

Article

Moderate Hydrogen Pressures in the Hydrogenation of Alkenes Using a Reactor with Hydrogen Gas Self-Inducing Impeller

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Abstract: The non-oxygenated oil product of the pyrolysis of polypropylene cannot be used directly as an engine fuel due to its high content of alkenes. However, high pressure of hydrogen gas is commonly employed in the hydrotreatment of alkenes to produce alkanes. A semi-batch hydrogenation reaction using a hydrogen gas self-inducing impeller to internally recirculate the hydrogen gas has been implemented in the present work to provide small hydrogen gas bubbles so that the gas dispersion in the liquid phase is intensified. This technique is expected to improve the contact of hydrogen, oil, and the Ni/Al₂O₃ catalyst, which in turn alleviates high pressures of hydrogen gas. The hydrogenation reaction was performed at 185 °C with an impeller speed of 400 rpm. The pressure was varied from 2 to 8 bar. At the pressure of 2 bar, the main reactions are the hydrogenation of alkenes and cyclization of alkenes leading to cycloalkane formation, while at the pressures of 4, 6, and 8 bar, the main reactions are dimerization or oligomerization and hydrogenation of alkenes. The hydrogenation reaction shifts the carbon chain length in the oil towards the carbon chain length attributed to diesel fuel with more branching as the hydrogen pressure is increased. The gas inducement technique employed in the present work has succeeded in saturating almost all alkenes at moderate pressures (below 9 bar), lower than the pressures used by previous researchers, i.e., above 9 bar.

Keywords: hydrogenation; non-oxygenated oil; pressure; self-inducing impeller; Ni/Al₂O₃ catalyst



Citation: Supramono, D.; Yoandi, I.; Fauzi, M.R. Moderate Hydrogen Pressures in the Hydrogenation of Alkenes Using a Reactor with Hydrogen Gas Self-Inducing Impeller. *Processes* **2022**, *10*, 1662. <https://doi.org/10.3390/pr10081662>

Academic Editor: Sandra M. A. Cruz

Received: 11 July 2022

Accepted: 16 August 2022

Published: 21 August 2022

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1. Introduction

The oil obtained from the pyrolysis of polyolefin plastics cannot be used directly as fuel due to its high alkene composition [1]. The presence of alkenes is undesirable because the oil combustion produces NO_x emissions, especially when the oil contains short-chain alkenes [2], and their content reduces the cetane number. In general, the cetane number of compounds with a similar carbon chain length increases in the following order: *n*-alkanes > alkenes > cycloalkanes > aromatic alkyls [3]. Hydrogenation of alkenes is required to obtain oil containing a high composition of alkanes. The hydrogenation can be classified as a hydrotreatment process, which involves a reactant of molecular hydrogen gas.

Hydrotreatment is commonly applied in petroleum reforming to remove oxygen, nitrogen, and sulfur through hydrodeoxygenation (HDO), hydrodenitrogenation (HDN), and hydrodesulfurisation (HDS), respectively. It is also widely applied in the hydrotreatment of biomass components, leading to green chemical synthesis, and in the upgrading of pyrolysis oil (bio-oil) to form high-quality hydrocarbon fuels. However, the contents of nitrogen and sulfur are negligible in bio-oil, but that of oxygen is high. Therefore, HDO is often used in bio-oil upgrading [4]. Hydrotreatment may also be applied to the deoxygenation of triglycerides for their conversion to hydrocarbon fuels, either through HDO or decarbonylation [5]. The review by Kim et al. on the hydrogenation of biomass-derived oxygenates shows that, in general, all processes involving the hydrogenation of olefins and HDO of oxygenates require high pressure of hydrogen gas, mostly above 20 bar [6]. Therefore, attempts at reducing hydrogen gas pressures are advocated due to safety issues and operating costs associated with high gas pressures.

Hydrotreatment involving hydrogen gas in general uses either trickle bed reactors or stirred tank reactors [7]. In the former type, the mixture of liquid and hydrogen gas passes down a fixed bed of catalyst, where bubbles are generated by the collision between the fluids and catalyst particles, while in the latter type, hydrogen-rich recycled gas is delivered as bubbles through a mixture of oil and catalyst particles, where bubbles are generated due to the velocity gradient between the gas and liquid phase. In both types of reactors, it is difficult to generate small bubbles of hydrogen gas because the gas–liquid mixture or gas emerges from ejectors or channels that have a low pressure drop between 1 and 3 bar [7,8]. Measurement of the air bubble size ejected from a channel into water shows that it varies from 0.3 to 1.6 mm from channels [9] or 0.5 to 3 mm from ejectors [10]. With such a low pressure drop, it is difficult to generate a high shear rate between gaseous and liquid phases to induce bubble breakup to form small gas bubbles [7,8]. Moreover, large-sized bubbles easily rise to the liquid surface, which consequently lowers the holdup of bubbles in the liquid phase [11], whereas high gas holdup or high gas dispersion in the liquid phase is of importance to improve the kinetics of the gas–liquid reactions [12].

Additionally, the mass diffusion of polar bio-oil reactants to the surfaces of catalysts used in HDO is inhibited by the presence of non-polar products of HDO. To address this disadvantage, a supercritical condition in HDO reactors is frequently used in hydrotreatment [13]. This condition allows the reduction of the viscosity and the diffusivity of the liquid phase to close to those of the gaseous phase, which eases the mass transfer of polar reactants to non-polar products. However, to reach this supercritical condition, the reactors need high pressure [14].

High kinetics of the hydrogenation reaction may require high pressure of hydrogen gas in order to increase the solubility of the gas in the liquid phase [15]. However, the solubility is low in the liquid phase, containing a high composition of olefins [15], and an external hydrocarbon solvent is added to reactors to improve the solubility [16]. According to Trinh et al. [17], alkanes with long carbon chains can improve the solubility of hydrogen gas.

Some workers have investigated the hydrogenation of alkenes using semi-batch reactors. Lylykangas et al. [18] and Lylykangas et al. [19] described the hydrogenation of isooctenes of 2,4,4-trimethyl-1-pentene and 2,4,4-trimethyl-2-pentene to form isooctane (2,2,4-trimethylpentane), and these papers addressed the use of a catalyst of Ni/Al₂O₃ and Pt/Al₂O₃, respectively. The maximum conversions of isooctenes on Ni/Al₂O₃ and Pt/Al₂O₃ catalysts under the conditions studied were 45% and 78%, respectively, for 5 h reactions at pressures of 20 and 40 bar, respectively. Similar feeds were used by Sarkar et al. [20] using a Pd/Al₂O₃ catalyst, chosen due to its low deactivation by coking, at pressures that varied from 9.6 to 28.6 bar. They found that the hydrogenation conversion of 2,4,4-trimethyl-1-pentene and 2,4,4-trimethyl-2-pentene reached 100% as the reaction was carried out at pressure above 21.70 bar for 2 h.

For the use of Ni catalysts in the deoxygenation of fatty acids derived from triglycerides, Lercher et al. [21] concluded that the highest activity is obtained by using supports of reducible ZrO₂, TiO₂, and CeO₂ instead of Al₂O₃, which improves the adsorption of fatty acids to the catalysts, attributed to the presence of oxygen vacancies. On the other hand, in order to suppress coke deposition and maintain the stability of the catalysts, Papageridis et al. [22] combined CaO-MgO basic sites with a catalyst of Ni/Al₂O₃. This prepared catalyst lowers the amount and the crystallinity of coke deposits and suppresses catalyst sintering.

The present work proposes the utilization of the high-pressure nature of hydrogen gas in the hydrogenation reactor in a different method from the previous researchers, where high shear strain was generated within the liquid phase, which allows the breakup of hydrogen gas bubbles [23] to improve their dispersion in the liquid phase near the impeller to complete the hydrogenation reactions of olefins [24]. To the best of the authors' knowledge, there has been no previous research on hydrogenation reactions occurring in a reactor involving self-inducement of the gas into the liquid phase, utilizing a high pressure drop to improve gas dispersion in the liquid phase, where the effect of gas pressure is the

main concern. Pressures of hydrogen gas chosen in the present work were within the range or lower than those applied by Lylykangas et al. [18,19] and Sarkar et al. [20].

The catalyst of Ni/Al₂O₃ used in the present work was prepared by co-precipitation, and was similar to that used by Lylykangas, which was a commercial catalyst. According to Champon et al. [25], the commercial catalyst of Ni/Al₂O₃ is conventionally prepared by co-precipitation and the content of Ni is approximately 14–17 wt%, which is close to the Ni content in a stoichiometric mixture. In order to suppress coke deposition and maintain the stability of the catalyst, the Ni and Al contents in the catalyst are directed to be in a stoichiometric mixture, which theoretically is associated with a mole ratio of Al to Ni of 2.0 [26], calcination is carried out at temperatures of more than 750 °C for CO₂ reforming [27], and the reaction is carried out at a low temperature [28]. A long-term stability test of the Ni/Al₂O₃ catalyst prepared by calcination at 750 °C has been performed by He et al. [27] for a continuous CO₂ reforming reaction with methane for 200 h at 700 °C. The conversion of methane was observed consistent at 82% during the reaction. For the catalyst preparation comparison, in the present work, the catalyst was prepared by co-pyrolysis and calcined at 700 °C for 5 h. The experiment on pyrolysis steam reforming carried out by Dao et al. [29] demonstrated that using a catalyst of Ni/Al₂O₃ prepared by co-pyrolysis and calcined at 750 °C for 3 h, coke deposits accumulated during the steam reforming reaction of the plastic pyrolysate at 800 °C, reaching 2.92%. For reaction condition comparison, in the present work, the hydrogenation reaction was carried out at 185 °C, which is a much lower temperature compared to that in the experiment of Dao et al., and the reaction used a Ni/Al₂O₃ catalyst prepared with a similar method, i.e., co-precipitation and calcination at 700 °C for 5 h. Therefore, it should be noted that in the present work, coke deposition was estimated to be low and the catalyst could maintain its stability.

The gas self-inducement method employed the recirculation of hydrogen gas inside the reactor by applying a hydrogen gas self-inducing impeller. In this case, the hollow shaft of the impeller took in high pressure hydrogen gas from the gas space above the liquid phase and delivered the gas into the tips of the flat blades of the turbine impeller. The description of the gas self-inducing impeller and hydrogen gas recirculation through the impeller, the hole on the top of the impeller shaft, and the inner side of the shaft are shown in Figure 1a, while the positions of the impeller in the liquid phase and the hole on the shaft in the gas phase are shown in Figure 1b. The incorporation of this technique into the reactor system is shown in Figure 2. The suction occurred due to the low pressure created by the high shear rate in the liquid near the tip of the blades due to crossflow [30]. With the negative pressure of the liquid in the wake of the blades, the bubbles are ingested into the wake [31] and undergo high breakup into small bubbles due to the high vorticity and pressure fluctuation in the wake region [31,32]. The rising velocity of small bubbles is very small [33], thus facilitating high dispersion and holdup of the bubbles in the liquid phase [34]. The high small bubble dispersion allows a high probability of contact between hydrogen gas and solid catalyst particles. Julcour et al. [35] report that the diffusion of hydrogen gas through the liquid–gas interface is limited even if the surface tension in the interface reduces as the hydrogen gas pressure is increased [36]. Therefore, the dispersion of hydrogen gas in oil, in addition to hydrogen gas' solubility in oil, can improve the extent of the hydrogenation reaction. Higher pressure in the reactor resists the buoyancy force of hydrogen gas bubbles in the oil, which improves the gas dispersion and the existence of hydrogen gas in the oil. The action of the impeller, which breaks up hydrogen bubbles into smaller bubbles due to high shear stress in the liquid, improves the possibility of hydrogen bubbles' contact with the catalyst [37]. The present work aimed to study the effect of the gas pressure in the hydrogenation reaction of oil obtained from polypropylene pyrolysis using a gas self-inducing impeller on the oil yield and composition. It was expected that by applying a self-inducing impeller, moderate hydrogen pressure would be sufficient for hydrogenation because it is sufficient to ensure the high dispersion of hydrogen gas bubbles for hydrogenation reactions. This technique involving the internal recirculation of

gases can be applied to hydrodeoxygenation reactions of bio-oil in the future that require catalysts and high hydrogen pressure.

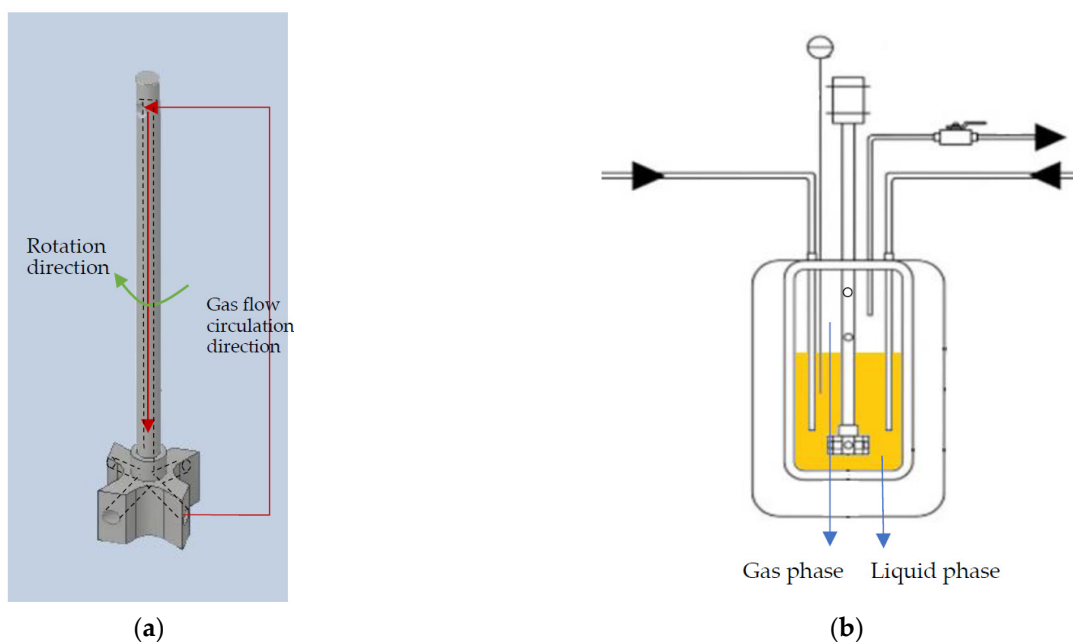


Figure 1. (a) Description of gas flow circulated from self-inducing impeller through liquid phase up to the hole at the top side of the impeller shaft to the vertical hole inside the shaft. (b) Description of gas and liquid phase areas where the gas self-inducing impeller is in the liquid phase and the hole at the top side of the impeller shaft is in the gas phase.

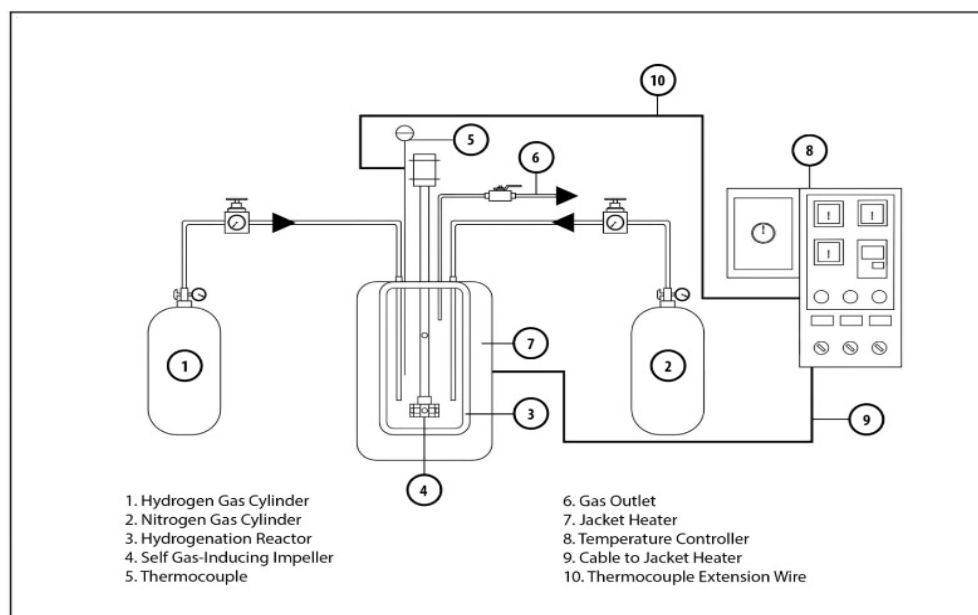


Figure 2. Schematic diagram of hydrogenation reactor with gas self-inducing impeller.

2. Materials and Methods

The present research was carried out in two stages, i.e., thermal slow pyrolysis of polypropylene beads and hydrogenation of oil feed obtained from the pyrolysis. Nitrogen gas was used as the carrier gas in the slow pyrolysis process. The second stage used Ni/Al₂O₃ catalyst. Preparation of Ni/Al₂O₃ catalyst was carried out by a co-precipitation method similar to the one proposed by Seo et al. [38]. First, 5 g of aluminum nitrate

nonahydrate (Merck) and 0.678 g of nickel nitrate hexahydrate (Merck) were dissolved in 40 mL of distilled water (solution A). Ammonia solution of 1 M and solution A were added into 50 mL of distilled water at room temperature, where the ammonia solution was added dropwise to keep the pH value of the mixture at 9, until a blue precipitate was formed (solution B). The precipitate was then aged at room temperature for 24 h. The solid precipitate was recovered from solution B by centrifugation, after which the solid was washed with ethanol, dried overnight at 100 °C, and calcined at 700 °C for 5 h to produce Ni/Al₂O₃ catalyst as a black powder. In this procedure, nickel loading was limited to 20 wt% in the catalyst.

Slow pyrolysis of polypropylene beads was carried out for the preparation of alkene feed to the hydrogenation reactor with the following conditions: mass of polypropylene feed per batch of 250 g, heating rate of 5 °C/min, final temperature of 500 °C, nitrogen pressure of 3 bar, holding time for 10 min at 500 °C, and impeller stirring speed of 400 rpm. The pyrolysis was carried out triplicate to ensure consistency of oil yield.

The oil feed was hydrogenated with the following conditions: oil feed volume of 100 mL, reactor temperature of 185 °C, mass of Ni/Al₂O₃ catalyst of 3 g, impeller stirring speed of 400 rpm, and hydrogen gas pressure varied at 2, 4, 6, and 8 bar. Hydrogenation reactions were carried out for 120 min. Experiments were started by flushing inside the reactor already filled with 150 mL of the reactant. Thereafter, the reaction was initiated by changing the gas feed to hydrogen. Oil feed and oil products of hydrogenation reactions were then analyzed by FTIR spectroscopy using the Thermo Scientific Nicolet iS 5 (Thermo Fisher Scientific, Madison, WI, USA), GC-MS spectroscopy using the PerkinElmer Clarus 680/600T (PerkinElmer Life and Analytical Sciences, CT, USA), and H-NMR spectroscopy using the Agilent NMR spectrometer at 500 MHz (Agilent Technologies, Santa Clara, CA, USA).

3. Results and Discussion

3.1. Catalyst Characterization

Data in Table 1 show that Ni loading is within the stoichiometric composition, i.e., within the range 14–17 wt% [26]. The average pore size of 6.1 nm indicates that the catalyst contains mesopores. In order to improve the activity of the hydrogenation reaction catalyst, for bulky molecules such as the ones in the feed for the reactor in the present work, the catalyst should be mesoporous [39] to overcome the steric hindrance due to branching in polypropylene pyrolysate and the chain length, which spans from C₆ to C₂₀ [40].

Table 1. Physicochemical properties of Ni/Al₂O₃ catalyst.

Ni Loading (%) ^a	Surface Area (m ² /g) ^b	Pore Volume (cm ³ /g) ^c	Average Pore Size (nm) ^d
16.4	250	0.47	6.1

^a Determined by ICP-AES measurement. ^b Determined by BET surface area. ^c Determined by BJH pore volume.

^d Determined by BJH average pore diameter.

3.2. FTIR Analysis

FTIR spectroscopy analysis was conducted to study the chemical bonds present in the oil feed and oil product samples. FTIR spectroscopy results are shown in Figure 3. From this analysis, qualitative comparison of the double bond abundances in the samples can be obtained. The wavelengths chosen for comparison are $\lambda = 1649\text{ cm}^{-1}$ and $\lambda = 886\text{ cm}^{-1}$, which correspond to the transmittance of C=C stretching and bending, respectively. The transmittances at different wave numbers shown in Table 2 indicate that double bond compositions decrease with increasing hydrogenation reaction pressure.

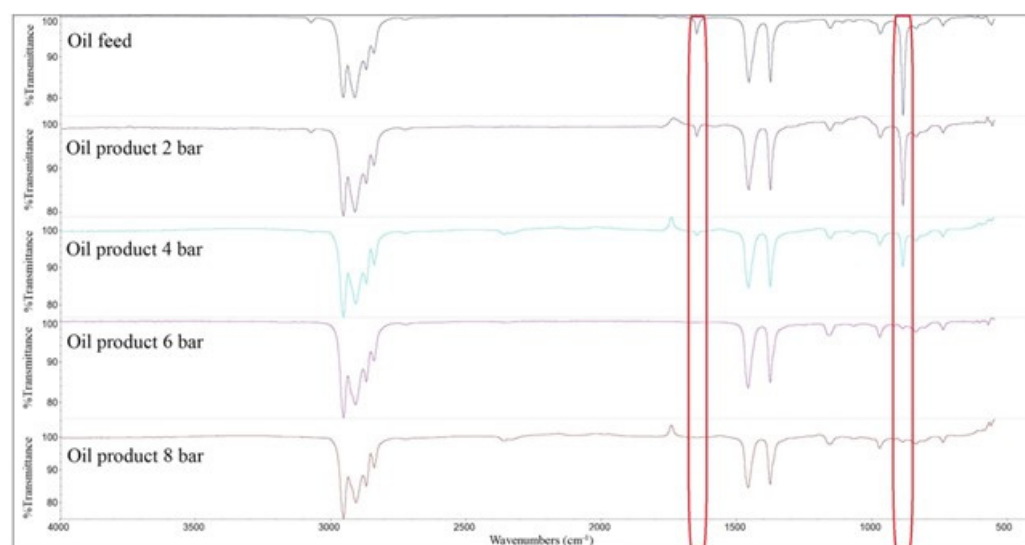


Figure 3. FTIR spectrum for oil feed and oil products for various hydrogenation pressures.

Table 2. Comparison of FTIR transmittance of C=C bonds.

Sample	%Transmittance at $\lambda = 1649 \text{ cm}^{-1}$	%Transmittance at $\lambda = 886 \text{ cm}^{-1}$
Oil feed	95.95	75.58
Oil product 2 bar	97.22	81.39
Oil product 4 bar	98.55	90.19
Oil product 6 bar	100	100
Oil product 8 bar	100	100

3.3. GC–MS Analysis

GC–MS analysis was carried out to investigate the compositions of different carbon chain groups in each functional group of alkanes, alkenes, and cycloalkanes both in the oil feed and in oil products. The carbon chain range from C_6 to C_{12} represents the range in gasoline, while that from C_{13} to C_{20} represents the range in diesel fuel. Figure 4 exhibits the effect of hydrogen gas pressure on the fractions of carbon chain ranges equivalent to those in gasoline and diesel fuel. In general, there is a reduction in the alkene composition for the carbon chain range of C_{13} – C_{20} as the pressure of hydrogen is increased. However, for the alkane composition, this carbon range is predominant in the oil product from the reactor run at 4, 6, and 8 bar, whereas, for the cycloalkane composition, this range is predominant in the oil feed and in the oil product from the reactor operated at 2 bar.

For the case of cycloalkane composition attributed to the carbon chain range of C_{13} – C_{20} , by comparison, the composition of cycloalkanes in the oil product of the hydrogenation reactor operated at 2 bar is higher than that in the oil feed, but also higher than that in the oil product of the reactor run at 4 bar. It is suspected that by running the reactor at 2 bar, with a possible lack of hydrogen solubility and dispersion in the liquid phase, the surface of the catalyst is partly encapsulated by carbon deposition. Bokx et al. [41] concluded that the carbon deposition on the nickel catalyst leads to the formation of a metastable carbide intermediate with nickel, which forms filamentous carbon by decomposition. Seo et al. [38] found that carbon deposition on the catalyst of Ni/ Al_2O_3 prepared by co-precipitation during 1000 min of steam reforming of methane is approximately 42 wt%, which is much higher compared to that for the catalyst prepared by impregnation, which reaches 3.8 wt%. The present catalyst was prepared by co-precipitation with a procedure similar to that carried out by Seo et al. [38]. Therefore, the possibility of carbon deposition

on the catalyst is significant. According to Trunschke et al. [42], a composite between the carbon deposit and transition metal oxides of titanium and zirconium pretreated at 1000 °C induces the dehydrogenation of *n*-octane to alkenes and promotes the aromatization of C₈ alkenes to ethylbenzene and *o*-xylene. In a similar work, Hoang et al. [43] found that a zirconia/carbon composite pretreated at various temperatures from 700 °C to 1600 °C promoted the aromatization of *n*-hexane and *n*-octane. Catalysts used by Truschke et al. [42] and Hoang et al. [43] equally reduced the acid sites' strength of the catalysts and favored aromatic compound formation. With the remaining nickel crystallite exposed to the surface of the Ni/Al₂O₃ catalyst, Savva et al. [44] pointed out that the catalyst of Ni/Al₂O₃ prepared by co-precipitation could convert benzene into cyclohexane. Therefore, during the hydrogenation reaction of the oil feed at the pressure of 2 bar, there could be the formation of cycloalkanes from alkanes and alkenes via aromatics as intermediate products, resulting in a high composition of cycloalkanes.

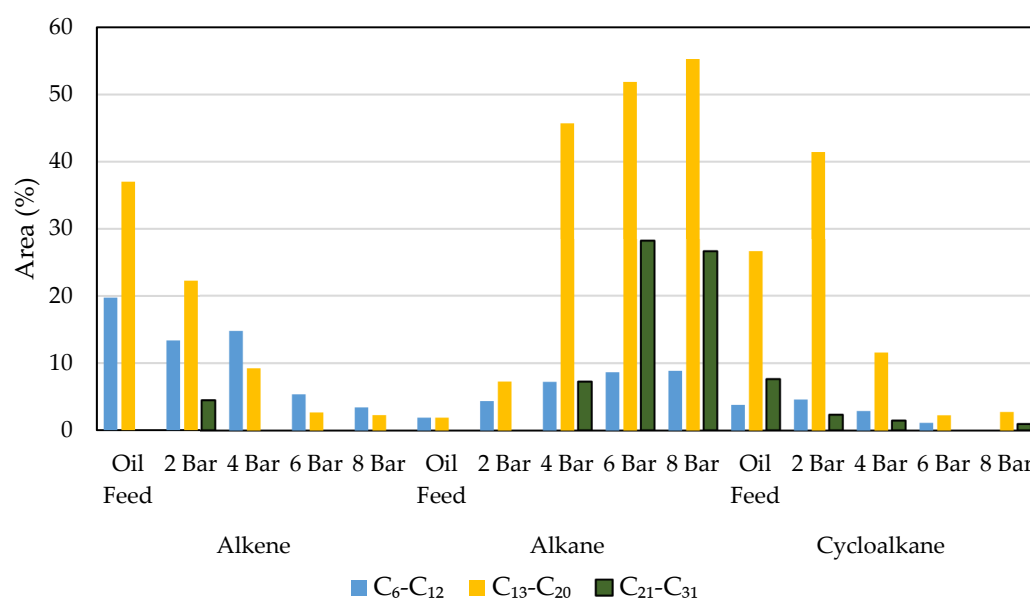


Figure 4. Composition of carbon chain groups in oil feed and oil products.

For the case of the cycloalkane composition attributed to the carbon chain range of C₁₃–C₂₀ within the pressure range of 4 to 8 bar, Figure 4 shows that the increase in pressure from 4 bar reduces the composition of cycloalkanes, which indicates that there is ring opening of the cyclic structure of cycloalkanes. D'Ippolito et al. [45] suggested that there is an optimum ratio of the acid sites to the metal activity characterized by Lewis acid to achieve the ring opening of cycloalkanes. Moreover, Sultan et al. [46] concluded that a catalyst of Ni/Al₂O₃ with Ni weight content of more than 7.9% exhibits cracking over cyclohexane. A similar trend also occurs for the carbon chain range of C₆–C₁₂ by comparing the compositions at different pressures but in a much smaller magnitude. The present catalyst contains around 16.4% weight, therefore favoring cyclohexane cracking. The presence of more molecular hydrogen promotes the ring opening of both cyclohexane and cyclopentane [47].

For the case of alkane composition, Figure 4 exhibits the emergence of the C₂₀–C₃₁ fraction in oil products produced by the reactor operated at pressures of 4, 6, and 8 bar. By comparison with the carbon chain range in the oil feed, these long carbon chain products must be a result of dimerization or oligomerization. According to Olivier-Bourbigou et al. [48] and Joshi [49], the dimerization or oligomerization reaction is initiated by the presence of alkenes and catalyzed by cationic nickel species. The Lewis acid of the Al₂O₃ support can be a strong activator for the generation of nickel cations [49].

The arguments above suggest that both dimerization or oligomerization, as described by the formation of long carbon chains in the reactor operated at 4, 6, and 8 bar, and the cyclization of alkenes, as described by the formation of cycloalkanes in the reactor operated

at 2 bar, require alkenes. Therefore, alkenes in the oil feed may undergo dimerization or oligomerization when the reactor pressure is at 4, 6, and 8 bar, and cyclization when the reactor pressure is at 2 bar, in addition to their hydrogenation.

3.4. H-NMR Analysis and Oil Characteristics

H-NMR analysis shows the abundance of hydrogen in the following groups: methyl, methylene, methine, allyl, and vinyl. The presence of double bonds in H-NMR analysis can be measured by the abundances of hydrogen bonded to vinyl and allyl moieties, as shown in Table 3. These moieties are alkenes, whose abundances are measured by GC-MS spectroscopy, as shown in Table 4. Using H-NMR spectroscopy with a sampling frame size of 256 points, Table 3 shows that the double bonds of vinyl and allyl practically no longer exist in oil products obtained from the hydrogenation reactions operated at 6 and 8 bar. It seems that when the H abundance was less than 1%, the spectroscopy was unable to detect the presence of double bonds considering that H abundances of 1.55% in vinyl could still be detected.

Table 3. Comparison of % abundance of H bonded via different chemical bonds.

Sample	Methyl (%)	Methylene (%)	Methine (%)	Allyl (%)	Vinyl (%)
Oil feed	51.52	14.18	22.15	6.79	5.36
Oil product 2 bar	59.80	14.39	18.35	3.03	4.43
Oil product 4 bar	63.63	15.54	19.27	0.00	1.55
Oil product 6 bar	66.38	17.73	15.89	0.00	0.00
Oil product 8 bar	68.32	16.68	15.00	0.00	0.00

Table 4. Comparison of total % area of different functional groups.

Sample	Alkane (%)	Cycloalkane (%)	Alkene (%)	Branching Index
Oil feed	3.81	37.96	56.74	0.5341
Oil product 2 bar	11.62	48.32	40.06	0.7803
Oil product 4 bar	60.13	15.88	23.99	0.7842
Oil product 6 bar	88.60	3.37	8.03	0.8940
Oil product 8 bar	90.64	3.65	5.01	0.9756

H atoms bonded to methylene exist as part of the cycloalkane ring, and the rest as part of the straight chains of alkanes or alkenes. Table 3 shows that H bonded to the methylene composition is much lower than that to methyl. This indicates that oil products contain a significant amount of branched substituents. Methyl may be attached to a main carbon chain at the end of the main chain or at the end of branched chains. However, Figure 4 shows that alkanes, alkenes, and cycloalkanes mostly contain C₁₃–C₂₀ carbon chains, and methyl is expected to be attached to the branched chains. The H-NMR results given in Table 3 can be used to estimate the degree of branching of carbon chains in oil, which gives a qualitative description of the ratio of the average number of CH₃ at the ends of branched carbon chains to the main carbon chains. Yan et al. [50] proposed the notion of the branching index (BI) as follows:

$$BI = \frac{\frac{1}{3}S_{CH_3}}{\frac{1}{2}S_{CH_2} + S_{CH}} \quad (1)$$

where S_{CH} , S_{CH_2} , and S_{CH_3} are the abundances of hydrogen atoms bonded to methane, methylene, and methyl, respectively.

Table 4 shows that the hydrogenation reaction forms more branching structures compared to that in the oil feed, and the higher-pressure reaction induces higher branching. On

the other hand, Figure 4 shows that the carbon chains in oil produced by hydrogenation at 4, 6, and 8 bar are dominated by those of C₁₃–C₂₀ (diesel fuel). The branching indices of the oil feed and oil products are all above that of commercial diesel of 0.40 [51]. This high branching is disadvantageous if the fuel products are to be used as alternatives to diesel fuel due to their low cetane number [52]. To avoid producing high branching in oil products, it may be advised to use polyethylene, which has few branched substituents, as a co-feed in the pyrolysis. [53].

4. Conclusions

Hydrogenation of a non-oxygenated oil feed shifts the carbon chain length in the oil towards the diesel fuel carbon chain with more branching as the hydrogen pressure is increased. Using a gas self-inducing impeller in the hydrogenation reactor allows the saturation of alkenes to be achieved at moderate pressures (below 9 bar), lower than the pressures used by previous researchers, i.e., above 9 bar.

Author Contributions: D.S.: Conceptualization, methodology, formal analysis, validation, writing—original draft, writing—review and editing. I.Y.: Methodology, investigation, formal analysis, writing—original draft. M.R.F.: Resources, visualization, project administration. All authors have read and agreed to the published version of the manuscript.

Funding: The present research work was funded by the Directorate of Research and Community Engagement, Universitas Indonesia through the scheme of Internationally Indexed Publications with contract number NKB-1721/UN2.RST/HKP.05.00/2020.

Institutional Review Board Statement: Not applicable.

Informed Consent Statement: Not applicable.

Acknowledgments: The authors would like to express their gratitude to the Directorate of Research and Community Engagement, Universitas Indonesia, for its financial support.

Conflicts of Interest: The authors declare no conflict of interest.

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