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Research Article

Optimizing Maleic Anhydride Production from Benzene's Cost by Preheating Inlet Air in Fired Heater and Modifying Distillation Column Operating Condition

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Abstract

The need for materials with superior properties from two different ingredients has recently attracted industries. Maleic anhydride or $C_4H_2O_3$ is an intermediate product often used to mix materials with distinct characteristics. One way to produce maleic anhydride is by reacting benzene and air. Unfortunately, the production of maleic anhydride is classified as energy-consuming due to the plant equipment used such as fired heaters and distillation columns for the production. In this process modification, optimization is carried out on the fired heater by using the residual heat from combustion as an energy source for the preheating of incoming air. In addition, optimization was also carried out on the distillation column by changing the operating variables, especially the reflux ratio and column temperature. Simulations were carried out using Aspen HYSYS V11 and comparisons were made between the energy required and the profit obtained from the original process to the modified process. The simulation results showed a reduction in energy cost by 0.771% on the fired heater and an increase in profit by 47.47% on the distillation column. Therefore, this modification reduces the energy cost while maximizing the profit made from maleic anhydride production using benzene and air.

Keywords: Maleic anhydride; Process modification; Heat efficiency; Profit maximation

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1. Introduction

The demand of natural rubber polymers combined with natural starch that have different properties and characteristics, can lead to the development of polymers with novel features [1]. To achieve compatibility between non-polar natural rubber polymers and polar natural starch, modifications are necessary [2]. The use of maleic anhydride additives can enhance the interaction between the two materials [3]. Maleic anhydride can be synthesized from benzene or n-butane. Benzene is widely available as a raw material and catalysts that have high selectivity, making it easier to achieve the optimal conversion. Maleic Anhydride (MAN) is a highly significant chemical in the global market due to its wide range of applications [4]. In domestic markets, benzene is significantly cheaper than maleic anhydride. Maleic anhydride ($C_4H_2O_3$) is an important intermediate in the industry [5]. This chemical is primarily used as a raw material for the production of unsaturated polyester resin. The resin is then used as a raw material in the manufacture of fiberglass to make it strong yet lightweight and resistant to corrosion, such as in the production of boats, cars, electronic equipment, and more [6].

Maleic anhydride has a molecular formula of $C_4H_2O_3$ or is also called 2,5-furandione which is an

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organic compound [7]. It is an organic compound that contains a dual functionality composed of an electron acceptor carbon-carbon double bond and a reactive anhydride group [8]. This compound can be synthesized by oxidizing benzene (C_6H_6) [9]. Maleic anhydride is considered an intermediate product, serving as a raw material for the production of other compounds. Its wide range of applications arises from its unique structure and the reactivity of the double bonds at the alpha and beta positions. The chemical structure and high reactivity of maleic anhydride derivatives make them suitable for the production of various types of resins and chemical transformations of organic reagents [10]. Furthermore, MA can be recycled and reused by simple dehydration and vacuum-drying [11]. Maleic anhydride is utilized across multiple industrial sectors, serving as a key raw material for products like unsaturated polyester resins, surface coatings, lubricant additives, and agricultural chemicals such as fumaric acid and succinic acid. More recently MA has also become an important ingredient in biodegradable plastics, one of the key developments for environmental protection in the future [12,13]. In this case, its production process heavily relies on operational equipment units to become reactions, one example of which is the reactor, distillation column, and fired heater.

This paper aims to evaluate the economic and environmental performance of the technology for manufacturing maleic anhydride (MAN) by focusing on eco-efficiency and life cycle impacts, including utility plant considerations. Eco-efficiency assesses the environmental impacts of a process in relation to its economic performance. Such an analysis is typically based on the quantitative evaluation of eco-indicators, which represent the relationship between an environmental factor (e.g., air emissions) and an economic factor (e.g., production rate or net profit). The benzene technology for MAN manufacture studied in this paper were originally proposed in [14]. The process was simulated by [15] in UniSim® Design Suite R390.1, so we could obtain the required data to estimate the costs and determine their eco-indicators (Figure 1). For the benzene technology, the same inlet feed conditions and equipment parameters used by [16] were considered. It was simulated under steady state conditions by using the non-random two liquid (NRTL) thermodynamic model (enthalpy, phase equilibrium, etc.) [16].

Fired heaters are an essential component of most process plants. They are primarily used to heat all types of hydrocarbons and also hot coils, steam or air. Fired heater required big amount of energy consumption, for that reason a lot of engineers tried to optimizing the process by improve its thermal efficiency. This improvement can save thousand of dollars and extending the equipment's life [17].

2. Methods

2.1 Simulation Software

Aspen HYSYS V11 is used to study, simulate and optimize maleic anhydride industrial process from benzene oxidation. In this study, Antoine package is used as the thermodynamic model in the simulator.

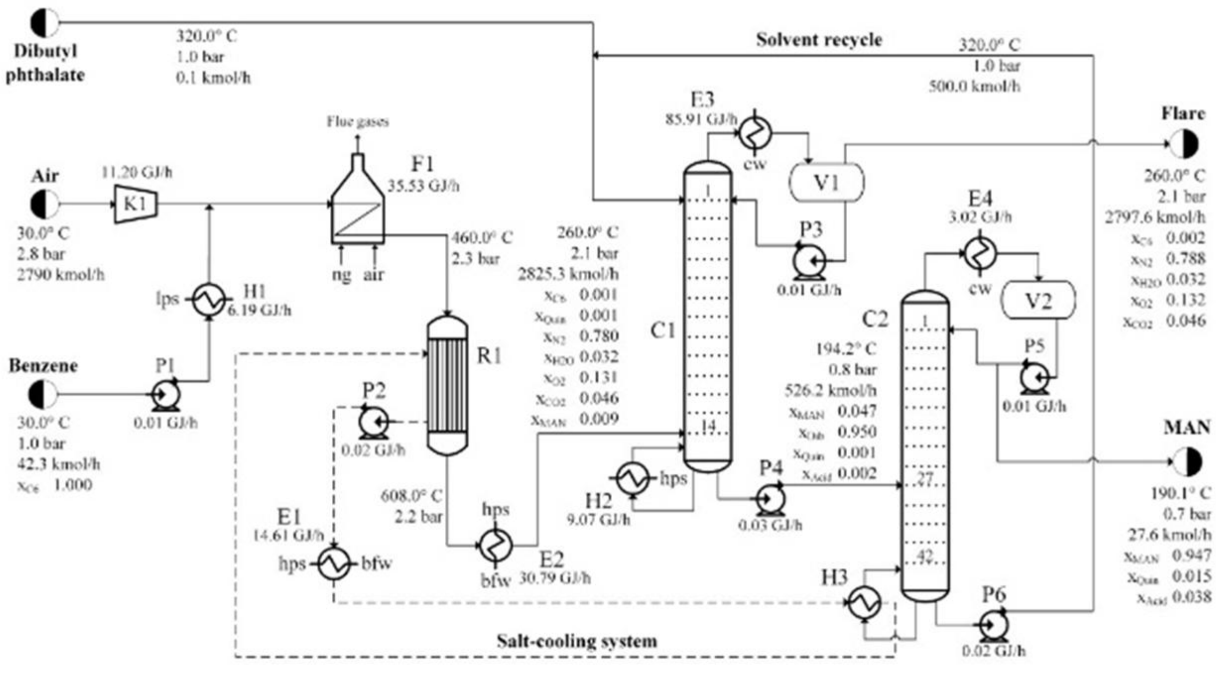

Figure 1. MAN basic process diagram [15]

2.2 Fired Heater Optimization

In MAN production, fired heater used to increase the temperature from feed temperature to reaction temperature. The ideal MAN production reaction temperature is 400 °C and the feed went in at 140 °C. This process will consume a lot of energy duty and cost to maximize the process. From this reason, fired heater process need to be optimized.

Fired heater optimization process can be achieved by increasing the thermal efficiency. Thermal efficiency of fired heater will be improved by preheating the inlet air using heat exchanger from combustion outlet cooling process. Preheating the inlet air will improve the combustion efficiency and reach higher temperature with shorter time. This process will make the fired heater duty become lower and conduct minimum hydrocarbon's amount.

To maximize the thermal efficiency, air inlet will be preheated using heater from environment temperature, which is around 30 °C to 100 °C. The composition, molar flow and pressure will be set the same as the original data. Energy duty from this heater exchanged with combustion output cooler process, which can reduce the overall energy duty.

2.3 Distillation Column Optimization

In this simulation, distillation is carried out in two stages. The first column distillation is used to separate all impurities contained in the maleic anhydride solution in dibutyl phthalate. This separation produces a top product of benzene and air mixture with a bottom product of pure maleic anhydride solution in dibutyl phthalate.

In the second column distillation, there is a separation between maleic anhydride and dibutyl phthalate. The top product is maleic anhydride with purity >95.0 %wt that is ready for sale. The bottom product is dibutyl phthalate solvent with a purity level of >99.0 %wt which can be recycled back to the solvent stream entering the absorber.

To maximize the operating conditions of the distillation column, profit-oriented objective is utilized based on the purity and flow rate of the maleic anhydride produced. Production profit is calculated based on the calculation of maleic anhydride sales revenue, equipment operating expenses, raw material costs, and solvent recycle costs. The maximum profit is granted by changing two variables of distillation column operating conditions using the 'Optimizer' function in Aspen HYSYS V11.

$$\text{Profit} = \text{Sales} - \text{Operational Costs} \tag{1}$$

Sales includes the revenue of maleic anhydride calculated using the following formula:

$$\text{MAN Sales} = \text{F} \times \text{Purity} \times \text{Price} \tag{2}$$

Operational costs include the cost of electricity used in the distillation column with an estimated cost (State Electricity Company/PLN price reference for business units) per kWh of $0.07, initial feed cost of $0.15 (MAN solution in DBH) and adjusting to the purity of the incoming feed, solvent recycle cost calculated using an estimated 2.00 %wt DBH lost in each cycle.

$$\text{DBH Cost} = 2.00\% \times \text{F} \times \text{Purity} \times \text{Price} \tag{3}$$

Overall operational costs are expressed as the following

$$\text{Operational Costs} = \text{Electricity Cost} + \text{Raw Material Cost} + \text{Solvent Recycle Cost} \tag{4}$$

The calculation will yield the net profit of maleic anhydride sales per kilogram from the distillation column operating unit.

3. Result and Discussion

3.1 Fired Heater Optimization

Based on the original process simulation, the following fuel, air inlet, and combustion outlet rate, composition and temperature data were obtained from Aspen HYSYS V11 (Tables 1-3). From the original process simulation, thermal efficiency obtained from Aspen HYSYS V11 (Figure 2). From the original simulation result, fired heater's thermal efficiency that can be achieved in this simulation is 75.91% with total cost $24.820/h (Figure 2).

Based on the optimized simulation result, the following fuel, air inlet, and combustion outlet rate, composition and temperature data were obtained from Aspen HYSYS V11 (Tables 4-6). From the optimized process simulation, thermal

Table 1. Original process air inlet

Parameter	Value
Flowrate (kg/h)	2865.3042
Oxygen mole fraction (%mole)	0.21
Nitrogen mole fraction (%mole)	0.79
Temperature (°C)	30

Table 2. Original process fuel inlet

Parameter	Value
Flowrate (kg/h)	170.8450
Ethane Mole Fraction (%mole)	0.97
Propane Mole Fraction (%mole)	0.03
Temperature (°C)	30

efficiency and energy duty obtained from Aspen HYSYS V11 (Figure 3). From the optimized simulation result, fired heater's efficiency that can be achieved in this simulation is 76.50%. With total cost $23.924/h.

Comparing the result between original and optimized simulation, fired heater's efficiency increased by preheating the inlet air. Based on result displayed in Figures 2 and 3., fired heater efficiency from original process is 75.91% to modified process increased 0.771% become 76.5%. With the same composition between the original and modified process (Tables 7-8), lower amount of hydrocarbon and air inlet achieved, from 3036.1492 to 2926.07 kg/h. Else than that, fuel cost used in original process reduced from $24.820/h become 23.924/h.

Higher air inlet temperature, will increase the thermal efficiency and lowering the overall process cost [18]. From heat balance equation on preheater, higher air combustion temperature will result lower fuel amount and heat flow. Which means O_2 and fuel required for combustion process will be decrease. This leads to energy and cost efficiency, since the amount of fuel will be decrease and the energy that required to achieve the temperature become lower [19-21].

3.2 Distillation Column Optimization

Based on the original process simulation, the following purity and flow rate data were obtained

Table 3. Original process combustion outlet.

Parameter	Value
Flowrate (kg/h)	3036.1492
Ethane mole fraction (%mole)	0.0167
Propane mole fraction (%mole)	0
Oxygen mole fraction (%mole)	0
Nitrogen mole fraction (%mole)	0.6420
Water mole fraction (%mole)	0.2043
Carbon monoxide mole fraction (%mole)	0.1370
Temperature (°C)	390.0788

Table 4. Modified process air inlet.

Parameter	Value
Flowrate (kg/h)	2761.4192
Oxygen mole fraction (%mole)	0.21
Nitrogen mole fraction (%mole)	0.79
Temperature (°C)	100

Table 5. Modified process fuel inlet.

Parameter	Value
Flowrate (kg/h)	164.6508
Ethane mole fraction (%mole)	0.97
Propane mole fraction (%mole)	0.03
Temperature (°C)	30

Table 6. Modified process combustion outlet.

Parameter	Value
Flowrate (kg/h)	2926.07
Ethane mole fraction (%mole)	0.0167
Propane mole fraction (%mole)	0
Oxygen mole fraction (%mole)	0
Nitrogen mole fraction (%mole)	0.6420
Water mole fraction (%mole)	0.2043
Carbon monoxide mole fraction (%mole)	0.1370
Temperature	392.4527

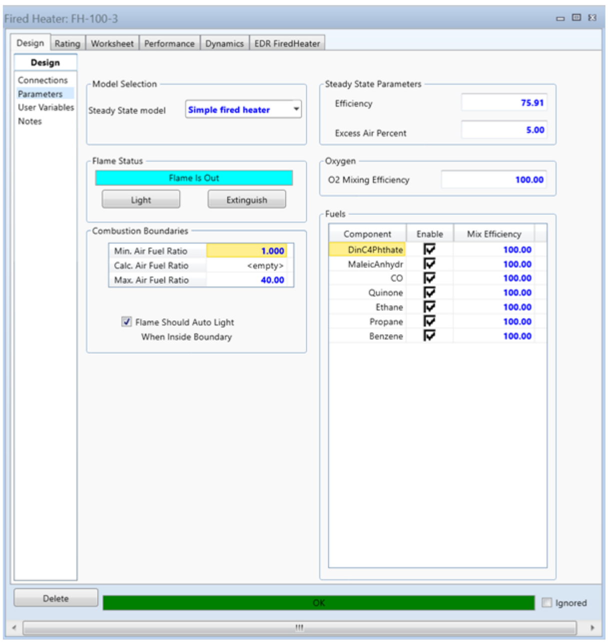

Figure 2. Original process thermal efficiency.

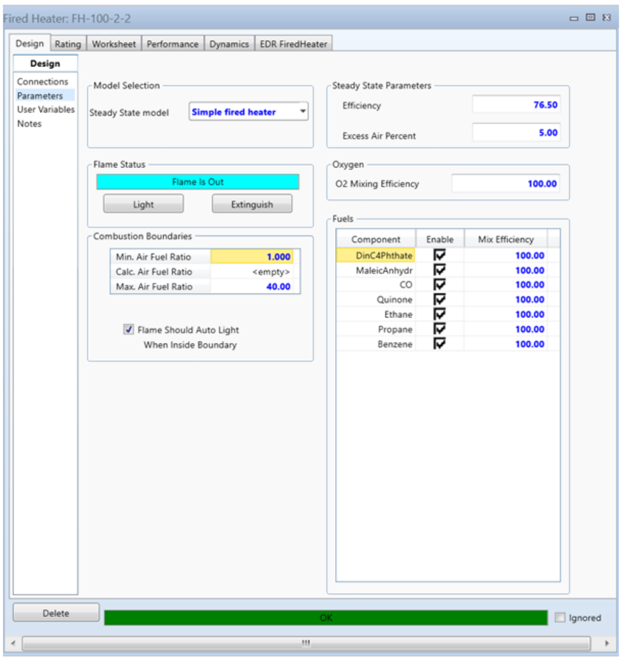

Figure 3. Modified process thermal efficiency.

(Table 9) from Aspen HYSYS V11. The following (Table 10) are specifications of the initial operating conditions used before optimization. The following is the profit calculation in the original simulation based on the data results obtained from Aspen HYSYS V11 (Table 11). From the original simulation results, the product sales profit is $83.27 for every kilogram of maleic anhydride produced. The operating variable conditions used are reflux ratio of 0.1290 and reboiler temperature of 339.3 ^{o}C.

Based on the simulation of the optimized process, the purity level and flow rate data were obtained from Aspen HYSYS V11 (Table 12). The following are specifications of operating conditions at the optimal point using the optimizer feature (Table 13). From the results of the optimized simulation, the product sales profit is $122.80 for each kilogram of maleic anhydride produced (Table 14). The operating variable conditions are at the optimal point. For the reflux ratio, a lower bound of 0.1235 and an upper bound of 0.1580 were installed and the optimal value was obtained at 0.1237. In addition, the reboiler temperature is set at a lower bound of 339.8 ^{o}C and an upper bound of 340.0 ^{o}C and the optimal value is obtained at 339.9 ^{o}C.

When compared to the original simulation. The optimized simulation is able to increase product purity and production profit. The maleic anhydride produced initially has a purity of 96.66 %wt and after optimization it increases by 1.14 %wt to 97.80 %wt. However, there is a decrease in overhead flow rate from 1135 kg/h to 1111 kg/h in the modified process.

Low reflux values tend to indicate more minimal energy requirements [22]. Generally, an increased reflux value will result in a linear increase in energy consumption [23]. However, there is a limit to the maximum distillate concentration obtained from the addition of reflux. At a certain point, increasing reflux will not have a significant impact on the purity of the product obtained [24]. This is consistent with the simulations results where the initial reflux value of 0.1290 drops to 0.1237 indicating that there is an optimum point to achieve equivalent or even more maleic anhydride purity with less energy.

Table 9. Purity and flow rate of maleic anyhdride in the original simulation.

Parameter	Value
Flowrate (kg/h)	1135
Purity (%wt)	96.66

Table 10. Operating conditions of the distillation column in the original simulation.

Parameter	Value
Reflux ratio	0.1290
Reboiler Temp. (^{o}C)	339.3

Table 7. Cost calculation for original process.

Parameter	Value	Fixed Cost	Value (in $)
Ethane Price (kg/h)	0.4595	Pure ethane price	47.493
Propane Price (kg/h)	0.0146	Pure propane price	205.263
		USD/h	24.820

Table 8. Cost calculation for modified process

Parameter	Value	Fixed Cost	Value (in $)
Condenser Duty (kW)	39.74	Cooling Cost	0.07
Reboiler Duty (kW)	653.5	Heating Cost	0.07
Overhead Production Rate (kg/h)	1135		
MAN Purity (%wt)	96.66	Pure MAN Price/kg	1.50
Bottom Production Rate (kg/h)	8501		
DBH Purity (%wt)	99.97	Pure DBH Price/kg	2.10
Feed Flow Rate (kg/h)	9636	Feed Cost	0.12
		Profit/kg MAN	83.27

Table 11. Profit calculation for original process.

Parameter	Value	Fixed Cost	Value (in $)
Ethane Price (m^3/h)	0.4428	Pure ethane price	47.493
Propane Price (m^3/h)	0.0141	Pure propane price	205.263
		USD/h	23.924

In the second variable regarding reboiler temperature, there is an increase in temperature from the starting point of 339.3 ^{o}C to 339.9 ^{o}C at the optimal point. The reboiler temperature itself has a direct impact on the separation efficiency and purity of the product obtained [25]. Broadly speaking, the fraction value of a component will increase if there is a heating treatment or the temperature is increased [26]. This is in accordance with the existing simulation where the increase in reboiler temperature produces a product with a higher purity to 97.80 %wt from the original 96.66 %wt.

4. Conclusions

The modified process successfully demonstrated an increase in energy efficiency and maximization of production profit. Based on the results obtained, there was an increase in energy efficiency at the fired heater from 75.91% in the original process to 76.50% in the modified process. In addition, the profit from purification in the distillation column also increased from $83.27/kg MAN to $122.80/kg MAN. Optimization of energy utilization through the use of residual heat and operating variable of distillation column can be used to improve the overall process.

CRediT Author Statement

Author Contribution: Cornellius P. Rongkang: Conceptualization, Methodology, Project Administration, Review and Editing. Inayah S. Putri: Conceptualization, Investigation, Writing, Review and Editing. Jason H. Kartawidjaja: Conceptualization, Methodology, Data Curation, Software, Review and Editing. Putri Dwi A. Lestari: Conceptualization, Investigation, Writing, Review and Editing. All author have all read and agreed to the published version of the manuscript.

Table 12. Purity and flow rate of maleic anyhdride in modified simulation.

Parameter	Value
Flowrate (kg/h)	1111
Purity (%wt)	97.80

Table 13. Operational conditions of the distillation column in modified simulation.

Parameter	Lower Bound	Optimal Value	Upper Bound
Reflux ratio	0.1235	0.1237	0.1580
Reboiler Temp. (^{o}C)	339.8	339.9	340.0

Table 14. Profit calculation for modified process.

Parameter	Value	Fixed Cost	Value (in $)
Condenser Duty (kW)	36.35	Cooling Cost	0.07
Reboiler Duty (kW)	572.0	Heating Cost	0.07
Overhead Production Rate (kg/h)	1111		
MAN Purity (%wt)	97.80	Pure MAN Price/kg	1.50
Bottom Production Rate (kg/h)	6761		
DBH Purity (%wt)	100.00	Pure DBH Price/kg	2.10
Feed Flow Rate (kg/h)	7873	Feed Cost	0.15
		Profit/kg MAN	122.80

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