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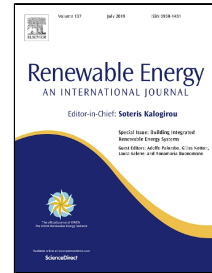
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## Platforms planning and process optimization for biofuels supply chain

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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.

43 First, even though Indonesia is a net paddy importer during the last 10 years, the country  
44 always increases its paddy production by average 3.5% per year [1]. As a result, the  
45 increased capacity of paddy production also increases raw material supply for biofuel.  
46 Second, unlike oil palm industry that is mostly situated in Sumatra (65%) and Kalimantan  
47 (26%), rice fields in Indonesia spread across the archipelago, from Sumatra, Java, and  
48 Sulawesi. Rice production data from Statistics Central Bureau (BPS) shows that rice  
49 plantations are available in 34 provinces, which is more beneficial in distributing the  
50 biomass and biofuel and therefore lower supply chain costs [2]. Third, from environmental  
51 issues, in places such as Sumatra, in order to enable palm oil plantations, a hectare of  
52 rainforest is destroyed and 174 tons of carbon emissions are released into the atmosphere,  
53 which destructs the ecosystems and violates our main objective of protecting the nature [3].  
54 Fourth, fermentation yield of oil palm waste such as empty fruit bunches is less than that of  
55 rice straws [4]. In addition, Lahijani and Zainal [5] shows that  $H_2/CO$  ratio of empty fruit  
56 bunches gasification is between 0.4 and 0.6, which is less than rice straws gasification (>  
57 90%), which makes rice straws more appropriate than oil palm waste to produce liquid  
58 biofuel. All of those reasons support our decision to use rice straws as a potential raw  
59 materials of biofuel.

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

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

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

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

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

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

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

127

128 **2. Research methods**

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

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

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

148 The given information includes

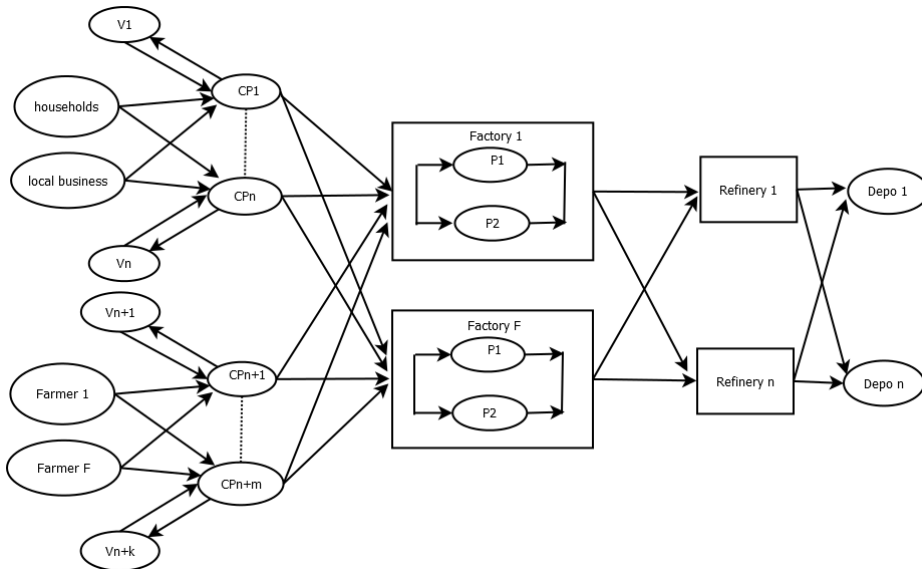
- 149 1. Raw materials locations and availability (mass and types)
- 150 2. Biofuel (and other byproducts) demand, specifications, and prices
- 151 3. Utility and other raw materials (e.g., hydrogen, acid, water, etc.) availability and  
152 prices
- 153 4. Delivery time windows of transporter, volumes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

167 Determine:

- 168 1. A detailed routings schedule for each transporters

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

173



174

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

176

177 Figure 1 shows how the supply chains works. Households, local food business, and farmers  
 178  $f \in \{1, \dots, F\}$  are suppliers of biomass materials. A set of transporters collect the biomass from  
 179 each collecting point  $CP_n$   $n \in \{1, \dots, N\}$  for farmland biomass,  $m \in \{1, \dots, M\}$ , and distributed  
 180 them to several available processing factories  $F_F$  that has several technology options  $p \in$   
 181  $\{1, \dots, P\}$ , before finally sent to depots  $n \in \{1, \dots, D\}$ .

182

## 183 2.1 Goal and scope definition

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

191

### 192 2.1.1 Technology options of biodiesel synthesis

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



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

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

206

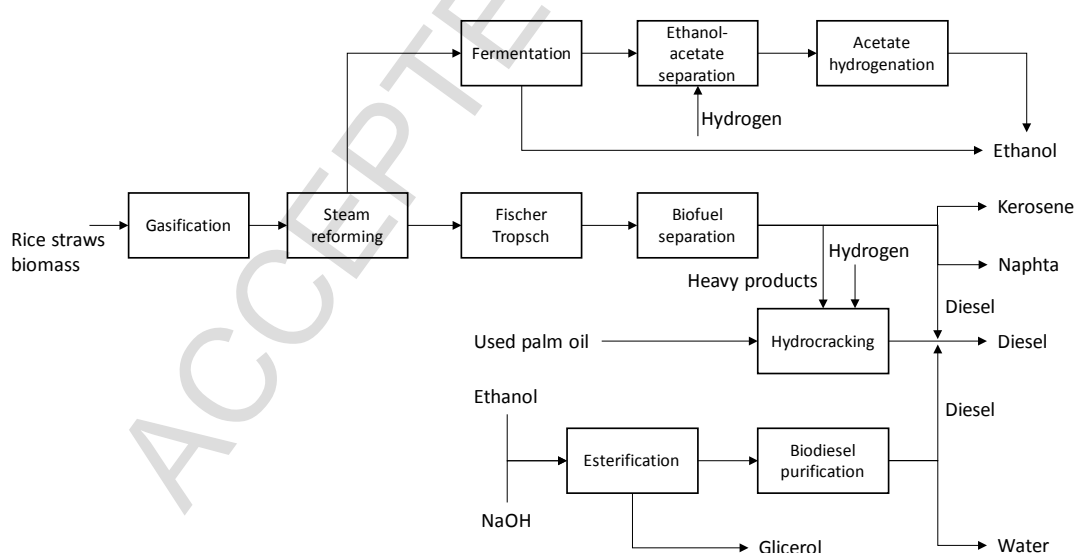
## 207 2.1.2 Technology description of hydrocracking and Fischer Tropsch of biodiesel

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

214 The second and third paths convert biomass and waste palm separately. If the supply chain  
215 chooses both paths, heavy products from FT process enter (sulfide  $\text{NiMo}/\text{Al}_2\text{O}_3$ ) at based  
216 catalyst hydrocracker at 237-357 °C [21]. Raw vegetable oil enters hydrocracker at 370 –  
217 390 °C over conventional hydrotreating catalysts (sulfide  $\text{NiMo}/\text{Al}_2\text{O}_3$ ), and has operating  
218 temperature 350 – 390 °C, depending on the products [18]. Therefore, each feedstock (rice  
219 straws and used palm oil) has its own hydrocracking facility, and thus they are not mixed  
220 during the conversion process.

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

225



226

227

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

230

## 231 2.2 Biofuel supply chain model

232

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

235

### 236 2.2.1 Raw materials collection

237 Vehicle routing to pick up the raw materials from sources  $i$  to destination  $j$ . Transportation  
238 costs are constrained by using a set of equations (1-7). Such a route from  $i$  to  $j$  can be only  
239 passed once for each trucks of total available  $m$  trucks from a biofuel factory (1,2). Eqs. (3)  
240 constraints the amount of BET transported by each ship from depot  $k$  to  $k'$ , after observing  
241 the demands at source  $DEM_k$  and destination  $DEM_{k'}$  as well as ship capacity  $CAP$ . Eq. (4)  
242 does not permit a vehicle backs to its original source without finishing the trip. Therefore at  
243 minimum  $v(S)$  number of transporters must be used to visit all vertices  $S$  except the source  
244 vertices  $S \subset V' \setminus \{1\}$ . Thus the trip is started by visiting vertices other than source  $|S| \geq 2$   
245 (Eq. 5). The availability of transporter depends on the availability of production  $Y_{it}$  (Eqs.  
246 6,7). The objective is to minimize total distance  $Tot\_col\_cost_{ij}$  as accumulation of individual  
247 collecting cost,  $ColCost_{ij}$  at a distance  $Dist_{ij}$  (8) from  $i$  to  $j$ , where  $i,j$  are members of  
248 vertice  $V$ .

$$249 \sum_{i=1} x_{jk} = m \quad (1)$$

$$250 \sum_{j=1} x_{kj} = m \quad (2)$$

$$251 Q_j = q_{jk} + \sum_k q_{kk'} \quad (3)$$

$$252 q_{kk'} = (q_{jk} + DEM_{k'} - CAP) + CAPx_{kk'} + (CAP - DEM_k + DEM_{k'})x_{k'k} \quad (4)$$

$$253 \sum_{i,j \in S} x_{ij} \leq |S| - v(S) \quad S \subset V' \setminus \{1\}; |S| \geq 2, \quad (5)$$

$$254 x_{ij} \in \{0,1\} \quad i,j = 1,\dots,n; i \neq j \quad (3)$$

$$255 x_{ij} \leq Y_{it} \quad (7)$$

$$256 Tot\_col\_cost_{ij} = \sum_{i,j \in V} Dist_{ij} ColCost_{ij} x_{ij} \quad (8)$$

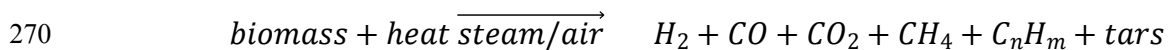
257

### 258 2.2.2 Biofuel factory decision

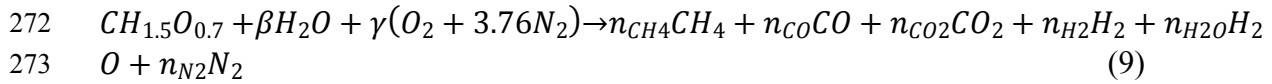
259 There are several options of biofuel process synthesis from biomass (Figure 2). The  
260 pathways include several stages of pre-treatment, that is followed by a sequence of Fischer  
261 Tropsch process and hydrocracking (FT process) and or hydrocracking process (HC). This  
262 article focuses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

#### 263 A. Syngas production for FT synthesis

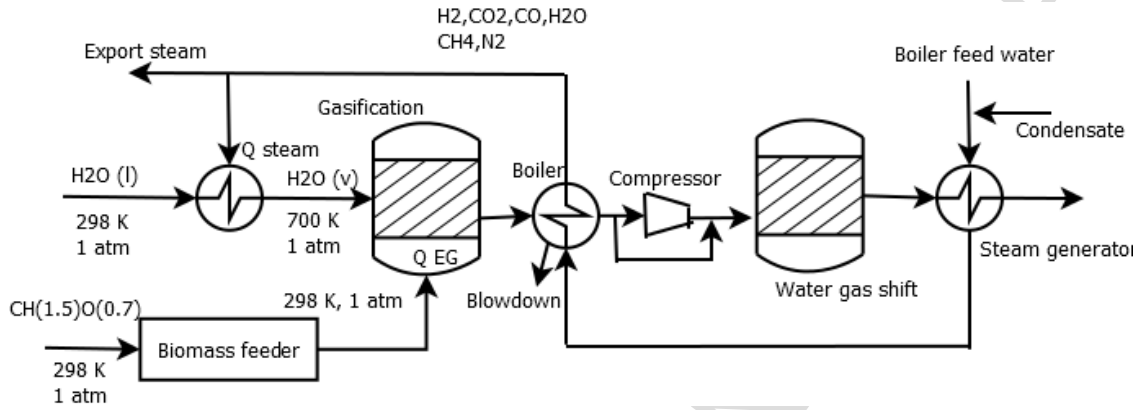
264 For biodiesel production from rice straws, syngas is the main component of biofuel that is  
265 obtained from heating the biomass with hot streams of steam and air. The process includes  
266 pretreatment that reduce the size of biomass to be easier to transport and treat in a  
267 gasification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



271 In specific biomass can be formulated as follows:



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



279

280 Fig 3. Syngas from biomass production

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

286 While some of boiler feed water is used in steam reforming, 5% blowdown is suggested to  
 287 keep the content of salt in the water not exceed the allowable limit (Eq.12). In addition, a  
 288 ratio of 2:1 for steam-methane inside WGS is considered to eliminate carbon deposit on  
 289 catalyst and therefore boiler feed water mass can be determined according to Eq. (13). Figure  
 290 3 of Mahishi and Goswani [21] is used to compose Eqs.(14-18). Eq. (19) calculates water  
 291 fraction in the gasification outlet. Considering steam separator efficiency for process and  
 292 reformer,  $J_{SS}$ , condensate mass flows  $F_{cond}$  must equal to the water content of WGS stream  
 293 (Eq.20). The outputs of WGS,  $F_{WGS,out}$ , will be the input of solid separator,  
 294  $F_{Solid\ separator,in}$  (Eq.21). Steam separation efficiency  $\epsilon_{SS}$  depends on separator temperature  
 295  $T_{SS}$  and its steam partial pressure  $P_{SS}y_{H_2O,WGSout}$  (Eq.22). Mixing temperature between  
 296 BFW and condensate,  $T_{mix}$ , is approximated from BFW temperature at 383 °C and  
 297 condensate temperature  $T_{SS}$  (Eq.23). Eqs.(24-26) bound operating temperatures in  
 298 industrial units [22].

$$299 \quad (F_{BFW} + F_{cond})Cp_{BFW}(T_{int} - T_{mix}) = Q_{preheater} \quad (10)$$

$$300 \quad Q_{preheater} = F_{WGS,out}C_p(T_{WGSout} - T_{SS}) \quad (11)$$

$$301 \quad F_{BFW} = 1.05(F_{stm} + 2 \times 18(n_{CH_4,WGSin} + \beta_{WGSin})) \quad (12)$$

$$302 \quad n_{H_2O} = \beta_{H_2O} - n_{CO} \quad (13)$$

$$303 \quad n_{N_2} = 0.79\gamma_{O_2} + 3.76N_2 \quad (14)$$

$$304 \quad n_{CO} = 0.75 - \beta_{H_2O}(0.4/0.25) \quad (15)$$

$$305 \quad n_{H_2} = 0.3734 - 2.303 \log(\beta_{H_2O}) + 1.2794 \quad (16)$$

$$306 \quad n_{CO_2} = 0.17 - \beta_{H_2O}(0.65 - 0.17/0.25) \quad (17)$$

$$307 \quad n_{CH_4} = 0.0707 - \exp(-1.281\beta_{H_2O}) \quad (18)$$

$$308 \quad y_{H_2O,WGSout} = \frac{n_{H_2O}}{n_{H_2O} + n_{N_2} + n_{CO} + n_{H_2} + n_{CO_2} + n_{CH_4}} \quad (19)$$

$$309 \quad F_{cond} = F_{WGS,out}(y_{H_2O,WGSout}(1 - \varepsilon_{SS})) \quad (20)$$

$$310 \quad F_{SS,out} = F_{Solid\ separator,in} = F_{WGS,out}(1 - (1 - \varepsilon_{SS}))y_{H_2O,WGSout} \quad (21)$$

$$311 \quad \varepsilon_{SS} = \left(\frac{T_{SS} - 273}{100}\right)^4 \frac{101.325}{P_{SS}y_{H_2O,WGSout}} \quad (22)$$

$$312 \quad T_{mix} = \frac{383F_{BFW} + F_{cond}T_{SS}}{F_{BFW} + F_{cond}} \quad (23)$$

$$313 \quad 400\text{ K} \leq T_{WGSin} \leq 530\text{ T} \quad (24)$$

$$314 \quad 1375\text{ K} \leq T_{WGSout} \leq 1675\text{ T} \quad (25)$$

$$315 \quad 298\text{ K} \leq T_{SS} \leq 400\text{ T} \quad (26)$$

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

320 Steam reforming (SMR):



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

$$323 \quad (F_{BFW} + F_{cond})Cp_{BFW}(516 - T_{int}) + \left(\frac{F_{BFW}}{1.05} + F_{cond}\right)\lambda_{BFW} = Q_{Boiler} \quad (28)$$

$$324 \quad Q_{Boiler} = F_{GAS,out}C_p(T_{GASout} - T_{WGS,in}) \quad (29)$$

325 Since the gas stream contains CO<sub>2</sub> and carbon, the stream is cleaned by using three different  
326 ways, hot cleaning, cold cleaning or partial swing absorber (PSA). In our simulation model,  
327 hot cleaning at 94 °C is used to remove also suspended solid.

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

$$329 \quad Syngas\_cost_i = \left(\frac{1}{Q_{Boiler}} + \frac{1}{n_{CO}} + \frac{1}{n_{H_2}_i}\right)$$

330

### 331 **B. Syngas fermentation**

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

334

### 335 **C. Fischer Tropsch model**

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

$$339 \quad -r_{CO} = 0.01 \times T - 2.01 \quad (30)$$

$$340 \quad \alpha_n = \begin{cases} 0.292 & n = 1 \\ -0.0317n + 1.0362 & 2 \leq n \leq 7 \\ 0.8 & n \geq 8 \end{cases}$$

341

$$342 \quad -r_{CO} = \frac{r_1}{1 - \alpha_1} \left( \sum_{i=1}^N \left( \prod_{j=1}^i \alpha_j \right) \right) \quad (31)$$

$$343 \quad r_1 = \frac{0.001059 P_{CO}^{-0.86} P_{H_2}^{1.32}}{\left( 1 + 0.46 P_{H_2} / P_{CO} \right)}$$

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

$$347 \quad -r_{FT} = \frac{k C_{H_2} C_{CO}}{(1 + K C_{CO})^2} \quad (32)$$

$$348 \quad k = 0.4 \exp \left( \frac{-37400}{RT} \right) \left[ \frac{m^6}{kg_{cat} mol s} \right] \quad (33)$$

$$349 \quad K = 5.10^{-9} \exp \left( \frac{68500}{RT} \right) \left[ \frac{m^3}{mol} \right] \quad (34)$$

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

$$353 \quad m_n = (1 - \bar{\alpha}) \bar{\alpha}^{n-1} \quad (35)$$

$$354 \quad \text{For } \bar{\alpha} = \left( 0.233 \left( \frac{y_{CO}}{y_{CO} + y_{H_2}} \right) + 0.633 \right) (1 - 0.0039(T - 533))$$

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

$$356 \quad m_{CH_4} = (1 - \bar{\alpha}) \quad (\text{Methane})$$

$$357 \quad m_{C_2 - C_4} = \sum_{n=2}^4 (1 - \bar{\alpha}) \bar{\alpha}^{n-1} \quad (\text{paraffin's})$$

$$358 \quad m_{C_5 - C_{10}} = \sum_{n=5}^{10} (1 - \bar{\alpha}) \bar{\alpha}^{n-1} \quad (\text{Gasoline})$$

$$359 \quad m_{C_{11} - C_{17}} = \sum_{n=11}^{17} (1 - \bar{\alpha}) \bar{\alpha}^{n-1} \quad (\text{Kerosene})$$

$$360 \quad m_{C_{18} - C_{24}} = \sum_{n=18}^{24} (1 - \bar{\alpha}) \bar{\alpha}^{n-1} \quad (\text{Diesel})$$

$$361 \quad m_{C_{25} - C_{34}} = \sum_{n=25}^{34} (1 - \bar{\alpha}) \bar{\alpha}^{n-1} \quad (\text{Wax})$$

$$362 \quad (36)$$

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

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

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

367

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

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

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

$$375 \text{ Conversion}(\%) = 1 / ((-15.99418699 \times \ln T_{hydrocracker}) + 96.69166996) \quad (37)$$

$$376 \text{ Diesel}_{select}(\%) = -5.333545685 \cdot 10^{-3} T_{diesel} + 2.840289671 \quad (38)$$

$$377 \text{ Kerosene}_{select}(\%) = 16877.53081 e^{(-4196.195605 / T_{kerosene})} \quad (39)$$

$$378 \text{ Naphta}_{select}(\%) = 5173904.255 e^{(-6432.000463 / T_{Naphta})} \quad (40)$$

379 We use experimental data of Bezergianni et al. [18] for the conversion and selectivity of  
380 used cooking oil into diesel as a function of the temperature. The authors did a  
381 hydrocracking experiment that were performed at  $P = 2000$  psig (13789.5 kPa),  $350 \text{ K} \leq$   
382  $T_{reaktor} \leq 390 \text{ K}$ ,  $LHSV = 1.5 \text{ h}^{-1}$  and  $H_2/\text{oil ratio} = 6000 \text{ scfb}$  ( $1068 \text{ nm}^3/\text{m}^3$ ). From the  
383 experiment data regression analysis, this article formulates products conversion from rice  
384 straws material as follows:

$$385 \text{ Conversion}(\%) = 1.795596872 \cdot 10^{-1} \ln(T_{hydrocracker} - 293.178567) \quad (41)$$

$$386 \text{ Diesel}_{select}(\%) = 3.023415023 \cdot 10^{-1} e^{398.7415509 / T_{hydrocracker}} \quad (42)$$

$$387 \text{ Kerosene}_{select}(\%) = -472.7896012 / T_{hydrocracker} + 1.402337459 \quad (43)$$

$$388 \text{ Naphta}_{select}(\%) = -218.0836957 / T_{hydrocracker} + 0.6796343336 \quad (44)$$

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

$$393 m_i = \text{Conversion}(\%)_i \times i_{select}(\%) \sum_{i \in HC} M_{HC} \quad (45)$$

$$394 M_{HC} = n_{CO} \times MW_{CO} + \left( \frac{m}{2} + n \right) H_2 \times MW_{H_2} \quad (46)$$

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

$$397 m_i = \sum_{ricestraws, cooking\ oil \in RM} (\text{Conversion}(\%)_i \times i_{select}(\%) \sum_{i \in HC} M_{HC}) RM\%$$

$$398 M_{HC} = \sum_{ricestraws, cooking\ oil \in RM} \left( n_{CO} \times MW_{CO} + \left( \frac{m}{2} + n \right) H_2 \times MW_{H_2} \right) RM\%$$

399

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

401 1.9 is implemented to determine hydrogen consumption. Mohanty et al. [25] formulated H<sub>2</sub>  
 402 consumption rate per unit mass of component j to be cracked into lighter component i as  
 403 follows:

$$404 \quad (H_2/HC)_j = \frac{\sum_{i=1}^{i=j-2} \frac{m_j}{1+(C/H\ ratio)_i} - \frac{1}{(C/H\ ratio)_j} \sum_{i=1}^{i=j-2} \frac{m_j (C/H\ ratio)_i}{1+(C/H\ ratio)_i}}{\sum_{i=1}^{i=j-2} \frac{m_j (C/H\ ratio)_i}{1+(C/H\ ratio)_i} \left(1 + \left(\frac{1}{(C/H\ ratio)_j}\right)\right)} j = \text{kerosene, diesel, FT output}$$

405 (47)

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

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

$$414 \quad \sum_{j \in J} \Delta H_{Rj} = \Delta H_{Rj}^0 + \left( \sum_{i=1}^{j-2} \Delta H_i m_i - \Delta H_j (m_C + m_H) - \Delta H_{H_2} (H_2/HC)_j \right) \quad (48)$$

$$415 \quad (m_C)_i = \text{Conversion}(\%)_i \times i_{select}(\%) \frac{C_i}{C_i + 1}$$

$$416 \quad (m_H)_i = \text{Conversion}(\%)_i \times i_{select}(\%) \frac{1}{C_i + 1}$$

417 Hydrogen enthalpy at standard condition (25 C and 1 atm) and reaction temperature  $\Delta H_{H_2}$   
 418  $= H_{H_2}^T - H_{H_2}^0$  are calculated from the available heat capacity data and regression data of  
 419 heat capacity as a function of temperature. Heavier and lighter products enthalpy at reaction  
 420 temperature and pressure is calculated from Peng-Robinson equation of state (EOS) and Lee  
 421 and Kassler Table. Standard enthalpy is calculated from heat capacity data The specific heat  
 422 of kerosene is 2.01 kJ/kg.K, diesel is 1.9 kJ/kg.K, naphta is assumed equal to gasoline at  
 423 2.1 kJ/kg.K, and hydrogen is 8.7 kJ/kg.K.

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

$$425 \quad dT/dW = - \sum_{j \in J} \frac{m_j}{\sum_{j \in J} m_j} \Delta H_{Rj} k_{Rj} / \sum_{i \in J-2} m_i C_{pi} \quad (49)$$

$$426 \quad k_{Rj} = 10^7 \exp(21000/RT_{reactor})$$

$$427 \quad (0.494 + 0.0052T_{BPi} - 0.00002185T_{BPi}^2 + 0.000000321T_{BPi}^3)$$

428 (50)

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

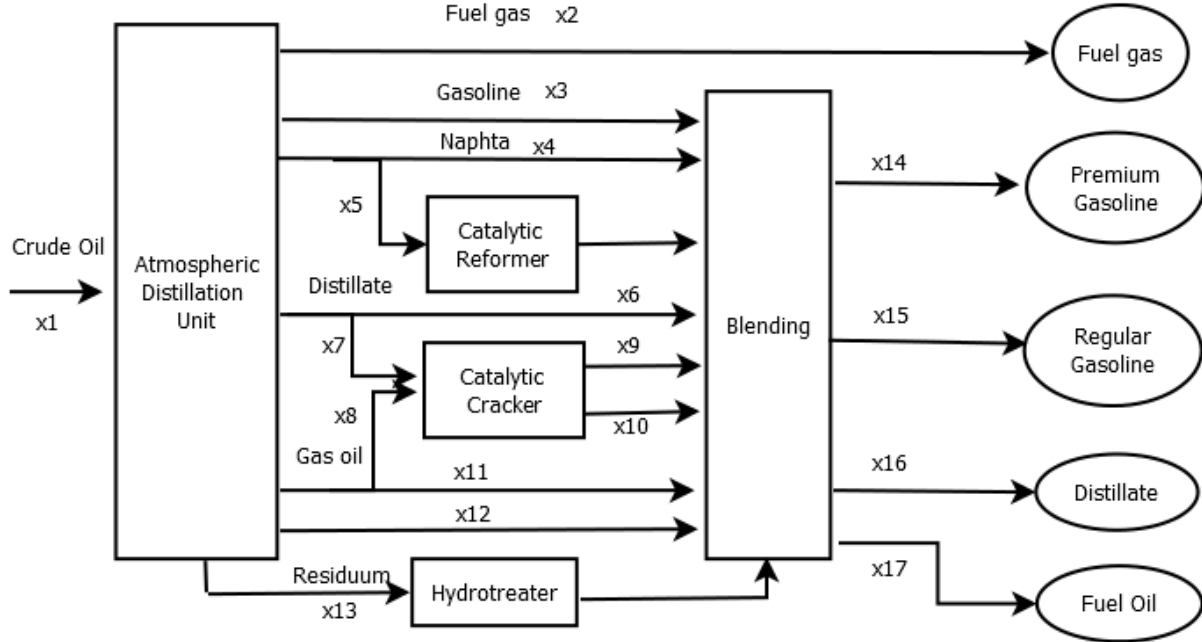
432

## 433 E. Product upgrading

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

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

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



441  
442

443 Fig 4. Process flow of oil refinery [21]

444

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

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

$$457 \quad x_{C/C'} = A_{C,S,p} x_{S/C'} \quad (50)$$

458

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



$$464 \quad \sum_p A_{C,S,p} PROD_{p,S} + u_S \geq 0 \quad (51)$$

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

$$468 \quad \sum_p A_{C',S,p} PROD_{p,S} + u_S - \sum_p A_{C',S,p} BLENDING_{C'SC} \geq 0 \quad (52)$$

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

$$470 \quad TRANSPORT_C = \sum_S \sum_p A_{C,S,p} BLENDING_{S,CI,FP} \quad (53)$$

471

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

$$475 \quad \sum_C \sum_S \alpha_{C,S,Q} \times \frac{\sum_S \sum_p A_{C',S,p} BLENDING_{C'SC}}{TRANSPORT_C} \geq Quality_{FP,Q}^{LB} \quad (54)$$

477 Since not all processes  $p$  ( $p \in P$ ) can be utilized by each processing unit  $U$  ( $u \in U$ ), total  
478 processing capacity of the unit  $u$ ,  $K_u$  is the summation of its individual processing capacity  
479 at certain crude oil  $S$ ,  $Z_{pS}$ , in a condition that the unit is capable of doing process  $p$ ,  $Y_{pU}$   
480  $= 1$ .

$$481 \quad \sum_p Y_{pU} \sum_S Z_{pS} \leq K_u \quad (55)$$

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

$$484 \quad u_S \leq u_S^{UB} \quad (56)$$

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

$$488 \quad \pi_{BLENDING} = \sum_C p_C x_C - \sum_S p_S u_S - \sum_{C'} \sum_S p_{C'} u_{C'S} - \sum_p Cp_p \sum_S Z_{pS} \quad (57)$$

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

## 500 F. Hydrogen management

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

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

$$513 \quad \sum_i x_i K_i = 1, \quad \sum_i y_i / K_i = 1$$

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

520 In 1 unit mass of hydrocarbon products, there are 0.5 – 0.8 unit mass of Nitrogen and 0.05 –  
 521 0.1 unit mass of Sulphur. Each 1 mole of Sulphur (MW = 32) and Nitrogen (MW = 14)  
 522 requires 1 mole of  $H_2$  (MW = 2). Thus for 1 unit mass of hydrocarbon products, hydrogen  
 523 demand for hydrodesulphurization and hydrodenitrogenation is  $\left(\frac{0.05}{32} + \frac{0.5}{14}\right) \times 2 \leq H_2 \leq$   
 524  $\left(\frac{0.1}{32} + \frac{0.8}{14}\right) \times 2$  that is about 14.876 % of the total hydrogen consumption. Thus there is  
 525 572% more consumption for other processes (hydrocracking, and saturation) [22].

526 In addition, the reactors produce C1-C4 occurs during hydrogen injection within a range 1 –  
 527 4 % wt of oil feed  $F_u$ . If the composition of  $F$  originally is  $z_{iu}$ , then for each 1 unit mass of  
 528 hydrocarbon  $i$ , hydrogen is produced as much as %wt. $H_2$  [27], the vapor mass  $V_u$  and  
 529 liquid mass  $L_u$  at each producer unit  $u$ , together with its compounds vapor-liquid  
 530 composition ( $y_{iu}x_{iu}$ ) can be formulated as follows:

$$531 \quad F_u z_{iu} = x_{iu} L_u + y_{iu} V_u$$

$$532 \quad V_u = ((C1 - C4)\% + \%wt.H_2) \times F_u + \sum_i y_{iu} F_u$$

$$533 \quad L_u = F_u - V_u$$

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

$$535 \quad q_u = \%wt.H_2 \times F_u / V_u$$

536

537 Considering that hydrogen production from catalytic cracker cannot meet hydrogen demand,  
 538 another source of hydrogen from biomass is provided which produce 99,99% pure  $H_2$   
 539 through header  $h$ . The gas is mixed with hydrogen from production units PU (i.e., catalytic  
 540 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.

$$576 \quad C_{H_2} = \sum_{u \in U} OC_{Q_u} + OC_{P_u} \quad (70)$$

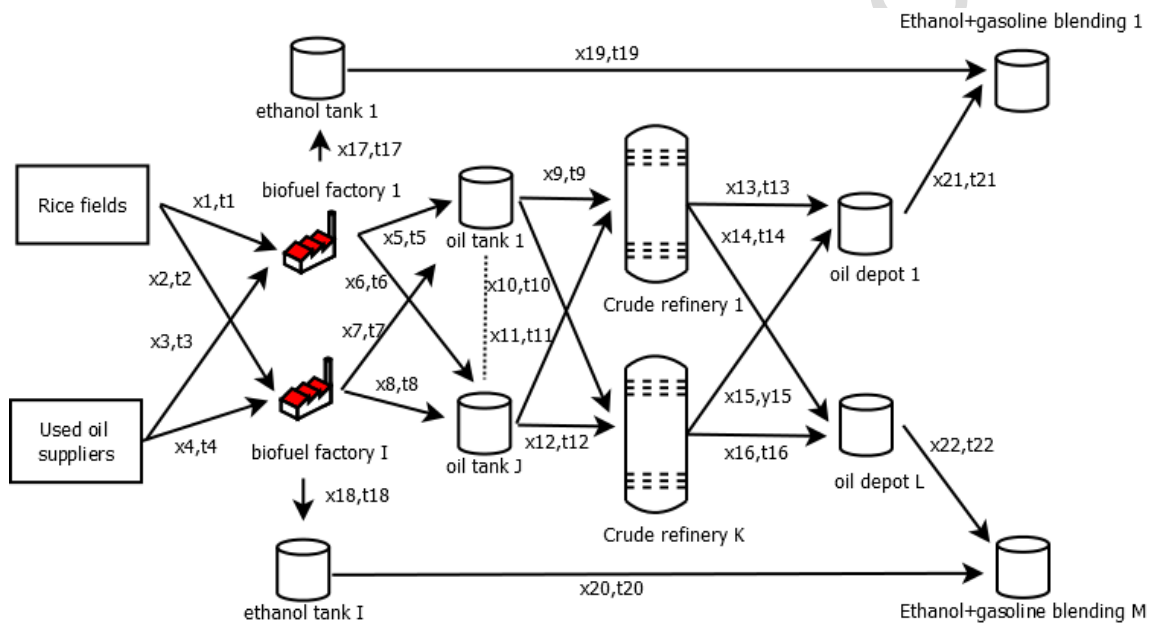
577 Eqs.(58-70) are used to manage the use and production of hydrogen in a hydrocarbon  
578 refinery.

579

## 580 2.2.5 Supply chain design

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

589



590

591

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

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

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

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

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

$$610 \sum_{v \in V} \sum_{r \in \rho_{\tau\tau}^r} \sum_{t \in T} VH_{vr} \geq \sum_{r(\tau,\tau')} \rho_{\tau\tau}^r \quad (71)$$

$$611 \sum_{v \in V} \sum_{L \in \rho_{\tau\tau}^r} \sum_{t \in T} VH_{\tau L} \leq VMAX_\tau \quad (72)$$

$$612 \sum_{r \text{ first}_{\tau\tau'} \in \rho_{\tau\tau}^r} \sum_{v \in V} VH_{vr} = \sum_{r \text{ last}_{\tau\tau'} \in \rho_{\tau\tau}^r} \sum_{v \in V} VH_{vr} \quad (73)$$

$$613 \sum_{v \in V} \sum_{r \in \rho_{\tau\tau}^r} \sum_{t \in T} T_{rv} \phi_{vr} \geq a_\tau \quad (74)$$

$$614 \sum_{v \in V} \sum_{r \in \rho_{\tau\tau}^r} \sum_{t \in T} T_{rv} \phi_{vr} \leq b_\tau \quad (75)$$

$$615 \sum_{r \in \rho_{\tau\tau}^r} DIRECT_{r(\tau,\tau')} + \sum_{r \in \rho_{\tau\tau 1}^r} TRANSSHIP_{r(\tau\tau 1\tau')} + BO_{\tau\tau'} = DEM_{\tau't} \quad (76)$$

$$616 \sum_{r \in \rho_{\tau 0\tau}^r} TRANSSHIP_{r(\tau 0\tau\tau 1)} = \sum_{r \in \rho_{\tau\tau 1}^r} TRANSSHIP_{r(\tau\tau 1\tau')} = \sum_{r \in \rho_{\tau 1\tau}^r} DIRECT_{r(\tau 1,\tau')} \quad (77)$$

$$617 \sum_{r \in \rho_{\tau 0\tau}^r} DIRECT_{r(\tau 0,\tau)} + \sum_{s \in 1}^S prob(s) DEM_{\tau 1t} + \sum_{r \in \rho_{\tau 0\tau}^r} TRANSSHIP_{r(\tau 0\tau\tau 1)} =$$

$$618 \sum_{v \in V} PAYLOAD_v VEHICLE_{vrt} - Z_{r\tau} \quad (78)$$

$$619 \sum_{r \in \rho_{\tau\tau}^r} DIRECT_{r(\tau,\tau')} + \sum_{r \in \rho_{\tau\tau 1}^r} TRANSSHIP_{r(\tau\tau 1\tau')} + \sum_{r \in \rho_{\tau\tau}^r} DIRECT_{r(\tau,\tau')} - \sum_{r \in \rho_{\tau 1\tau}^r} DIRECT_{r(\tau 1,\tau)} -$$

$$620 \sum_{r \in \rho_{\tau 1\tau}^r} TRANSSHIP_{r(\tau 0\tau\tau 1)} - \sum_{r \in \rho_{\tau 1\tau}^r} DIRECT_{r(\tau 1,\tau)} = Z_{r(\tau-1)} - Z_{r\tau}$$

$$621 \quad (79)$$

$$622 SS_{\tau'} = \sum_{r \in \rho_{\tau\tau}^r} DEM_{\tau't} T_{rv} + z\sigma\sqrt{T_{rv}} \quad (80)$$

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

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

640 The supply chain objective is formulated as follows:

$$\begin{aligned}
SC_{cost} = & \sum_{r \in R} \sum_{v \in V} \sum_{t \in T} shipcost_{rv} \cdot shiptime_{rv} \cdot V H_{vrt} + Y_{r(n_1, n_2)} \sum_{r \in \Omega} \sum_{\rho_{rr}^r \in \Omega} DIRECT_{r(\tau, \tau')} + S_{r(\tau, \tau')} * q \\
& + (1 - \phi_{r((\tau, \tau'))}) \sum_{r \in \Omega} \sum_{\rho_{rr1}^r \in \Omega} TRANSSHIP_{r(\tau, \tau1, \tau')} \cdot q
\end{aligned} \tag{81}$$

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

$$\min Tot\_col\_cost_{ij} + Syngas\_cost_i FT\_cost_i + SC_{cost} + C_{H2} - \pi_{BLENDING} \tag{82}$$

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).

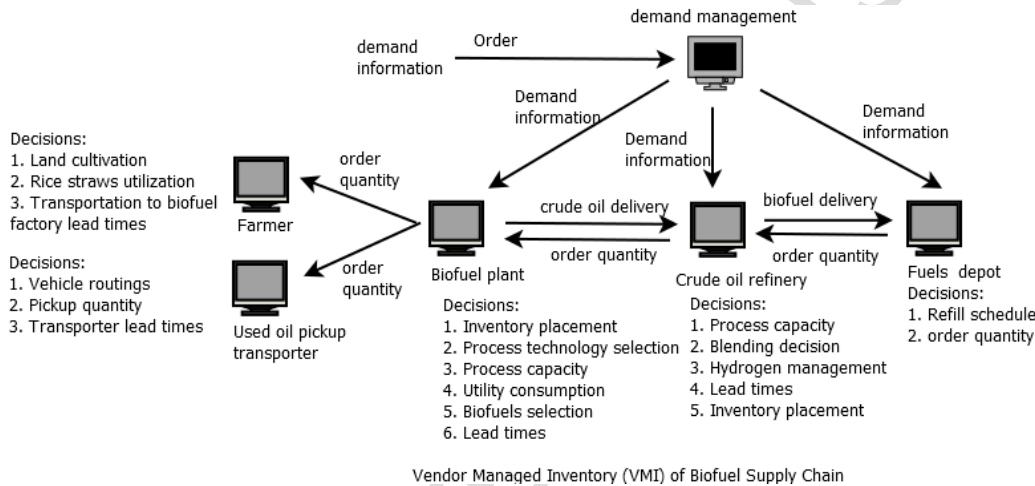


Fig 6. Supply chain information and decisions flows

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

667 refinery location is listed in Table 5 [30].

668

669 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

670

671 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

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

	Cilacap	Balongan	Tuban
Indramayu	2	1	3
Bojonegoro	2	3	1

673

674 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

675

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

Depot	Average Lead Time (Loading & Voyage in days), $d_i$	$\sigma L$	$a_i$	$b_i$
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

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

678 delivery time windows.

### 679 **3. Results and Discussion**

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

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

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

689

#### 690 **3.1 Raw materials supply**

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

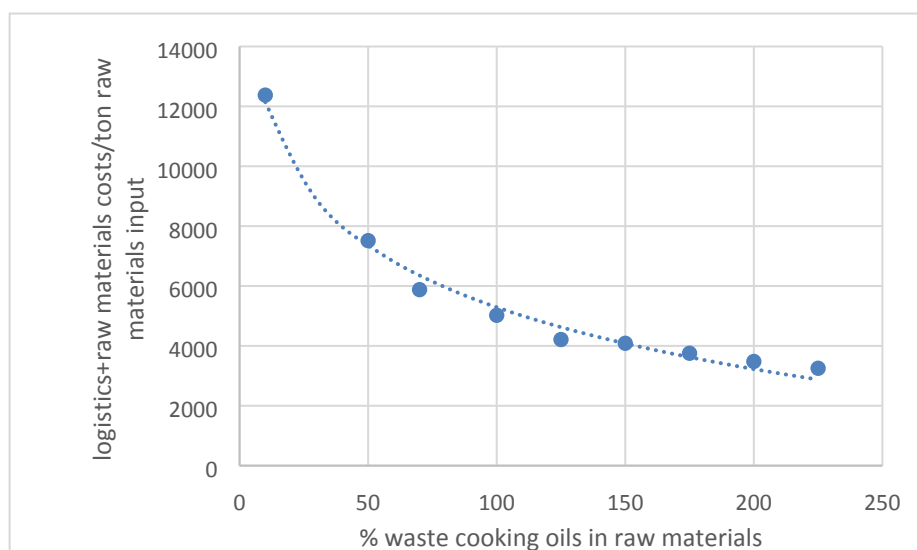
$$694 \frac{\text{logistics} + \text{RM costs}}{\text{ton raw material}} = -2979 \times \ln (\% \text{ cooking oil in raw materials}) + 19007$$

695 (83)

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

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





709

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

712

### 713 3.2 Biofuel synthesis plant

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

716

717

718

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

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

721

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

Fischer Tropsch Reactor	Calculation	Martin and Grossmann [14]
Reactor outlet	533	493.15
CH <sub>4</sub> (% 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 <sub>2</sub> /CO ratio	1.8	1.7
Probability of chain growth	0.879	0.91

724

725 Table 8. Optimum operating condition of hydrocracker of oils from rice straws (Gasification  
726 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

727

728 Table 6 lists operating condition required to produce syngas at maximum hydrogen and CO  
729 yields. In addition the formation of methane is reduced since the gas does not have any  
730 benefit to syngas yield. Similarly Table 7 and 8 exhibit suggested optimum conditions of  
731 Fischer Tropsch and hydrocracker that slightly different to literatures [19,25] and plant data.  
732 The discrepancies are due to the differences of raw materials (switchgrass, vs rice straws),  
733 catalysts, and hydrogen consumptions. Nevertheless, the differences are still within the  
734 allowable ranges, for instances FT reactor temperature range (623 – 668 K), and  
735 hydrocracker reactor H<sub>2</sub> consumption. The output composition depends on the value of  $\bar{\alpha}$ , as  
736 well as H<sub>2</sub> and CO mole ratio. Therefore, raw material H,C and O composition are the main  
737 factors in determining the operating conditions of FT synthesis.

738 In addition to gasification and used oil hydrocracking platforms, the following fermentation  
739 platform is used to produce ethanol as an alternative clean energy source (see Figure 5).  
740 Gasification is used to produce syngas that can be converted into ethanol by fermentation  
741 [16]. Table 9 shows standard operating conditions to synthesize ethanol by syngas platform.  
742 The conditions are simulated by using DWSIM software to obtain technically acceptable  
743 outcomes. While ethanol yield is 99,1%, it is sufficient to be directly mixed with gasoline at  
744 final depots. Therefore the next section of product upgrading is only implemented for  
745 non-ethanol biofuel.

746

747 Table 9. Operations variables of the syngas fermentation platform

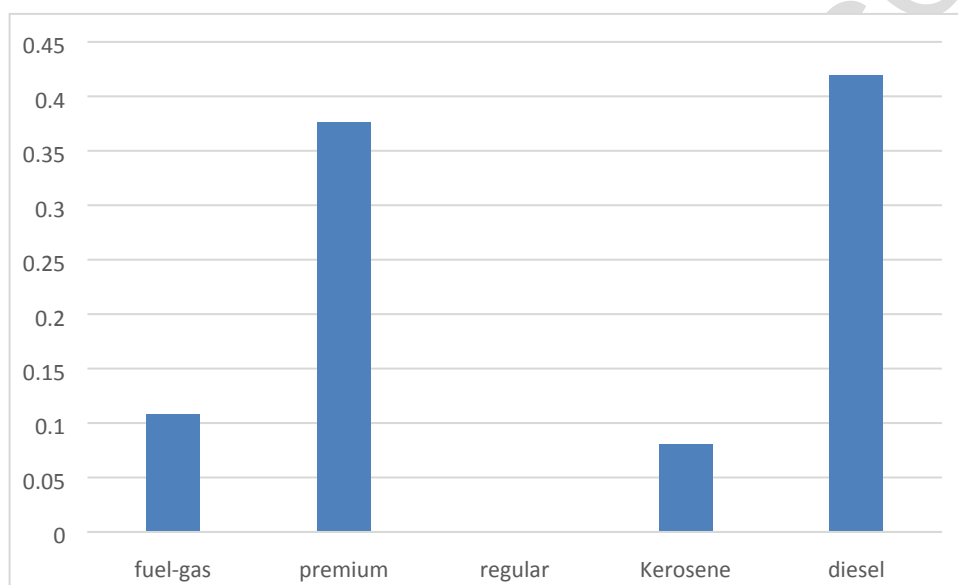
Variables	Value
T gasifier (K)	1375
T mix (K)	561.6
CO <sub>2</sub> in syngas (mol)	0.724
H <sub>2</sub> in syngas (mol)	0.211
CO in syngas (mol)	0.288

H <sub>2</sub> O in syngas (mol)	0
CH <sub>4</sub> 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 <sup>-1</sup> )	0.87
Ethyl acetate conversion to ethanol	0.991

748

749 **3.3 Product upgrading**

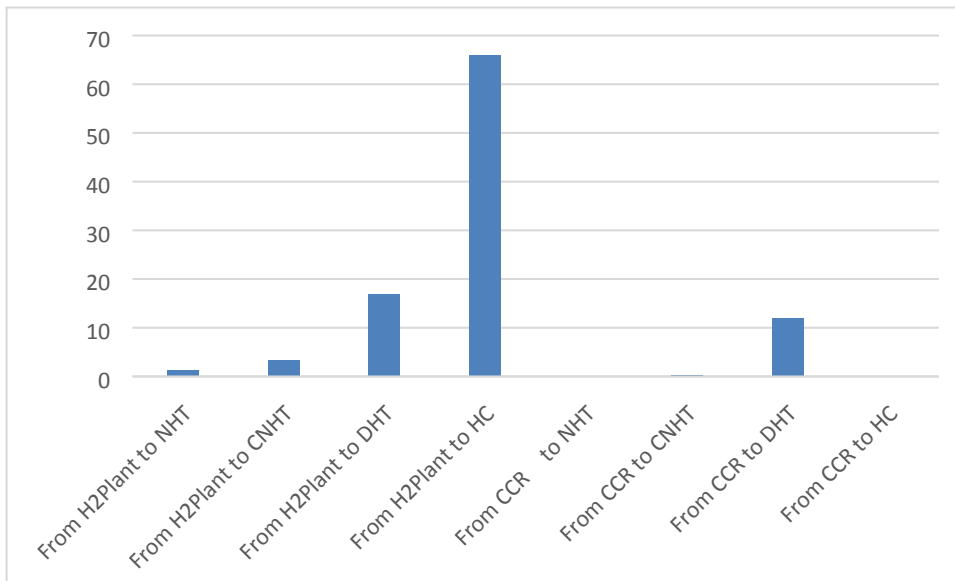
750 In addition to oils synthesis operating conditions, the following product upgrading is  
 751 necessary to produce higher purity of biofuel. The data of flash calculation [31] and the  
 752 following composition of product blending is obtained.



753

754 Fig 8. Blending composition of refinery outputs

755



756

757 Fig 9. Hydrogen allocation (in percentage) of refinery

758 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

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

### 761 3.4 Product distribution

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

766 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

767

### 768 3.5 Summary of economic and social perspectives

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

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

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

$$785 \pi_1 = aq_1 - bq_1^2 - bq_1q_2 - c_1q_1$$

$$786 \pi_2 = aq_2 - bq_2^2 - bq_1q_2 - c_2q_2$$

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

$$791 q_1 = \frac{a - 2c_1 + c_2}{3b}, q_2 = \frac{a - 2c_2 + c_1}{3b}, p = \frac{a + c_1 + c_2}{3}$$

$$792 \pi_1 = \frac{(a - 2c_1 + c_2)^2}{9b}, \pi_2 = \frac{(a - 2c_2 + c_1)^2}{9b}$$

793

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

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

803 Another determinant of choosing platform is that carbon emission level. In crude oil refinery  
 804 the yield of CO<sub>2</sub> is 0.33 kg/kg crude oil [34]. On the other side, CO<sub>2</sub> yield is between 3 Kg  
 805 CO<sub>2</sub>/liter ethanol and 3.24 Kg CO<sub>2</sub>/liter ethanol. By considering ethanol density is 789

806 kg/m<sup>3</sup>, the yield is between 3.8 Kg CO<sub>2</sub>/Kg ethanol and 4.1 Kg CO<sub>2</sub>/Kg ethanol [35]. From  
807 logistics cost, in one hand, oceangoing chemical tankers range from 5,000 tonnes  
808 deadweight (DWT) to 35,000 DWT in size. On the other hand, crude oil can be shipped at  
809 capacity 80,000 to 550,000 DWT. This implies that transport capacity of oil tanker is much  
810 larger than ethanol carrier. In conclusion ethanol global shipping might be considered  
811 carefully prior to execution due to the above factors and therefore hydrocarbons from  
812 biomass is a more attractive option.

813 The social contribution analysis measures the impact of the biofuel supply chain on the  
814 country national development. From social responsibility perspective, the ethanol supply  
815 chain contributes to job creation. Solecki et al [36] estimated per gallon basis, a million  
816 gallons of production capacity generates 2.24 permanent jobs, 10.29 construction jobs, and  
817 nearly 15 indirect jobs and nationally contribute to 0.09 % of GDP to chemical industry  
818 sector. On the other hand, crude oil refinery, based on a study of South Africa oil refinery  
819 data, employs 0.98 employee/barrel or 31394.7 jobs per million gallons. The information  
820 signifies that by taking average salary of employee is 700 US\$ and total income of oil  
821 industry, then the supply chain contribution to the country GDP from oil industry is about  
822 0.17 %.

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

825

#### 826 **4. Conclusions**

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

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

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

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

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

857

### 858 **Appendix A: Mixed integer nonlinear programming (MINLP) model**

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

$$861 \max_{x,y} f(x) + f(y) + f(x,y) \quad (A1)$$

$$862 \text{ s.t. } g(x) + g(y) + g(x,y) \leq 0 \quad (A2)$$

$$863 h(x) + h(y) + h(x,y) = 0 \quad (A3)$$

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

#### 874 *2.3.1 Model of variables $f(x,y), g(x,y), h(x,y)$ convexification*

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

877 While the polynomial and/or signomial variables exist in Eq.(A1), it should be  
 878 transformed according to the results of convex analysis. The nonlinear transformation  
 879 constraints for the geometric variables are first rewritten based on formulation in [37]:  
 880

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

882 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  $r_n \in R$

883 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  $\sum_n s_n \leq 1$

884

885 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

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

889  
 890 Based on three properties of signomial terms [37], a signomial part of  $Z_{ij} = x_i^a x_j^b$  requires  
 891 variables transformation to replace each non-convex terms from  $x_i^a$  to  $X_i^1$  and from  $x_j^b$  to  
 892  $X_j^1$  in the objective function and constraint in with a convex underestimation of it and then  
 893 finding the solution of the convex relaxation (Properties 4 and 5) based on formulation in  
 894 [37]. The relaxation will not interfere convex part of the objective function and constraints in  
 895 Eqs.(2a,2b). Therefore the gaps between original and relaxed problems cannot exceed the  
 896 allowable error  $\varepsilon$  (Property 6).

897  
 898 Property 4:  $h^{CONV}(x_i, Z)$ ,  $g^{CONV}(x_i, Z)$  be convex  $\forall x \in \{x^L, x^U\}$ .

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

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

902  
 903 The underestimation of original problem (1) by transformation according to properties 1 to 3  
 904 is followed by property 4 to 6 to get piecewise linear approximation of the original problem.  
 905 The discretization needs variable bounds as follows:

$$906 e^{\sum_{i \in I} r_{ij} \ln(x_i)^L} \leq e^{\sum_{i \in I} r_{ij} \ln(x_i)} \leq e^{\sum_{i \in I} r_{ij} \ln(x_i)^U} \quad (A4)$$

907 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  
 908 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  
 909 follows:

$$910 e^{Y_j} \geq Y_j^L + \frac{Y_j - Y_j^L}{Y_j^U - Y_j^L} (e^{Y_j^U} - e^{Y_j^L}) \quad (A5)$$

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

$$916 diff_{i}^{max} = e^{Y_j} - Y_j^L - \frac{Y_j - Y_j^L}{Y_j^U - Y_j^L} (e^{Y_j^U} - e^{Y_j^L}) \quad (A6)$$

917 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  $diff_{i}^{max}$   
 918 as the allowable error between original and relaxed solution.

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

$$923 diff_{is}^{max} = e^{Y_{js}} - Y_{js}^L - \frac{Y_{js} - Y_{js}^L}{Y_{js}^U - Y_{js}^L} (e^{Y_{js}^U} - e^{Y_{js}^L}) \quad (A7)$$

924 The discretization steps can be summarized as follows:

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

$$926 diff_{is}^{max} = e^{Y_{js}} - Y_{js}^L - \frac{Y_{js} - Y_{js}^L}{Y_{js}^U - Y_{js}^L} (e^{Y_{js}^U} - e^{Y_{js}^L})$$



- 927 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  
 928 original solution.  
 929 3. Solve Eq.(5) and find  $e^{Y_{js}^U}$   
 930 4. For discretization of  $Y_j$ , generate S discrete value of  $Y_{js}$ , in order to get  $\Delta_{Bs} = Y_{js}^U -$   
 931  $Y_{js}^L$   
 932 5. Continue the process until  $\sum_{s \in S} \Delta_{Bs} = Y_j^U$

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

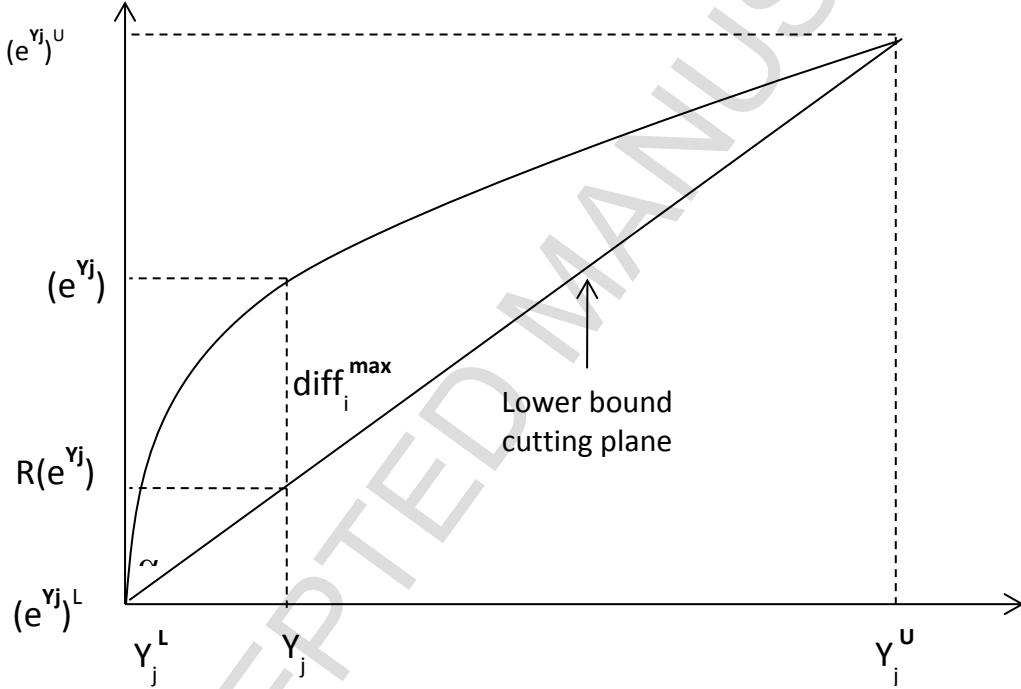
$$935 \min \sum_{j \in J} \left( \sum_{i \in I} e^{r_{ij} \ln(x_i)^U} - \sum_{i \in I} e^{r_{ij} \ln(x_i)^L} \right) y_{si} \quad (\text{A8})$$

$$936 \text{ s.t. } y_{si} \geq y_{(s+1)i} \quad (\text{A9})$$

$$937 \sum_s \left( \sum_{i \in I} e^{r_{ij} \ln(x_i)^U} - \sum_{i \in I} e^{r_{ij} \ln(x_i)^L} \right) y_{si} = \sum_{i \in I} r_{ij} \ln x_{ip} \quad (\text{A10})$$

$$938 \ln x_p = \Delta_1 \theta_1 + \dots + \Delta_p \theta_p \quad (\text{A11})$$

$$939 y_s \in \{0,1\}, \theta_p \in \{0,1\} \quad (\text{A12})$$



940  
 941

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

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

946

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### 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.