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# Comparative Analysis Between PI Conventional and Cascade Control in Heater-PFR-Series

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**Abstract**

*The processes of heater-reactor-series are often used in the chemical process industries. Since the reactor outlet temperature increases as a result of the exothermic reaction, this temperature needs to be maintained below the upper temperature limit in order to reduce the formation of by-product and to prevent fouling in the next process heat exchangers. The goals of this work are to propose the new cascade controller and to compare with the previous Proportional-Integral (PI) controller in heater-plug-flow-reactor-series (Heater-PFR-series) to produce benzene through the reaction of hydrodealkylation of toluene (HDA). The three control models will be discussed in this paper. Model A is the previous PI controller to keep the reactor inlet temperature, model B is the new PI controller to maintain the reactor outlet temperature, and model C is the new Cascade controller to keep both reactor outlet and inlet temperatures. The main disturbance of feed flowrate was made to examine the three control models. The three control models were rigorously examined in UniSim Design R451. The PI control parameters were tuned by using "autotuner" mode of UniSim. As shown in dynamic simulation study, the three control models with its tuning parameters gave the fast and stable responses. Integral of the absolute value of the error (IAE) at the reactor outlet temperature for model A, B, and C are 3305, 2487, and 931 °C, respectively. This study revealed that model C (Cascade) acted very well to the main disturbance change in feed flowrate, responses of model C were better and faster than those in A and B.*

**Keywords:** *cascade control; dynamic simulation; PFR; PI conventional; UniSim*

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## INTRODUCTION

The process of heater-reactor-series is frequently met in the chemical process industries, such as isomerization process to convert normal butane into isobutane, hydrodealkylation of toluene (HDA) to produce benzene, and vinyl acetate process (Luyben *et al.*, 1999). However, the outlet and inlet reactor temperatures really affect the reactor performance and the produced products. The reactor temperature that is out of range can cause the problems, for example the reactor temperature should not be higher than upper limit to avoid the formation of by-product or damage to the catalyst. And vice versa, the reactor temperature that is less than lower limit can reduce the rate of reaction, so that the main-product will decrease. When the disturbance enters the reactor, it has to propagate through the reactor and make the reactor outlet temperature deviate from its desired value before feedback controller takes corrective action. This motivates us to study the control strategies for a heater-reactor-series process, with implementation of feedback control (FBC) conventional and cascade control strategies, in order to achieve the desired operating conditions, and the fast and stable responses.

There have been many contributions to the design of cascade control. To name but a few, far back to year of 1988, Yu (1988) has proposed a design procedure based on the parallel cascade control structure for disturbance-rejection. Yu (1988) stated that the proposed approach offered a simple and effective alternative for disturbance-rejection. Urrea-Garcıa *et al*. (2015) has proposed the control structure which allows the controller to adapt to temperature error variations along the tubular reactor. They explored the application of a variable control structure for tubular reactors, based on multiple temperature measurements. Ahmed *et al*. (2013) has utilized cascade control strategy to control the temperature inside a jacketed exothermic continuous stirred tank reactor (CSTR). Recently, Ma *et al* (2019) have studied on the output tracking with disturbance attenuation for cascade control systems subject to network constraint. They investigated the simultaneous design of the primary and the secondary controllers to achieve the output tracking performance of the networked cascade control system (NCCS) with disturbance attenuation.

Along with the growth of chemical process industries, the use of computational tools such as Matlab, Scilab, Aspen-Hysys, UniSim is therefore very important to carry out dynamic simulation and explore its dynamic behavior. Several studies (Hermawan and Haryono 2012, Hermawan and Puspitasari 2018, and Hermawan *et al*. 2016) have used Scilab to carry out steady state and dynamic simulation in a 10 L mixing tank. Hermawan (2005) has used Hysys to carry out both steady state and dynamic simulations in HDA process with energy integration schemes. Wongsri and Hermawan (2005) also utilized Hysys to examine the proposed control structure in a complex HDA plant. Wongsri and Hermawan (2005) has proposed heat pathway heuristics (HPH) in conjunction with the plantwide control procedure given by Luyben *et al*. (1997) to model heat pathway management systems and the control configuration of a complex energy integrated HDA plant. The comparison between SIMULINK and HYSYS simulation and control in CSTR has been presented by Taimor (2016). He pointed that SIMULINK as laboratory tool presented higher satisfaction with HYSYS. Recently, Hermawan (2020) has comprehensively explained the use of UniSim in steady state mode.

The goal of this work is to device control configurations in heater-plug-flow-reactor-series (Heater-PFR-Series) to produce benzene through the HDA reaction. Proportional-Integral (PI) conventional and Cascade control configurations will be used and compared for controlling the reactor outlet/inlet temperature. The three control models will be used in the heater-PFR-series. Model A is the previous PI controller to keep the reactor inlet temperature that be presented by Hermawan (2005), Wongsri and Hermawan (2005), and Luyben (2002). Model B is the new PI controller to maintain the reactor outlet temperature, and model C is the new Cascade controller to keep both reactor outlet and inlet temperatures. The main disturbance of feed flowrate would be made to examine the three control models. UniSim Design R451 simulator from Honeywell is utilized to carry out both the steady state and dynamic simulations.

## MATERIAL AND METHOD

This work will be carried out through literature study and process simulation using UniSim Design R451 simulator. The HDA process to produce benzene at high reactor temperature is chosen as a case study. In order to achieve the desired goals, this work is carried out through the following stages:

### Steady State Simulation

First, a steady-state model of Heater-PFR-Series is built in UniSim, using equipment design information, mainly taken from Hermawan (2005) and Luyben *et al* (1999). The Peng-Robinson model is chosen in this simulation for the calculation of the physical properties because of its reliability in predicting the properties of most hydrocarbon-based fluids over a wide range of operating conditions (Wongsri and Hermawan, 2005). UniSim flowsheet of Heater-PFR-Series is shown in Figure 1.

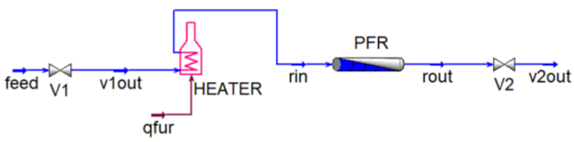


Figure 1. UniSim flowsheet of Heater-PFR-series

The feed stream with conditions as shown in Table 1 is heated in Heater until the target temperature of 1150 ºF. Then, the heated stream from Heater is flowed to the PFR for doing the two vapor-phase reactions as written in equation (1) and (2).

$$\underset{\text{Toluene}}{C_7H_8} + \underset{\text{Hydrogen}}{H_2} \rightarrow \underset{\text{Benzene}}{C_6H_6} + \underset{\text{Methane}}{CH_4} \quad (1)$$

$$\underset{\text{Benzene}}{2C_6H_6} \leftrightarrow \underset{\text{Diphenyl}}{C_{12}H_{10}} + \underset{\text{Hydrogen}}{H_2} \quad (2)$$

Benzene is a main-product, while Methane and Diphenyl are by-products.

The two kinetic expressions are modeled with standard Arhenius kinetic expression available in UniSim. The two-reaction rates ($R_1$ and $R_2$) are the functions of partial pressures (Luyben 2002) and given in equation (3) and (4).

Table 1. Feed conditions

| Stream Name | feed |
|---|---|
| Temperature [F] | 1100 |
| Pressure [psia] | 556 |
| Molar Flow [lbmol/hr] | 5000 |
| Component mole fraction: | |
| Hydrogen | 0.4169 |
| Methane | 0.4922 |
| Benzene | 0.0080 |
| Toluene | 0.0829 |
| Diphenyl | 0.0000 |

$$R_1 = \left(3.686\text{x}10^6\ e^{\frac{-90800}{RT}}\right) P_T P_H^{0,5} \qquad (3)$$

$$R_2 = \left(9\text{x}10^4 e^{\frac{-90800}{RT}}\right) P_B^2 - \left(2.553\text{x}10^5 e^{\frac{-90800}{RT}}\right) P_D P_H \qquad (4)$$

where the reaction base is partial pressure (psia), vapor phase, $R_1$ and $R_2$ have units of lbmole/cuft/minute. $P_T$, $P_H$, and $P_D$ are partial pressure of Toluene, Hydrogen, and Diphenyl, respectively (in psia). Activation energi ($E$) is in Btu/lbmole and the temperature is in Rankin.

For further dynamic simulation, we need some informations about equipment's specification. These spesifications is directly taken from Luyben *et al* 1999, Hermawan 2005, Luyben 2002, and Wongsri and Hermawan 2005. The specifications of PFR, and Heater are listed in Table 2, and 3, respectively. PFR has pressure drop of 17 psig and is operated adiabatically. Heater has volume of 300 cuft and pressure drop of 5 psig.

Table 2. PFR specification and operating condition

| | |
|---|---|
| Temperature (F) | 1150 |
| Pressure (psia) | 521 |
| Diameter (in) | 9.53 |
| Length (ft) | 57 |
| Pressure drop (psi) | 17 |
| Process | adiabatic |

Table 3. Heater's specification

| | |
|---|---|
| Model | Supplied duty |
| Pressure drop (psig) | 5 |
| Volume (cuft) | 300 |

The plumbing system, specifically control valve, need to be inserted in UniSim flowsheet before switching to dynamics (Luyben, 2002). We choose Masoneilan Valve with type of DP Globe, V-Port, and *quick opening*. The valve coefficient (Cv) can be obtained by using "size valve" menu available in UniSim. The control valve's specifications are listed in Table 4.

Table 4. Control valve's specification

| Valve | Stream | Press. Drop (psi) | Cv (USGPM) | Size (in) |
|---|---|---|---|---|
| V1 | feed | 30 | 471.2 | 10 |
| V2 | rout | 30 | 514.8 | 10 |

## Design of Control Configuration

The FBC with PI-Conventional and Cascade control model would be applied to Heater-PFR-Series. In general, there are three controllers in Heater-PFR-Series, they are Flow Controller (FC), Temperature Controller (TC), and Pressure Controller (PC), for maintaining the feed flowrate, the reactor outlet/inlet temperature, and the reactor outlet pressure, constant at its set-point, respectively. In order to compare with the previous work, the three control models, i.e. model A, B, and C, are proposed in this work.

### Model A

Figure 2 shows the model A in the control configuration of Heater-PFR-Series. Model A is PI controller in the reactor inlet temperature as given by the previous work of Luyben 1999, Hermawan 2005, Luyben 2002, and Wongsri and Hermawan 2005. The control configuration in model A has the following loops:

1. Feed stream is flow controlled by controller FC1. The output (OP) target of FC1 is valve V1.
2. The reactor inlet temperature ($T_{rin}$) is controlled by controller TC1, and its OP target is heater duty ($q_{fur}$). Direct Q model is selected to control $T_{rin}$ by manipulating a heat removal rate ($q_{fur}$). Range of $q_{fur}$ is in between 0 and 8 MMBtu/hr.
3. The reactor outlet pressure is controlled by controller PC1. The OP signal changes the position of valve V2, which manipulates the reactor outlet flowrate ("rout" flowrate).

In Model A, controller TC1 is used to keep the reactor intlet temperature ($T_{rin}$) constant at its set-point and lets the reactor outlet temperature ($T_{rout}$) changes as the input disturbance changes. Therefore, this can cause the reactor outlet temperature to rise higher than the upper temperature limit due to the exothermic reaction. This comes up the problems, namely the formation of by-products and fouling in the next process heat exchangers (Luyben *et al* 1999).

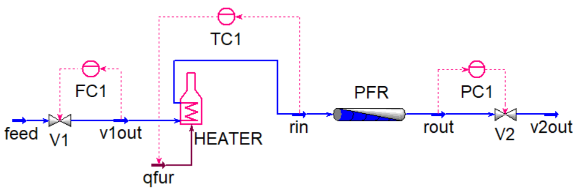


Figure 2. Model A in the control configuration of Heater-PFR-Series

### Model B

The model B in the control configuration of Heater-PFR-Series is shown in Figure 3. The model B is similar with model A, except in the controller TC1. In model B, controller TC1 is used to keep temperature $T_{rout}$ constant at its set-point and lets temperature $T_{rin}$ changes as the input disturbance changes. Therefore, controller TC1 will react only the process has been upset (Smith and Corripio, 1997).

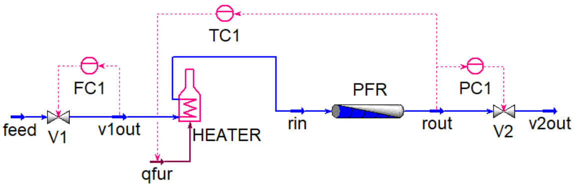


Figure 3. Model B in the control configuration of Heater-PFR-Series

### Model C

Model C is shown in Figure 4. The model C is similar with models A and B, except in the controller TC1. In order to overcome the disadvantage of PI-Conventional, a secondary measurement and a secondary feedback controller should be employed. These secondary instruments are used to measure and control temperature $T_{rin}$, so that it recognizes the upset condition sooner. This approach is called Cascade control (Seborg *et al*., 2011). In Cascade control configuration, there is one manipulated variable, i.e. heater duty $q_{fur}$ and two measurements, i.e. temperatures $T_{rout}$ and $T_{rin}$.

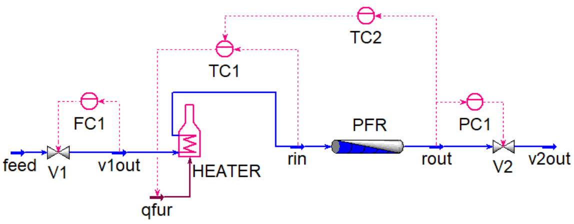


Figure 4. Model C in the control configuration of Heater-PFR-Series

### Dynamic Simulation

First of all, before switching to dynamics, the equpments' size and the plumbing must be specified. The Unisim Dynamic Assistant can be used to know the lack of required data or information in dynamic simulation. Dynamic Assistant will give some suggestions or considerations so that dynamic simulation can be carried out well.

Another important thing to support dynamic simulation is the value of FBC parameters, such as proportional gain ($K_c$), integral time constant ($\tau_I$), and derivative time constant ($\tau_D$). The value of these parameters greatly affects the stability of the control system. The PI controller would be used in Heater-PFR-Series. Therefore, the only two parameters, i.e. $K_c$ and $\tau_I$, need to be tuned properly. The PI controller parameters are tuned by using "autotuner" mode of UniSim, and its results are directly used in dynamic simulation.

In order to examine the three models of control configurations and evaluate the resulted PI controller parameters, the closed loop dynamic simulation should be carried out. In addition, the closed loop dynamic simulation is also aimed to examine the robustness of the three models of control configurations to a change in input disturbance. In this work, the feed flowrate ($f_{feed}$) is selected as a main disturbance variable because it often occurs in the chemical plant especially when it wants to increase or decrease the rate of production. In addition, the other variable disturbance, i.e. the feed temperature ($T_{feed}$) is also made just to explore its dynamic behavior. The two disturbances are made based on step fuction.

## RESULTS AND DISCUSSION

### Steady State Simulation Results

For heating of feed stream 5000 lbmole/hour from 1100 ºF to 1150 ºF, the heater duty ($q_{fur}$) is 4.05 MMBtu/hour. After heating process, the feed stream enters PFR to carry out the two reactions. The conversion of the first and second reaction are $X_1$=69.60% and $X_2$=24.22%, respectively. Since the first reaction is exothermic (heat of reaction -18000 Btu/lbmole), and the second is slightly edothermic (heat of reaction 3500 Btu/lbmole), the reactor outlet temperature rises to 1222 ºF.

These steady state simulation results of PFR are compared with those in Hermawan (2005) and Luyben (2002). The HYSYS simulator was used in both Hermawan (2005) and Luyben (2002). These results are the same as those given by Hermawan (2005) and Luyben (2002). The steady state simulation results of Heater-PFR can be viewed in UniSim workbook as shown in Figure 5.

| Name | feed | rout | v1out | v2out | rin |
|---|---|---|---|---|---|
| Vapour Fraction | 1,0000 | 1,0000 | 1,0000 | 1,0000 | 1,0000 |
| Temperature [F] | 1100 | 1222 | 1100 | 1222 | 1150 |
| Pressure [psia] | 556,0 | 504,0 | 526,0 | 474,0 | 521,0 |
| Molar Flow [lbmole/hr] | 5000 | 5000 | 5000 | 5000 | 5000 |
| Mass Flow [lb/hr] | 8,500e+004 | 8,500e+004 | 8,500e+004 | 8,500e+004 | 8,500e+004 |
| Heat Flow [MMBtu/hr] | -3,132 | 0,9235 | -3,132 | 0,9235 | 0,9235 |
| Master Comp Mole Frac (Hydrogen) | 0,416900 | 0,360172 | 0,416900 | 0,360172 | 0,416900 |
| Master Comp Mole Frac (Methane) | 0,492200 | 0,549897 | 0,492200 | 0,549897 | 0,492200 |
| Master Comp Mole Frac (Benzene) | 0,008000 | 0,063760 | 0,008000 | 0,063760 | 0,008000 |
| Master Comp Mole Frac (Toluene) | 0,082900 | 0,025203 | 0,082900 | 0,025203 | 0,082900 |
| Master Comp Mole Frac (BiPhenyl) | 0,000000 | 0,000969 | 0,000000 | 0,000969 | 0,000000 |

Figure 5. Workbook of Heater-PFR-Series

### Controller Acting and Tuning Results

Another important thing for dynamic simulation is selection of controller action. There are two types of controller actions, they are reverse and direct. The flow controller (FC1) would be reverse acting since an increase in flow should result in moving the valve V1 toward the closed position (increasing PV decreases OP). The reactor intlet/outlet temperature controllers (TC1/TC2) would also be reverse acting, increasing the reactor intlet/outlet temperature (PV) decreases the heater duty (OP). Unlike FC1, TC1, and TC2, the reactor outlet pressure controller (PC1) would be direct acting since an increase in pressure should result in moving the valve V2 toward the open position (increasing PV increases OP). The actions of controllers in all models are shown in Table 5 to 7.

Table 5. Controller's parameters in model A

| Controller | Action | Kc | $\tau_I$ (minute) | Set-point |
|---|---|---|---|---|
| FC1 | Reverse | 0.102 | 4.12E-3 | 5000 lbmole/hr |
| TC1 | Reverse | 1.74 | 8.27E-2 | 1150 °F |
| PC1 | Direct | 4.08 | 3.11E-2 | 504 psia |

Table 6. Controller's parameters in model B

| Controller | Action | Kc | $\tau_I$ (minute) | Set-point |
|---|---|---|---|---|
| FC1 | Reverse | 0.102 | 4.12E-3 | 5000 lbmole/hr |
| TC1 | Reverse | 0.639 | 1.68 | 1221 °F |
| PC1 | Direct | 4.08 | 3.11E-2 | 504 psia |

Table 7. Controller's parameters in model C

| Controller | Action | Kc | $\tau_I$ (minute) | Set-point |
|---|---|---|---|---|
| FC1 | Reverse | 0.102 | 4.12E-3 | 5000 lbmole/hr |
| TC1 | Reverse | 1.74 | 8.27E-2 | ***Remote set-point*** |
| TC2 | Reverse | 0.132 | 1.83 | 1221 °F |
| PC1 | Direct | 4.08 | 3.11E-2 | 504 psia |

The controller parameters, i.e. proportional gain ($K_c$), and integral time constant ( $\tau_I$), must be tuned well before carry out dynamic simulations. The values of $K_c$ and $\tau_I$ resulted by "autotuner" mode of UniSim for model A, B, and C are listed in Table 5, 6, and 7, respectively. These results are then directly used in dynamic simulation. Other than that, the ranges of sensor/transmitter should also be determined, and the ranges are listed in Table 8.

Table 8. Range of CV (PV)

| Controller | Range | Unit |
|---|---|---|
| FC1 | 3000 - 7000 | lbmole/hr |
| TC1 | 900 - 1400 | °F |
| TC2 | 900 - 1400 | °F |
| PC1 | 404 - 604 | psia |

**Dynamic Simulation Results**

In order to examine the three models of control configurations and to evaluate the resulted controller parameters, two disturbance system are made and discussed as follows:

**Dynamic Responses to a Set-point Changes in Controller FC1**

The main disturbance of set-pont changes in controller FC1 as shown in Figure 6 can be made as follows:

- At time equal 10 minutes, the set-point of feed flow controller FC1 is changed from 5000 lbmole/hour to 6000 lbmole/hour.
- At time equal 60 minutes, the set-point of FC1 is changed from 6000 lbmole/hour to 4000 lbmole/hour.

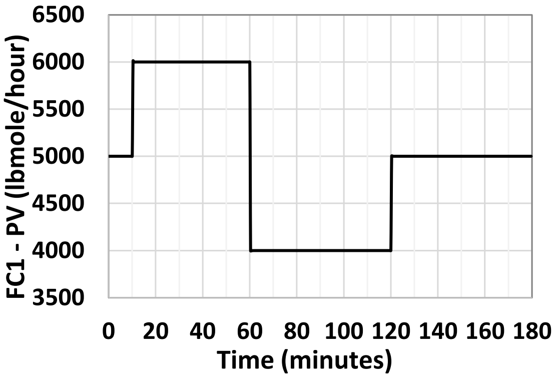


Figure 6. Set-point Changes in Controller FC1

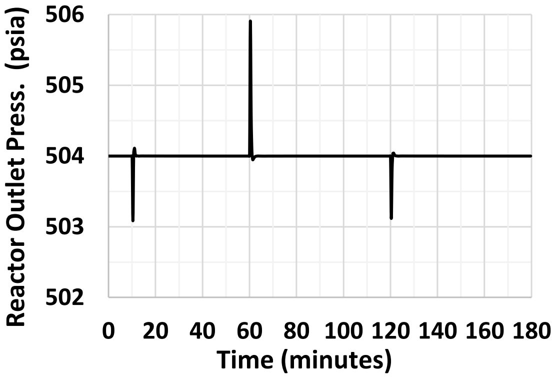


Figure 7. Dynamic Response of Reactor Outlet Pressure to Set-point Changes in Controller FC1

- At time equal 120 minutes, the set-point of FC1 is changed from 4000 lbmole/hour to 5000 lbmole/hour.

Dynamic responses of control system in Heater-PFR-Series to set-point changes in controller FC1 are shown in Figure 7 to 12. The reactor outlet pressure $P_{rout}$ (Figure 7) decreases (and increases) as feed flowrate increases (and decreases), but controller PC1 can return the pressure $P_{rout}$ to its set point of 504 psia quickly.

Figure 8 and 9 show dynamic response of the reactor outlet ($T_{rout}$) and inlet ($T_{rin}$) temperature to set-point changes in controller FC1. In general, first the temperatures $T_{rout}$ and $T_{rin}$ drop as the feed flowrate increases, and then their values can be returned to their set-points by increasing heater duty ($q_{fur}$) from to 4.05 to 5.40 MMBtu/hour as shown in Figure 10. As can be seen in Figures 11 and 12, the conversions of the firt and second reactions ($X_1$ and $X_2$) decrease as the reactor temperature decreases. When the feed flowrate decreases from 6000 to 4000 lbmole/hour at time equals 60 minutes (Figure 6), the temperatures $T_{rout}$ and $T_{rin}$ (Figures 8 and 9), and the conversions $X_1$ and $X_2$ (Figures 11 and 12) increase quickly, and their values can be returned to their set-points by decreasing the heater duty (Figure 10).

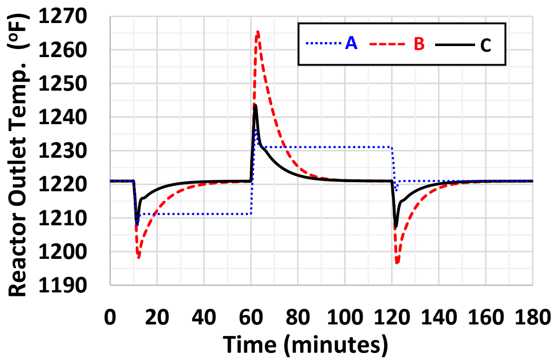


Figure 8. Dynamic Response of Reactor Outlet Temperature to Set-point Changes in Controller FC1

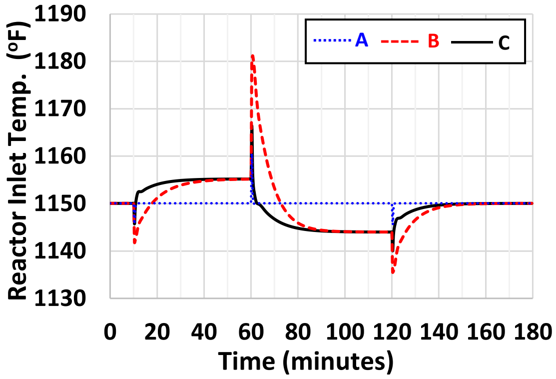


Figure 9. Dynamic Response of Reactor Intlet Temperature to Set-point Changes in Controller FC1

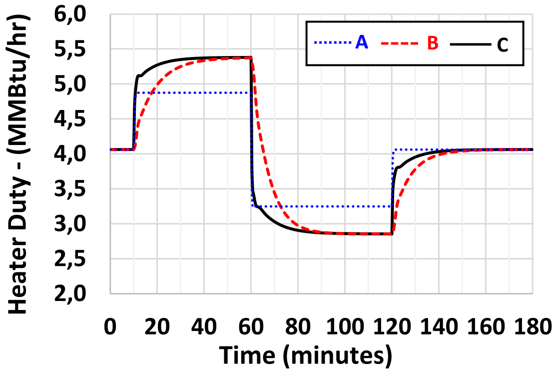


Figure 10. Dynamic Response of Heater Duty to Set-point Changes in Controller FC1

The comparison of dynamic responses between the previous work in model A (PI controller of the reactor inlet temperature) and the current work in model B (PI controller of the reactor outlet temperature) and C (Cascade controller of both the reactor outlet and inlet temperatures) are represented by the dotted, dashed and solid line, respectively (Figures 8 to 12). But model B produces a bigger overshoot than model A and C. The performance of control models can be determined by calculating the integral of the absolute value of the error (IAE). The calculated IAE at the reactor outlet temperature for model A, B, and C are 3305, 2487, and

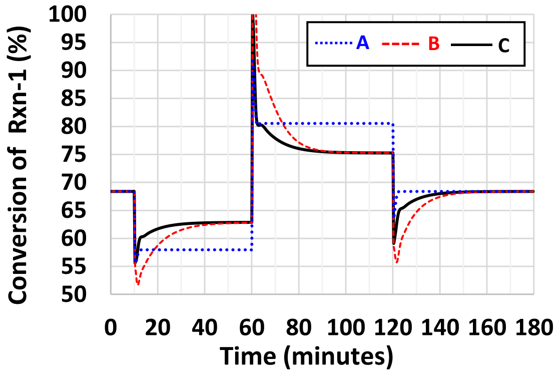


Figure 11. Dynamic Response of Conversion of Reaction-1 to Set-point Changes in Controller FC1.

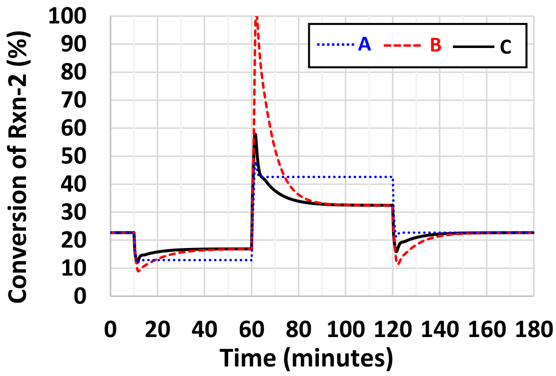


Figure 12. Dynamic Response of Conversion of Reaction-2 to Set-point Changes in Controller FC1

931 ºC, respectively. This indicates that the dynamic responses of model C are faster than models A and B.

At time 60 to 120 minutes, the conversion $X_2$ resulted by model A is bigger than that resulted by model C (Figure 12). This implies that PI conventional produces more by-product (Diphenyl) than Cascade. Cascade can improve the response of PI Conventional by measuring temperature $T_{rin}$ and taking control action before its effect has been felt by the reacting mixture in PFR. Under cascade configuration, the effect of the disturbance on the primary plant is consequently suppressed by the fast dynamics of the secondary controller. This agree with those in Stephanopoulos (1984), Smith and Corripio (1997), and Ma *et al* (2019).

**Dynamic Responses to Disturbance Changes in Feed Temperature**

In addition, the other disturbance, i.e. the feed temperature was also made just to explore the dynamic behaviour of the three control models. Disturbance changes in the feed temperature as shown in Figure 13 can be made as follows:

- At time equal 10 minutes, the feed temperature is changed from 1100 ºF to 1110 ºF.
- At time equal 60 minutes, the feed temperature is changed from 1110 ºF to 1090 ºF.
- At time equal 120 minutes, the feed temperature is changed from 1090 ºF to 1100 ºF.

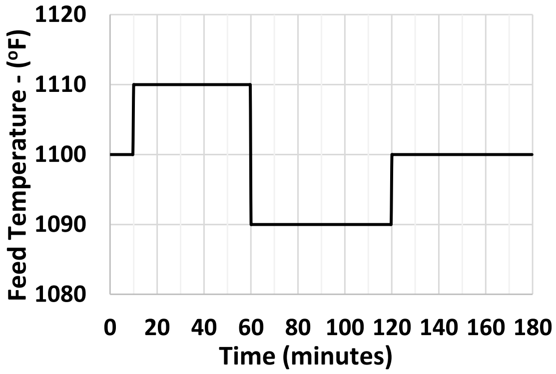


Figure 13. Disturbance Changes in Feed Temperature

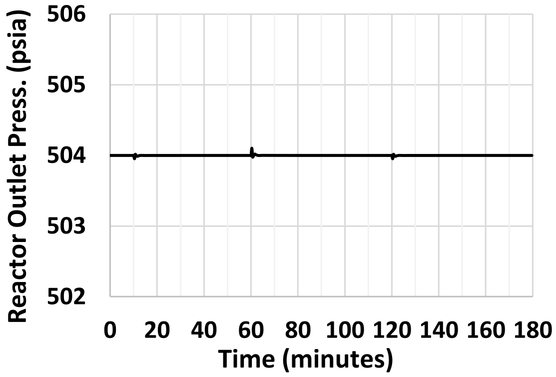


Figure 14. Dynamic Response of Reactor Outlet Pressure to Disturbance Changes in Feed Temperature

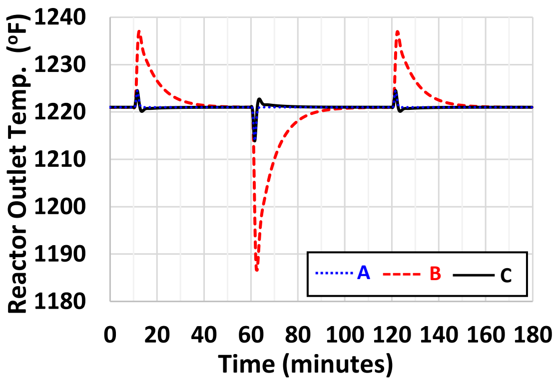


Figure 15. Dynamic Response of Reactor Outlet Temperature to Disturbance Changes in Feed Temperature

Figures 14 to 19 show dynamic responses of control system in Heater-PFR-Series to disturbance changes in the feed temperature. Dynamic response of the reactor outlet pressure $P_{rout}$ (Figure 14) is very fast. Controller PC1 can maintain $P_{rout}$ at its set point of 504 psia well.

Dynamic responses of the reactor outlet and inlet temperature ($T_{rout}$ and $T_{rin}$) to disturbance changes in the feed temperature are shown in Figure 15 and 16, respectively. In general, first the temperatures $T_{rout}$ and $T_{rin}$ increase as the feed temperature increases, and then their values can be returned to their set-points by decreasing heater duty ($q_{fur}$) from to 4.05 to 3.25 MMBtu/hour as shown in Figure 17. The conversions of the firt and second reactions ($X_1$ and $X_2$) increase as the reactor temperature increases. When the feed temperature decreases from 1110 to 1090 ºF at time equals 60 minutes (Figure 13), the temperatures $T_{rout}$ and $T_{rin}$ (Figures 15 and 16), and the conversions $X_1$ and $X_2$ (Figures 18 and 19) decrease quickly. However, the temperature controller can return values of the process variables to their set-points by increasing the heater duty (Figure 17).

Figures 15 to 19 also show the comparison of dynamic responses between models A, B and C. The dotted, dashed, and solid line in Figures 15 to 19 represent the responses of model A, B, and C, respectively. The dynamic responses of models A and C are very close. The PI conventional in model B produces a bigger overshoot than models A and C.

Again, the dynamic reponses of Cascade are faster than PI Conventional. As shown in dashed line in Figure 18, model B produces a bigger overshoot than model C, the conversion $X_1$ resulted by model B is smaller than that resulted by model C. This implies that PI Conventional produces less main product (Benzene) than Cascade.

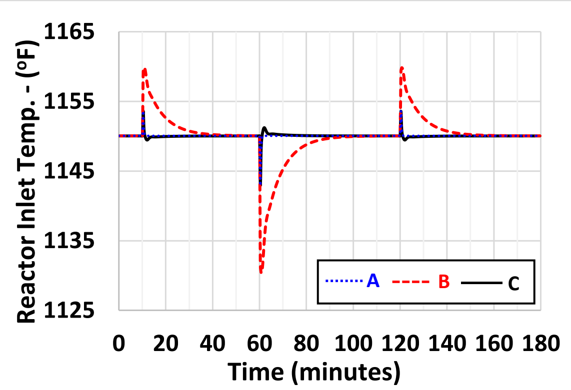


Figure 16. Dynamic Response of Reactor Inlet Temperature to Disturbance Changes in Feed Temperature

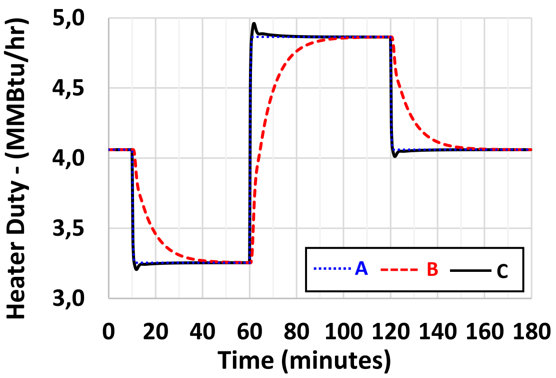


Figure 17. Dynamic Response of Heater Duty to Disturbance Changes in Feed Temperature

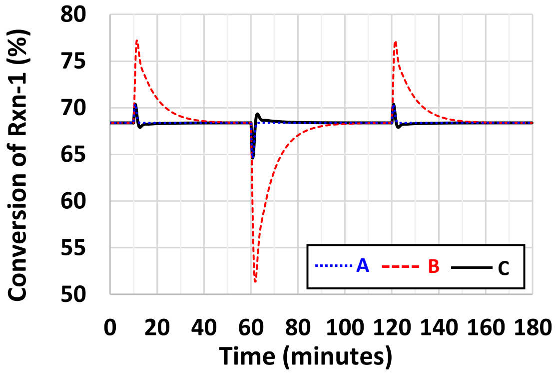


Figure 18. Dynamic Response of Conversion of Reaction-1 to Disturbance Changes in Feed Temperature

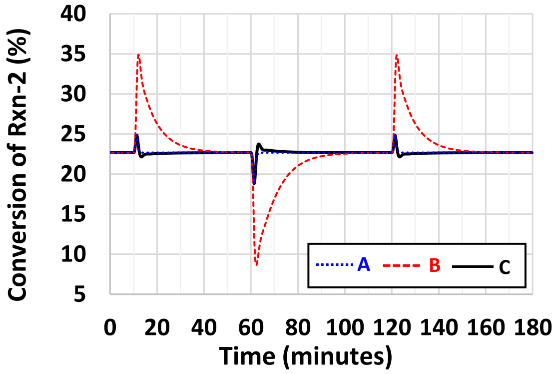


Figure 19. Dynamic Response of Conversion of Reaction-2 to Disturbance Changes in Feed Temperature

As explained in Seborg *et al.* (2011), a disadvantage of PI Conventional is that corrective action for disturbances does not begin until after the controlled variable deviates from the set point. Therefore, in order to overcome this problem, a secondary measurement point and secondary temperature feedback controller should be employed in the PFR. This approach is known as cascade control.

Cascade control system is often utilized when the output of the desired primary process is controlled by the output of a secondary control process. Under this configuration, the effect of the disturbance on the primary plant is consequently suppressed by the fast dynamics of the secondary controller (Ma *et al* 2019).

## CONCLUSION

Study on the HDA process to produce benzene, tuning of PI parameters, dynamic simulation and control in Heater-PFR-series have been successfully done through the closed loop simulation using UniSim Design R451 simulator. The new control model of cascade control has been applied to Heater-PFR-Series and compared with the previous model of PI conventional control. The closed loop dynamic behaviors of the control models have also been explored and compared.

According to our dynamic simulation results, the resulted controller gains ($K_c$) and integral time constants ($\tau_I$) were able to produce the fast and stable responses to the main disturbance of set-point changes in flow controller and the other disturbance change in feed temperature. The most robust control is obtained when a cascade control is employed. Cascade control can improve the response of PI conventional by measuring the reactor inlet temperature and taking control action before its effect has been felt by the reacting mixture in PFR.

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## NOTATION

$E_{1,2}$ : the activation energy of reaction 1, and 2 in Btu/lbmole.
$f_{feed}$ : the feed flowrate in lbmole/hour.
$K_c$ : controller gain
$P_B$ : the partial pressure of Benzene in psia
$P_D$ : the partial pressure of Diphenyl in psia
$P_H$ : the partial pressure of Hydrogen in psia
$P_{rout}$ : reactor outlet pressure in psia
$P_T$ : the partial pressure of Toluene in psia
$q_{fur}$ : the heater duty in MMBtu/hour
$R_{1,2}$ : the reaction rates of reaction 1, and 2 in lbmole/cuft/minute.
$R$ : constant of gas universal used in equation $R_{1,2}$.
$T$ : Gas temperature used in equation $R_{1,2}$ in Rankin.
$T_{feed}$ : the feed temperature in ºC
$T_{rin}$ : the reactor inlet temperature in ºC.
$T_{rout}$ : the reactor outlet temperature in ºC.
$X_1$ : the conversion of reaction-1 in %
$X_2$ : the conversion of reaction-2 in %
$\tau_I$ : integral time constant in minute
$\tau_D$ : derivative time constant in minute

## ABBREVIATION

CV : Controlled Variable
DV : Disturbance Variable
FBC : Feedback Control
FC : Flow Controller
HDA : Hydrodealkylation of Toluene
IAE : The Integral of the Absolute value of the Error
MV : Manipulated Variable
OP : Output
PFR : Plug Flow Reactor
PC : Pressure Controller
PI : Proportional Integral
PV : Process Variable
SP : Set Point
TC : Temperature Controller

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