

Process Integration and Economic Analysis of Bio-Oil Platform for the Production of Methanol and Combined Heat and Power

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Abstract

Process to process material and heat integration strategies for bio-oil integrated gasification and methanol synthesis (BOIG-MeOH) systems were developed to assess their technological and economic feasibility. Distributed bio-oil generations and centralised processing enhance resource flexibility and technological feasibility. Economic performance depends on the integration of centralised BOIG-MeOH processes, investigated for cryogenic air separation unit (ASU) and water electrolyser configurations. Design and operating variables of gasification, heat recovery from gases, water and carbon dioxide removal units, water gas shift and methanol synthesis reactors and CHP network were analysed to improve the overall efficiency and economics. The efficiency of BOIG-MeOH system using bio-oil from various feedstocks was investigated. The system efficiency primarily attributed by the moisture content of the raw material decreases from oilseed rape through miscanthus to poplar wood. Increasing capacity and recycle enhances feasibility, e.g. 1350 MW BOIG-MeOH with ASU and 90% recycle configuration achieves an efficiency of 61.5% (methanol, low grade heat and electricity contributions by 89%, 7.9% and 3% respectively) based on poplar wood and the cost of production (COP) of methanol of 318.1 Euro/t for the prices of bio-oil of 75 Euro/t and electricity of 80.12 Euro/MWh, respectively. An additional transportation cost of 4.28-8.89 Euro/t based on 100 km distance between distributed and centralised plants reduces the netback of bio-oil to 40.9-36.3 Euro/t.

Keywords: gasification; methanol; polygeneration; bio-oil; low grade heat; biorefinery

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1. Introduction

Methanol is globally one of the most important chemicals as well as energy carrier with application to fuel cell vehicles, flexible fuel vehicles and biodiesel production. In recent times, the global demand for methanol has increased by over 43% from 31.4 million tonne in year 2001 [1]. Methanol is produced from syngas using a range of resources, including currently exploited natural gas [2-4], and also other resources such as coal [5,6] and biomass [7-9]. Comparative studies undertaken by Williams et al. [7], Katofsky [8] and Hamelinck and Faaij [9-11] have demonstrated lower energy efficiencies of 54-58% and higher COP of methanol of 14-17 USD/GJ based on biomass feedstock, compared to 67.5% and 6-14.5 USD/GJ from natural gas, respectively. With rapidly growing interest in the development and demonstration of biomass pyrolysis processes, bio-oil or pyrolysis oil can become a promising infrastructure-compatible intermediate for the production of methanol [12-15]. Fast pyrolysis process involves thermal decomposition reactions that occur in a few seconds, at modest temperature conditions (~500°C) and in the absence of oxygen or oxygen-deficient environment. The capital and operating costs of a biomass fast pyrolysis plant have been estimated to be 48.3 million USD and 9.6 million USD, respectively for producing 426 tons/day of bio-oil from 550 dry tons/day of wood chip [16]. A number of significant advantages exist in fast pyrolysis as a pretreatment step for converting biomass into liquids, ~8-12 times more bulk density; removal of char along with alkali metals; lower tar content; ease and lower cost in handling, storing and transporting liquids to the production as well as reception sites [12]. In addition, bio-oil has an advantage over crude oil during transportation, due to its inert and non-toxic properties and separation into heavy organic fractions that will sink, instead of spreading over the water surface [17]. Consequently, this will not cause severe environmental pollution during a spill or pipeline leakage, as in the incident of Gulf of Mexico oil spill in 2010 [18].

Alongside all these advantages, there exist flexibility in biomass selection for distributed generations of bio-oil and opportunities for industrial scale processing of bio-oil in centralised chemical or fuel production plants, such as, methanol synthesis [12,19]. Brammer et al. [15] and Bridgewater et al. [13] investigated into the commercial competitiveness of bio-oil systems for a range of applications, heat, power and CHP. The economic attractiveness depends on many factors, including the scale, the location and its associated economic and logistic factors, e.g. availability and price against competitive fossil resources and industrial practices etc. It is clear from their studies that the centralised applications of bio-oil enhance economic competitiveness. Building upon these concepts [12-15], this study explores process integration opportunities for the conversion of bio-oil from distributed pyrolysis plants as a commodity through a centralised methanol synthesis plant. The evaluation of commercial opportunities is undertaken by quantitative assessment of the number of commercial / demonstrated process units required. These include gasification, methanol synthesis and oxygen supply units that may be exploited in distributed and centralised scales respectively. The approach here was also to use a range of scales, 1 MW through 675 MW to 1350 MW based on thermal input of bio-oil to a specific application and carry out economic assessment and sensitivity to important cost factors, such as the price of electricity and the transportation cost of bio-oil (in the context of the UK and EU) [13,14,20] which determine the acceptable buy-in price of bio-oil for conversion.

Bio-oil can be resourced from wood such as poplar. The UK has a total poplar plantation area of approximately 11337 ha (0.01% of global): England (88.2%), Wales (4.8%), Scotland (4.3%) and Northern Ireland (2.7%), with mean yield of 7.3 odt/ha per year [21-23]. Due to the limited local supply of poplar wood, indigenous resources such as miscanthus and oilseed rape can be used. The plantation area of miscanthus in the UK has been reported to be 12700 ha, with yield of 9-10 odt/ha per year [24]. Oilseed rape has a plantation area of 85711 ha, with yield of 2.5-3.3 t/ha [24]. The process under consideration was based on poplar wood and further established for the above energy crop and agricultural waste.

The process-related elements in this work include facilitation of partial oxidation within bio-oil gasification to achieve thermally efficient performance in fluidised bed and entrained flow bed gasifiers [25-27], process design and variability studies for feed conditioning for methanol synthesis reactor, using heat integration strategy [7-9,28,29]. The effect of increased thermal efficiency of the gasifier is the reduction in mole fraction of hydrogen in the product gas from the gasifier, which can be compensated by exothermic water gas shift reactor (alternatively supplied from water electrolysis), operating at lower temperatures. In addition to achieving the desired molar stoichiometric number (SN) of the feed gas to methanol synthesis reactor, the degree of water and carbon removal from the product gas by incorporation of appropriate technologies on the yield of methanol and plant efficiency has been analysed. Emphasis was given on total site CHP network synthesis and hot water recovery and thereby improving overall site energy efficiency. The objectives of this paper were thus to investigate technical and economic feasibility of bio-oil based methanol synthesis technologies with a view in commercial deployment over short-term future.

2. Process Simulation and Sensitivity Studies

Two BOIG-MeOH process configurations with cryogenic ASU and water electrolyser that comprised of gasification, gas cleaning and processing, methanol synthesis and power generation are depicted in Figures 1 and 2, respectively. The process operating variables specified in ASPEN Plus simulation are presented in Table 1.

Figure 1

Figure 2

Table 1

Table 2 summarises the technology developers and capacities of the main units [30-33]. A single state-of-the-art pyrolyser unit of 200 t/d throughput of biomass produces 150 t/d of bio-oil [12]. A single train entrained flow gasifier [30] can process 2000 t/d of bio-oil from 14 such

pyrolysis plants, to produce ~600 t/d of methanol. An oxygen-blown gasifier is opted since air will lower the heating value of the resulting product gas. The oxygen requirement for single train gasifier is 860 t/d, requiring one train cryogenic ASU unit [33]. A Lurgi methanol synthesis reactor [31] thus may produce 2270 t/d of methanol from 7290 t/d of bio-oil from 49 pyrolysis units, and 4 integrated gasification and ASU units, in a centralised 1350 MW (thermal input of bio-oil) BOIG-MEOH process.

Table 2

Bio-oil was modelled using three representative components, i.e. acetic acid, acetol and guaiacol, indicated in Table 3, while RGibbs reactor (works on Gibbs free energy minimisation principles) was used to model the gasifier (Table 1). The simulation results of the bio-oil gasification were validated against literature results [34], with residual sum of square (RSS) obtained within an acceptable range (Table 3). Water electrolyser modelled as another RGibbs reactor (Table 1), is operated at 130°C and 30 bar [34] (Figure 2). Nitrogen generated from the cryogenic ASU (Figure 1), modelled as a component separator, Sep (Table 1), can be used for purging the reactor vessels or can be sold.

Table 3

A thermally neutral performance of the oxygen-blown entrained flow bio-oil gasifier (stream 1 to GASIFIER in Figures 1-2) can be attained at ~1100°C. This is achieved by preheating oxygen (stream 4) feed from ASU to a temperature of 630°C [35] using very high pressure (VHP) steam generated by the heat recovery from the product gas of the gasifier. This way heat recovered at a lower level can be utilised in supplying endothermic heat of reforming reactions at a higher gasifier temperature. Bio-oil may not need to be preheated so as to prevent degradation and formation of char before entering the gasifier [26,36]. The product gas (stream 6) comprising of CO, CO₂, H₂ and H₂O as the main components, is likely to be free from tar, ash, nitrogen and sulphur, as bio-oil is relatively clean [12].

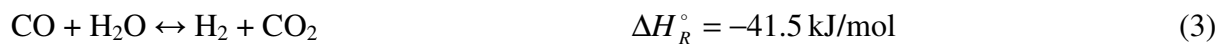
Subsequently, heat from the product gas was recovered into the generation of VHP steam (SYNGCOOL). A water-gas shift reactor (WGS in Figure 1) was needed prior to gas cooling for water removal in the ASU configuration. These are mainly to attain an appropriate molar stoichiometric number (SN): $(H_2 - CO_2) / (CO + CO_2) = 2$ [4,8] in the feed gas to the methanol synthesis reactor. In electrolyser configuration (Figure 2), the stoichiometric ratio was met by supplying H_2 from the electrolyser, requiring no WGS reactor. In both configurations, the product gas was cooled to just above its dew point for water removal (H2OREM). The dry gas has a much enhanced performance in the methanol synthesis reactor, analysed later. The water removed can be sent to wastewater treatment plant, comprising of physical, chemical or biological processes, and recovered as boiler feed water after 10% purge [37], the cost of which was considered later. CO_2 in the product gas is removed using Sulfinol unit (combined physical and chemical solvent process) (CO2SEP) by 99% and 85% on molar basis, in electrolyser and ASU configurations respectively [38]. The CO_2 captured was compressed through CO2COMP to supercritical CO_2 at 80 bar, suitable for storage [39]. Subsequently, the clean syngas was compressed through SYNGCOMP to 100 bar prior to the methanol synthesis reactor (METHANOL).

A Lurgi isothermal and gaseous phase fixed bed methanol synthesis reactor can be suitable for the operating conditions provided in Table 1. The sensitivity analyses of methanol synthesis in terms of temperature and pressure conditions were performed, presented in the following section. The pressure of the gaseous product stream from METHANOL was brought down from 100 bar to 40 bar, by expanding through SYNGEXP (Figures 1 and 2). The feed gas compressor and the product gas expander can be run on the same shaft. A minimum outlet pressure of 40 bar was achievable while keeping the outlet at gaseous phase. A flash column (METSEP) was further used to separate the gaseous and liquid products at 40°C and 24 bar [29]. 98% recovery (molar basis) of methanol in the liquid product (stream 21) from METSEP can be achieved. The offgas containing unreacted gases such as H_2 , CO , CH_4 etc. (stream 20) was fed to the gas turbine (GASTURB) for power generation.

A trace amount of natural gas (stream 22) was required for achieving a stable operating region and to refrain from modification of the gas turbine combustor [40]. The resulting Wobbe Index of the gas turbine (measure of interchangeability of fuel gases and for comparing the combustion energy among fuel gases with different compositions) was validated against industrial study by Shah et al. [41], in order to ensure that it was within $\pm 10\%$ of the stable combustion region. Air was supplied to the gas turbine combustor (GTCOMB), after compression to 14 bar via AIRCOMP. GTCOMB, operating at 1200°C and 14 bar, did not require de-rating (reduction in burning temperature and compressor power ratio to allow a stable combustion), because the LHV of the gas feed into the combustor was found to be greater than 6 MJ/m^3 [38,40]. The exit temperature and pressure of the exhaust gas from the GASTURB are approximately 740°C and 2 bar, respectively. The heat content of the exhaust gas from GASTURB was recovered into the generation of VHP steam in heat recovery steam generator (HRSG).

2.1 Sensitivity Studies of Methanol Synthesis Reaction

The principal reactions involved in the methanol synthesis reactor, modelled as a stoichiometric reactor (RStoic) in ASPEN Plus, are provided in equations (1)-(3) [2]. Methanol synthesis reactors operate within 50-150 bar pressure and 230°C - 270°C temperature, in the presence of Cu/Zn/Al catalyst [10].



Low temperature and high pressure favour the reactions, according to Le Chatelier's principle. The temperature and pressure effects on the components' mole fractions are illustrated in Figure 3(a) and (b), respectively. The efficiency is increased from 50 bar, 300°C (20.9%) through 100 bar, 250°C (48.3%) to 150 bar, 210°C (67.4%). However, at a very low temperature the rate of

reaction is very slow, while a very high pressure necessitates high compression energy and contemplation for equipment for safety reasons. Therefore, to attain the required rate of reaction and moderate compression energy, 250°C and 100 bar are recommended [10].

Figure 3

The ideal value of *SN* is 2, however a *SN* of slightly higher than 2 is used in practice, in order to control the by-product formation [3]. *SN* greater than 2 is attained by higher hydrogen and lower CO₂ molar fractions in the syngas. The *SN* must be adjusted prior to the methanol synthesis reaction to ensure higher conversion and selectivity to methanol. A small amount of CO₂, 1-2%, that may act as a promoter for the primary reaction (equation (1)) and help in maintaining catalyst activity, is also required in the feed gas [8,10].

Water in the feed stream to methanol synthesis reactor is undesirable due to its dilution effect and facilitation of water-gas shift reaction (equation (3)), resulting in CO₂ instead of methanol. *SN* does not account for the molar ratio of water in the feed. Figure 4 indicates that the removal of water from 0% to 90% on molar basis results in an increase in carbon efficiency, defined in equation (4), from 38.7% to 72.5%, while *SN* remains unchanged at 2.1. Thus, higher degree of water removal is desired alongside higher carbon efficiency, hence moles of methanol produced, whilst conforming to the specification of *SN*.

$$\text{Carbon efficiency (\%)} = \frac{\text{moles of methanol produced}}{\text{moles of carbon oxides (CO + CO}_2\text{) in feed}} \times 100\% \quad (4)$$

Figure 4

Furthermore, Table 4 presents a sensitivity analysis of the effect of temperature of WGS and the degree of CO₂ removal on *SN* for ASU configuration (Figure 1). The water-gas shift reaction in equation (3) is favoured at low temperatures, thereby improving the H₂/CO molar ratio. A WGS operating temperature of 450°C, which provides reasonable performance of its catalyst

and reaction, and the removal of CO₂ by 85% on molar basis, can attain a *SN* of 2.03 and CO/CO₂ molar ratio of 4.14 in the feed gas.

Table 4

3. CHP Network Design and Energy Efficiency

The supply and the target temperatures as well as the heat duties of the streams and process units were extracted from the integrated flowsheet simulations, shown in Tables 5(a) and (b) for ASU and electrolyser cases in Figures 1 and 2, respectively. Heat utilisation and consumption were classified based on the heat duties / temperature levels, in Tables 5(a) and (b), respectively [42,43].

Table 5

The VHP and MP steam that can be generated by the maximum recovery of the sensible heat from SYNGCOOL and HRSG and the product gas of WGS (HE2 in Figure 1), was estimated using the composite curve analysis, within the specified minimum approach temperature, 20°C [43,44]. The targeted steam main levels are VHP steam (100 bar, 650°C superheated), MP steam (15 bar, 200°C saturated) and LP steam (5 bar, 152°C saturated), in line with industrial processes [45]. The analysis is presented in Figures 5(a), (b) and (c), respectively. Isothermal conditions in WGS and METHANOL were maintained by extracting their respective exothermic heat of reactions into MP steam generation (using enthalpy balance) on the shell side of the reactor.

Figure 5

The heat demands were placed by O₂ preheater (HE1), HE4 (in ASU configuration), Sulfinol unit and methanol distillation unit. The VHP steam was used in HE1 and HE4 for preheating based on composite curve analysis. The steam requirements by the Sulfinol and methanol distillation units, not simulated in ASPEN Plus, were estimated using industrial data, 1.42 kg of LP steam/kg of acid gas and 0.45 kg of LP steam/kg of methanol, respectively [46,47].

The CHP network is depicted in Figure 6. The back-pressure steam turbines ST1 and ST2 are operated between VHP and LP steam levels and ST3 between MP and LP steam levels respectively. The condensing turbine, ST4 was further used to generate power after fulfilling the LP steam requirement by Sulfinol and methanol distillation units. 80% isentropic and 95% mechanical efficiencies were assumed.

Figure 6

The energy balance around two systems, comprising of the process site (Figures 1 and 2) and the CHP network (e.g. Figure 6), was analysed for 1 MW, 675 MW and 1350 MW capacities with once-through mode, and recycle mode (90%) for 1350 MW capacity, in Tables 6 and 7, respectively. The recycle mode refers to the recycling of the offgas from the methanol synthesis reactor (stream 20 in Figures 1 and 2). It leads to higher production of methanol, in relation to the generation of power. The power requirement by the cryogenic ASU unit was estimated based on 235 kWh/t O₂ requirement by the site (Figure 1) [48]. The power requirement by the water electrolyser unit was estimated to be 4049 kWh/t in ASPEN Plus simulation (Figure 2).

Table 6

Table 7

The efficiency of an overall site was calculated based on the LHV of the bio-oil and the production of methanol and electricity, for both with and without the generation of low grade heat (Tables 6 and 7). The LHV of methanol was assumed to be 20 MJ/kg. In the ASU configuration in Figure 1, the surplus low grade heat included the heat from HE3, HE5, excess heat from HE2 and heat released from the condenser (after condensing turbine ST4). The heat available from HE2, HE3, excess from SYNGCOOL (after the heat recovery using composite curve analysis) and condenser (after condensing turbine ST4) provided the low grade heat for the electrolyser configuration. The surplus low grade heat from the site can be recovered into hot water generation, which is particularly important in the current EU scenario where solar energy is targeted for

domestic hot water generation. The electrolyser configuration consumes 3.6-5.3 times of power higher than the ASU configuration. Obviously, this is attributed to the high power requirement by the water electrolyser unit, 75-79% of the total power consumption on site, whilst the ASU only consumes 15.6-24.1%, respectively. The electrolyser configuration with recycle mode produces more methanol, compared to equivalent ASU cases; however at a cost of additional power requirement. The net energy efficiency of the sites increases with increasing capacity, due to higher proportional increment in power generation from syngas expander and gas and steam turbines from 0.14 MW through 117.4 MW to 276.2 MW, for 1 MW, 675 MW and 1350 MW cases (ASU configuration), respectively.

The analysis of bio-oils derived from different biomass sources, poplar [34], miscanthus [49] and oilseed rape [50], is presented in Table 8(a). The heating values, chemical composition and moisture content have strong influence on the efficiency of BOIG-MeOH system. The efficiency increases from poplar (48.3%) through miscanthus (66.9%) to oilseed rape (68.2%), respectively (Table 8(b) presented for 1350 MW BOIG-MeOH with ASU configuration). The net energy efficiency is thus strongly influenced by their moisture contents, 30%, 10.1% and 9.5% more than their carbon and hydrogen wt%, 63%, 72.5% and 89.5%, respectively in Table 8(a).

Table 8

4. Economic Analysis

The objective of the detailed economic analysis was to evaluate the netback of the bio-oil feedstock and the COP of methanol, and their sensitivity with respect to the price fluctuation of certain components, capital, operating and transportation costs.

4.1 Capital Costs

The capital cost of the processes under consideration (Figures 1 and 2) has been evaluated in terms of the direct and indirect capital costs. The inside battery limit (ISBL) cost data, the base costs, scale factors and base scales, of the major process units, the outside battery limit (OSBL)

costs and the indirect capital costs were estimated based on the works of Tijmensen et al. [38] and Hamelinck and Faaij [9], in Table 9. A scale factor, R was applied in the cost and size correlation in equation (5) [51], for evaluating various sizes of individual equipment, unless otherwise specified. Multiple units were considered if a desired capacity of a unit exceeds its maximum size, e.g. 400 MW HHV for the gasifier [38]. It was assumed that the Sulfinol unit accounted for 10% of the total capital cost, that is a proprietary information [38]. The cost of ASU was estimated based on the parameters provided in Table 9 (item 8) and using equation (5), while that of the water electrolyser was estimated based on 7500 SEK(Swedish Krona)/kW, Saxe and Alvfors [52], which is equivalent to 825 Euro/kW in 2007.

$$\frac{COST_{size2}}{COST_{size1}} = \left(\frac{SIZE_2}{SIZE_1} \right)^R \quad (5)$$

$SIZE_1$ and $COST_{size1}$ represent the capacity and the cost of a base unit, whilst $SIZE_2$ and $COST_{size2}$ represent the capacity and the cost of the unit after scaling up/down, respectively.

Table 9

The capital costs of individual equipment were further levelised to year 2009 by applying a cost index (equation (6)), adopted from Chemical Engineering Plant Cost Index (CEPCI) [53].

$$\text{Present cost} = \text{Original cost} \times \left(\frac{\text{Index at present}}{\text{Index when original cost was obtained}} \right) \quad (6)$$

Figure 7 illustrates the breakup of capital costs of individual equipment in a BOIG-MeOH configuration. The oxygen supply units remain one of the major contributors to the total capital cost, 18.6% and 52.5% for ASU and electrolyser configurations respectively. In addition to lower overall efficiency (albeit higher methanol production), the electrolyser configuration results in 65.2% higher capital cost than that of ASU configuration.

Figure 7

4.2 Operating Costs

The fixed operating costs were estimated based on a percentage of the total indirect capital costs (TIC), given in Table 10. The cost of the personnel was estimated based on the study by Tijmensen et al. [54]. The fixed operating costs were evaluated according to the work of Sinnott [55]. Variable operating costs include the costs of natural gas [56] considered in all cases, electricity [56] in electrolyser configuration and steam [55] in 1350 MW, recycle mode ASU configuration, respectively. Finally, the percentage of miscellaneous cost was adopted from the work of Sinnott [55].

Table 10

4.3 Discounted Cash Flow, Netback of Bio-oil and Cost of Production of Methanol Analysis

An instance of discounted cash flow analysis is presented that provides an annualised capital charge of 13% for a discount rate of 10% and 15 operating years. The start-up period is 2 years, with 25% and 75% of the total capital cost were distributed in the -1^{st} and 0^{th} year (0^{th} year indicates the plant start-up year) [54].

The netback indicates the value of a feedstock obtained from selling its products at their market prices and is determined using expression (7) [43]. The netback thus sets the maximum acceptable buy-in price of a feedstock. The market price of the feedstock thus must be less than this price for an economic processing.

$$\text{Netback} = \text{Value from products} - (\text{Annualised capital cost} + \text{Annual operating cost}) \quad (7)$$

The contract price of methanol in the European region as posted by Methanex is 250 Euro/t (valid from 1^{st} April to 30^{th} June, 2010) [57]. The price of electricity was adopted from DECC [56], reported at 7.284 pence/kWh (2009) (equivalent to 80.12 Euro/MWh, assuming 1 GBP = 1.1 Euro), excluding the Climate Change Levy (CCL). The CCL only applies to industrial sectors where taxable supplies such as electricity, coal and petroleum are charged [58]. The full rate of CCL for

the electricity is reported at 0.47 pence/kWh (equivalent to 5.17 Euro/MWh) [56]. A comparison of annualised capital charge, annual operating cost and the value of products and the netback of bio-oil (both, with and without CCL) between the ASU and electrolyser configurations for 1 MW, 675 MW and 1350 MW capacities is presented in Table 11. The COP [59] of methanol was calculated using equation (8), based on the price of bio-oil at 75 Euro/t [60], and the cost of electricity at 80.12 Euro/MWh in year 2009 [56].

COP of methanol

$$= \frac{\text{Annualised capital cost} + \text{Annual operating cost} + \text{Cost of bio - oil} - \text{Cost of electricity generated}}{\text{Annual production of methanol}} \quad (8)$$

Table 11

Higher netback (maximum acceptable buy-in price) of bio-oil and lower COP of methanol are desired. The netback of bio-oil for the best two cases, 1350 MW ASU configurations with and without recycle, is 45.2 Euro/t and 6.65 Euro/t, without the consideration of CCL, respectively. The COP of methanol from these two cases is 318.1 Euro/t and 469.3 Euro/t respectively. The inclusion of CCL has a negligible impact on economics, especially for the economically attractive cases, e.g. 0.4 Euro/t for 1350 MW ASU with recycle configuration. The generic trend is the increasing economic feasibility with increasing capacity and by the introduction of recycle mode. None of the electrolyser cases is economically feasible. The capital cost, operating cost and the cost of bio-oil (assuming price of bio-oil at 75 Euro/t, without incorporating transportation cost) contribute to 31.1%, 17.5% and 51.4%, respectively to the total annual investment of the economically attractive case, i.e. 1350 MW ASU with recycle configuration.

4.4 Transportation Cost of Bio-Oil

The transportation cost of bio-oil is an important factor to analyse, when bio-oil produced from distributed pyrolysis plants needs to be transported to centralised BOIG-MeOH process. The data and analysis of bio-oil transportation from distributed pyrolysis plants to centralised sites,

based on the studies by Bridgwater et al. [13], Rogers and Brammer [14] and Pootakham and Kumar [20], presented in Table 12 are applicable to the UK and EU contexts. The cost of transporting 303.75 t/h bio-oil (equivalent to 2.43 million t/y; or 2.24 million m³/d; or 3.89×10⁷ GJ/y) to 1350 MW BOIG-MeOH site was estimated using the distance rate [13,20] and zone costing approaches [14]. The results provided in Table 12 used both fixed and variable costs depending upon the distance or thermal value of bio-oil [13,14,20]. (Note: 1 USD = 0.8 Euro and 1 GBP = 1.1 Euro were assumed). The zone costing approach uses the number of round trips in a day within the distributed-centralised region as the basis to define a transport zone. Thus, zone 1 is the outermost zone where only one round trip is possible in a day, carrying the highest total cost amongst all zones, 64.2 million Euro/y or 26.42 Euro/t, compared to 8.6 million Euro/y or 3.54 Euro/t for zone 6 that implies 6 round trips in a day, respectively. It is hence beneficial to implement more round trips in a day so as to reduce the transportation cost. Tanker (truck) with load ranges of 24-44 tonnes and 60 m³ per truck and pipeline capacity of 560 m³/d were considered for a distance of 100 km between distributed pyrolysers and centralised BOIG-MeOH sites. The resulting transportation cost is 4.28-8.89 Euro/t (or 10.4-21.6 million Euro/y for 1350 MW case) that reduces the netback of bio-oil from 45.2 Euro/t (Table 11) to 40.9-36.3 Euro/t. In terms of COP of bio-oil, transportation adds 5.7-11.9% extra on 75 Euro/t.

Table 12

By incorporating the transportation cost of bio-oil over a distance of 100 km, the netback of bio-oil in 1350 MW recycle and ASU configuration, would further be reduced to 40.9-36.3 Euro/t (compared to 45.2 Euro/t in Table 11) at the least, using the same basis.

4.5 Analysis of economic feasibility

The effect of market price of methanol on the netback of bio-oil was examined for the following prices of methanol, 106.7 Euro/t, 250 Euro/t and 299.6 Euro/t, and two sets of electricity price, 46.61 Euro/MWh (2005) and 80.12 Euro/MWh (2009) [56], for 1350 MW BOIG-MeOH

system with ASU and recycle configuration, in Figure 8(a). The estimated COP of methanol from natural gas is 4.7 Euro/GJ, which is equivalent to 106.7 Euro/t [61]. Hamelinck and Faaij [9] estimated the COP of methanol of 8.6-12 USD/GJ (equivalent to 214.7-299.6 Euro/t) from biomass. The maximum netback of bio-oil thus obtained is 66.9 Euro/t for the given maximum market prices of methanol and electricity. Based on the illustration in Figure 8(a), the minimum price of methanol is 153.2 Euro/t and 146.9 Euro/t for the price of electricity of 46.61 Euro/MWh and 80.12 Euro/MWh, respectively, corresponding to zero netback of bio-oil.

Figure 8(b) demonstrates the effect of the cost of bio-oil on the COP of methanol, for 0 Euro/t, 75 Euro/t and 150 Euro/t of the cost of bio-oil. The acceptable range of the cost of bio-oil is between 0-75 Euro/t; whilst 150 Euro/t of the cost of bio-oil may provide flexibility in the distance between distributed and centralised plants. An increase in the electricity price by 33.51 Euro/MWh from 2005 to 2009 can reduce the COP of methanol by 6.3 Euro/t. Escalation in the price of electricity and methanol from renewable resources in future can thus stimulate deployment of these systems.

Figure 8

If 5% reduction in the operating cost of BOIG-MeOH system is achieved, 0.9% lower COP of methanol than that reported in Table 11 can be attained. 20% reductions in operating cost result in 2.8% lower COP of methanol. Reduction in capital cost brings more significant improvements, e.g. 5% and 20% reductions in capital cost lower the COP of methanol by 1.6% and 5%, respectively. Following the assumptions in the works of Faaij et al. [51], the capital cost of 1350 MW ASU and recycle configuration can be reduced by 54.4% in the short-term, e.g. 2020, after 20th plant is built (838.2-382.4 million Euro/y), due to technological learning. After 100th plant is built, 69.1% reductions in capital cost or 242 Euro/t of COP of methanol compared to 318.1 Euro/t for the first plant are achievable. It may be noted that the learning effects imply projections based on today's value of a given currency.

5. Conclusions

The techno-economic feasibility of BOIG-MeOH system has been analysed using process variability and integration approaches. The performance of methanol synthesis reactor in terms of the yield of methanol was analysed with respect to the stoichiometric number in the feed gas and water removal, with the latter being very critical to achieve higher yield of methanol. The electrolyser configuration can achieve higher amount of methanol production compared to the ASU configuration. This is attributed to additional hydrogen supply from water dissociation, requiring no water-gas shift reaction, however, at the cost of higher electricity and capital cost requirements by the water electrolyser unit. An energy efficient gasification process has been integrated in the context of overall CHP generation network that exploits heat recovery from the syngas cooler, HRSG, WGS and methanol synthesis reactor into the generation of surplus steam and thereby steam turbine power, oxygen preheating to avoid high temperature heat of reforming reactions within gasification and low grade heat recovery. The cost competitiveness and energy efficiency of the systems increase with increasing capacity and by the introduction of recycle. Between ASU and electrolyser configurations, the latter configuration is not economically feasible at this stage. The bio-oil gasification route provides more flexibility in terms of biomass selection and pyrolysis in distributed generations and processing of bio-oils using commercially available centralised scale gasification and methanol synthesis reactor, etc. technologies. This led to the performance analyses of centralised BOIG-MeOH system based on various biomass feedstocks, oilseed rape, miscanthus and poplar wood, resulting in highest to lowest energy efficiency obtained. These variations can be attributed to the characteristics of bio-oil, particularly the moisture content. The netback of bio-oil can be reduced with the incorporation of the transportation cost between the distributed pyrolysis and centralised BOIG-MeOH plants. A more competitive industrial scale process may be realised with rising price of electricity and reduction in capital and operating costs.

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Nomenclatures

R Scale factor, equation (5)

SN Stoichiometric number for the feed to methanol synthesis

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