Future directions for CHP plants using biomass and waste – adding production of vehicle fuel and other chemicals

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Abstract. In Northern Europe, the production of many biobased CHP plants is getting affected due to the enormous expansion of wind and solar power. In addition, heat demand varies throughout the year, and existing CHP plants suffer economically. By using the existing CHP plants for the production of chemicals, they can be operated more hours and can provide operational and production flexibility and thus increase efficiency and profitability. In this paper, we look at possible solution by converting an existing CHP plant into integrated biorefinery by retrofitting pyrolysis and gasification including gas upgrading and conversion of bio oil to vehicle grade liquid biofuels. Waste and biomass CHP were combined with pyrolysis through G-valve to liquefy biomass and then react the liquid with hydrogen produced in a gasifier upfront to CHP plant. The results show that a pyrolysis plant with 18 ton/h feed handling capacity (90 MWth), when integrated with gasification for hydrogen requirement and CHP plant for heat can produce 5.2 ton/h of gasoline/diesel grade biofuels. The system integration gives positive economic benefits too but the annual operating hours can impact the economic performance.

1 Introduction

In Northern Europe, there are many combined heat and power (CHP) plants to fulfill the heat and electricity demand. Many of them are operating with biomass and waste as fuel. Since the enormous expansion of wind and solar power in Germany, Denmark and Sweden, we see strong competition between these and the thermal power of all kind. The wind and solar power cannot be controlled and thus have priority to the electric grid, and then coal-fired and nuclear power plants. Waste and biomass-fired CHP plants also back their production.

Furthermore, the variation in heat demand throughout the year also forced these CHP plants to operate on part load. Part load operation of CHP plant with too few operational hours not only make the economic performance worse but also decrease the overall process energy efficiency [1]. By integrating the CHP plants with thermochemical processes the for production of biofuels/chemicals they can be operated more hours with the high energetic return and may also the increased profitability provided that the additional integrated processes like gasifier and pyrolysis are not too expensive [2], [3].

Both gasification and pyrolysis processes can treat biomass and waste into a variety of biofuels in standalone processes. However, these processes are carried out at a high temperature and require heat at various unit operations. The heat required has to be produced onsite. Existing CHP plants can provide the required heat for these thermochemical processes during offpeak hours and increase its overall operating capacity throughout the year. In this paper, we look at a novel solution to integrate pyrolysis and gasification including gas upgrading with existing CHP plants. Three criteria's have been selected to design the integrated processes. (1) All the heat required to run the pyrolysis and gasification processes should be produced and provided within the whole integrated system. (2) The capacity of integrated plants is selected in such a way that they will not affect the heat and power production in existing CHP plants. (3) The hydrogen required to upgradation of bio-oil from pyrolysis is produced in the upfront gasification system.

2 Process integration concept

The process integration concept is presented in Figure 1. we combine the CHP with pyrolysis to liquefy biomass and then react the liquid with hydrogen produced in a gasifier. The gas from the gasifier then is upgraded by membrane separation with liquid absorbent. The gasifier is located separately upstream of existing CHP plant. The G-valve in the CFB boiler is used as a pyrolysis unit. The pyrolysis is taking place in the G-valve where heat is coming from the hot sand from the combustor. Vapors obtained from the pyrolysis are condensed to bio-oil. The bio-oil is a thick brown liquid and composed of complex oxygenated hydrocarbons with water in it. Bio oil can be upgraded via hydrotreating to vehicle fuels (gasoline/diesel). In this process, oxygen in bio-oil is removed as a water and carbon dioxide by reacting it with hydrogen at high pressure and moderate temperature. Amount of hydrogen required for hydrotreating is usually 4-5 wt. % of total bio oil weight [4]. The required hydrogen can be purchased or produced onsite by the steam reforming of off-gas from hydrotreater but this gas is usually not enough so natural gas from the grid is required to produce the required hydrogen. The purchasing of hydrogen or natural gas from the grid takes a toll on the economic performance of the whole process [4, 5]. The upgradation of bio-oil requires hydrogen as additional raw material. The upfront gasifier is designed to produce the required hydrogen. The gasifier requires an oxidizing agent which can be air or steam. Use of air can dilute the synthesis gas with nitrogen which will increase downstream process complexity and costs. So steam is used as a gasifying medium and imported from CHP plant.

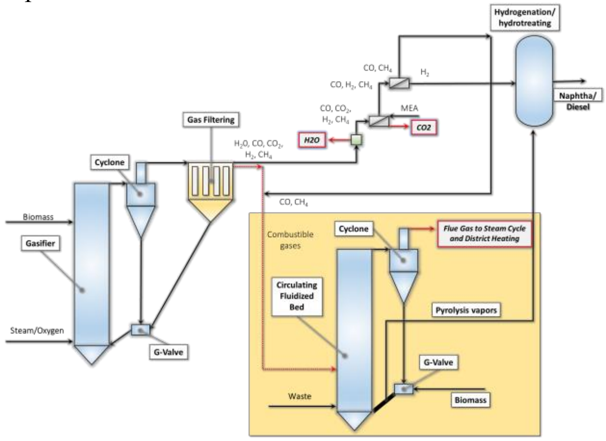

Figure 1: Process scheme with pyrolysis integrated with CHP G-valve and gasifier integration at the upfront of CHP plant (highlighted portion shows the existing boiler of CHP plant)

3 Process modeling and simulation

A circulating fluidized bed boiler located in Malarenergi, Västerås, Sweden is considered as a base existing power plant. The boiler operates on refuse-derived fuel and has a capacity of 180 MWth. The studied systems were simulated using Aspen plus. The existing CHP plant was modeled and simulated by dividing among different sections such as a boiler with heat recovery steam generation and steam turbine previously developed by authors [2]. Pyrolysis and gasification process was modeled and integrated with existing CHP plant to estimate the mass and energy balances. The product distribution of the pyrolysis reactor was modeled by using the method previously developed by authors [2, 7] while the bio-oil upgradation methodology has been adopted from Dutta et al. [4]. The hydrogen required for hydrotreating of bio-oil was imported from the gasification process upfront to CHP plant. The gasification and hydrogen separation process has been modeled previously by authors [2][7].

In the reference CHP plant, the steam enters the turbine at 470 ºC and 74 bar. For the mass integration, steam was extracted at the second stage of the steam turbine at 250 ºC and 4 bar to use as an oxidizing agent in the gasifier. The char and unreacted gases in gasifier were combusted in the existing boiler as additional fuel. Similarly, the bio char and non-condensable gases from pyrolysis in G-valve of CFB boiler are also combusted as additional fuel in the existing boiler. For heat integration, the pinch temperature of 5K has been considered.

4 Results

Technical performance:
For integration case, it was assumed that half the capacity of the boiler can be pyrolyzed for liquid fuels, i.e., approximately 90 MW. The feed selected for the thermochemical conversion was either wood pellets or waste with a high heating value of 18 MJ/kg. This implies that the pyrolysis process needs 18 ton/h (5 kg/s) of biomass feed. The amount of bio-oil produced was 12.4 ton/h and require 0.6 ton/h of hydrogen for hydrotreating. Gasification of wood pellets produces the required hydrogen. From the simulation, it has been estimated that 12.6 ton/h (3.5 kg/s) of feed is required in the gasification process. The steam to biomass ratio of 0.6/1 was selected in the simulations, so 7.5 ton/h (2.1 kg/s) of steam also needs to be imported from existing CHP plant. The upgradation of bio-oil produced 2.6 ton/h of biofuel with naphtha range hydrocarbons and 2.6 ton/h with diesel range hydrocarbons. The integrated process also produces approximately 39 MW excess heat from the burning of additional fuel in the CFB boiler containing bio char, non-condensable gases from pyrolysis, and syngas from gasification after separation of hydrogen. It also implies that the boiler needs less amount of fuel to run on full load when integrated with pyrolysis/gasification (46 ton/h instead of 60 ton/h).

It has also been observed that steam to biomass ratio has a big effect on the volume fraction of synthesis gas and subsequently the amount of hydrogen produced as a result. The amount of biomass to produce the required hydrogen varied by changing the steam to biomass ratio (figure 2).

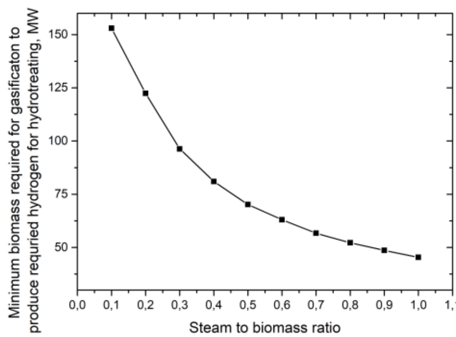

Figure 2: Effect of steam to biomass ratio on the minimum gasification capacity requirement to produce the hydrogen needed in hydrotreating

As shown by the results that the demand for hydrogen is (0.6 ton/h), i.e. 600 kg/h equivalent to 300 kmol H_2/h. We can go backward and look at two different alternatives. In alternative 1, we assume a water gas shift reactor (WGS) and alternative 2 with a membrane with a thin layer of Palladium and Tantal on a porous ceramic tube. In the WGS reactor, we assume 82% conversion of methane [8] and 96 % conversion of CO [9]. For membrane separation process to produce hydrogen, we assume that 90% of the hydrogen is passing the membrane with quality of > 99% H_2, and we can also neglect permeation of other components. In Table 1 we see the gas components as kmol/h from the gasifier for both alternatives, i.e., (1). After the WGS reaction and (2). After the membrane filtration.

Table 1: Gas composition as kmol/h for the major components neglecting the nitrogen and steam (kmol/h)

	CO_2	CO	CH_4	H_2	H_2O
After gasifier	190.3	135.7	53.9	28.6	247.8
1. After WGS reactor	364.7	5.4	9.7	335.6	29.2
2. After membrane filtration					
Permeate				302	
Reject	364.7	5.4	9-7	33.6	29.2

300 kmol H_2/h correspond to 6720 nm^3 H_2/h. According to [10] the permeate flux through the Palladium membrane is 30.5 m^3/m^2.h at dP= 10 bar, and twice as much at 20 bar dP (pressure difference over the membrane). If we assume 10 bar dP the membrane surface area would be 220 m^2.

Economic assessment:

The cost for the Pd-coated membrane is estimated by [10] to be 574 – 765 €/m2. For 220 m2 the installation cost for membrane with modules would be 126 – 168 k€. To this we need a gas compressor for the flow 14700 nm3/h increasing the pressure to 10 bar [11]. With a two-stage gas compressor the power demand would be some 1700-1850 hp. The average cost for a complete gas compressor station in the US is 1712 \$/hp according to Zhao and Rui [12]. For our demand the compressor cost estimates to approximately 2.57 – 2.79 M€. We also need to include a heat exchanger for gas cooling and reheating to some 300 ºC. According to Harvey et al. [13], gas-gas heat exchangers have an average cost around or 430 – 540 \$/kW. We have a gas flow out of the reactor around 657 kmol/h or 14 700 nm^3/h without nitrogen or some 23 000 nm^3/h with nitrogen. If we assume, 400 ºC cooling demand amounts to 3675 – 5750 kW. This means cost for the heat exchanger would be around 1.42 – 2.74 Million €. For the actual gasifier plant, we already have an estimate for a 90 MW plant at 50 M€ including ceramic filter and condenser earlier [14]. In our the direct demand for production of the hydrogen needed is approximately 25 MW (HHV in the produced gas), or 38 MW with 25 % combusted an input of 33 MW fuel, with some margin 38 MW for the gasifier, or 24.7 M€. There will be additional costs for buildings and installations connecting the gasifier, gas upgrading and rebuild of G-valve to become the pyrolyzer. A rough estimate would be around 10 M€. For the upgradation of biooil to liquid gasoline and diesel grade biofuels, the capital cost is estimated from Jones et al. [5]. The capital cost for hydrotreating and hydrocracking of biooil is estimated as 18-23 M€. Total capital expenses used as in this study is estimated as a summation of both plants i.e., 68 M€. With the annuity factor of 0.081, we get annual capital cost of 5.5 M€/y. Table 2 describes the production costs with biomass and waste as an input fuel. We neglected the cost of the existing CHP plant. We assume that the plant should operate 7000 hours in an annual year. However, during the winter the demand for heat may be quite high, and the focus will be on delivery of heat to consumers. Then the operational hours can be expected to be lower for fuel production. We thus make also a calculation for 5000 h operations per year reported above. For less operating hours, we used the same capital investment cost, but electric power and staff costs will be less. Similarly, the consumption of biomass also affects.

Table 2: Operating and production costs for liquid fuel

	38 MW gasifier + Gvalve	
Annual operating hours, h	**7000**	**5000**
Capital costs gasification plant, M€	45	45
Capital costs hydrotreating and hydrocracking of bio oil, M€	23	23
Total capital costs, M€	68	68
Annualized capital cost (0.08 annuity), M€/y	5.5	5.5
Operating costs		
Electricity costs ((1.27 – 1.38 MW) @ 40€/MWh, k€/y	0.37	0.27
Membrane replacement, k€/y	0.5	0.5
Staff costs, M€/y	1	0.7
Operating costs, M €/y	7.37	6,67
Fuel costs		
(1) Waste (-12 €/MWh), M€/y	-3.2	-2.8
(2) Biomass (14 €/MWh), M€/y	14.9	10.71
Total operating costs with waste as fuel, M€/y	4.17	4.01
Total operating costs with biomass as fuel, M€/y	22.27	17.52
Production costs		
Production costs (€/kg) liquid fuel with waste as feed	0.12	0.15
Production costs (€/kg) liquid fuel with biomass as feed	0.63	0.69

The production costs with less operating hours produced fuel at higher production costs. This also shows that any plant which operates at less capacity than designed gives less economic benefits. Further economic indicators such as net present value and payback period and internal rate of return have also been estimated by following the procedure described in [15].

Table 3: Economic indicators of integrated process

Economic indicators	**With biomass as fuel**		**With waste as fuel**	
	7000 h	5000 h	7000 h	5000 h
Internal rate of return, %	9	5	10	6
Payback period, years	7.1	10	6.1	9
Net present value, M€	86	17	118	20

Discussion

In reality, we should add part of the cost for the existing CFB combustor, although much of the capital cost is already paid off and heat and power also produced during the production of liquid hydrated fuel. The capital cost is around 200 M€ for a CFB with fuel preparation, exhaust gas train and steam turbine including buildings and installation costs. With the same annuity, i.e., 0.081, it means 16 M€/y. If all this cost is also added to the production of liquid fuel in the studied system, it will give production costs of 1.14 €/kg. However, most of this cost was not be added because then the production of heat and power should also be added to the costs.

Conclusion and further work

The production of district heat is always a primary product in an existing CHP plant. The production of power mainly depends on the price fluctuation of electricity; it is usually most interesting to produce as much electric power instead of liquid fuels if the price of electricity is high. However, with the increased competition of power from alternative renewable resources it would be feasible to produce liquid hydrated fuel all the year round i.e.m during a normal winter it means we could produce liquid fuel with 7000h/y, but during harsh winter significantly less, but not below 5000 h/y operating hours.

The combination of gasification and pyrolysis system to produce liquid fuel seems feasible both economically and technically, but the complexity increases with such a system. But the future complex scenarios with multiple renewable energy resources present to meet the demand requires complex solutions for implementation. To reduce the complexity of whole system integration, we can produce bio-oil from pyrolysis at one facility and then upgrading it at another refinery. Another option to decrease the complexity of integration is to produce liquid biofuels CHP plant integrated with gasification with Fischer-Tropsch reactor.

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References

[1] E. Sermyagina, J. Saari, J. Kaikko, and E. Vakkilainen, *Appl. Energy*, **183**, (2016).
[2] C. A. Salman, M. Naqvi, E. Thorin, and J. Yan, *Energy Convers. Manag.*, **152**, May, (2017).
[3] C. A. Salman, M. Naqvi, E. Thorin, and J. Yan, *Appl. Energy*, **226**, January, (2018).
[4] A. Dutta *et al.*, March 2015, (2015).
[5] S. Jones *et al.*, *PNNL Rep.*, **PNNL-23053**, (2013).
[6] C. A. Salman, S. Schwede, E. Thorin, and J. Yan, *Appl. Energy*, **204**, (2017).
[7] M. Naqvi, E. Dahlquist, and J. Yan, *Biofuels*, **7269**, April, (2016).
[8] Y. Lv, X. Yu, S.-T. Tu, J. Yan, and E. Dahlquist, *J. Memb. Sci.*, **362**, 1–2, (2010).
[9] F. T. Report *et al.*, August, (2008).
[10] P. K. T. Liu and A. Anderson, *Annu. Rep.*, (2007).
[11] D. C. and M. F. J. Torkelson, N. Ye, Z. Li, *Tech Rep.*, (2008).
[12] Y. Zhao and Z. Rui, *Int. J. Oil, Gas Coal Technol.*, **8**, 1, (2014).
[13] R. Hackl and S. Harvey, **Hackl-2013**, (2013).
[14] E. Dahlquist, M. Naqvi, E. Thorin, J. Yan, K. Kyprianidis, and P. Hartwell, *Appl. Energy*, **204**, (2017).
[15] J. Turton, Richard; C. Bailie, Richard; B. Whiting, Wallace; A. Shaeiwitz, 2009.