LAPPEENRANTA UNIVERSITY OF TECHNOLOGY
Faculty of Technology
Master’s Degree Programme in Chemical and Process Engineering

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Simulation and optimization of IGCC technique for power generation and hydrogen production by using lignite Thar coal and cotton stalk

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Supervisor: Professor Andrzej Kraslawski

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ABSTRACT

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Simulation and optimization of IGCC technique for power generation and Hydrogen production by using Thar lignite coal and cotton stalk

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The purpose of this study was to simulate and to optimize integrated gasification for combine cycle (IGCC) for power generation and hydrogen (H_2) production by using low grade Thar lignite coal and cotton stalk. Lignite coal is abundant of moisture and ash content, the idea of addition of cotton stalk is to increase the mass of combustible material per mass of feed use for the process, to reduce the consumption of coal and to increase the cotton stalk efficiently for IGCC process. Aspen plus software is used to simulate the process with different mass ratios of coal to cotton stalk and for optimization: process efficiencies, net power generation and H_2 production etc. are considered while environmental hazard emissions are optimized to acceptance level.

With the addition of cotton stalk in feed, process efficiencies started to decline along with the net power production. But for H_2 production, it gave positive result at start but

after 40% cotton stalk addition, H_2 production also started to decline. It also affects negatively on environmental hazard emissions and mass of emissions/ net power production increases linearly with the addition of cotton stalk in feed mixture. In summation with the addition of cotton stalk, overall affects seemed to negative. But the effect is more negative after 40% cotton stalk addition so it is concluded that to get maximum process efficiencies and high production less amount of cotton stalk addition in feed is preferable and the maximum level of addition is estimated to 40%. Gasification temperature should keep lower around 1140 °C and prefer technique for studied feed in IGCC is fluidized bed (ash in dry form) rather than ash slagging gasifier.

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Symbols, subscripts and acronyms

Symbols	**Definition**	**Units**
P	pressure	bar, atm, MPa
P_w	power	kW, MW, GW, TWh
S	particle size	µm
T	temperature	˚F, ˚C, K,
m	mass	Btu,kg, ton, megatonne billiontonne
$m^{\cdot}$	mass flow rate	kg/s, mg/s, ppm
v	volume	ft^3, m^3
$v^{\cdot}$	volumetric flow	cum/s, ft^3/s,
n	mole	kmole, lbmole
$n^{\cdot}$	molar flow	kmole/s
E	enthalpy	MW, kW
HHV	high heating value	Btu/lb, kJ/kg, MJ/kg, kCal/kg
W	heat duty of stream	kJ/kg
CV	calorific value	MJ/m^3
t	time	s, minute, hour

Subscripts

g	product gas
f	feed
flue	flue gas
net	net (overall)
aux	auxiliary
CCE	combine cycle efficiency
THE	thermal efficiency

Acronyms

wt	weight fraction
wb	wet basis
db	dry basis
ar	as received
HHV	high heating value
CGE	cold gas efficiency
HGE	hot gas efficiency
THE	thermal efficiency
CCE	combine cycle efficiency
IGCC	Integrated gasification combine cycle
ASU	air separation unit
FP	feed preparation
PR	particle removing
AGR	acid gas removal
HRSG	heat recovery and steam- generation unit
WGS	wet gas shift reactor
PSA	pressure swing adsorption
GT	gas turbine
ST	steam turbine

1 INTRODUCTION

The object of the study is the simulation and optimization of integrated gasification combined cycle (IGCC) technique for the production of electricity and hydrogen (H_2) by using Thar lignite coal and cotton stalk as feed stock. Optimization deals with the process efficiencies and environmental hazardous emissions. For the environmental hazard emissions, emission of nitrogen dioxide NO_2, nitrous oxide NO (collectively called as NOx), sulfur dioxide (SO_2), sulfur mono and tri oxide SO, SO_3 respectively (collectively called as SOx) in pound per million (ppm) level and carbon dioxide, CO_2 emissions in kg/s is considered. The emissions are optimized to acceptable level. For the process efficiencies production of syngas (carbon monoxide CO + hydrogen H_2), H_2 gas production, cold and hot gas efficiencies (CGE,HGE), combine cycle efficiency (CCE), overall thermal efficiency (THE), net power generation and auxiliary power consumptions etc. are considered.

Aspen Plus simulation software is used to develop model for the co-gasification of IGCC technique. For the gasification “shell entrained flow gasifier” is used with optimum maximum temperature is set to get the ash in molten, slag form while for the steam generation from the flue gas “combine cycle” is used. In this study, the simulation of whole IGCC with combine cycle for steam generation has been done for different compositions of biomass and coal to optimize the process for maximum attainable process efficiencies with less environmental hazardous emissions and costing of fuel. The main ideas of the thesis are the efficient utilization of cotton stalk with low grade lignite coal for the production of electricity and H_2 gas, to identify the best available feed mixture to get maximum process efficiencies with environmental consideration and efficient utilization of biomass, cotton stalk for gasification.

Thar lignite coal contains low mass of combustible material per mass of feed coal. With the addition of biomass in specific mass of feed stream, mass flow rate of combustible gas will increase. In order to study the impact of addition of biomass 5 mixtures with different compositions are selected. With the variation of ratios of biomass/coal in feed mixture, marked effects are observed in the process parameters including the consumption of air required for gasification, auxiliary power consumption, total net power production and efficiencies of product gas like CGE, HGE and TGE etc. Sensitive analysis is used to study these effects. For gasification, first feed mixture is pre-dried to desired moisture content before to introduce into the gasifier. Pre drying has greater effect on net power production and process efficiencies. In order to optimize the overall process, sensitive analysis is applied to attain allowed feed moisture. Similarly oxygen for gasification and ratio of biomass/coal in feed mixture, NOx and SOx emissions to accepted level etc. are optimized.

For gasification, shell gasification with dry feed ash slagging is selected in view of flexibility in moisture, and ash content in the feed streams, also it has more efficiency then the slurry feed even for the high moisture of the feed stream.

2 POWER PRODUCTION IN PAKISTAN

Among the South Asian countries Pakistan is leading in the shortage of electricity and short fall counts nearly 30% of the maximum or peak load demand. Currently the peak requirement is recorded to 19228 megawatt (MW) while installed capabilities to meet the demands are only 15232 MW whereas the demand is predicted to rise in future at a very high speed. The tremendous growth in population leads to demand rise to 9% while the supply of electricity power is 7%, also according to CIA World Factbook, Pakistan economy is increasing with the pace around 4.8 % so it is assumed that energy demand will rise to 22448MW by 2012 and if the pace remains the same then it would lead to 60000MW by 2020. Currently due to limited oil and gas reseves and political figth on hydel energy, Pakistan is facing shortage of power production and importing large quantities of oil and gas from Middle East countries. Although for the current use Pakistan has good gas resources but it is not enough for the need and for the future prospectus. These situations demand to find out the other sources for the production of electricity (Business Recorder, 2011; Hasan, 2008; Central Intelligence Agency, 2003; Adeel, 2007; Zeeshan, 2010).

2.1 Overview of electricity production in Pakistan

In Pakistan electricity is produced mainly by three sources, majorly dominant by fossil fuels (including oil and gas) followed by hydroelectricity and then by nuclear where as the contribution of coal is only 0.18%. The contrbiution of the sources has been shown in figure 1. Paksitan spends around US$ 11 billion per annum to import the oil for the production of electricity. Pakistan rely heavily on the oil and gas for the power production and keeping in view the increase of price of oil and gas and with the increase in demand of electricity, if Pakistan will continue rely on oil and gas then it will affect directly on the foreign reserves as well as the development of Pakistan. Therefore it is necessary to search new ways for the production of electricity (Pakistan, 2010; Khanji, 2008).

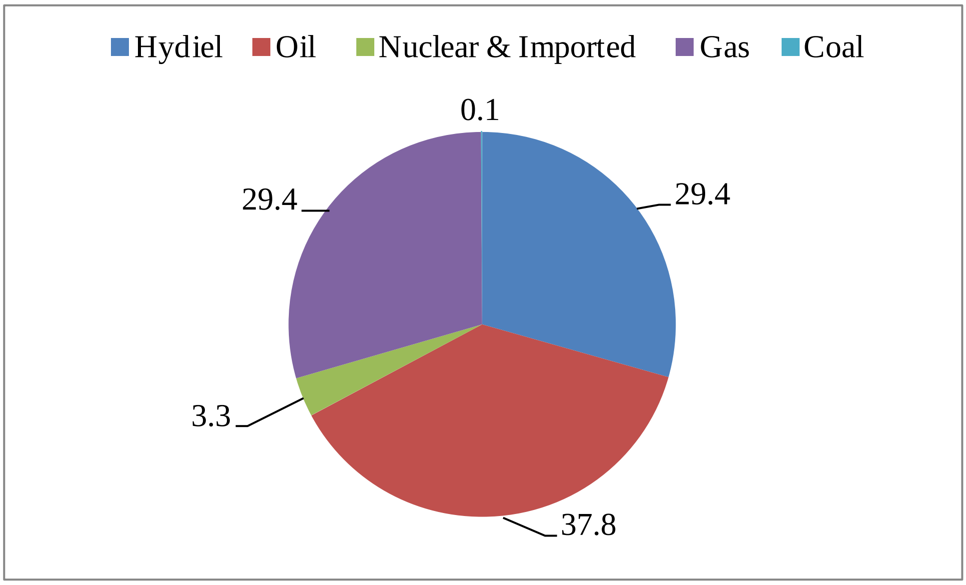


Figure 1 Contribution of sources in Pakistan’s power production (Pakistan, 2010)

2.2 Oil and gas in the production of electricity

According to “Pakistan Energy Year Book 2010”, in 2010 contribution of oil remained 37.8% in overall electric production (Pakistan, 2010). With the increase of transportation, population, electricity demand and other factors the consumption of oil increased during recent years and it is predicted to be increase more in future as well but unlike with the consistent demand, the oil production remained low resulting in the import of oil.

The demand of oil is subjected to increase around 17% for the period of 2010-2011 and expected to increase more than 19 milliontonne by the year 2017-2018, while the gap of demand and supply will be fulfilled by importing oil from Middle East countries (Adeel, 2007).

2.3 Gas in the production of electricity

For gas supply, although consumptions and demands are nearly same and currently there is not so much import but in future it is predicted that if Pakistan will use gas resources in the same way then it will have to import, especially from neighboring countries like Iran and Saudia Arabia etc.

During the year 2010, shortage of gas supply is observed in domestic and transportation sectors resulting of drop of gas supply for power production. For the year 2009, 48 % of total gas was supplied to power sector which reduced to 44% for 2010. This leads to shortage of power production and significant decline, 12% of power shortage is observed (Express Tribune, 2011).

2.4 Coal in the production of electricity

Unlike other countries, the contribution of coal in power production is very low in Pakistan. According to “Pakistan Energy Year Book 2010”, coal contribution remained only 0.1% of overall electric production for 2010 whereas coal generates about 40% of electricity globally. China alone contributes 78% coal for the energy production while USA 60%. In recent past due to some political tension between west countries and Iran, some standoff in Arabian Gulf, coal has got a special importance and with the development of IGCC to several byproducts and electricity, its importance cannot be denied. It is expected that coal consumption will increase by 75% from 2000 to 2030.

According to IEA estimation in 2008, with the current projects demand, potential reserves of oil will last for 41 years; natural gas can last for 67 years while coal can last for 192 years. These facts urge to utilize and focus more on coal rather than to rely on oil and gas. Pakistan is focusing more on coal utilization and with the 2030 plan, utilization of coal for power production will increase from current

200 mega watt (MW) to 19910 MW while consumption of coal in overall energy mix will increase from 5% to 19% by 2030 and this will further increase to 50 % by 2050. Pakistan has very large quantities of coal reserves and it is estimated to 185,457 million total, whereas alone Thar has 175,506 milliontonne coal reserves (Report Buyer, 2010; Raheem, 2008; Pakistan, 2010).

2.4.1 Thar Coal

Thar coal field was discovered by British Overseas Development Agency and Sindh Arid Zone Development in 1998. The coal field covers the area of 9000sq km and having the potential of around 175.5 billiontonne of coal whereas 2357 milliontonne have been measured with the total area of 358.5 sq km. Thar coal field can be divided into 4 blocks which is given by

1. (Block 1) Sihar Vikian-Varvai
2. (Block 2) Singharo-Bhitro
3. (Block 3) Saleh Jo Tar
4. (Block 4) Sanolba

Proximate analysis of Thar coal on as receive basis (ar) and weigth basis wt% of different blocks are shown in Table 1

S.No.	Area	Moisture %	Ash %	Volatile Matter %	Fixed Carbon %	Sulphur%	High Heating Value Btu/lb
1	Block 1	44.07	6.18	33.04	22	0.92	6398
2	Block 2	49.01	5.18	26.5	19.35	1.05	5780
3	Block 3	45.41	6.14	28.51	19.56	1.12	5875
4	Block 4	43.02	6.57	29.04	21.61	1.2	5971

Table 1 Proximate analysis of thar coal on (ar) basis (Mohammad et al., 2010).

The ASTM rank for Thar coal is from lignite to sub-bituminous with the mositure content 29.6-55.5%, volatile matter 23.1-36.6%, fixed Carbon 14.2-34.0%, ash content 2.9-11.5%, while sulfur content 0.4-2.9% where as high heating value (HHV) on moisture mineral material free basis (mmmf) is estimated to 6,244 - 11,045 Btu/lb (Pakistan, 2010; Anila Sarwar, 2011).

Thar coal has variations in the properties and varies from low ash content 2.9 to high value 11.5. High ash and mositure content of coal make the overall gasification uneconomical and with respect to operation not feasible. To avoid this hurdle it will be a good option to use co-gasification that is to gasify the coal having higher ash with biomass having low ash content.

Despite of the fact Thar coal has high moisture cotent and required pre drying for gasification, the high volatile component and low fuel ratio (ratio of fixed carbon/volatile matter) makes it suitable for gasification and so for power generation through IGCC. For good quality coal, the typical value of fuel ratio varies from 2.5 to 4 while lower rank coal has ratio of 1.5 or below to 1.5.

This theory is also supported by one research for different qualities of coal from higher to low grade in Japan. According to that research, low grade brown coal with HHVon dry basis 24280-29470 kJ/kg and fuel ratio less than 1.5 gasified easily than the high rank bituminous coal with HHV of 33910-35160 and fuel ratio 2.5 (Takao et al., 2011).

2.5 Overview of biomass for gasification

Biomass consists of polymers and organic compounds. Major compounds include lignin and carbohydrates (cellulose and hemicelluloses) etc. The ratio of ingredients depends on species and so as the resulting properties also vary with the species. Normally, biomass having low ash is preferable for gasification but for others biomasses like straws and grasses etc. have significantly higher ash content. They can be employed easily with co-gasification.

The conversion of organic compounds into fuel gas mixture is also very good. For typical biomass around 80-85% thermodynamically efficiency is reported. Biomass gasification produces clean fuel gas which can be used in IGCC and efficiency can be further increased by the use of combine cycle. The impurity in the product gas after gasification can be easily removed. These overall advantages make it better for gasification rather than combustion (Arkansas Economic Development Commission, 2010).

2.6 Biomass for power production

Biomass continues to gain more attention as a fuel alternative and great renewable energy source for power production. According to REN 21 report, the total power capacity by biomass was estimate to 62 gigawatt (GW) with United State leading to 10.4 GW whereas in the European Union biomass usage increased to 10.2% for the period between 2008-2009, and total production is estimated to 87.4 Tetra watt hour (TWh). It is forecasted that by 2020, biomass demand will touch to 44% and the major demand will be in energy sector. Similarly in USA four times more energy production is forecasted by utilizing biomass (Heather, 2011).

2.7 Prospectus of biomass for the production of electricity in Pakistan

Pakistan has great potential of biomass in form of crops residues, woods and wastages including (animal, human and municipal waste). According to one report, every day around 50 kilotonne of solid wastes and 1500 m^3 of woods are generated daily in the country where as 225 kilotonne of crops residues are estimated daily (Khanji, 2008).

2.7.1 Municipal wastages

As discussed earlier on average daily 50 kilotonne solid wastages are being produced in Pakistan whereas only in Karachi more than 7 kilotonne of solid wastages are produced and of which 60% dumped to open air away from city while 4 kilotonne is used. If they are properly land filled then they can produce great amount of energy. In UK about 28 megatonne of wastages produce daily and by recovering only 11% of this UK is producing 190MW (Nayyer et al, 2005).

2.7.2 Poultry farm and other animal processing units

In Pakistan there is vast network of poultry farm. According to one report by Pakistan Poultry Association (PPA), there are more than 15 million layer-chicken and 528 megatonne broilers chicken birds produced in 2003. But it is more important to note that just around maximum 10 % poultry farm have the membership, so actual figure will be much more than of it and with the time span the growth has increased tremendously. According to one unofficial survey in 2005, only in Karachi more than 500 kilotonne poultry wastages/year have been produced. In UK, 400 kilotonne poultry wastages/year produce 38.5 MW (Nayyer et al, 2005).

2.7.2 Wheat straw

Wheat is the largest crop of Pakistan. For the period of 2009-2010, total production of wheat was recorded to 23.31 megatonne while for the period 2010-2011 production is forecasted to 24 megatonne (FAO, 2011)

According to one report, by gasification 1 pound of straw produces 23.9 ft^3 of producer gas with HHV 7750 Btu/hr while average calorific value is recorded to

125 Btu/ft.^{3} On average one 1 acre of wheat land produces about 3000 pounds of wheat straw while 1 acre of wheat land can produce 71700 ft^3 of product gas (Sadaka, 2008) .

Application of wheat straw for gasification is very much limited due to presence of high amount of ash, 14% maximum on dry basis (db). Ash mainly consists of potassium and chlorine which are not suitable for fuel in power production. But the effect can be minimized by co-gasification with other biomass or with coal and by varying process parametres: including temperature of the reaction, variation in steam and air flows etc (Vera et al., 2010).

2.7.3 Rice husk

Rice is ranking third largest crop for Pakistan's after wheat and cotton. Total production of the rice for period of 2010-2011 is estimated to 5.7-6.1 megatonne where as exported 4.5 megatonne. For the period 2009-2010, the production is recorded to 6.7 megatonne. (dawn, 2011)

Rice husk is the byproduct of rice industry which is obtained after the separation of brown rise from paddy. It contains high percentage of volatile which makes it energy efficient while although the ash is higher and maximum to 20 wt% but ash is mostly consists of silica which is environmental friendly thus all in all rice husk is very good option for gasification (Kuen et al.1998)

In Indian state Bihar, a mini power plant is generating 1 kwh of electricity by using 1.5 kg of rice husk. On average the weight of husk produce is around 20% means 1 ton paddy per hour rice mill can produce 200 kg of husk while the exact value depend on nature/or variety of rice where as the heat generated for complete combustion is about 3000 kCal/kg and requires about 4.7 kg of air/kg of rice husk. By gasification of rice husk using equivalence ratio (ER) about 0.3-

0.4, product gas having heat energy of 3.4-4.8 MJ/m^3 can be produced (Belonio, 2005)

2.7.4 Cotton stalk

Cotton and related to cotton industries contribute 61% share of Pakistan total export. Total production of the cotton for period of 2010-2011 is estimated to 14.01 million bales with total 3.2 million hectare area for its cultivation while for the period 2009-2010 the production is recorded to 13.36 megatonne with the same irrigated area as of 2010-2011 (pccc, 2011).

In India on average 3 metric tons of cotton stalks are produce for 1 hectare of the cotton field where as the heating value is recorded to 17.40 MJ/kg on (wb) with moisture 12% and the maximum ash content for cotton stalk is noted to 7%. The problem of cotton for gasification is its bulky nature and highly branchy. This leads to difficulties in transportation and storage. This problem can be solved out by densification (Tandon et al, 2009).

3 GASIFICATION

Gasification technologies are getting a lot of importance to produce environmentally clean and energy efficient power generation by using a variety of fuels including: coal, biomass, oil and gas etc (Moreea, 2000). Gasification is a process which involves the conversion of carbonaceous materials (solid fuels) into combustible or synthesis gas (gaseous fuels) by partial oxidation. By principle, it involves the series of chemical reactions of carbon present in the biomass or in the coal with air, pure oxygen (O_2), carbon dioxide(CO_2), or the combination of these at temperature around 1000 °C or higher and produces gaseous fuel which then can be used to produce heat and electricity or as a feed material for the synthesis of other products like, methane(CH_4), H_2, ammonia(NH_3), sulfuric Acid (H_2SO_4) etc (John et al, 2005).

Unlike combustion, gasification involves the partial oxidation. Gasification can be termed as incomplete combustion due to requirement of less amount of O_2 which is 50-70% less than actual theoretical amount of O_2 required for complete combustion (John Rezaiyan, 2005). Incomplete combustion produces product gas having different pollutant formations than combustion. In gasification sulfur converts to hydrogen sulfide (H_2S) rather than SO_2, while nitrogen (N_2) converts to NH_3 rather than NOx formation in pure combustion. Product gas is the combination of CO, H_2, CO_2, water (H_2O), N_2, CH_4, other gaseous products, tar, char and Ash etc.

Gasification involves the reduction of O_2 to the formation of CO_2 or H_2O and reduction in carbon to hydrogen (C/H) mass ratio. As the gasification is the partial oxidation process so the amount of the CO and H_2 is dominant while CO_2 and H_2O are produced in less quantity. The quantity of N_2 and heating value of product gas depends on the oxidant. If the oxidant is purely air and/or steam then the product gas will have lower calorific value (CV) in the range of 4 and 6 MJ/m^3 (107-161 Btu/ft^3). The lower CV is due to presences of the N_2 in the product gas which dilute it and lower the CV. With pure O_2 and/or steam used as oxidant, resultant product gas having medium CV in the range of 10 and 20 MJ/m^3 (268-537 Btu/ft^3). It is due less amount of N_2 in product gas. The highest CV gas is produced by using natural gas 37 MJ/m^3. Low CV product gas can be used as industrial fuel and for power generation while medium gas can be used as fuel gas, raw material for the production of NH_3, methanol and gasoline etc. and for power generation. So both low and medium CV producer gas can be employed in IGCC technique. O_2 has more advantage than air used as oxidant. With O_2 more coal/biomass converts into product gas higher CV which is lower with air. But for the production of pure O_2, additional equipments and units are required like air separators and compressors etc. which is assumed to utilize 10-15% of the gross power generated (Marano et al., 2002; Moreea, 2000).

The composition of the product gas depends on several factors including:

1. Characteristics of the fuel (fuel composition, extent of reaction and moisture content etc.)
2. Process parameters (operating temperature and pressure etc.)
3. Oxidizing material (O_2 or Air or combination of O_2 with steam etc.)
4. Mode of flow of fuel and oxidizing materials in the gasifier that is (co-current, counter current etc.)
5. Type of gasifier etc. (Mustafa et al., 2009).

Therefore it is very hard to define the product gas composition theoretically

3.1 Gasification Phenomenon

Gasification process consists of mainly three steps which occur in sequences.

1. Pyrolysis
2. Gasification
3. Combustion.

All these steps occur in series and there are no sharp boundaries among them. For biomass or other feed having higher moisture, normally dehydration (by preheating or drying) is required before gasification.

Feed streams having higher moisture in the range of 25 to 60 % and even for some of biomass moisture content found to be 90 %, if they will be applied directly to gasifier then it results to great loss of energy to overall process. First energy will consume to dry the moisture content and then to gasify the dried biomass. To remove one kg of water from biomass 2260 kJ energy is required which decreases the energy efficiencies of the process. To make the process energy efficient, biomass is normally preheated or dried to moisture content range up to 10 to 20 % and then introduces in the gasifier. Remaining 10-20% moisture is removed in the gasifier where the heat from the exothermic reactions increases the temperature of the gasifier and at 100 °C remaining water detached from the biomass and after that with the increase of the temperature the volatile attached with the biomass started to devolatize which continues to the temperature 200 °C. There are several reactions involved in gasification and some important reactions are discussed below.

Carbon reactions

$C + CO_2 \leftrightarrow 2CO$ 172 kJ/mol (1)

$C + H_2O \leftrightarrow CO + H_2$ 131 kJ/mol (2)

$C + 2H_2 \leftrightarrow CH_4$ −74.8 kJ/mol (3)

$C + 0.5\ O_2 \rightarrow CO$ 111 kJ/mol (4)

Oxidation reactions

$C + O_2 \rightarrow CO_2$ −394 kJ/mol (5)

$CO + 0.5O_2 \rightarrow CO_2$ −284 kJ/mol (6)

$CH_4 + 2O_2 \leftrightarrow CO_2 + 2H_2O$ −803 kJ/mol (7)

$H_2 + 0.5\ O_2 \rightarrow H_2O$ −242 kJ/mol (8)

Shift reaction

$CO + H_2O \leftrightarrow CO_2 + H_2$ −41.2 kJ/mol (9)

Methanation reactions

$2CO + 2H_2 \rightarrow CH_4 + CO_2$ −247 kJ/mol (10)

$CO + 3H_2 \leftrightarrow CH_4 + H_2O$ −206 kJ/mol (11)

$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O$ −165 kJ/mol (12)

Steam-Reforming reactions

$CH_4 + H_2O \leftrightarrow CO + 3H_2$ 206 kJ/mol (13)

$CH_4 + 0.5\ O_2 \rightarrow CO + 2H_2$ −36 kJ/mol (14)

(Prabir, 2010)

3.2 Pyrolysis

Pyrolysis involves the thermal decomposition of biomass/coal large hydrocarbons into smaller gas molecules. In this phase there is no significant chemical reaction takes place with the oxidant material. For biomass, hemicelluloses, cellulose and lignin break into char, tar and volatile (Mustafa et al., 2009).

Char produced by biomass is normally not pure carbon but contains few amount of hydrocarbon mixture. There is also a basic difference in the char produced by biomass and coal gasification. Biomass char has higher porosity in the range of 40 to 50 % whereas the coal char has lower porosity in the range of 2 to 18% (Prabir, 2010).

3.3 Gasification

Gasification involves all major chemical reactions of the whole process. In the presence of oxidizing materials, reaction takes place between the available hydrocarbons and oxidizing materials. The available char reacts with the oxidizing materials and produces product gas and ash, while the destination of tar depends more on the temperature and nature of the gasifier. At higher temperature, in moving bed and entrained flow gasifier, tar starts to cracking and re-polymerization. If tar cracks then it results into gas and light oil while on re-polymerization it converts to either char or gas or in heavy tar. For low temperature operations, like in moving bed gasifiers the exit gas temperature is lower so tar exits with the exit gas but for none slagging or dry ash gasifiers like British Gas Lurgi (BGL), they can be recycled to process and can be further gasified (Moreea, 2000).

The major important reactions occurs in gasifications are given and discussed below.

3.3.1 Volatile combustion reactions

These are the oxidation reactions and release a lot of energy which is sufficient for the whole endothermic reactions in gasification

$$C + 0.5\ O_2 \rightarrow CO \qquad (4)$$

$$CO + 0.5\ O_2 \rightarrow CO_2 \qquad (5)$$

$$H_2 + 0.5\ O_2 \rightarrow H_2O \qquad (8)$$

(Prabir, 2010).

3.3.2 Boudouard reaction model

The reaction of char and CO_2 is known as boudouard reaction which is given by reaction 1. The reaction is endothermic and by increase in temperature favors the forward reaction yielding more production of CO. As compare to reaction number 2 and 3, it is much slower (Prabir, 2010).

3.3.3 Water-gas reaction

The reaction of char with H_2O is known as water gas reaction.

$$C + H_2O \leftrightarrow CO + H_2 \tag{2}$$

As the reaction is endothermic so with the increase in temperature, reaction will go in forward direction yielding more production of CO and H_2, but it is also true presence of H_2 affect char gasification and it is studied that in the presence of 30% H_2, char gasification can be reduced to maximum factor of 15 (Prabir, 2010).

3.3.4 Methanation reaction

Methanation reaction involves the formation of CH_4, by following reaction

$$C + 2\,H_2 \rightarrow CH_4 \tag{3}$$

For complete conversion of carbon in feedstock, boudouard and water gas reaction can be summed up into two homogenous reactions given by CO shift reaction and shift reforming reactions (Prabir, 2010).

3.3.5 CO Shift Reaction

The CO shift reaction involves the production of H_2 at the expense of CO, it is given by,

$$CO + H_2O \leftrightarrow CO_2 + H_2 \qquad (9)$$

It is slightly exothermic and affected by temperature. With the decrease in temperature the reaction goes to forward direction resulting production of CO_2 and H_2 where as the pressure does not affect so much. The rate of reaction is very high above 1000 °C and achieves equilibrium rapidly while at lower temperature it has higher equilibrium constant resulting higher yield of H_2 but with lower rate of reaction. With the use of catalyst the rate of reaction can also be increased (Prabir, 2006).

3.3.6 Steam Reforming reaction.

The reaction of steam and methane is given by

$$CH_4 + H_2O \leftrightarrow CO + 3\ H_2 \qquad (13)$$

which indicates that reaction is pressure dependent and with the decrease in pressure, reaction will go in forward direction and there will be more syngas formation also as the reaction is endothermic so with the increase in temperature the reaction will go in forward reaction. For that reason gasification is favorable at high temperature and pressure (Prabir, 2010; Mustafa et al., 2009).

3.4 Char combustion

Most of the reactions involved in the gasification are endothermic while the heat energy involved in the gasification are provided by the char combustion, given by reaction number 4 and 5

$$C + 0.5\ O_2 \rightarrow CO \qquad (4)$$

$$C + O_2 \rightarrow CO_2 \qquad (5)$$

(Prabir, 2010).

3.5 Gasification technologies

Gasification technologies are based on the types of gasifiers used which are discussed below.

Mainly there are three major gasification techniques.

- Moving bed or fixed bed gasifier
- Fluidized bed gasifier
- Entrained flow gasifier.

3.5.1 Moving bed gasifier

Moving or fixed bed gasifier operates on counter direction flow operation. Carbonaceous feedstock is introduced from the upward direction while oxidizing agent is introduced from bottom of the gasifier, resulting maximum interaction for the reaction. While coal is moving downward the temperature increases so first it will devolatize, then gasified and combusted in the bottom section with the upcoming oxidant material. The residence time for the gasification is 0.5 to 1 hour while pressure remains 30 bar to 100 bar where as tar produces in dry or in molten form depending on the temperature profile. If the gasifier temperature is within range of 1200-1300 °C, the tar will be in dry (non slag) while if the temperature 1500-1600 °C or higher resulting tar will be in slag form (non dry). Steam requirement also depends on the tar condition. For the slag ash process, it requires normally more steam than the dry ash (Moreea, 2000; Toshi'ichi et al., 1991).

3.5.1 Fluidize bed gasifier

In fluidized bed gasifier, carbonaceous feedstock of the particle size in the range of <5-6 mm is introduces along with the supporting materials (including sand, char or there mixtures). They both are fluidized by the incoming oxidant and product gas. Unlike moving bed gasifier where heat profile vary with respect to the length of the gasifier, in the fluidized bed temperature of the feed coal/biomass changes very fast, resulting drying, devolatization and gasification etc. thorough out the reactor. The volatile released is cracked and the product gas contains nearly zero or very few concentration of the tar and other hydrocarbon. Normally temperature inside the gasifier remains 900-1100 °C. The temperature range is kept lower to avoid the ash to slag formation and defluidization of the bed (Moreea, 2000).

For the IGCC normally higher temperature and pressure is more efficient. In order to utilize fluidized bed, with IGCC, two techniques can be used. Ash agglomeration, developed by Institute of Gas Technology (IGT) and second one is modification to High Temperature Wrinkler (HTW) gasifier developed by Rheinbraun. Ash agglomeration involves the agglomeration of small molted ash which sticks together to form larger and denser agglomerate which cannot be fluidized and then come down to an extractor place particularly for their removal (Toshi'ichi et al., 1991).

3.5.3 Entrained flow gasifier

In Entrained flow gasifier, the carbonaceous feedstock is introduced along with the oxidizing materials from the bottom of the gasifier at higher temperature from 1200 °C to above, with the particle size of 75µm. Normally the temperature and pressure of the gasifier is kept higher for the maximum conversion of the char and for the slag formation of the tar while the residence time is of few seconds. Fluxing agent is also added into gasifier to lower the melting point of the slag which then removed from the bottom of the gasifier easily. This gasifier is suitable for all types of coal including bituminous coal which creates caking problem (Moreea, 2000; Toshi'ichi et al., 1991).

The major advantages of entrained flow gasifier can be summarized as

- To handle all kinds of coal as feed including high quality bituminous and low quality like lignite coal with high moisture and ash content.
- Product gas (or syngas) contains negligible amount of tars and oil.
- Carbon conversion is higher.
- Lower methane formation which is good for the synthesis gas related products.
- Higher temperature operation causing high reaction rate and so high product output.rate.

Although higher temperature operation has advantages but it also involves relatively higher amount of O_2 to attain the higher gasifier temperature (as slagging temperature is normally higher). The high outlet temperature results into more conversion of carbon chemical energy to sensible heat which in turns raises the reactor temperature but this involves in the decline of product gas efficiencies including CGE and HGE.

Depending on the operation entrained flow gasifier can be further divided into three major technologies, they are GE Energy (formerly Texaco), Shell and E-Gas (Conoco Philips).

In comparison to shell and GE operations the basic differences are feeding of coal into gasifier and ash removal system. In GE the coal is slurrized by using water typically 35% w/w where as in shell the coal feed is dried typically moisture range to 2%. In GE, slagging along with product gas send for quenching, taken out from gasifier and then droped into box attached to bottom of the gasifier. Finally, it cool down, recover the energy and send to cleaning section from there it can be recycled or taken out from system. In shell process, slag at gasifier temperature is recycled to feed after grinding. In shell, initially cooled syngas is employed to decrease the product gas temperautre where as in GE either the combination of radiant and convection cooling is used or with water, quenching is done. With the addition of water in GE, the ratio of H_2/CO in the product gas remains near to 1 where as in shell the ratio remains in the range of 0.5 (Maurstad, 2005).

3.6 Co-gasification

Co-gasification deals with the joint conversion of at least two carbonaceous fossil fuels into useful gas along with the release of heat energy. Co-gasification follows the same technique of the gasification with the advantage of the addition of the biomass (normally waste biomass is employed to make useful use of it). Willem-Alexander Power Plant at Belgium and ELCOGAS power plant in Spain etc. are the examples of the cogasifcation plant. Co-feeding gives the better ratio of H_2/CO which decline with the addition of biomass leads to increase of carbon conversion along with the increment of CGE. The advantage of co-feeding is in improvement in synergetic effect. But the ratio of co-feeding is very critical for the composition of the flue gas and finally for syngas (Luis, 2011)**.**

An experiment is conducted for co-gasification of birch wood and daw mill coal and results showed that with the co-gasification the reactivities of fuel increases along with the product gas while decline in tar and ammonia formation (Collot et al 1999).

3.7 Worldwide gasification

According to U.S. Department of Energy's (DOE) 2010 worldwide gasification database, there are total 144 plants are operating with 412 gasifiers and with the capacity of producing 70817 MW of syngas. While currently 11 plants, with 17 gasifiers, are under construction and along with it 37 plants, with 76 gasifiers with capacity of about 51288 MW are in planning and in designing stage which hopefully will be in operation in 2011-2016. So it is expected that by the end of 2016, total 192 plants with 505 gasifiers and capacity of 122106 MW will be in operation worldwide.

Currently on the utilization of product (syngas) by gasification, production of chemical is dominant with 45 % while production of liquid fuel is 38% followed power production 11% and then gaseous fuel 6%.

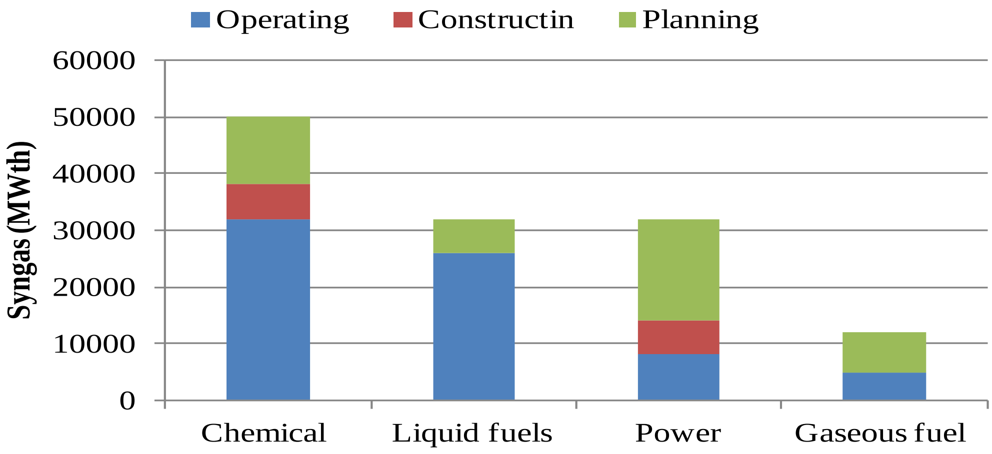


Figure 2 Statistic of production of syngas (MW) for different industries (U.S. Department of Energy National energy technology laboratory, 2010)

It is very much clear from figure 2 that after 5 years with the additional plant in operation, power plant will lead to 38% followed to chemical production 28%, gaseous fuel 18 % and liquid fuel 17%.

On the basis of primary feed for the gasification, coal is leading with 51% share while petroleum provides 25 % followed natural gas 22% and then Petcoke (U.S. Department of Energy National energy technology laboratory, 2010).

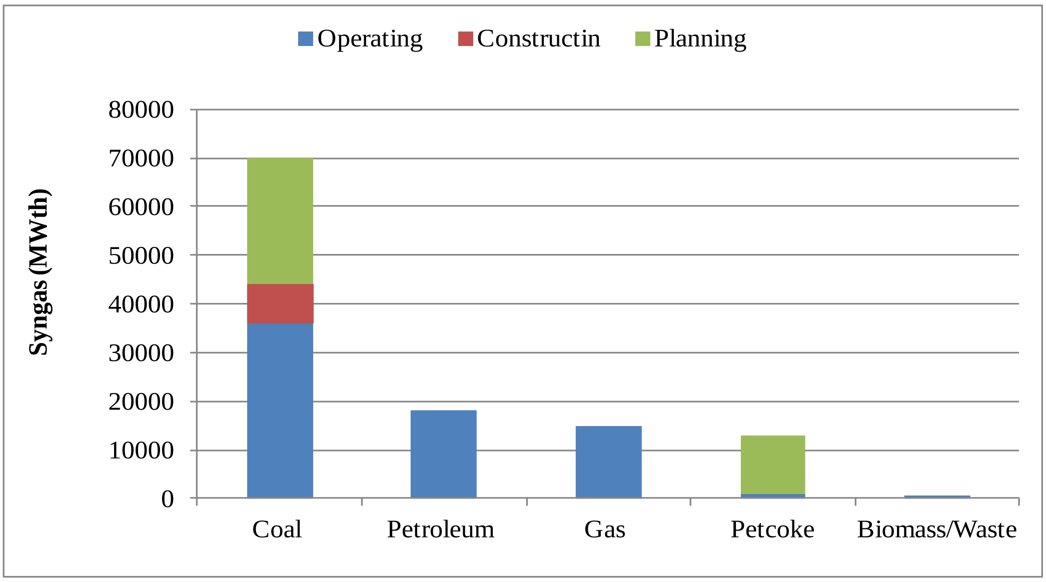


Figure No. 3 Statistic of usage primary feed for the production of syngas (MW) by gasification (U.S. Department of Energy National energy technology laboratory, 2010)

From the figure number 3, it is concluded that contribution of biomass/waste as a primary feed for the gasification is very low but today with the environmental and economical point of view co-feeding of biomass with coal or with other feed will dominant in near future.

3.8 Integrated Gasification Combined Cycle (IGCC)

IGCC stands for integrated gasification combine cycle, where combine cycle system involves use of combination of gas and steam turbine. Now a days due to less complexicity of combine cycle, it is considerd to most desireable for power generation and one of the most efficient power production technologies for the carbonaceous feedstock (Toshi'ichi Takematsu et al., 1991). By using IGCC technique, not only power can be produced but it also produces steam, H_2 and other useful products. Sulfur is also produce as valuable product with marketed quality (Zhu, 2004).

IGCC has higher efficiency than the typical steam cycle and also the costing for CO_2 capturing is lower than coal combustion system. IGCC has proved to be more advantageous over traditional method of power production by combustion. It produces smaller quantity of solid residues, producing less environmental hazards, economically feasible, and lower water consumptions etc.

Currently its cost is about 20 % more than the typical conventional plant but its efficiency is around 45% which is expected to increased to 50% by 2020 and Department of Energy (DOE) of the U.S.A is targeting to enhance the efficiency of combine cycle with IGCC to 52 % on HHV which is much more than modern supercritical coal fired power plant. Conventional coal fired plant maximum efficiency is estimated to 40% while normal efficiency is 32% whereas IGCC power plant usually has efficiency around 42% and more than it. But for the second generation of IGCC, it is predicted will be has same capital cost as conventional pressurized combustion plant with great process efficiency. According to Wood GC, capital costing for IGCC with carbon capture sequestration (CCS) is lower as compare to conventional production of electricity by coal combustion and reported to saving of dollar 200-400/kW. But this calculation is based on 90% of carbon capturing if it increase to greater than 90% than cost will be more (Toshi'ichi et al., 1991; Prabir, 2010; David et al., 2011)

A typical flow diagram of IGCC is shown in figure 4. In the flow sheet mainly three sections are discussed including gasification, combine cycle and air separation. Air separation is use to produce pure O_2 to get medium CV gas. Coal and/or biomass is send to gasifier where it is reacted with oxidizing material, in our case it is 95% pure O_2. Product gas produces in gasification is then passed through cleaning steps and cooling to get steam and then finally burn into gas turbine combustor. The power acquired from the system is the summation of steam and gas turbines.

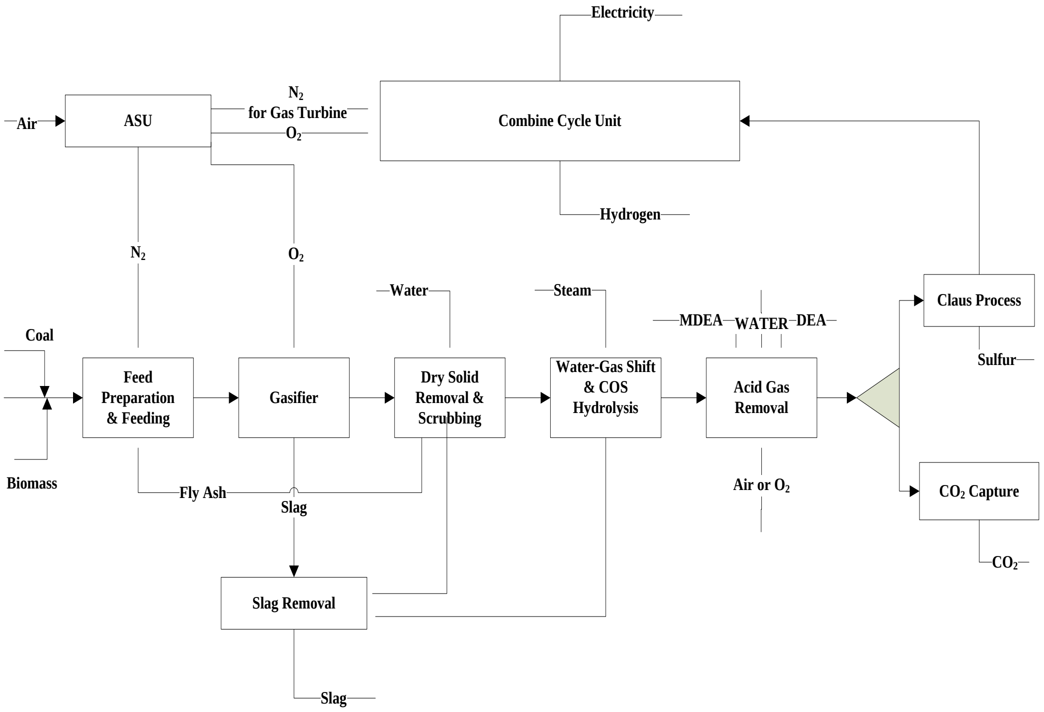


Figure 4 Flow sheet of IGCC with CO_2 captures (Po-Chuang Chena et al.,2010).

There are two typical scenarios of IGCC technique, one with post CO_2 capture the second is without post CO_2 capture.

In the post CO_2 capture, the clean gas contains higher amount of CO (around 56-62%) is send for shift reaction (for the production of CO_2 and H_2) which then proceed by the removal of CO_2 and compressed for the storage and sequestration, while clean gas is used for H_2 (99.8% purity) and power production. The post capture of CO_2 is economical at higher pressure rather than after combustion to capture CO_2 where as in the second scenario the clean gas is send directly for the power production in Gas Turbine (Liang-shih, 2010).

4 MODELLING AND SIMULATION OF IGCC

IGCC Process can be divided into three functional blocks.

1. Raw material (biomass and/or coal) preparation and gasification block.
2. Gas cooling, cleaning and removal of undesired products block.
3. Power generation block.

4.1 Raw material processing and gasification

Raw material processing starts with crushing, grinding of the feed material and then product gas formation after gasification. They are discussed below.

4.1.1 Air separation unit

Normally in IGCC O_2 is used with purity 95-99% by volume. Air separation unit is used to purify the O_2 to the desirable range and then send for gasification while N_2 is used in pre-drying of the feed, to carry the feed to the reactor and remaining N_2 is utilized in Gas turbine to lower the combustion temperature. Air separation is done by using distillation and absorber column with high and low pressure to get desired purity (Liang-shih, 2010). The flow sheet is shown in Appendix I (1/11).

4.1.2 Drying of high moisture of feed and convey to gasification

In the first step the feed is broken down to the particle size 75 µm and then dried to desirable range and after pressurizing to the gasifier pressure, it is send for the gasification by O_2/or Air (Liang-shih, 2010).

Too high moisture and ash content normally requires larger volume process equipments in order to handle the larger moisture due to lower energy and dense feed. With the increase of moisture and/ or ash content the demand of pure O_2 also increases to maintain the operating temperature. By increasing O_2 the more of heat available in coal will convert into thermal heat to maintain the gasifier temperature and so cold gas efficiency will reduce. By increasing O_2, overall process efficiency decreases (Maurstad, 2005).

For shell gasifier with high grade coal and dry feeding system, normally moisture content keep low below 2% but for low grade coal with very high moisture content, it can be managed to moisture content less than 10% (Ke Liu et al., 2010).

4.1.3 Gasifier

Gasifier is modeled at higher temperature in the range of 1200 to 1600 °C and higher pressure between 22-40 bars and product gase exits from the gasifier at the designed temperature and pressure. The temperature is set to get ash in slag form and in order to flow out slag from the reactor temperature is kept 100-150 °C higher than the ash fusion temperature. The gas is then directed to syn gas cooling and cleaning sections where as the slag is seperated from the bottom of the reactor (National Energy Technology Laboratory, 2010; Ke Liu et al., 2010). The flow sheet for feed preparation and gasification is given in Appendix I (2/11).

4.2 Gas cooling, cleaning and removal of impurities

4.2.1 Gas cooling

The product gas after the gasification is first quenched with recycle product gas by passing though cooler and then by passing syngas cooler. The purpose of heat transfer through recycle product gas is to lower the temperature of the raw product gas to transfer the molten ash into solid material which is not threat of fouling, corrosion and erosion of the syngas cooler. Normally temperature of raw producer gas drops to 800-900 °C from the cooler and then finally to 240 °C by producing superheated steam of pressure 50-180 bar in the syngas cooler. Syngas cooler is the combination of convection, economizer and super heater (Liang-shih, 2010).

4.2.2 Particulate removing

In cleaning stage, first the fly ash is removed by passing through the candle filter or cyclone separator and any trace of filter fly ash pass in the product stream then it is captured in wet scrubber. Cyclone is very good for high temperature operation where as candle filter is good for the temperature range of 300-500 °C. The recovered fly ash and/or slag is either recycled to process or it can be taken out from the process depending on the process.

Particulate removal involves the removal of chloride (HCl), sulfide (H_2S), NH_3 and other contaminants. Spent water with the particulate is removed from the system and then sends to treatment system where it is passes through the gravity settler and removed from the system in the cake form which then can be reused to process or can be then emit out from system depending on the process demand. Water available after the filtration is reused for gasification where as the clean scrubbed gas is send to water gas shift reactor and then to carbonyl sulfide (COS) hydrolysis (Liang-shih, 2010; National Energy Technology Laboratory , 2010).The flow sheet is given in Appendix-I (3/11)

4.2.3 Wet scrubbing

After dry solid removal, the raw gas is quenched and sends to wet scrubbing for the final particle removal. To ensure the complete finest/smallest particle removal normally process is carried below the syngas dew point temperature which in turns makes finest particle to act as nuclei for condensation (Liang-shih, 2010).

4.2.4 Water gas shift and COS Hydrolysis

There are two process involve for water gas shift process sweet and sour shift process. Sweet shift process involves the introduction of steam while sour process utilizes the saturated scrubbed gas for shift reaction. In sweet process, gas is condense to remove all moisture content and then reheated and again addition of steam takes place. This whole makes process thermally inefficient as compare to sour process. Normally for coal gasification sour process is generally practiced.

The purpose of water gas shift reactor is to adjust the steam/CO ratio as per demand of the process requirement. Normally it is kept higher than 2 in order to avoid carbon deposition and also for higher conversion of CO to CO_2. The reaction is given by

$$CO + H_2O \leftrightarrow CO_2 + H_2 \qquad -41.2\ \text{kJ/mol} \qquad (9)$$

The reaction is exothermic and equimolar, so effect of pressure is low while temperature affects a lot. But at lower temperature, the rate of reaction is very slow so in order to increase the rate of reaction generally the reaction is carried out in two stages. In the first stage, high temperature (300-500 °C) is maintained for higher conversion and it results maximum conversion of CO and formation of H_2 in the presence of Iron catalyst. The second step is low temperature shift reaction which is carried out at temperature (210-270°C) in the presence of copper-based catalyst. These two steps shift conversion resulted to maximum conversion of CO and formations of H_2.

In COS hydrolysis, sulfur in product gas converts into H_2S. It is normally carried out at 170-210 °C. Mostly sulfur in the coal converts to H_2S but around 3-10% forms organic COS which is then converted to inorganic H_2S by passing scrubbed gas from water through reactor. The flow sheet is given in Appendix-I (4/11). The reaction is given by

$$COS + H_2O \rightarrow H_2S + H_2O \qquad (15)$$

(Liang-shih, 2010; National Energy Technology Laboratory , 2010)

4.2.5 Mercury removal

The syngas is then cooled to nearly 40 °C and then passed through the activate carbon bed mercury (Hg) removal. On passing by carbon bed which is impregnated with sulfur, it converted to HgS

$$Hg + S \rightarrow HgS \qquad (16)$$

(Liang-shih, 2010).

4.2.6 Acid gas removal

Syngas contains the impurities including H_2, CO_2 and small amount of unconverted CO and H_2O etc. Acid gas mainly consists of H_2S and CO_2. The removal of the acid gas is done individually and simultaneously both at the same depending on the nature of the process. There are different ways to treat the acid sour syngas but Selexol process is selected.

For the Selexol process two interconnected absorber is used one for the removal of sulfur while other for the CO_2. First lean gas is introduced from the bottom in H_2S absorber and moving in counter current direction while passing through H_2S absorber maximum amount of the lean gas is freed with H_2S while containing maximum amount of CO_2, it is then send to CO_2 absorber with part of it recycle to H_2S absorber. While the solvent coming out from the bottom of the H_2S absorber is H_2S rich and send to H_2S stripper. Similarly H_2S free gas enters from the bottom of the CO_2 absorber and move perpendicular to upward direction. The process ensures the removal of clean gas with CO_2 content to minimum extend. Part of the free H_2S and CO_2 gas is recycled to the system (National Energy Technology Laboratory, 2010; Liang-shih, 2010).

The flow sheet for H_2S and CO_2 are given in Appendix I (5/11 and 6/11) respectively.

4.2.7 Claus process

Claus process involves the conversion of H_2S to elemental S by the oxidation of H_2S with air. The reaction is given by 17.

$$2\ H_2S + O_2 \rightarrow 2S + 2H_2O \qquad (17)$$

The reaction proceeds into two steps, in first steps about one third of H_2S oxidizes with O_2 to produces SO_2 and H_2O by reactions 18.

$$2H_2S + 3O_2 \square\!\rightarrow 2SO_2 + 2H_2O \qquad (18)$$

The temperature is normally kept higher 1800-2800 F with pressure sets higher than the atmospheric pressure. The gas emits from the first claus reactor and goes for cooling while steam results due to heat exchange is directed to HRSG.

In the second stage remaining two third of H_2S reacts with available SO_2 and produces elemental sulfur at 360 F and 10 psig pressure by following reaction.

$$2\ H_2S + SO_2 \leftrightarrow 3S + 2H_2O \tag{19}$$

The gas is directed to separator where elemental sulfur (with 99% purity) is produced while unconverted gas is send to gas turbine section (Jayakumar, 2008).The flow sheet of Claus process given in Appendix I (7/11)

4.2.8 Pressure swing adsorption

Pressure swing adsorption is used to get maximum purity of H_2 gas. It is employed to purify further H_2 which is then taken as product and can be used as raw material for other products. It consists of multiple fixed absorbers of fixed beds including silica gel, activated carbon etc. As H_2 is being high volatile and having low polarity so while passing the raw gas through the absorber it remains unattached to the bed and results to H_2 with purity of 99.99% where as the impure gas containing CO, CO_2 and H_2 is purged from the absorber and then send for the combustion in the gas turbine section. A Pressure swing adsorption is able to recover 80-92% of H_2 gas (Liang-shih, 2010).

4.2.9 CO_2 capture and storage

CO_2 separation is done by pre combustion which is found to be more efficient and cost saving. It is easier to remove the impurities from the fuel gas before the combustion of syngas rather than conventional concept of post-combustion. But for pre combustion following additional operations is required.

The first operation in IGCC for pre-combustion of CO_2 is shift gas reaction of fuel gas. Fuel gas has major components CO and H_2 and with addition of steam it converts into H_2 and CO_2.

$$CO + H_2O \leftrightarrow CO_2 + H_2 \qquad -41.2\ \text{kJ/mol} \qquad (9)$$

The reaction is exothermic and heat releases in the reaction is utilized in HRSG for steam generation and in power production.

The second process is the separation of CO_2 from the fuel gas by using acid gas removal method.

The third additional unit required is to compress the CO_2 gas which is needed to reduce the capacity volume of the CO_2 for the cheaper and efficient transportation to the storage destination. Normally 140 bar is typically used to compressed the CO_2. CO_2 can be injected to underground reservoirs. The other available option can be saline aquafier (Jennie, 2005; Kehlhofer et al., 2009; Ke Liu et al., 2010).

4.3 Power generation block

Combine power cycle system is use for power generation. Combine cycle is the combination of power cycle of gas turbine (GT) and steam turbine (ST). Overall for the energy generation, IGCC contains heat recovery steam generator (HRSG), GT and ST.

4.3.1 Gas turbine

Gas turbine is very important in power generation unit and produces 60-75% of the total power production. GT section consists of air compressor to compress the air in the range of 14 to 30 bar for fuel combustion, combustor where combustion takes place of flue gas and then flue gas expands in GT to generate power. The flue gas exits at around 450-650 °C. The final exit temperature of flue gas depends on various factors like turbine efficiency, pressure ratio and flue gas inlet temperature etc. (Kehlhofer et al., 2009).

The flow sheet of Gas turbine section given in Appendix I (8/11)

4.3.2 Heat Recovery Unit and Steam Generation

HRSG is very important part in power generation in IGCC. HRSG consists mainly of three heat exchangers: economizer, evaporator and super heater. They are connected in series, first water enters into economizer where it is heated to its boiling point without changing its phase and then goes to evaporator where the latent heat of water is supplied to convert water into vapor form or saturated steam and then to super heater which is used to increase the sensible heat of the saturated steam and converted to superheated steam. The whole process depends on the heat duty of hot flue gas (Liang-shih, 2010).

The flow sheet of HRSG section given in Appendix I (9/11)

4.3.3 Steam turbine

Steam turbine utilizes super heated steam to generate electricity, when high pressure steam passes through the steam turbine then heat energy from steam transform into mechanical work and produces power (Liang-shih, 2010).

5 SIMULATION PART

Six samples are selected based on different compositions of cotton stalk, biomass and coal whereas Mix-3 is taken as base case, with feed ratio of 40/60 wt%.

5.1 Simulation approach

All the important unit operations including ASU, FP, PR, WGS, AGR, CR, GT, HRSG and PSA are modeled as hierarchy in the flow sheet. “Hierarchy” means separate block which is very useful for simulation and especially if the simulation is lengthy and having large flow sheet. The flow sheet of overall unit operation of the process is given in Appendix I (11/11).

Most of the input parameters are designed by using calculators including O_2 and steam for gasification, air in ASU, N_2 for drying of the feed, amount of steam in WGS reactions in WGS section, flow rate of methyl diethanolamine (MDEA) and diethylamine (DEA) solution in AGR unit, required amount of air for combustion in CS and in GT section, amount of N_2 and steam for diluting etc. Similarly numbers of design spec are used to adjust the parameters. Most prominent parameters are requirement of steam in heat exchangers, steam flow rate in ST, final temperature of the super heated steam, exit temperature of heat exchanger before the strippers in the AGR sections etc.

Working gasification temperature is set by using sensitive analysis. Overall sensitive analysis is applied to study the impact of air or ER on the gasification temperature so as on CGE, HGE, CCE, TE, H_2 production, CO_2 production and net work done etc. Sensitive analysis is also done to study the impact of gasification temperature on the NOx and SOx emissions.

5.1.1 Physical properties

Peng Robinson with Bostone Mathias (PR-BM) is selected for most of unit operations including gasification, cleaning and GT sections etc. For HRSG system, STEAMNBS while for AGR section, UNIFAC method is used. For feed materials including biomass coal and ash is defined as non conventional compound while for enthalpy calculation HCOALGEN and for density model DCOLIGT is selected (Mar et al., 2008; Technology, 2010).

UNIFAC mehotd is used as it is giving results closer to real values for AGR section. Heat of combustion (HCB) of coal and biomass is calculated by using Dulong and modified Dulong formulas respectively in (db) and wt%, which is given by equation (1) and (2) in Appendix II (1/5) and used in simulation.

5.1.2 Input data

Important parameters of the feed mixtures are shown in table 2, where sample Mix-6 is pure coal and Mix-7 is biomass. The proximate and ultimate analysis of biomass and coal is given in (db) and wt%. Data for the feed mixtures are taken from journal articles.

	Mix-1	Mix-2	Mix-3	Mix-4	Mix-5	Mix-6	Mix-7
Input Data							
Biomass %	50	60	40	25	75	0	100
Coal %	50	40	60	75	25	100	0
C (% ar)	55.32	53.68	56.97	59.44	51.20	63.56	47.08
H (% ar)	5.86	5.61	6.12	6.50	5.22	7.14	4.58
O (% ar)	28.66	31.37	25.96	21.90	35.43	15.14	42.19
N (% ar)	1.58	1.47	1.68	1.84	1.31	2.11	1.04
S (% ar)	0.74	0.59	0.89	1.11	0.37	1.49	0.00
Cl (% ar)	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ASH (% ar)	7.83	7.29	8.38	9.20	6.47	10.56	5.10
MS (% ar)	20.71	17.37	24.05	29.07	12.36	37.42	4.00
VM (% ar)	66.03	68.05	64.01	60.97	71.09	55.91	76.15
FC (% ar)	26.14	24.66	27.62	29.83	22.44	33.53	18.75
HCB (db), MJ/KG	23.28	22.09	24.47	26.24	20.32	29.21	17.35
Feed (Kg/s)	30.76	30.76	30.76	30.76	30.76	30.76	30.76
Gasification Temperature, C	1522	1522	1522	1525	1517	1528	1524
Gasification Pressure, bar	25.00	25.00	25.00	25.00	25.00	25.00	25.00
Air/Feed ratio (mb)	0.53	0.51	0.56	0.60	0.47	0.67	0.43

Table 2 Input parameters for the feed mixtures. (Anila et al., 2011; Munir et al., 2010)

5.2 Air separation (ASU unit)

ASU is modeled with absorbers and distillation columns equipped with reboiler and working at high and lower pressure 6 and 1 bar respectively producing concentrated O_2 and N_2. O_2 is used for gasification in FP unit while N_2 is used in GT section. In actual process 95% purity of O_2 is used for gasification so in our simulation this will use. Results of N_2 and O_2 streams from ASU is shown in table 1, Appendix II (2/5)

5.3 Feed preparation and gasification (FP unit)

In simulation, whole gasification system is designed with two blocks: decomposition unit (RYield reactor) and gasifier (RGibbs reactor). In RYield reactor there is no requirement of the kinetics of the reactions. The coal, biomass and ash is defined as non conventional compound, which is send to decomposition unit to convert it into conventional form by specifying the yield with respect to proximate and ultimate analysis of coal and biomass where as the energy released from the decomposition of nonconventional compound is used in the gasifier to make the process adiabatic. RGibbs reactor works on the basis of minimizing gibbs free energy (Plus, 2011) and by using conditions, equilibrium state is developed, although in actual practice it is very difficult to develop the equilibrium in very short time but there are number of models that are verified by experimental results with the same postulates so here it is also true. For the unconverted 1% carbon in gasifier, inert value 0.01 is specified for carbon resulted to solid carbon in product stream which then react with ash to form slag. Unreacted is treated as soot and is taken as fly ash in simulation. The temperature of the reactor is set to 1520 °C by varying the oxidant flow rate in the reactor by using sensitive analysis while pressure to 25 bar.

Temperature is set in view to keep the ash in slag form and it is better to keep the temperature 100-125°C higher then ash fusion temperature. Cotton stalk, ash fusion temperature reported to in the range of 1400-1450 °C while for lignite coal 1300 °C (Anuradda, 2009; Bassam et al., 2010; Ke Liu et al., 2010).

Product gas after gasification is send to HRSG system for the production of super heated steam till the temperature drop to 800 C and then further cool down by recycling the product gas after cleaning. The cleaning process is done at 22 bar. Product gas compositions are given in table 2 Appendix II (2/5),

5.4 Wet gas shift reactor (WGS unit)

After the removal of fly ash and unconverted carbon from the product gas at temperature 240 °C, it is directed to WGS reactor for the shift gas reaction which results to conversion of CO and H_2O into H_2 and CO_2. For wet gas shift reaction saturated steam at 22 bars is used. Addition of number of mole of steam is fixed to 1.5 with mole flow of CO in the product gas by using calculator in simulation. To increase the conversion of CO, the reaction is proceed into two steps, the first reactor with low temperature 230 °C while second reactor with high temperature 450 °C. Nearly 85-90% conversion of CO is achieved by using this sequence.

Available COS in the product gas is hydrolyzed to H_2S and CO_2 by using steam at 22 bar. Similarly calculator is used to fix the amount of steam for the conversion of COS and equimolar flow of steam is used with COS. Unreacted COS is directed to GT section while product gas is send to AGR unit. Stream results for the inlet and outlet streams in WGS are shown in table 3 Appendix II (3/5).

5.5 Acid gas removal (AGR unit)

AGR unit consists of two process units, H_2S and CO_2 removal sections. In H_2S section, MDEA solution with mass fraction water 0.5 and MDEA 0.5 are used. This mass ratio is found to be feasible for operational and economical point of view. The quantity of MDEA is fixed by using calculator with mass ratio 1.5 to product gas entering to the H_2S absorber. Similarly for CO_2 removal section DEA solution is used with mass fraction H_2O 0.5, MDEA and DEA 0.25 respectively. The quantity of the DEA solution is also fixed by using calculator. DEA solution with mass ratio of 6.58 to the product gas entering to CO_2 absorber is used.

The mass fractions of MDEA and DEA in their respective solutions are used for higher process efficiency with minimum costing (Daniele et al., 2002).

These solutions are easily recoverable from their respective stripper sections with the stream lean-11 and lean -21 and then they can be recycled. Heating in the reboiler is done by using steam producing in HRSG section.

From the AGR section, 84-85 % CO_2 removal is achieved which is then compressed to 155 bar and then send for sequestration. H_2S is sent to CS for further treatment while half of clean gas is sent to PSA unit for H_2 production and half to GT section. Saturated steam from HRSG is used in the reboiler. From the AGR unit, pressurized CO_2 at 155 bar is taken out for CO_2 sequestration. Process parameters for AGR unit is given in table 4 Appendix 2 (3/5) while recovered H_2S and CO_2 streams results are shown in table 5 Appendix -2 (4/5).

Mole Frac	PG before cleaning	Clean Gas
H_2O	9.78	1.44
N_2	1.62	1.92
O_2	0.00	0.00
H_2	32.50	84.23
CO	51.66	2.76
CO_2	4.11	9.65
METHANE	0.00	0.002
COS	0.02	0
H_2S	0.31	0
Total Flow kmol/sec	1.97	1.89
Total Flow kg/sec	37.88	14.22
Total Flow cum/sec	7.06	2.49
Temperature C	800	72.1
Pressure bar	25	22

Table 3 Comparison of product gas after gasification and clean gas for base case

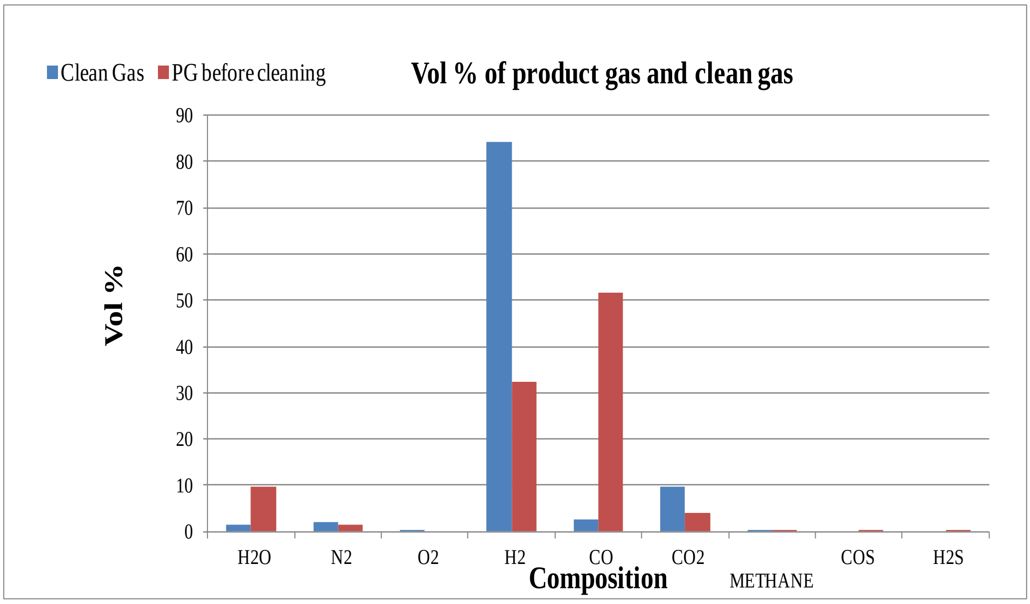


Figure 5 Vol% of product and clean gas for base case.

The tail gas from H_2S stripper is treated in the CS section. H_2S reacts with air and produces SO_2 which then converted to elemental sulfur and sulfur is taken out as by product from the system. The flow of air is controlled by using calculator. Quantity of air required for complete combustion is fixed by using calculator with 99% H_2S conversion. Unconverted H_2S is directed to GT section. Stream results are given in table 6 Appendix II (4/5).

5.6 Claus process (CS unit)

CS process is designed by using adiabatic reactor for the combustion of H_2S into elemental sulfur. The addition of air is fixed with the equimolar air required for the combustion of H_2S by using calculator.

5.7 Pressure swing adsorption (PSA unit)

PSA is simply designed by using separator (with H_2 purity 85%) where as tail gas is directed to GT section. The stream results are given in table 7 Appendix II (5/5).

5.8 Gas turbine section (GT unit)

GT section is designed with process pressure at 22 bar. Water addition is fixed to minimize the H_2 final volume concentration to 60% in fuel gas. Steam is also used to increase the temperature of the fuel gas to 260 °C, to get good HHV of fuel gas before combustion which results to increase in the enthalpy of stream from 68708.93 kJ/kmole to 1022109 kJ/kmole, it is highlighted in table 8 Appendix II (5/5). 6% excess air is supplied for the complete combustion of fuel gas. Combustion of fuel gas results to high temperature flue gas which is decreased to 1200 °C by using pure N_2 gas from ASU. The flow rate of N_2 gas is fixed by using calculator.

Both N_2 and Air after compressing to 22 bar results to rise in temperature which then decreased to 70 °C by passing through heat exchanger with water. This also leads to less requirement of steam to increase the temperature of flue gas to 260 °C. Finally flue gas after expanding through GT exits at temperature around 505-525 °C and leads to HRSG system.

N_2 and H_2O addition is used to dilute and saturate the fuel gas. Addition of steam is carried out to raise the temperature of fuel gas to 260 °C. Fuel gas' inlet temperature to GT at around temperature 1200 °C, while flue gas outlet temperature from GT is 525 °C and then finally from HRSG at 105 °C is modeled to Elcogas plant model (Treviño, 2000)

5.8.1 Environmental consideration

The most important factors for environmental considerations are NOx and SOx emissions in the final flue gas after combustion in GT. NOx and SOx emissions depend on many factors. They are discussed below

H_2 concentrated flue gas

In IGCC technique with pre-combustion CO_2 capture, fuel gas creates problem for GT. Fuel gas before combustion in GT section is mostly dominant with H_2 gas with volume concentration of 70-85%. Highly H_2 concentrated fuel gas is threat to high flame temperature which in turns produces more NOx and SOx which is not good for the environmental consideration and also higher H_2 is not good for GT life. Flame temperature is directly related to NOx emissions. The maximum allowed concentration of H_2 for GT is 65% (Jennie, 2005).

Fuel to air ratio for combustion

By keeping high fuel to air ratio, by means of partial combustion NOx and SOx emissions can be controlled but it leads to unburned CO and H_2 in the final flue gas so ratio 1 and more than 1 is preferable for complete combustion (Kehlhofer .et al., 2009).

Air temperature for combustion

Air temperature increases after compression, air at high temperature can then increase flame temperature so it is better to transfer the heat from air and also from N_2 (which is injecting after combustion to control the outlet temperature of flue gas) (Kehlhofer .et al., 2009).

Combustion process residence time

Sufficient residence time is required for complete combustion. The easiest way to reduce the flame temperature is by diluting with N_2 and saturation by using fuel or steam. Flame temperature so as emissions can be reduced either by dilution or by saturation and combination of both of them which is preferable. By saturation NOx level can be minimized to low level as 40 ppm with (15% O_2). Although saturation increases plant output but it also leads to efficiencies loss and also required high water consumption. Practically water has more advantages as compare to steam for saturation. With the saturation by using water or steam, H_2 concentration can also be decreased to less than critical 65% (Kehlhofer .et al., 2009).

Fuel gas is saturated by spraying water or steam at higher temperature than the fuel gas. Normally water at higher pressure from HRSG or intermediate pressure steam produced in the HRSG system is usually employed for this purpose (Zhu, 2004).

The maximum allowed NOx and SOx level for 6% dry O_2 for Elcogas coal is 200 PPM while for saturation purpose water is employed and then by indirect heating the temperature of flue gas increased to 260 °C (Treviño, 2000; Francisco, 2005). Complete material balance around GT is shown in table 8 Appendix II (5/5).

5.9 HRSG

HRSG system is designed to produce steam at three pressure levels: low pressure steam 6.5 bar, intermediate pressure steam 35 bar and high pressure steam 127 bar. Heat is recovered from flue gas (GT section outlet gas with temperature 505-535 °C) and temperature of flue gas is reduced to 105 °C where as the product gas from the gasification is used to superheat the high pressure saturated steam at 127 bar and exit from the GT section at temperature 240 °C to PR section. To make flow sheet simple, only steam cycle has been shown in simulation. Steam flow rate is specified by using calculator to keep the final exit temperature of flue gas to 105 °C while the superheated steam temperature is also set by using calculator which maintains the outlet product gas temperature to 240°C.

5.10 Process efficiency

All efficiencies are calculated on HHV basis. HCB for feed mixture is calculated on (db) while for other streams on (wb). Energy balance, net power production and consumption, and process efficiencies are shown in table 1 Appendix IV (1/3) whereas material balance and production is shown in table 2 Appendix IV (2/3) while in table 3 Appendix IV (3/3), hazard emissions including NOx and SOx, captured CO_2 and units production of the process are discussed.

5.10.1 Calculation of HHV

HHV can be defined as the amount of heat released by burning unit mass or volume of solid fuel. To calculate HHV, initial temperature of fuel and final temperature of the product is taken to 25°C. HHV includes latent heat of vaporization of water in the products. It is possible to calculate the HHV on the Aspen plus and HHV hierarchy is simulated for feed, product gas, flue gas, and flue gas after combustion etc. Flow sheet to calculate HHV is shown in table 10 Appendix I (10/11).

5.10.2 Cold gas efficiency

Cold gas efficiency is normally used to define the efficiency of the gasifier. It is normally calculated when the producer gas from gasifier is directly used to run internal combustion engine. It can be related with the formula as

$$\eta_{\text{coldgas}} = \frac{m_g HHV_g}{m_f HHV_f} \tag{20}$$

where

m_g = Product gas flow rate (kg/s)

HHV_g = Heating value of the producer gas (kJ/kg)

m_f = Feed flow rate (kg/sec)

HHV_f = Feed high heating value (kJ/kg)

η_{coldgas} = Cold gas efficiency (Prabir, 2010).

5.10.3 Hot gas efficiency

Hot gas efficiency includes the heat duty with sensible heat of the producer gas where as sensible heat calculated at the gasifier exit temperature. Heat duty calculation is same with the cold gas efficiency.

$$\eta_{\text{hot gas}} = \frac{m_g H_g + m_g HHV_g}{m_f HHV_f} \tag{21}$$

where

m_g = Product gas flow rate (kg/s)

H_g = Sensible heat duty of producer gas (kJ/kg)

HHV_g = High heating value of the producer gas (kJ/kg)

m_f = Feed flow rate (kg/sec)

HHV_f = Feed high heating value (kJ/kg)

$\eta_{\text{hot gas}}$ = Hot gas effeciency (Prabir, 2010).

5.10.4 Overall Combine Cycle Efficiency

Overall combine cycle efficiency can be defines as the ratio of summation of net work done by gas turbine section and HRSG to the heat supplied by product gas and flue gas entering to HRSG system.

$$\eta_{\text{CCE}} = \frac{W_g + W_{flue}}{m_f HHV_f} \qquad (22)$$

η_{CCE} = Overall combine cycle efficiency

W_g = Sensible heat duty of producer gas (kJ/kg)

W_{flue} = Heating value of the flue gas (kJ/kg)

m_f = Feed flow rate (kg/sec)

HHV_f = Feed high heating value (kJ/kg) (Prabir, 2010).

5.10.5 Thermal Net Efficiency

Thermal efficiency can be defined by using the relation

$$\eta_{\text{THE}} = \frac{Pnet}{m_f HHV_f} \qquad (23)$$

η_{THE} = Thermal efficiency

P_{net} = Net Power production (MW)

m_f = Feed flow rate (kg/sec)

HHV_f = High heating value of the feed gas (kJ/kg) (Emun, 2010).

6 OPTIMIZATION

For optimization total 5 mixtures are selected with different compositions on wt% to study the effect of feed on process efficiencies, net power production and consumption, H_2, NOx, SOx and CO_2 production etc. First the cotton stalk (biomass) to coal ratio in feed mixture will be changed to study the effect of cotton stalk addition and then the variation of gasification temperature by varying ER will be studied.

6.1 Impact of variation of biomass to coal mass ratio

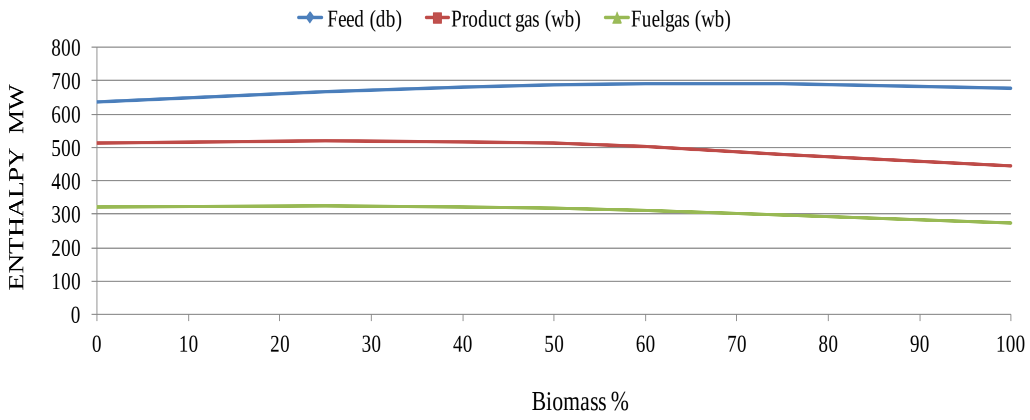
Biomass has low HHV value 16.657 MJ/kg as compare to coal 29.207 MJ/kg but it contains more combustible material 0.91 kg of combustible material/kg of biomass in wt% where as 100 % coal has 0.63 kg of combustible material/kg of coal in wt%. The mixture of coal and biomass can produce good process efficiencies. In order to study the effect of biomass addition following things are studied.

6.1.1 HHV of process stream

In order to study the impact of biomass addition, enthalpy (MW) of feed mixture on (db), product gas on (wb) and fuel gas (entering to GT section) on (wb) are calculated and presented in table 4

	ENTHALPY , MW		
Biomass %	Feed (db)	Product gas (wb)	Fuelgas (wb)
0	636.47	514.77	321.06
25	667.87	521.10	324.41
40	680.73	518.14	322.32
50	686.61	512.42	318.41
60	690.72	501.96	311.49
75	692.63	480.68	297.47
100	678.97	444.31	273.54

Table 4 Enthalpy (MW) of feed mixture (db), product gas (wb) and fuel gas (wb)



Firgure 6 Graph of enthalpy (MW) of feed mixture (db), product gas (wb) and fuel gas (wb)

From the table 4 and figure 6, it is clear that with the increase of biomass compositions in the feed mixture, enthalpy of feed increases linearly which is due to addition of biomass having more combustible material/feed, which results to more mass production in product gas so ultimately enthalpy in MW of product gas which is the multiple product of HHV (MJ/kg) and mass (kg/s) will increase till the point will achieve where by adding more biomass enthalpy will decline due to lower HHV. From figure 6, it is cleared that after 75% of biomass addition in the feed mixture HHV started to decline.

But it also notable that biomass is concentrated with O_2 not with C as shown in table 2, it contains nearly three time more O_2 than coal which has 15% of $O_{2.}$ and carbon is 47% on wt% while for coal is 63.5% on wt%, so initially with increasing the biomass this leads to increase C in feed mixture due to more combustible material/kg of feed and will increase HHV of product but at specific ratio of biomass to coal with the addition of biomass, C level in mixture will start to decrease and so enthalpy of product and fuel gas. This can be clearly experienced after 40 % biomass addition in feed mixture, enthalpy of both product and flue gas streams started to decline.

From the figure 6, the ratio of Biomass to coal 40:60 is considered to feasible from process efficiency point of view and therefore it is taken as base case, it is more explained in the later sections.

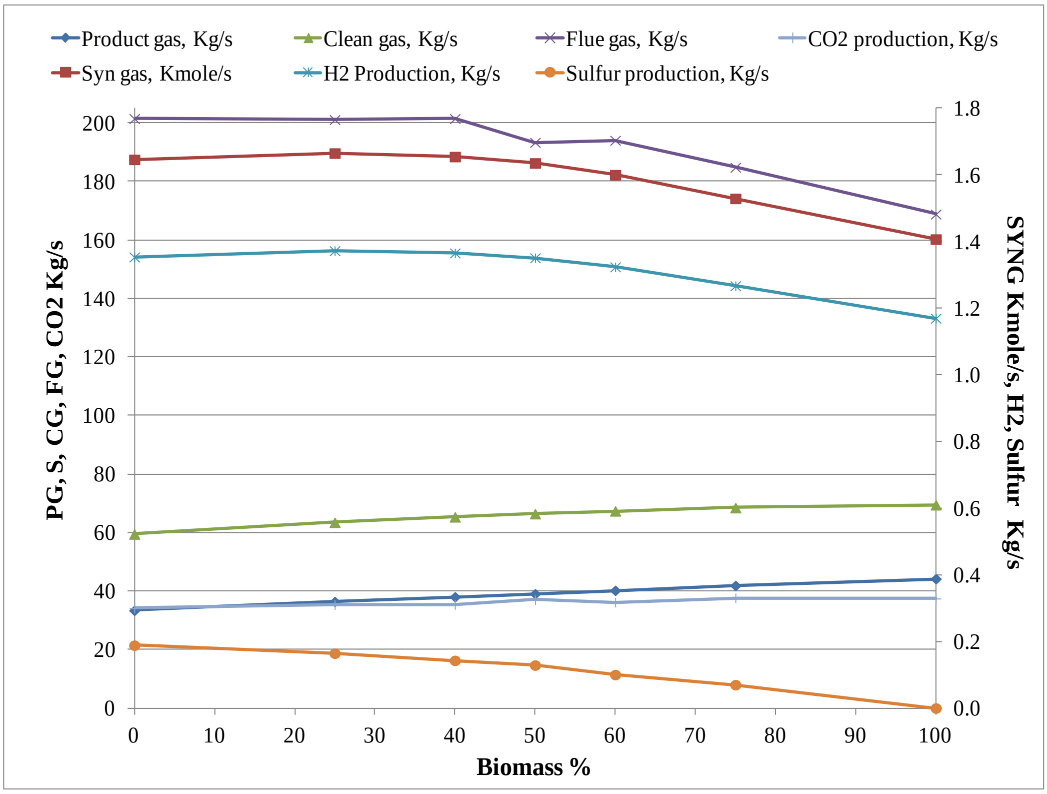
6.1.2 Impact on production rate, and in other streams

The addition of biomass in feed mixture for specific mass quantity results into increase in production rate and flow of clean gas and CO_2 production where as results for other streams are also as shown in table 5 and figure 7.

Production, kg/s

	Biomass %	Product gas	Syngas, kmole/s	Clean gas	Flue gas	H_2	Sulfur	CO_2
Mix-6	0	33.39	1.65	59.48	201.47	1.35	0.19	34.18
Mix-4	25	36.40	1.67	63.45	201.11	1.37	0.16	35.33
Mix-3	40	38.02	1.65	65.34	201.46	1.37	0.14	35.39
Mix-1	50	39.07	1.64	66.42	193.33	1.35	0.13	37.00
Mix-2	60	40.22	1.60	67.38	194.03	1.32	0.10	36.17
Mix-5	75	41.88	1.53	68.48	184.80	1.27	0.07	37.68
Mix-7	100	44.19	1.41	69.47	168.80	1.17	0.00	37.45

Table 5 Impact of biomass addition on flow rate of product gas, syngas, clean gas, flue gas, H_2, sulfur and CO_2



Firgure 7 Graph of impact of biomass addition on flow rate of product gas, syngas, clean gas, flue gas, H_2, sulfur and CO_2

With biomass addition, product gas and clean gas (along with CO_2 production) increase linearly as it is illustrated earlier due to increase in mass of combustible material/kg of feed. But for syngas, it will increase at start to the 40% of biomass addition, but after that with the addition of biomass there will be more O_2 and less C in the feed mixture (and this trend will rises linearly with further addition of biomass) so there will be less formation of syngas in the product stream. H_2 production has same trend as of syngas production.

It is interesting to note that clean gas is increasing linearly with the addition of biomass, as biomass has nearly 4% ash content as shown in table 2, where as the coal has double quantity of ash. This leads to increase in the mass flow of clean gas with the addition of biomass in the feed mixture.

Similarly as there is no sulfur content in biomass so with the addition of biomass sulfur content in the process streams will decline.

6.1.2 Impact on Unit production rate, and in other streams

With increasing biomass in feed mixture for specific mass quantity, units' production is also affected. The main process units' results are given in table 6

Unit Production, Kmole/s

Biomass %	Gasification	ASU	Gas Cleaning	GT	HRSG
0	1.83	5.02	1.88	1.21	7.50
25	1.92	5.09	1.88	1.19	7.51
40	1.97	4.99	1.89	1.21	7.50
50	1.99	5.16	1.85	1.18	7.21
60	2.00	4.89	1.83	1.17	7.23
75	2.01	4.68	1.75	1.12	6.89
100	2.03	4.26	1.64	1.06	6.28

Table 6 Impact of biomass addition on units' production of FG, ASU, PR, AGR, GT and HRSG

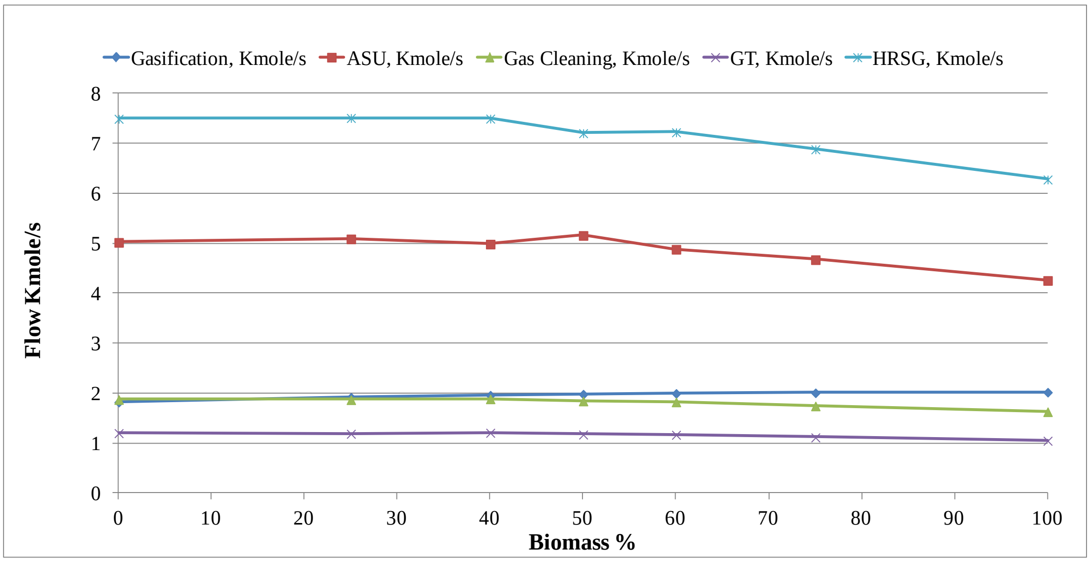


Figure 8 Graph of impact of biomass addition in the feed mixture on the production of FG, ASU, PR, AGR, GT and HRSG

With the addition of biomass in feed stream, as biomass has low moisture content which is nearly 4% as compare to coal which has 37% leads to decrease of moisture of feed mixture. With the decrease in moisture content less N_2 will required in drying so as the less supply from ASU. As shown in table 2 Appendix IV (2/3), for pure coal it require more N_2 in GT section to minimize the temperature of fuel gas entering to GT as compare to biomass, it means that with the addition of biomass less N_2 required in GT section so this overall decrease the N_2 demand in process so less production in ASU section.

Gas cleaning production will increase at start but after 40% biomass addition as the product gas started to concentrate more with CO_2 so this will lead to decrease in mass flow rate of CO_2 free clean gas. Around 82-85% CO_2 will remove from the clean gas stream and remaining will lead to GT and then to HRSG. So overall process stream flow rate after 40% of biomass addition in the feed mixture for the remaining units after gas cleaning will decline.

6.1.3 Impact on net power consumption and production

With addition of biomass in feed mixture for specific mass quantity, net power consumption and production affected a lot along with the production of H_2 gas.

Net Power consumption and production , MW

Biomass %	Aux. Power	GT	HRSG	Net Power	H_2 ,Kg/s
0.00	71.01	113.10	59.34	101.43	1.35
25.00	74.03	113.53	57.45	96.69	1.37
40.00	72.02	113.32	57.06	98.36	1.37
50.00	74.32	108.85	59.91	94.45	1.35
60.00	71.61	109.25	55.82	92.91	1.32
75.00	73.10	104.27	53.41	84.91	1.27
100.00	67.27	95.76	44.64	73.13	1.17

Table 7 Net power consumption and production and H_2 gas production for different feed mixtures

Initially, with the addition of biomass in feed mixture results into increment in enthalpy of product and flue gases so this results into increase in GT power production but with the addition of biomass, fuel gas started to decline in H_2 concentration and flame temperature also started to drop, so less H_2O and N_2 will required for the dilution and saturation. This will lead to decline in the flue gas flow rate after combustion in gas turbine which will decrease the enthalpy of the flue gas. In summation, mass flow rate will started to decline in GT and HRSG sections and net overall power production will decline. Similarly as the above stated reason auxiliary power consumption will also reduce.

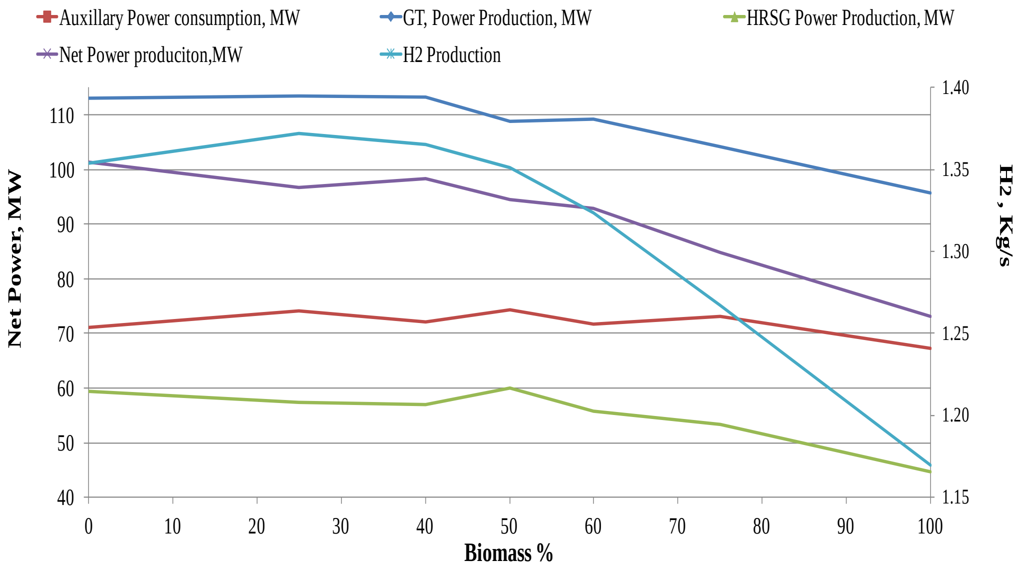


Figure 9 Graph of impact of biomass addition on net power consumption and production and H_2 production.

6.1.4 Impact on process efficiencies

Addition of biomass in feed mixture results into decrease in process efficiencies. To understand the process efficiencies CGE, HGE, CCE, and THE are calculated and given in table 8.

	Biomass %	CGE %	HGE %	CCE %	THE %
Mix-6	0.00	80.88	95.58	53.71	15.94
Mix-4	25.00	78.02	93.31	52.70	14.48
Mix-3	40.00	76.12	91.77	52.86	14.45
Mix-1	50.00	74.63	90.60	53.00	13.76
Mix-2	60.00	72.67	89.04	52.99	13.45
Mix-5	75.00	69.40	86.35	53.01	12.26
Mix-7	100.00	65.44	83.92	51.33	10.77

Table 8 CGE, HGE, CCE and THE of the process for different feed mixtures

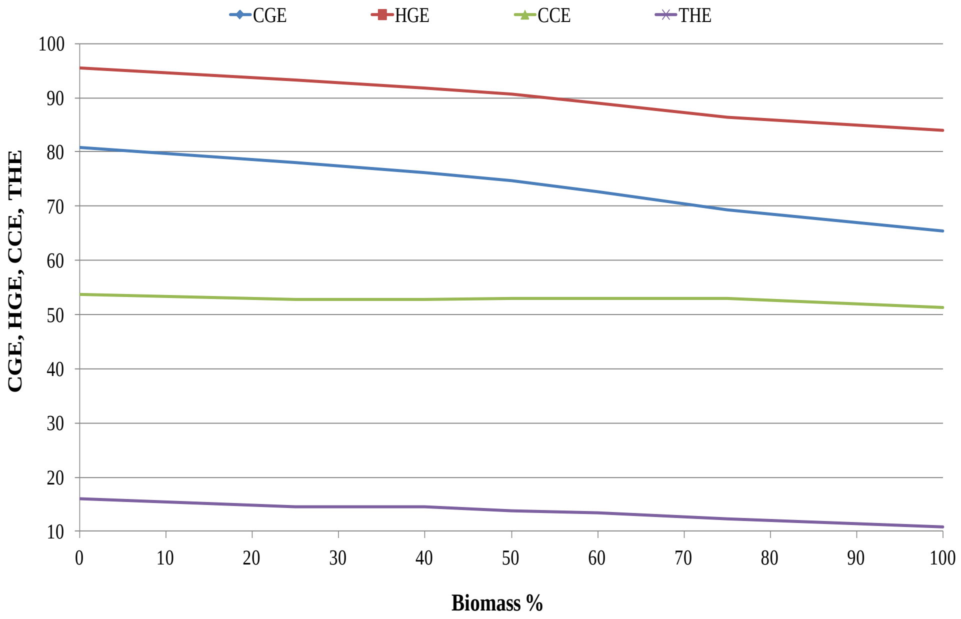


Figure 10 Impact of biomass addition on CGE, HGE, CCE and THE of the process

With the addition of biomass all the efficiencies started to decline and the effect is more after 40% addition of biomass in feed mixture. This behavior is due to addition of low HHV biomass which causes to decline net heat production of the system so as the CGE of the product streams and as the HGE and THE depends on the CGE so they also started to decline along with CGE.

6.1.5 Impact on emission/net power production

	Emissions/Net Power		
Biomass %	NOx, mg/MW	SOx, mg/MW	CO_2 capture, Kg/MW
0.00	371.61	3.34	0.31
25.00	391.78	3.14	0.37
40.00	384.00	2.75	0.36
50.00	389.09	2.57	0.39
60.00	391.90	2.25	0.39
75.00	408.72	1.75	0.44
100.00	432.15	0.00	0.51

Table 9 NOx, SOx and CO_2 capture production per unit net power production for different feed mixtures

With the addition of biomass in feed mixture, NO_X and CO_2 level started to increase due to availability of more O_2 and less C concentrated biomass feed and as the with the addition of biomass the net power production started to decline so this whole affects to increase production of emissions/net power production. The case is very much different for SOx as biomass has no sulfur in the composition so it's giving good results.

6.2 Variation of gasification operating parameters

IGCC process mainly deals with gasification process which in turns depends on gasification temperature. Gasification temperature can be changed by varying inlet feed flow, steam and air use for gasification and moisture drying etc. With increasing mass flow rate the temperature will increase as carbon content to burn will increase which leads to increase gasification temperature and with increase of feed inlet, flue gas flow rate will also increase which will increase net power production through GT (Wayne et al., 2009; Emun et al., 2010).

Moisture content in the dry feed should be less than 10% as per requirement for shell gasifier so optimization is done by varying air flow rate for gasification to study the impact on gasification temperature and so overall production. Air variation for gasification can be done by varying equivalence ratio.

Equivalence ratio

Equivalence ratio (ER) is the ratio of actual oxidant used in combustion to feed with stoicheometric oxidant required for complete combustoin to feed in mass basis where oxidant can be air or O_2. ER can be related by the formula.

$$\mathrm{ER} = \frac{Act_{Oxidant}}{Stoic_{Oxidant}}$$

where

Act $_{\text{Oxidant}}$ = Actual amount of oxidant used for gasification
Stoic $_{\text{Oxidant}}$ = Stoicheometric amount of oxidant used for gasification

Normally ER 0.2–0.4 is used for gasification, too low value of ER<0.2 results into incomplete gasification while ER greater than 0.4 results for more production of CO_2 and H_2O at the cost of desirable products, syngas. With higher value of ER the producer gas started to go toward combustion and at ER 1 to greater than 1 it started to complete combustion. ER also affects on carbon conversion efficiency. It is stated that with the ER up to the range of 0.26, carbon conversion normally increases and after 0.26 carbon cooconversion started to decline (Prabir, 2006; Wayne et al., 2009).

To study the impact of ER on the process parameters base case is selected as feed.

6.2.1 Impact of ER on the reaction temperature and syngas production

Sensitive analysis is done for ER from 0.3-0.4 to study the effect on gasification temperature and syngas production. ER range is selected to get the gasification temperature in the range of 1000-1800°C. The results are shown in table 1,-Appendix III (1/4) while graphs is shown as

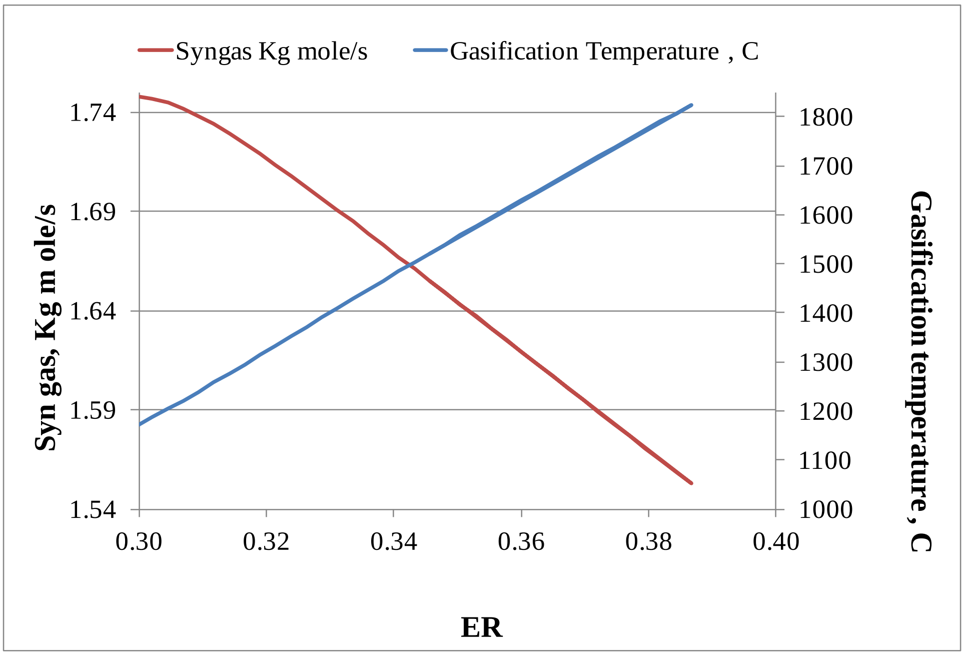


Figure 10 Effect of ER on gasification temperature, °C and syn gas production (kmole/s)

With the increase of ER, there will be more O_2 available, so more C will burn to increase the reactor temperature and results to production of more CO_2 and H_2O instead of syngas (CO and H_2) thus overall process efficiencies will be decreased. This graph also tells that for getting maximum syngas production, the temperature of the reactor should keep lower by keeping low ER.

6.2.2 Impact of ER on net power production, CGE, HGE and CCE

The effect of ER, on net overall power production, CGE, HGE and CGE is studied by using sensitive analysis on selected ER. The results are shown in table 2, Appendix III (2/4) where as graph is shown below.

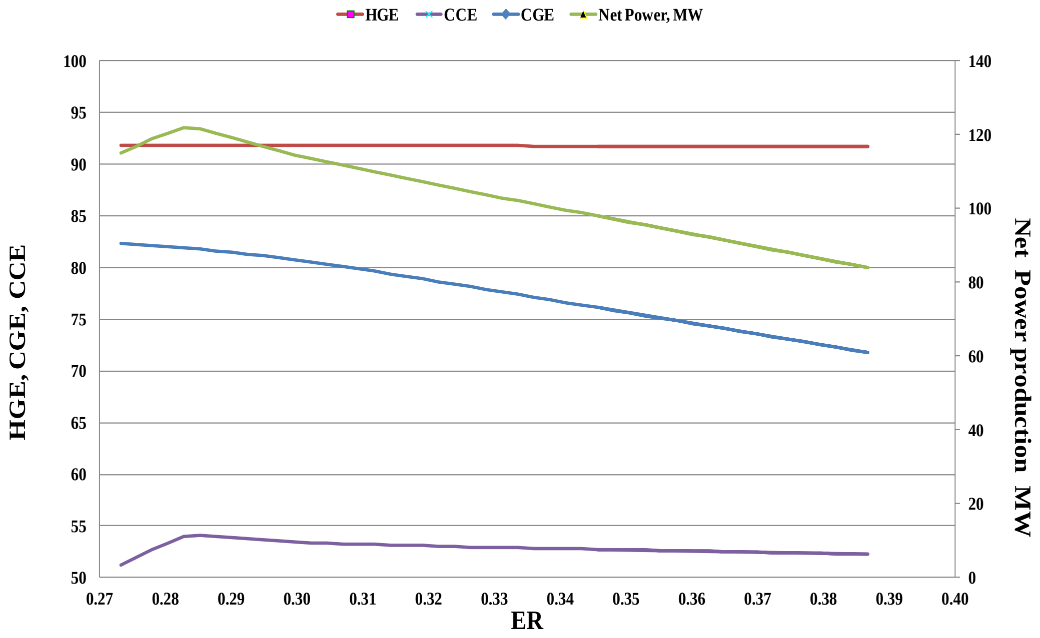


Figure 11 Effect of ER on net power production (MW), CGE, HGE, and CCE

From the figure 11, it is very much clear that with ER, overall process efficiencies are going to decrease. Initially with the ER all efficiencies and net power production increase and there is peak of net power production which is around 120 MW is achieved at ER 0.2852, and so as CGE and HGE. Similarly from figure 10 at low ER, we have the maximum syngas production and one more sensitive analysis is performed for H_2 production and it also gave close to highest production of H_2 at this ER value which is given by table 3, Appendix III (3/4). So it is concluded that ideally for process operation ER should be 0.2852 and correspondent temperature 1140 °C should be working temperature for the gasification.

6.2.3 Impact of ER on the emission

To find out the relation of the impact of ER on emissions, the relation of emissions release/net power production is used. The result of sensitive analysis of effect of ER on the emissions is shown in table 4, Appendix III (4/ 4). It is clear from the figure that with the increase ER as net power production is decreased which leads to increase of emission productions per net power. Also with the increase of ER, the more feed will burn to CO_2 and slightly more NOx will be produced.

6.3 Use of air in GT section for dilution

In order to study the impact of diluting effect of N_2 flow on the final flue gas exiting from the GT section, a sensitive analysis is performed by replacing N_2 with air at same temperature and pressure. The results are same with N_2 for power production for GT, HRSG and net power production with slightly differences. But the NOx level was very much high at around 800 ppm which is unacceptable.

7 CONCLUSIONS

Based on sensitive analysis, comparative studies on process efficiencies, HHV, enthalpy, hazard emissions, net power consumption and production along with H_2 productions, some conclusions have been reached which are stated below.

It is very hard to give simple one optimum feed composition which is suitable from every point of view but with the simulation and sensitive analysis of the cases, feed ratio of cotton stalk to coal 40:60 is found to be most compromising one. There is no any doubt that coal alone can generate more power due to more HHV than cotton stalk but coal has lower mass of combustible material than biomass and at the same time a lot of energy required for pre drying of coal while cotton stalk has low moisture content. With the pre drying of coal for acceptable range 9%, auxiliary power consumption will increase which will lower the net power production of plant so addition of cotton stalk gives positive results concerning power production in gas turbine and overall H_2 gas production which increases with the addition of cotton stalk to certain level. Optimum results obtained with addition of 40% cotton stalk in feed mixture. Similarly with the addition of cotton stalk, reduction in auxiliary power consumption and environmental emissions (like NOx and SOx) are also achieved but emissions productions/net power production increase due to less net power production with the addition of cotton stalk. These overall make the study challenging and interesting. To improve process efficiency of cogasifcation of low grade Thar coal with cotton stalk following further future works are recommended.

The maximum process efficiencies, power and H_2 production is observed at temperature 1100-1130 °C by keeping low ER. But at this temperature ash will not slag, so fluidized bed gasifier will be good option. In fluidized bed ash should remain in solid form so operating temperature should be lower from 900-1100 °C and in our case the optimum temperature is closed to 1100°C.

All the process efficiencies depend more on the carbon availability in the feed mixture, with the addition of cotton stalk which has low carbon content there is no any significant improvement observed in net production of gas turbine and HRSG. It will be good option to try other biomasses like rise husk, wheat straw etc. having more carbon content.

If still slagging gasifier will be priority then it is better to use lower acceptable temperature on which ash will slag. Optimum ash fusion temperature of feed mixture should find out practically and gasification temperature should be operated at that temperature. Higher gasifier temperature is resulted by increasing ER. Higher ER means more volume and hence large size reactor will be needed, so large gas cleaning sections and large volume equipments will have to use in other unit operations.

For low temperature gasification operation, with ash in slag form can be possible with flux agent. Flux agent lowers down the melting point of ash. In this case the ash can be obtained in slag form with lower gasifier temperature and also flux agent can also be easily removed from gasifier with ash. It is also recommended to study the IGCC process using shell gasifier with flux agent and keep the gasification temperature to 1100 °C range.

Steam injection is not employed in gasification due to presence of available excess 9% bound moisture after drying in the feed mixture. But for H_2 rich syngas steam injection is preferable. But at the same too much moisture can decrease CGE and overall process efficiency. It will be very good option to study the effect of it.

8 SUMMARY

The study has been done to simulate IGCC technique by using cotton stalk with low grade Thar coal for power generation and H_2 production and to optimize the process from environmental and process efficiencies concerned. Simulations of the different feed mixtures of biomass and coal is done with Aspen plus and they are also optimized to get maximum process efficiency and acceptable level of NOx and SOx emissions. With the addition of biomass there is marked improvement is observed in H_2 production while due to low HHV of cotton stalk, overall process efficiencies declined but the benefit of higher mass of combustible material per mass of feed is obtained and also auxiliary net power consumption reduced due to low moisture and ash content in cotton stalk. 40% addition of cotton stalk in feed mixture is found to be optimum having acceptable environmental hazard emissions and considerably prefer net overall power production along with H_2 production.

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APPENDIX I (1/11)

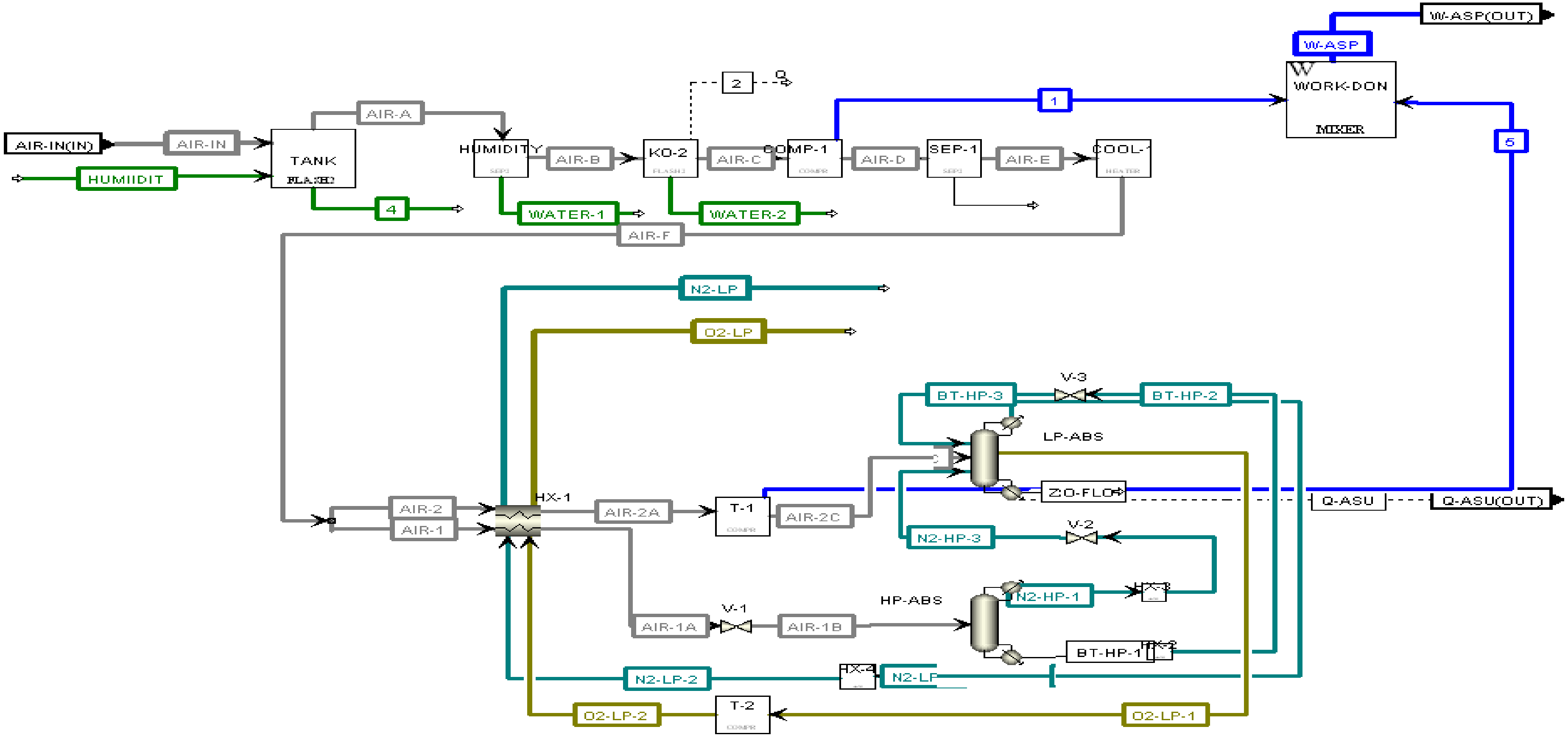


Figure 1 Flow sheet of air seperation unit (ASU).

Appendix I (2/11)

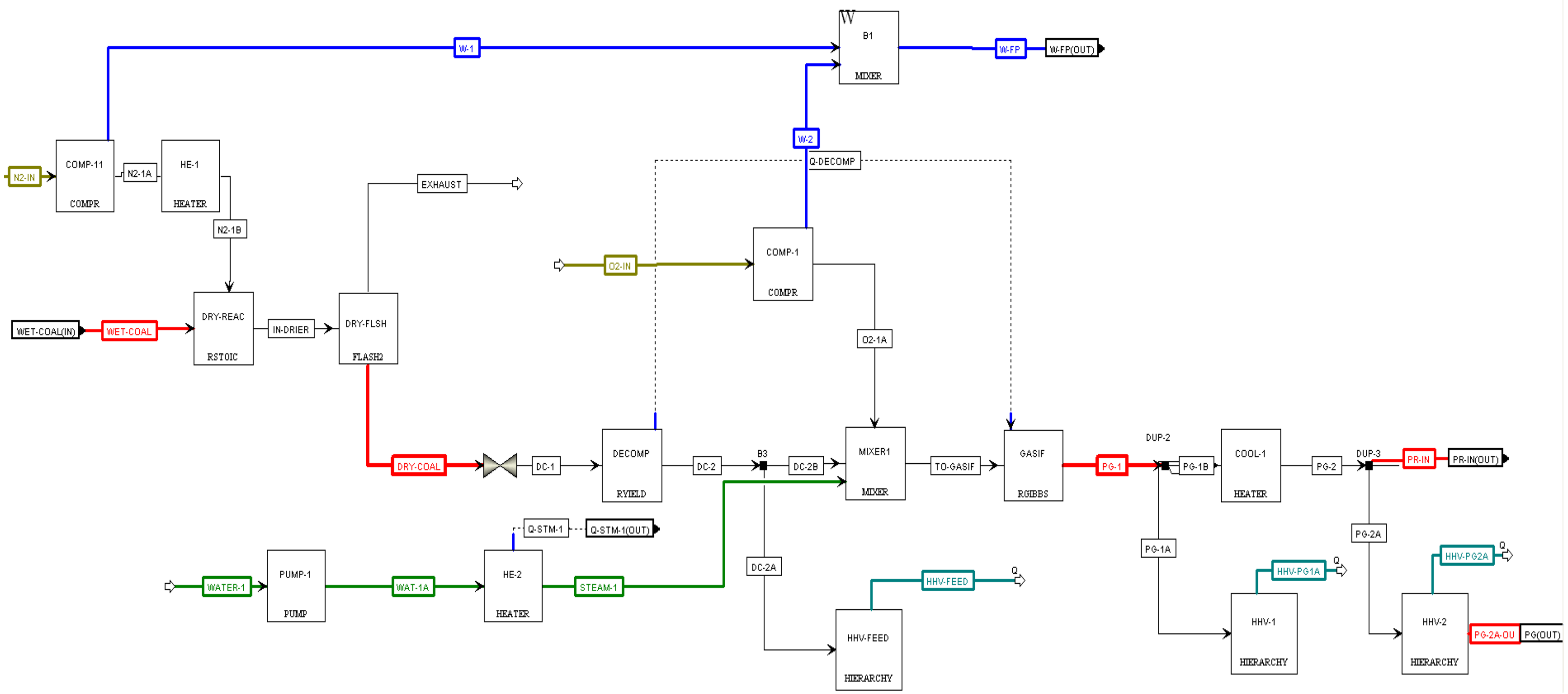


Figure 2 Flow sheet of feed preparation and gasificaiton (FP)

Appendix I (3/11)

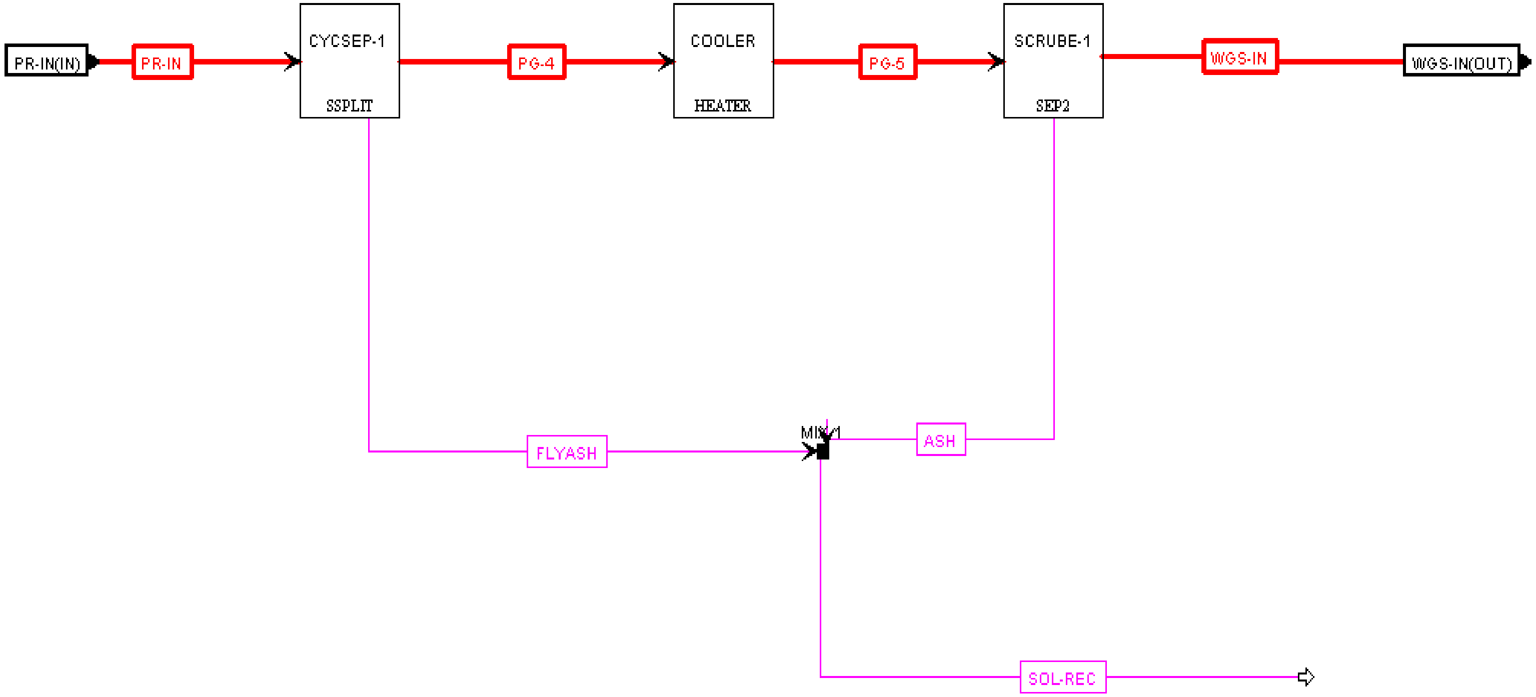


Figure 3 Flow sheet of solid particulate removal section (PR).

Appendix I (4/11)

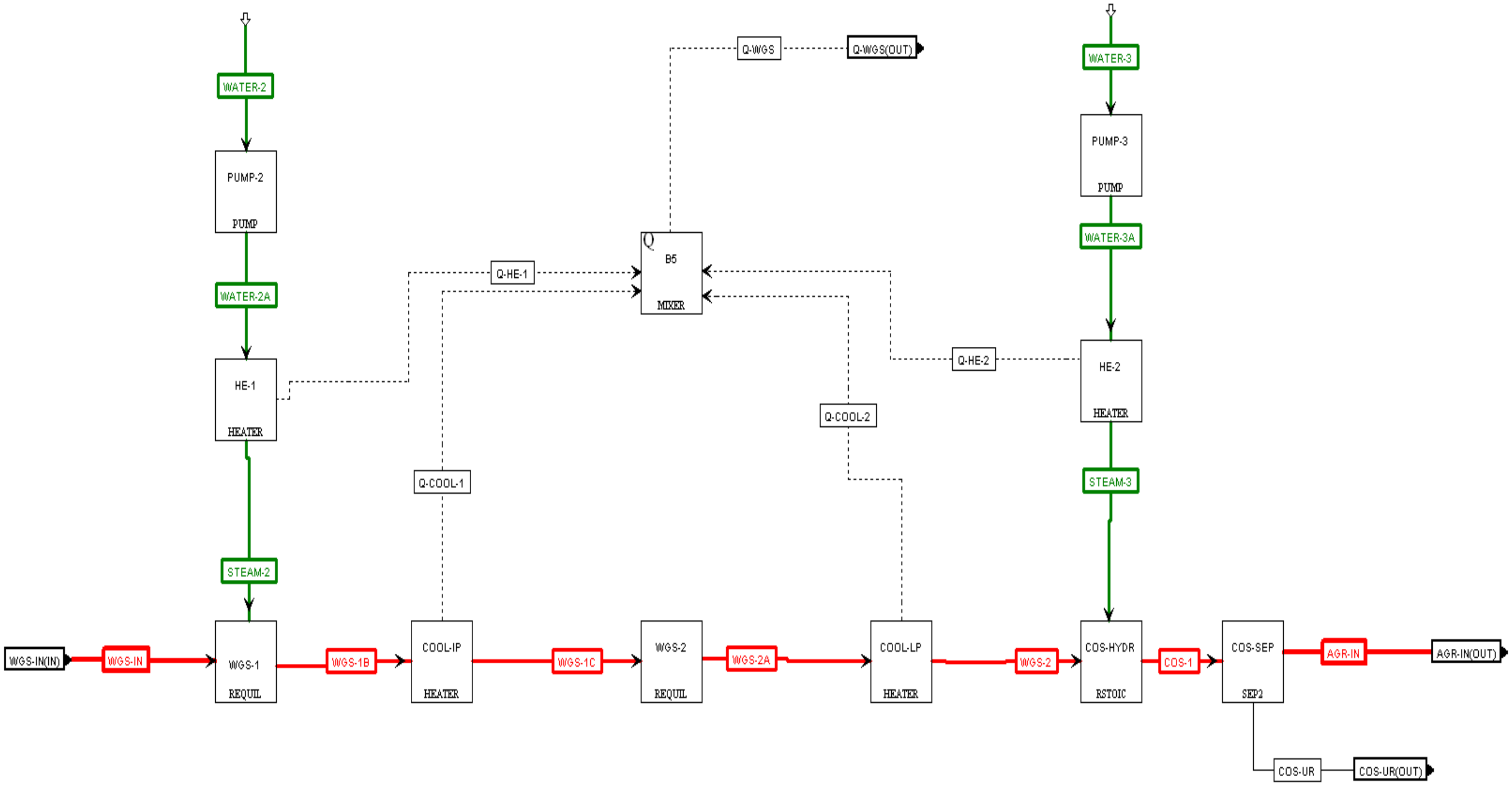


Figure 4 Flow sheet of water gas shift (WGS) reaction system.

Appendix I (5/11)

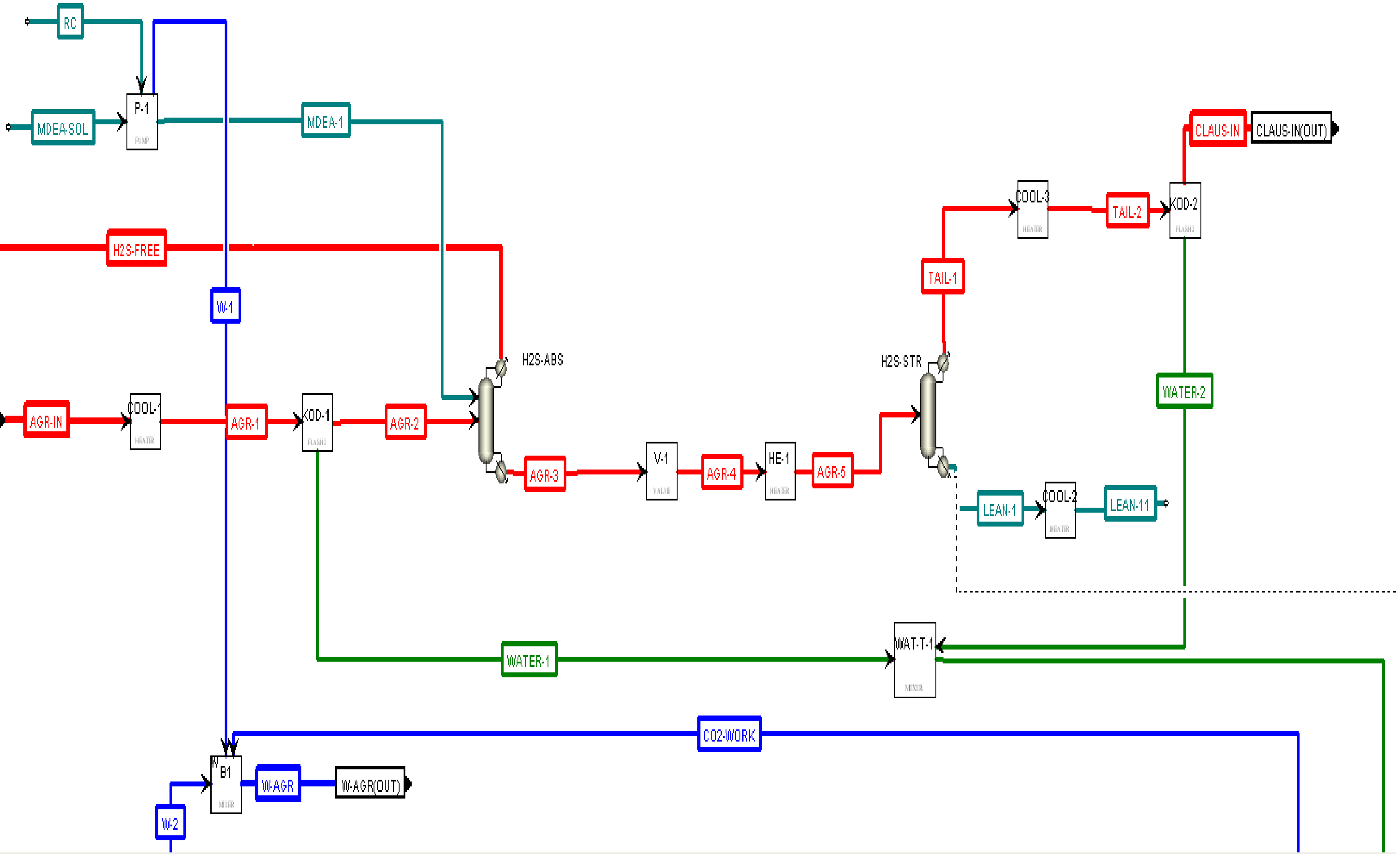


Figure 5 Flow sheet of acid gas removal section (AGR) with H_2S removal section

Appendix I (6/11)

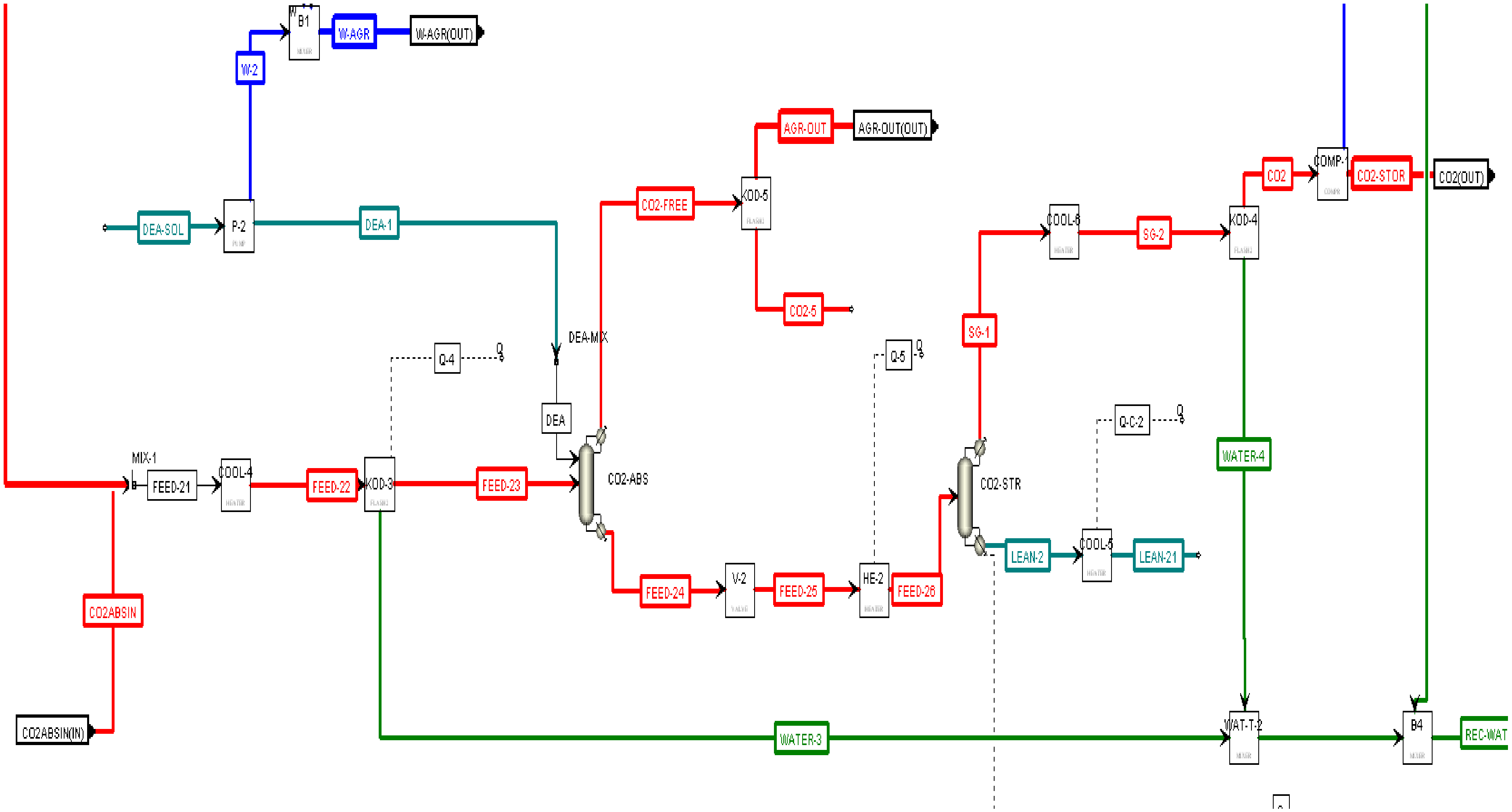


Figure 6 Flow sheet of acid gas removal section (AGR) with CO_2 removal section

Appendix I (7/11)

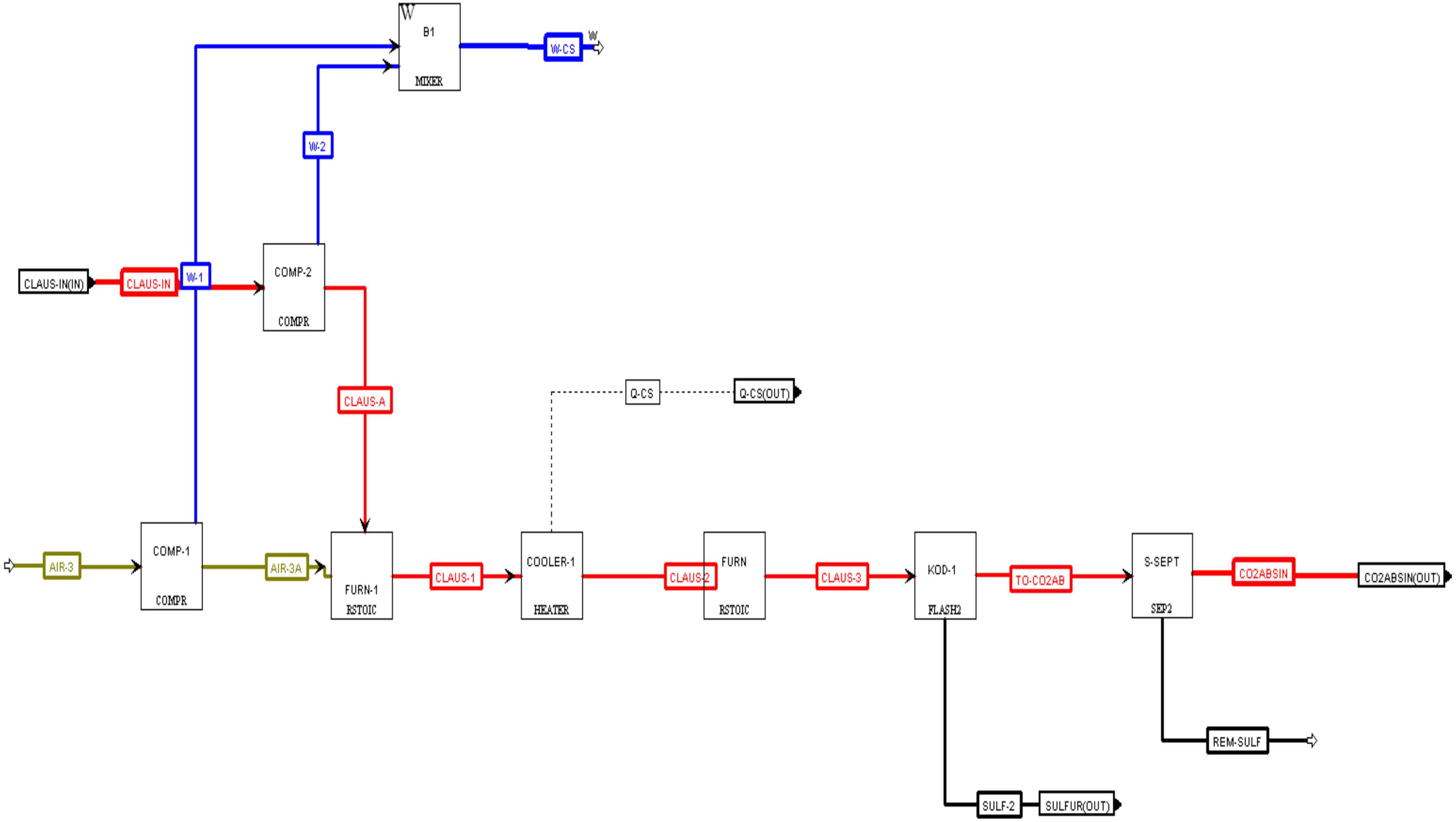


Figure 7 Flow sheet of claus process (CS)

Appendix I (8/11)

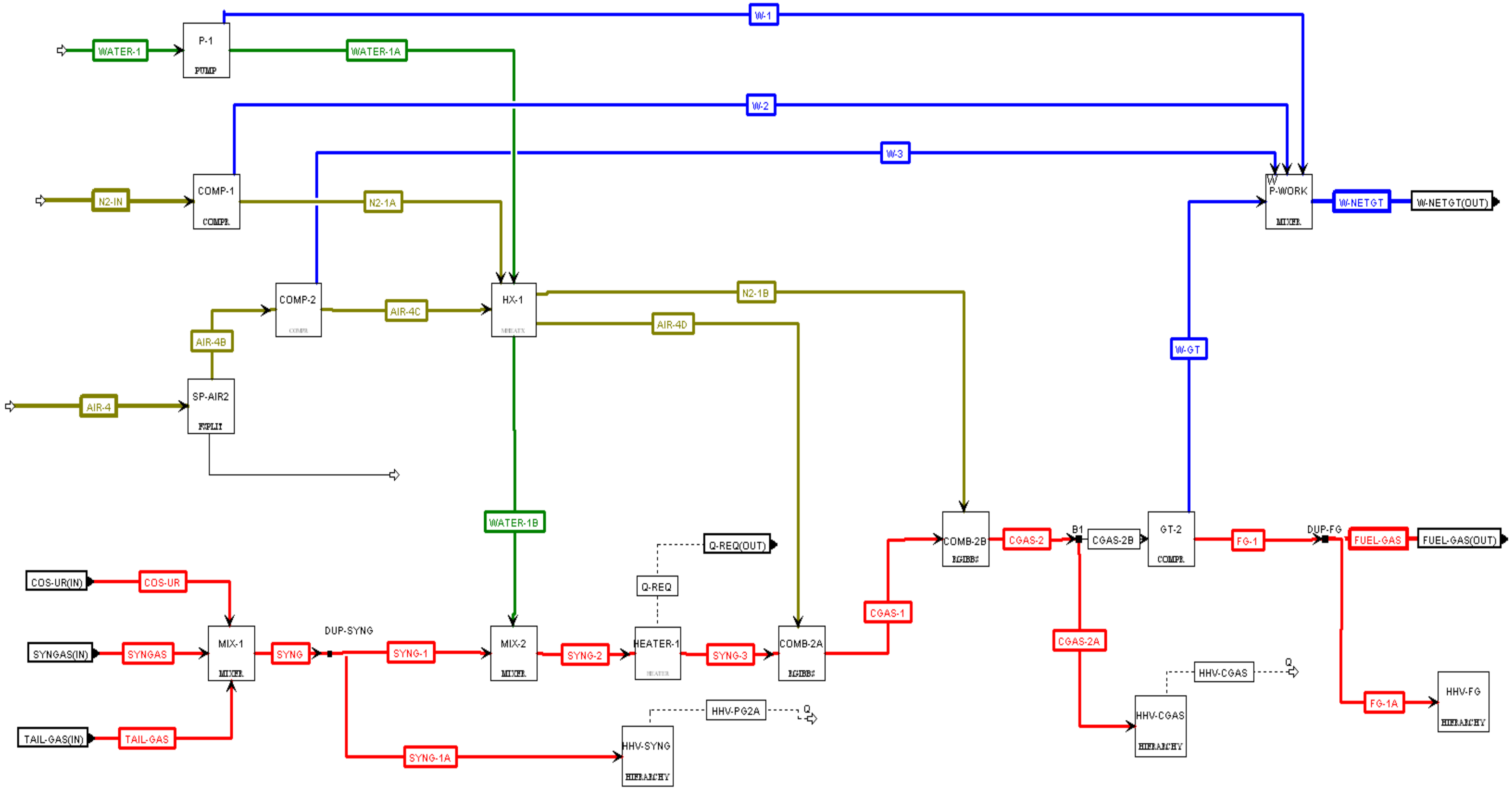


Figure 8 Flow sheet of gas turbien (GT) section

Appendix I (9/11)

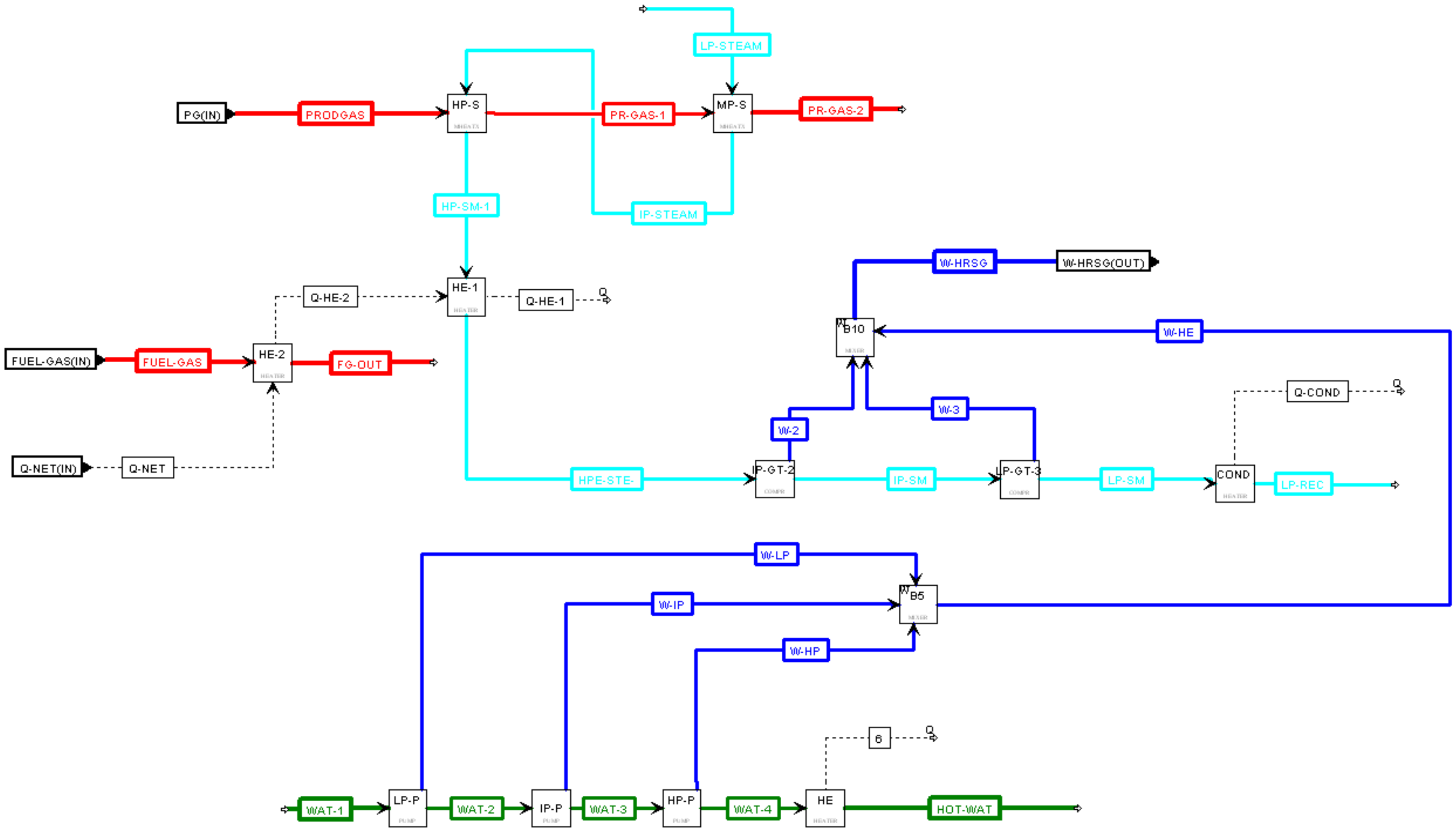
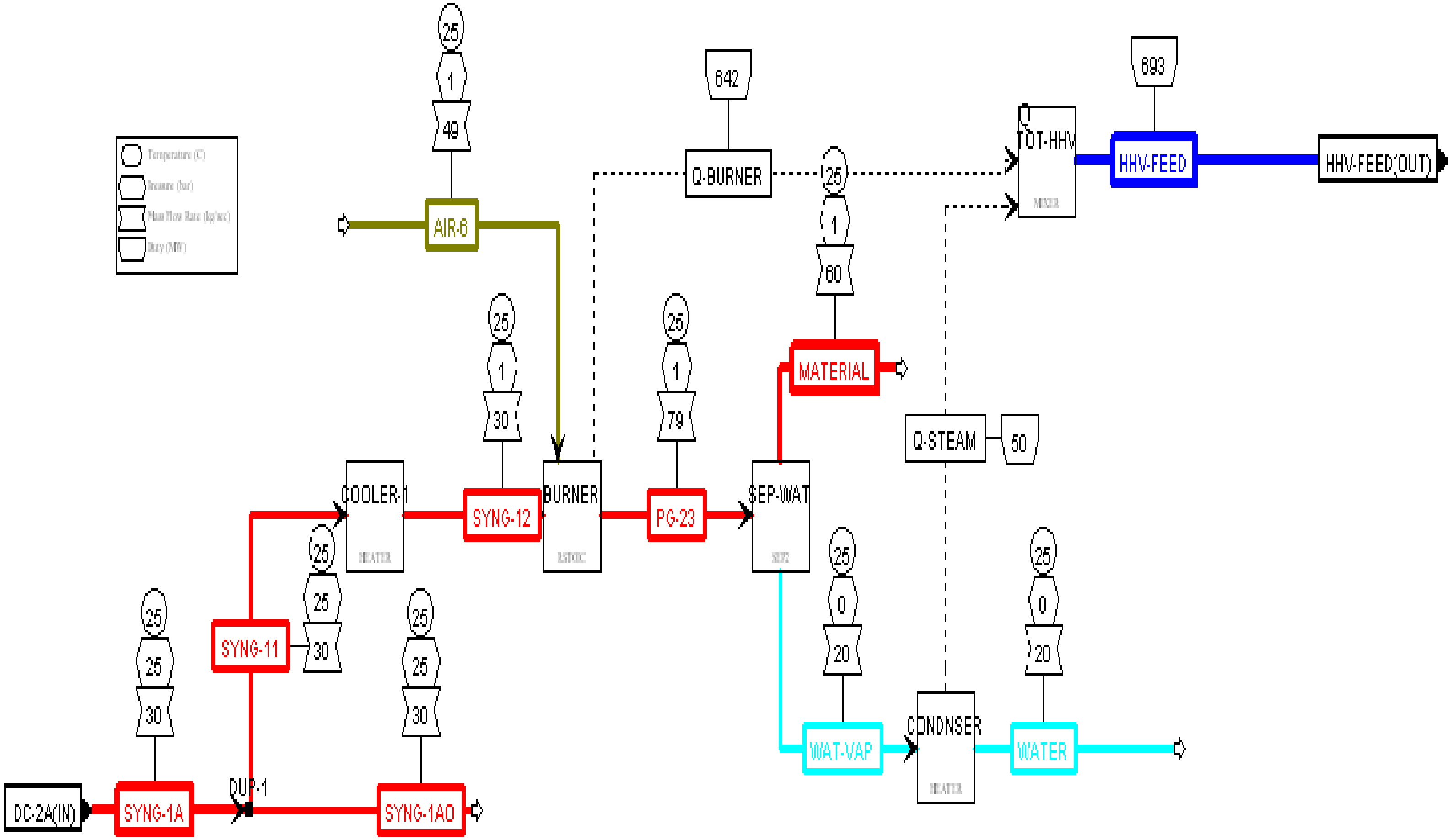


Figure 9 Flow sheet of HRSG

Appendix I (10/11)



Firgure 10 HHV calculation flowsheet

Appendix I (11/11)

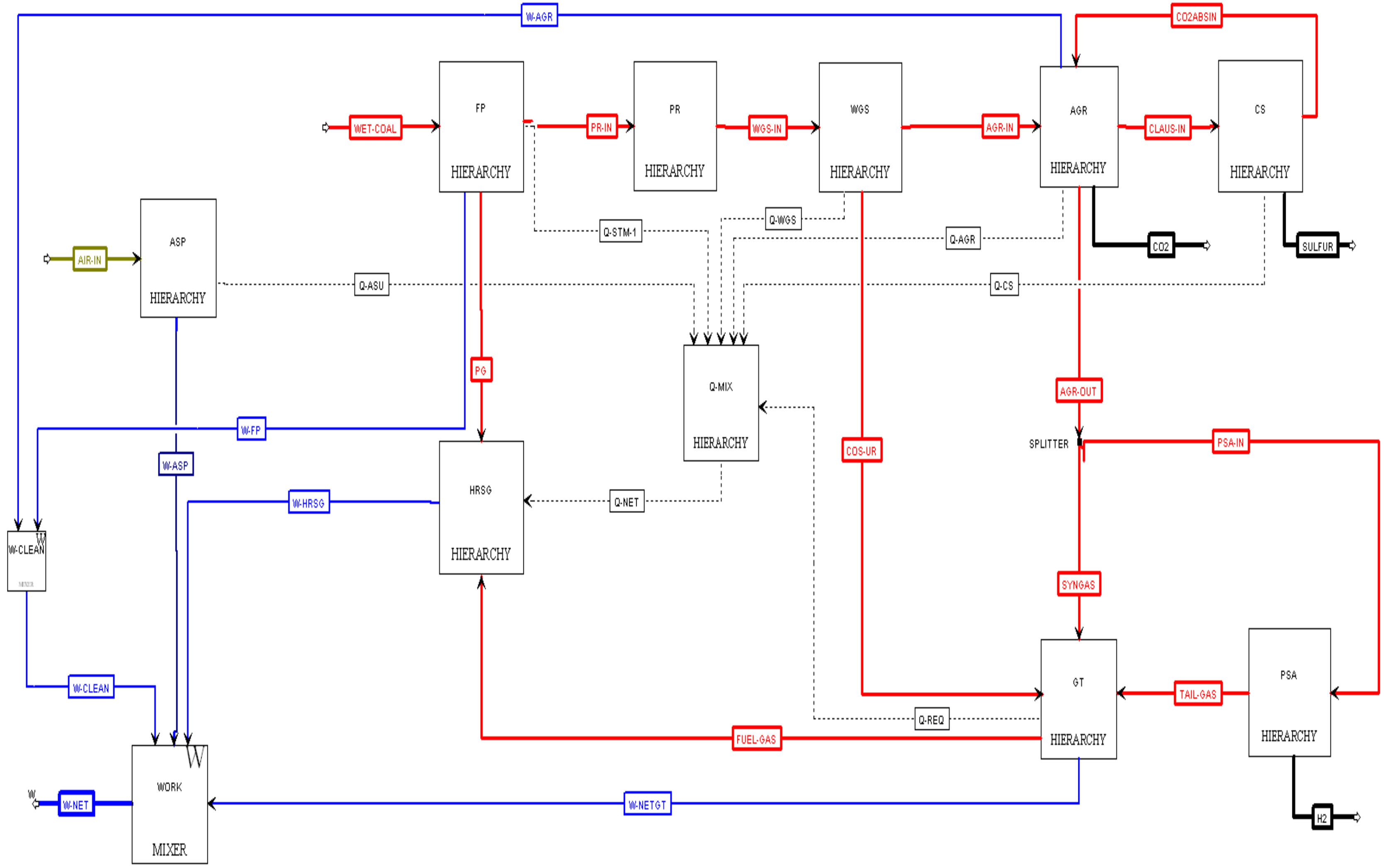


Figure 11 Flow sheet of overall process.

APPENDIX II (1/5)

Dulong formula

$$HHV = 33.823C + 144.249(H - O/8) + 9.418S \tag{1}$$

where HHV is in (MJ/Kg) and C, H, S are on (ar) basis

Modified Dulong formula

$$HHV = 3.491C + 1.178H + 0.1005S - 0.103O - 0.0151N - 0.0211ASH \tag{2}$$

where HHV is in (MJ/Kg) and C, H, S are on (ar) basis

Appendix II (2/5)

	Stream - O_2	Stream -N_2
Vol % N_2	0.00	0.98
Vol % O_2	1.00	0.02
Total Flow kmol/sec	1.0	4.0
Total Flow kg/sec	32.0	112.0
Total Flow cum/sec	21.87	80.03
Pressure bar	16.45	16.45
Temperature C	1.10	1.20

Table 1 N_2 and O_2 streams from ASU.

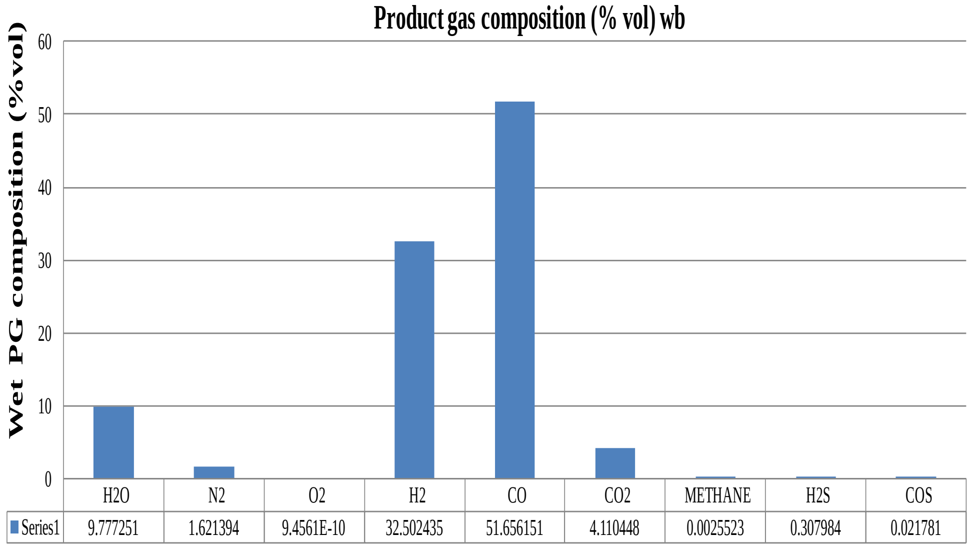


Table 2 Product gas composition on (wb) for base case

Appendix II (3/5)

Mole Fraction	WGS-IN	AGR-IN
H_2O	9.78	21.80
N_2	1.62	0.91
O_2	0.00	0.00
NO_2	0.00	0.00
H2	32.50	45.67
C	0.00	0.00
CO	51.66	1.74
CO_2	4.11	29.69
METHANE	0.003	0.0014
H_2S	0.31	0.19
COS	0.02	0
Total Flow kmol/sec	2.22	3.60
Total Flow kg/sec	42.63	67.39
Total Flow cum/sec	4.32	6.35
Temperature C	240.00	200.08
Pressure bar	22.00	22.00

Table 3 Product gas on (wb) before and after shift reaction for base case

Absorber Section	H_2S Removal	CO_2 Removal
Pressure bar	22	22
Inlet PG Temp. C	67	50
Outlet PG Temp. C	38	73
Inlet Solution Temp. C	40	70
Outlet CO_2 Temp. C	45	72
Stripper Section		
Pressure bar	1	1
Inlet Solution Temp. C	80	99
Outlet Solution Temp. C	98	102
Outlet Tail gas Temp. C	80	99

Table 4 Process parameters for AGR section.

Appendix II (4/5)

	H_2S Removal		CO_2 Removal	
	IN, Kmole/s	OUT,Kmole/s	IN, Kmole/s	OUT, Kmole/s
H_2O	0.761	0.015	0.029	0.036
N_2	0.032	0.002	0.040	0.004
O_2	0.000	0.000	0.000	0.000
H_2	1.594	0.000	1.594	0.000
CO	0.061	0.004	0.060	0.008
CO_2	1.036	0.330	0.989	0.785
METHANE	0.000	0.000	0.000	0.000
H2S	0.006	0.004	0.002	0.002
MDEA	0.000	0.000	0.000	0.000
DEA	0.000	0.000	0.000	0.000
Total Flow kmol/sec	3.49	0.36	2.71	0.83
Total Flow kg/sec	65.34	15.10	50.15	35.56
Total Flow cum/sec	6.17	8.91	3.62	0.38
Temperature C	200.10	30.00	80.38	566.31
Pressure bar	22.00	1.00	22.00	155.00

Table 5 Stream results for H_2S and CO_2 removal section

Mole Frac	Sulfur
H_2O	0.02
S	0.98
Total Flow kmol/sec	0.00
Total Flow kg/sec	0.14
Total Flow cum/sec	0.00
Temperature C	257
Pressure bar	22

Table 6 Sulfur gas composition

Appendix II (5/5)

	H_2
Total Flow kmol/sec	0.68
Total Flow kg/sec	1.37
Total Flow cum/sec	0.89
Temperature C	73.29
Pressure bar	22.00

Table 7 H_2 gas composition

Mole Frac	Before saturation	After saturation	After combustion	After dilution to GT	Flue gas from GT
H_2O	2.24	22.19	35.82	16.73	16.73
N_2	2.99	2.38	56.23	79.70	79.70
O_2	0.000	0.000	0.835	0.431	0.431
NO2	0.00	0.00	0.0002567	0.0000620	0.0000620
NO	0.00	0.00	0.2545400	0.0166840	0.0166840
S	0.000	0.000	0.000	0.000	0.000
SO2	0.000	0.000	0.000	0.000	0.000
SO3	0.000	0.000	0.000	0.000	0.000
H_2	75.442	60.045	0.133	0.000	0.000
CL2	0.00	0.00	0.00	0.00	0.00
HCL	0.00	0.00	0.00	0.00	0.00
C	0.00	0.00	0.00	0.00	0.00
CO	4.30	3.42	0.13	0.00	0.00
CO_2	15.03	11.96	6.59	3.13	3.13
METHANE	0.00	0.00	0.00	0.00	0.00
H_2S	0.00	0.00	0.00	0.00	0.00
AMMONIA	0.00	0.00	0.00	0.00	0.00
NIO	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00
METHANOL	0.00	0.00	0.00	0.00	0.00
Total Flow kmol/sec	1.21	1.53	3.49	7.50	7.50
Total Flow kg/sec	12.85	18.47	89.00	201.56	201.56
Total Flow cum/sec	1.53	2.32	29.83	41.95	447.32
Temperature C	63.56	144.91	1981.76	1200.00	515.24
Pressure bar	22.00	22.00	22.00	22.00	1.10
Vapor Frac	0.99	0.97	1.00	1.00	1.00
Liquid Frac	0.01	0.03	0.00	0.00	0.00
Solid Frac	0.00	0.00	0.00	0.00	0.00
Enthalpy kJ/kmol	-68708.93	-102109.63	-36897.19	-12939.22	-37347.45

Table 8 Material balance around GT section

APPENDIX III

Appendix III (1/4)

ER	TPG1 C	Syn Gas Kg mole/s
0.27	1050.19	1.68
0.28	1057.56	1.69
0.28	1065.41	1.70
0.28	1073.84	1.71
0.28	1082.95	1.72
0.29	1092.73	1.73
0.29	1103.36	1.73
0.29	1114.87	1.74
0.29	1127.34	1.74
0.29	1140.83	1.75
0.30	1155.22	1.75
0.30	1170.59	1.75
0.30	1186.81	1.75
0.30	1203.76	1.74
0.31	1221.32	1.74
0.31	1239.34	1.74
0.31	1257.70	1.73
0.31	1276.32	1.73
0.32	1295.23	1.72
0.32	1314.14	1.72
0.32	1333.11	1.71
0.32	1352.09	1.71
0.33	1371.07	1.70
0.33	1390.02	1.70
0.33	1408.92	1.69
0.33	1427.76	1.68
0.34	1446.55	1.68
0.34	1465.27	1.67
0.34	1483.92	1.67
0.34	1502.50	1.66
0.35	1521.00	1.65
0.35	1539.42	1.65
0.35	1557.77	1.64
0.35	1576.05	1.64
0.36	1594.15	1.63
0.36	1612.28	1.62
0.36	1630.34	1.62
0.36	1648.30	1.61
0.36	1666.29	1.61
0.37	1684.11	1.60
0.37	1701.86	1.59
0.37	1719.54	1.59
0.37	1737.15	1.58
0.38	1754.68	1.58
0.38	1772.14	1.57
0.38	1789.54	1.56
0.38	1806.86	1.56
0.39	1824.12	1.55

Table 1 Effect of ER on gasification temperature and syngas production

Appendix III (2/4)

ER	Net Power, MW	CGE	HGE	CCE
0.27	114.99	82.35	91.79	51.16
0.28	116.96	82.25	91.79	51.93
0.28	118.82	82.14	91.80	52.67
0.28	120.42	82.03	91.80	53.34
0.28	121.84	81.91	91.80	53.96
0.29	121.64	81.78	91.81	54.13
0.29	120.36	81.64	91.81	53.99
0.29	119.11	81.48	91.81	53.86
0.29	117.92	81.32	91.81	53.74
0.29	116.75	81.15	91.81	53.63
0.30	115.73	80.96	91.81	53.56
0.30	114.59	80.76	91.81	53.45
0.30	113.63	80.56	91.81	53.40
0.30	112.70	80.34	91.81	53.35
0.31	111.68	80.11	91.81	53.27
0.31	110.78	79.88	91.81	53.23
0.31	109.91	79.64	91.80	53.19
0.31	109.05	79.40	91.80	53.16
0.32	108.19	79.15	91.80	53.13
0.32	107.34	78.91	91.80	53.10
0.32	106.34	78.66	91.80	53.02
0.32	105.47	78.41	91.79	52.98
0.33	104.62	78.15	91.79	52.95
0.33	103.78	77.90	91.79	52.92
0.33	102.94	77.65	91.78	52.89
0.33	102.11	77.39	91.78	52.87
0.34	101.28	77.14	91.78	52.84
0.34	100.45	76.89	91.78	52.81
0.34	99.63	76.63	91.77	52.79
0.34	98.80	76.38	91.77	52.76
0.35	97.98	76.12	91.77	52.74
0.35	97.15	75.87	91.76	52.71
0.35	96.33	75.62	91.76	52.68
0.35	95.51	75.36	91.76	52.66
0.36	94.63	75.11	91.75	52.61
0.36	93.83	74.85	91.75	52.59
0.36	93.03	74.60	91.75	52.57
0.36	92.21	74.34	91.74	52.55
0.36	91.39	74.09	91.74	52.52
0.37	90.58	73.83	91.74	52.49
0.37	89.76	73.58	91.73	52.47
0.37	88.89	73.32	91.73	52.42
0.37	88.07	73.07	91.73	52.40
0.38	87.25	72.81	91.72	52.37
0.38	86.44	72.56	91.72	52.34
0.38	85.62	72.30	91.72	52.32
0.38	84.81	72.05	91.71	52.29
0.39	84.00	71.79	91.71	52.26

Table 2 Effect of ER on net power production, CGE, HGE and CCE.

Appendix III (3/4)

ER	H_2 ,Kg/s
0.27	1.38
0.28	1.39
0.28	1.40
0.28	1.41
0.28	1.41
0.29	1.42
0.29	1.42
0.29	1.43
0.29	1.43
0.29	1.43
0.30	1.44
0.30	1.44
0.30	1.44
0.30	1.43
0.31	1.43
0.31	1.43
0.31	1.43
0.31	1.42
0.32	1.42
0.32	1.41
0.32	1.41
0.32	1.41
0.33	1.40
0.33	1.40
0.33	1.39
0.33	1.39
0.34	1.38
0.34	1.38
0.34	1.37
0.34	1.37
0.35	1.37
0.35	1.36
0.35	1.36
0.35	1.35
0.36	1.35
0.36	1.34
0.36	1.34
0.36	1.33
0.36	1.33
0.37	1.32
0.37	1.32
0.37	1.31
0.37	1.31
0.38	1.31
0.38	1.30
0.38	1.30
0.38	1.29
0.39	1.29

Table 3 Effect of ER on H_2 production

Appendix III (4/4)

	Emission/Net power production		
	CO2	NOx	SOx
ER	Kg /MWatt	mg /MW	mg /MW
0.27	0.29	0.05	1.86
0.28	0.28	0.07	2.86
0.28	0.28	0.10	3.86
0.28	0.28	0.17	4.86
0.28	0.28	0.58	5.86
0.29	0.28	99.18	6.86
0.29	0.28	165.15	7.86
0.29	0.28	208.89	8.86
0.29	0.29	242.14	9.86
0.29	0.29	267.57	10.86
0.30	0.29	288.73	11.86
0.30	0.30	311.47	12.86
0.30	0.31	324.78	13.86
0.30	0.31	335.46	14.86
0.31	0.31	344.07	15.86
0.31	0.32	351.01	16.86
0.31	0.32	356.64	17.86
0.31	0.32	361.26	18.86
0.32	0.32	365.13	19.86
0.32	0.33	368.43	20.86
0.32	0.33	371.32	21.86
0.32	0.34	373.89	22.86
0.33	0.34	376.23	23.86
0.33	0.34	378.40	24.86
0.33	0.35	380.46	25.86
0.33	0.35	382.44	26.86
0.34	0.35	384.36	27.86
0.34	0.36	386.24	28.86
0.34	0.36	388.11	29.86
0.34	0.36	390.19	30.86
0.35	0.37	392.16	31.86
0.35	0.37	394.07	32.86
0.35	0.38	395.97	33.86
0.35	0.38	397.86	34.86
0.36	0.39	399.76	35.86
0.36	0.39	401.75	36.86
0.36	0.39	403.72	37.86
0.36	0.40	405.71	38.86
0.36	0.40	407.73	39.86
0.37	0.41	409.78	40.86
0.37	0.41	411.86	41.86
0.37	0.42	414.24	42.86
0.37	0.42	416.40	43.86
0.38	0.43	418.60	44.86
0.38	0.43	420.84	45.86
0.38	0.44	423.40	46.86
0.38	0.44	425.65	47.86
0.39	0.45	428.07	48.86

Table 4 Effect of ER on CO_2, NOx and SOx production per net power production

APPENDIX IV (1/3)

	Mix-1	Mix-2	Mix-3	Mix-4	Mix-5	Mix-6	Mix-7
Bio Mass wt %	50	60	40	25	75	0	100
Coal wt %	50	40	60	75	25	100	0
Coal input, HHV (KJ/Kg)	29207.88	29207.88	29207.88	29207.88	29207.88	29207.88	29207.88
Biomass input, HHV (KJ/Kg)	16657.65	16657.65	16657.65	16657.65	16657.65	16657.65	16657.65
Mix, HD db (MW)	686.61	690.72	680.73	667.87	692.63	636.47	678.97
Feed, Kg/s	30.76	30.76	30.76	30.76	30.76	30.76	30.76
Product gas, HD wb (Mwatt)	512.42	501.96	518.14	521.10	480.68	514.77	444.31
Fuel gas, HD wb (Mwatt)	318.41	311.49	322.32	324.41	297.47	321.06	273.54

Equivalence ratio (ER) used	0.35	0.36	0.35	0.34	0.36	0.32	0.35
O_2 per kg of Wet Coal	0.46	0.46	0.47	0.47	0.46	0.46	0.42

ENERGY BALANCE

Net Internal Power use, Mwatt							
ASU	33.37	38.75	39.57	40.00	37.10	39.85	33.80
Gas Cleaning	40.95	32.86	32.45	34.03	36.00	31.16	33.47
Net Cosnusmption	74.32	71.61	72.02	74.03	73.10	71.01	67.27

Net Internal power production, Mwatt							
Gas Turbine	108.85	109.25	113.32	113.53	104.27	113.10	95.76
HRSG	59.91	55.82	57.06	57.45	53.41	59.34	44.64

Net Overall	Mix-1	Mix-2	Mix-3	Mix-4	Mix-5	Mix-6	Mix-7
Work Power, MW	94.45	92.91	98.36	96.69	84.91	101.43	73.13

	Mix-1	Mix-2	Mix-3	Mix-4	Mix-5	Mix-6	Mix-7
CGE %	74.63	72.67	76.12	78.02	69.40	80.88	65.44
HGE %	90.60	89.04	91.77	93.31	86.35	95.58	83.92
CCE %	53.00	52.99	52.86	52.70	53.01	53.71	51.33
THE %	13.76	13.45	14.45	14.48	12.26	15.94	10.77

Table 1 Energy balance, net power production, and efficiencies of process with corresponding feed streams

Appendix IV (2/3)

	Mix-1	Mix-2	Mix-3	Mix-4	Mix-5	Mix-6	Mix-7
Coal input, HHV (KJ/Kg)	29207.88	29207.88	29207.88	29207.88	29207.88	29207.88	29207.88
Biomass input, HHV (KJ/Kg)	16657.65	16657.65	16657.65	16657.65	16657.65	16657.65	16657.65
Mix, HD db (MW)	686.61	690.72	680.73	667.87	692.63	636.47	678.97
Feed, Kg/s	30.76	30.76	30.76	30.76	30.76	30.76	30.76
Product gas, HD wb (Mwatt)	512.42	501.96	518.14	521.10	480.68	514.77	444.31
Fuel gas, HD wb (Mwatt)	318.41	311.49	322.32	324.41	297.47	321.06	273.54
Equivalence ratio used	0.35	0.36	0.35	0.34	0.36	0.32	0.35
O_2 per kg of Wet Coal	0.46	0.46	0.47	0.47	0.46	0.46	0.42
Steam per kg of Wet Coal	0.00	0.00	0.00	0.00	0.00	0.00	0.10
MATERIAL BALANCE							
Consumption , Kg/s							
Air, Kg/s	218.67	209.13	214.53	218.00	200.04	215.31	182.76
N_2, Kg/s	106.79	107.97	112.45	112.23	102.61	112.87	92.16
O_2, Kg/s	14.18	14.14	14.30	14.43	14.00	14.27	12.89
H_2O for GT section, Kg/s	5.10	5.52	5.63	6.13	5.33	5.56	5.00
H_2 production, Kg/s	1.35	1.32	1.37	1.37	1.27	1.35	1.17
Net Water consumption for steam, K	111.73	114.82	108.63	107.13	103.07	110.56	89.74
Production, Kg/s							
Product gas, Kg/s	39.07	40.22	38.02	36.40	41.88	33.39	44.19
Syn gas, Kmole/s	1.64	1.60	1.65	1.67	1.53	1.65	1.41
Clean gas, Kg/s	66.42	67.38	65.34	63.45	68.48	59.48	69.47
Flue gas, Kg/s	193.33	194.03	201.46	201.11	184.80	201.47	168.80
H_2 production, Kg/s	1.35	1.32	1.37	1.37	1.27	1.35	1.17
Sulfur production, Kg/s	0.13	0.10	0.14	0.16	0.07	0.19	0.00
CO_2 production, Kg/s	37.00	36.17	35.39	35.33	37.68	34.18	37.45

Table 2 Material balance of the process with corresponding feed streams

Appendix IV (3/3)

	Mix-1	Mix-2	Mix-3	Mix-4	Mix-5	Mix-6	Mix-7
POLUTTANT EMISSIONS							
Nox and SOx, production/power production							
NOx, mg/s	36748.67	36410.90	37769.55	37883.75	34703.84	37693.47	31603.94
NOx, ppm	190.08	187.65	187.48	188.37	187.80	187.09	187.22
NOx, mg/Mwatt	389.09	391.90	384.00	391.78	408.72	371.61	432.15
Sox, mg/s	242.51	209.24	270.84	303.55	148.36	339.01	0.00
SOx, ppm	1.25	1.08	1.34	1.40	0.80	1.68	0.00
SOx, mg/Mwatt	2.57	2.25	2.75	3.14	1.75	3.34	0.00
CO_2 capture, Kg/Mwatt	0.39	0.39	0.36	0.37	0.44	0.31	0.51
UNIT PROCESS PRODUCTION							
Unit production/power production							
Gasification, Kmole/s	1.99	2.00	1.97	1.92	2.01	1.83	2.03
ASU, Kmole/s	5.16	4.89	4.99	5.09	4.68	5.02	4.26
Gas Cleaning, Kmole/s	1.85	1.83	1.89	1.88	1.75	1.88	1.64
Fuel gas, Kmole/s	1.18	1.17	1.21	1.19	1.12	1.21	1.06
Flue gas, Kmole/s	7.21	7.23	7.50	7.51	6.89	7.50	6.28
Unit production/power production							
Gasification, gmmole/Mwatt	21.03	21.55	19.99	19.91	23.72	18.02	27.71
ASU, gmmole/Mwatt	54.67	52.59	50.74	52.68	55.10	49.54	58.29
Gas Cleaning, gmmole/Mwatt	19.58	19.71	19.24	19.40	20.62	18.52	22.36
HRSG, gmmole/Mwatt	76.34	77.80	76.27	77.72	81.12	73.92	85.92

Table 3 Pollutant emissions, CO_2 capture and unit production of the process with corresponding feed stream