

Techno-economic analysis of a peatland gasification process for methanol synthesis and power production

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Abstract

In 2015, uncontrolled burning of tropical peatland in Indonesia led to a noxious haze that plagued the region. This paper aims to prevent the deadly haze from its source by harvesting the peat and utilizing it in a controlled manner. A conceptual design of a peat gasification process to produce electricity and methanol is proposed. Five selection criteria, namely, methanol yield, reduction in CO₂ emissions, usable energy, capital cost and payback period are used to compare four design alternatives for a peat gasification process. Two types of gasifiers (i.e., updraft fixed-bed and dual fluidized-bed) and two types of methanol synthesis reactors (i.e., gas-phase and liquid-phase) are paired to create four design alternatives that are modeled in Aspen Plus. The chosen design is then refined with detailed modeling of the power generation section and energy integration. The final design consumes 1,000 tonne peat/day, produces 214 tonne methanol/day, and generates 56 MW of electricity. Furthermore, GHG emissions are reduced by 22.4%.

Keywords

Techno-economic, Peat Gasification, Aspen Plus simulation, Methanol synthesis

1. Introduction and background

In recent years, Indonesia has witnessed its ongoing haze crisis dramatically worsen. By definition, haze is an atmospheric phenomenon mainly associated with the encounter of sunlight with pollution particles (EPA, 1999). Problems arise from the fact that some of the pollutants that form haze are linked to severe health complications and environmental damage. In Indonesia, the haze is perpetuated by the uncontrolled burning of the nation's tropical peatlands. Historically, this haze was only a domestic issue, but a palm oil boom beginning three decades ago has made the issue transboundary (Carlson, et al., 2012; Varkkey, 2012). Most recently, in 2015, the noxious haze intensified in part due to El Niño and Indonesia's daily emissions skyrocketed (Harris, et al., 2015). Much of Southeast Asia was also blanketed for months.

The burning of peatland is primarily rooted in agribusiness. Every year, there is mounting evidence of illegal peat and forest fires (also known as "hotspots") on properties for commercial oil palm plantations (Varkkey, 2012; Sizer, et al., 2013; BBC, 2015), meaning that the fires are intentional acts of deforestation. This is said to be caused by two main parties: corporations and independent farmers. The 2009 Law for Protection and Management of Environment allows two hectares per head of family to be burned for planting local vegetation (PR Indonesia, 2009). However, illegal burning is on the rise and neither party is taking ownership of their actions. There has also been a tendency for the Indonesian government to deflect criticisms of inaction and this has sparked mass controversy particularly within the Association of Southeast Asian Nations (ASEAN). A comprehensive review of literature reveals that some research has been done in policy development and overhaul (Varkkey, 2013; Atkinson, 2014), but engineering solutions are not emerging from either the academic space or industry. There is a pressing need for innovative designs to address the haze crisis.

Without significant action, the haze crisis is set to worsen as plantations expand and underground fires linger during long dry seasons (BBC, 2015). The true economic impact of the haze crisis is difficult to quantify and values

circulating through global media cover a wide range of estimates. By way of example, Dr. Herry Purnomo at the Centre for International Forestry Research (CIFOR) estimated that US\$4bn were lost in terms of “agriculture production, destruction of forest lands, health, transportation, tourism and other economic endeavours” as of September 2015 (Koswaraputra, 2015). The World Bank assigned a higher value of US\$16bn (The Guardian, 2015). Purnomo also estimates that at least US\$10bn is needed to cope with unending forest fires in some of Indonesia’s worst affected areas (i.e., its hotspots). President Joko Widodo’s administration has also provided early estimates totalling 300 trillion (~US\$22bn) to 475 trillion rupiah (~US\$34bn) to mitigate the crisis (Chan, 2015). Although these values vary, it is evident that tackling this issue is a large financial endeavour.

It is also evident that corporations and rural farmers are not being incentivized adequately or engaged actively. This likely contributes to a lack of public and private interest in design-based engineering solutions to the haze problem. This is reflected by the lack of literature sources in that space. Additionally, it is difficult to make a true comparison of the costs of mechanized land clearing versus swidden agriculture due to the various private and social costs involved. For instance, Varma (2003) attempted to analyze the costs and benefits of the slash-and-burn technique during Indonesia’s haze crisis in 1997 (Varma, 2003). He estimated losses of US\$20.1bn, but this was met with doubt from others in the academic community (Tacconi and Vayda, 2006). Nevertheless, Purnomo from CIFOR, simplifies the matter in a startling light: mechanized land clearing costs roughly two million rupiah (~US\$145) per hectare, whereas slash-and-burn costs about US\$7 per hectare.

Indonesia has the fourth largest area of peatland in the world, following the Russian Federation, Canada, and the United States of America (World Energy Council, 2013). With over 20.7 million hectares, Indonesia’s peatlands are the most extensive in the tropical zone. They span several regions of the country, but are located largely in the sub-coastal lowlands of Kalimantan and Sumatra. The heat of combustion, proximate analysis, and ultimate analysis for peat are required to fully characterize the peat for process decision. This information can be amalgamated from a variety of resources (Lucas, 1982; Andriesse, 1988; Rice and MacCarthy, 1991; Usup, et al., 2004; Chen and Agus, 2010). A summary of normalized parameters for Indonesian peat are provided in Table 1.

Table 1. Parameters for Indonesian peat.

Calorific Value (dry basis)	20.3 MJ/kg		
Proximate Analysis		Ultimate Analysis (dry basis)	
Moisture Content	83.1 wt.-%	C	48.62 wt.-%
Fixed Carbon (dry basis)	33.7 wt.-%	H	4.86 wt.-%
Volatile Matter (dry basis)	53.6 wt.-%	N	0.93 wt.-%
Ash (dry basis)	12.7 wt.-%	Cl	0.01 wt.-%
		S	0.16 wt.-%
		O	32.72 wt.-%

Historically, peat has been used as a fuel for heating and power production either by itself or co-fired with coal (Boron, et al., 1987). Its usage has declined over the past decades due to environmental concerns over peat harvesting practices and the introduction of cheaper and more widely available fossil fuels. However, ongoing peat production and consumption still continues today with most activity concentrated in Europe (Ramo, 2013; World Energy Council, 2013). A relatively small area of global peatland has been used to extract peat for energy. In the 1980s, a feasibility study was carried out regarding the use of peat for electricity generation in central Kalimantan, but no project resulted (World Energy Council, 2013). There are some exceptions, one example being a small peat-fired power plant that once operated in southern Sumatra. With the rise of biomass utilization technologies, there is growing potential for repurposing peat and the ongoing Indonesian haze crisis presents an urgent need to prevent further uncontrolled burning of peatlands. One solution is to harvest and utilize the peat to create both energy and chemicals. Two potential technologies for such purposes, namely, anaerobic digestion and gasification were explored in detail in the early stages of this project. Ultimately, gasification was selected for further study.

The fundamental output from the gasification process is synthesis gas (hereafter referred to as syngas), which is rich in hydrogen and carbon monoxide. As a fuel, it can be used for heating or producing electricity through conventional power generation configurations. Syngas is also a useful precursor that can be used in the synthesis of valuable fuels and chemicals. Some potential products include methanol, ethanol, hydrogen gas, ammonia, fertilizer,

and liquid fuels. In general, Indonesia's heating needs are not as extensive as their electricity needs. This is in part due to its geographic position in a tropical climate region; average temperatures in Indonesia are in the range of 28-30 °C annually (NWS, 2016). The need for residential heating is therefore relatively small. Heat demand still exists in industrial sectors and this energy demand must be met by local energy potential locations. In other words, heat sources need to be close to where they are utilized (World Energy Council, 2013). Electricity distribution is not as limited as heat distribution in this regard. From an electricity perspective, generation capacity growth in Indonesia cannot keep up with demand growth (EIA, 2015). This leads to power shortages and a low electrification ratio as compared to countries with similar income levels. By way of example, the Papua province only provides electricity to 43% of its population. Through synthesis processes and fermentation processes, syngas can be converted to numerous chemical products. Four in-demand chemical commodities, namely, methanol, ethanol, hydrogen, and ammonia were explored in detail in the early stages of this project. Other products such as fertilizers and liquid fuels have not yet reached the same commercial depth or viability as those mentioned previously. Ultimately, methanol was selected for further study.

Energy demand is quickly increasing in developing countries like Indonesia (Figure 2). The main consumers are transportation and industrial sectors followed by residential and commercial sectors (Hasan, Mahlia, and Nur, 2012). Most of the energy supply comes from crude oil, biomass, natural gas and coal with some coming from geothermal and hydropower. However, coal and the other non-renewable resources account for 75% of final energy consumption (Hasan, Mahlia, and Nur, 2012). Coal is the main competitor of biomass energy production as it is cheap and abundant domestically. Moreover, Indonesia has an estimated 21.13 billion short tons of coal reserve potential (Hasan, Mahlia, and Nur, 2012). However, there is large potential for Indonesia to sustainably collect and convert biomass to energy. Suntana, et al. (2009) found that more than 10 million Indonesian households or about 12,000 villages could be supplied annually with electricity from biomass. Additionally, Indonesia's National Energy Policy introduced a number of important changes in 2014 that included an ambitious plan to modify the energy mix to 30% coal, 22% oil, 23% renewable resources, and 25% natural gas by 2025 (IEA, 2014). Most notably, renewables are being promoted for greater use.

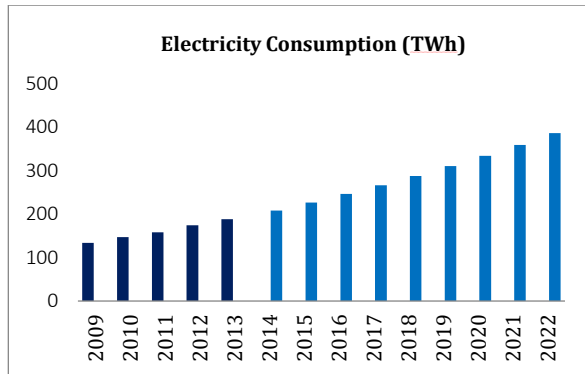


Figure 2. Indonesia's electricity demand [actual: 2009-2013, projected: 2014-2022] (IEA, 2013).

Methanol is a global commodity with production in Asia, North and South America, Europe, Africa and the Middle East. While there is an increasing demand for methanol in chemical applications, there is a stronger increase in demand for methanol in energy applications (Boyd, 2015). This is largely driven by the increase in demand for cleaner energy (Methanol Institute, 2011). Methanol itself can be used in applications such as transportation fuel, wastewater denitrification, fuel cell hydrogen carrier, biodiesel transesterification, or electricity generation (Methanol Institute, 2011). Methanol is also positioned as a fundamental raw material due to its versatility as a feedstock. One study shows that methanol could be used in Indonesia to help meet its gasoline needs while reducing carbon emissions by 8-35% (Suntana, et al., 2009). China is a particularly promising market for methanol based on growth trends in the 2000s. This is in part due to the rising demand for fuel blending (Yang and Jackson, 2012). According to Boyd (2015), Asia as a whole will represent 70% of global methanol demand by the end of 2016. Within that, China will represent 80% of the global demand growth. More recent reports have stated that China meets more than 10 % of its demand through imports of which 12 % is from Southeast Asia (Agarwal, 2015).

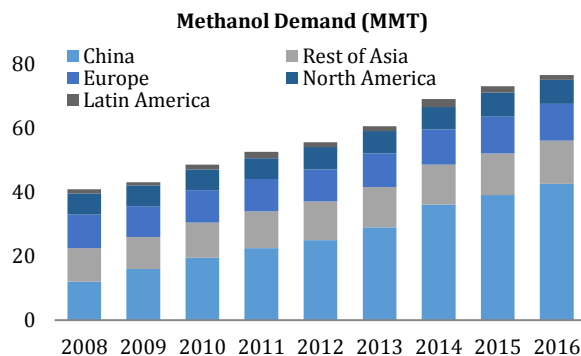


Figure 3. Global methanol demand [actual: 2008-2012, projected: 2013-2016] (CMAI, 2012)

2. Simulation and Results

A base design is constructed and various technologies involved in a polygeneration biomass gasification process are explored. Four design alternatives are formulated and modelled in Aspen Plus (Dyment, Reimers and Beck, 2013). The design alternatives are evaluated against a set of selection criteria and one alternative is selected for further development.

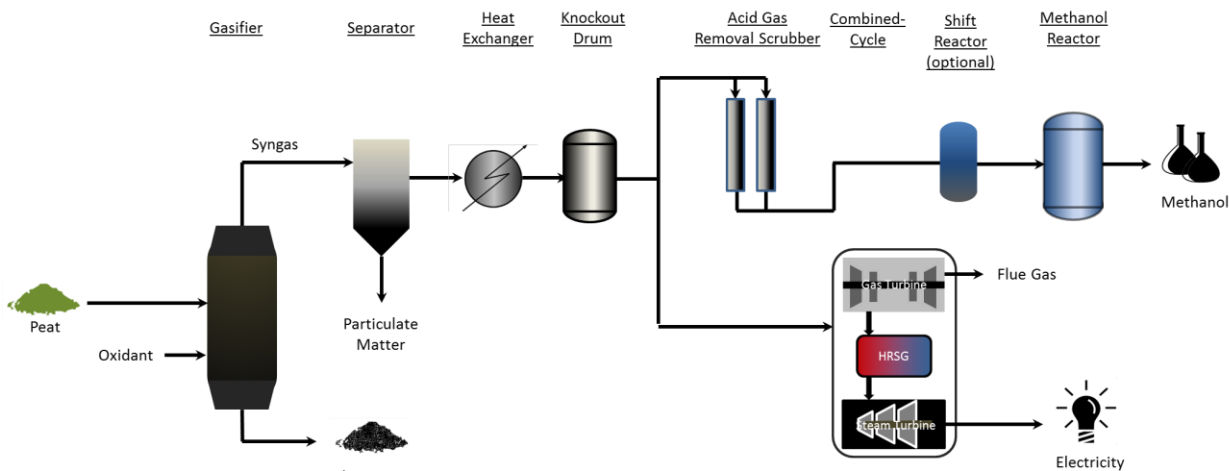


Figure 4. Process flow diagram for the base case design

Pre-dried peat with a moisture content of 15% is fed to the gasifier along with an oxidant and temperature regulator. This level of moisture is typical for thermochemical operation (Andriessse, 1988; Capareda, 2013). The two gasifier outputs are syngas and bottom ash. Unreactive metal oxides as well as unconverted peat or char are removed as bottom ash. Some fine particles known as fly ash are carried out in the syngas. The syngas cleaning and conditioning stage within which harmful particulate matter is removed, thereby reducing if not preventing the haze. Syngas is also cooled during the cleaning and conditioning stage as the methanol synthesis pathway strictly requires a low temperature, the lower gas temperature reduces equipment costs as well as operating and maintenance costs, heat can be recovered by generating steam, which can be used elsewhere in the process. Thereafter, syngas is split into two utilization pathways: methanol synthesis and power generation.

Regardless of what technology is used for power generation, the theoretical principles are similar. Chemical energy in syngas is converted to electrical energy either directly via electrochemical reactions, or indirectly through the use of mechanical devices (e.g., turbines). The flue gas is treated before being released into the environment, subject to local regulations. Methanol synthesis catalysts can either be solid or liquid. These catalysts, however, are prone to

damage by the sulphuric components often found in syngas (Radovic and Vannice, 1987). In light of this, acid gas removal is needed to condition the syngas before it is fed to the methanol reactor. Other pre-treatment units like a CO-shift reactor are sometimes needed depending on the type of methanol synthesis approach. The process design alternatives are summarized here. Two gasifiers (i.e., updraft fixed-bed and dual fluidized-bed) and two methanol synthesis reactors (i.e., gas-phase and liquid phase) are chosen. Each gasifier is then paired with each methanol synthesis reactor to generate four design alternatives: 1) Design Alternative 1: fixed-bed gasifier and gas-phase methanol synthesis reactor, 2) Design Alternative 2: fixed-bed gasifier and liquid-phase methanol synthesis reactor, 3) Design Alternative 3: fluidized-bed gasifier and gas-phase methanol synthesis reactor, and 4) Design Alternative 4: fluidized-bed gasifier and liquid-phase methanol synthesis reactor.

This section describes how the Aspen Plus simulations for all four design alternatives are constructed. All models follow the same process map and process flow shown in Figure 4. Capitalized terms represent stream names and unit names in the models, whereas italicized terms are units in Aspen Plus. The Redlich-Kwong-Soave cubic equation of state with Boston-Mathias alpha function (RKS-BM) is used as the global method for calculating physical properties (Swanson, et al., 2010; Begum, et al., 2013). The process feed stream, DRYPEAT, is characterized by a custom-defined Non-Conventional (NC) Solid. Peat is assumed to be chemically similar to coal in its early stage of formation (Boron, et al., 1987; Matar, Mirbach and Tayim, 1989). Therefore, the method of HCOALGEN (i.e., coal enthalpy model) is used to specify peat's heat of combustion, heat of formation, and heat capacity. Similarly, the method of DCOALIGT (i.e., coal density model) is used for calculating its density. Both HCOALGEN and DCOALIGT make use of the proximate analysis and ultimate analysis for peat.

The operating moisture content of peat is specified as 15% (Andriessse, 1988; Capareda, 2013). This assumes that the peat has been pre-treated and dried prior to operation. A mass flowrate of 1000 tonne peat / day is used as a basis, which is reasonable for a large-scale biomass utilization process (Witjaksono, 2013). For peat to be able to undergo chemical reactions in Aspen+, it has to be decomposed into conventional species based on its proximate and ultimate compositions. Peat is most appropriately represented by a combination of ash, H₂O, C, H₂, N₂, Cl₂, S, and O₂. Ash is also modelled as a NC Solid and it represents all trace-components (e.g., metal oxides) that are assumed to be non-reactive. This decomposition process is facilitated by an *RYield* unit labelled DECOMP. The reactor yield is determined based on the moisture content, ash content and ultimate analysis of peat by a 'Calculator' labelled DECYIELD. Now, the process feed stream, under the label ELEMENT, is prepared for gasification. Note that there is a makeup energy stream labelled BALANCE, which accounts for enthalpy losses during decomposition. ELEMENT, BALANCE, and oxidant are then fed to the gasification section of the model.

The primary output from the gasification process is syngas, which is cleaned and conditioned in the next stage of the model. First, syngas undergoes solid separation in a *Sep* unit labelled SOLIDSEP. This is where the harmful particulate matter that produces haze is filtered out of the process. Other unreacted solids and both bottom ash and fly ash are also separated here. Note that in reality the different types of ash are separated differently. Thereafter, syngas is cooled using a *Heater* unit labelled COOLING1. Any condensate is then removed using a knock-out drum labelled KODRUM. After cooling, the syngas is split into two streams for power generation and methanol synthesis. The split fraction is 0.5. This suffices for comparing design alternatives, but should be optimized in the future. A largely simplified representation of power generation is used for comparing the design alternatives. This is because modelling it in detail will not favour one design alternative over another unless stream (material and/or energy) integration exists. Integration, as an element of process modelling, is time-consuming and was therefore only performed for the final design. The simplification is given by a combustion process within which syngas is burned with excess air. This is housed in an *RGibbs* unit labelled COMBUST. Flue gas is the sole output from this combustion process and it contains recoverable heat that can be used for combined-cycle power generation. An efficiency of 61% is used for calculating the power output (Kiameh, 2012).

For the methanol synthesis pathway, the syngas is compressed (COMP1) and cooled (COOLING2) for acid gas removal (AGR1). The Rectisol process is used to remove sulphur and carbon dioxide. AGR1 is a *RadFrac* column with no condenser or reboiler, which is representative of an actual absorber. Due to time constraints, the complete loop for solvent regeneration is not modelled, but its most critical element, that is the pressure-swing operation, is included. This is needed for calculating the amount of solvent that is lost and needs to be replenished. Following acid gas removal, clean syngas (denoted as CLEANGAS) is fed to the methanol synthesis section of the model. Lastly, the synthesized methanol product (RAWMEOH) is then purified in a distillation column, which is modelled using a *RadFrac* column labelled DISTIL with a Partial-Vapor-Liquid condenser. From the distillation, outputs include a

water-rich stream labelled DISTBTM, a nearly pure methanol stream labelled PUREMEOH, and a gaseous stream labelled VENT.

For alternative1 the air separation unit for obtaining pure oxygen is not modelled. In literature, the different zones of the gasifier are sometimes ordered sequentially (Begum, et al., 2013). Using this setup, both the feed and oxidant enter the gasification zone and then syngas exits from the combustion zone. However, upon further research, it was found that this approach was not appropriate for updraft fixed-bed gasifiers, since the oxidant should enter the combustion zone, and syngas should exit from the gasification zone (Basu, 2010). These two zones were modelled as follows: a) Oxygen and steam enter the combustion section (an *RGibbs* unit labelled COMB), and b) syngas exits from the gasification zone (an *RGibbs* unit labelled GASF). To better represent the actual path travelled by the gases (i.e., GASFOUT → TOCOMB → TOGASF), an internal loop is constructed between the gasification zone and combustion zone. Before syngas cleaning and conditioning, the syngas is fed into a steam reformer (an *RGibbs* unit labelled REFORMER) to crack methane. The reformer would also crack gaseous tar, but tar production cannot be captured in the model when using *RGibbs* units for the gasifier. This, however, should have negligible effect on the model results since the tar concentration is typically in the order of a few g/Nm³ (Basu, 2010).

For fixed-bed gasification with gas-phase methanol synthesis, a second stage of acid gas removal (AGR2) is required after syngas cleaning and conditioning. The modelling of AGR2 is similar to that of AGR1, but with a different temperature and number of stages. A CO-shift reactor is not required as the mole fraction of CO in the treated syngas is already lower than the desired value of 0.16 (Hamelinck and Faaij, 2001). The syngas is then fed into the gas-phase methanol reactor (an *REquil* unit labelled MEOHPDN). The reactor product (denoted as GASMEOH) is routed through a ‘Flash2 unit’ labelled FLASH to separate the raw methanol from unreacted syngas and inert gases. The majority of the gases are recycled into the reactor and the rest is purged to prevent inert gas buildup. Design alternative 2 has the same process flow as alternative 1 except for the methanol synthesis section. Instead of gas-phase methanol (GPMEOH) synthesis, this alternative utilizes liquid-phase methanol (LPMeOH) synthesis. The key differences between synthesis methods are the restrictions on the H₂ to CO ratio of the incoming syngas and the operation temperature of the reactor. In regard to the ratio, LPMeOH synthesis has more flexibility and does not need a CO-shift reactor. As compared to its gas-phase counterpart, LPMeOH also operates at a lower temperature. As a result, both gaseous and liquid methanol are formed. The gaseous methanol is condensed to join the liquid stream, and together they represent the raw methanol product. For design alternative 3 the process is unchanged with reference to Design Alternative 1 up to and including the decomposition of peat. The ELEMENT stream is then split, most of which is fed to a steam-fired gasification chamber (an *RGibbs* unit labelled GASF) to produce syngas. The balance is fed to an adjacent air-fired combustion chamber (an *RGibbs* unit labelled GASFCOMB), where it is burned to generate heat for the gasification chamber. Note that a steam reformer is not modelled for this alternative because the syngas contains minimal methane (<0.3 mol-%). Thereafter, the downstream processes mirror those in Design Alternative 1, except a CO-shift reactor is needed in this case. This is because the syngas has a high CO concentration of 47 mol-%, which greatly exceeds values recommended in literature (Hamelinck and Faaij, 2001). For design alternative 4, instead of gas-phase methanol (GPMEOH) synthesis, this alternative utilizes liquid-phase methanol (LPMeOH) synthesis.

Table 2 shows a summary of simulation results based on the Aspen Plus models for all design alternatives. These results include production and consumption of methanol and energy as well as emission data for carbon dioxide. The results are then used to compare the four design alternatives. Net heat energy surplus is calculated by taking into account all the heat sources and heat sinks in the process. The percentage reduction in CO₂ emission is calculated as the percentage of carbon in peat that is captured into ash and methanol. The elemental carbon ratio in peat from above (i.e., 0.4133 tonne C / tonne peat) is derived from the composition data in Table 1 on a wet basis.

Table 2. Summary of simulation results for all design alternatives.

Design Alternative	1	2	3	4
Methanol production rate (tonne/day)	226.7	298.3	162.4	204.0
Methanol consumption by first acid gas removal unit (tonne/day)	16.6	16.6	0.1	0.1
Methanol consumption by second acid gas removal unit (tonne/day)	24.0	N/A	22.5	N/A
Net methanol production rate (tonne/day)	186.1	281.7	139.8	203.9

Usable energy from syngas combustion (MW)	102.2	102.2	106.9	106.9
Total shaft work required for compressors (MW)	1.1	1.1	5.3	5.3
Net heat energy surplus (MW)	38.7	42.2	29.9	31.2
Reduction in CO ₂ emission (%)	18.9	27.6	15.7	21.7
Emission rate of CO ₂ (tonne/day)	1,229	1,097	1,278	1,187

Five selection criteria, namely, methanol yield (tonne methanol / tonne peat), reduction in CO₂ emissions (%), usable energy (MW), capital cost (million US\$) and payback period (years) are used to compare the four design alternatives and select the best option. These criteria are chosen and weighted to reflect the design goals. Note that a higher rating indicates better performance with respect to each criterion (e.g., a higher capital cost rating indicates a lower capital cost). The first three criteria were based on the simulation results. Design Alternative 2 (DA 2) earned the highest rating in all three of these criteria for two main reasons: pure oxygen generates a higher quality syngas than steam, and liquid-phase methanol synthesis captures more carbon than its gas-phase counterpart. The last two criteria required preliminary costing for each of the alternatives. Design Alternative 4 (DA 4) rated highest in both capital cost and payback period. This is primarily because it does not require either an air separation unit or the second acid gas removal column, both of which have large upfront costs and are costly to maintain. A summary of the simulation results, preliminary costing, and ratings is provided in Table 3. Both results and costs are listed within the ‘Value’ column. As can be seen in Table 4, alternative 4 (i.e., fluidized-bed gasification and liquid-phase methanol synthesis) scores highest overall and is therefore chosen for the final design.

Table 3. Summary of values (i.e., simulation results and cost calculations), ratings (R: 1-4) and scores (S) for the four design alternatives.

Criteria	Weight	DA 1			DA 2			DA 3			DA 4		
		Value	R	S	Value	R	S	Value	R	S	Value	R	S
#1	0.1	0.37	2	0.2	0.56	4	0.4	0.28	1	0.1	0.41	3	0.3
#2	0.2	18.9	2	0.4	27.6	4	0.8	15.7	1	0.2	21.7	3	0.6
#3	0.1	95.6	3	0.3	98.9	4	0.4	89.8	1	0.1	91.1	2	0.2
#4	0.2	239.6	1	0.2	229.3	2	0.4	219.6	3	0.6	211.6	4	0.8
#5	0.4	13.3	1	0.4	9.0	2	0.8	7.7	3	1.2	6.2	4	1.6
Total Score		1.5			2.8			2.2			3.5		

After identifying alternative 4 (final design in Figure5) as the highest-scoring option with respect to the selection criteria, its model was fine-tuned. More specifically, units were added, removed or modified to improve model accuracy. Parameters for certain units were changed to achieve better outcomes (e.g., higher methanol yield, lower equipment costs, etc.). Moreover, a detailed model of the power generation section was constructed to replace the simplified version used for comparing design alternatives. This allowed for extensive heat integration throughout the model to make the process more energy efficient. Upon further review of the simulation results, it was discovered that the cooled raw syngas (COOLGAS) has a relative humidity of less than 5%. It is therefore unlikely that condensation will occur, so the knockout drum is omitted from the model. Also, a polytropic multistage compressor (AGRCOMP) is used in place of the isentropic single-stage compressor (COMP1) prior to acid gas removal. Lastly, after adjusting parameters of the methanol distillation column, it was found that the column was unnecessary for the desired level of purification. A flash drum is used instead to reduce capital expenditure.

Thermodynamic limitations prevent the model from achieving the desired exhaust temperature of 550 °C with a single-stage setup, so a dual-stage combustion gas turbine is used. Compressed fuel is fed into two combustion chambers (RGibbs units labelled COMBUST1 and COMBUST2) in the gas turbine, where it is mixed and burned with compressed air. The combustion products then expand through two turbine stages (Compr units labelled GAST1 and GAST2) and produce a combined shaft work output. This is directly translatable to power. The exhaust gas from the gas turbine (labelled GASTOUT) then passes through a Heat Recovery Steam Generator (HRSG), where boiler feed water streams are preheated and steam is generated or reheated. Flue gas from the combustion chamber of the

gasifier is also routed through the HRSG. To model this section, three sequential heat exchangers (*HeatX* units labelled HRSG1, HRSG2 and HRSG3) are used. It is assumed that the flue gas after HRSG3 will undergo treatment before being released into the environment.

The steam turbine consists of three stages, namely, high pressure, intermediate pressure and low pressure. These are modelled as *Compr* units labelled STEAMT1, STEAMT2 and STEAMT3 respectively. High pressure steam first passes through STEAMT1 and expands to become intermediate pressure steam. A small quantity of steam is drawn for use in the methanol synthesis section, but the majority is reheated in HRSG2 to the same temperature as before expansion and fed to STEAMT2 (Boyce, 2010). Then, the steam passes through STEAMT3. Note that non-condensing steam turbines are used and it is assumed that the low pressure steam at the outlet of STEAMT3 can be used for auxiliary heating before being condensed back to boiler feed water. Any non-condensables will be vented. After constructing the power generation section for the final design, energy integration was performed by recovering heat from high-temperature streams and extracting energy from unreacted syngas. The two most apparent instances of heat recovery are as follows: 1) Hot flue gas from the gasifier is routed to the HRSG power generation section, so that its thermal energy can be utilized to produce steam, and 2) The purge stream from the methanol synthesis section is routed as fuel to the gas turbine, so that its CO and H₂ can be combusted to generate power. With respect to energy extraction from unreacted syngas, heat is transferred to boiler feed water streams (labelled BFW1, BFW2 and BFW3) before they enter the HRSG for further heating. This integration occurs in three areas, namely, raw syngas (RAWGAS) cooling, compressor intercooling before acid gas removal, and fuel compressor intercooling before the gas turbine.

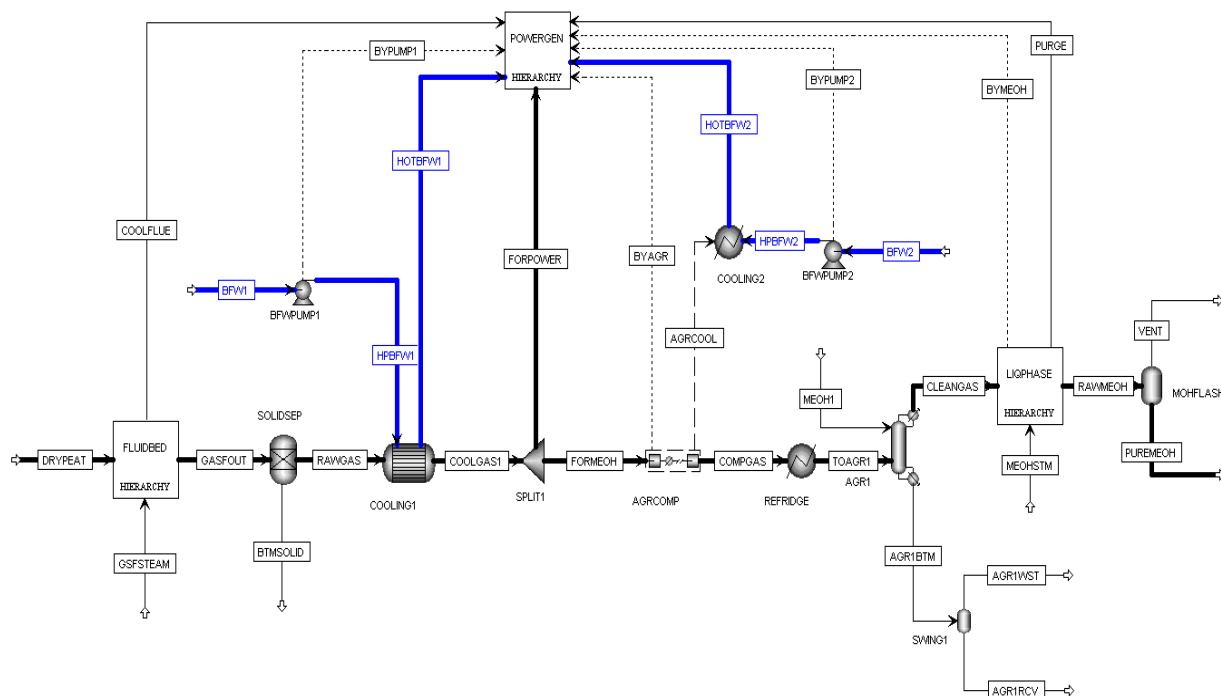


Figure 5. High-level view of the final design in Aspen Plus.

After process refining, the simulation results for the final design are updated in Table 4. Note that as a result of heat integration as well as detailed modelling of the power generation section, usable energy is no longer a useful performance measure. Instead, actual generation and consumption of electricity from the process can now be obtained directly from the Aspen model. Results relevant to the environmental performance of the process are summarized in Table 4. The reduction in CO₂ emission and the emission rate of CO₂ are calculated in the same way as for the design alternatives. The concentrations of SO₂ and NO_x are also reported in Table 5.

Table 4. Summary of simulation results for the final design.

Methanol production rate (tonne/day)	213.6
Methanol consumption by acid gas removal unit (tonne/day)	≈ 0 (negligible)

Net methanol production rate (tonne/day)	213.6
Electricity generation from the gas turbine (MW)	39.1
Electricity generation from the steam turbine (MW)	24.9
Electricity consumption by pumps (MW)	0.8
Electricity consumption by compressors (MW)	7.1
Net electricity output (MW)	56.1

Table 5. Summary of environmental performance for the final design.

Reduction in CO ₂ emission (%)	22.4
Emission rate of CO ₂ (tonne/day)	1,178
Concentration of SO ₂ in untreated flue gas (mg/m ³)	235
Concentration of NO in untreated flue gas (mg/m ³)	10
Concentration of NO ₂ in untreated flue gas (mg/m ³)	2,770

For peat-to-methanol, there is limited information in literature for validation. Methanol yield was calculated as roughly 0.24 tonne methanol/tonne peat for a plant that was built three decades ago (ACS, 1981). More generally, for biomass-to-methanol, the yield can be as high as 0.57 tonne methanol/tonne biomass (NREL, 1995). From the Aspen Plus model for the final design, with 500 tonnes of peat used for methanol synthesis per day, 214 tonnes of methanol is produced daily. This gives a methanol yield of roughly 0.43 tonne methanol/tonne peat, which is within the observed range of 0.24 to 0.57. This demonstrates that the methanol yield for the final design is reasonable when compared to literature sources. For an integrated gasification combined cycle (IGCC), efficiencies of between 39 and 58 (% LHV of feed) have been cited for existing plants (Higman and Burgt, 2008; Kehlhofer, et al., 2009). More recent technology can achieve efficiency as high as 61% (Kiameh, 2012; GE Power&Water, 2015). From the Aspen Plus model for the final design, with 500 tonnes of peat used for power generation per day and an HHV of 20.3 MJ/kg for peat, the energy stored in incoming peat is around 117 MW. With a net power output of 56 MW, the efficiency of the proposed process is roughly 48 (% HHV of peat). The LHV data for peat is not available, but this is not concerning since it can be assumed with reasonable certainty that the LHV efficiency will be within the range of 39-61%.

The base case for the economic assessment of the final design is given by the following parameters: 1) Peat flowrate = 1,000 tonne/day, and 2) Peat price = \$17.2/tonne. The peat price assumes the peat itself is supplied for free, but pre-treatment costs (e.g., drying, handling, and storage, etc.) amount to \$17.2/tonne peat (Western University, 2012). For the final design, preliminary equipment sizing and costing was performed using the simulations results from the Aspen Plus model. The capital costs for unit operations not in the model (e.g., flue gas treatment, methanol recovery after acid gas removal, etc.) are estimated by multiplying the total capital cost for modelled units by 5%. Thereafter, a factor of 20% is applied to the total fixed capital to include working capital. A relatively simple estimation of operating cost is employed in this study. Only costs of feedstock and water are considered, so other elements like labour and storage should be considered in the future. As mentioned, peat assumes a price of \$17.20/tonne. The cost of water is cited as \$1.10/tonne (Towler and Sinnott, 2013). Once these are determined, the payback period can be found by calculating the annual revenues from the sales of methanol and electricity. A constant electricity price of \$0.0817/kWh (PWC, 2013) and an inflation rate of 5% (Trading Economics, 2016) are assumed. Methanol price is forecasted using a simple polynomial model based on historical contract pricing data for the Asia-Pacific region (Methanex, 2016). The annual revenue therefore changes year to year. Note that a constant methanol price of \$255/tonne was assumed for simplification when comparing design alternatives. Annual profit for year-round operation can then be found by subtracting the O&M costs from the revenues. It is assumed that an actual availability of 80% (i.e., the plant operates 80% of the time) is appropriate once planned and unplanned shutdowns are accounted for (Higman and Burgt, 2008). The payback period can be assessed visually by plotting cumulative operating profit over time. This is illustrated in Figure 6. Different feedstock pricing scenarios have also been included in the graph.

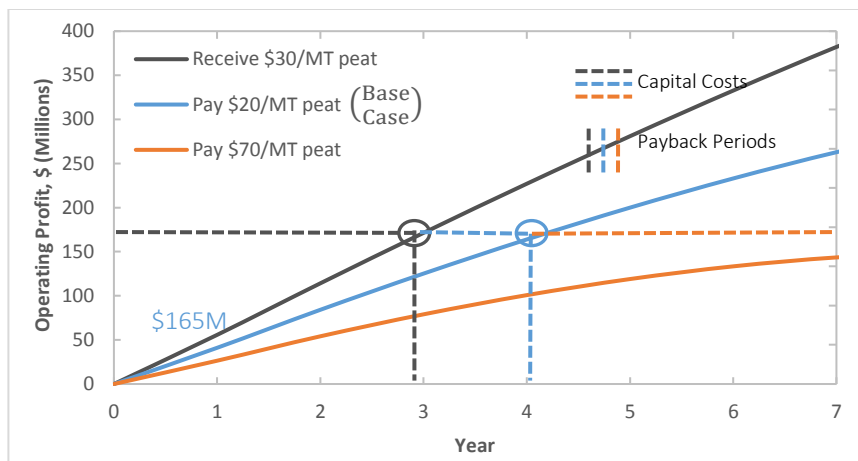


Figure 6. Cumulative operating profit for different peat prices.

3. Conclusions

The fluidized-bed gasifier and liquid-phase methanol synthesis reactor (alternative 4) is the best design based on preliminary costing and simulation results. Four design alternatives were generated by pairing two gasifiers (fixed-bed and fluidized-bed) with two methanol synthesis reactors (gas-phase and liquid-phase). The fluidized-bed gasifier with liquid-phase methanol reactor performs the best with respect to capital cost and payback period. These selection criteria are evaluated using preliminary equipment sizing and cost calculations. Although it is ranked below the fixed-bed gasifier with liquid-phase methanol synthesis in the remaining criteria, Design Alternative 4 still performs strongly in CO₂ reduction, methanol yield and usable energy. These selection criteria are evaluated based on the Aspen Plus simulation results. After ranking the design alternatives against these selection criteria with different weightings, Design Alternative 4 has the highest score and is therefore selected as the final design. Methanol yield and power generation efficiency of the final design is reasonable and aligns with values cited for plants of this size. The final design was refined for a high methanol yield and high electricity output. The final design converts 1000 tonne peat/day into 214 tonne methanol/day and 56 MW of electricity. With 500 tonnes of peat for methanol synthesis, this gives a methanol yield of roughly 0.43 tonne methanol/tonne peat, which is within the observed range of 0.24 to 0.57 for other biomass-to-methanol plants. The energy efficiency is around 48%, which is in the range of 39-61% cited for existing integrated gasification combined-cycle plants.

The greatest environmental risks are air pollution from carbon dioxide emissions and sulphur and nitric oxides. The selected design of a dual fluidized-bed gasifier and liquid-phase methanol synthesis reactor effectively prevents the haze. By harvesting the peat and utilizing it in a controlled manner, emission of particulate matter is essentially eliminated, thereby preventing the formation of haze. This is because the dense nature of the synthesis gas resulting from gasification enables easy removal of particulate matter. In the final design, SO₂ and NO₂ emissions have concentrations of 235 mg/m³ and 10 mg/m³ respectively, both of which are below emission limits. In addition, by capturing the carbon into methanol, CO₂ emission is reduced by 22.4% as compared to the open burning of peatland. Payback period is reasonable for various peat prices and flowrates. The final design has a total capital cost of \$165M and a payback period of roughly four years based on a peat price of \$17.20/tonne, an electricity price of \$0.0817/kWh, and methanol pricing extrapolated from historical data.

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