Performance and cost analysis of hydrogen production from steam reforming and dehydrogenation of ethanol

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Abstract. Mitigation of carbon dioxide (CO₂) emission has been a worldwide concern. Decreasing CO₂ emission by converting it into higher value products such as methanol can be a promising way. However, hydrogen (H₂) cost and availability are one of key barriers to CO₂ conversion. Ethanol can be a sustainable source for H₂ due to its renewable nature and easy conversion to H₂-rich gas mixtures through ethanol steam reforming process. Nevertheless, steam reforming of ethanol generates CO₂. Hence, this research is focused on different methods of H₂ productions about a 1,665.47 t/y from ethanol for supplying to methanol plants was performed using Aspen PLUS V10. The ethanol steam reforming process required the lowest required ethanol feed for a certain amount of H₂. In contrast, the ethanol steam reforming process presented significant amount of CO₂ emission from reaction and electricity consumption. But the ethanol dehydrogenation of ethanol not only generates H₂ without CO₂ emission from the reaction but also ethyl acetate or acetaldehyde, which are value chemicals. However, ethanol dehydrogenation processes in case II and III presented relatively higher cost because by-products (ethyl acetate or acetaldehyde) were rather difficult to be separated.

1. Introduction
Nowadays, carbon dioxide (CO₂) emission is the main factor affecting the environment because CO₂ emissions from industrial activities and the burning of fossil fuels will pump an estimated 36.8 billion metric tons of carbon dioxide into the atmosphere. And total carbon emissions from all human activities, including agriculture and land use, will likely cap off at about 43.1 billion tons in 2019 [1]. Therefore, mitigation of carbon dioxide (CO₂) emission has been a worldwide concern due to CO₂ emissions is a key contributing factor to the high emissions of green-house gases (GHG) globally. The report was presented 76.7% of GHG emissions emanate from emissions of CO₂ [2].

Methanol is a versatile chemical. It is important intermediate in manufacturing many other chemicals such as formaldehyde and acetic acid, both of which are important feed stocks for the polymer industry. Moreover, methanol can be converted into dimethyl ether (DME) which serves as a replacement for liquefied petroleum gas (LPG) or compressed natural gas (CNG) [3, 4]. However, for each ton of methanol produced from syngas generates 0.6 - 1 tons of CO₂ [5]. CO₂ can react with hydrogen (H₂) to form methanol. Therefore, conversion of CO₂ into methanol can be a promising way to mitigate CO₂ emission. However, H₂ source and availability is one among barriers of CO₂ conversion. For CO₂ conversion to value produce, H₂ must be green H₂ from renewable sources. For conventional H₂
production, steam reforming of methane is the most widely used technique and covers over 50% of the world’s H\textsubscript{2} production [6]. The key reactions are:

\begin{align*}
\text{CH}_4 + \text{H}_2\text{O} & \leftrightarrow \text{CO} + 3\text{H}_2, \quad \Delta H_0 = +203 \text{ kJ/mol} \quad (1) \\
\text{CO} + \text{H}_2\text{O} & \leftrightarrow \text{CO}_2 + \text{H}_2, \quad \Delta H_0 = -41.2 \text{ kJ/mol} \quad (2)
\end{align*}

Steam reforming of methane requires high operating temperatures (700-1100 °C) and still promotes CO\textsubscript{2} emission [6]. Electrochemically water splitting has been used in green hydrogen production. However, it was reported that the water split H\textsubscript{2} fed methanol production from CO\textsubscript{2} plant is still uneconomically feasible [6-9]. Bio-based chemicals e.g. bioethanol for H\textsubscript{2} production process has become interesting. The use of ethanol for this purpose represents an opportunity to produce H\textsubscript{2} from renewable sources.

Bioethanol takes the widest slice of the production of biofuels worldwide. It is derived from the fermentation of polysaccharides, obtained from the processing of certain agricultural products, almost exclusively from sugar cane and corn [10]. In 2018, bioethanol is becoming over supply in Thailand. Ethanol consumption growth is expected to slow down due to increasing of electricity’s involvement in the transport [11, 12]. To increase bioethanol consumption, bioethanol can be used to produce H\textsubscript{2} through ethanol steam reforming process as shown in Equation (3):

\begin{equation}
\text{C}_2\text{H}_5\text{OH} + 3\text{H}_2\text{O} \leftrightarrow 6\text{H}_2 + 3\text{CO}_2, \quad \Delta H_0 = +174 \text{ kJ/mol} \quad (3)
\end{equation}

As presented in Equation (3), although steam reforming of ethanol generates CO\textsubscript{2}, it was reported that biomass energy and carbon capture and storage lead to a net removal of atmospheric CO\textsubscript{2} [13]. Alternative to steam reforming of ethanol was ethanol dehydrogenation for H\textsubscript{2} production. Moreover, dehydrogenation of ethanol not only generates H\textsubscript{2} without CO\textsubscript{2} emission but also ethyl acetate or acetaldehyde, which are value chemicals [14, 15], as presented in Equations (4)-(5).

The dehydrogenation of ethanol to acetaldehyde:

\begin{equation}
\text{C}_2\text{H}_5\text{OH} \rightarrow \text{CH}_3\text{CHO} + \text{H}_2, \quad \Delta H_0 = +16.45 \text{ kJ/mol} \quad (4)
\end{equation}

The dehydrogenation of ethanol to ethyl acetate:

\begin{equation}
2\text{C}_2\text{H}_5\text{OH} \rightarrow \text{CH}_3\text{COOCH}_3 + 2\text{H}_2, \quad \Delta H_0 = +5.98 \text{ kJ/mol} \quad (5)
\end{equation}

In this study, different processes of hydrogen productions from ethanol has been simulated for supplying to CO\textsubscript{2} to methanol plants. Steam reforming of ethanol is compared with dehydrogenation of ethanol. Dehydrogenation of ethanol not only generates H\textsubscript{2} without CO\textsubscript{2} emission but also ethyl acetate, which is value chemical. Techno-economic analysis of the H\textsubscript{2} production process was carried out in term of H\textsubscript{2} productivity, energy consumption and cost. Capital investment and operating cost were estimated. The study was based on the amount of H\textsubscript{2} available to convert CO\textsubscript{2} to methanol on conventional process of methanol synthesis.

2. Simulation method

A comparative study on hydrogen production from ethanol processes (ethanol steam reforming and ethanol dehydrogenation) was performed using Aspen PLUS V10. The ethanol dehydrogenation was divided into 2 processes. Two different processes of ethanol dehydrogenation reaction were described in 2 main products as follows: ethanol dehydrogenation to acetaldehyde with hydrogen production and ethanol dehydrogenation to ethyl acetate with hydrogen production. The low-pressure sections using properties database was used along with the Non-Random Two Liquid (NRTL) fluid package and henry’s law involved in supercritical components. This model has been applied to a full set of experimental data with a good accuracy for ethanol steam reforming process (case I) [10, 16]. However, NRTL model could not suitable predict high pressure (P >10 bar) in ethanol dehydrogenation processes (case II and case III). The Predictive Redlich-Kwong-Soave (PSRK) equation of state provides satisfactory predictions of the ethanol dehydrogenation at high temperature and pressure [14].
The capacity of the H\textsubscript{2} synthesis unit was based on the approximately amount of H\textsubscript{2} (about 1665.47 t/y), which was available for CO\textsubscript{2} conversion to methanol plant. This capacity was related to the work of Khunathorncharoenwong et al. [17], that presented CO\textsubscript{2} utilization of 11,565 t/y and methanol production of 7,427 t/y.

2.1. Ethanol steam reforming process
For H\textsubscript{2} production process through ethanol steam reforming, the kinetic models used in the reformer was based on power-rate law expression (Eq. (6)). The ethanol steam reforming on CeO\textsubscript{2}-supported Pt/Ni catalyst in a fixed bed with isothermal reactor has been proposed by Vincenzo Palma et al. [16].

Table 1 presents the main possible reactions to describe the steam reforming of ethanol, suggesting a set of reactions as in Equations (3), (7)-(9) with an activation of energy and kinetic constant.

\[ k = A \exp \left( \frac{-E_a}{RT} \right) \]  

The rate expression was reported at constant temperature of 527 ℃ under atmospheric pressure for use in case I.

| Reaction | Activation energy, \( E_a \) (kJ mol\(^{-1}\)) | Kinetic constant, \( k \) (m\(^3\) kmol\(^{-1}\) s\(^{-1}\)) |
|----------|----------------------------------|----------------------------------|
| \( \text{C}_2\text{H}_5\text{OH} + 3\text{H}_2\text{O} \leftrightarrow 6\text{H}_2 + 3\text{CO}_2 \) | (3) | 17 | 21.2 |
| \( \text{C}_2\text{H}_5\text{OH} \leftrightarrow \text{C}_2\text{H}_5\text{O} + \text{H} \) | (7) | 74 | 45,231 |
| \( \text{C}_2\text{H}_5\text{O} \leftrightarrow \text{CH}_3 + \text{CO} \) | (8) | 181 | 6,781 |
| \( \text{CO} + \text{H}_2\text{O} \leftrightarrow \text{H}_2 + \text{CO}_2 \) | (9) | 74 | 5,132 |

2.2. Ethanol dehydrogenation to ethyl acetate
In this study, the results from Carotenuto et al. [14] was used to describe the ethanol dehydrogenation to ethyl acetate on a copper/copper-chromite based catalyst with isothermal reactor. The involving reactions are reported as presented in Eqs. (10)-(12). The endothermic reaction (\( \Delta H_0 = +5.98 \text{ kJ/mol} \)) in gas phase was based on a constant temperature of 260 ℃ and a pressure of 20 bar.

\[ \text{C}_2\text{H}_5\text{OH} \rightarrow \text{CH}_3\text{CHO} + \text{H}_2 \]  
\[ \text{C}_2\text{H}_5\text{OH} + \text{CH}_3\text{CHO} \rightarrow \text{CH}_3\text{COOCH}_2\text{CH}_3 + \text{H}_2 \]  
\[ 2\text{CH}_3\text{CHO} \rightarrow \text{Other products} \]

The conversion of ethanol was reported at 70.63%. The selectivity of ethyl acetate was 4.25% (Eq. (11)) while the selection of acetaldehyde was 0.52% Eq. (10).

2.3. Ethanol dehydrogenation to acetaldehyde
In case III, simulation of H\textsubscript{2} production from ethanol dehydrogenation to acetaldehyde was carried out. According to the work of Matsumura et al. [18], the reaction was done over silica catalyst with isothermal reactor. The conversion of ethanol was reported at 34.3% while the selectivity of acetaldehyde was 96% and the selectivity of ethylene was 2%. Ethanol dehydrogenation proceeded at reaction temperature of 396.85 ℃ and pressure of 20 bar. Moreover, the ethanol before fed to reactor was preheated at 896.85 ℃ due to the endothermic reaction.

2.4. Cost estimation
The following section shows the cost analysis related to H\textsubscript{2} production processes. The cost of the processes was estimated based on the capital and the operating costs.
2.4.1. Capital cost. The capital cost of H\textsubscript{2} production process was estimated by using the Module costing technique. The sizing of equipment for the estimation was obtained the simulation data from Aspen Plus V10 software. The bare module cost (C\textsubscript{BM}) was calculated from the purchased equipment and installation costs. The chemical engineering plant cost index (CEPCI) was used to accommodate inflation rate. Working capital was assumed at 15% of total capital investment [19]. Fixed cost investment and total capital investment are presented in Equations (13) and (14), respectively.

Fixed cost investment (FCI):

\[
FCI = 1.18 \sum_{i=1}^{n} C_{BM_i}
\]  

Total capital investment (TCI):

\[
TCI = \frac{100}{85} FCI
\]  

2.4.2. Cost of manufacturing (COM). The overall of H\textsubscript{2} production process, including the cost of manufacturing without depreciation (COM\textsubscript{d}) was calculated by fixed cost investment, operating labor cost (C\textsubscript{OL}), utility cost (C\textsubscript{UT}), waste treatment (C\textsubscript{WT}) and raw material cost (C\textsubscript{RM}) by using Equation (15):

\[
COM_d = 0.180FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM})
\]  

The raw material cost of ethanol was in the range of 0.53-1.06 USD/L [20] This price was based on 99 wt% of ethanol. The utility cost was presented in Table 2.

| Utility          | Unit  | Cost   |
|------------------|-------|--------|
| Cooling Water    | USD/t | 0.0148 |
| LP steam (5 bar, 160 °C) | USD/t | 29.29 |
| MP steam (10 bar, 184 °C) | USD/t | 29.59 |
| HP steam (41 bar, 254 °C) | USD/t | 29.97 |
| Natural gas      | USD/MWh | 11.43 |
| Electricity      | USD/kWh | 0.06  |

Electricity was generated in a power plant based on different fuel: natural gas and coal which releases different amount of CO\textsubscript{2} during electricity generation. For natural gas, the CO\textsubscript{2} emission was estimated at 0.417 kg/kWh during electricity generation process. On the contrary, for electricity generation using coal as fuel, CO\textsubscript{2} was released at 1.002 kg/kWh [21].

The operating labor cost was assumed at 50,000 USD/y [19] with a single operator works 49 weeks/y, 5 shifts/week and 8 h/shift. The number of operating labors (N\textsubscript{OL}) can be determined using Equation (16):

\[
N_{OL} = (6.29 + 0.23N_{np})^{0.5}
\]  

where N\textsubscript{np} is the number of equipment such as compressors, reactors, heat exchangers and towers.

2.4.3. Revenues. H\textsubscript{2} price of 4,872.88 USD/t (1.15 USD/100 SCF) [20] was assumed in the cost. This revenues of H\textsubscript{2} can reduce the cost of raw materials for methanol synthesis. For the value product, acetaldehyde price was 1,005.30 USD/t [20] and ethyl acetate price was 1,110.00 USD/t [20] on industrial requirement (Ethyl acetate > 99.5 wt% and ethanol < 0.2 wt%). The value product can sell to decrease the total cost of manufacturing per ton of H\textsubscript{2}.
3. Results and discussion

3.1. Process description

3.1.1. Ethanol steam reforming process (case I). As presented in Figure 1(a), the ethanol feed stream and the H₂O feed stream were mixed with the recycled gas (M-100) and pumped to 2 bar (K-100). The mixed feed was preheated to 220 °C by heat exchanger (E-100) with high pressure (HP) steam at inlet temperature 250 °C. The feed gas was fed to the isothermal plug flow reactor operated at 527 °C. The product stream was then further cooled (E-101) to 70 °C by cooling water at inlet temperature 20 °C. The stream out of the cooler was condensed for phase splitting at a flash vessel. (V-100) The vapor product from the top of flash vessel was mixed H₂ and CO₂.

3.1.2. Ethanol dehydrogenation process to ethyl acetate (case II). As presented in Figure 1(b), the ethanol feed was pumped to 10 bar in a single stage (K-100) and preheated to 115 °C (E-100) by medium pressure steam (MP) at inlet temperature 175 °C. After that, the stream was pumped to 20 bar (K-101) and preheated to 220 °C (E-100) by high pressure steam (HP) at inlet temperature 250 °C before the stream was fed to the conversion reactor (R-100). The operating condition inside the conversion reactor was maintained at 260 °C and 20 bar. The unreacted ethanol together with product were removed from the reactor to cooler (E-102). The product stream was cooled to 35 °C (E-102) by cooling water at inlet temperature 20 °C. The majority of the ethyl acetate and unreacted ethanol were then separated from H₂ product in flash vessel (V-100). The majority of ethyl acetate mixed with unreacted ethanol from bottom stage. Ethyl acetate and ethanol are azeotrope and cannot be easily separated using a simple distillation column. Complicated refractive distillation is required and will be further studied. It should be noted that ethyl acetate price is 1,110.00 USD/t [20].

3.1.3. Ethanol dehydrogenation process to acetaldehyde (case III). As presented in Figure 1(c), the ethanol feed was mixed with the recycle ethanol in mixer (M-100). The mixed feed was pumped to 5 bar and was preheated to 239.85 °C by heat exchanger with high pressure steam (HP) at inlet temperature 250 °C. After preheat, the ethanol feed was changed to gas phase. The feed gas was compressed to 20 bar and was preheated to 896.85 °C (E-101). The feed gas was fed to conversion reactor operated at 896.85 °C and 20 bar. Product stream was cooled to 50 °C (E-102) with cooling water at inlet temperature 20 °C and the pressure was decreased to 2 bar using valve (VLV-100). After that, the H₂ product was separated from acetaldehyde and unreacted ethanol by flash vessel (V-100). H₂ was removed from the top of the column at operating -22 °C and was heated to room temperature by heat exchanger with low pressure steam (LP) at inlet temperature 125 °C, while acetaldehyde and unreacted ethanol were removed from the bottom. The mixture of acetaldehyde and unreacted ethanol from the bottom stage was heated to 50 °C with low pressure steam at inlet temperature 125 °C. This stream was fed to a distillation column. The atmospheric column with 17 theoretical trays (T-100) was used to purify 99.46 wt% of acetaldehyde from ethanol.
Figure 1. Process flow diagram of: (a) Ethanol steam reforming process (case I), (b) Ethanol dehydrogenation process to ethyl acetate (case II), and (c) Ethanol dehydrogenation process to acetaldehyde (case III).

3.2. Performance analysis

3.2.1. Comparison of required ethanol feed. For ethanol feed was related to mass balance on ethanol steam reforming reaction and ethanol dehydrogenation reaction to produce H₂ (about a 1665.47 t/y.) which was available for CO₂ conversion to methanol [17]. The ethanol feed to each process are shown in Figure 2. It can be seen that steam reforming required least ethanol feed than the steam reforming process since the steam reforming reaction provides the highest stoichiometric H₂. Moreover, in case II, ethanol was not able to be recycled due to separation difficulty. Separation of ethanol and ethyl acetate in case II is among the difficult separation due to azeotrope mixture. Instead of ethanol should be recycled for deceasing of the ethanol make-up.
Figure 2. Required ethanol feed for ethanol steam reforming and ethanol dehydrogenation.

(a)

Figure 3. (a) The hydrogen productivities from ethanol steam reforming and ethanol dehydrogenation (b) CO₂ emission of ethanol steam reforming and ethanol dehydrogenation.
3.2.2. Comparison of H$_2$ productivity and CO$_2$ emission. In this study, productivity of H$_2$ from 3 cases are presented in Figure 3(a) for comparative purposes. The amount of H$_2$ which was required for methanol synthesis is about 1,665.47 t/y. Therefore, the productivity of H$_2$ from all cases can be used in methanol production with about 7,427 t/y of methanol productivity. For ethanol steam reforming, the stream of H$_2$ production was mixed with CO$_2$ which was by-product from the reaction. In contrast, the ethanol dehydrogenation can produce H$_2$ without CO$_2$ containing. CO$_2$ emission from each process is reported in Figure 3(b). Steam reforming in case I produced significant amount of CO$_2$ from the reaction while dehydrogenation in case II and III did not produce CO$_2$ from the reaction. Electricity consumption in processes was converted into CO$_2$ emission according to type of fuel used to produce electricity (natural gas and coal).

3.3. Cost analysis

The overall H$_2$ production process cost is consisted of capital investment and the annualized cost of manufacturing. The capital investment cost of different processes is shown in Table 3 while the cost of manufacturing without depreciation (COM) is shown in Table 4. As presented in Table 3 and 4, case I presented the lowest fixed capital investment cost and cost of manufacturing while case II was relatively most expensive. Ethyl acetate and ethanol are azeotrope and cannot be easily separated using a simple distillation column. Ethyl acetate separation must be improve in order to gain more economical return.

For case II, although acetaldehyde could be separated, the price of acetaldehyde is much cheaper than ethyl acetate in case II. However, based on the fact that current retail price of H$_2$ is about 4,872.88 USD/t, case II and case III presents a potential as H$_2$ production process for CO$_2$ conversion.

### Table 3. Capital investment costs of steam reforming and dehydrogenation of ethanol process.

| Parameters               | Unit  | Case I         | Case II        | Case III         |
|--------------------------|-------|----------------|----------------|------------------|
| H$_2$ Productivity       | t/y   | 1,735.46       | 1,652.50       | 2,309.52         |
| Fixed capital investment (FCI) | USD   | 287,931.18     | 422,354.90     | 4,087,806.01     |
| Total capital investment (TCI) | USD   | 338,742.56     | 496,888.12     | 4,809,183.54     |

### Table 4. Cost of manufacturing of ethanol steam reforming and ethanol dehydrogenation process.

| Parameters               | Unit  | Case I         | Case II        | Case III         |
|--------------------------|-------|----------------|----------------|------------------|
| Raw material: Ethanol    | USD/y | 738,496.63     | 5,601,204.03   | 3,752,929.44     |
| Water                    | USD/y | 124.72         | -              | -                |
| Total raw materials costs| USD/y | 738,621.35     | 5,601,204.03   | 3,752,929.44     |
| Utility                  |       |                |                |                  |
| Cooling Water            | USD/y | 9,974.40       | 41,486.77      | 13,361.39        |
| LP steam                 | USD/y | -              | -              | 475,415.39       |
| MP steam                 | USD/y | -              | 475,415.39     | -                |
| HP steam                 | USD/y | 1,022,129.85   | 993,268.74     | 2,378,757.86     |
| Natural gas              | USD/y | -              | -              | 557,385.87       |
| Electricity              | USD/y | 204,487.56     | 297,463.52     | 350,503.69       |
| Total utility costs      | USD/y | 1,236,591.81   | 1,568,696.98   | 3,775,424.26     |
| Operating labor          | USD/y | 600,000.00     | 650,000.00     | 700,000.00       |
| Total cost of manufacturing| USD/y | 4,119,186.38   | 10,669,502.13  | 11,254,016.95    |
| Cost of manufacturing per ton of H$_2$ | USD/t | 2,373.54       | 4,619.79       | 5,155.48         |
| Revenue of selling value products per ton of H$_2$ | USD/t | 2,373.54       | 4,619.79       | 3,533.84         |

4. Conclusion

The performance and cost analysis of H$_2$ production from ethanol processes were compared between ethanol steam reforming (case I) on CeO$_2$-supported Pt/Ni catalyst at constant temperature of 527 °C.
under atmospheric pressure, ethanol dehydrogenation to ethyl acetate (case II) on a copper/copper-chromite based catalyst based on a constant temperature of 260 °C and a pressure of 20 bar and ethanol dehydrogenation to acetaldehyde (case III) on silica catalyst at reaction temperature of 396.85 °C and pressure of 20 bar. Performance analysis including required ethanol feed, H\textsubscript{2} productivity and CO\textsubscript{2} emission was carried out while cost analysis including capital investment and product cost was performed. For performance analysis, ethanol steam reforming process required the lowest required ethanol feed, followed by ethanol dehydrogenation to acetaldehyde and ethanol dehydrogenation process to ethyl acetate, respectively. The ethanol steam reforming process showed that the lowest fixed capital investment and cost of manufacturing. However, the process presented significant amount of CO\textsubscript{2} emission from reaction and electricity consumption. In contrast, dehydrogenation of ethanol not only generated H\textsubscript{2} without CO\textsubscript{2} emission from the reaction but also ethyl acetate or acetaldehyde, which are value chemicals. However, ethanol dehydrogenation processes in case II and III presented the higher cost because by-products (ethyl acetate or acetaldehyde) were rather difficult to be separated and required further process improvement.

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**References**

[1] Harvey C and Gronewold N 2019 CO\textsubscript{2} Emissions Will Break Another Record in 2019.
[2] Abokyi E, Appiah-Konadu P, Abokyi F and Oteng-Abayie E F 2019 *Energy Reports* **5** 1339-53
[3] Ghosh S and Seethamraju S 2019 *Chem. Eng. Process* **145** 107673
[4] Makoś P, Słupek E, Sobczak J, Zabrocki D, Hupka J and Rogala A 2019 Dimethyl ether (DME) as potential environmental friendly fuel. In: *E3S Web of Conferences: EDP Sciences* p 00048
[5] Blumberg T, Morosuk T and Tsatsaronis G 2017 *Appl. Sci.* **7** 1213
[6] Abbas S, Dupont V and Mahmud T 2017 *Int. J. Hydrog. Energy* **42** 18910-21
[7] Kajaste R, Hurme M and Oinas P 2018 *AIMS Energy* **6** 1074-102
[8] Kim Y-M, Kim-Lohsoontorn P and Bae J 2011 *J. Electrochem. Sci. Technol.* **2** 32-8
[9] Kim-Lohsoontorn P, Paichitra C, Vorathamthongdee S and Seeharaj P 2015 *Chem. Eng. J.* **278** 13-8
[10] Rossetti I, Compagnoni M and Torli M 2015 *Chem. Eng. J.* **281** 1024-35
[11] Nicely R and Counselor A 2018 2018 Biofuels Annual in Thailand
[12] Rossetti I, Tripodi A and Ramis G 2020 *Int. J. Hydrog. Energy* **45** 10292-303
[13] Darde V, Van Well W J, Fosboel P L, Stenby E H and Thomsen K 2011 *Int. J. Greenh. Gas Con.* **5** 1149-62
[14] Carotenuto G, Tesser R, Di Serio M and Santacesaria E 2013 *Catal. Today* **203** 202-10
[15] Luo S, Song H, Philo D, Oshikiri M, Kako T and Ye J 2020 *Appl. Catal. B: Environ.* 118965
[16] Palma V, Castaldo F, CiamPELLi P and Iaquaniello G 2014 *Appl. Catal. B: Environ.* **145** 73-84
[17] Khunathorncharoensangwong N, Charoensappanimit P, Assabumrungrat S and Kim-Lohsoontorn P 2020 *Int. J. Hydrog Energy* In corrected Proof
[18] Matsumura Y, Hashimoto K and Yoshiida S 1989 *J. Catal.* **117** 135-43
[19] Turton R, Bailie R C, Whiting W B and Shaeiwitz J A 2008 *Analysis, synthesis and design of chemical processes: Pearson Education*
[20] Reporter C M 2019 Asia chemicals outlook 2019
[21] Administration U S E I 2018 U.S. electric utility and independent power electricity generation and resulting CO\textsubscript{2} emissions by fuel in 2018