Analysis and Cost Estimation of Methanol / Glycerin Distillation Unit Process System

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Abstract. An analysis on the process system of a distillation unit was carried out through stages consisting of data collection, computational design and simulation, calculation of the degrees of freedom, economic design criteria assessment, and estimation of cost. The analysis was employed to determine design variables and economical aspects that needed to be considered in the distillation unit design. In this article, field data was obtained from the Biodiesel Plant, PT. Cemerlang Energi Perkasa, Plant 1, Apical Group Dumai, optimized as basis for calculating and assuming certain values or process variables in the preparation of this process system analysis. The degrees of freedom obtained for the distillation column, which has been reduced by the number of variables is $C + 2$. Through calculation of estimated cost, it is found that total value of fixed capital investment is \$ 6,093,147.43 considering cost of main equipment of \$ 1,594,230.10 as well as total direct cost value of the methanol/glycerin distillation unit of \$ 3,905,863.75. Cost of the process operation is estimated at \$ 234,615,524.6 with minimum product price of glycerin \$ 3.39/lb. It is hoped that through this analysis, construction of distillation units, especially methanol/glycerin separation, can be well arranged and optimized.

1 Introduction

The continuous increase in world population, rapid industrialization, urbanization, and the immense strength of economic growth bring about the increase in consumption of fossil fuels to meet the mounting energy demand [1]. The perpetual emissions from fossil fuel burnings encourage the need for an appropriate and sustainable alternative for fossil fuels [1]. Biodiesel is regarded as the best substitute for diesel for it is environmentally friendly, sustainable, and its use in diesel engines requires little or no modification [2]. Biodiesel is composed of a combination of alkyl esters and long chains of fatty acids complied to required standards [3]. Biodiesel is a renewable fuel made from plant-based oils and animal fats [4]. These types of oils are far more environmentally friendly than conventional diesel as they emit less carbon gas [5].

Palm oil is one of the raw materials that can be utilized to produce biodiesel. The palm tree (*Elaeis guineensis*) is a seasonal plant which is potential raw material of fatty acids, and

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other common components in the food and chemical industries [6]. In Indonesia, palm oil has significant contribution towards domestic growth. Palm oil exports (2017) attained US \$ 23 billion [7]. BPS data present that palm oil supplied GDP (Gross Domestic Product) of Indonesia around 17% in agriculture in 2014. Additionally, based on agriculture census in 2013, around 2 million (M) farmers are involved in palm oil industry. The Indonesian Palm Oil Association (GAPKI) and other related industries employ up to 7.8 million workers within the palm oil process chain [7].

In this article, transesterification between palm oil and methanol was done to produce biodiesel consisting of methyl esters and by-products in the form of glycerin. The glycerin by-product has adequate sale value and demand in the market. Thus, its separation from the biodiesel and residue materials such as methanol using distillation is highly valuable [8]. Data from the field was obtained and simulated as a basis for design and analysis. This article aims to analyze the processes in the distillation unit to get optimal estimation of the process cost through the analysis of degrees of freedom and economic design criteria.

2 Methodology

Data collection needed to carry out simulations and analyses were obtained from actual field data, literature study data, and process simulation data. The actual field data consist of glycerin quality log sheet data and the operating conditions of the 166C1 column of the Glycerin Purification process from 19-23 September 2018 at 08.00 WIB, optimized so that the simulation as shown in Figure 1 is obtained.

Fig. 1. Methanol/glycerin distillation unit simulation result.

The simulation results exhibit optimal value of mass flow in the range of 4000 kg/hour with the main equipment consisting of pumps, tanks, heaters, and distillation columns. The distillation column has no condenser and only uses partial reboiler as a mixed heating media.

Analysis of the process system of the distillation unit was done by calculating its degrees of freedom, followed by the design of economic criteria and estimation of cost.

3 Degree of Freedom

The degree of freedom (DoF) of a process is defined as a group of process variables that can be regulated by the designer, operator or control system encompassing things such as temperature, pressure, composition, flow rate and component flow rate [9]. The degree of freedom is obtained by calculating all the variables used in the system minus the number of

independent equations associated with these variables [10]. In general, the degree of freedom can be determined by the following equation:

$$N_D = N_V - N_E \tag{1}$$

where, N_v is the number of degree of freedom, N_v is the number of variables and N_e is the number of equations involved in the system.

Analysis of the degree of freedom of a distillation column can be done using the mass and enthalpy balance approach as well as phase equilibrium equations and stoichiometric relationships that express the normalization of limits on some intensive variables, such as the concentration of components in the liquid and vapor phases. [11,12]. There are three types of variables in the separation process, namely intensive variables which include composition, temperature or pressure; extensive variables which include flow rate or heat transfer; along with equipment parameters such as the number of equilibrium stages [11]. Physical properties such as enthalpy or K values are not calculated.

An important aspect to be noted in the analysis of the degree of freedom of a process system is how mass flows and heat are described. Single phase flows containing C components must have a complete set of intensive variables that determine their condition, consisting of C - 1 mole fraction (or other concentration variables), temperature and pressure [11]. Based on Gibbs phase rules, for a single-phase system, the number of intensive variables needed to determine the state of the flow can be obtained by the formula $f = C - \sigma + 2 = C + 1$ variable. In addition, it should be noted that the conditions should not be the only aspect determined, but also the total flow rate as an extensive variable, to be able to determine the number of all phases in the flow. Furthermore, the mole fraction is also used to perfect the variables that are contained in the flow, meaning that the total of all variables in one stream, for example in the liquid phase flow, is C + 3 as shown in Table 1. The equation concerning the limit of mole fraction needs to be added to the list of equations to indicate the relationship between the concentrations of each substance present in the flow.

$$\sum_{i=1}^C \text{mole fraction} = 1,0 \tag{2}$$

Table 1. Variables in the liquid phase flow.

Variable Type	Number of variable
Liquid mole fraction $x_1, x_2, \dots$	C
Pressure, P	1
Temperature, T	1
Total Flow Rate, L	1
Total Variables	C + 3

In this paper, the distillation column to be analyzed for its degrees of freedom is the distillation column in the methanol/glycerin separation process in Plant 1 Biodiesel PT.

Shining Mighty illustrated in Figure 2. The methanol/glycerin distillation column consists of three stages and two feed streams that enter at different stages with different mass flow. Incoming feed is sprayed into the column with the temperature at the top column lower than the temperature at the middle column to prevent the presence of glycerin carry-over in the top product.

Fig. 2. Methanol/Glycerin Distillation Column Scheme.

Analysis of degree of freedom in the distillation column was first performed separately on each of the constituent elements consisting of three equilibrium stages equipped with one reboiler which would later go through more thorough analysis. The analysis was done by determining the number of variables and the number of equations involved in each of these sections, followed by calculating the total degrees of freedom.

The starting stage of the distillation column is the feed stage 1 which has two inflows and two outflows as shown in Figure 3, in which the feed is in the liquid phase. We assumed that this stage is ideal and adiabatic meaning implying that there is no heat transfer between the stage and the surrounding environment. Thus, the number of variables in the feed stage 1 is the number of variables associated with each stream, i.e. $C + 3$, multiplied by the number of existing streams.

$$N_V = 4(C+3)=4C+12$$

(3)

Fig. 3. Feed stage 1.

Equations related to variables in feed stage 1 include mass and enthalpy balances, phase equilibrium relationship, mole fraction limit, temperature and pressure equilibriums from the flow as summarized in Table 2.

Table 2. Variables in the liquid phase flow.

Equation type	Number of variable
Pressure equilibrium $P_{VOUT} = P_{LOUT}$	1
Temperature equilibrium $T_{VOUT} = T_{LOUT}$	1
Phase equilibrium relation $(y_i)_{VOUT} = K_i(x_i)_{LOUT}$	C
Component mass balance $F_1(x_i)_{F_1} + V_{IN}(y_i)_{V_{IN}}$ $= L_{OUT}(x_i)_{L_{OUT}}$ $+ V_{OUT}(y_i)_{V_{OUT}}$	C – 1
Total mass balance $F_1 + V_{IN} = L_{OUT} + V_{OUT}$	1
Enthalpy balance in adiabatic process $H_{F_1}F_1 + H_{V_{IN}}V_{IN} = H_{L_{OUT}}L_{OUT} + H_{V_{OUT}}V_{OUT}$	1
Mole fraction/concentration limit of components $\sum_{i=1}^c (x_i)_{F_1} = 1.0$ (e. g. for stream/flow F_1)	4
Total equations	2C + 7

Once the number of variables and equations contained in feed stage 1 is determined, the degree of freedom can be calculated using equation (4) as follows:

$$N_D=N_V - N_E=(4C+12)-(2C+7)=2C+5$$

(4)

The next stage is the feed stage 2 located in the center of the distillation column. In contrast to feed stage 1, the stage 2 has a total of five streams with three inflows and two

outflows, where there is an increase in liquid phase inflow originating from the stage above as illustrated in Figure 4.

Fig. 4. Feed stage 2.

The equations involved in feed stage 2 are presented in Table 3, whilst the number of variables is as follows:

$$N_V=5(C+3)=5C+15$$

(5)

Table 3. Variables in the liquid phase flow.

Equation type	Number of variable
Pressure equilibrium $P_{VOUT} = P_{LOUT}$	1
Temperature equilibrium $T_{VOUT} = T_{LOUT}$	1
Phase equilibrium relation $(y_i)_{VOUT} = K_i(x_i)_{LOUT}$	C
Component mass balance $F_1(x_i)_{F_1} + V_{IN}(y_i)_{V_{IN}}$ $= L_{OUT}(x_i)_{L_{OUT}} + V_{OUT}(y_i)_{V_{OUT}}$	C - 1
Total mass balance $F_2 + L_{IN} + V_{IN} = L_{OUT} + V_{OUT}$	1
Enthalpy balance in adiabatic process $H_{F_2}F_2 + H_{L_{IN}}L_{IN} + H_{V_{IN}}V_{IN}$ $= H_{L_{OUT}}L_{OUT} + H_{V_{OUT}}V_{OUT}$	1
Mol fraction/concentration limit of components $\sum_{i=1}^C (x_i)_{F_1} = 1.0$ (e.g. for stream/flow F_1)	5
Total equations	2C + 8

Thus, the degrees of freedom of feed stage 2 is

$$N_D=(5C+15)-(2C+8)=3C+7$$

(6)

The next stage to be analyzed is the last stage of the distillation column located at the bottom called the stripping section. Similar to feed stage 1, this section only has a total of four streams, i.e. two inflows and two outflows as shown in Figure 5, making the number of variables to be $4C + 12$.

Fig. 5. Feed stage 3.

The equations involved in stage 3 include equations that also appear in feed stage 1, provided that the feed flows are instead fluid inflows originating from feed stage 2. The equations are summarized in Table 4 below.

Table 4. Equations in stage 3.

Equation Type	Number of equation
Pressure equilibrium $P_{VOUT} = P_{LOUT}$	1
Temperature equilibrium $T_{VOUT} = T_{LOUT}$	1
Phase equilibrium relation $(y_i)_{VOUT} = K_i(x_i)_{LOUT}$	C
Component mass balance $L_{IN}(x_i)_{F_1} + V_{IN}(y_i)_{V_{IN}} = L_{OUT}(x_i)_{LOUT} + V_{OUT}(y_i)_{V_{OUT}}$	C - 1
Total mass balance $L_{IN} + V_{IN} = L_{OUT} + V_{OUT}$	1
Enthalpy balance in adiabatic process $H_{F_1}F_1 + H_{V_{IN}}V_{IN} = H_{LOUT}L_{OUT} + H_{V_{OUT}}V_{OUT}$	1
Mol fraction/concentration limit of components $\sum_{i=1}^C (x_i)_{F_1} = 1.0$ (e.g. for stream/flow F_1)	4
Total equation	$2C + 7$

$$N_D=(4C+12)-(2C+7)=2C+5$$

7)

In the partial reboiler section (Figure 6), there are three mass flows and one heat flow, expressed by the number of variables $3(C + 3) + 1$, where the addition of +1 is to characterize the heat transfer that occurs in the reboiler. It is assumed that there occurs perfect evaporation, meaning that the flow leaving the reboiler is at thermodynamic equilibrium, reflected through the equations in Table 5.

$$N_V=3(C+3)+1=3C+10$$

8)

Fig. 6. Reboiler.

Table 5. Equations in reboiler.

Equation Type	Number of equation
Pressure equilibrium $P_{VOUT} = P_{LOUT}$	1
Temperature equilibrium $T_{VOUT} = T_{LOUT}$	1
Phase equilibrium relation $(y_i)_{VOUT} = K_i(x_i)_{LOUT}$	C
Component mass balance $L_{IN}(x_i)_{F_1} = L_{OUT}(x_i)_{LOUT} + V_{OUT}(y_i)_{VOUT}$	C – 1
Total mass balance $L_{IN} = L_{OUT} + V_{OUT}$	1
Enthalpy balance in adiabatic process $H_{F_1}L_{IN} + H_{Lout}L_{out} = H_{LOUT}L_{OUT} + Q$	1
Mol fraction/concentration limit of components $\sum_{i=1}^C (x_i)_{F_1} = 1.0$ (e. g. for stream/flow F_1)	3
Total equations	2C + 6

Hence, the total degrees of freedom found in this partial reboiler is:

$$N_D=(3C+10)-(2C+6)=C+4$$

9)

After each element has been separately analyzed for its degrees of freedom, all results were summarized into one table (Table 6), containing an account of the number of equations and the corresponding variables so as to facilitate the next calculation process.

Table 6. Number of variables and equations of each element in the distillation column unit.

Element	N _V	N _E
Feed stage 1	4C + 12	2C + 7
Feed stage 2	5C + 15	2C + 8
Stage 3 (bagian stripping)	4C + 12	2C + 7
Reboiler	3C + 10	2C + 6
Total	16C + 49	8C + 28

All sections are then combined into one column, where there are six recurring streams, namely two streams in feed stage 1, two streams under feed stage 2, the flow entering the reboiler, and the flow coming out of the reboiler. The number of variables in the distillation column unit can be obtained by subtracting the total variables from all elements by the number of repetitive flows. Thus, we come to the following result:

$$(N_{V'})_{unit} = \sum (N_{V'})_{element} - 6(C+3) = (16C+49) - (6C+18) = 10C+31 \tag{10}$$

The number of independent equations of the distillation column unit can be obtained in a similar manner:

$$(N_E)_{unit} = \sum (N_E)_{element} - 6 = 8C+28-6 = 8C+22 \tag{11}$$

The number of degrees of freedom from the distillation column is, therefore, calculated as follows:

$$(N_D)_{unit} = (N_{V'})_{unit} - (N_E)_{unit} = (10C+31) - (8C+22) = 2C+9 \tag{12}$$

All sections are then combined into one column, where there are six recurring streams, namely two streams in feed stage 1, two streams under feed stage 2, the flow entering the reboiler, and the flow coming out of the reboiler. The number of variables in the distillation column unit can be obtained by subtracting the total variables from all elements by the number of repetitive flows. Thus, we come to the following result:

$$(N_v)_{unit} = \sum (N_v)_{element} - 6(C+3) = (16C+49) - (6C+18) = 10C+31 \tag{13}$$

Although the above analysis reveal that the degrees of freedom of the distillation column in the methanol/glycerin separation unit is $2C + 9$, it is necessary to consider whether there are known variables or specific values that will affect the final value of the existing degrees of freedom. Through observing data in the field, certain variable values have been collected and summarized in Table 7.

Table 7. Specified variables in the methanol/glycerin separating process by distillation column in Plant 1 Biodiesel PT. Brilliant Mighty Energy.

Variable type	Number of variable
Distillation glycerin product quality: Total glycerin, methanol, H ₂ O, Ash, Salt, MONG (Matter Organic Non-Glycerol), %w	C
Operational condition of Methanol/Glycerin Distillation Column : • Pressure of bottom column, <i>mbar g</i> • Pressure of middle column, <i>mbar g</i> • Pressure of top column, <i>mbar g</i> • Flow of glycerin to storage, <i>kg/hr</i> • Temperature of glycerin to storage, °C • Temperature of bottom column, °C	1 1 1 1 1 1
• Feed mass flow, <i>kg/hr</i>	1
Total variables of known values	C + 7

Thus, the final number of degrees of freedom from the distillation column reduced by the number of variables with known values is:

$$(N_D)_{final} = (N_D)_{unit} - (C+7) = (2C+9) - (C+7) = C+2 \tag{13}$$

4 Economic Design Criteria

Chemical process industries commonly use the distillation unit in thermal separation method. This separation unit is still going on as favour method for over forward decades. However, despite its versatility and extensive use, the distillation column requires considerable energy as the processes taking place hold more than 50% of the plant's operating costs. Meanwhile, complexity of design in the distillation column requires many kind of design variables than other operating units. The design variables in the distillation columns depend upon operating pressure, number of columns, feed entry location, reflux ratio, and composition specifications for the product as prominent variables in the column [13].

The distillation column design can be optimized economically involving not only trays number, feed location, but also operating conditions. The economic optimization of the involves determining the number of trays and feed locations to minimize the total investment and operating costs [14]. Basically, the calculation of the economic optimization for the distillation column can be executed by calculation of both the capital cost of equipment's and

the energy cost of the reboiler [14]. The main parts of the equipment in the distillation column are commonly the column vessels with L/D ratio and two heat exchangers, i.e. reboiler and condenser. Additional support units such as pumps, valves, and reflux drums are usually insignificant at the conceptual design stage. The trays cost denotes usually smaller than that of column and heat exchanger. The basic economic of the distillation column shown in Table 8 provides economic relationship, size and parameters used.

Table 8. Economic foundation in the distillation unit [14,15].

Condensers Heat-transfer coefficient = 0.852 kW/(K m ²) Typical differential temperature = 13.9 K Capital cost = 7296 (area) ^{0.65} Area in m ²
Reboilers Heat-transfer coefficient = 0.852 kW/(K m ²) Typical differential temperature = 34.8 K Capital cost = 7296 (area) ^{0.65} Area in m ²
Column vessel capital cost = 17.640 (D)^{1.066} (L)^{0.802} Diameter and length in meters
Column vessel capital cost = 17.640 (D)^{1.066} (L)^{0.802} Diameter and length in meters
Energy costs LP steam (6 bar, 87 psia, 160°C, 433 K) = \$7.78/GJ MP steam (11 bar, 160 psia, 184°C, 457 K) = \$8.22/GJ HP steam (42 bar, 611 psia, 254°C, 527 K) = \$9.88/GJ Electricity = \$ 16.8/GJ Refrigeration Chilled water at 5°C, returned at 15°C = \$4.43/GJ Refrigerant at -20°C = \$7.89/GJ Refrigerant at -50°C = \$13.11/GJ
TAC = (capital cost / payback period) + energy cost
Payback period assumed to be a certain year

The items in Table 8 are generally used in condensing and boiling of the hydrocarbon systems. The total condenser heat transfer coefficient is greater than the reboiler. Reboiler has a higher fouling tendency because the temperature used is high (coking or polymerization). There are various objective functions that can be used for economic optimization, not excluding the concept of "time value of money". Cases in point are "net-present-value" and "discounted cash flow". The application of this method requires many assumptions in which limited by the accuracy of these assumptions. These assumptions encompass expectation of forward sales, raw materials and products prices, and construction schedules.

Therefore, TAC (Total Annual Cost) is used, as shown in the table. This calculation combines energy costs and annual capital costs. The TAC unit is \$ year. The unit of capital investment is \$. The annual capital cost unit is reflected in \$/year, while the cost can be calculated by dividing the capital cost by the appropriate return payback period.

5 Cost Estimation

The distillation process of the methanol/glycerin mixture in the distillation unit has an optimal capacity of 4000 kg/hour with the detail of the main equipment used in this unit shown in Table 9. The simulation results provide the value of the capacity or mass flow of each equipment, where cost I (total delivered cost of the major equipment items) can be calculated using the following equation:

$$I=I_B \left(\frac{Q}{Q_B}\right)^M$$

(14)

Table 9. Distillation unit main equipment, capacity, and estimated total delivered cost.

Equipment	Capacity (kg/h)	M	I (\$/day)
Tank	240	0,66	172.178,11
Pump 1	2480	0,68	1.265,75
Pump 2	1300	0,68	194,51
Distillation Column	2900	0,79	1.177.765,23
Heater	1350	0,48	58.806,15
Reboiler	4070	0,25	168.420,34
Total Delivered Cost of Main Equipment			1.594.230,10

The estimated total direct cost is a percentage of the total value of the delivered cost of the major equipment, denoted as I_d. Direct cost consists of several aspects ranging from the construction process to the installation of equipment in the distillation unit. Some aspects of direct cost are shown in Table 10 [16].

Table 10. Direct cost aspects and percentage range from the design process [16].

Direct cost	Range	fi
Installation	0.10-0.20	0.13
Insulation	0.10-0.25	0.15
Piping	0.50-1	0.7
Foundations	0.03-0.13	0.08
Building	0.07	0.07
Structure	0.05	0.05
Fireproofing	0.06-0.1	0.08
Electrical	0.07-0.15	0.11
Painting and clean up	0.06-0.1	0.08
$\sum f_i$		1.45

Total direct cost is formulated as:

$$Total\ direct\ cost=(1+\sum f_i)I_E \tag{15}$$

The total direct cost value of the methanol/glycerin distillation unit is \$ 3,905,863.75. With the value of direct cost found, the value of indirect cost consisting of 30% contractors overhead and profit, 13% engineering fee, and 13% contingency, fi total value of 1.56 is obtained. Total fixed capital investment (FCI) is found through the following calculation:

$$Total\ FCI=(1+\sum f_i)fI_E \tag{16}$$

The obtained total value of fixed capital investment is \$ 6,093,147.43 as the value/cost required to create construct a set of methanol/glycerin distillation unit. The distillation unit operation cost is calculated using the values and variables shown in Table 11.

Table 11. Distillation unit main equipment, capacity, and estimated total delivered cost.

Variable	Value/Cost (\$)
Q (Production rate)	70,080.,00 lb/year
E	0.1
D	0.1
Im	0.2
T	0.5
A	0.15
Investment	6,093,147.43
Manufacturing cost (aI)	913,972.12
Utility (b)	3
CI	23,461,552.46

The cost (C) obtained from the calculation is \$ 234,615,524.6. The minimum glycerin product price (P) is assumed to be \$ 3.39/lb, adjusted to the price of glycerin in the market in the year 2020. Through this determined value, the utility value (b) is found to be 3. All these values are in accordance with the following equation:

$$C=aI+bQ+cL$$

(17)

$$P=(C+0,5\ I_F)/Q$$

(18)

6 Conclusion

Through the analysis of the process system of the methanol/glycerin distillation unit, the final amount of degrees of freedom from the distillation column, reduced by the number of variables of known values, is C + 2. From the cost estimation calculations, it is found that the total value of fixed capital investment is \$ 6,093,147.43 with a total cost of the main equipment of \$ 1,594,230.10 as well as total direct cost of the methanol/glycerin distillation unit of \$ 3,905,863.75. The cost value of the process operation is obtained at \$ 234,615,524.6 with a minimum product price of the resulting glycerin of \$ 3.39/lb. From these data, it is hoped that optimization both in design and cost for the methanol/glycerin distillation unit can be carried out optimally.

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