

1 **Techno-Economic Performance Analysis of Bio-Oil based Fischer-Tropsch**
2 **and CHP Synthesis Platform**

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5 **Abstract**

6 The techno-economic potential of the UK poplar wood and imported oil palm empty fruit
7 bunch derived bio-oil integrated gasification and Fischer-Tropsch (BOIG-FT) systems for the
8 generation of transportation fuels and combined heat and power (CHP) was investigated. The
9 bio-oil was represented in terms of main chemical constituents, i.e. acetic acid, acetol and
10 guaiacol. The compositional model of bio-oil was validated based on its performance through a
11 gasification process. Given the availability of large scale gasification and FT technologies and
12 logistic constraints in transporting biomass in large quantities, distributed bio-oil generations
13 using biomass pyrolysis and centralised bio-oil processing in BOIG-FT system are technically
14 more feasible. Heat integration heuristics and composite curve analysis were employed for once-
15 through and full conversion configurations, and for a range of economies of scale, 1 MW, 675
16 MW and 1350 MW LHV of bio-oil. The economic competitiveness increases with increasing
17 scale. A cost of production of FT liquids of 78.7 Euro/MWh was obtained based on 80.12
18 Euro/MWh of electricity, 75 Euro/t of bio-oil and 116.3 million Euro/y of annualised capital cost.

19 *Keywords:* gasification; Fischer-Tropsch; polygeneration; bio-oil; energy efficiency; biorefinery

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22 **1. Introduction**

23 The UK's biofuel supply accounted for 3.33% of the total road transport fuel in the past
24 year that exceeded the Government's target of 3.25% [1]. However, this is way below the EU
25 Renewable Energy Directive's mandates of 20% renewable energy targets, including 10%
26 biofuel mix by 2020 [2]. The UK is obliged to meet a target of 15% share of energy from
27 renewable sources by 2020 [2]. At present, the biofuel resources are largely the first generation
28 arable crops that supply majority of bioethanol and waste oils and oily crops that supply
29 biodiesel. The contribution of biomass towards the world's future energy supply can be in the
30 range of 20-50% from currently exploited less than 10% of the total energy supply in the
31 industrialised countries [3]. This implies moving towards energy integrated and efficient
32 lignocellulosic biorefinery systems.

33 Gasification followed by Fischer-Tropsch (FT) is a promising route for producing
34 transportation fuel and CHP from lignocellulosic biomass, in a biorefinery fashion. Integrated
35 biomass gasification FT systems were analysed for technical feasibility and economics,
36 alternative process configurations and prioritising R&D activities for their commercialisation [4-
37 7]. Studies have demonstrated up to 50% overall energy efficiency and economic acceptability
38 achievable due to reduction in capital investment and learning effect. However, transportation,
39 handling, storage, availability and supply in bulk and low bulk energy density of biomass remain
40 a major obstacle in the development of large scale biomass based processing. For these logistic
41 reasons, biomass can be processed into more convenient, cleaner (without tar, char and ashes)
42 and transportable forms such as liquid bio-oils, through fast pyrolysis or liquefaction process.
43 Liquid bio-oil has a higher energy density compared to solid biomass by up to 7 times. This

44 reduces logistics and transportation difficulties, and storage space requirement associated with
45 biomass [8]. Fast pyrolysis involves thermal decomposition reactions that occur in a few seconds,
46 at modest temperature conditions and in the absence of oxygen or in oxygen lean environment.
47 Aston University, UK and Dynamotive undertakes extensive research and development of
48 biomass pyrolysis technologies for the production of bio-oil [9-11]. The Energy research Centre
49 of the Netherlands (ECN) focuses on the thermochemical production routes from biomass to
50 syngas and FT diesel production [7,12]. Other specialist biomass gasification and pyrolysis
51 research centres include BTG [13,14] and Indian Institute of Petroleum [15]. Gasification of bio-
52 oil [16] and bio-oil / char (bio-slurry) [17,18] have received considerable interest, and research
53 has been carried out extensively by FZK, Dynamotive, Future Energy and BTG.

54 The reference bio-oil used is from the UK poplar wood and oil palm empty fruit bunch
55 produced elsewhere. The area available for growing UK poplar wood is ~0.8 million ha [19].
56 Indonesia (7 million ha) and Malaysia (4.5 million ha) are the major producers of oil palm [20].
57 Oil palm empty fruit bunch can also be processed into bio-oil [21] that can be imported to UK.
58 Bio-oil can be collected from various distributed biomass pyrolysis plants and imported and
59 subsequently processed in centralised BOIG-FT plants [11]. In a large centralised BOIG-FT
60 plant, similar to a petro-refinery accepting oils from different locations within specifications, the
61 bio-oil can be converted into different products through a series of processing operation. Even
62 though gasification and FT technologies are available at centralised scale, bio-oil is a premature
63 energy commodity and overall BOIG-FT systems for various biomass resources must be energy
64 integrated and analysed using energy and process integration tools, which are the main aims of
65 this study.

66 The creation of an advanced biorefinery framework utilising bio-oils into biofuels is
67 potentially a very effective way to lower our dependency on crude oil based refinery [22].
68 However, there are techno-economic barriers to overcome for the commercial deployment of
69 such biorefining technologies. The main objective of this paper is to achieve competitive process
70 efficiency using heat integration and CHP network design strategies and enhance economic
71 feasibility by the employment of full conversion configurations (in addition to once-through
72 configurations) and economies of scale ranging from 1 MW, through 675 MW to 1350 MW. The
73 study presents systematic modelling and operability studies of BOIG-FT processes based on their
74 impacts on overall system performance. The viability of importing bio-oil from other countries
75 using shipping cost analysis has also been assessed.

76 **2. Modelling and Simulation**

77 *2.1 Process Description*

78 The BOIG-FT system under consideration can be divided into four main sections:
79 gasification, gas cleaning and processing, FT synthesis reactor and power generation, as depicted
80 in Figure 1. The process models used in ASPEN Plus simulation are presented in Table 1.

81 **Figure 1**

82 **Table 1**

83 Bio-oil (stream 2 in Figure 1) is converted into product gas through the high-temperature
84 gasification (~1100°C), GASIFIER (modelled using Gibbs free energy minimisation method in
85 ASPEN Plus), using oxygen (stream 1). Entrained flow gasifier (technology developers
86 including GE, Shell, E-Gas etc.) can be employed for gasifying bio-oil [13]. An oxygen-blown

87 gasifier is opted since air will lower the heating value of the resulting product gas. Oxygen is
88 assumed to be supplied from oxygen plant, the cost of which was included in the economic
89 analysis. Oxygen was preheated to 480°C using high pressure (HP) steam generated on site. In
90 addition to the reduction in the operating temperature of the gasifier, preheating oxygen can also
91 achieve thermally neutral condition in the gasifier. It is not necessary to preheat bio-oil since it
92 will degrade and promote char formation at high temperature. Product gas (stream 4) from the
93 gasifier contains significant amount of CO and H₂ which are the main reactants for the FT
94 process, and it has been assumed to be free from nitrogen, sulphur, tar and ash. Subsequently, the
95 product gas passes through a series of cleaning and processing units, in order to meet the
96 stringent conditions required by the FT synthesis process. These include water removal columns,
97 H2OREM1 and H2OREM2. H2OREM1 was used to manipulate the steam required by the
98 water-gas shift reaction whilst H2OREM2 was used to reduce the amount of water content in
99 syngas which may influence the conversion of FT reaction. The water removed can be sent to a
100 wastewater treatment plant, e.g. physical, chemical or biological processes, the cost of which has
101 been considered in this study. A water-gas shift reactor (WGS) was used for adjusting the H₂/CO
102 molar ratio to about 2, detailed in section 2.4. A Sulfinol unit (a combined physical and chemical
103 solvent process) [4], represented as a CO₂ separator (CO2SEP) was needed to remove CO₂ by
104 99% on molar basis. The Sulfinol unit can also be used for co-capturing H₂S and COS.
105 Subsequently, the CO₂ captured was compressed in CO2COMP to 80 bar for storage.

106 The clean syngas is sent to the FT reactor (FTREACT) for producing hydrocarbon liquids.
107 The type of FT reactor configuration is highly dependent on the desired product distribution [23].
108 The low temperature Fischer-Tropsch (LTFT) was employed for higher diesel range production
109 (C₁₂-C₁₈). A LTFT tubular fixed bed reactor with cobalt-based catalyst available from Shell,

110 Malaysia can be employed [24]. The operating temperature of FT reactor is 240°C and the
111 pressure is 25 bar, within typical ranges of 200-250°C and 25-60 bar, respectively. Cobalt
112 catalyst is active at lower temperature operations and for high diesel and wax yields. A flash
113 column (VLFLASH) was used to separate the gaseous stream 20 and liquid product stream 21
114 which were cooled to 40°C.

115 The offgas containing light gases such as CH₄, C₂H₆ etc. (stream 23) are fed to gas
116 turbine (GASTURB) for power generation. To enable a stable combustion, Wobbe Index of the
117 gas turbine was validated to ensure that it was within ±10% compared to the base case provided
118 by Shah et al. [25]. The Wobbe Index is a measure of interchangeability of fuel gases and for
119 comparing the combustion energy among fuel gases with different compositions. Co-firing with
120 trace amount of natural gas was further undertaken in the gas turbine to increase the Wobbe
121 Index of the feed gas to the gas turbine. Co-firing was advantageous since it could eliminate any
122 need for modifying the gas turbine and combustor for low heating value syngas [26]. Air
123 containing 79 mol% N₂ and 21 mol% O₂ was compressed to 14 bar through AIRCOMP and
124 supplied to the gas turbine combustion chamber (GTCOMB) to assist the combustion reaction of
125 FT offgas. Since the LHV of the syngas into the combustor was greater than 6 MJ/m³, de-rating
126 (reduction in burning temperature and compressor power ratio to allow stable combustion) of the
127 gas turbine combustor was not required for an operating condition of 1200°C and 14 bar in the
128 combustor. The exit temperature and pressure of the exhaust gas from the gas turbine (stream 29
129 in Figure 1) are 744°C and 2 bar respectively. An enthalpy of -2.3×10^3 kJ/kg and entropy of 1.2
130 kJ/kg-K (with respect to the reference temperature and pressure of 25°C and 1.013 bar) were
131 available with this exhaust gas. The heat content in the exhaust gas from the gas turbine was
132 recovered into the generation of high pressure superheated steam using heat recovery steam

133 generator (HRSG). The carbon balance across the centralised BOIG-FT system (Figure 1) shows
134 0.19 kmol/GW of net carbon removal from the site. This is a net result of CO₂ capture, which has
135 a carbon negative impact, and addition of natural gas, which has a carbon positive impact.

136 *2.2 Selection of Model Components for Bio-Oil*

137 The model component of bio-oil was represented by validating its gasification
138 performance. The bio-oil represented using one (dextrose), two (acetic acid and guaiacol), three
139 (acetic acid, acetol and guaiacol) and four (acetic acid, acetol, guaiacol and furfural, C₅H₄O₂)
140 chemical components was assessed. These models were tested by comparing the syngas
141 composition obtained from simulation of bio-oil gasification, against industrial reference cases
142 [27]. The feed and product composition and gasification process operating conditions are
143 presented in Table 2. The composition of bio-oil, represented by different sets of model
144 components, was estimated by balancing the C, H and O elements in the product gas component.
145 The combinations were selected systematically by taking into consideration that the aqueous
146 phase comprises acetic acid and acetol (soluble in water), while guaiacol and furfural (insoluble
147 in water) are present in the lignin phase of bio-oil. The rationale was to distribute the masses of
148 C, H and O according to the atomic ratio, e.g. for 3-component model, acetic acid (C₂H₄O₂)
149 carries 2/12 of the mass of C, whilst acetol (C₃H₆O₂) and guaiacol (C₇H₈O₂) carry 3/12 and 7/12
150 of the mass of C, respectively. An example of the mass balance in relation to C, H and O
151 elements for 3-component model (acetic acid, acetol and guaiacol) is presented in Table 3.

152 **Table 2**

153 **Table 3**

154 The chi-square (χ^2) test was performed for justifying statistical significance of bio-oil
155 representation. The statistical significance gives an indication of how likely the hypothesis is
156 wrong. In the present context, the null hypothesis is defined as “the selected model component(s)
157 is (are) acceptable for representing bio-oil”. Firstly, χ^2 for each model was calculated, using
158 equation (1), and is summarised in Table 2. $y_{i,S}$ and $y_{i,R}$ denote the mole fraction of component i
159 (components in product gas) obtained from simulation and reference cases, respectively.

$$160 \quad \chi^2 = \sum_{i=1}^N \frac{(y_{i,S} - y_{i,R})^2}{y_{i,R}} \quad (1)$$

161 By assuming a significance level (the probability of rejecting the hypothesis) of 0.05, and
162 taking into account that there are 4 degrees of freedom, the critical value was determined to be
163 9.49 from chi-square distribution table [28]. As can be seen from Table 2, the χ^2 for each model
164 falls below the critical value of 9.49. Therefore, it can be concluded that there is no sufficient
165 evidence to reject the hypothesis.

166 The lower χ^2 value obtained for the 3-component model presented in Table 2, signifies
167 that the 3-component model is more acceptable compared to the others. The residual sum of
168 square (RSS) statistical method demonstrated in equation (2) was also applied to describe the
169 degree of discrepancy between the simulation and the reference results in terms of compositions
170 of syngas obtained from bio-oil gasification. The RSS obtained for each model is presented in
171 Table 2.

$$172 \quad \text{Residual sum of square (RSS)} = \sum_{i=1}^N (y_{i,S} - y_{i,R})^2 \quad (2)$$

173 The 3-component model (acetic acid, acetol and guaiacol) with the lowest χ^2 value (0.018)
174 and the least RSS value (0.41) can thus adequately represent bio-oil for its performance through
175 the gasification process under consideration.

176 2.3 Sensitivity Studies of Gasification

177 The sensitivity studies in terms of the effect of the gasifier temperature, pressure and
178 oxygen-to-feed (O_2/F) molar ratio on H_2 and CO concentration in the product gas, were studied.
179 Figure 2(a) presents changes in the composition of the product gas with the gasification
180 temperature, for a constant pressure of 30 bar and O_2/F molar ratio of 0.54. It can be seen that H_2
181 and CO contents in the product gas rise substantially from 500°C to 1000°C. H_2 starts declining
182 and CO increases at a lower rate, above 1000°C. It is thus proposed that the gasifier temperature
183 should be kept above 1000°C. Also, the concentration of CH_4 is less than 1 vol% at this
184 temperature [13]. A relatively higher proportion of CH_4 is advantageous if the gas is used into
185 power generation, however it is undesirable for an application into FT synthesis reaction [13].
186 Lower CO_2 composition can also be observed when the temperature increases. FT synthesis
187 imposes a stringent condition on H_2/CO molar ratio of 2. The impact of the gasifier temperature
188 on H_2/CO molar ratio in the product gas is depicted in Figure 2(b). The H_2/CO molar ratio in the
189 gas from the gasifier normally falls below 2. Hence, the ratio needs to be adjusted using water-
190 gas shift reactor in between the gasifier and the FT reactor.

191 **Figure 2**

192 The pressure has a negligible impact on the product gas composition due to the equimolar
193 stoichiometric gasification reactions, where pressure has less effect on changing the equilibrium
194 composition. However, in order to avoid any compression work for the subsequent FT reaction at

195 25 bar, the gasifier was operated at a higher pressure (Table 1). The gas composition is highly
196 sensitive to the changes in O₂/F molar ratio, as shown in Figure 3(a). It is intended to keep the
197 O₂/F molar ratio as low as possible (e.g. less than 0.5) primarily to maintain increasing H₂ and
198 CO contents in the product gas. As illustrated in Figure 3(b), lower O₂/F molar ratio results in
199 higher H₂/CO molar ratio in the product gas.

200 **Figure 3**

201 *2.4 Modelling and Sensitivity Studies of FT Synthesis*

202 The H₂/CO molar ratio in the syngas feed to the FT reactor is a highly influential
203 parameter for dictating the reaction rate. Higher H₂/CO molar ratio results in a higher selectivity
204 for lighter hydrocarbons due to higher probability of chain termination. Water gas-shift reactor
205 should be used prior to FT reactor to adjust H₂/CO molar ratio since cobalt catalyst has
206 negligible water-gas shift activity. Ideally, the required H₂/CO molar ratio is 2 (practically is
207 about 2.15) when cobalt catalyst is used [29].

208 The FT synthesis reactions are highly exothermic. The main reactions involve the
209 production of paraffin (equation (3)) and olefin. The production of straight-chain paraffins from
210 C₁ to C₃₀ was modelled using equation (3) in a stoichiometric reactor (RStoic in Table 1) in
211 ASPEN Plus.



213 The fractional conversion of each reaction was estimated based on the weight distribution
214 of each FT product obtained via Anderson-Schulz-Flory (ASF) distribution model [30,31], in
215 Excel spreadsheet. This model assumes a constant chain growth probability, α . Thus, $(1-\alpha)$

216 equals to the probability of termination of the carbon chain. In general, FT reaction involves two
217 principal mechanisms (excluding chain initiation step):

218 • Chain growth by absorbing CO and H₂ (stepwise addition of CH₂) to form longer carbon
219 chain length.

220 • Termination by desorption from the catalyst to form paraffin or olefin.

221 The ASF model which relates the weight fraction of the FT product, w with the chain
222 growth probability α , is described in equation (4). n denotes the carbon number.

$$223 \quad w_n = \alpha^{n-1} (1-\alpha)^2 n \quad (4)$$

224 Typical values of α fall between 0.7-0.9 [31]. An empirical equation (5), in which α can
225 respond to changes in temperature, by Song et al. [32], was adopted. A value of α of 0.76 at a FT
226 reactor operating temperature of 240°C is used in the simulation.

$$227 \quad \alpha = \left(A \frac{y_{CO}}{y_{H_2} + y_{CO}} + B \right) [1 - 0.0039(T - 533)] \quad (5)$$

228 A and B are the parameters with values of 0.2332±0.0740 and 0.6330±0.0420, respectively [32].

229 y_{CO} and y_{H_2} denote the mole fractions of CO and H₂, respectively in the feed stream to the FT

230 reactor. T is the operating temperature of the FT reactor, K.

231 By applying ASF relation in equation (4), the weight and molar distributions of each FT
232 product were obtained. An overall conversion of 80% of the syngas (by taking CO as the limiting
233 reactant) was assumed. The fractional conversion of CO in each reaction was determined by
234 solving equation (6) in the spreadsheet environment integrated to the Aspen simulation
235 framework.

236 Total CO converted = $\sum_{n=1}^{n=30} y_{p,n} \times F_p \times n$ (6)

237 $y_{p,n}$ is the mole fraction of paraffin with carbon number n produced. F_p is the total molar flow
238 rate of paraffin produced, kmol/s.

239 The simulation results in terms of the weight and molar distribution are featured in Figure
240 4(a) and (b), respectively. By adopting equation (5), a sensitivity analysis in terms of the effect
241 of temperature on the product distribution can be evaluated, in Figure 5. It implies that a higher
242 diesel range product can be attained at a lower temperature.

243 **Figure 4**

244 **Figure 5**

245 **3. Performance Analysis of BOIG-FT System of Various Capacities**

246 The BOIG-FT systems with various capacities were considered for 80% conversion of
247 CO in the FT feedstock. In addition, the 675 MW case was studied for 60% CO conversion. For
248 full conversion FT reactor configuration, CO conversion of 40% per pass was assumed, so that
249 an overall 80% CO conversion can be obtained.

250 *3.1 Combined Heat and Power (CHP) Integration and Energy Efficiency*

251 Strategies for overall heat integration of the BOIG-FT system under consideration have
252 been proposed, in order to maximise energy recovery within the site and thereby improve cost-
253 effectiveness of the system. The sensible heat from the site can be graded based on the
254 temperature level and the heat content. The high grade surplus heat from processes can be used
255 into process heating and the generation of high pressure steam to be used for other processes.

256 The low grade surplus heat from the site can be utilised into low level process-to-process heating.
257 The CHP integration strategy was demonstrated for an integrated coal gasification combined
258 cycle study by Ng et al. [33]. The process heat supply and demand are listed in Table 4.

259 **Table 4**

260 Three levels of steam can be generated from the site, i.e. high pressure (HP) steam at 40
261 bar and superheated to 500°C, medium pressure (MP) steam at 15 bar and slightly above the
262 saturated temperature at 200°C and low pressure (LP) steam at 5 bar and 152°C, respectively.
263 The high grade heat from SYNGCOOL + H2OREM1 and HRSG was used in the generation of
264 HP steam, whilst the exothermic heat from FTREACT was utilised into the production of MP
265 steam. The excess HP and MP steam was let down to generate low pressure (LP) steam and
266 additional power.

267 The hot and cold composite curves [34] were analysed for estimating the maximum
268 amount of steam generation from SYNGCOOL + H2OREM1 and HRSG, presented in Figure
269 6(a) and (b), respectively. In each of these figures, the cold composite curve was shifted
270 horizontally towards the hot composite curve (i.e. by increasing the mass flow rate of steam) in
271 order to reach a minimum approach temperature of 20°C. The shifting of composite curves
272 ensures a maximum heat recovery that can be attained from a site. The SYNGCOOL and
273 H2OREM1 can physically be one unit. The temperatures and heat duties of the hot streams
274 (SYNGCOOL + H2OREM1 and HRSG) and the cold streams (steam generation) were obtained
275 from ASPEN Plus simulation. For the HP steam generation, water at 25°C was heated to the
276 saturated temperature of 250.4°C at 40 bar, and then superheated to 500°C. The amount of steam

277 generation from FTREACT was estimated from its heat of reaction at isothermal condition and
278 the enthalpy of vaporisation of MP steam at 200°C at 15 bar.

279 **Figure 6**

280 The heat demands by HE1, HE2 and HE4 were satisfied by the HP steam, and the
281 remaining steam was sent to a steam turbine ST1 for power generation. The composite curves for
282 these heat balances were analysed to determine the amount of HP steam needed. The steam
283 required by the preheaters and reboilers in Sulfinol unit for solvent regeneration was evaluated
284 based on 1.42 kg/kg of LP steam to acid gas [35]. The remaining HP and MP steam was let down
285 through the steam turbines ST2 and ST3 (back pressure turbines) generating additional power.
286 The excess LP steam from the exit of the steam turbines after fulfilling the heat requirement by
287 the Sulfinol unit was utilised in condensing turbine ST4 for power generation. The site-wide
288 steam and power network is depicted in Figure 7. The steam turbines were assumed to perform at
289 80% isentropic efficiency and 95% mechanical efficiency. The low grade heat surplus from
290 WGS, H2OREM2, HE3, HE5 and condenser can be used for generating hot water for domestic
291 heating systems and small-scale industrial plants.

292 **Figure 7**

293 Energy balances performed for different capacities of the BOIG-FT system (1 MW, 675
294 MW and 1350 MW) are summarised in Table 5 for once-through configuration and Table 6 for
295 full conversion configuration, respectively. The LHV of FT liquids was assumed to be 45 MJ/kg.
296 The net heat generation accounts for the entire low grade heat surplus from the site that does not
297 involve steam generation. The C₅₊ liquid yield and selectivity between various BOIG-FT system
298 cases under consideration were analysed and compared. Additionally, a comparison of energy

299 efficiency based on the lower heating values (LHV) of the resulting FT liquids and the amount of
300 heat and power generation is also presented.

301 **Table 5**

302 **Table 6**

303 The 675 MW BOIG-FT system with once-through configuration was studied for 80% and
304 60% conversions respectively. The lower conversion case has the advantage of generating more
305 power, e.g. approximately 2.4% higher than the system with 80% conversion. This is due to
306 higher amount of offgas produced from the FT reaction that is utilised in the gas turbine.
307 However, the BOIG-FT system with 60% conversion has efficiency 15% lower than the system
308 with 80% conversion (Table 5). This is due to a much lower C₅₊ liquid yield, approximately
309 22.2% lower than the system with 80% conversion. Therefore, higher conversion of the FT
310 reaction is more favourable for obtaining higher efficiency. Further, it can be concluded that
311 higher efficiency can be achieved at higher capacity of the BOIG-FT system, while the full
312 conversion configuration is generally more efficient than the once-through configuration. The
313 efficiency which includes the net heat is higher than the efficiency without the net heat,
314 suggesting that the recovery of the low grade heat into useful by-products such as hot water can
315 be beneficial.

316 The proximate and ultimate analyses of bio-oils obtained from poplar and oil palm empty
317 fruit bunch are presented in Table 7(a). The performance of BOIG-FT system using bio-oils from
318 poplar and oil palm empty fruit bunch is compared in terms of energy efficiency, defined as the
319 LHV of FT liquid, electricity with / without heat with respect to the LHV of bio-oil, summarised
320 in Table 7(b). The LHV of FT liquid was assumed to be 45 MJ/kg. The efficiency of FT liquid

321 and CHP generated using bio-oil from oil palm empty fruit bunch is higher than using poplar, i.e.
322 50.5% compared to 38.9%, respectively. This is due to the lower moisture content in the oil, i.e.
323 30% and 18.6% for bio-oils from poplar and oil palm empty fruit bunch, respectively. The FT
324 liquid, low grade heat and electricity contributed to 54-66%, 22.5-33.2% and 11.6-12.8% of the
325 total output energy, for the 1350 MW case using either poplar or oil palm empty fruit bunch as
326 feedstocks. The surplus low grade heat from the site can be recovered into hot water generation,
327 which is highly essential in the UK energy scenario. The energy efficiency of biomass
328 gasification with FT synthesis has been reported to be 42-50% [5], which in turn reflects that the
329 bio-oil system is competitive against the biomass system.

330 **Table 7**

331 *3.2 Economic Analysis*

332 Detailed capital cost, operating cost and discounted cash flow (DCF) analyses performed
333 for 1 MW, 675 MW and 1350 MW LHV cases with once-through and full conversion
334 configurations are presented. The economic analysis also includes an analysis of the netback of
335 bio-oil. In addition, the import economics of bio-oil derived from oil palm empty fruit bunch was
336 evaluated for viability in the UK context.

337 *3.2.1 Capital Costs*

338 The capital costs of a BOIG-FT system include the direct (ISBL and OSBL) and indirect
339 capital (design and installation costs for constructing a site as well as the costs forecasted for
340 some unforeseen circumstances). The cost data, base costs, scale factors and base scales of major
341 process units in ISBL were adapted from Tijmensen et al. [5], summarised in Table 8. A scale

342 factor R was applied for different sizes in the cost and size relationship, in equation (7).
 343 Maximum size of a gasifier is 400 MW HHV. Thus, multiple units were taken into account for
 344 the gasification capacity exceeding its maximum size. The cost of Sulfinol unit is generally
 345 proprietary information and has been assumed to account for 10% of the total capital cost [4].
 346 The cost of SYNGCOOL was estimated the same way as for the HRSG. This also includes the
 347 base cost, scale factor, base scale and progress ratio.

$$348 \quad \frac{COST_{size2}}{COST_{size1}} = \left(\frac{SIZE_2}{SIZE_1} \right)^R \quad (7)$$

349 $SIZE_1$ and $COST_{size1}$ represent the capacity and the cost of the base system, whilst $SIZE_2$ and
 350 $COST_{size2}$ represent the capacity of the system after scaling up/down and its corresponding cost,
 351 respectively. R is the scaling factor.

352 **Table 8**

353 Cost index method was applied, as given in equation (8), for levelising the cost taken
 354 from to the recent year 2009. The cost index adopted is from Chemical Engineering Plant Cost
 355 Index (CEPCI) [36]. The CEPCI for the years 1999 and 2009 are reported at 390.6 and 524.2,
 356 respectively.

$$357 \quad \text{Present cost} = \text{Original cost} \times \left(\frac{\text{Index at present}}{\text{Index when original cost was obtained}} \right) \quad (8)$$

358 *3.2.2 Operating Costs*

359 The operating costs of the BOIG-FT system were evaluated by considering the fixed and
 360 variable operating costs and other miscellaneous costs such as sales expense. Fixed operating

361 costs are independent of the production rate, and estimated based on percentage of total indirect
362 capital costs (TIC) [37], given in Table 9. The cost allocated for personnel is based on the work
363 by Tijmensen et al.[5].

364 **Table 9**

365 *3.2.3 Discounted Cash Flow and Netback of Bio-oil Analysis*

366 The discounted cash flow (DCF) analysis was performed to estimate the annual capital
367 charges. The cumulative discounted cash flow is expressed as the net present value (NPV) in a
368 DCF analysis. A discount rate of 10% for 15 operating years has been assumed. The construction
369 / start-up period was assumed to be 2 years, where 25% and 75% of the total capital cost were
370 distributed in the -1 and 0th year (0th year indicates plant start-up year) [5]. An annualised capital
371 charge of 13.1% can be incurred, for the conditions specified above.

372 The netback indicates the value of a feedstock obtained from selling its products at
373 market price after the deduction of associated costs. The netback thus sets the maximum
374 acceptable cost (market price) of a feedstock. The cost / market price of a feedstock needs to be
375 less than its netback in order to make profits from its processing. The netback of bio-oil was
376 calculated by applying equation (9).

377
$$\text{Netback} = \text{Value from products} - (\text{Annualised capital cost} + \text{Annual operating cost}) \quad (9)$$

378 The products from the BOIG-FT system are the FT liquids and electricity. It has been
379 assumed that the system operates for 8000 hours per year. The value of the FT liquids was
380 assumed to be 42.6 Euro/MWh, which is its cost of production from wood [38]. The price of
381 electricity was adopted from DECC, reported at 7.284 pence/kWh in the year 2009 (equivalent to

382 80.12 Euro/MWh, assuming 1 GBP = 1.1 Euro), and excluding Climate Change Levy (CCL)
383 [39]. CCL only applies to industrial sector where taxable supplies such as electricity, coal and
384 petroleum are charged. The full rate of CCL for electricity is reported at 0.47 pence/kWh
385 (equivalent to 5.17 Euro/MWh). A comparison of costs, i.e. annualised capital charge, annual
386 operating cost, value from products and thus the netback of bio-oil (with and without CCL) for 1
387 MW, 675 MW and 1350 MW cases with once-through and full conversion configurations is
388 presented in Table 10.

389 **Table 10**

390 The netback of bio-oil is an effective way for examining the economy of scale. The
391 BOIG-FT system is economically competitive for larger capacities, i.e. the netback of bio-oil
392 ranges from 11.4-20.9 Euro/t (excluding CCL) for 675 MW and 1350 MW capacities for both
393 configurations. The full conversion configuration is more attractive than the once-through
394 configuration. In addition, lower conversion is not desirable, as demonstrated in 675 MW case
395 with 60% conversion, which has a negative netback of bio-oil of -7.96 Euro/t (excluding CCL).

396 *3.2.4 Import Economics of Bio-oil*

397 Secured and ample supply of bio-oil is required for a large scale BOIG-FT system.
398 Import of bio-oil from other countries at an acceptable cost may be an option to fulfil the rising
399 demand for biofuels in the UK. Limited cost information on importing bio-oil from other
400 countries to the UK is available, mainly due to the fact that bio-oil is yet to be established as an
401 energy commodity in the UK and European countries. Hence, the total delivered cost of bio-oil
402 that depends on various aspects, such as loading and discharging, rail and road transportation,
403 labour, taxes, shipping etc., is yet not fully estimated. The delivered cost of bio-oil to Rotterdam,

404 The Netherlands from Canada, Brazil, South Africa, Ukraine and Baltic was studied by Bradley
405 [40]. This cost was 6-10.6 Euro/GJ, using 4700 t tankers (ship) and without return trip.

406 Shipping cost is the major component of the delivered cost of bio-oil, especially when the
407 distance between the countries is significant. This study estimates the shipping cost of bio-oil
408 derived from oil palm empty fruit bunches, from Malaysia to the UK. Previous study provided an
409 estimation of the shipping cost of bio-oil from Vancouver to Rotterdam to be 4.96 Euro/GJ, for a
410 distance of 14400 km [40]. Thus a shipping cost of 5.2 Euro/GJ (equivalent to 83.2 Euro/t) is
411 incurred for transporting bio-oil from Port Kelang, Malaysia to Port of Immingham, UK, over a
412 distance of 15000 km, assuming a linear relationship between the distance and the shipping cost.
413 For 2.43 million t/y of bio-oil import to the 1350 MW BOIG-FT system, 202.2 million Euro/y
414 are incurred from shipping of bio-oil that is 4 times the original operating cost of 52.8 million
415 Euro/y (Table 10). The high cost of shipping of bio-oil from other countries rich in biomass
416 resources such as South East Asia is not economically viable at present. However, this cost can
417 be reduced considerably (e.g. by approximately half) by introducing larger tanker for shipment,
418 i.e. 61 Euro/t to 30 Euro/t [40]. Correspondingly, 41.6 Euro/t of shipping cost indicates a total
419 cost of 101 million Euro/y for 1350 MW system.

420 By incorporating shipping costs of 83.2 Euro/t and 41.6 Euro/t, the netback of bio-oil for
421 1350 MW case with full conversion is further reduced to -62.3 Euro/t and -20.7 Euro/t,
422 respectively, from 20.9 Euro/t in Table 10. Furthermore, a maximum shipping cost of 20.8 Euro/t
423 for the 1350 MW full conversion case is estimated, (based on zero netback of bio-oil) implying a
424 maximum locus of radius within 3774 km from the UK. It is not cost-effective to ship bio-oil

425 across a long distance under the current economic scenario, albeit bio-oil from oil palm empty
426 fruit bunch achieves higher efficiency than poplar (Table 7(b)).

427 *3.3 Analysis of Performances*

428 A lower temperature FT reaction at 200°C which corresponds to a higher chain growth
429 probability, α of 0.87 (determined using equation (5)) was also studied, in addition to the base
430 case with FT reaction temperature at 240°C and α at 0.76. Higher C₅₊ selectivity of 79.4% and
431 yield of 63.5% can be achieved from lower temperature FT reaction, compared to the selectivity
432 of 66.3% and yield of 52.9% from 675 MW BOIG-FT system with once-through configuration
433 (Table 5). This suggests that the lower temperature FT reaction can have higher yields of diesel,
434 which is consistent with the observation mentioned in section 2.4.

435 The best case with 1350 MW capacity and full conversion configuration was taken as the
436 basis for further economic sensitivity studies. The relationship between the costs of production of
437 FT liquids and bio-oil is illustrated in Figure 8, by taking the prices of electricity of 46.61
438 Euro/MWh and 80.12 Euro/MWh for the years 2005 and 2009, respectively, as the basis [39].
439 The cost of production of FT liquids was predicted by subtracting the value of electricity from
440 the total unit annualised cost of 169.1 million Euro/y (Table 10) incurred by the 1350 MW
441 BOIG-FT system with full conversion configuration, including the following costs of bio-oil, (i)
442 0 Euro/t, (ii) 9.8 Euro/t (2005) and 20.9 Euro/t (2009) for a given cost of production of FT liquid
443 from wood-based FT plant: 42.6 Euro/MWh [38], (iii) 75 Euro/t and (iv) 150 Euro/t. The
444 reported cost of production of bio-oil is between 75-300 Euro/t [41]. Even for a minimum
445 reported cost of bio-oil of 75 Euro/MWh and an electricity price of 80.12 Euro/MWh in 2009,
446 the cost of production of FT liquids is expected to be at 78.7 Euro/MWh, which is higher than

447 that currently produced from coal and biomass, i.e. 29.2-43.15 Euro/MWh and 46 Euro/MWh,
448 respectively [5,42]. A lower electricity price, e.g. 46.6 Euro/MWh in 2005 further enhances the
449 cost of FT liquids to 86.1 Euro/MWh for the same cost of bio-oil of 75 Euro/MWh. If the
450 operating cost of BOIG-FT system can be reduced by 5-20% due to higher energy saving and
451 infrastructure sharing, the COP of FT liquid can be reduced by 0.9-3.7%. 2-8% reduction in COP
452 of FT liquid can be achieved by 5-20% reduction in capital cost. It has also been projected that a
453 reduction in capital by 69% can be achieved after 100th is built, due to technological learning,
454 based on the progress ratio of individual units given by Faaij et al. [43]. Also, economic
455 incentives may be created by versatile price structure for energy commodities through the UK
456 Renewable Obligation Certification, which currently offers a single electricity price from all
457 renewable resources.

458 **Figure 8**

459 **4. Conclusions**

460 This paper presents simulation based techno-economic analysis of centralised scale
461 BOIG-FT systems, using bio-oil as feedstock. Bio-oil has been modelled generically and
462 adequately, using three representative chemical components, i.e. acetic acid, acetol and guaiacol,
463 based on its performance through gasification. A comprehensive study has been followed
464 thereafter for deciding on operating conditions for individual processes as well as overall system.
465 Site-wide heat integration was applied for improving the efficiency of the overall system while
466 achieving cost savings through CHP generation. Case studies were performed for 1 MW, 675
467 MW and 1350 MW, with once-through and full conversion FT reactor configurations. Higher FT
468 conversion was preferable for obtaining higher C₅₊ liquid selectivity. This way, higher efficiency

469 and also lower cost of production of FT liquid can be attained. The 1350 MW BOIG-FT system
470 with full conversion configuration was found to result into the lowest cost of production of FT
471 liquid amongst all the cases studied. The import of bio-oil from other countries by shipping to
472 the UK adds 2-4 times more than the operating cost. This is therefore not feasible under the
473 current economic climate. The BOIG-FT system under research is considered to be a very
474 effective lignocellulosic biorefinery system and commercially deployable within the next 10
475 years. The capital cost of the BOIG-FT system is thus expected to decrease according to the
476 learning curve effect. It is envisaged that a more competitive cost of production of biofuels from
477 bio-oil can be realised in the future, in the light of increasing value of electricity and reduction in
478 capital and operating costs.

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482 **Nomenclatures**

A	Parameter in equation (5), with values of 0.2332 ± 0.0740
B	Parameter in equation (5), with values of 0.6330 ± 0.0420
F_p	Total molar flow rate of paraffin produced (kmol/s)
N	Carbon number
T	Temperature (K)
w_n	Weight fraction of Fischer-Tropsch product with carbon number n
$y_{i,S}$	Mole fraction of component i (components in product gas) obtained from simulation

$y_{i,R}$	Mole fraction of component i (components in product gas) from reference case
$y_{p,n}$	Mole fraction of paraffin with carbon number n produced
y_{CO}	Mole fraction of CO in Fischer-Tropsch feed stream
y_{H_2}	Mole fraction of H ₂ in Fischer-Tropsch feed stream
A	Probability chain growth

483

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