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Platforms planning and process optimization for biofuels supply chain

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Abstract

Sustainable biofuel supply chain is a key to sustainable manufacturing and the future of production. Greener production is now becoming an order qualifier for the global competition. Modeling biofuel supply chains that achieve economic, social, and environmental feasibility is a challenge. This article develops biofuel platform planning and optimization that unifies biofuel product, production process and networks design into an umbrella of sustainable supply chain planning. A design of biofuel supply chain networks under various production paths is considered. The modeling results show that an optimum region of composition ratio between rice straws and waste cooking oils can be set within the range from 0% to 50%. Bio-diesel is favored over ethanol by occupying over 40% of the total biofuel outputs. However, ethanol yield is 99.1% and therefore it is sufficient to be directly mixed with gasoline at final depots. In terms of social contribution, it is estimated that the supply chain contribution to the case country GDP is about 0.17%. Looking at the above statistics, future research on global economic impacts and competitiveness of biofuel production is suggested.

Keywords: biofuel; optimization; rice straws; used cooking oil; supply chains

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Abstract

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Keywords: biofuel; optimization; rice straws; used cooking oil; supply chains

1 Introduction

This article aims to address the challenge of finding a sustainable solution of biofuel supply chain design from possible sources of domestic used cooking oil and rice agriculture. Without losing the generality of the proposed solution, Indonesia is taken as a case country that has abundant resources of used cooking oil and rice field biomass. The reason is that nowadays Indonesia has been steadily increasing internal oil consumption coupled with a 1/3 drop in domestic production, turning Indonesia into a net oil importer just 12 years after its peak of production. Indonesia has withdrawn from OPEC, because they have no more exportable oil to offer the world market. Therefore, the biofuel production might transfer Indonesian capability to return its position as an oil exporter.

Recently Indonesian policy is to mix 20% of biodiesel from palm oil with 80% of petroleum diesel. Further increase of biodiesel content to 30% in future is also under way. However, using fresh palm oil is less sustainable, since the supply chain has to compete with household demand. Furthermore, this article would like to generate the circular economy to convert wastes into more renewable energy. Recently, rice straws are less utilized and mostly burned or converted into fertilizer, which generates additional carbon emissions. Most end users recycle palm oil that can deteriorate food consumers, which creates further health care cost for the country. Therefore, we transform those by-products into energy.

For Indonesia as one of the largest palm oil exporters, oil palm waste could be an attractive option of biomass. However, we exclude the possibility due to several reasons.

First, even though Indonesia is a net paddy importer during the last 10 years, the country always increases its paddy production by average 3.5% per year [1]. As a result, the increased capacity of paddy production also increases raw material supply for biofuel. Second, unlike oil palm industry that is mostly situated in Sumatra (65%) and Kalimantan (26%), rice fields in Indonesia spread across the archipelago, from Sumatra, Java, and Sulawesi. Rice production data from Statistics Central Bureau (BPS) shows that rice plantations are available in 34 provinces, which is more beneficial in distributing the biomass and biofuel and therefore lower supply chain costs [2]. Third, from environmental issues, in places such as Sumatra, in order to enable palm oil plantations, a hectare of rainforest is destroyed and 174 tons of carbon emissions are released into the atmosphere, which destructs the ecosystems and violates our main objective of protecting the nature [3]. Fourth, fermentation yield of oil palm waste such as empty fruit bunches is less than that of rice straws [4]. In addition, Lahijani and Zainal [5] shows that H_2/CO ratio of empty fruit bunches gasification is between 0.4 and 0.6, which is less than rice straws gasification (> 90%), which makes rice straws more appropriate than oil palm waste to produce liquid biofuel. All of those reasons support our decision to use rice straws as a potential raw materials of biofuel.

The choice between used cooking oil and rice straws as sources of biofuel are based on the following reasons. Used cooking oils are available widely across the country with regards to the consumption of the oils for daily cooking activities. In addition, according to the National Academy of Sciences Proceedings, the potential of rice straw as a raw material is that it may reduce the combine-climate-change and health costs from \$469 million to \$208 million [6]. Thus, the potential applications are vast, since the products can complement and even substitute crude oil based fuels. For farmers, the opportunity of supplying biofuel plant increases economic value of farmland by diversifying the outputs from land utilization to fuels production.

While fuel production from biomass is promising, due to its technology complexity and capital investment, biomass conversion to biofuel cannot directly be sold as transportation fuel. It does not meet Euro-4 specifications, nor does it comply with any of the fuel categories in the World-wide Fuel Charter. In academia, very few researchers have access to unrefined Fischer-Tropsch (FT) products in sufficient quantities to do meaningful research on Fischer-Tropsch refining. Studies in refining should ideally have both academic and industrial significance. Furthermore, the naphtha, which is in the motor-gasoline boiling range, has to be sold as paraffin's in the chemicals market, or as cracker feedstock, due to its poor transportation fuel properties.

Some contributions on the modeling of biomass to biofuel production address the above challenge by increasing the quality of process. De Clerk [7] highlighted FT refinery design to optimize the production of on-specification motor-gasoline, jet fuel and diesel fuel from biomass. The FT receives input from biomass gasification of biomass. Other contributors assessed both technical and economic aspects of gasification in order to measure its environmental impacts [8]. You and Wang [9] did a techno-economic optimization in order to trade off among capital, operating, transportation, and storage costs of distributed

processing networks. The contribution highlights the importance of operational costs sharing that considers not only biomass to liquids production but also liquids supply chain. Zhang et al. [10] develop a simulation model of a biomass feedstock supply chain for a biofuel facility in order to monitor the quality and moisture content, and to evaluate supply chain economic performance and environmental impacts. Zhang et al. [11] developed a facility location optimization model of biofuel facilities. The objective is to minimize inventory, travel distance, energy consumption, and GHG emissions. The authors combine optimization and simulation model in order to update costs and distance parameters. The solution of biofuel supply chain design needs to consider both process and networks design.

For the country under study, the design of biofuel supply chain has another challenge in terms of the unreliability of supply chains infrastructure and suppliers. The first instance, Indonesian farmers often use rice straws for fertilizer, feeding cattle, or even electricity production [12]. Therefore, no commercial rice straws to biodiesel conversion but sugar cane in Indonesia and even worldwide, can be used as references [13,14]. The second instance, while used cooking oils are mostly used for producing detergent or other forms of soaps, most of them are re-used for cooking that leads to cancer disease. Those practices contribute to air pollution in terms of releasing carbon dioxide to the atmosphere and non- safety food distribution to the society. Therefore we could then generalize the challenge to raise a question on how can a bio-fuel supply chain give positive impacts on economic growth, carbon emissions reduction and social development of local community.

To address those challenges, we propose a supply chain design for biofuel supply chains that accommodate flexible biofuel synthesis from two different sources of feedstock, domestic used cooking oil, and rice agriculture. The supply chain includes 1) paddy plantation and transportation, 2) used cooking oil collection and transportation, 3) woody biomass transportation, biomass processing, biofuel distribution. Supply chain optimization is a way to realize long term vision of an organization towards sustainable production that emphasizes on the environmental, societal and economic aspects of a firm's actions [15]. At a glance, the inclusion of impact assessment leads to what is called circular supply chain (CSC) and furthermore circular economy. In circular economy, any contributors to the economic system must be able to return any disposed product into something recyclable. For instances, 1) process waste must be able to be reused by the process, while the process must be able to suffice its own energy needs taken from their by-products, 3) any process must be able to be assessed in terms of their carbon footprint and energy consumption, and 4) the supply chain must be able to generate sustainable economy in the area where the facilities are established. Therefore, biodiesel supply chain must benefit to farmers, factory as well as employment and incomes in the country.

The remainder of this study is organized as follows: A discussion of the literature related to biodiesel conversion is presented in the beginning, after which the proposed CSC model is presented. Optimal control is used to optimize producer profit. Afterwards, the simulation result is presented, followed by the highlights of the managerial implications. Finally, this study presents a summary of the findings of the method and recommendations for its further development and practical application.

2. Research methods

A preliminary study of biofuel production from rice straws and used cooking oils in Indonesia is developed to assess supply chain sustainability. The reasons are provided as follows: 1) Consumption on cooking oils in Indonesia are ever increasing, 2) Rice straws are mostly burned after paddy harvesting and , 3) the country use less biomass than any other countries in the region, 4) None of biomass consumptions are used for generating liquid bio-fuels, which have a better economic value. Considering those reasons, Indonesia biomass assessment may results in fruitful implementation.

This article proposes unified product, process and supply chain planning to obtain optimum configuration of biofuel supply chain. For alcohol platform, we follow the design of rice-straws based ethanol from Kristianto and Zhu [16]. For biofuel from rice straws platform, the scope of the study includes rice plantation and harvests, bio-fuel synthesis, and logistics activities to deliver the product from farm lands to export terminals. For used cooking oil supply chains, the scope of the study includes collecting activity, residues transportation to the nearest biofuel plants, biofuel delivery storage facility and transportation to export terminals, and shipping to customers depots.

The supply chain process planning is solved by combining process synthesis simulation and optimization. Due to the nature of the rice fields, households consumption and biodiesel plant environments are changing over time, stochastic model of the supply chain planning is formulated to give a realistic representation about the system under studied.

The given information includes

1. Raw materials locations and availability (mass and types)
2. Biofuel (and other byproducts) demand, specifications, and prices
3. Utility and other raw materials (e.g., hydrogen, acid, water, etc.) availability and prices
4. Delivery time windows of transporter, volumes and crude types of their loads
5. Configuration details (numbers of biofuel factory, crude distillation units (CDUs), storage tanks, jetties and their interconnections) of the supply chain
6. Holdup in the pipeline and limits on flow rates from the biofuel station and jetties to tanks and from tanks to CDUs
7. Limits on CDU processing rates
8. Storage tank capacities, their initial inventory levels and initial volume fractions of crudes in tanks
9. Information about modes of crude segregation in storage and processing
10. Information about impurities limits during storage and processing
11. Economic data such as sea waiting costs, unit transportation costs, operating costs, etc.
12. Production demands during the scheduling horizon. These are normally available from the monthly production plan of the refinery.

Determine:

1. A detailed routings schedule for each transporters

2. Inventory and composition profiles of storage tanks
3. Technical performances (yield, conversion, biofuel distribution).
4. Logistic networks with multi-modal transportation (ship, train, pipe, etc.).
5. Benefit to costs ratio of each platforms

Fig.1 System boundary of biofuel synthesis from used cooking oil and rice straws

Figure 1 shows how the supply chains works. Households, local food business, and farmers $f \in \{1, F\}$ are suppliers of biomass materials. A set of transporters collect the biomass from each collecting point CP_n $n \in \{1, N\}$ for farmland biomass, $m \in \{1, M\}$, and distributed them to several available processing factories F_F that has several technology options $p \in \{1, P\}$, before finally sent to depots $n \in \{1, D\}$.

2.1 Goal and scope definition

Figure 1 shows the system boundary of the fuel synthesis from used oil and rice straws. The goal of this study is to assess the feasibility of global biofuel supply chain from biomass and used cooking oil. The assessment includes economic and environment impacts based on CO_2 and energy performance of production process from rice plantation or and from cooking oils collection to biodiesel transfer to user machines. The functional unit is one unit mass of raw materials. The scope of the study includes raw materials transportation, storage, production, crude fuels transportation, purification, final liquid fuels transportation and distribution.

2.1.1 Technology options of biodiesel synthesis

There are several options of synthesizing biodiesel, which offers several advantages and disadvantages in points of view of economy, environment and society. The first path is transesterification of palm oils that react with alcohol in the presence of potassium hydroxide (KOH) or sodium hydroxide (NaOH) as a catalyst that breaks down the oils into alkyl ester and releases glycerol [17]. The second path is the hydrogenation of oil palms that yield biodiesel through Ni, Mo or zeolite catalysts [18]. The third path is to use woody biomass or

farmland biomass as Fischer Tropsch (F-T) synthesis process raw materials, and hydrogenate the heavier hydrocarbons to produce lighter fractions such as biodiesel and gasoline [19].

This article constructs decision making models to choose at least one among those three alternatives. However, having more than one platform is also possible as long as economically feasible. Since sulfur content (19 % to 20 %) of rice straws might be harmful for the catalyst of syngas process, biomass based biodiesel process synthesis needs sulphur removal prior to FT synthesis [20].

2.1.2 Technology description of hydrocracking and Fischer Tropsch of biodiesel

According to Figure 2, the first path is the transesterification of palm oils, converting the used palm oil into diesel directly. The byproduct of the process is glycerol and water that can be used as secondary products. There is no need to have raw material pretreatments that can reduce raw material size and clean it, as required by Fischer Tropsch (FT) process in the third path. The FT process is, however, not necessarily needed to convert cellulose and hemicellulose into simpler monomers, since all compounds will be converted into syngas.

The second and third paths convert biomass and waste palm separately. If the supply chain chooses both paths, heavy products from FT process enter (sulfide NiMo/Al₂O₃) at based catalyst hydrocracker at 237-357 °C [21]. Raw vegetable oil enters hydrocracker at 370 – 390 °C over conventional hydrotreating catalysts (sulfide NiMo/Al₂O₃), and has operating temperature 350 – 390 °C, depending on the products [18]. Therefore, each feedstock (rice straws and used palm oil) has its own hydrocracking facility, and thus they are not mixed during the conversion process.

The outputs of HC and FT are hydrocarbons of diesel, gasoline, and kerosene. However, prior to FT gasification is needed to produce syngas, and it was finally hydrogenated to crack heavier hydrocarbons into smaller and shorter chain products. On the other hand, HC can be directly hydrocracked at a mild temperature and pressure to produce mostly the biodiesel.

Fig 2. Process flow diagram of ethanol synthesis from rice straw under sugar and syngas fermentation platforms.

2.2 Biofuel supply chain model

The supply chain is modeled to minimize total operating costs, from farmers to final biofuel blending operations.

2.2.1 Raw materials collection

Vehicle routing to pick up the raw materials from sources i to destination j . Transportation costs are constrained by using a set of equations (1-7). Such a route from i to j can be only passed once for each trucks of total available m trucks from a biofuel factory (1,2). Eqs. (3) constraints the amount of BET transported by each ship from depot k to k' , after observing the demands at source DEM_k and destination $DEM_{k'}$ as well as ship capacity CAP . Eq. (4) does not permit a vehicle backs to its original source without finishing the trip. Therefore at minimum $v(S)$ number of transporters must be used to visit all vertices S except the source vertices $S \subset V' \setminus \{1\}$. Thus the trip is started by visiting vertices other than source $|S| \geq 2$ (Eq. 5). The availability of transporter depends on the availability of production Y_{it} (Eqs. 6,7). The objective is to minimize total distance $Tot_col_cost_{ij}$ as accumulation of individual collecting cost, $ColCost_{ij}$ at a distance $Dist_{ij}$ (8) from i to j , where i, j are members of vertex V .

$$\sum_{i=1} x_{jk} = m \quad (1)$$

$$\sum_{j=1} x_{kj} = m \quad (2)$$

$$Q_j = q_{jk} + \sum_k q_{kk'} \quad (3)$$

$$q_{kk'} = (q_{jk} + DEM_{k'} - CAP) + CAP x_{kk'} + (CAP - DEM_k + DEM_{k'}) x_{k'k} \quad (4)$$

$$\sum_{i,j \in S} x_{ij} \leq |S| - v(S) \quad S \subset V' \setminus \{1\}; |S| \geq 2, \quad (5)$$

$$x_{ij} \in \{0,1\} \quad i, j = 1, \dots, n; i \neq j \quad (3)$$

$$x_{ij} \leq Y_{it} \quad (7)$$

$$Tot_col_cost_{ij} = \sum_{i,j \in V} Dist_{ij} ColCost_{ij} x_{ij} \quad (8)$$

2.2.2 Biofuel factory decision

There are several options of biofuel process synthesis from biomass (Figure 2). The pathways include several stages of pre-treatment, that is followed by a sequence of Fischer Tropsch process and hydrocracking (FT process) and or hydrocracking process (HC). This article focuses on Fischer Tropsch process and hydrocracker modeling and optimization.

A. Syngas production for FT synthesis

For biodiesel production from rice straws, syngas is the main component of biofuel that is obtained from heating the biomass with hot streams of steam and air. The process includes pretreatment that reduce the size of biomass to be easier to transport and treat in a gasification reactor. The following steps include a gasification process that is followed by water shift gas (WGS) and steam reforming. The entire process from pretreatment to steam reforming can convert essentially all of the biomass, including lignin, to syngas.

In specific biomass can be formulated as follows:

The formula $CH_{1.53}O_{0.66}$ is obtained from C,H,O composition of rice straws [15]. We consider only syngas components H_2, CO, CO_2, CH_4 to determine utilities demands (β , the mole of steam per mole biomass, γ is mole oxygen and therefore $\frac{100}{21}\gamma$ moles of air per mole biomass. Tars include many aromatic hydrocarbon but their quantity are significantly smaller than those main components. Therefore we exclude those from the calculation.

Fig 3. Syngas from biomass production

Prior to entering steam reformer, a steam generator produces steam by mixing boiler feed water (BFW) and condensate F from the water gas shift (WGS) reactor (Eq.10). The mixing temperature of both streams T_{mix} is further increased further to T_{int} after passing the mixture into a preheater so that additional heat from WGS reactor, $Q_{preheater}$ increases the temperature (Eq.11).

While some of boiler feed water is used in steam reforming, 5% blowdown is suggested to keep the content of salt in the water not exceed the allowable limit (Eq.12). In addition, a ratio of 2:1 for steam-methane inside WGS is considered to eliminate carbon deposit on catalyst and therefore boiler feed water mass can be determined according to Eq. (13). Figure 3 of Mahishi and Goswani [21] is used to compose Eqs.(14-18). Eq. (19) calculates water fraction in the gasification outlet. Considering steam separator efficiency for process and reformer, J_{SS} , condensate mass flows F_{cond} must equal to the water content of WGS stream (Eq.20). The outputs of WGS, $F_{WGS,out}$, will be the input of solid separator, $F_{Solid\ separator,in}$ (Eq.21). Steam separation efficiency ϵ_{SS} depends on separator temperature T_{SS} and its steam partial pressure $P_{SS}y_{H_2O,WGSout}$ (Eq.22). Mixing temperature between BFW and condensate, T_{mix} , is approximated from BFW temperature at 383 °C and condensate temperature T_{SS} (Eq.23). Eqs.(24-26) bound operating temperatures in industrial units [22].

$$(F_{BFW} + F_{cond})Cp_{BFW}(T_{int} - T_{mix}) = Q_{preheater} \quad (10)$$

$$Q_{preheater} = F_{WGS,out}C_p(T_{WGSout} - T_{SS}) \quad (11)$$

$$F_{BFW} = 1.05(F_{stm} + 2 \times 18(n_{CH_4,WGSin} + \beta_{WGSin})) \quad (12)$$

$$n_{H_2O} = \beta_{H_2O} - n_{CO} \quad (13)$$

$$n_{N_2} = 0.79\gamma_{O_2} + 3.76N_2 \quad (14)$$

$$n_{CO} = 0.75 - \beta_{H_2O}(0.4/0.25) \quad (15)$$

$$n_{H_2} = 0.3734 - 2.303 \log(\beta_{H_2O}) + 1.2794 \quad (16)$$

$$n_{CO_2} = 0.17 - \beta_{H_2O}(0.65 - 0.17/0.25) \quad (17)$$

$$n_{CH_4} = 0.0707 - \exp(-1.281\beta_{H_2O}) \quad (18)$$

$$y_{H_2O,WGSout} = \frac{n_{H_2O}}{n_{H_2O} + n_{N_2} + n_{CO} + n_{H_2} + n_{CO_2} + n_{CH_4}} \quad (19)$$

$$F_{cond} = F_{WGS,out}(y_{H_2O,WGSout}(1 - \varepsilon_{SS})) \quad (20)$$

$$F_{SS,out} = F_{Solid\ separator,in} = F_{WGS,out}(1 - (1 - \varepsilon_{SS}))y_{H_2O,WGSout} \quad (21)$$

$$\varepsilon_{SS} = \left(\frac{T_{SS} - 273}{100} \right)^4 \frac{101.325}{P_{SS}y_{H_2O,WGSout}} \quad (22)$$

$$T_{mix} = \frac{383F_{BFW} + F_{cond}T_{SS}}{F_{BFW} + F_{cond}} \quad (23)$$

$$400\ K \leq T_{WGSin} \leq 530\ T \quad (24)$$

$$1375\ K \leq T_{WGSout} \leq 1675\ T \quad (25)$$

$$298\ K \leq T_{SS} \leq 400\ T \quad (26)$$

Produced steam is used in gasification that is followed by steam reforming or partial oxidation to eliminate hydrocarbons. This article uses steam reforming by considering that oxygen from air will also have nitrogen that needs more treatment during hydro-treating (HT) process as follows:

Steam reforming (SMR):

The energy balance across steam reformer can be formulated as follows:

$$(F_{BFW} + F_{cond})Cp_{BFW}(516 - T_{int}) + \left(\frac{F_{BFW}}{1.05} + F_{cond}\right)\lambda_{BFW} = Q_{Boiler} \quad (28)$$

$$Q_{Boiler} = F_{GAS,out}C_p(T_{GASout} - T_{WGS,in}) \quad (29)$$

Since the gas stream contains CO₂ and carbon, the stream is cleaned by using three different ways, hot cleaning, cold cleaning or partial swing absorber (PSA). In our simulation model, hot cleaning at 94 °C is used to remove also suspended solid.

The objective function of syngas plant-i is then formulated as follows:

$$Syngas_cost_i = \left(\frac{1}{Q_{Boiler}} + \frac{1}{n_{CO}} + \frac{1}{n_{H_2}} \right)_i$$

B. Syngas fermentation

Syngas gas fermentation is used to convert syngas into ethanol. The detail of the process synthesis and heat-energy balances have been detailed in [16].

C. Fischer Tropsch model

During FT reactions, syngas is converted into hydrocarbons and their chains n growth at different rates r_n that depends on CO conversion rate, $-r_{CO}$, and probability of growth α_n from chain length k to n [23] :

$$-r_{CO} = 0.01 \times T - 2.01 \quad (30)$$

$$\alpha_n = \begin{cases} 0.292 & n = 1 \\ -0.0317n + 1.0362 & 2 \leq n \leq 7 \\ 0.8 & n \geq 8 \end{cases}$$

$$-r_{CO} = \frac{r_1}{1 - \alpha_1} \left(\sum_{i=1}^N \left(\prod_{j=1}^i \alpha_j \right) \right) \quad (31)$$

$$r_1 = \frac{0.001059 P_{CO}^{-0.86} P_{H_2}^{1.32}}{\left(1 + 0.46 P_{H_2} / P_{CO} \right)}$$

While the above reaction rates has CO as a limiting reactant, by assuming constant volume reaction, FT reaction rate $-r_{FT}$ is linear correlated to r_{CO} and is calculated based on CO and H_2 consumption, C_{H_2}, C_{CO} and absorption and reaction rates K, k as follows [24]:

$$-r_{FT} = \frac{k C_{H_2} C_{CO}}{(1 + K C_{CO})^2} \quad (32)$$

$$k = 0.4 \exp \left(\frac{-37400}{RT} \right) \left[\frac{m^6}{kg_{cat} mol s} \right] \quad (33)$$

$$K = 5.10^{-9} \exp \left(\frac{68500}{RT} \right) \left[\frac{m^3}{mol} \right] \quad (34)$$

By the end of growth cycle, hydrocarbons mass fraction distribution can be calculated according to the most well-known and simplest product distribution model is the Anderson-Schulz-Flory (ASF) distribution given below

$$m_n = (1 - \bar{\alpha}) \bar{\alpha}^{n-1} \quad (35)$$

$$\text{For } \bar{\alpha} = \left(0.233 \left(\frac{y_{CO}}{y_{CO} + y_{H_2}} \right) + 0.633 \right) (1 - 0.0039(T - 533))$$

Thus for methane, paraffin, gasoline, diesel, and wax have the following mass fractions:

$$m_{CH_4} = (1 - \bar{\alpha}) \quad (\text{Methane})$$

$$m_{C2-C4} = \sum_{n=2}^4 (1 - \bar{\alpha}) \bar{\alpha}^{n-1} \quad (\text{paraffin's})$$

$$m_{C5-C10} = \sum_{n=5}^{10} (1 - \bar{\alpha}) \bar{\alpha}^{n-1} \quad (\text{Gasoline})$$

$$m_{C11-C17} = \sum_{n=11}^{17} (1 - \bar{\alpha}) \bar{\alpha}^{n-1} \quad (\text{Kerosene})$$

$$m_{C18-C24} = \sum_{n=18}^{24} (1 - \bar{\alpha}) \bar{\alpha}^{n-1} \quad (\text{Diesel})$$

$$m_{C25-C34} = \sum_{n=25}^{34} (1 - \bar{\alpha}) \bar{\alpha}^{n-1} \quad (\text{Wax})$$

(36)

Due to the fact that paraffin's and wax are not the main products, they must be cracked into lighter products such as gasoline, kerosene and diesel.

The objective function of syngas plant-i is then formulated as follows:

$$FT_cost_i = \left(\frac{1}{Q_{Boiler}} + \frac{1}{n_{CO}} + \frac{1}{n_{H_2}} \right)_i$$

D. Hydrocracking for both FT synthesis and cooking oil hydro-treating

Hydrocracking of FT synthesis and cooking oil hydro-treating products occur in different hydrocrackers. Furthermore, the outputs of both hydrocrackers enter refinery process to get diesel, kerosene, and naphtha.

We use experimental data of Martin and Grossmann [19] for the conversion and selectivity of FT process into liquids as a function of the temperature. From the regression analysis, this article formulates products conversion from rice straws material as follows:

$$Conversion(\%) = 1/((-15.99418699 \times \ln T_{hydrocracker}) + 96.69166996) \quad (37)$$

$$Diesel_{select}(\%) = -5.333545685 \cdot 10^{-3} T_{diesel} + 2.840289671 \quad (38)$$

$$Kerosene_{select}(\%) = 16877.53081 e^{(-4196.195605/T_{kerosene})} \quad (39)$$

$$Naphta_{select}(\%) = 5173904.255 e^{(-6432.000463/T_{Naphta})} \quad (40)$$

We use experimental data of Bezergianni et al. [18] for the conversion and selectivity of used cooking oil into diesel as a function of the temperature. The authors did a hydrocracking experiment that were performed at $P = 2000$ psig (13789.5 kPa), $350 \text{ K} \leq T_{reaktor} \leq 390 \text{ K}$, $LHSV = 1.5 \text{ h/l}$ and $H_2/\text{oil ratio} = 6000 \text{ scfb}$ ($1068 \text{ nm}^3/\text{m}^3$). From the experiment data regression analysis, this article formulates products conversion from rice straws material as follows:

$$Conversion(\%) = 1.795596872 \cdot 10^{-1} \ln(T_{hydrocracker} - 293.178567) \quad (41)$$

$$Diesel_{select}(\%) = 3.023415023 \cdot 10^{-1} e^{398.7415509/T_{hydrocracker}} \quad (42)$$

$$Kerosene_{select}(\%) = -472.7896012 / T_{hydrocracker} + 1.402337459 \quad (43)$$

$$Naphta_{select}(\%) = -218.0836957 / T_{hydrocracker} + 0.6796343336 \quad (44)$$

For each raw materials (rice straws and used cooking oil), the mass fraction of HC reactor products m_i are obtained by multiplying conversion% of mix products and specific product selectivity and input syngas M_{HC} . For 1 unit mass of oil, oil products $i \in (\text{Gasoline, kerosene, diesel, wax})$ mass fraction can be determined as follows:

$$m_i = Conversion(\%)_i \times i_{select}(\%) \sum_{i \in HC} M_{HC} \quad (45)$$

$$M_{HC} = n_{CO} \times MW_{CO} + \left(\frac{m}{2} + n \right) H_2 \times MW_{H_2} \quad (46)$$

Considering raw material composition between used cooking oil and rice straws, Eqs.(45-46) can be modified as follows:

$$m_i = \sum_{ricestraws, cooking\ oil \in RM} (Conversion(\%)_i \times i_{select}(\%) \sum_{i \in HC} M_{HC}) RM\%$$

$$M_{HC} = \sum_{ricestraws, cooking\ oil \in RM} \left(n_{CO} \times MW_{CO} + \left(\frac{m}{2} + n \right) H_2 \times MW_{H_2} \right) RM\%$$

While H_2 is the main compound to control reactor temperature and product yields, H/C ratio

1.9 is implemented to determine hydrogen consumption. Mohanty et al. [25] formulated H_2 consumption rate per unit mass of component j to be cracked into lighter component i as follows:

$$(H_2/HC)_j = \frac{\sum_{i=1}^{i=j-2} \frac{m_j}{1 + (C/H \text{ ratio})_i} - \frac{1}{(C/H \text{ ratio})_j} \sum_{i=1}^{i=j-2} \frac{m_j (C/H \text{ ratio})_i}{1 + (C/H \text{ ratio})_i}}{\sum_{i=1}^{i=j-2} \frac{m_j (C/H \text{ ratio})_i}{1 + C/H \text{ ratio}} \left(1 + \left(\frac{1}{(C/H \text{ ratio})_j} \right) \right)} j = \text{kerosene, diesel, FT output} \quad (47)$$

Since HC reaction is exothermic, temperature rise across HC reactor must be controlled in such a way that not exceed the upper limit of catalyst active temperature. Thus hydrocracking heat of reaction $\sum_{j \in J} \Delta H_{Rj}$ is calculated from reactants that is involved into cracking $j \in J$. The reaction heat is equal to heat of reactions of cracking products $i \in I$ that rises the temperature as a function of catalyst weight, W , at dT/dW .

The ΔH_{Rj} components are standard heat of reaction of heavier oils j that are cracked into lighter oils I , and heat of reaction at reaction temperature ΔH_{Rj}^T . ΔH_{Rj}^T is composed from product enthalpy plus hydrogen enthalpy minus lighter oil products enthalpy.

$$\sum_{j \in J} \Delta H_{Rj} = \Delta H_{Rj}^0 + \left(\sum_{i=1}^{j-2} \Delta H_i m_i - \Delta H_j (m_C + m_H) - \Delta H_{H_2} (H_2/HC)_j \right) \quad (48)$$

$$(m_C)_i = \text{Conversion}(\%)_i \times i_{\text{select}(\%)} \frac{C_i}{C_i + 1}$$

$$(m_H)_i = \text{Conversion}(\%)_i \times i_{\text{select}(\%)} \frac{1}{C_i + 1}$$

Hydrogen enthalpy at standard condition (25 C and 1 atm) and reaction temperature $\Delta H_{H_2} = H_{H_2}^T - H_{H_2}^0$ are calculated from the available heat capacity data and regression data of heat capacity as a function of temperature. Heavier and lighter products enthalpy at reaction temperature and pressure is calculated from Peng-Robinson equation of state (EOS) and Lee and Kassler Table. Standard enthalpy is calculated from heat capacity data The specific heat of kerosene is 2.01 kJ/kg.K , diesel is 1.9 kJ/kg.K , naphta is assumed equal to gasoline at 2.1 kJ/kg.K , and hydrogen is 8.7 kJ/kg.K .

Thus temperature change dT per mass of catalyst dW , dT/dW is calculated as follows [20]:

$$dT/dW = - \sum_{j \in J} \frac{m_j}{\sum_{j \in J} m_j} \Delta H_{Rj} k_{Rj} / \sum_{i \in J-2} m_i C_{pi} \quad (49)$$

$$k_{Rj} = 10^7 \exp(21000/RT_{\text{reactor}}) (0.494 + 0.0052T_{BPi} - 0.00002185T_{BPi}^2 + 0.000000321T_{BPi}^3) \quad (50)$$

For $R = 8.314 \text{ J/mol.K}$, $T_{\text{reactor}}, T_{BPi}$ are reactor and boiling temperature of lighter product i (K). Since HC occurs in a plug flow reactor, catalyst weight is related to its geometry (volume, weight, porosity, etc).

E. Product upgrading

Product upgrading is an effort to get a higher purity of biofuel. The process includes crude oil distillation and hydrogen treatment to get a higher product purity. From Crude Distillation Unit simulation by DWSim simulation software, it is obtained that condenser

and reboiler duties are 285 kW/kg and -225.44 kW/kg respectively and optimal HC product compositions is obtained from 15 stages atmospheric distillation.

In addition to reboiler and condenser duties, hydrogen consumption for product upgrading, FT synthesis, and hydro-cracking can be modeled according to the flows shown in Figure 4.

Fig 4. Process flow of oil refinery [21]

In our problem, processing units are $U \in \{ADU, CR, CC, HT, BL, FG, PG, RG, D, FO\}$, where ADU= atmospheric distillation unit, CC = catalytic cracker, CR = Catalytic reformer, HT =Hydro-treater, BL = Blending, are used to process crude oil $S \in (S_1, S_2)$, S_1 = rice straws, S_2 = used cooking oil, that contains compounds $C \in (FG = \text{Fuel gas}, PG = \text{Premium gasoline}, RG = \text{Regular gasoline}, D = \text{Distillate}, FO = \text{Fuel oil})$ according to process $P \in (AD = \text{Atmospheric distillation}, CCR = \text{Catalytic cracking}, CRF = \text{Catalytic reforming}, HTR = \text{Hydro-treating}, CCG = \text{Catalytic cracking of gas oil}, BLD = \text{Blending})$. The final product as $FP \in (FGP = \text{Fuel gas}, PGP = \text{Premium gasoline}, RGP = \text{Regular gasoline}, DP = \text{Distillate}, FOP = \text{Fuel oil})$.

Mass balance constraints can be used for replacing $g(x)$. Mass balance constraints are obtained from [26], in terms of composing matrix elements, $A_{C,S,p}$, that represents yields of product C, either from crude oil S or intermediate products C', that enter process p [21].

$$x_{C/C'} = A_{C,S,p} x_{S/C'} \quad (50)$$

In addition, logical constraints are assigned (i.e., $Y_i = 1$) only if unit u_S is operated and capable of doing specific process p. with an input-output coefficient matrix. Variable u_S is the total incoming mass of component c within crude oil S, and $PROD_{p,S}$ is the production capacity of component of crude oil S in a process p. The following equation (51) can be associated to logical constraint [21].

$$\sum_p A_{C,S,p} PROD_{p,S} + u_S \geq 0 \quad (51)$$

Similarly, the intermediate products C' from crude oil S are mixed to produce final products C at a level $BLENDING_{C'SC}$, with other intermediates has the following material balances before entering blending process [21].

$$\sum_p A_{C',S,p} PROD_{p,S} + u_S - \sum_p A_{C',S,p} BLENDING_{C'SC} \geq 0 \quad (52)$$

Thus transportation capacity of final products are equal to blending capacity [21].

$$TRANSPORT_C = \sum_S \sum_p A_{C,S,p} BLENDING_{S,CI,FP} \quad (53)$$

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In addition to material balance, quality balance is also considered to achieve quality standard of final products, in terms of quality attribute based on crude oil source and intermediate product, $\alpha_{C,S,Q}$. Thus the quality balance can be written as follows [21]:

$$\sum_C \sum_S \alpha_{C,S,Q} \times \frac{\sum_S \sum_p A_{C',S,p} BLENDING_{C'SC}}{TRANSPORT_C} \geq Quality_{FP,Q}^{LB} \quad (54)$$

Since not all processes p ($p \in P$) can be utilized by each processing unit U ($u \in U$), total processing capacity of the unit u , K_u is the summation of its individual processing capacity at certain crude oil S , Z_{pS} , in a condition that the unit is capable of doing process p , $Y_{pU} = 1$.

$$\sum_p Y_{pU} \sum_S Z_{pS} \leq K_u \quad (55)$$

It is also noted that shipping of crude oil from crude oil synthesis plant cannot exceed transportation capacity, u_S^{UB} .

$$u_S \leq u_S^{UB} \quad (56)$$

Thus the profit function of crude oil refining plant can be formulated as the selling price of the final product C , p_C , minus purchasing of crude oil and intermediate products, $p_S, p_{C'}$, and processing cost at processing unit u , Cp_p [21].

$$\pi_{BLENDING} = \sum_C p_C x_C - \sum_S p_S u_S - \sum_C \sum_S p_{C'} u_{C'S} - \sum_p Cp_p \sum_S Z_{pS} \quad (57)$$

Processing cost depends on steam to reforming process and hydrogen consumption that are used to upgrade the quality of final products. Steam reforming needs steam to be reacted with biomass to produce hydrogen. The detail of steam reforming is modeled by using DWSIM 4.0 by taking methane (CH_4) as a hydrogen source. Methane is obtained from top product of crude oil distillation. Hydrogen is produced by producer units (PU) in 1) Hydrogen plant (H_2) and 2) catalytic reformer (CR), and consumed by Hydrocracker (HC), diesel hydrotreater (DHT), kerosene hydrotreater (KHT), cracked naphtha hydrotreater (CNHT), naphtha hydrotreater (NHT), and Hydrodealkylation (HDA). Thus consumer units (CU) are HC, DHT, KHT, CNHT, NHT, and HAD. There is about 5 percent weight (5 % wt) of hydrogen needed by the overall hydro-treating (HT) process.

500 F. Hydrogen management

Hydrogen is used in three phase of hydro-treating, 1) Feed and hydrogen mixing, 2) Reaction (hydrotreating or hydrocracking), and 3) Flash separation. Reaction in the reactor removes sulphur, nitrogen and metals content and meanwhile saturates olefins and some aromatics.

During the reaction, C1-C4 compounds are also formed within a range 1 – 4 % wt. Hydrogen consumption for each consumer units are based on Lambert et al. [27] and the results of flash calculation (how much to go to gas stream and liquid stream). K-values are necessary to be found prior to the calculation of hydrogen and hydrocarbons (i = Naphtha, diesel, kerosene, and wax) vapor-liquid equilibrium across refinery networks. This article uses Wilson formulation that requires critical temperature (T_{ci}) and pressure (P_{ci}) and compressibility factor (ω_i). By considering computational complexity and negligible influence of hydrocarbon compositions, K-values can be assumed to be constant. The composition of liquid x_i and vapor y_i are therefore:

$$\sum_i x_i K_i = 1, \quad \sum_i y_i / K_i = 1$$

For each production unit u , crude oil from rice straws is assumed contains 0.5 – 0.8% of Nitrogen and 0.05 – 0.1% of Sulphur. In addition, crude oil from used cooking oil is assumed contains 0.003 % of Nitrogen and 0.01 % of Sulphur. For simplicity, used cooking oil Nitrogen and Sulphur contents are ignored. By assuming that Nitrogen and Sulphur are distributed evenly into hydrocarbon products, the composition of those compounds do not change in each streams.

In 1 unit mass of hydrocarbon products, there are 0.5 – 0.8 unit mass of Nitrogen and 0.05 – 0.1 unit mass of Sulphur. Each 1 mole of Sulphur (MW = 32) and Nitrogen (MW = 14) requires 1 mole of H_2 (MW = 2). Thus for 1 unit mass of hydrocarbon products, hydrogen demand for hydrodesulphurization and hydrodenitrogenation is $\left(\frac{0.05}{32} + \frac{0.5}{14}\right) \times 2 \leq H_2 \leq \left(\frac{0.1}{32} + \frac{0.8}{14}\right) \times 2$ that is about 14.876 % of the total hydrogen consumption. Thus there is 572% more consumption for other processes (hydrocracking, and saturation) [22].

In addition, the reactors produce C1-C4 occurs during hydrogen injection within a range 1 – 4 % wt of oil feed F_u . If the composition of F originally is z_{iu} , then for each 1 unit mass of hydrocarbon i , hydrogen is produced as much as %wt. H_2 [27], the vapor mass V_u and liquid mass L_u at each producer unit u , together with its compounds vapor-liquid composition ($y_{iu}x_{iu}$) can be formulated as follows:

$$F_u z_{iu} = x_{iu} L_u + y_{iu} V_u$$

$$V_u = ((C1 - C4)\% + \%wt.H_2) \times F_u + \sum_i y_{iu} F_u$$

$$L_u = F_u - V_u$$

The hydrogen purity at producer unit u , q_u is therefore as follows:

$$q_u = \%wt.H_2 \times F_u / V_u$$

Considering that hydrogen production from catalytic cracker cannot meet hydrogen demand, another source of hydrogen from biomass is provided which produce 99.99% pure H_2 through header h . The gas is mixed with hydrogen from production units PU (i.e., catalytic cracking, vacuum gas oil flash) (Eq.60). The amount of hydrogen supplied to process unit

541 $U \in \{PU, CU\}$, Q_u^{UB}, Q_u^{LB} and its purity q_u^{UB}, q_u^{LB} is however are limited at a certain level of
542 upper and lower bounds (Eqs.58 & 59).

$$543 \quad Q_u^{LB} \leq Q_u \leq Q_u^{UB} \quad (58)$$

$$544 \quad q_u^{LB} \leq q_u \leq q_u^{UB} \quad (59)$$

$$545 \quad Q_u = \sum_{h \in H} Q_{uh} + \sum_{u' \in CU} Q_{u'e} \quad (60)$$

546 Hydrogen is supplied to hydrogen header h , Q_h by both production unit PU , Q_{uh} and
547 consumer units CU , $Q_{u'h}$.

$$548 \quad Q_h = \sum_{u \in PU} Q_{uh} + \sum_{u' \in CU} Q_{u'h} \quad (61)$$

549 From a header h , hydrogen is delivered to production units and the excess is purged to fuel
550 gas refinery FGR , $Q_{h.FGR}$ (Eq.62). Therefore there is no accumulation of hydrogen mass
551 inside headers.

$$552 \quad Q_h = \sum_{u \in PU} Q_{hu} + Q_{h.FGR} \quad (62)$$

553 The quality of hydrogen from header h , q_h is determined by the quality of hydrogen
554 production unit u , q_u , and the quality of hydrogen that is produced by consumer units u' ,
555 $q_{u'}^{out}$. Thus the actual hydrogen mass flows at a header h , $Q_h q_h$ is the summation of the two
556 previous streams (Eq.63).

$$557 \quad Q_h q_h = \sum_{u \in PU} Q_{uh} q_u + \sum_{u' \in CU} Q_{u'h} q_{u'}^{out} \quad (63)$$

558 The total supply of hydrogen Q_u^{in} , is the accumulation of hydrogen supply from production
559 unit u to consumer unit u' and vice versa, $Q_{uu'}$ and $Q_{u'u}$, from header to both consumer
560 unit u' and producer unit u , Q_{hu} .

$$561 \quad Q_u^{in} = \sum_{h \in H} Q_{hu} + \sum_{u' \in PU} Q_{u'u} + \sum_{u' \in CU} Q_{u'u} \quad (64)$$

562 If the quality of the product is included, the actual mass flows of hydrogen supply can be
563 formulated as follows:

$$564 \quad Q_u^{in} q_u^{in} = \sum_{h \in H} Q_{hu} q_h + \sum_{u' \in PU} Q_{u'u} q_{u'} + \sum_{u' \in CU} Q_{u'u} q_{u'}^{out} \quad (65)$$

565 Similarly, the excess hydrogen and its quality, is formulated as follows:

$$566 \quad Q_u^{out} = \sum_{h \in H} Q_{uh} + \sum_{u' \in CU} Q_{uu'} + Q_{u.FGR} \quad (66)$$

$$567 \quad Q_u^{out} q_u^{out} = \sum_{h \in H} Q_{uh} q_u + \sum_{u' \in CU} Q_{uu'} q_{u'}^{out} + Q_{u.FGR} q_u \quad (67)$$

568 For each consumer units CU , H_2/Oil ratio $Ratio_u$ is determined in advance based on input
569 oil at each consumer unit x_U [27], and the following equation holds.

$$570 \quad Q_u^{in} q_u^{in} \geq Ratio_u x_U \quad (68)$$

571 In order to get better product quality, the quality of hydrogen input to consumer units must
572 be higher than the output quality of the units.

$$573 \quad q_u^{in} \geq q_u \text{ for } u \in CU \quad (69)$$

574 Finally, the objective of hydrogen management is to minimize costs of quality as well as
575 production.

$$576 \quad C_{H2} = \sum_{u \in U} OC_{Q_u} + OC_{P_u} \quad (70)$$

Eqs.(58-70) are used to manage the use and production of hydrogen in a hydrocarbon refinery.

2.2.5 Supply chain design

The SC design include transportation planning, inventory planning, product planning, production and delivery planning. Transportation planning determines raw materials and products routings as well as the amount of delivery vehicle used by the supply chain, inventory planning determines safety stock allocation for each products, product planning determines how much to produce biofuel (diesel, gasoline, jet fuel), production planning determines how much to produce biofuel and orders raw materials, and delivery planning determines timing and directions of biofuel delivery. The inventory planning can be optimized with regards to minimum.

Fig 5. Supply chain networks of biofuel from biomass and used cooking oil

We follow supply chain design formulation by Kristianto and Gunasekaran [28]. Figure 5 is the superstructure of biofuel supply chain. There are two options of transportation mode, ship and train. We follow a model of ship and train scheduling [28].

Suppose we have a set of feasible vehicle routings Ω that consists of location vertices V^r and transportation arcs A^r at a specific route r , to compose a directed graph $G^r = (V^r, A^r)$. For each G^r , there are two possible decisions. The graph can be chosen as a route to transport biofuel by vehicle $v \in V$ from source τ to destination τ' , that arrives at destination τ' after taking a journey at known lead times T_{rv} (therefore binary variable $\phi_{vr} = 1$ and $\phi_{vr} = 0$ for otherwise) at transportation cost $tcost_{vr}$, by allowing backorders are allowed between τ and τ' as many as $BO_{\tau\tau'}$. The transportation time T_{rv} cannot exceed the maximum allowable arrival time b_v . In addition to direct routes, transshipped routes that pass intermediate destination τ_1 are allowed and indexed within feasible set of extreme points p of $\forall p \in \rho_{\tau\tau'}^r$ for direct routes and $\forall p \in \rho_{\tau\tau_1\tau'}^r$ for transshipped routes. By combining feasible routes and

vehicle, at period t , the feasible route can be rewritten as $Z_{\tau\tau'}^{rt}$. Finally, due to working hours regulation, the number of visited port is limited to φ_{τ} .

Kristianto and Gunasekaran [28] establishes the following set of constraints for ships scheduling and transportation mode choice.

$$\sum_{v \in V} \sum_{r \in \rho_{\tau\tau'}^r} \sum_{t \in T} VH_{vr} \geq \sum_{r(\tau, \tau')} \rho_{\tau\tau'}^r \quad (71)$$

$$\sum_{v \in V} \sum_{L \in \rho_{\tau\tau'}^r} \sum_{t \in T} VH_{\tau L} \leq VMAX_{\tau} \quad (72)$$

$$\sum_{r \text{ first } \tau\tau' \in \rho_{\tau\tau'}^r} \sum_{v \in V} VH_{vr} = \sum_{r \text{ last } \tau\tau' \in \rho_{\tau\tau'}^r} \sum_{v \in V} VH_{vr} \quad (73)$$

$$\sum_{v \in V} \sum_{r \in \rho_{\tau\tau'}^r} \sum_{t \in T} T_{rv} \phi_{vr} \geq a_{\tau'} \quad (74)$$

$$\sum_{v \in V} \sum_{r \in \rho_{\tau\tau'}^r} \sum_{t \in T} T_{rv} \phi_{vr} \leq b_{\tau'} \quad (75)$$

$$\sum_{r \in \rho_{\tau\tau'}^r} DIRECT_{r(\tau, \tau')} + \sum_{r \in \rho_{\tau\tau 1}^r} TRANSSHIP_{r(\tau\tau 1 \tau')} + BO_{\tau\tau'} = DEM_{\tau' t} \quad (76)$$

$$\sum_{r \in \rho_{\tau 0 \tau}^r} TRANSSHIP_{r(\tau 0 \tau \tau 1)} = \sum_{r \in \rho_{\tau \tau 1}^r} TRANSSHIP_{r(\tau \tau 1 \tau')} = \sum_{r \in \rho_{\tau 1 \tau'}^r} DIRECT_{r(\tau 1, \tau')} \quad (77)$$

$$\sum_{r \in \rho_{\tau 0 \tau}^r} DIRECT_{r(\tau 0, \tau)} + \sum_{s \in 1}^S prob(s) DEM_{\tau 1 t} + \sum_{r \in \rho_{\tau 0 \tau}^r} TRANSSHIP_{r(\tau 0 \tau \tau 1)} = \sum_{v \in V} PAYLOAD_v VEHICLE_{vrt} - Z_{r\tau} \quad (78)$$

$$\sum_{r \in \rho_{\tau\tau'}^r} DIRECT_{r(\tau, \tau')} + \sum_{r \in \rho_{\tau\tau 1}^r} TRANSSHIP_{r(\tau\tau 1 \tau')} + \sum_{r \in \rho_{\tau\tau'}^r} DIRECT_{r(\tau, \tau')} - \sum_{r \in \rho_{\tau 1 \tau'}^r} DIRECT_{r(\tau 1, \tau')} - \sum_{r \in \rho_{\tau 1 \tau}^r} TRANSSHIP_{r(\tau 0 \tau \tau 1)} - \sum_{r \in \rho_{\tau 1 \tau}^r} DIRECT_{r(\tau 1, \tau)} = Z_{r(\tau-1)} - Z_{r\tau} \quad (79)$$

$$SS_{\tau'} = \sum_{r \in \rho_{\tau\tau'}^r} DEM_{\tau' t} T_{rv} + z\sigma\sqrt{T_{rv}} \quad (80)$$

Eq.(71) states that the number of ships or trains in use to deliver from τ to τ' at period t , VH_{vr} , cannot less than the number of feasible routes $\rho_{\tau\tau'}^r$ but cannot higher than the amount of available vehicle, $VMAX_{\tau t}$ (Eq.72). At any location τ and route r , the number of vehicle must be equal between arrival and departure (Eq.73). For any vehicle that passes through a certain route r , the vehicle cannot deliver product before the earliest arrival time at location τ' , $a_{\tau'}$ (Eq.74) and the latest arrival $b_{\tau'}$ (Eq.75).

The demands at destination τ' and period t , $DEM_{\tau' t}$, can be met from direct delivery from source τ , $DIRECT_{r(\tau, \tau')}$, transshipped delivery $TRANSSHIP_{r(\tau\tau 1 \tau')}$, through location $\tau 1$, or from previous period backorder from source τ , $BO_{\tau\tau'}$ (Eq.76). All products delivered to a certain transshipment point $\tau 1$ must be delivered directly to final destination τ' (Eq.77). The amount of products that are moving in the supply chain networks must not exceed the capacity or payload of the vehicle after being compensate by its allowance $Z_{r\tau}$ (Eq.78). The unused capacity of a vehicle that gives service to both direct and transshipped routes must be available to negate backorders (Eq.78). Eq.(79) states that backup capacity must be available to cover possible future backorders. Eq.(80) decides on safety stock of total product at location τ to cover demand variations at destination τ' through route r . towards destination τ' at a certain delivery lead times, T_{rv} .

The supply chain objective is formulated as follows:

$$SC_{cost} = \sum_{r \in R} \sum_{v \in V} \sum_{t \in T} shipcost_v.shiptime_{rv} V H_{vrt} + Y_{r(n_1, n_2)} \sum_{r \in \Omega} \sum_{\rho_{rr}^r \in \Omega} DIRECT_{r(\tau, \tau')} + S_{r(\tau, \tau')} * q$$

$$+ (1 - \phi_{r((\tau, \tau'))}) \sum_{r \in \Omega} \sum_{\rho_{rr1}^r \in \Omega} TRANSSHIP_{r(\tau, \tau1, \tau')} . q$$
(81)

Finally optimization model of biofuel supply chain is formulated as follows:

$$\min Tot_col_cost_{ij} + Syngas_cost_i FT_cost_i + SC_{cost} + C_{H2} - \pi_{BLENDING}$$
(82)

Subject to : Eqs.(1) – (81).

2.3 Information exchange and decision making process in the supply chain

The following flowchart describes the solution process of biofuel supply chain modeled by Eqs.(1-74).

Fig 6. Supply chain information and decisions flows

The supply chain demand information can be distributed so that each location can make their own decision based on the information. In that way, each location can decide by itself various local decisions that would not impacts the entire supply chain. There are two types of variables, integer that are related to supply chain, transportation and biofuel production networks design, and continuous that are related to process variables, order quantity, inventory placements, hydrogen management and lead times decisions. The details of the algorithm is exhibited in Appendix A.

2.3 Model data

Various sources of data for raw materials for instances Samuel [29] presented the mechanism of rice straws collection from farmers to collecting points. In addition Statistical Central Bureau of Indonesia provides data of used cooking oil suppliers, rice straws price.

In distributing the refined oil, this article uses logistics data of five main PERTAMINA depots, Plumpang, Semarang, Surabaya, Manggis, and Tanjung Gerem (Tables 1 to 4), oil

refinery location is listed in Table 5 [30].

Table 1. Location indexes [30]

Location	Location number	Location	Location number
Indramayu	1	Tj. Gerem	6
Bojonegoro	2	Pengapon	7
Cilacap	3	Surabaya	8
Balongan	4	T.T Manggis	9
Tuban	5	IJG	10

Table 2. Distances from Biofuel Refinery Points to Destination Depots (Ranked) [30]

	TgGerem	Jakarta	Semarang	Surabaya	TTManggis
Cilacap	1	2	3	4	2
Balongan	3	1	2	4	5
Tuban	5	4	2	1	3

Table 3. Distances from Biofuel Plants to Refinery Plants (Ranked) [30]

	Cilacap	Balongan	Tuban
Indramayu	2	1	3
Bojonegoro	2	3	1

Table 4. Vessel Capacity and Costs [30]

Depot	Kl	Transportation cost/kL (Rp)
MR1	34000	130
MR2	34000	130
MR3	34000	130
MR4	34000	150
pipeline	18000	100

Table 5. Supply Lead Time Uncertainty Calculation Summary [30]

Depot	Average Lead Time (Loading & Voyage in days), d_i	σL	a_i	b_i
Tj. Gerem	4	Uniform(1;2)	0	76.125
Pengapon	4		0	76.125
Surabaya	4		0	76.125
T.T Manggis	4		0	76.125
IJG	4		0	76.125
Cilacap	4		0	72.125
Balongan	4		0	72.125
Tuban	4		0	72.125
Indramayu	4		0	4
Bojonegoro	4		0	4

In Table 5, the supply chain considers delivery lead times uncertainty as a parameter of

delivery time windows.

3. Results and Discussion

The objective of this section is to exhibit the performance of supply chain networks with regards to the following performance measures:

1. Technical analysis with regards to the available references
2. Economic analysis of each platforms
3. Logistics cost and carbon emissions
4. Supply chain impacts to gross domestic product (GDP)

Therefore raw materials supply, biofuel synthesis, and supply chain are three major areas that are concerned. The following three sub-sections analyze those three areas according to their performance measures.

3.1 Raw materials supply

Optimization model Eqs.(82) is solved with regards to Eqs.(1-81) by changing the percentage of used cooking oil as a raw material. The results are fitted to get an equation as follows:

$$\frac{\text{logistics} + \text{RM costs}}{\text{ton raw material}} = -2979 \times \ln (\% \text{ cooking oil in raw materials}) + 19007 \quad (83)$$

The data fitting has coefficient of determination 98,98 % and that's means waste cooking oils content in raw materials is meaningful to be analyzed since the variable has a strong correlation with both logistics and raw materials costs. The value implies that the use of cooking oil in addition to other raw materials (i.e., straws) could reduce the total costs of the biofuel supply chain.

However, Figure 7 reveals the fact that if the materials composition rice straws and waste cooking oils is over 1:1, logistics and raw material costs of the supply chain would be reduced significantly. Thus, an optimum region of composition ratio between rice straws and waste cooking oils can be set within the range from 0% to 50%. The supply chain can use the composition range as a decision variable in order to cope with supply uncertainty. For instance, whenever the rice straws supply decreases due to some reasons such as seasonal supply, the supply chain can add more percentage of used cooking oil into the composition of raw materials, and vice versa.

Fig 7. Raw material logistics and procurement costs at different used cooking oil-rice straws ratio

3.2 Biofuel synthesis plant

In a processing facility the following operations variables for syngas, FT synthesis HC reactor, and distillation towers are suggested.

Table 6. Operations variables of the syngas of oils from 1 mole rice straws (Gasification platform)

Variables	Value
T gasifier (K)	1375
T mix (K)	561.6
CO ₂ in syngas (mol)	0.72
H ₂ in syngas (mol)	0.21
CO in syngas (mol)	0.29
H ₂ O in syngas (mol)	0
CH ₄ in syngas (mol)	0.05

Table 7. Optimum operating condition of Fischer-Tropsch reactor of oils from rice straws (Gasification platform)

Fischer Tropsch Reactor	Calculation	Martin and Grossmann [14]
Reactor outlet	533	493.15
CH ₄ (% mass)	4	0
Parrafin (% mass)	9.9	0
Gasoline (% mass)	26.8	50

Diesel (% mass)	34.8	50
Wax (% mass)	24.5	0
Selectivity to diesel (%)	74	83
H ₂ /CO ratio	1.8	1.7
Probability of chain growth	0.879	0.91

Table 8. Optimum operating condition of hydrocracker of oils from rice straws (Gasification platform)

Hydrocracker reactor	Calculation	Mohanty et al [20]	Plant Data
Hydrogen feed mass flow	0.015	0.015368	0.017815
Reactor outlet	668.15	693	714
Gasoline (% mass)	16.5	16.17	15.8
Diesel (% mass)	62.8	48.7	50.5
Kerosene (% mass)	24.6	30.53	29.4
lighter products (% mass)	1.00E-05	4.51	4

Table 6 lists operating condition required to produce syngas at maximum hydrogen and CO yields. In addition the formation of methane is reduced since the gas does not have any benefit to syngas yield. Similarly Table 7 and 8 exhibit suggested optimum conditions of Fischer Tropsch and hydrocracker that slightly different to literatures [19,25] and plant data. The discrepancies are due to the differences of raw materials (switchgrass, vs rice straws), catalysts, and hydrogen consumptions. Nevertheless, the differences are still within the allowable ranges, for instances FT reactor temperature range (623 – 668 K), and hydrocracker reactor H₂ consumption. The output composition depends on the value of $\bar{\alpha}$, as well as H₂ and CO mole ratio. Therefore, raw material H,C and O composition are the main factors in determining the operating conditions of FT synthesis.

In addition to gasification and used oil hydrocracking platforms, the following fermentation platform is used to produce ethanol as an alternative clean energy source (see Figure 5). Gasification is used to produce syngas that can be converted into ethanol by fermentation [16]. Table 9 shows standard operating conditions to synthesize ethanol by syngas platform. The conditions are simulated by using DWSIM software to obtain technically acceptable outcomes. While ethanol yield is 99,1%, it is sufficient to be directly mixed with gasoline at final depots. Therefore the next section of product upgrading is only implemented for non-ethanol biofuel.

Table 9. Operations variables of the syngas fermentation platform

Variables	Value
T gasifier (K)	1375
T mix (K)	561.6
CO ₂ in syngas (mol)	0.724
H ₂ in syngas (mol)	0.211
CO in syngas (mol)	0.288

H ₂ O in syngas (mol)	0
CH ₄ in syngas (mol)	0.049
CO uptaking rate mmol/gr enzyme.h)	0.376
Ethanol yields mM/h)	1.121
Acetic acid yields (mM/h)	18.814
Clostridium Ljunhdahlia bacteria growth rate (h ⁻¹)	0.87
Ethyl acetate conversion to ethanol	0.991

748

749 **3.3 Product upgrading**

750 In addition to oils synthesis operating conditions, the following product upgrading is
 751 necessary to produce higher purity of biofuel. The data of flash calculation [31] and the
 752 following composition of product blending is obtained.

753

754 Fig 8. Blending composition of refinery outputs

755

Fig 9. Hydrogen allocation (in percentage) of refinery

Table 10. Utility consumptions from alcohol and hydrocarbons platforms

	Ethanol, Kayleen et al. [32]	Hydrocarbons, Wang et al. [33]
Steam (kJ/kg)	6.80E+00	6.49E+02
Electricity (kJ)	4.47E+02	1.07E+02
Process water (gallons/ton rice straws)	3.24E+02	0.00

Table 10 shows that ethanol has less energy demands from steam. However, the platform has also less fuel market share than hydrocarbons.

3.4 Product distribution

From logistics point of view, transportation routings cluster delivery of ethanol across the country according to Table 9 and Table 10 to minimize both transportation costs and carbon emissions. Both results warrant the availability of fuel at minimum inventory levels and transportation costs.

Table 11. Transportation routings from biofuel plant to fuel depots

Transportation planning	1st stop location	2nd stop location	3rd stop location
Transport itinerary 1	1	5	4
Transport itinerary 2	2	3	4
Transport itinerary 3	3	10	4
Transport itinerary 4	3	7	6
Transport itinerary 5	4	6	9
Transport itinerary 6	4	6	10
Transport itinerary 7	4	7	10
Transport itinerary 8	6	5	6
Transport itinerary 9	6	8	7
Transport itinerary 10	7	6	9
Transport itinerary 11	9	5	3
Transport itinerary 12	10	4	2

3.5 Summary of economic and social perspectives

Table 10 shows that the operations benefit to costs ratio of ethanol surpasses hydrocarbons. However, if economic competitiveness is taken into consideration, the result would be the reverse. Moreover, diesel has a competitive advantage over ethanol in terms of environmental benefit. A question on how to manage both products needs an explanation from game theory model that explain firms actions to satisfy sustainability aspects of sustainable production [15].

Let suppose that ethanol needs to be mixed with gasoline with a ratio 0.1. Figure 7 shows that only 27 % of raw materials (from rice straws only) can be converted into ethanol. In addition Table 10 exhibits that gasoline account for 37.6 % of total outputs. By assuming that ethanol is only supplied to biofuel depots, this implies that at maximum 3.76 % of biomass can be converted into ethanol and 96.24 % into hydrocarbons. In this case, ethanol is diesel competitor but a complementary product of gasoline.

While ethanol price is slightly cheaper than diesel but higher than gasoline, it plays a crucial role in determining the selling prices of diesel and gasoline. For an example, the following Cournot game when ethanol plant has a cost advantage over hydrocarbons, where $c_1 < c_2$ and so that firm profits are

$$\pi_1 = aq_1 - bq_1^2 - bq_1q_2 - c_1q_1$$

$$\pi_2 = aq_2 - bq_2^2 - bq_1q_2 - c_2q_2$$

Cournot game is adopted due to the fact that both diesel and ethanol are substitutes. The decision is determined based on the availability of products, rather than price level. Thus by solving the first-order conditions (FOC) of profit maximization simultaneously, then the Cournot equilibrium can be obtained as follows:

$$q_1 = \frac{a - 2c_1 + c_2}{3b}, q_2 = \frac{a - 2c_2 + c_1}{3b}, p = \frac{a + c_1 + c_2}{3}$$

$$\pi_1 = \frac{(a - 2c_1 + c_2)^2}{9b}, \pi_2 = \frac{(a - 2c_2 + c_1)^2}{9b}$$

In economic perspective, the above equations mean that 1) equilibrium price is achieved by considering both products prices, 2) and production levels and profits are determined by operating costs difference between diesel and ethanol.

Considering operating costs, utility consumptions of hydrocarbons is less than ethanol (Table 10). The fuel required for steam and electricity generation is derived mainly from process wastes which include: refinery gas, residual fuel oils (fuel oil, vacuum wastes and asphalt wastes) and FCC coke [33]. This implies that operating costs of diesel is less than ethanol and therefore should be produced more over ethanol. Therefore, the biofuel production might return Indonesia position as an oil exporter.

Another determinant of choosing platform is that carbon emission level. In crude oil refinery the yield of CO₂ is 0.33 kg/kg crude oil [34]. On the other side, CO₂ yield is between 3 Kg CO₂/liter ethanol and 3.24 Kg CO₂/liter ethanol. By considering ethanol density is

kg/m³, the yield is between 3.8 Kg CO₂/Kg ethanol and 4.1 Kg CO₂/Kg ethanol [35]. From logistics cost, in one hand, oceangoing chemical tankers range from 5,000 tonnes deadweight (DWT) to 35,000 DWT in size. On the other hand, crude oil can be shipped at capacity 80,000 to 550,000 DWT. This implies that transport capacity of oil tanker is much larger than ethanol carrier. In conclusion ethanol global shipping might be considered carefully prior to execution due to the above factors and therefore hydrocarbons from biomass is a more attractive option.

The social contribution analysis measures the impact of the biofuel supply chain on the country national development. From social responsibility perspective, the ethanol supply chain contributes to job creation. Solecki et al [36] estimated per gallon basis, a million gallons of production capacity generates 2.24 permanent jobs, 10.29 construction jobs, and nearly 15 indirect jobs and nationally contribute to 0.09 % of GDP to chemical industry sector. On the other hand, crude oil refinery, based on a study of South Africa oil refinery data, employs 0.98 employee/barrel or 31394.7 jobs per million gallons. The information signifies that by taking average salary of employee is 700 US\$ and total income of oil industry, then the supply chain contribution to the country GDP from oil industry is about 0.17 %.

In concluding this section, hydrocarbons from mix materials of used oil and biomass is more promising economically and gives the supply chain a higher level of sustainability.

4. Conclusions

The proposed modeling is capable of providing an alternative solution of unified product and platforms planning within an umbrella of supply chain management. The solution considers economic, environmental and social impacts of the supply chain that brings some implications to management decision making. One of them is that the model guides management to control biofuel production process at a certain level of operational standard and utility consumptions (Tables 6 – 10). The implementation of the control yields product quality and blending (Figure 8) by consuming optimum utilities, which minimize the level of carbon emissions. The operational control also benefits to both supply chain and society by adding more employment that contributes to GDP.

The present contribution highlights the following results. 1) Biofuel platform planning and products distribution give positive impacts on economic growth, carbon emissions reduction and social development of local community. 2) Hydrocarbon biofuel are favored over ethanol in minimizing overall carbon emissions and utility consumptions, 3) The contribution of biofuel supply chain to the country GDP may increase the biofuel supply chain sustainability. In concluding our discussion, we proposes recommendation for future research that might enhance and generalize the results of this article. For instance, the development of decision support tools that embeds the sustainable biofuel supply chain on energy grid in such a way that supports sustainable industry.

In terms of scientific contributions, we develop a new algorithm that can solve the complexity of platform planning and process optimization (see Appendix A). Our simulation

shows that the model can calculate Fischer-Tropsch reactor and hydrocracker optimum operating conditions close enough to the experimental results (Tables 7 and 8). Furthermore, combining process platform and supply chain planning is necessary to consider problem complexity and complicity, which is formulated in such a way that computational tool can address the challenge. In conclusion, our generic model could be a starting point to unify product, process and supply chain design for biofuels.

Further research may investigate a number of remaining issues. The supply chain should involve biomass supply networks to mitigate supply disruptions and biofuel price fluctuations. In addition, the extension to global operations of biofuel supply chain may be useful to be investigated as well.

Appendix A: Mixed integer nonlinear programming (MINLP) model

The solution method can be summarized by following model of mixed integer nonlinear programming (MINLP):

$$\max_{x,y} f(x) + f(y) + f(x,y) \quad (A1)$$

$$\text{s.t. } g(x) + g(y) + g(x,y) \leq 0 \quad (A2)$$

$$h(x) + h(y) + h(x,y) = 0 \quad (A3)$$

The objective function (A1) represents either profit maximization or cost minimization. For most of cases, inequality constraint (A2) is used to represent quality bounds (lower or upper) and equality constraint (A2) for material balance along the process flows. In each objective (1a) and constraints (A2,3), the first two terms stand for integer variables $f(y), g(y), h(y)$ (i.e., choice of machines, production sequence, etc) and continuous variables $f(x), g(x), h(x)$ (i.e., flow rates, temperatures, pressures, mixing ratio, etc). The last terms, $f(x,y), g(x,y), h(x,y)$ are complicating variables and may be correlated and therefore non-convex. There are many efficient methods for solving the first two terms but not the non-convex and correlated terms. The following subsections propose solution strategies for handling non-convex variables and solving the formulation.

2.3.1 Model of variables $f(x,y), g(x,y), h(x,y)$ convexification

The nonconvex variables can exists in forms of polynomial and/or signomial variables in the objective functions and constraints.

While the polynomial and/or signomial variables is exists in Eq.(A1), it should be transformed according to the results of convex analysis. The nonlinear transformation constraints for the geometric variables are first rewritten based on formulation in [37]:

Property 1: the system of $ae^{(r_1x_1 + \dots + r_nx_n)}$ is convex for R_n^+ , $a \geq 0$, and $r_n \in R$

Property 2: the system of $\frac{ae^{(r_1x_1 + \dots + r_nx_n)}}{x_1^{s_1}x_2^{s_2}\dots x_n^{s_n}}$ is convex for R_n^+ , $a, s \geq 0$, and $r_n \in R$

Property 3: the system of $ax_1^{s_1}x_2^{s_2}\dots x_n^{s_n}$ is convex for R_n^+ , $a \leq 0, s \geq 0$, and $\sum_n s_n \leq 1$

Proofs: $ax_1^{s_1}x_2^{s_2}\dots x_n^{s_n}$ can be rewritten as $ae^{(s_1\ln x_1) + \dots + (s_n\ln x_n)}$. The $\ln$ function is concave

on R_n^+ . It is therefore the value of a must be negative to change it to convex function (Property 3). While any exponential function is always convex and increasing on R (Property 1), the division of two convex functions is another convex function (Property 2).

Based on three properties of signomial terms [37], a signomial part of $Z_{ij} = x_i^a x_j^b$ requires variables transformation to replace each non-convex terms from x_i^a to X_i^1 and from x_j^b to X_j^1 in the objective function and constraint in with a convex underestimation of it and then finding the solution of the convex relaxation (Properties 4 and 5) based on formulation in [37]. The relaxation will not interfere convex part of the objective function and constraints in Eqs.(2a,2b). Therefore the gaps between original and relaxed problems cannot exceed the allowable error ε (Property 6).

Property 4: $h^{CONV}(x_i, Z)$, $g^{CONV}(x_i, Z)$ be convex $\forall x \in \{x^L, x^U\}$.

Property 5: $h^{CONV}(x_i, Z) \leq h(x_i)$, $g^{CONV}(x_i, Z) \leq g(x_i)$

Property 6: $\forall x \in \{x^L, x^U\}$ for $\varepsilon \geq 0$, $\exists \{x^L, x^U\} \subseteq \{x^L, x^U\}$ with $\delta(\varepsilon) = \|x^u - x^l\|_2^2 > 0$ such that $h^{CONV}(x_i, Z) - h(x_i) \leq \varepsilon$.

The underestimation of original problem (1) by transformation according to properties 1 to 3 is followed by property 4 to 6 to get piecewise linear approximation of the original problem. The discretization needs variable bounds as follows:

$$e^{\sum_{i \in I} r_{ij} \ln(x_i)^L} \leq e^{\sum_{i \in I} r_{ij} \ln(x_i)} \leq e^{\sum_{i \in I} r_{ij} \ln(x_i)^U} \quad (A4)$$

While $e^{\sum_{i \in I} r_{ij} \ln(x_i)}$ is nonlinear, the underestimation of $e^{\sum_{i \in I} r_{ij} \ln(x_i)}$ can be formulated by replacing $\sum_{i \in I} r_{ij} \ln(x_i) = Y_j$ and $e^{Y_j} = Z_j$ and finding linear interpolations (LI) of $e^{a \ln(x_i)}$, as follows:

$$e^{Y_j} \geq Y_j^L + \frac{Y_j - Y_j^L}{Y_j^U - Y_j^L} (e^{Y_j^U} - e^{Y_j^L}) \quad (A5)$$

The right hand side (RHS) is the under-estimator and linear relaxation $Rel(e^{\sum_{i \in I} r_{ij} \ln(x_i)})$ of the original signomial equation $e^{\sum_{i \in I} r_{ij} \ln(x_i)}$, which has difference value to its original. The difference between left hand side (LHS) and RHS of Eq.(LI-1) reaches maximum at $Y_j = \sum_{i \in I} r_{ij} \ln(x_i)^{diff^{max}}$, where the maximum distance between RHS and LHS of Eq. LI-1 can be formulated as follows:

$$diff_i^{max} = e^{Y_j} - Y_j^L - \frac{Y_j - Y_j^L}{Y_j^U - Y_j^L} (e^{Y_j^U} - e^{Y_j^L}) \quad (A6)$$

While Y_j^L and $e^{Y_j^L}$ are known, Y_j^U and $e^{Y_j^U}$ can be obtained iteratively by setting $diff_i^{max}$ as the allowable error between original and relaxed solution.

The values of $Y_j^U - Y_j^L$ is equal to incremental steps $s = 1$ to S are employed to get local upper bound at step s , $Y_{ijs}^U = \sum_{i \in I} r_{ij} \ln(x_{is})^U$. Each step s has length $\Delta_{Bs} = \sum_{i \in I} r_{ij} \ln(x_{is})^U - \sum_{i \in I} r_{ij} \ln(x_{is})^L$ and at the next iterations $s+1$, the upper bound at step s to be the lower bound, $Y_{ijs}^U = Y_{ijs(s+1)}^L$. Eq.(4) can be reformulated as follows:

$$diff_{is}^{max} = e^{Y_{js}} - Y_{js}^L - \frac{Y_{js} - Y_{js}^L}{Y_{js}^U - Y_{js}^L} (e^{Y_{js}^U} - e^{Y_{js}^L}) \quad (A7)$$

The discretization steps can be summarized as follows:

1. Set the allowance $diff_{is1}^{max}$ and find Δ_{Bs} and $e^{Y_{js}^U}$ by solving

$$diff_{is}^{max} = e^{Y_{js}} - Y_{js}^L - \frac{Y_{js} - Y_{js}^L}{Y_{js}^U - Y_{js}^L} (e^{Y_{js}^U} - e^{Y_{js}^L})$$

2. If $Y_{js}^L \geq Y_j^L$, use Y_{js}^L as the global lower bound and the relaxed solution is equal to the original solution.
3. Solve Eq.(5) and find $e^{Y_{js}^U}$
4. For discretization of Y_j , generate S discrete value of Y_{js} , in order to get $\Delta_{Bs} = Y_{js}^U - Y_{js}^L$
5. Continue the process until $\sum_{s \in S} \Delta_{Bs} = Y_j^U - Y_j^L$

Considering the complicating variables $f(x,y), g(x,y), h(x,y)$, those terms in problem (1) can be reformulated as integer programs (IP) as follows:

$$\min \sum_{j \in J} \left(\sum_{i \in I} e^{r_{ij} \ln(x_i)^U} - \sum_{i \in I} e^{r_{ij} \ln(x_i)^L} \right) y_{si} \quad (A8)$$

$$\text{s.t. } y_{si} \geq y_{(s+1)i} \quad (A9)$$

$$\sum_s \left(\sum_{i \in I} e^{r_{ij} \ln(x_i)^U} - \sum_{i \in I} e^{r_{ij} \ln(x_i)^L} \right) y_{si} = \sum_{i \in I} r_{ij} \ln x_{ip} \quad (A10)$$

$$\ln x_p = \Delta_1 \theta_1 + \dots + \Delta_p \theta_p \quad (A11)$$

$$y_s \in \{0,1\}, \theta_p \in \{0,1\} \quad (A12)$$

Fig A1. A construction of piecewise linear approximation based on $diff_i^{\max}$

Three incremental variables from pre-processing steps are used as decision parameters. If incremental step $s+1$ is chosen to be positive ($y_{(s+1)i} = 1$), incremental steps s and the following steps before s are also chosen to be positive ($y_{si}, y_{(s-1)i}, \dots, y_{s1} = 1$) (Eq.A9).

References

- [1] Food and Agriculture Organization of the United Nations. (2017). Production. Rice market monitor, 20(1). 1-41.
- [2] Statistics Indonesia (BPS). (2018). Production of Paddy by Province (ton), 1993-2015

(Indonesia).

- [3] C.L. Illsley. (2018). Where Are Indonesia's Palm Oil Plantations Located?, Economics. Downloaded from <https://www.worldatlas.com/articles/where-are-indonesia-s-palm-oil-plantations-located.html> on 6 January 2019.
- [4] Suwannarangsee, S., Bunternngsook, B., Arnthong, J., Paemane, A., Thamchaipenet, A., Eurwilaichitr, L., ... & Champreda, V. (2012). Optimisation of synergistic biomass-degrading enzyme systems for efficient rice straw hydrolysis using an experimental mixture design. *Bioresource technology*, 119, 252-261.
- [5] Lahijani, P., & Zainal, Z. A. (2011). Gasification of palm empty fruit bunch in a bubbling fluidized bed: a performance and agglomeration study. *Bioresource technology*, 102(2), 2068-2076.
- [6] Hill, J., Polasky, S., Nelson, E., Tilman, D., Huo, H., Ludwig, L., ... & Bonta, D. (2009). Climate change and health costs of air emissions from biofuels and gasoline. *Proceedings of the National Academy of Sciences*, 106(6), 2077-2082.
- [7] de Klerk, A. (2011). Fischer-Tropsch fuels refinery design. *Energy & Environmental Science*, 4(4), 1177-1205.
- [8] Swanson, R. M., Platon, A., Satrio, J. A., & Brown, R. C. (2010). Techno-economic analysis of biomass-to-liquids production based on gasification. *Fuel*, 89, S11-S19.
- [9] You, F., & Wang, B. (2011). Life cycle optimization of biomass-to-liquid supply chains with distributed-centralized processing networks. *Industrial & Engineering Chemistry Research*, 50(17), 10102-10127.
- [10] Zhang, F., Johnson, D. M., & Johnson, M. A. (2012). Development of a simulation model of biomass supply chain for biofuel production. *Renewable Energy*, 44, 380-391.
- [11] Zhang, F., Johnson, D., Johnson, M., Watkins, D., Froese, R., & Wang, J. (2016). Decision support system integrating GIS with simulation and optimisation for a biofuel supply chain. *Renewable Energy*, 85, 740-748.
- [12] Seabra, J. E., & Macedo, I. C. (2011). Comparative analysis for power generation and ethanol production from sugarcane residual biomass in Brazil. *Energy Policy*, 39(1), 421-428.
- [13] Perales, A. V., Valle, C. R., Ollero, P., & Gómez-Barea, A. (2011). Technoeconomic assessment of ethanol production via thermochemical conversion of biomass by entrained flow gasification. *Energy*, 36(7), 4097-4108.
- [14] Van der Heijden, H., & Ptasiński, K. J. (2012). Exergy analysis of thermochemical ethanol production via biomass gasification and catalytic synthesis. *Energy*, 46(1), 200-210.
- [15] Tseng, M. L., Tan, R. R., & Siriban-Manalang, A. B. (2013). Sustainable consumption and production for Asia: sustainability through green design and practice. *Journal of*

Cleaner Production, 40, 1-5.

- [16] Kristianto, Y., & Zhu, L. (2017). Techno-economic optimization of ethanol synthesis from rice-straw supply chains. *Energy*, 141, 2164-2176.
- [17] Antczak, M. S., Kubiak, A., Antczak, T., & Bielecki, S. (2009). Enzymatic biodiesel synthesis—key factors affecting efficiency of the process. *Renewable energy*, 34(5), 1185-1194.
- [18] Bezergianni, S., Kalogianni, A., & Vasalos, I. A. (2009). Hydrocracking of vacuum gas oil-vegetable oil mixtures for biofuels production. *Bioresource Technology*, 100(12), 3036-3042.
- [19] Martín, M., & Grossmann, I. E. (2011). Process optimization of FT-diesel production from lignocellulosic switchgrass. *Industrial & Engineering Chemistry Research*, 50(23), 13485-13499.
- [20] Mahishi, M. R., & Goswami, D. Y. (2007). An experimental study of hydrogen production by gasification of biomass in the presence of a CO₂ sorbent. *International Journal of Hydrogen Energy*, 32(14), 2803-2808.
- [21] Fernandes, F. A., & Teles, U. M. (2007). Modeling and optimization of Fischer–Tropsch products hydrocracking. *Fuel Processing Technology*, 88(2), 207-214.
- [22] Rajesh, J. K., Gupta, S. K., Rangaiah, G. P., & Ray, A. K. (2001). Multi-objective optimization of industrial hydrogen plants. *Chemical Engineering Science*, 56(3), 999-1010.
- [23] Visconti, C. G., & Mascellaro, M. (2013). Calculating the product yields and the vapor–liquid equilibrium in the low-temperature Fischer–Tropsch synthesis. *Catalysis today*, 214, 61-73.
- [24] Lee, T. S., & Chung, J. N. (2012). Mathematical modeling and numerical simulation of a Fischer–Tropsch packed bed reactor and its thermal management for liquid hydrocarbon fuel production using biomass syngas. *Energy & Fuels*, 26(2), 1363-1379.
- [25] Mohanty, S., Kunzru, D., & Saraf, D. N. (1990). Hydrocracking: a review. *Fuel*, 69(12), 1467-1473.
- [26] Kendrick, D., Meeraus, A., & Suh, J. S. (1981). Oil refinery modeling with the gams language (No. UT/CES-RR-14). Texas Univ., Austin (USA). Center for Energy Studies.
- [27] Lambert, G. J., Schoeber, W. J. A. H., & Van Helden, H. J. A. (1994, April). The hydrogen balance in refineries. In *Foster Wheeler Heavy Oil Processing and Hydrogen Conference*, Noordwijk, The Netherlands.
- [28] Kristianto, Y., & Gunasekaran, A. (2018). A global optimization for sustainable multi-domain global manufacturing. *Computers & Operations Research*. Vol. 89, 307-323.
- [29] Samuel, V. (2013). Environmental and socioeconomic assessment of rice straw conversion to ethanol in Indonesia: the case of Bali.

- [30] Simatupang, T., & Pardhana, H.W. (2011). Improving Inventory Management and Supply Chain of Diesel Fuel in Pertamina 5 Main Depots. *The Asian Journal of Technology Management* Vol. 3 No. 1 (2010) 31-43.
- [31] Singh, B. B., Bi-linear Data Reconciliation and Rigorous Optimisation of Refinery Hydrogen Networks with Hydrogen Plant. Process Integration Research Consortium, Manchester, 2006.
- [32] Kaylen, M., Van Dyne, D. L., Choi, Y. S., & Blasé, M. (2000). Economic feasibility of producing ethanol from lignocellulosic feedstocks. *Bioresource Technology*, 72(1), 19-32.
- [33] Wang, M., Lee, H., & Molburg, J. (2004). Allocation of energy use in petroleum refineries to petroleum products. *The International Journal of Life Cycle Assessment*, 9(1), 34-44.
- [34] Szklo, A., & Schaeffer, R. (2007). Fuel specification, energy consumption and CO₂ emission in oil refineries. *Energy*, 32(7), 1075-1092.
- [35] Karras, G. (2011). Oil refinery CO₂ performance measurement. CBE-UCS final report.
- [36] Solecki, M., Dougherty, A., & Epstein, B. (2012). Advanced Biofuel Market Report 2012, Meeting US Fuel Standards.
- [37] Pörn, R., Björk, K. M., & Westerlund, T. (2008). Global solution of optimization problems with signomial parts. *Discrete optimization*, 5(1), 108-120.

Highlights

1. Biofuels platform planning and optimization was developed.
2. A case study of biofuels supply chain under various production paths was conducted.
3. The combined raw materials increase the supply flexibility and supply chain responsiveness.
4. Hydrocarbon biofuels are favored over ethanol in minimizing overall carbon emissions.