Application of Reverse Flow Reactor for Vent Gas Emission Reduction in Catalytic Oxidation Unit at Purified Terephthalic Acid Plant

F M Hanafiah and Y W Budhi*
Department of Chemical Engineering, Faculty of Industrial Technology, Institut Teknologi Bandung, Jalan Ganesha 10 Bandung 40132, Indonesia

E-mail: Y.Wibisono@che.itb.ac.id

Abstract. PTA (Purified Terephthalic Acid) manufacturing process generates waste gases containing VOC (Volatile Organic Compounds) that have negative health effects, one of it is benzene. Catalytic oxidation method consists of a recuperative (need external energy to reach the reaction temperature) and reverse flow reactor (RFR) which uses the principles of autothermal (requires no external energy). Fuel consumption for heating the feed gas is a major operating cost in the catalytic oxidation, so that the fuel reduction becomes important. The objective of this study is to assess the technical possibility of RFR to lower the fuel consumption; and to assess the operating conditions for autothermal conditions. The mathematical model consisted of the unsteady state mass and energy balances and was numerically solved using a software package of FlexPDE version 6.32. The recuperative system was simulated in the steady state condition, while RFR with the same amount of catalyst was simulated in the unsteady state condition. As the key parameter, the switching time was varied to consider the performance of RFR. At various switching times and inlet concentrations, the autothermal was achievable even heat extraction was required to prevent the catalyst overheat. At the feed gas linear velocity of 1.2 m/s and the switching time of 7.5 s or higher, RFR provides energy saving that is equivalent to US$ 0.1208/ton feed gas up to US$ 0.7248/ton of feed gas compared to recuperative system.

1. Introduction
Terephthalic acid (TA) is an organic compound with formula C₆H₄(CO₂H)₂ and widely used as commodity chemical. Terephthalic acid in a purified condition is well known as Purified Terephthalic Acid (PTA). Almost all PTA productions in the world are used as raw material for the polyethylene terephthalate fibre (PET polyester) manufacturer. From 1999 to 2009, PTA worldwide demand grows 7-8% per year and mostly it is influenced by Asia market which is up to 70% [1].

The exhaust gases of the PTA process typically contain CO, methyl bromide, benzene, and others VOC (Volatile Organic Compounds). According to the US Environmental Protection Agency (EPA), VOC refers to any compounds of carbon, excluding carbon monoxide, carbon dioxide, carbonic acid, metallic carbides or carbonates and ammonium carbonate. The VOC may participate in atmospheric photochemical reactions, except those designated by EPA as having negligible photochemical reactivity. VOC indicate negative health effects, such as respiratory disorders, allergies, immune effects in infants and children, and carcinogenic [2]. Therefore, the VOC must be treated before discharging to the atmosphere. Worldwide, the environmental regulation shows more stringent, especially in USA and
Europe. The PTA plant requires to reduce VOC containing in its exhaust gases. Some PTA plants in Indonesia install a catalytic converter unit to reduce the VOC before venting the exhaust gas.

The application of carbon adsorption may capture the VOC in the exhaust gas up to 97% recovery but is not enough for most PTA plants. Thermal oxidation may reduce VOC and CO containing in the vent gases up to 99% or even more. However, more fuel consumption is required to reach the operating temperature above 800 °C. The catalytic oxidation process destructs VOC and CO simultaneously and offers a fuel consumption saving, due to lower operating temperatures, typically around 275–500 °C [3].

The main operating cost in the catalytic oxidation is furnace fuel. The fuel reduction is always important, meanwhile achieving the environment and health concerns. Various technologies for the abatement of the VOC concerning catalyst and reactor developments and controlling methods have been applied. The selection method is based on operating cost and initial capital cost, VOC type and concentration, as well as gas flow rate. In general, the PTA plant exhaust gas flow rate has capacity above 20 m³/s and has VOC concentration in the range of 500–1000 ppm, while the expected conversion is at least 98% [3–4]. Thus, the most feasible options for industrial applications are thermal and catalytic oxidations. Given the operating cost consideration, the catalytic oxidation is mostly preferred [5].

The catalytic oxidation for regenerative type by heat storage media (direct heating and cooling) is also known as reverse flow reactor (RFR). Figure 2 shows the reverse flow reactor concept. The heat transfer effectiveness may reach 85% or higher when using the regenerative type of reactor. The most important advantage is no external heat supply (fuel) needed when the autothermal operation is achieved. This kind of operation may reach up to 99% conversion [5].

The working principle of RFR as shown in Figure 2: in the first half cycle, the feed gas flows through valve V1 then enters the reactor to carry out the reaction and generate heat released by exothermic reaction. The released heat is stored in the catalyst bed or inert solid in the reactor. The reaction products leave the reactor through V4. During this time period, the valves V2 and V3 are closed. In the second half cycle, valves V1 and V4 are closed while valve V2 and V3 are open, so that the feed gas enters the reactor from the bottom to the top. When the feed gas enters the reactor, this stream is heated up by the heat that is previously stored in the catalyst bed or inert solid [6–7].

In this study, benzene was considered as representative compound of VOC. Benzene is widely agreed as dangerous compound and must be removed because it can cause carcinogenic and harmful to health
[8]. In addition, when benzene has been converted above 98\%, it can be assured other VOCs also have been converted (above 98\%) [5]. The exception is for CH₄ in very diluted conditions, which is difficult to be oxidized and it requires a relatively higher temperature [10-12].

In general, the catalytic oxidation of VOC uses platinum group catalyst. Several other studies also used 0.5\% Pt/γ-Al₂O₃ as catalysts for benzene oxidation [5, 9, 13-17]. The oxidation of benzene can be written as follows:

\[ \text{C}_₆\text{H}_₆ + 7.5 \text{O}_₂ \rightarrow 6 \text{CO}_₂ + 3 \text{H}_₂\text{O} \quad \Delta H = -3259 \text{ kJ/mol} \]

The mechanism of the catalytic oxidation of benzene follows two-stage redox model, also known as the Mars-Van Krevelen Model. It is based on several studies comparing power-law kinetic model, Langmuir-Hishelwood model, Eley-Rideal model, and Mars-Van Krevelen model. It is found that the Mars-Van Krevelen model indicates most proper model in this case [14].

Based on Mars-Van Krevelen model, hydrocarbon oxidation is assumed to occur through a surface redox cycle. The reaction rate of the benzene oxidation is as follows:

\[ r = \frac{k_{ol} k_i C_{O₂} C_{benzena}}{k_{ol} C_{O₂} + v_i k_i C_{benzena}} \]  

(1)

\[ v_i \] is the oxygen stoichiometric coefficient in the benzene oxidation equation, which is 7.5. \( C_{O₂} \) is the oxygen concentration in mol\%, and \( C_{benzena} \) is the benzene concentration. Based to the research [15]

\[ k_{ol} = 1.069 \times 10^{-4} e^{\left(\frac{-3795}{RT}\right)} \quad \text{and} \quad k_i = 5.565 \times 10^{-2} e^{\left(\frac{-22582}{RT}\right)} \]

When using RFR, some of the difficult reaction to react without external heat source becomes feasible. The reaction also sustains even with lower feed temperature and gets higher conversion when compared to the reaction performed in an adiabic reactor with a one directional flow. During start up, the RFR may need preheating to bring the catalyst bed at the reaction temperature, but after the reaction takes place, it withstands without external heat source (autothermal). This because the solid inert or catalyst bed has three times volumetric heat capacity larger than the gas volumetric heat capacity [18].

After reaching the autothermal, heat extraction may be needed to maintain the reactor temperature. The two most potential and efficient methods: first, partial flow withdrawal for heat extraction by an external heat exchanger, and the flow is not returned to the reactor. The second is full flow withdrawal for heat extraction by an external heat exchanger, and all the flow is returned to the reactor [19].

The switching time of flow reversal is a key performance to maintain the autothermal operation. The exothermic energy must be stored in the solid material inside the reactor. In this case, the reactor is a fixed bed reactor with or without a solid inert section. In general, the inert (either monolith or packed bed) is placed on the catalyst both sides to store the heat and preheat the feed [10, 18-23]. The study purpose is to examine the technical feasibility of RFR for optimizing the process equipment configuration and get lower fuel consumption. Meanwhile assess the autothermal operating conditions.

2. Methodology

This research consists of six stages: data collection, modelling, model validation, steady state simulation, unsteady state simulation and result analysis. The flow condition used in this study is shown in Table 1 which is typical flue gas at PTA plants with medium production capacity [3]. The FlexPDE Software v6.32s [24] is used with the following equations:

The mass balance equation [25]:

\[ \frac{\partial C_i}{\partial t} = -U_i \frac{\partial C_i}{\partial z} + D_{eff} \frac{\partial^2 C_i}{\partial z^2} + R_{i(x,t)} \]  

(2)

The energy balance equation:

\[ \epsilon \rho \xi \frac{\partial T_s}{\partial t} = -U_i \epsilon \rho \xi \frac{\partial T_s}{\partial z} - \alpha_0 a_v (T_g - T_s) \]

\[ \left[ (1 - \epsilon) \epsilon \rho \xi \frac{\partial T_s}{\partial t} + \epsilon \rho \xi \frac{\partial T_s}{\partial z} \right] = \alpha_0 a_v (T_g - T_s) + (-\Delta H) R_{i(x,t)} + \frac{\partial^2 T_s}{\partial x^2} \]

(3)

(4)
\[ U(t) = K(t)u \]  
\[ \text{[K has (+) value for right direction flowrate and (-) for left direction flowrate]} \]

Initial condition:
\[
C_i = C_{i,\text{input}}
\]
\[
T_g = T_{g,\text{input}}
\]
\[
T_{s,\text{ inert}} = T_{s,\text{ one direction steady state}}
\]
\[
T_{s,\text{ katalis}} = T_{s,\text{ one direction steady state}}
\]

Boundary condition:
\[
\frac{\partial C_i}{\partial z} = 0, \frac{\partial T_s}{\partial z} = 0 \text{ at } z = 0
\]
\[
\frac{\partial T_s}{\partial z} = \frac{\partial T_g}{\partial z} = 0 \text{ at } z = L
\]

The modelling assumptions are:
- The reactor is assumed to be dynamic, pseudo homogeneous and one-dimensional (axial).
- Gas is assumed to be an ideal gas at operating conditions with constant physical properties (to P & T).
- The fixed bed reactor is assumed as a plug flow reactor, axial dispersion is not considered.
- Pressure drop along the reactor is ignored. Constant pressure along the reactor.
- The reactant adsorption and the product desorption are assumed to be unaffected to the gas velocity in the reactor due to dilution with the inert gas.
- \( \text{N}_2 \) and other VOC present in the feed are not involved in the reaction and assumed as inert.

Table 1. Off Gas Composition and Condition Data as Feed of the Reactor.

| Units     | Value  |
|-----------|--------|
| Pressure  | kg/cm²G |
| Temperature | °C  |
| Mass Flow rate | kg/hour |
| Composition |        |
| Nitrogen   | % mass |
| Oxygen     | % mass |
| Water      | % mass |
| Carbon dioxide | % mass |
| Benzene    | % mass |
| Methanol   | % mass |
| Other VOCs | % mass |
| Total      | % mass |

Model is validated by comparing the model in this research to a previous research. Initial simulations are carried out in steady conditions and used as initial conditions for unsteady state. The results of steady-state simulation are concentration profile \( (C) \) along the reactor and temperature profile \( (T) \) along the reactor. The purpose of unsteady conditions simulation is to know the reactor temperature profile during a reverse flow operation. This simulation is performed with and without catalyst length optimization and inert solids presence. This simulation results a temperature profile against switching time, temperature profile against concentration, and heat extracted to maintain the temperature.

3. Result and Discussion

3.1. Model Validation

![Figure 3. The profile of the benzene inlet concentration against the reaction conversion at 300°C temperature and gas superficial velocity = 0.147 m/sec. The symbol (●) refers to experimental results [15]; the symbol (□) refers to the simulation in this study.](image-url)
The previous study results [15] describes reactor feed inlet concentration against conversion at temperature 300°C in a one directional reactor at steady state. The comparison is under the same operating conditions and reactor dimensions. The result in Figure 3 shows a similarity and has average error of 16.75%. So, it is concluded that the model is usable to simulate the case in this study.

3.2. Steady State One Directional Reaction Simulation

One directional flow reaction simulation in steady state is the base case for other simulations. In this simulation is used the dimension and operating condition of industrial scale reactor. Commonly, the industrial scale reactor is designed with a margin to the theoretical sizing. This is to overcome the process fluctuation and the catalyst deactivation effect so the interval between catalyst replacements is long and economic. In this study the industrial scale reactor is used including the over design. The catalyst bed length is 1.95 m and reactor diameter is 2.11 m. The maximum allowable catalyst bed length is 2.9 m.

The one directional reaction simulation in steady state checks the reaction effect to the catalyst bed effective length. The catalyst bed effective length is the catalyst length where the reaction occurs and achieves 100% conversion on new catalyst conditions. The effective length may be shorter than the catalyst bed overall length. This simulation varies the reactor inlet temperature and intends to know the effective length effect. The reactor inlet concentration is 15 ppm benzene.

The temperature is varied from 80°C up to 400°C. The temperature prior entering the catalytic oxidation system is 80°C. The HE-1 in Figure 1 functions as pre-treatment to heat up the feed up to 100°C (to vaporize the water). Because bromine presence in the feed stream may form HBr upon meet with water (liquid phase). HBr is highly corrosive. Heating up to 400°C can be achieved with the furnace. HE-2 is a recuperative heater to reduce the energy load of the furnace.

Referring to Figure 4, as the inlet reactor temperature increase, shortening the reactor effective length. The effective length is about 24%-37% of the total reactor length, indicating a large over design to encounter the catalyst deactivation. Commonly Pt/Al₂O₃ catalysts normal lifetime is 5 years [9].

Figure 4. The reactor’s inlet temperature profile against the effective length to obtain 100% conversion at the initial concentration of 15 ppm and the linear gas velocity of 1.2163 m/sec.

Figure 5. The supplied energy profile to meet a suitable inlet temperature with the effective length to get 100% conversion at 15 ppm initial concentration and a linear gas velocity of 1.2163 m/sec. The profile of inlet temperature against the reactor effective length (solid line), the additional energy required to increase the temperature from 80°C to the targeted inlet temperature (dashed line).

By the inlet temperature increases, the reactor effective length is shorter, but it also must be compensated by the higher fuel consumption in the furnace to increase the temperature. Figure 5 shows
the energy required to increase the temperature to 400°C. Exothermic reaction is the major source of the supplied energy, and partly supplied from the furnace. The recuperative heater output can reach a temperature of 280°C, while the furnace-supplied heat is to increase from 280°C to the required reaction temperature, and this heat is from the fuel combustion that may be a natural gas, diesel, or others.

Figure 6 shows the required additional energy from the furnace to raise the temperature from 280°C, about 3 to 19 MMBtu/hour, equivalent to a heating cost of US$ 160,000 to US$ 1 million per year (US$ 0.1208/ton feed gas up to US$ 0.7248/ton feed gas). This refers to the natural gas price US$ 6/MMBtu as stated in the Indonesian Ministry of Energy and Mineral Resource Decree Number 8 Year 2020. Hence alternative method for benzene catalytic oxidation may gains saving. So, the RFR is more effective than the recuperative method. Figure 7 shows the inlet benzene concentration increases along with the reactor effective length. The required reactor length is equivalent to 30%-45% of the total typical industrial reactor length, which means the over design is large. This is a design optimization potential and combines it with the RFR concept.

### 3.3. Simulation of RFR without Catalyst Length Optimization and Inert Solid Presence.

This part explains the RFR simulation when using industrial scale reactor (includes the over design), with bed catalyst length of 1.95 m and reactor diameter 2.11 m. There is no catalyst length optimization and inert solids as heat transfer medium, so all the solid in the reactor are Pt/Al₂O₃ catalyst. The RFR simulation is developed by varying the switching time (ST) and the benzene inlet concentration; besides observing the heat extraction possibility. Variations in flowrate are not conducted because by changing the flowrate for same reactor size, it changes the residence time in the reactor, which is the characteristic of variations in flowrate indirectly reflected in ST variations.

![Figure 6](image1.png)

**Figure 6.** The additional energy to get suitable reactor inlet temperature at various reactor inlet temperature (gas linear velocity 1.2163 m/sec). The energy required to increase the temperature from 80°C (solid line), the energy required to increase the temperature from 280°C (dashed line).

![Figure 7](image2.png)

**Figure 7.** The reactor inlet concentration Profile against minimum reactor length to achieve 100% conversion (linier gas velocity 1.2163 m/sec).

![Figure 8](image3.png)

**Figure 8.** Simple Diagram of the proposed RFR.
Figure 8 is the proposed RFR system. The reactor feed is also the heating product on HE-1 to keep the temperature at 100°C. So that the whole simulations on RFR is fed at 100°C. The HE-3 is a “Cooler or Boiler” to extract excess heat avoiding reactor overheat. The HE-3 cold side can be a heat transfer fluid, cooling water or boiler feed water. This dynamic simulation determines the effect of Switching Time (ST) for selecting the right operating parameters to maintain reactor operation without external heat and no overheating. The linear velocity is 1.2163 m/sec with the residence time 2.4336 seconds.

In Figure 9 shows a significant temperature rise at small ST and peaking at ST about 5 seconds (twice the residence time), then the temperature decreases and stabilize at ST 7.5 seconds (three times the residence time) and above. While the ST profile against the required heat extraction to not exceed 400°C, has similar profile to the ST profile for the highest temperature, the different is the heat extracted profile tend to be flat on ST 15 seconds and above. The temperature limit of 400°C is taken as the reactor operating envelop. When temperatures exceeding 550°C may cause agglomeration or sintering [9].

It is simulated a reverse flow reaction with a variation in the inlet benzene concentration with the same reactor dimension as the previous simulation. The simulation results is in Figure 10 and shows the higher temperature as the inlet benzene concentration increase into the reactor. Because more inlet concentration means more benzene reacts thus produces more exothermic heat. The temperature rise in this simulation is not too extreme, since the chosen ST is a stable ST obtained from the previous sub-chapter on ST variation (7.5 seconds). The heat extraction required also increase with the inlet benzene concentration increased; this is certainly related to the more benzene reaction.

3.4. Reverse Flow Reaction Simulation with Catalyst Length Optimization and Inert Solid Presence.

This section simulates RFR with industrial scale reactor, the differences with previous sub-chapter is the optimization of catalyst length, which only 50% (0.975 m) of catalyst length compare to the previous sub-chapter. The 50% catalyst length is based on the sub-chapter 3.2, where the largest effective length obtained is 45% of the total catalyst length; it is rounded to 50% of the total catalyst length. In addition, on the right and left sides of the catalyst is placed inert solid (25% of the catalyst length in the previous sub-chapter or 0.4875 m). Thus, the overall length of the catalyst and the inert solids in this simulation is equal to the total catalyst length in the simulations at the previous. This optimization aims to reduce the catalyst amount; the 50% catalyst length decrease provides significant saving due to expensive Pt.
price. The average Pt price in January to August 2020 on the London Bullion Exchange is about $871.07/troy ounce or equal to $28005.61/kg. In Figure 11 shows the proposed RFR system. The system is very similar to the system proposed in Figure 8, the difference is the presence of inert solids in the reactor. The linear velocity is 1.2163 m/sec.

In Figure 12, there is a significant temperature rise at small ST, and peaking at ST about 5 sec. Then the temperature decreases and flats at ST 7.5 seconds and above, which is at the highest temperature about 650°C. The decrease of heat extraction required to avoid exceeding 400°C after ST 5 seconds is because longer ST means more heat stored utilization to heat up the feed. While the cause of the temperature increase in ST less than 5 seconds is the small ST causes a high heat accumulation in the reactor. From the data then it is concluded that the suitable ST is 7.5 second and above, because the highest temperature achieved stable at about 650°C, means more controlled reaction.

Another interesting thing found (as shown in Figure 13) is when ST is smaller than the reactor residence time there are two peak temperatures on the right and left side of the reactor. When ST is greater than or equal to the reactor residence time then only one peak temperature is found, one peak temperature is the same as the profile description proposed by Liu [23]. The comparison of the temperature profile illustrates if ST is smaller than the reactor residence time, then fluid unable to flow from one side to other side. When the direction of fluid flow is reversed, there is a fluid trapped in the centre of the reactor, this is the cause of the two peak temperatures in the profile.
The next is simulation of reverse flow reaction with a benzene inlet concentration variation at the same reactor dimension as the previous simulation. The simulation results are presented in Figure 14, and it shows as the benzene inlet concentration rises, the higher peak temperature can be achieved. Because more inlet concentration means more benzene reacts and produce exothermic heat. The temperature rise in this simulation is not too extreme, since the chosen ST in this simulation is a stable ST obtained from the previous (7.5 seconds). The heat extraction required also increase with more inlet benzene concentration; this is certainly related to the more exothermic heat produced.

![Figure 14. Benzene inlet concentration profile against peak temperatures achieved (solid line) and Benzene inlet concentration profile against heat extraction required to avoid exceeding 400°C (dashed line).](image)

From the simulation of ST variations and benzene inlet concentrations it shows the autothermal is happened in benzene oxidation, even heat extraction is required to allow good temperature control in the reactor. Thus, this section main conclusion is by optimizing the length of the catalyst, the reactor operation still takes place with 100% conversion and auto thermally. Additionally, potential capital expenditure reduction is also observed because of optimization of the catalyst length.

4. Conclusion

The study focuses on industrial scale reactor. The taken reactor size affect to the conclusion. There is a large over-design factor in industrial scale reactors to counter catalyst deactivation during operation.

In one directional flow operation:

a. As the reactor inlet temperature increase, the catalyst effective length is decreased. The results show by increasing reactor inlet temperature from 80°C to 400°C, the effective length decreased from 0.73 m to 0.47 m or decrease 36%.

b. As the benzene inlet concentration increases there is an increase in effective length. The study shows by increasing benzene inlet concentration from 12 ppm to 20 ppm, then the effective length increased from 0.59 m to 0.88 m or increase 49%.

In reverse flow reactor:

a. ST 7.5 second and above should be used. The heat extracted to stabilize the RFR at 400°C is in the ranges of 5x10⁵ to 5.5x10⁶ J/(m³. second).

b. As the benzene inlet concentration increases, the higher peak temperature. At benzene inlet concentration 12 ppm to 20 ppm, the peak temperature ranging from 475°C to 800°C. The heat extracted to stabilize the RFR at 400°C is ranging 2.75x10⁶ to 3.75x10⁶ J/(m³. second).

In RFR with Catalyst Length Optimization and Inert Solid Presence:

a. ST 7.5 second and above should be used. The heat extracted to stabilize the RFR at 400°C in the ranges of 1x10⁶ to 4.5x10⁶ J/(m³. second).

b. As the benzene inlet concentration increases, the higher peak temperature. At benzene inlet concentration of 12 ppm to 20 ppm, the peak temperature varies from 525°C to 850°C. The heat extracted to stabilize the RFR at 400°C is ranging 1.15x10⁶ to 1.275x10⁶ J/(m³. second).

The benzene catalytic oxidation can operate in autothermal by RFR and may get energy savings of US$ 0.1208/ton feed gas up to US$ 0.7248/ton feed gas compared to recuperative types. Heat extraction may be needed to avoid catalyst overheat. By optimizing the catalyst effective length (50% of the reactor length), the industrial scale reactor still can achieve 98% conversion and auto thermally.
References
[1] Ainsworth S 1994 Purified Terephthalic Acid Capacity Takes Off C&E News 72 19–20
[2] Code of Federal Regulations, 40: Chapter 1, Subchapter C, Part 51, Subpart F, 51100.
[3] Belmonte F, Abrams K, and Oppenheim J 2001 Catalytic Vent Gas Treatment System For Abatement of Volatile Chemical Emissions, US Patent 6,207,120 B1.
[4] Kirk R and Othmer D 1978 Encyclopedia of Chemical Technology Volume 17 (New York: John Wiley & Sons)
[5] Hermia J and Vigneron S 1993 Catalytic Incineration for Odor Abatement and VOC Destruction, Catalyst Today 17 249-258.
[6] Budhi Y W, Jaree A, Hoebink J H B J, Schouten J C 2004 Simulation of reverse flow operation for manipulation of catalyst surface coverage in the selective oxidation of ammonia Chemical Engineering Science 59 5365-5377.
[7] Matros Y and Bunimovich G 1996 Reverse-Flow Operation in Fixed Bed Catalytic Reactors Catalysis Reviews: Science and Engineering 38 1-68
[8] Longo D, Kasper D, Jameson J, Fauci A, Hauser S, Loscalzo J 2004 Harrison’s Principles of Internal Medicine, 16th edition (New York: McGraw-Hill Professional) p 618
[9] Heck R, Farrauto R and Gulati S 2009 Catalytic Air Pollution Control: Commercial Technology (New Jersey: Wiley and Sons)
[10] Budhi Y W, Effendy M, Bindar Y, Subagjo S 2014 Dynamic Behavior of Reverse Flow Reactor for Lean Methane Combustion Journal of Engineering and Technological Sciences 46 (3) 299-317.
[11] Subagjo S, Budhi Y W, Effendy M, Bindar Y, 2014 Homogenity of Continuum Model of an Unsteady State Fixed Bed Reactor for Lean CH₄ Oxidation Journal of Engineering and Technological Sciences 46 (2) 195-210
[12] Nuryaman A, Gunawan A Y, Sidarto K A, Budhi Y W 2012 Singular Perturbation Problem for Steady State Conversion of Methane Oxidation in a Reverse Flow Reactor Journal of Mathematical and Fundamental Sciences 44 (3) 275-284
[13] Sawyer J E and Abraham M A 1994 Reaction Pathway During The Oxidation of Ethyl-Acetate on Platinum Alumina Catalyst Industrial and Engineering Chemistry Research 33 2084-2089
[14] Ordonez S, Bello L, Sastre H, Rosal R, and Diez F V 2002 Kinetics of The Deep Oxidation of Benzene, Toluene, n-Hexane, and Their Binary Mixture Over a Platinum on γ-Alumina Catalyst Applied Catalyst B: Environmental 38 139-149
[15] Lee S B, Kang, S L, Lee J D, Hong I K 1999 Catalytic Oxidation Kinetics of Vapor Phase BTX over Pt/γAl₂O₃ Journal of Industrial and Engineering Chemistry 5 No.3 170-176.
[16] Rusu A O and Dumitru E 2003 Destruction of Volatile Organic Compounds by Catalytic Oxidation Environmental Engineering and Management Journal 2 273-302
[17] Santos V P 2010 Catalytic Oxidation of Volatile Organic Compounds Doctoral thesis (Porto: University of Porto) p 9–12
[18] Salomons S 2003 Modelling the Behaviour of a Reverse-Flow Catalytic Reactor for the Combustion of Lean Methane Master degree thesis (Edmonton: University of Alberta) p 6-14.
[19] Marin P, Ordonez S, and Diez F V 2009 Procedures for Heat Recovery in The Catalytic Combustion of Lean Methane–Air Mixtures In A Reverse Flow Reactor Chemical Engineering Journal 147 356–365
[20] Budhi Y W 2020 Reverse Flow Reactor with Side Feeding as a Novel Strategy to Create Dynamic Oxygen Coverage and Heat Propagations in Lean Ammonia Oxidation Chemical Engineering and Processing: Process Intensification 144 108064
[21] Budhi Y W, Jaree A, Hoebink J H B J, Schouten J C 2004 Reverse flow operation with reactor side feeding: analysis, modeling, and simulation Industrial and Engineering Chemistry Research 43 6955-6963.
[22] Budhi Y W 2020 Reverse Flow Reactor with Side Feeding as a Novel Strategy to Create Dynamic Oxygen Coverage and Heat Propagations in Lean Ammonia Oxidation Chemical Engineering
and Processing: Process Intensification 144 108064

[23] Liu B 2000 Experimental and Modelling Study of Reverse Flow Catalytic Converters for Natural Gas/Diesel Dual Fuel Engine Pollution Control Doctoral thesis (Edmonton: University of Alberta)

[24] PDE Solutions Inc 2014 FlexPDE 6 Version 6.35 2/17/2014 (Spokane Valley: PDE Solutions Inc)

[25] Suharti P H, Budhi Y W and Bindar Y 2011 Kajian Pemanfaatan Reverse Flow Reactor Untuk Oksidasi Katalitik Uap Bensin Jurnal Teknik Kimia Indonesia 10 26-36