A sustainability case study of a biomass power plant using Empty Fruit Bunch in Malaysia

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Contents

Abstract

Energy demand is ever increasing around the world with limitation of fossil fuel. Alternative source to generate the energy is becoming increasing popular with net zero emission strategy adapted by most of the countries at the climate change conference at Glasgow. One of the main targets to utilise biomass for clean energy production for countries where biomass is available abundantly. Malaysia is considered as a tropical weather country and the country is rich with palm oil tree farming with one of the biggest palm oil producers after Indonesia. Residue of

the palm oil tree is rich in calorific value and can be considered for the energy generation using biomass gasification. Empty fruit Bunch and other biomass from the palm tree can be utilized to generate the electricity centrally or as a decentralized power plant in remote location in Malaysia. Energy Malaysia has performed the feasibility study on the electricity production from EFB. Aspen plus process simulation performed with suitable fluid package to simulate the entire process of electricity generation and the gasification plant considered to be run on the generated electricity from the plant. Surplus of 10 MW electricity can be supplied to the grid for the use in community or local industries. Proposed solution can be further expanded to decentralised electricity generation in the remote area of the Sarawak region and remote main land Malaysia.

Keywords: sustainability, energy, biomass power plant, zero-emission.

Introduction

Energy plays pivotal part in the social and economic growth of the country. The energy utilisation in any country is expected to increase with population growth. In 2009, Malaysia's total population was estimated to be about 28.3 million. The total population is projected to grow at annual rate of 1.3% over the outlook period, reaching just below 40 million by 2035. Electricity demand is expected to increase significantly from 96.3 TWh in 2009 to 206 TWh in 2035 with the GDP growth of 4% for next 25 years [1]. Therefore, government requires to provide adequate basic energy supply for entire population to meet up to their domestic needs such as cooking, lighting their home and chilling up their perishable goods and other needs that require electricity. In order to meet the demand, the Malaysian Parliament approved a sophisticated system of feed in tariffs since 2011. This is expected to accelerate renewable energy development in the economy.

Problem: Empty fruit Bunch (EFG) is one of the residues from palm oil fruit after the oil extraction and the EFB residue generate strong smell if left to decompose on the farm. Malaysia is a tropical country and the weather is humid and rainy, this EFB residue can be utilised to generate electricity via biomass power plant. As shown in Figure-1 EFB, Palm Shell and Faber can be utilised as a biomass.

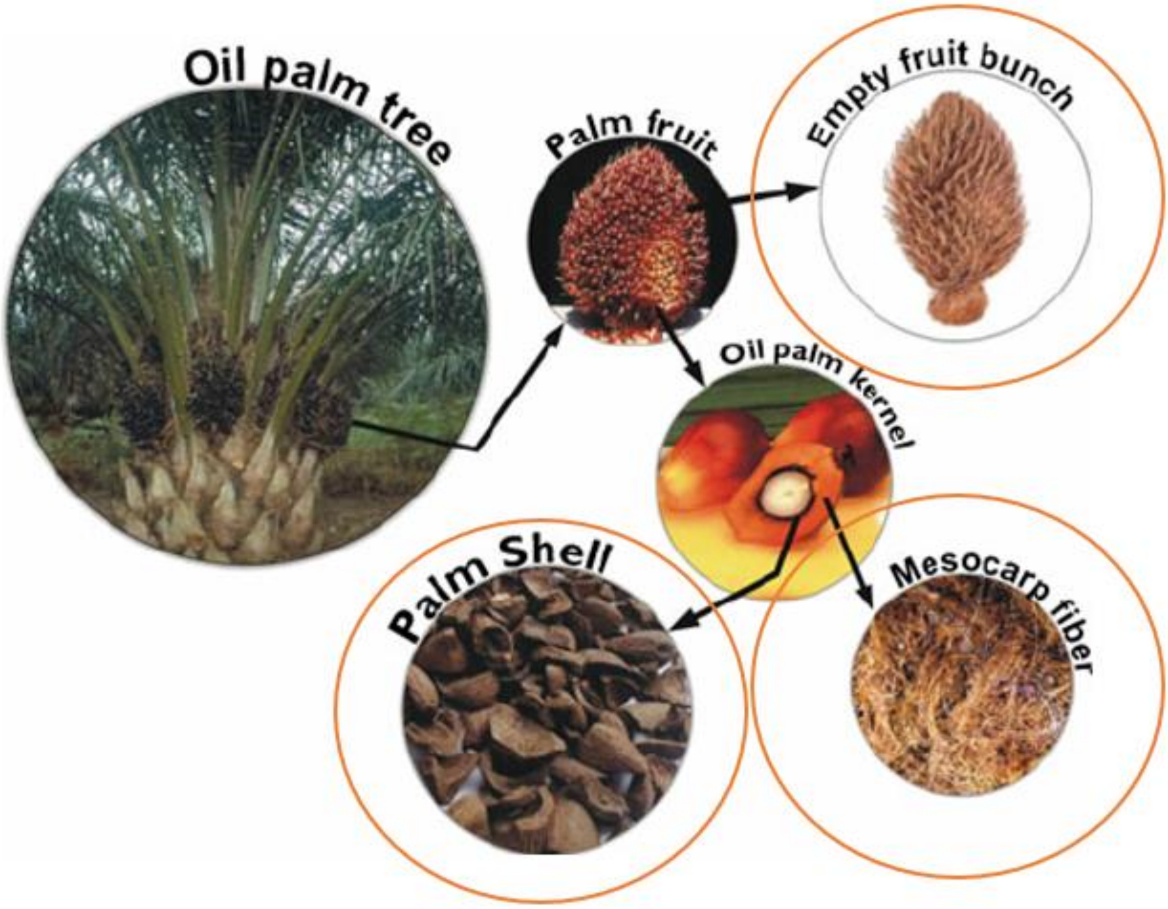

Figure-1 Biomass residue from Palm Tree

Possible solution: Empty fruit bunch from palm oil tree is one of the most important economic oil crops in Malaysia. Malaysia is one of the leading agricultural commodity producers in the Southeast Asian region. Therefore, agricultural wastes are abundant and readily available [2]. The main agricultural based wastes in the country are biomass from EFB and pal tree residue. This waste from oil palm biomass accounted for 50,000 kilotons in the form Empty Fruit Bunch [EFB] [3]. Energy Malaysia decide to take advantage of the abundant source of EFB and generate the electricity. The organisation planned to run the plant in 2015 to generate the electricity from the biomass. Energy Malaysia decided to run the plant with the electricity generate from biomass and 10 MW surplus can be supplied to the grid. Dried EBF can be utilised to generate the syn gas via gasification and combustion cycle (Figure-4)

Energy Malaysia is also looking to develop the strategy to engage local communities for the decentralized electricity generation from EFB as described in Figure-2

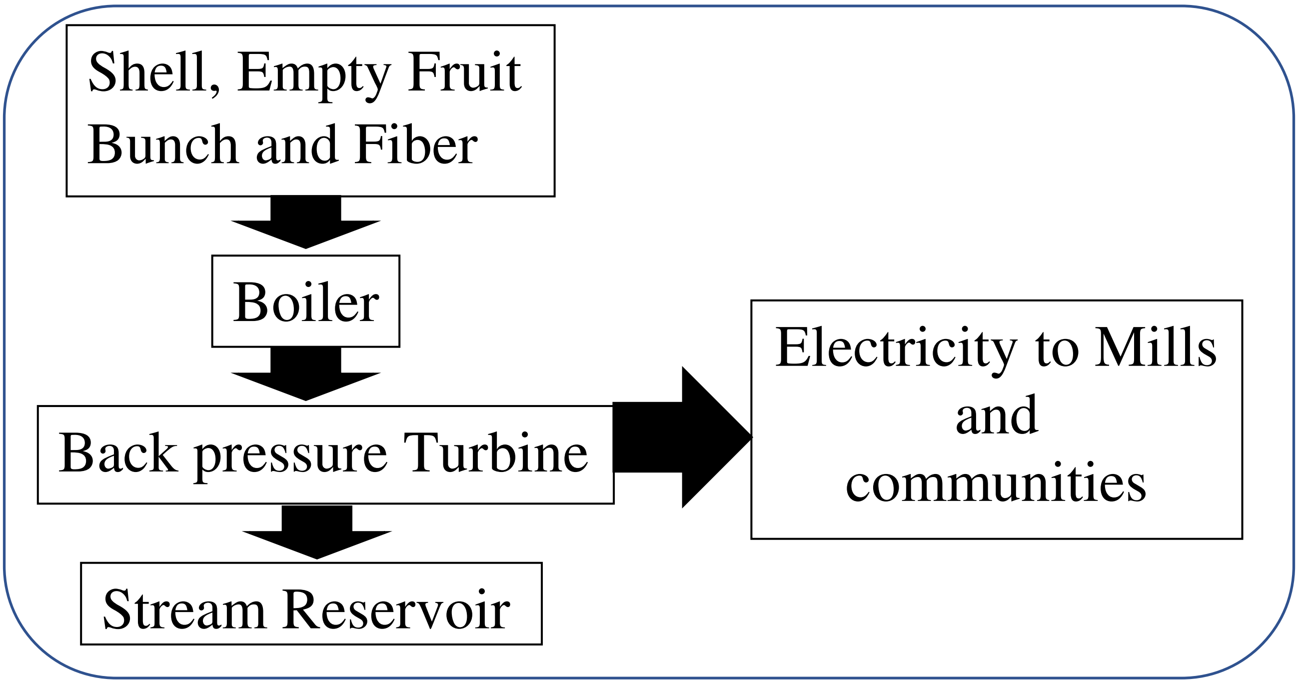

Figure-2 Decentralized electricity generation from palm oil mill plant to rural communities.

Process description

The as-received Empty Fruit Bunches (EFB) have a very high amount of moisture content almost 60 % (wt.) and hence need to be solar dried for about 28-30 hours. After drying, the moisture content remains less than 10 % in the EFB. The remaining weight of EFB after drying is around 134260 tonnes/year, which is available for electricity production. The composition of solar dried EFB is given in Table 1 [1]. The dried EFB is fed to the gasifier using a screw feeder. The Fluidized Bed Reactor (FBR) has been used for gasifying EFB in presence of steam into a medium heating value synthesis gas, which is further processed and utilized for generation of electricity in the power plant. The reactor operates in temperature range of around 850 ^{0}C. A bed of catalyst particles (such as olivine or nickel based supported catalysts) is used for cracking of tar compounds formed during gasification into gaseous products and also for better heat and mass transfer during the process. A cyclone separator is used for separating the solid particles (i.e., ash, catalyst and some amount of char) from the volatile products leaving the reactor. The char separated with ash is moved to another fluidized bed reactor for combustion at around 990 ^{0}C using preheated air. The heated catalyst particles, coming with char and ash is being recycled back to the gasifier, after separation from flue gases using another cyclone separator. These heated particles provide the necessary heat energy required in the gasifier. This twin-fluidized bed gasification scheme is known as Silva-Gas Process and is described in [2].

After removal from gasifier, the produced synthesis gas is then cleaned for removal of particulates, remaining amount of fly ash and acidic gases. However, before conditioning, the syngas is cooled for removing sensible heat and making it appropriate for downstream cleaning processes. A dry scrubber (hot gas filter) is used for separation of fines from the volatiles coming out of heat exchanger at around 500 ^{0}C. After this, the gas temperature is reduced to around 40 ^{0}C for allowing the separation of acidic gases such as H_2S and NH_3. The Selexol process is preferred for removal of these components from the synthesis gas and being discussed in [2]. The heat removed from hot gas being utilized for generation of low-pressure steam. This steam would meet the requirement of gasification process as well as other heating requirements such as EFB drying process (if required) and any solvent regeneration processes in the plant.

After gas processing, the synthesis gas is compressed using a 2-stage compressor to pressure around 10 bars. The energy required for cooling the synthesis gas between stages is provided for preheating of combustor air. The compressed gas is burned with compressed air from turbine compressor in a burner at temperature of around 1070 ^{0}C, and then the combusted products are utilized in the gas turbine for power generation. The gas from turbine outlet is mixed with the flue gas from the char combustor. The mixed flue gas is at temperature higher than 650 ^{0}C and used for power generation through heat recovery steam generator. The pumped water is used for taking the heat energy from the hot gas stream and the produced medium pressure steam is utilized for generating power using steam turbine, which operates at vacuum conditions for maximizing the output. The surface condenser is required for condensing the process steam in vacuum conditions and for reusing the condensed water into the heat recovery steam generator.

The flue gas from heat recovery steam generator is at around 170 ^{0}C and being used for generating hot water in a heat exchanger. This hot water can be utilized as a source of heat energy in absorption-refrigeration cycle for plant cooling/refrigeration requirements. Afterwards, the gas is being processed for CO_2 recovery using solvent such as MEA (monoethanolamine). The vent gases are sent to the stack, while the solvent is being regenerated for re-use. The recovered CO_2 is further dried, compressed and stored in a suitable location such as depleted oil field for Enhanced Oil Recovery. However, the details of Carbon Capture and Storage (CCS) process are not discussed in scope of this project, and can be exercised at later on stages depending on final requirements.

Assumptions

1. The overall plant is in steady state.

2. The solar drying of EFB is assumed.

2. The energy required for pelletizing the solar- dried EFB for better handling and transportation, and screw feeding is not considered here.

3. The gasification process is assumed to be in equilibrium at such higher temperatures and hence, an equilibrium model has been selected for modelling this process.

4. The gasifier is assumed to maintain isothermal conditions during the process.

5. Tar formation is negligible.

6. Formation of gaseous hydrocarbons other than methane, such as ethane, propane is negligible at such higher temperatures.

Aspen Plus simulation procedure

The RKS-BM (Redlich-Kwong- Soave with Boston Mathias) thermodynamic package in ASPEN PLUS VERSION 7.2 is used for this study. In simulation, the solar-dried EFB is devolatalized using a RYIELD reactor, which is usually the first step of any gasification process, and the pure carbon is distributed between gasifier and combustor as per the energy requirements of the gasification process. The pure carbon amount routed to combustor is taken around 20%, while remaining goes to gasifier section. In next step, the devolatalized EFB is fed into a RSTOIC reactor (in gasifier section) for conversion of available sulphur and nitrogen into H_2S and NH_3, respectively. As the gasification reactions are considered to occur in equilibrium at such high temperature conditions, a RGIBBS reactor is used for conversion of devolatalized EFB into gaseous products based on restricted chemical equilibrium approach. The reactions used for gasification are shown in Table 2. Steam is introduced in the reactor according to the biomass feed rate, as steam/biomass ratio controls the formation of carbon dioxide over carbon monoxide during the process. The cyclone separator for separation of unreacted char (which already being sent to combustor in simulation), ash and catalyst particles is shown using substream splitter in ASPEN. A RGIBBS reactor with Gibbs free energy minimization method and no specific chemical equations is used for combustion of unreacted char (assumed as pure carbon in simulation). Preheated air is used for combustion purposes. This char is combusted for raising the catalyst temperature. The heat energy required for gasification is being supplied by hot catalyst particles recycled back from combustor to gasifier. The near atmospheric pressure conditions are maintained in both gasifier and combustor.

The synthesis gas from cyclone separator is cooled and then sent to hot gas filter for particulates removal. The filter is being treated using a separator block in ASPEN. After separation, the gas is further cooled to low temperature for using in acid gas removal process. The Selexol process for separation of acidic gases is being considered as only a component separator block during simulation. The details with mass and energy balance of this process has not been discussed here.

The synthesis gas, after cleaning, being compressed using a 2-stage isentropic compressor with isentropic and mechanical efficiencies of 80% and 95% for each stage, respectively. The gas is then burned in a burner using compressed air from turbine compressor. The air compressor has an isentropic efficiency of 85% and mechanical efficiency of 97%. The burner is also being modelled as a RGIBBS reactor using Gibbs free energy minimization method with no specific chemical reactions. The combustion products at high temperature and pressure are fed to an isentropic gas turbine with a pressure ratio of 0.1 and isentropic and mechanical efficiencies of 90% and 98%, respectively.

The flue gases from gas turbine outlet and char combustor are mixed. The heat energy from this mixed stream is being taken by a pumped water stream to generate medium pressure steam. The medium pressure steam is used for generating power using a steam turbine, modelled using an isentropic turbine with discharge pressure of 0.06 bar and isentropic and mechanical efficiencies of 75% and 97%, respectively. The water pumps in plant have efficiencies of around 75%.

The flue gases from the heat recovery steam generator are further utilized for generating hot water for cooling/chilling plant needs through absorption-refrigeration cycle. The heat and mass balance calculation for hot water and downstream configuration has not been shown, and can be modelled at later on stages depending on final requirements

Results and discussion

The process flow sheet and plant layout are shown in Figure 1 and 2, respectively. The stream summary is provided in Table 3. Based on simulation results, the output of gas turbine is 9.43 MW (after removing the compressor power required by burner air), and of steam turbine is 4.48 MW. The parasitic power requirement by synthesis gas compressor and water pumps is around 3.5 MW. Considering power requirement of around 0.4 MW for other additional loads such as EFB pelletizing and feeding (not considered currently), the power available for grid supply is 10 MW. The simulation details can be found from the ASPEN input file, attached with other support files.

This design was considered keeping in mind the client's requirement about zero emission biomass power plant. The advantage of using twin-fluidized bed gasification scheme is no additional source of heat energy required for maintaining the temperature in reactor. The fluidized bed combustor, by burning the unreacted char from gasifier, provides the heat energy via hot recycled catalyst particles. The usage of dry hot gas filter over wet scrubbing process favours the extraction of higher amount of specific heat energy from synthesis gas before making it available at low temperature conditions for acid gas removal process. The discussion about Selexol process has not been done in terms of energy balance. However, based on its commercial application, it is considered as one of the optimum process for removal of the acidic gases. The heat energy of synthesis gas between 2 stages of compressor is used for heating of combustor air, causing the reduction of energy requirement for air preheating processes. The steam turbine generator is operated at vacuum conditions using surface condenser for energy maximization through the steam cycle. The remaining energy from flue gases is utilized for plant cooling/chilling requirements via absorption-refrigeration cycle, before sending to low temperature operation of carbon dioxide removal. The Carbon Capture and Storage (CCS) scheme proposed for removal of carbon dioxide from the flue gases going to vent, leads to zero emissions from the plant. This carbon dioxide could also be sold in the carbon trading market via different mechanisms for increasing the overall profit from the plant. However, the cost estimation for CCS scheme as well as Selexol process and absoprtion-refrigeration cycle has not been performed, and can be considered at later on stages depending on final requirements.

Cost Analysis:

Capital Cost Estimate

Capital cost of the plant is estimated based on capacity of the plant and cost of the equipment utilized in the plant. Capacity factor uses ration of the capacity of existing equipment to the new equipment. Moreover, this ratio is multiplied by the cost of the existing equipment to the estimated cost of the new equipment. All cost was estimated according the 1990 dollars [3] and inflation rate of 2% is included in the cost to estimates the current cost of the plant.

EFB preparation and drying cost

Solar drying for 28-30 hours is used for raw EFB to decrease its moisture content level from 60% (wt.) to 8.75% (wt.). There is no cost involved in this process. The remaining weight of the EFB after drying is 370.13 tonnes /day. All the calculations are based on remaining weight of EFB after sun drying.

Gasification Cost

The integrated gasification combined cycle plant was simulated in ASPEN Plus software. The gasifier operates at atmospheric pressure and 850 ^{0}C temperature. Several independent sources [4-6] estimated the cost of the similar system. It is assumed that two gasifiers of equal capacity will be needed. The gasification section includes gasifier, char combustor, char combustor cyclone, ash cyclone, catalyst make up hopper, catalyst surge pots, start-up burner and blower, and steam supply valves.

Gas clean-up cost

Combination of dry scrubber (hot gas filter) and acid gases removal process is used for gas clean-up. The estimated cost is modified based on capacity of the plant. The Selexol process for separation of acidic gases is being used for further gas clean-up. The details with cost analysis are not discussed here for the process of acidic gas removal.

Gas Turbine and Heat Recovery Steam Generator

Costs for the gas turbines utilized in this study are determined from published data [7] and [3] or from the manufacturer. Again, installation factors are slightly reduced because of the modular nature of the gas turbines and associated equipment. The cost of the HRSG turbine is estimated from published sources [8] and [3]

Steam Turbine System

The steam turbine cycle cost was determined from capacity cost ($/kW) using number of references [9]. This cost includes cooling tower, feed water pumps, condenser etc.

The capital cost data for the plant is provided in Table 4.

Operating and Maintenance cost

Operating and maintenance cost of the plant is calculated based on 80% capacity factor. Empty Fruit Bunch (EFB) is available as a raw material for no cost or very low cost. It was also assumed that 3 operators are required per shift for smooth plant operation. The summary of the operating cost per year is tabulated in Table 5. Carbon capture and storage is not calculated under the operating cost.

Break-even analysis:

Initial investment: $22,121

Operating cost per year: $2,845

Energy cost per unit in Malaysia: $0.053/unit [10]

Estimated energy production per hour: 10 MWh

Estimated energy production per year: 10 MWh*1000 kWh/MWh*24hours/day*365 days/year = 87600000 kW or units

Total income: 87600000 units*$0.053/unit = $4642800/year

Capital investment + (operating cost/year) * x years = (Total income/year) * x years

22121 + 2845x = 4642800x

x = 22121 / (4642800-2845) = 0.00476 years = 1.74 days

So, with the above capacity and costs, if we assume that the plant will operate 24 hours a day all year round, then the break-even point can be reached in 2 days and the plant will start making profit after that.

Efficiency calculation:

Higher Heating Value (HHV) of EFB: 19.643 MJ/kg [11]

Annual usage: 134260 tonnes/year = 134260000kg/year

Energy input = 19.643 MJ/kg*134260000kg/year = 2.63726918×10^9 MJ/Year = 83.63 MJ/s

Energy output = 10 MWh =10 MJ/s

Efficiency = output/input = 10/83.63 = 0.12 = 12%

Conclusion

Above case study describes the possibility of using 134260 tonnes/year of dried EFBs to sustainably produce power from a biomass power plant in Malaysia. The main design constraint that was achieved successfully here was zero-emission biomass power production. The detailed process for achieving such outcome has been outlined in detail and simulated using RKS-BM thermodynamic package in ASPEN PLUS VERSION 7.2. From the simulation results and cost analysis, it can be clearly seen that this plant can become profitable after just 2 days of operation. Although, not quite efficient but with freely available EFBs which can otherwise pose a huge risk to agricultural land, this biomass energy plant can prove to be a sustainable solution.

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Appendix

Table 1: Proximate and Ultimate Composition of solar-dried EFB [1]

Proximate Analysis:	
Moisture (wt%, as received after solar drying)	7.95
Volatile Matter (wt%)	83.86
Fix Carbon (wt%)	10.78
Ash (wt%)	5.36
Ultimate Analysis:	
C (wt%)	49.07
H (wt%)	6.48
O (wt%)	38.29
N (wt%)	0.7
S (wt%)	0.1
Ash (K, P, Mg, particulates, etc.) (wt%)	5.36

Table 2: Steam Gasification reactions [2]

$C + 0.5O_2 \Longleftrightarrow CO$	Combustion Reaction
$CO + H_2O \Longleftrightarrow CO_2 + H_2$	Water Gas Shift Reaction
$CO + 3H_2 \Longleftrightarrow CH_4 + H_2O$	Steam Methane Reforming
$CO + 0.5O_2 \Longleftrightarrow CO_2$	Combustion Reaction
$H_2 + 0.5O_2 \Longleftrightarrow H_2O$	Combustion Reaction

Table 3: Stream Summary

	Dried EFB	Steam To Gasifier	Combustor Pre-heated Air	Syngas (after cooling/ cleaning)	Compressed Burner Air
Temperature C	25	145	200	40	334.3
Pressure bar	1.05	1.05	1.05	1.05	10
Mass VFrac	0	1	1	1	1
Mass SFrac	1	0	0	0	0
Total Mass Flow , kg/hr	15326.48	11494.86	23000	16941.56	75000
Volume Flow, cum/hr	12.121	20992.14	31377.43	29436.14	13190.943
Enthalpy, Gcal/hr	-25.74	-36.241	0.98	-22.883	5.717
Density , kg/cum	1264.458	0.548	0.733	0.576	5.686
Mass Flow, kg/hr					
N2	0	0	17642.91	0	57531.225
O2	0	0	5357.091	0	17468.775
H2	0	0	0	1462.809	0
H2O	0	11494.86	0	0	0
CO	0	0	0	8416.996	0
CO2	0	0	0	7058.124	0
CH4	0	0	0	3.632	0
S	0	0	0	0	0
CL2	0	0	0	0	0
NH3	0	0	0	0	0

H2S	0	0	0	0	0
C	0	0	0	0	0
BIOMASS	15326.48	0	0	0	0
ASH	0	0	0	0	0

	Flue Gas (Gas Turbine + combustor)	Steam from HRSG	Flue Gas To Absoprtion-Refrigeration Cycle
Temperature C	671.6	480.2	176
Pressure bar	1.05	14	1.05
Mass VFrac	1	1	1
Mass SFrac	0	0	0
Total Mass Flow , kg/hr	116326.1	20000	116326.1
Volume Flow, cum/hr	319819.4	4906.063	144754.6
Enthalpy, Gcal/hr	-74.787	-59.906	-91.463
Density , kg/cum	0.364	4.077	0.804
Mass Flow, kg/hr			
N2	75174.13	0	75174.13
O2	2705.134	0	2705.134
H2	0	0	0
H2O	13080.82	20000	13080.82
CO	0	0	0
CO2	25366.04	0	25366.04
CH4	0	0	0
S	0	0	0
CL2	0	0	0
NH3	0	0	0
H2S	0	0	0
C	0	0	0
BIOMASS	0	0	0

ASH	0	0	0

Table 4: Capital cost estimate for the plant

Capital Cost	
Plant Section Description	Installed equipment cost $K including 2% inflation
Wood Drying	0 (Sun Drying)
Gasification	4943
Gas Clean up	1881
Gas Turbine	6221
HRSG	2678
Steam Cycle	4415
Compressor	1983
Net Capital cost	22121

Table 5: Estimation of operating cost for the plant

Annual operating cost	
Empty Fruit Bunch	No cost
Water ($/T)	92
Operating Labor	445
Supervision and Clerical	430
Maintenance cost	1736
Other operating cost	142
Net operating cost	2845

Figure 3: Process Flow sheet

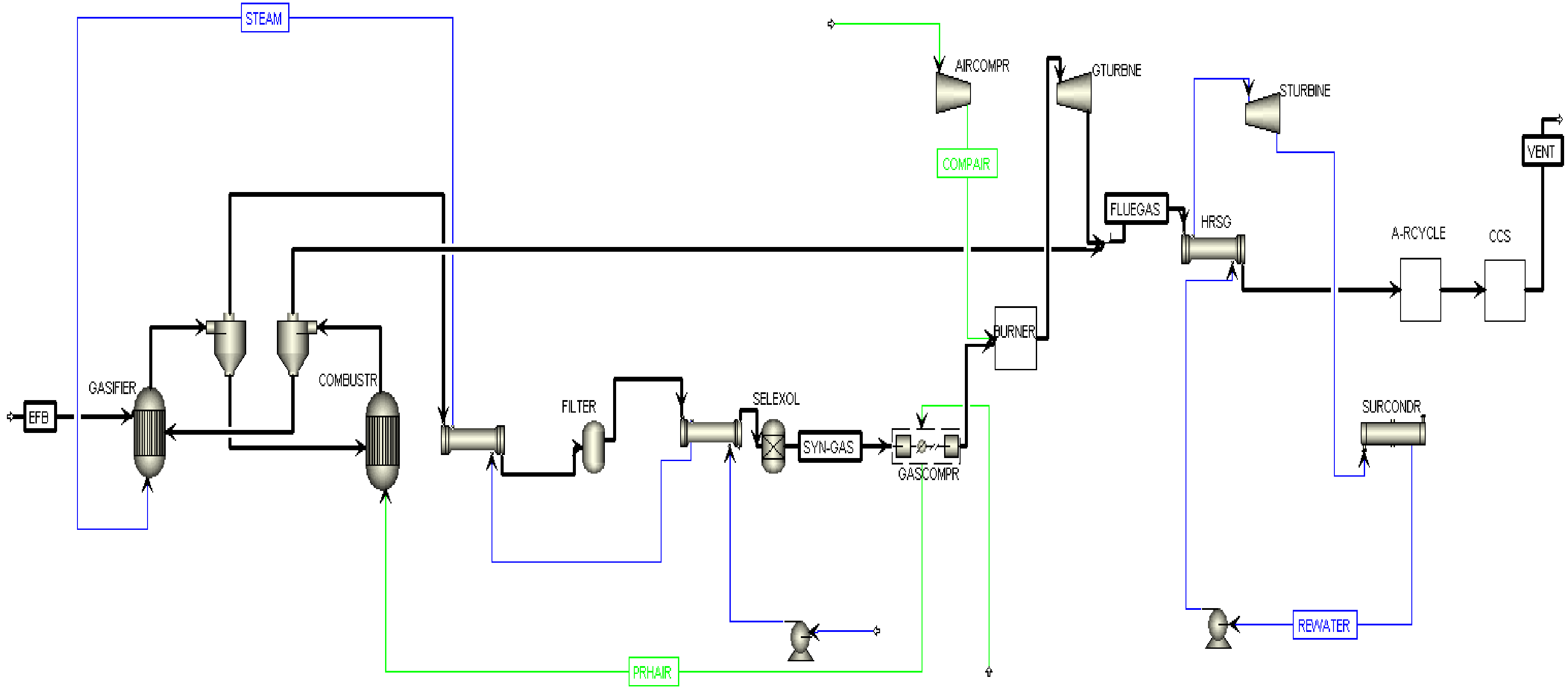

Figure 4: System Plant Layout