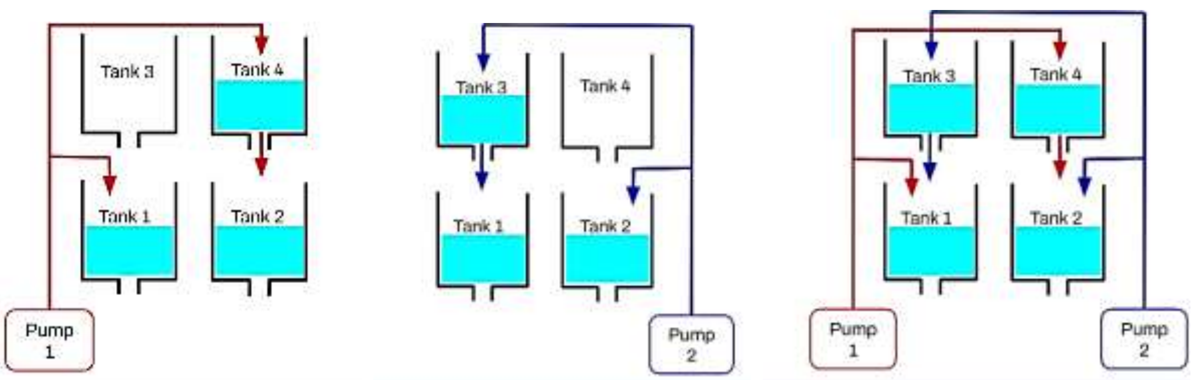


Figure 5. A Simulink model of the QTS constructed from individual tank transfer functions related via the characterized flow matrix

2.4 Controller Design and Implementation

Simulation tests were conducted in Simulink to assess the QTS's PID control. Figure 6 shows three scenarios that examined pump operation at the maximum 15V setting. In case 1, set point 1 was configured at 18 cm while set point 2 was held at 0 cm, activating only pump 1. Case 2 reversed this configuration to isolate the pump 2 control. Finally, case 3 challenged the controller by setting set point 1 at 18 cm and set point 2 at 14 cm, requiring coordinated multi-input multi-output (MIMO) regulation.



(a) Case 1: Setpoint 1 at 18 cm, Setpoint 2 at 0 cm

(b) Case 2: Setpoint 1 at 0 cm, Setpoint 2 at 18 cm

(c) Case 3: Setpoint 1 at 18 cm, Setpoint 2 at 14 cm

Figure 6. Simulation setup for performance analysis of controller for each case (a),(b), and (c)

2.4.1 Design and tuning for PID controller using Heuristic Method

PID controller uses proportional, integral, and derivative terms to regulate output by comparing the process variable with the set point, reducing errors, and ensuring smooth responses. K_p, K_i, and K_d gains can be adjusted to achieve balance, making them ideal for control applications. Mathematically, the control output, u(t), for the PID controller, is given by Eq. (5)

$$u(t) = K_p e(t) + K_i \int e(t)dt + K_d \frac{de}{dt} \tag{5}$$

A rigorous tuning procedure was undertaken to tune PID controller response characteristics systematically. The proportional, integral, and derivative parameters were methodically varied within the simulation environment to enhance transient response speed, minimize steady-state offset and reduce overshoot tendency. An iterative tuning in Table 2, where the approach involving repetitive simulations test and performance evaluation allowed gradual refinement of PID settings. The tuning workflow continued until the PID controller demonstrated robust compensatory behavior amid dynamic interactions within the multivariable QTS.

Table 2 PID Controller tuning parameters under testing scenarios for the QTS

Case/Controller	PID Controller 1			PID Controller 2		
	K_p	K_d	K_i	K_p	K_d	K_i
Case 1	20	0.03	0	0	0	0
Case 2	0	0	0	20	0.02	0
Case 3	20	0.035	10	20	0.025	15

2.4.2 Selection of MPC Parameter Using Heuristic Method

For the implementation of MPC, key algorithmic elements were rigorously optimized to maximize the closed-loop regulation of QTS. Central to the MPC approach are parameters governing the prediction horizon, the number of coordinated control moves considered, and cost function optimization criteria, which are tabulated in Table 3. An initial set of simulations without constraints, documented in Table 4, were performed to determine baseline MPC performance. This established the prediction and control horizons needed to adequately capture the interconnected tank system's multivariable dynamics. Subsequent simulations in Table 4 evaluated MPC under varying input and output constraints. Input constraints were imposed individually based on typical upper and lower bounds for pump flow rates. The maximum and minimum values were selected as 15 and ∞ respectively by observing the pump flows needed to achieve the set point minimal overshoot in unconstrained cases.

Table 3 Selection of sample times, prediction horizon, and control horizon of MPC under all cases

Cases	MPC Horizon		
	Sample Time (s)	Prediction Horizon, N_p	Control Horizon, N_u
Case 1	0.02	50	15
Case 2	0.02	50	15
Case 3	0.02	75	25

Lastly, combined input and output constraints were tested to study MPC performance under the most realistic operating restrictions. A review of the unconstrained response indicated target levels could be maintained. Therefore, these values were used as the maximum and minimum for the input and output constraints, respectively, 15 and 18 in Table 4. The gradual imposition of constraints in this manner investigated controller tolerance to operational restrictions while honing in on appropriate bounding values. Tight constraints challenged MPC to optimize control within practical physical limitations.

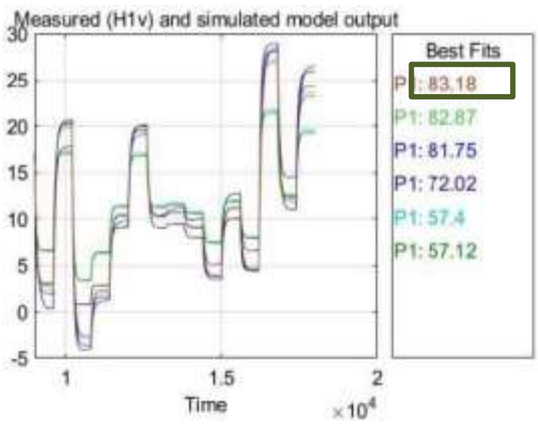
Table 4 Input and state constraint conditions imposed in MPC for control of the QTS

Cases	Without constraint				Input constraint				Input and output constraint			
	Input		Output		Input		Output		Input		Output	
	Min	Max	Min	Max	Min	Max	Min	Max	Min	Max	Min	Max
Case 1	-∞	∞	-∞	∞	0	15	-∞	∞	0	15	0	18
Case 2	-∞	∞	-∞	∞	0	15	-∞	∞	0	15	0	18
Case 3	-∞	∞	-∞	∞	0	15	-∞	∞	0	15	0	18

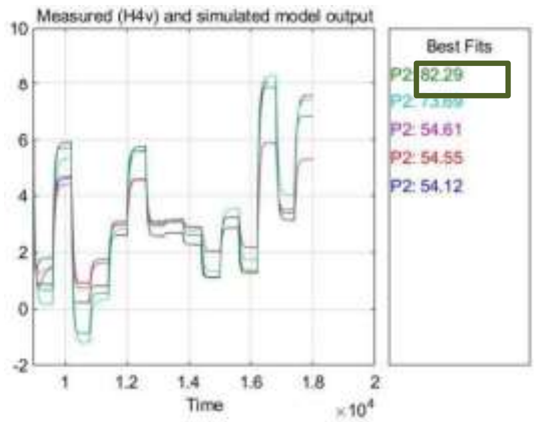
3. RESULTS AND DISCUSSION

3.1 QTS Open Loop Test and Transfer Function Modeling via System Identification Toolbox

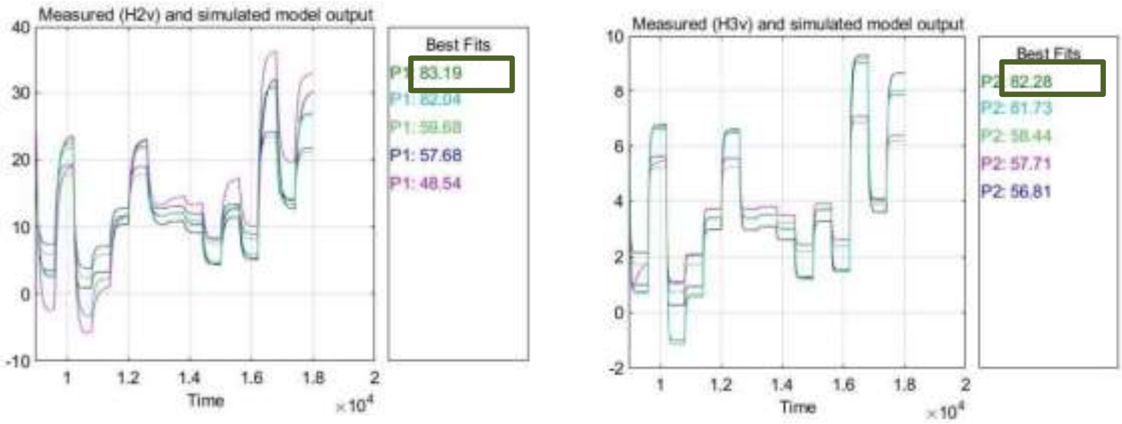
Open-loop step response tests were conducted by applying individual step inputs to each pump while keeping others at zero. Input and output data were measured using the QTS modelling, which accurately represents the system's dynamic behavior. The input-output time-domain data was analyzed using MATLAB's System Identification Toolbox to identify linear dynamic models. Initial model structures from first to second-order transfer functions were fitted to the test data. Optimization was performed to estimate model parameters that minimized the error sum of squares between actual and fitted responses.



(a) Case 1: Best Fits in Tank 1



(b) Case 1: Best Fits in Tank 4



(c) Case 2: Best Fits in Tank 2 (d) Case2: Best Fits in Tank 3

Figure 7. Estimation and validation with Best Fits for case 1 (a) and (b), and case 2 (c) and (d)

Upon achieving the best fit in Figure 7, the transfer functions were developed and arranged into a matrix configuration. Figure 8 depicts the interconnected nature of QTS, illustrating how the input to one pump affects the water level in a specific tank. This visualization aids in designing and implementing the transfer function model of QTS, enabling the management of complex interactions.

Tank 1 (1st order) **Tank 3** (2nd order)

$$G(s) = \begin{bmatrix} \dfrac{8.2976}{(1+59.928s)} & \dfrac{2.7739}{(1+30.855s)(1+1.3145s)} \\ \dfrac{2.4025}{(1+40.36s)(1+1.5053s)} & \dfrac{9.4503}{(1+82.713s)} \end{bmatrix} = \begin{bmatrix} G_{11} & G_{12} \\ G_{21} & G_{22} \end{bmatrix}$$

Tank 4 (2nd order) **Tank 2** (1st order)

Figure 8. Matrix configuration of QTS transfer function model

The QTS transfer function model was configured based on the matrix and was presented in Figure 9. This setup will simulate control analysis using PID and MPC controllers. The configuration offers a comprehensive framework to analyze the dynamic interactions between the tanks and pumps. This linear model will be used for controller design, simulation, and performance analysis to effectively manage the complex interactions between multiple coupled tanks and pump inputs.

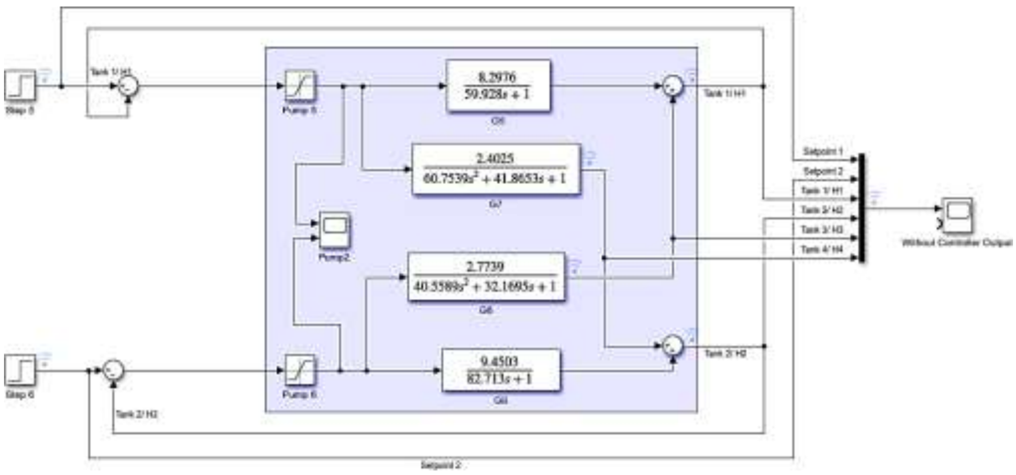
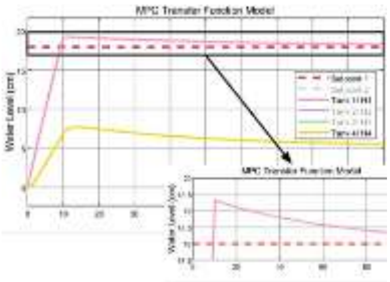


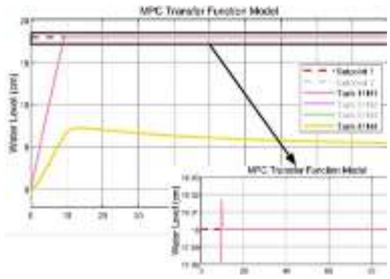
Figure 9. QTS transfer function model in Simulink

3.2 MPC Performance with and without Constraint

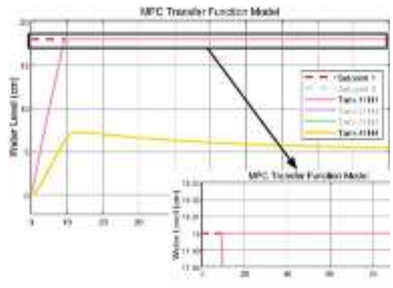
The MPC controller was first evaluated without constraints to serve as a baseline for comparison. In simulation trials encompassing three characteristic set point changes (cases 1 until 3), the controller tracked references acceptably but experienced an overshoot of up to 1.5 cm. In contrast, stabilization dynamics remained similar for cases 1 and 2, case 3 transitioned from rapid to gradual recovery. Input limitations from 0-15 V emulating realistic pump operation were enforced. A substantial decrease in overshoot exceeding 100% for cases 1 and 2 demonstrated enhanced precision in Figure 10. Concomitantly, post-disturbance stabilization accelerated, corroborating improved stability. For case 3, performance was also ameliorated markedly as overshoot and settling time reduced substantially.



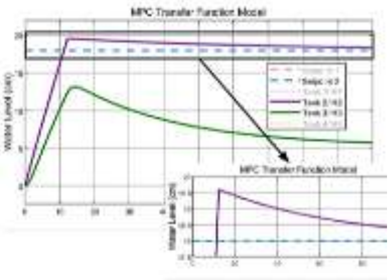
(a) Case 1: MPC response without constraint



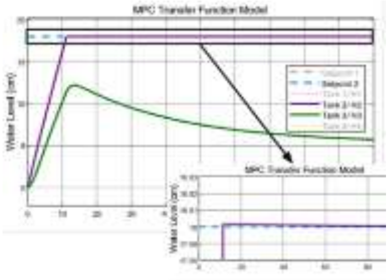
(b) Case 1: MPC response with input constraint



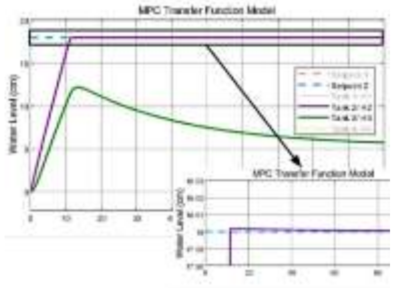
(c) MPC response with input and output constraint



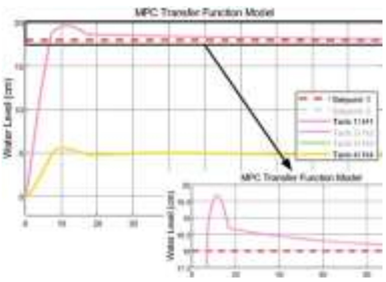
(d) Case 2: MPC response without constraint



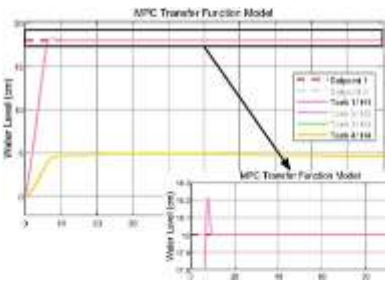
(e) Case 2: MPC response with input constraint



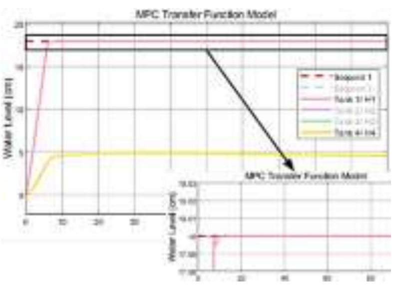
(f) Case 2: MPC response with input and output constraint



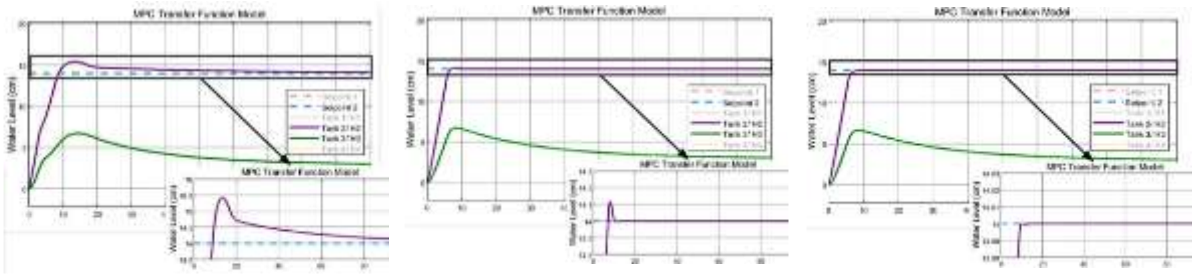
(g) Case 3 at Tank 1: MPC response without constraint



(h) Case 3 at Tank 1: MPC response with input constraint



(i) Case 3 at Tank 1: MPC response with input and output constraint



(j) Case 3 at Tank 2: MPC response without constraint

(k) Case 3 at Tank 2: MPC response with input constraint

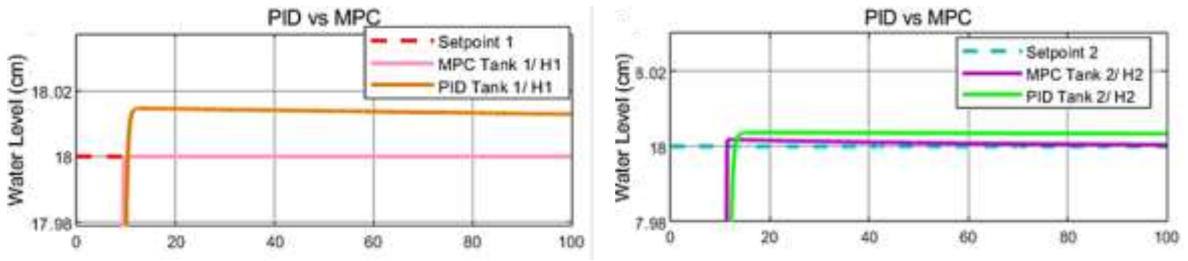
(l) Case 3 at Tank 2: MPC response with input and output constraint

Figure 10. Output responses of MPC in various conditions of constraint

Finally, combined input-output constraint bounded tank levels and flows optimally. Near-elimination of overshoot and negligible steady-state error validated MPC proficiency in optimizing control amid tight restrictions. Maximum level deviations fell below 0.1 cm, corroborating adept handling of QTS interactions to achieve high-fidelity set point tracking under constraints. Overall, constraints increasingly refined transient response while MPC exhibited skill adapting to limitations, a corroborating utility for industrial quadruple tank processes requisite of constrained precision control amid complex dynamics.

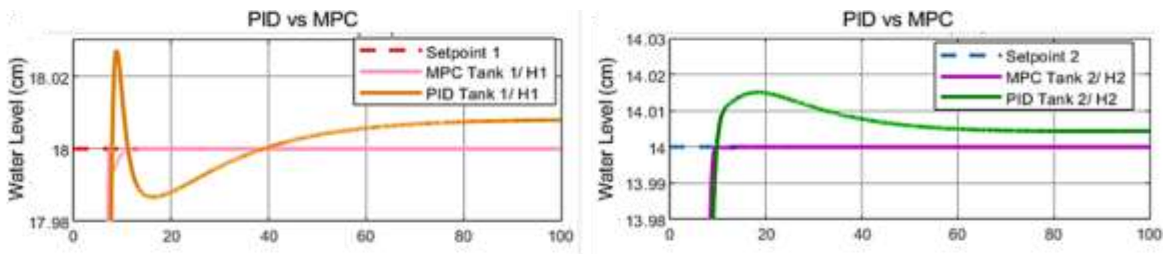
3.3 Output Comparison and Performance Analysis between PID Controller and MPC

The performance analysis revealed both controllers effectively regulated tank levels under different setpoints. However, MPC demonstrated better control qualities in certain cases. For Tank 1 in case 1, MPC attained slightly faster rise and settling times than PID. Both controllers achieved zero steady-state error, but MPC exhibited a marginal lower overshoot of 0.098%. In case 2 involving Tank 2, PID showed comparable rise and settling times to MPC. However, MPC controlled steady-state levels more accurately despite higher overshoot. As seen in Figure 11, MPC maintained levels closer to the setpoint.



(a) Case 1: Setpoint 1 at 18 cm

(b) Case 2: Setpoint 2 at 18 cm



(c) Case 3: Setpoint 1 at 18 cm and setpoint 2 at 14 cm

Figure 11. Output response comparison between PID and MPC in (a) Case 1, (b) Case 2, and (c) Case 3

MPC's predictive capability contributed significantly to its superior handling of the QTS's interactive multivariable behavior. By utilizing an internal process model, MPC could anticipate the influence of present inputs on future outputs across the entire networked tank system. This allowed it to optimize a sequence of moves, minimizing overall error systematically. In contrast, PID operated in a feedback mode without inherent knowledge of coupled system dynamics. As a result, PID was susceptible to interactions between manipulated and non-manipulated variables. This likely contributed to its marginally higher errors observed under setpoint changes.

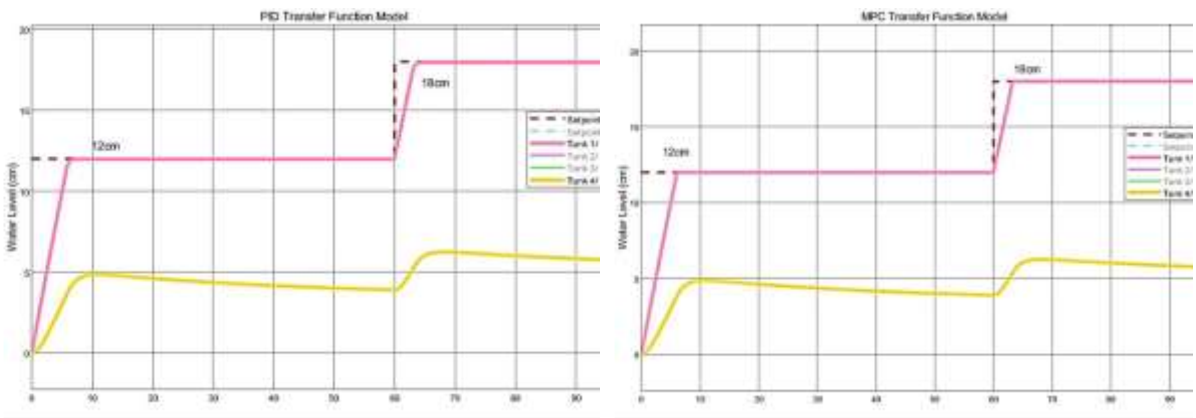
Additionally, MPC explicitly accounted for process input or state constraints, which helped regulate tank levels within tight bounds with virtually no steady-state error. PID lacked this constraint-handling ability, rendering its control less refined for the constrained QTS application. In summary, MPC excelled relative to PID, mainly because its model-based predictive control strategy could more effectively manage the multivariable nonlinearities inherent in the challenging QTS process.

Table 5 Report key closed-loop response metrics evaluated to assess the relative performance of PID and MPC methodologies

Performance Metrics	Case 1				Case 2				Case 3			
	Tank 1		Tank 2		Tank 1		Tank 2		Tank 1		Tank 2	
	PID	MPC	PID	MPC	PID	MPC	PID	MPC	PID	MPC	PID	MPC
Rise Time (s)	7.50	7.48	-	-	-	-	8.98	8.97	4.77	4.75	4.94	4.78
Settling Time (s)	9.22	9.16	-	-	-	-	11.06	10.99	6.62	6.16	7.19	6.69
Overshoot (%)	0.98	0.098	-	-	-	-	0.16	0.78	10.72	0.017	7.63	0.008
Steady State Error (cm)	0.013	0	-	-	-	-	0.003	0	0.130	0	0.300	0

3.4 Validation of the Robustness of the MPC during Case 1

The robustness of the MPC controller was validated by introducing single setpoint changes to tank 1 and comparing it with the PID controller. Initially, the setpoint was 12 cm. Figure 12 shows both controller response to the liquid level around this setpoint, and MPC successfully regulated over the control horizon.



(a) PID output response (b) MPC output response

Figure 12. Process variable response for set point changes under case 1 operation at tank 1

At 60 seconds, the setpoint was increased to 18 cm. The MPC recomputed optimal control moves to transition the level between the setpoints smoothly. As shown in Figure 12, there was no overshoot or instability. Table 6 shows the quantitative performance of both controllers at each setpoint. At 12 cm, the MPC had a rise time of 5.72 seconds, a settling time of 6.90 seconds, and a negligible overshoot of 0.124% with no steady-state error. The PID showed nearly similar dynamics but with a 0.5 cm steady-state error. When increased to 18 cm, the MPC and PID rise and settling times were comparable at 61.41 seconds and 63.08 seconds for MPC and 61.51 seconds and 63.25 seconds for PID, respectively. Only the MPC achieved zero steady-state error, demonstrating more accurate long-term regulation.

Table 6 Performance metrics characterizing the process response to set point changes at 12 cm and 18 cm under MPC

Performance Metrics/Step	12 cm		18 cm	
	PID	MPC	PID	MPC
Rise time (s)	5.82	5.72	61.51	61.41
Settling time (s)	6.99	6.90	63.25	63.08
Overshoot (%)	0	0.124	0	0
Steady-state error (cm)	0.5	0	0.5	0

While both controllers transitioned setpoints smoothly, the MPC exhibited tighter setpoint tracking through zero steady-state error. Figure 13 shows the comparison of MPC and PID in one output response. This validates its superior reference tracking robustness. Additionally, the MPC can self-tune to changing conditions, whereas PID requires retuning. In conclusion, direct comparison to PID confirms the MPC's enhanced setpoint regulation and adaptive capabilities, reinforcing its robustness and suitability for demanding process control applications.

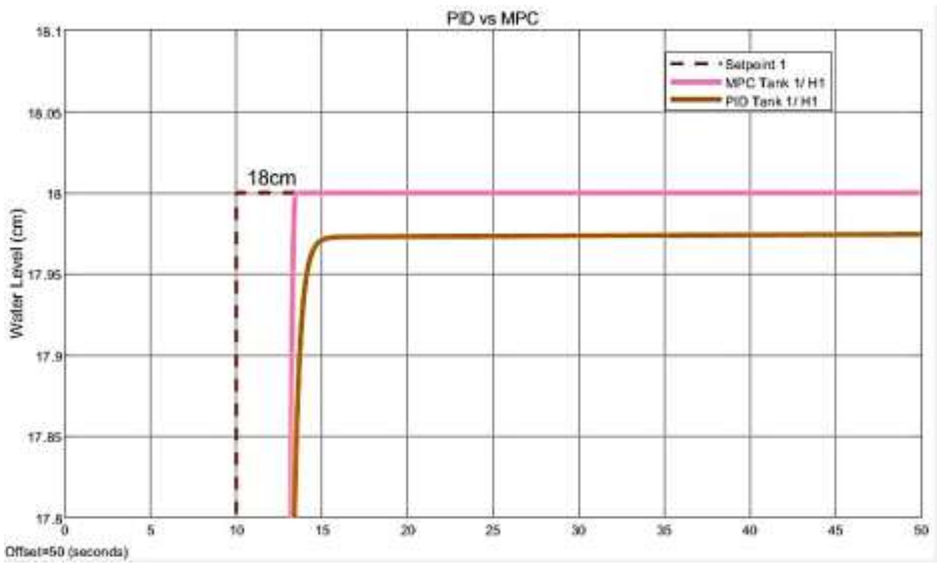


Figure 13. Comparison of output response between PID controller and MPC

4. CONCLUSION

In conclusion, this work demonstrated the efficacy of MPC for regulated control of a challenging QTS process characterized by interactive multivariable dynamics and operational constraints. An accurate linear process representation capturing dominant plant behaviour was obtained through rigorous system identification and model validation. Controllers were designed, tuned and analyzed via extensive simulation under diverse setpoint scenarios. Quantitative performance metrics revealed that MPC consistently outperformed PID, achieving faster response, lower error and overshoot due to its inherent predictive capabilities and constraint handling. Utilizing an internal dynamic model allowed MPC to systematically optimize multistep predictions accounting for input or state limitations and interactions across the networked process. MPC optimized the liquid level between changing targets with zero steady-state error while a consistent offset marred PID response. This research, therefore, highlights the critical advantages of MPC's optimization abilities in addressing demands for constraint control and adaptivity that conventional PID lacks. The findings represent a meaningful contribution by definitively demonstrating MPC's superiority in regulatory performance and flexibility

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