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Modelling and sensitivity analysis of a generic chemical process plant for plantwide control

Anna B. A. Andrade[a], Beatriz C. Andrade[a], Hugo M. Mafra[a], Rogério L. Pagano[a], Rodolpho R. Fonseca[a,*], José L. Guzmán[b]

[a] *Chemical Engineering Dpto, Federal University of Sergipe, Marcelo Deda Chagas Avenue, s/n, 49107-230 São Cristóvão, Brazil.*
[b] *Dpto. de Informatica, Universidad de Almería, ceiA3, Carretera Sacramento, s/n, 04120, Almería, España.*

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Abstract

This work presents the modelling and sensitivity analysis of a continuous chemical plant, using dynamic simulations in the open-source software Scilab/XCOS®. The study evaluates the steady state gains between disturbances and processes variables in order to identify the most significant cause and effect relationships in the system. Different operational scenarios were simulated through the application of step disturbances, allowing the analysis of the transient behavior of the main variables in open loop. The results highlight the influence of specific variables on process stability, providing technical foundations for the optimized selection of control loop strategies. This design approach contributes to structuring a more efficient plantwide control design, aiming the minimization of undesired variables interactions, increasing process safety and control.

Keywords: Modelling, Sensitivity analysis, Chemical industry, Plantwide, Process control.

1. Introduction

In the present, with global industrialization, a lot of products consumed in the everyday life is produced in an industrial plant. In this way, the capability to control chemical plants and its process variables is extremely necessary and applying plantwide control is highly desirable. This control strategy allows to maintain process variables in a desirable level, keeping operational conditions within the safety limits and optimizing the process (Downs and Vogel, 1993).

Most industrial processes can be considered complex because of the quantity and type of equipment utilized, their connections with each other and the existence of recycling streams. For example, a typical characteristic of industrial chemical processes that contributes to its complexity is the chemical reactions in the process and, regardless their rating and nature, whether it is an exothermic or endothermic reaction, the temperature control is essential to avoid undesirable effects like thermal runaway (Luyben, 2019), which can cause serious damages and plantwide shutdown.

Several studies address the challenges associated with modelling and controlling plants with recycle streams, particularly regarding the mitigation of the snowball effect, a phenomenon in which small variations in the feed load can trigger severe and disproportionate changes in the recycle streams (Luyben, 1994), and the strong coupling between process variables. In this context, fundamental hierarchies for integrated systems have been proposed, emphasizing inventory stabilization and efficient management of recycle streams (Skogestad, 2004). Subsequently, such methodologies were applied to complex reactor-separator systems, demonstrating that decentralized control combined with sensitivity analysis constitutes an effective strategy to reduce open-loop disturbance propagation and improve overall process operability (Araujo, 2008; Skogestad, 2004).

In chemical plants with recycle, small operational disturbances can be amplified throughout the system, compromising product purity, operational stability, and process safety. In this scenario, the present work is justified by the need to deepen the dynamic analysis of nonlinear processes through open-source tools, such as Scilab/XCOS® (Scilab, 2026). Based on a sensitivity analysis and in the methodology proposed by Luyben *et al.* (1997) for plantwide control design, this study focus on process modelling and analysis of disturbance variables influence on process dynamic behavior. This approach contributes to increased

*Author for correspondence: rodolphorf@academico.ufs.br

operational robustness and thermodynamic efficiency, allowing the plant to reach new steady states in open loop, even under critical disturbances in the thermal utilities.

2. **Methods**

The chemical plant described in this paper is proposed for further plantwide control studies. Therefore, its phenomenological model captures the dynamics of a continuous chemical plant with a recycle stream, whose unit operations are shown in the flowsheet in Figure 1. The process begins in the surge tank V-01, which combines the fresh feed of component A (stream nº1) with the liquid recycle stream (stream nº10). The resulting mixture is heated in the shell and tube heat exchanger E-01 before entering the reaction section, composed of the CSTRs in series R-01 and R-02. In these reactors, the additions of catalyst (stream nº4) and inhibitor (stream nº6) occur, respectively. The reactor effluent is then cooled in exchanger E-02 and directed to the flash separator vessel V-02. In this equipment, the vapor phase is recovered at the top (stream nº9), while the bottom liquid phase is divided into a purge stream (stream nº11) and a recycle stream that returns to V-01.

The entire mathematical modelling of the system was programmed and simulated in the Scilab/XCOS® and is presented in Appendix A. The model parameters are indicated in Appendix B. To solve the system of differential equations, the solver Runge-Kutta 4(5) was employed.

For the design of process control, the plantwide control methodology proposed by Luyben *et al.* (1997) was adopted. This method consists in nine sequential steps, as follows: 1º) establish the control objectives, 2º) determine control degrees of freedom, 3º) establish energy management system, 4º) set production rate, 5º) set product quality, handle safety, operational and environmental constraints, 6º) set control inventories and fixed recycle flow rate, 7º) check component balances, 8º) control individual unit operations and 9º) optimize economics or improve dynamic controllability.

This control design approach establishes a logical hierarchy that ensures tight control over mass and energy inventories, which is essential for stabilizing the overall balance in processes with a recycle stream. However, closed-loop simulations were not included in this report and will be addressed in future work, since this report focuses on process modelling and the sensitivity analysis of disturbance variables as a background for further control studies.

2.2. Selection of Variables for Sensitivity Analysis

The selection of disturbance variables for the sensitivity analysis was based on the systematic application of nine steps procedure for plantwide control design (Luyben *et al.*, 1997). The adoption of this methodology is particularly relevant given the high complexity of the plant under study, characterized by the presence of a recycle stream and the strong integration between unit operations, which allows the propagation of disturbances and imposes additional challenges to the stability and dynamic performance of the system.

In this context, the concentration of A in stream nº1 ($C_{V\text{-}01,A,1}$) was defined to guarantee the overall component balance, ensuring that each chemical species has a defined destination either by reaction or exit, thus avoiding undesired accumulations in the system. To address energy management and the control of individual units, the utility temperature in E-01 (T_{lps}) was selected to maintain the temperature constant in the feed of the first reactor. As a variable for manipulating the throughput, the catalyst concentration in stream nº4 ($C_{R\text{-}01,C,2}$) was chosen because it provides a direct and rapid effect on the reaction rate. The inhibitor concentration in stream nº6 ($C_{R\text{-}02,I,2}$) acts on the management of operational and safety constraints, being important to prevent thermal runaway and

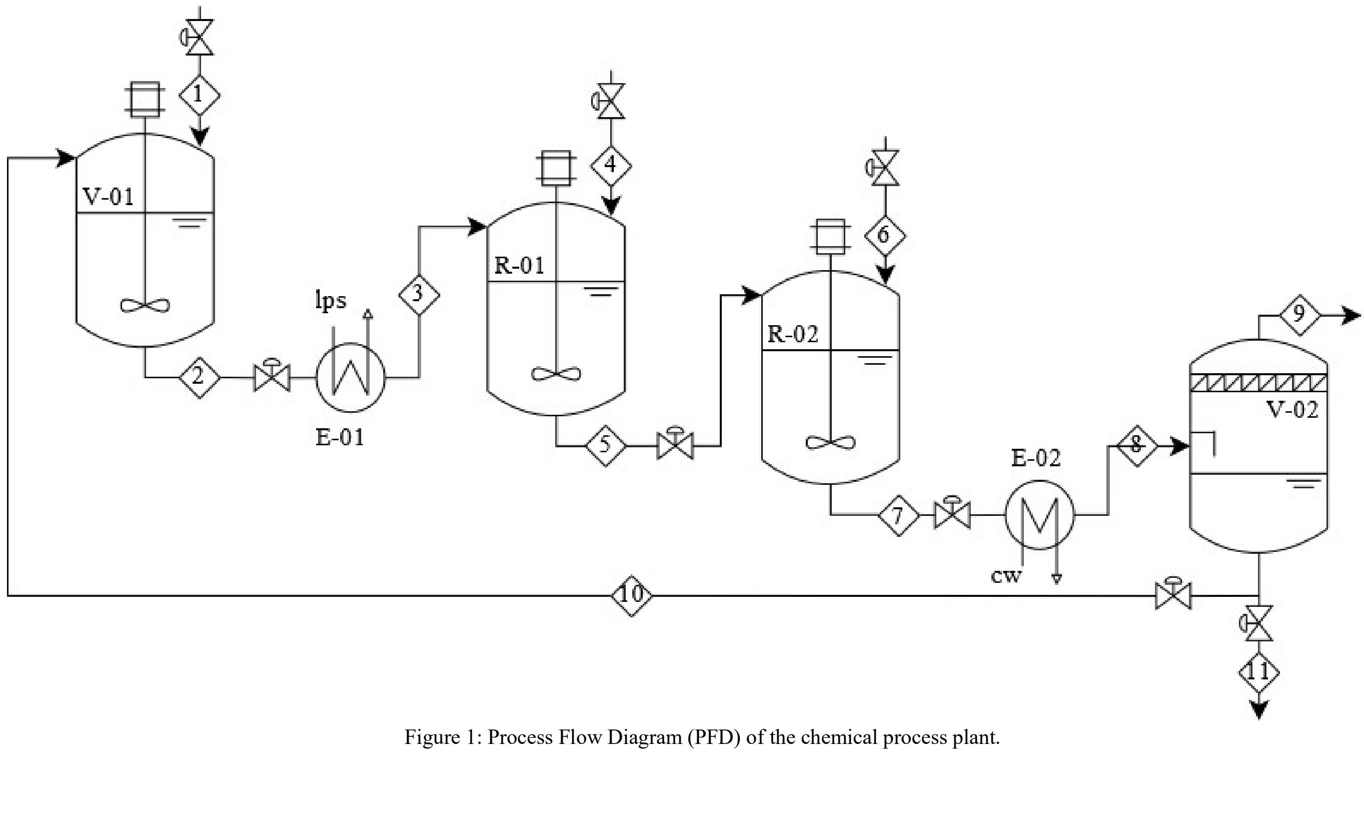

Figure 1: Process Flow Diagram (PFD) of the chemical process plant.

ensure process integrity. Finally, the utility temperature in E-02 (T_{cw}) and the pressure in V-02 (P) were selected for quality and inventory control, since they affect the concentration of the product B in stream nº9 ($C_{V\text{-}02,B,V}$).

2.3. Sensitivity analysis method

To perform the sensitivity analysis and evaluate the impact of disturbance variables on the PVs, step disturbances of +/-2% and +/-5% were applied to the nominal values of the disturbances. This approach allowed to observe the dynamic behavior of the plant in open loop, verifying its capacity in converging to a new steady state, once chemical processes can reach different steady states depending on operational conditions (Seborg *et al.*, 2016). The simulations were executed in Scilab/XCOS®, using Runge-Kutta 4(5) numerical integration method solver. A total time of 14400 s was considered with the disturbance step occurring at 7200 s.

The sensitivity analysis was based on the evaluation of the steady state gains (Kp) obtained between the system variables. Thus, the magnitude of the gain acts as a direct indicator of the degree of influence between the variables: the higher its absolute value, the more significant and pronounced is the impact of the disturbance variables on process variables.

3. Results and Discussion

3.1. Values of the variables at the initial steady state

The initial conditions were determined considering enough simulation time to reach steady state. Initial conditions are indicated in Table 1.

Table 1: Values of the process variables

Analyzed Variables	Values at steady state
$C_{R\text{-}01,B,o}$	3.294 [mol.m^{-3}]
$C_{V\text{-}01,I,o}$	0.008 [mol.m^{-3}]
$C_{V\text{-}02,B,V}$	0.286 [mol.m^{-3}]
$C_{V\text{-}02,B,L}$	0.476 [mol.m^{-3}]
$C_{V\text{-}02,I,L}$	0.011 [mol.m^{-3}]
$T_{V\text{-}02}$	346 K

3.2. Simulation Results

Considering the plantwide control method proposed by Luyben *et al.* (1997), the MVs and PVs were defined as indicated in Table 2. The sensitivity analysis developed in this paper aims to further contribute to the feedback control improvement considering other advanced control strategies, *e.g.* feedforward and ratio control.

Table 2: Proposed MVs and PVs for feedback control.

MV	PV		MV	PV
$F_{V\text{-}01,o}$	$h_{V\text{-}01}$		$F_{E\text{-}02,s}$	$T_{V\text{-}02}$
$F_{E\text{-}01,s}$	$T_{R\text{-}01}$		F_{purge}	$h_{V\text{-}02}$
$F_{R\text{-}01,o}$	$h_{R\text{-}01}$		$F_{R\text{-}02,2}$	$C_{V\text{-}01,I,o}$
$F_{R\text{-}01,2}$	$C_{B,5}$		$F_{V\text{-}01,1}$	$C_{B,9}$
$F_{R\text{-}02,o}$	$h_{R\text{-}02}$			

The results of sensitivity analysis are presented in Table 3. For improved visualization and discussion of the results, a subset of the data was considered, highlighting only the cases in which an absolute value of Kp higher than ± 0.6 was obtained in at least on step. The dynamic responses of the main process variables obtained from the sensitivity analysis are presented below in graphical form, from Figure 2 to Figure 6.

Table 3: Sensitivity analysis.

Disturbance	PV	Gains considering steps of			
		+2%	-2%	+5%	-5%
$C_{V\text{-}01,A,1}$	$C_{R\text{-}01,B,o}$	3.06	3.06	3.06	3.07
	$T_{V\text{-}02}$	-1.4	-1.5	-1.4	-1.5
T_{lps}	$C_{V\text{-}02,B,V}$	-2.74	-4.86	-2.03	-9.24
	$T_{V\text{-}02}$	80.9	81.3	80.1	82.0
$C_{R\text{-}02,I,2}$	$T_{V\text{-}02}$	1.4	1.5	1.4	1.5
T_{cw}	$C_{V\text{-}02,B,V}$	-1.72	-2.36	-1.41	-3.19
	$T_{V\text{-}02}$	45.6	45.8	45.5	45.9
P	$C_{V\text{-}02,B,V}$	0.82	0.72	0.90	0.66
	$T_{V\text{-}02}$	6.9	7.1	6.8	7.2

These figures illustrate the temporal evolution of selected processes variables under steps of ±2% and ±5% applied to the disturbance variables, allowing a clearer visualization of the transient behavior, the new steady state conditions, and the overall sensitivity of the plant in open loop.

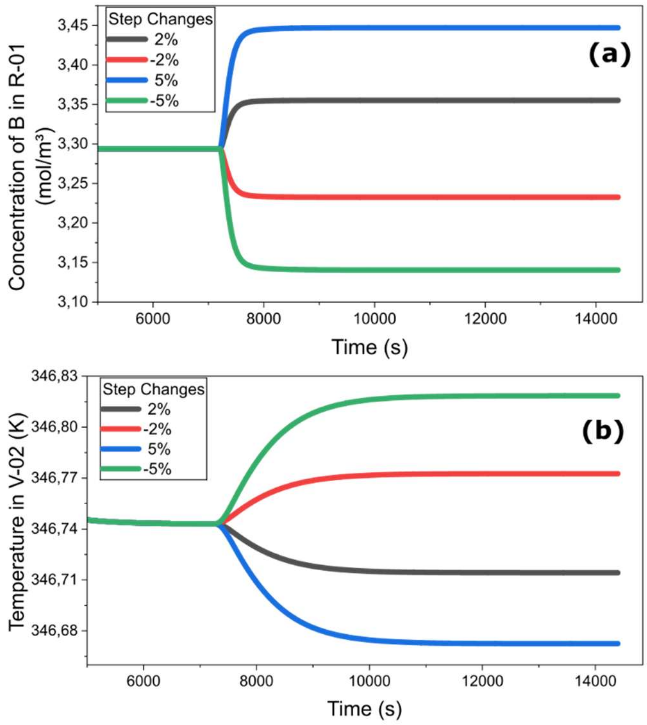

Figure 2: Simulation results for step in $C_{V\text{-}01,A,1}$. (a) Transient responses of $C_{R\text{-}01,B,o}$. (b) Transient responses of $T_{V\text{-}02}$

The plots include the pre-disturbance steady state condition before disturbances at t_d = 7,200 s and extend until the end of the simulation period at t_f = 14,400 s. Together with the results summarized in Table 3, these plots support a more comprehensive discussion of the process dynamics.

The comparative analysis of the steady state gains obtained in the simulations allowed inferences regarding the degree of linearity in the interactions between manipulated and process variables. It was observed that, for certain relationships, the

Kp values remained consistent and numerically close, regardless of the direction or magnitude of the applied disturbance, characterizing a predominantly linear behavior.

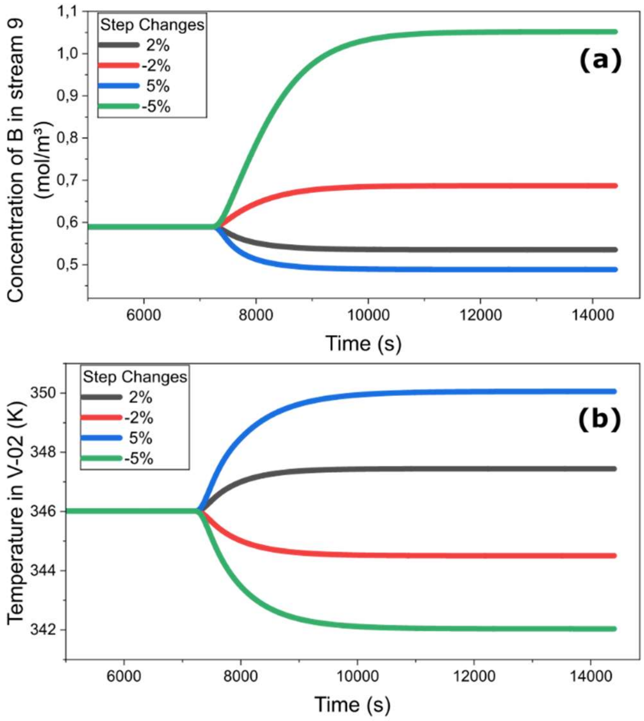

Figure 3: Simulation results for step in T_{lps}. (a) Transient responses of $C_{V\text{-}02,B,V}$. (b) Transient responses of $T_{V\text{-}02}$.

Conversely, responses that showed significant discrepancies in the magnitude of the gain according to the amplitude and sign of the disturbance evidence non-linear interactions. This pronounced non-linearity is inherent to the phenomenological complexity of the system under study. In this context, the non-linear behaviors observed in the influence of the utility temperatures (T_{lps} and T_{cw}) and pressure (P) on the top stream product concentration $C_{V\text{-}02,B,V}$ stand out, as well as the strong coupled interaction between the pressure and the temperature of the flash vessel $T_{V\text{-}02}$.

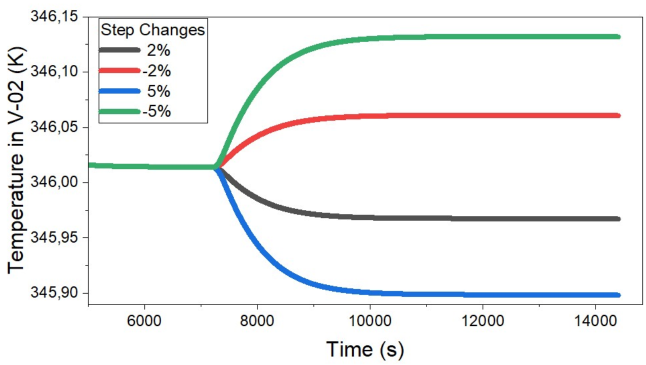

Figure 4: Transient responses of $T_{V\text{-}02}$ following step in $C_{I,6}$.

3.3. Sensitivity Analysis and Control Implications

Based on the results, it was observed that altering the concentration of $C_{V\text{-}01,A,1}$ directly impacts the $C_{R\text{-}01,B,o}$, since A is the precursor reactant for B. Besides that, variations in the feed load of A result in changes in the production rate of B, which alters the overall mass balance and the composition of the feed entering V-02. Thus, an increase in A decreases $T_{V\text{-}02}$, as shown in Table 3 by the high gain value between $C_{V\text{-}01,A,1}$ and $T_{V\text{-}02}$. It is observed because $C_{V\text{-}01,A,1}$ alters the thermal properties of the mixture and the overall boiling point of the liquid. Thermodynamically, this change in composition shifts the liquid-vapor equilibrium, altering the individual vapor pressures and, consequently, the equilibrium constants.

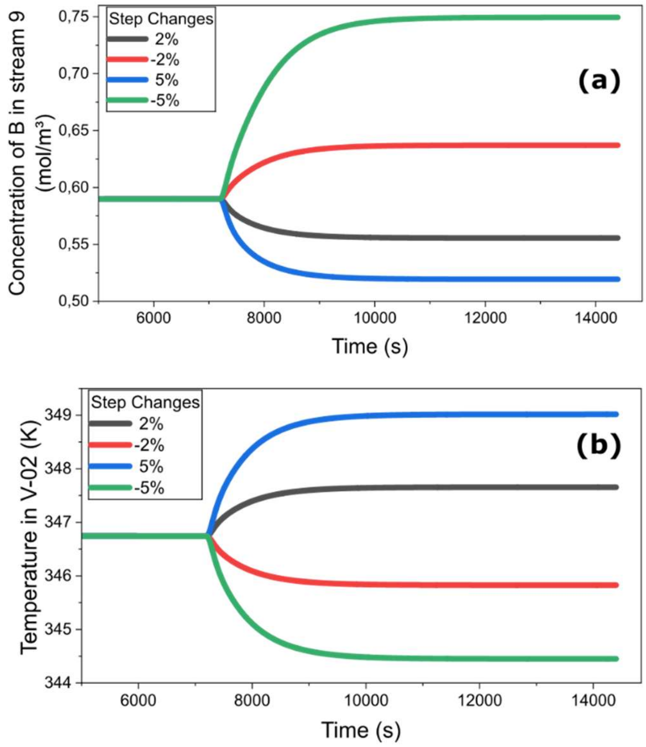

Figure 5: Simulation results for step in T_{cw}. (a) Transient responses of $C_{V\text{-}02,B,V}$. (b) Transient responses of $T_{V\text{-}02}$.

Regarding the impact of T_{lps} on process variables, the most sensitive variable is $T_{V\text{-}02}$. This occurs because the internal temperature of the vessel is the result of a thermal balance between the inlet and outlet streams. An increase in the T_{lps}, used for heating, propagates through the system, raising the $T_{V\text{-}02}$ and impacting the $C_{V\text{-}02,B,V}$. Given that the production of B is an endothermic reaction, heating favors the chemical kinetics, increasing the total mass of B that reaches the vessel. However, this thermal increase also significantly raises the overall vaporized fraction (β). Since concentration is a relative measure, component B undergoes a dilution effect in the vapor phase, as the total volume of vapor generated by the other components grows at a higher proportion than the increase in the production of B in the reactors. It can be observed that when a -5% disturbance is applied to the utility temperature, the steady state gain for $C_{V\text{-}02,B,V}$ is higher compared to the other cases, indicating a non-linear behavior. This occurs because component B is more volatile than the remaining components. Therefore, the more significant the reduction in the inlet temperature, the less the heavier components will vaporize, resulting in a higher concentration of B. The disturbance $C_{R\text{-}01,C,2}$ demonstrated a negligible impact compared to other disturbance variables. In contrast, $C_{R\text{-}02,I,2}$ presents a moderate impact on $T_{V\text{-}02}$, operating analogously to the variation of A by altering the reaction kinetics and the inlet thermal load on V-02.

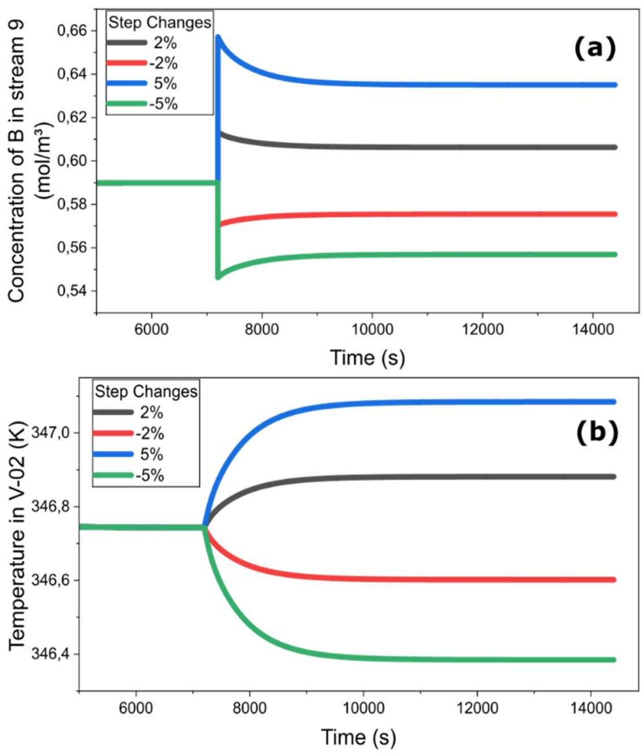

Figure 6: Simulation results for step in P. (a) Transient responses of $C_{V\text{-}02,B,V}$. (b) Transient responses of $T_{V\text{-}02}$.

The effect of T_{cw} is similar to T_{lps}, but with an inverse behavior due to its cooling function. An increase in this temperature raises $T_{V\text{-}02}$, which, as previously discussed, causes a decrease in $C_{V\text{-}02,B,V}$ due to the increase in the overall vaporized fraction and the consequent dilution of the component at the top. Conversely, a reduction in T_{cw} intensifies the heat exchange on E-02, reducing $T_{V\text{-}02}$ and favoring an increase in $C_{V\text{-}02,B,V}$.

It is important to analyze $C_{V\text{-}02,B,V}$ behavior in response to steps in P due to the characteristic curve shape presented in Figure 6(a). Unlike the other responses, the presence of an overshoot is observed immediately after the application of the disturbance, a typical response for lead-lag first order process when zero is lower than pole (Seborg *et al.,* 2016). However, this phenomenon is associated with a rapid compression or expansion of the gas phase volume, since an increase in pressure causes an immediate volume reduction, instantaneously raising the concentration and generating sharp peaks. Conversely, a pressure reduction has the opposite effect. After this initial peak, the system undergoes a slower accommodation, governed by the liquid-vapor thermodynamic equilibrium. Under this condition, an increase in pressure favors the partial condensation of the components, gradually reducing the concentration from the peak, while a decrease in pressure stimulates vaporization, partially recovering the initial drop. Furthermore, the P acts by shifting the thermodynamic equilibrium, modifying the partial pressures and the equilibrium constants of the components. This change alters the vaporization rate and, consequently, the temperature of the vessel. Thus, justifying the high steady state gains observed for these couple of variables in Table 3.

From the sensitivity analysis in open loop, it is possible to conclude that future projects should consider the application of control strategies that act directly on the disturbance variables with the greatest impact on the PVs. Considering the proposed MVs and PVs in Table 2, for $T_{V\text{-}02}$ and $C_{V\text{-}02,B,V}$ control, which are critical variables highly influenced by P, T_{cw} and T_{lps}, the use of PI/PID and feedforward control are recommended, once offset could be avoided with disturbance rejection. On the other hand, regarding level control loops, to avoid disturbance propagation through the plant as discussed by Luyben *et al.* (1997), P control should be considered.

4. Conclusions

In summary, the evaluation of steady state gains proved to be a fundamental tool for understanding the interactions among disturbances and process variables. Most notably, temperature variations in the utilities T_{lps} and T_{cw} exert the most profound impact on the overall dynamics of the plant in study. These thermal disturbances propagate intensely through the system, resulting in the highest observed gains for the flash vessel temperature. In addition, to highlighting these primary impacts, the analysis captured complex transient phenomena, such as pressure overshoot and compound dilution effects in the vapor phase. By rigorously mapping these cause-and-effect relationships, it was possible to translate the open-loop dynamic behavior into an assertive plantwide control strategy. Therefore, it is concluded that the strategic application of PID control associated with feedforward, guided by the highest system sensitivities to strictly regulate mass and energy inventories, can be an essential strategy for disturbance rejection.

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Appendix A

The subindices j, i, o, n, p, s, m, L, V on model refers to plant equipment tag on PFD, input stream, output stream, chemical specie, pipe and shell sides of heat exchangers, mean temperatures on heat exchangers sections, liquid and vapor phases on flash vessel, respectively.

Mass balance for surge tank and reactors

$$a_j \frac{dh_j(t)}{dt} = \sum_{i=1}^{2} F_{j,i}(t) - F_{j,o}(t) \tag{1}$$

Energy balance for surge tank

$$a_j h_j(t) \frac{dT_{j,o}(t)}{dt} = \sum_{i=1}^{2} F_{j,i}(t) T_{j,i}(t) - F_{j,o}(t) T_{j,o}(t) \tag{2}$$

Energy balance for CSTRs

$$a_j h_j(t) c_p \frac{dT_{j,o}(t)}{dt} = \sum_{i=1}^{2} F_{j,i}(t) c_p T_{j,i}(t) - F_{j,o}(t) c_p T_{j,o}(t) + a_j h_j(t) r_{j,1}(t) \Delta H_1 + a_j h_j(t) r_{j,2}(t) \Delta H_2 \tag{3}$$

Chemical reactions and kinects

$$A \overset{C}{\rightarrow} 2B \tag{4}$$

$$C + I \rightarrow 2D \tag{5}$$

$$r_{j,1}(t) = k_1 e^{\frac{-E_{a,1}}{RT_{j,o}(t)}} C_{j,A,o}(t) C_{j,C,o}(t) \tag{6}$$

$$r_{j,2}(t) = k_2 e^{\frac{-E_{a,2}}{RT_{j,o}(t)}} C_{j,I,o}(t) C_{j,C,o}(t) \tag{7}$$

Concentrations on surge tank

$$a_j h_j(t) \frac{dC_{j,n,o}(t)}{dt} = \sum_{i=1}^{2} F_{j,i}(t)\, C_{n,i}(t) - F_{j,o}(t) C_{j,n,o}(t) \tag{8}$$

Concentration of A, B, C, I and D on CSTRs

$$a_j h_j(t) \frac{dC_{j,A,o}(t)}{dt} = \sum_{i=1}^{2} F_{j,i}(t)\, C_{A,i}(t) - F_{j,o}(t) C_{j,A,o}(t) - a_j h_j(t) r_{j,1}(t) \tag{9}$$

$$a_j h_j(t) \frac{dC_{j,B,o}(t)}{dt} = \sum_{i=1}^{2} F_{j,i}(t)\, C_{B,i}(t) - F_{j,o}(t) C_{j,B,o}(t) + 2 a_j h_j(t) r_{j,1}(t) \tag{10}$$

$$a_j h_j(t) \frac{dC_{j,C,o}(t)}{dt} = \sum_{i=1}^{2} F_{j,i}(t)\, C_{C,i}(t) - F_{j,o}(t) C_{j,C,o}(t) - a_j h_j(t) r_{j,2}(t) \tag{11}$$

$$a_j h_j(t) \frac{dC_{j,I,o}(t)}{dt} = \sum_{i=1}^{2} F_{j,i}(t)\, C_{I,i}(t) - F_{j,o}(t) C_{j,I,o}(t) - a_j h_j(t) r_{j,2}(t) \tag{12}$$

$$a_j h_j(t) \frac{dC_{j,D,o}(t)}{dt} = \sum_{i=1}^{2} F_{j,i}(t)\, C_{D,i}(t) - F_{j,o}(t) C_{j,D,o}(t) + 2 a_j h_j(t) r_{j,2}(t) \tag{13}$$

Shell and tube heat exchanger modelling

$$\rho a_j h_j(t) c_p \frac{dT_{mj,p}(t)}{dt} = \rho c_p F_{j,p}(t) \left[T_{j,p,i}(t) - T_{j,p,o}(t) \right] + U a_j \Delta T_k(t) \tag{14}$$

$$\rho a_j h_j(t) c_p \frac{dT_{mj,s}(t)}{dt} = \rho c_p F_{j,s}(t) \left[T_{j,s,i}(t) - T_{j,s,o}(t) \right] - U a_j \Delta T_k(t) \tag{15}$$

$$\Delta T_k(t) = \left[T_{j,s,i}(t) - T_{j,p,o}(t) + T_{j,s,o}(t) - T_{j,p,i}(t) \right] / 2 \tag{16}$$

$$T_{j,p,o}(t) = 2T_{m,j,p}(t) - T_{j,p,i}(t) \tag{17}$$

$$T_{j,s,o}(t) = 2T_{m,j,s}(t) - T_{j,s,i}(t) \tag{18}$$

Mass balance for flash vessel

$$\rho_L . a_j \frac{d\,h_j(t)}{dt} = Q_i .\, \rho_i - Q_{o,L} .\, \rho_L - Q_{o,V} .\, \rho_V \tag{19}$$

Energy balance for flash vessel

$$\frac{d\,T_{V-02}(t)}{dt} = \frac{Q_i . \rho_i . cp_i}{\rho_L . V_L} T_i(t) - \frac{Q_{o,L} . \rho_L . cp_L}{\rho_L . V_L} T_{V-02}(t) + \frac{+Q_{o,V} . \rho_V . cp_V}{\rho_L . V_L} T_{V-02}(t) - \frac{Q_{o,V} . \rho_V . \Delta h}{\rho_L . V_L} + \frac{q}{\rho_L . V_L} \tag{20}$$

Mass fractions on flash vessel

$$\frac{\rho_i \cdot a_j \cdot h_j(t)}{\sum x \cdot M} \cdot \frac{d\,x_n}{dt} = \frac{Q_i \cdot \rho_i}{\sum z \cdot M} \cdot\ z_n\ - \frac{Q_{o,L} \cdot \rho_L}{\sum x \cdot M} \cdot\ x_n - \frac{Q_{o,V} \cdot \rho_V}{\sum K \cdot x \cdot M} \cdot K_n\big(T(t)\big) \cdot x_n \tag{21}$$

$$Q_{o,V} = \left\{ \left[Q_{j,i} . \left(\frac{z_A}{MM_A} + \frac{z_B}{MM_B} + \frac{z_C}{MM_C} + \frac{z_D}{MM_D} + \frac{z_I}{MM_I} \right) \right] - \frac{P.V_V}{R.T^2} \frac{dT}{dt} \right\} .\ MM_V \tag{22}$$

$$y_n(t) = K_n\big(T_{V-02}(t)\big) \cdot x_n(t) \tag{23}$$

$$K_n\big(T_{V-02}(t)\big) = \frac{P_n^{sat}(T_{V-02}(t))}{P} \tag{24}$$

$$P_n^{sat}\big(T_{V-02}(t)\big) = e^{\left(A_n - \frac{B_n}{T_{V-02}(t) + C_n}\right)} \tag{25}$$

$$\sum \frac{z_i(K_i - 1)}{1 + \beta(K_i - 1)} + \frac{z_C}{\beta} = 0 \tag{26}$$

Appendix B

Constant	Values	Constant	Values
$Q_{V\text{-}01,1}$	0.01 m³/s	$U_{E\text{-}01}$	1,000W/m².K
$Q_{V\text{-}01,2}$	0.035m³/s	$U_{E\text{-}02}$	1,000W/m².K
$Q_{U\text{-}01}$	0.005m³/s	R	8.314J/mol.K
$Q_{U\text{-}02}$	0.005m³/s	$E_{a,1}$	1,800J/mol
$Q_{R\text{-}01,4}$	0.002m³/s	$E_{a,2}$	900J/mol
$Q_{R\text{-}01,5}$	0.037m³/s	k_1	0.45m³/mol.s
$Q_{R\text{-}02,6}$	0.002 m³/s	k_2	1.35m³/mol.s
$Q_{R\text{-}02,7}$	0.039m³/s	ΔH_1	400,000J/mol
$Q_{V\text{-}02,10}$	0.025m³/s	ΔH_2	150,000J/mol
$Q_{V\text{-}02,11}$	0.014m³/s	P_{V02}	101.325kPa
$C_{V\text{-}01,A,1}$	6mol/m³	$\rho_{L_{V-02,L}}$	850kg/m³
$C_{R\text{-}01,C,4}$	10mol/m³	$\rho_{V_{V-02,V}}$	1.5kg/m³
$C_{R\text{-}02,I,6}$	10mol/m³	M_A	0.034kg/mol
$a_{V\text{-}01}$	1.5m²	M_B	0.017kg/mol
$a_{U\text{-}01}$	4.88375m²	M_C	0.028kg/mol
$a_{U\text{-}02}$	9.767325m²	M_I	0.002kg/mol
$a_{R\text{-}01}$	1.5m²	M_D	0.015kg/mol
$a_{R\text{-}02}$	1.5m²	$cp_{V-02,i}$	2.2kJ/kg.K
$T_{V\text{-}01,1}$	303.15K	$cp_{V-02,V}$	2.2kJ/kg.K
$T_{R\text{-}01,4}$	318.15K	A_A	16.8792
$T_{R\text{-}02,6}$	338.15K	B_A	3,803.98
$T_{E\text{-}01,s,i}$	473.15K	C_A	230.71
$T_{E\text{-}02,s,i}$	303.15K	A_B	16.5785
$V_{E\text{-}01,p}$	0.003099m³	B_B	3,626.55
$V_{E\text{-}01,s}$	0.08571m³	C_B	239.76
$V_{E\text{-}02,p}$	0.0062022m³	A_I	15.0717
$V_{E\text{-}02,s}$	0.152176m³	B_I	3,580.8
$V_{V\text{-}02}$	23.3m³	C_I	224.65
$\rho_{E\text{-}01}$	1,000kg/m³	A_D	16.1154
$\rho_{E\text{-}02}$	1,000kg/m³	B_D	3,483.67
$h_{E\text{-}01}$	4,186.8J/kg.K	C_D	204.64
$h_{E\text{-}02}$	4,186.8J/kg.K	$\Delta h_{V\text{-}02}$	1,000kJ/kg