Study on feasibility of coir dust as feedstock for entrained flow gasification system

In partial fulfilment of the requirement for the degree

of

Master of Technology

in

Thermal Engineering Specialization

by

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CERTIFICATE

This is to certify that the report entitled **"Study on feasibility of coir dust as feedstock for entrained flow gasification system"** submitted to the National Institute of Technology, Rourkela by **Debabrata Singh**, Roll No. **211ME3358** in partial fulfillment of the requirement for the degree of **Master of Technology** in **Department of Mechanical Engineering** with specialization in **Thermal Engineering** is a record of bona fide work carried out by him under my supervision and guidance.

Place: NIT Rourkela, **Date:** 2nd June, 2014.

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List of Symbols and Abbreviations

Symbol	Description
ρ	Density of the fluid (Kg/m^3)
g	Acceleration due to gravity (m/s^2)
ΔH_{fR}	Heat of formation of the reactants (KJ/Mole)
ΔH_{fp}	Heat of formation of the products (KJ/Mole)
M_R	Mass of the reactants (Kg)
M_p	Mass of the products (Kg)
C_{PR}	Specific heat of the reactants (Kcal/Kg-K)
C_{Pp}	Specific heat of the products (Kcal/Kg-K)
ΔT_R	Change of temperature of the reactants (^{0}C)
ΔT_p	Change of temperature of the products (^{0}C)
P_{CO2}	Partial pressure of carbon dioxide
P_{H2}	Partial pressure of Hydrogen
P_{CO}	Partial pressure of carbon monoxide
P_{H2O}	Partial pressure of water
K	Equilibrium constant
ER	Equivalence Ratio
PSD	Particle size distribution
MC	Moisture content

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Abstract

Experimental investigations of the influence of equivalence ratio on gas composition, adiabatic flame temperature, calorific value and rate of gas generation were performed using coir dust as feedstock in entrained flow gasification system. Experiments were carried out on a pilot scale (2.5 m high X 0.25 m i.d.) existing entrained flow biomass gasifier installed at IMMT, Bhubaneswar. Model calculations were made to find out the composition and other properties of the gas taking coir dust as feedstock. Results were realized through comparison of output from theoretical as well as experimental value obtained with varying equivalence ratio. The outputs obtained from Entrained flow gasifier are of high quality and the process can be industrialized.

Keywords: *Entrained flow gasification, adiabatic flame temperature, shift reaction & equivalence ratio.*

1. INTRODUCTION

1.1 Utilization of biomass

World energy consumption has been rising worldwide, as developing nations begin to industrialize and as developed nations buy more energy consuming appliances to make life more comfortable. If these trends continue, we may choke on account of less energy supply. Fossil fuels continue to dominate as the main sources of energy produced and consumed worldwide despite the growth of the share of renewables in both production and consumption, especially the monopoly in the transport sector. As fossil supplies decline, renewables will take their place although we are all concerned whether the energy supplies we have can sustain the developing world that is rapidly becoming energy hungry. At this important juncture we just might have found our efficient replacement *i.e.,* Syn Gas produced from Biomass Gasification. Biomass is a term used to denote carbon containing plant or animal product. Animal products (excreta) are considered biomass only when it contains mainly saccarides (cellulose, starch etc.) and peptides. Biomass thus, is an organic material, which has stored sunlight in the form of chemical energy in its biochemical activities. Biomass fuels include wood, wood waste, straw, manure, sugar cane, bagasse and many other by products from a variety of agricultural processes. Biomass, at present, is considered to be a major energy source besides fossil fuels like coal and petroleum. Energy, in its many useful forms, is a basic necessity that Influences standard of living and technological growth in all societies. The fossil fuel reserves have mostly been explored in twentieth century. The apprehension of facing a crisis in the next century due to exhaustion of all sorts of fossil fuels has drawn attention of the nations all over the world towards exploiting non-conventional energy sources particularly on renewable energy resources. Biomass as available in the society may be virgin or waste depending on its utility in the society. When burned, the chemical energy is released as heat. In fact, biomass continues to be a major source of energy

in the developing world because of its renewed ability. Biomass from plant products makes major part of the total biomass.

In developing countries like India, biomass in its natural form in dry condition is still widely used in rural communities as a major heat source, setting aside a small part of it from manure and compost. In this respect too, biomass is used very selectively and a large variety of biomass is allowed to perish in the environment. Leaving aside the selective type of biomass, which is lumpy and smokeless and used for domestic heating purposes, light grainy and powdery biomass finds no use. Moreover, their uses in industrial sector is very limited as they occur in a scattered manner and are not collected in an organized way due to their bulk volume and end-value. India produces about 420 million tons of biomass (fire wood: 220 million tons, and Agro Wastes including powdery biomass: 220 millions tons). Considering the vast availability, it may be expected that small industrial installation with an appropriate technology may either add value to it or exploit efficiently the potential heat energy present in it. Relatively new technologies have come up to help in this respect. Biomass, instead of being used in solid form directly is converted for use in gaseous form through gasification route.

1.2 Biomass Gasification

Biomass gasification is the technology to convert solid woody cellulose material into combustible gaseous product. It is a developing energy technology along with various systems for utilization of biomass. This technology produces gases and makes biomass very important source of a versatile fuel. There are two ways of gasifying biomass. Unlike the first route of bacterial conversion named as Bio-methanation where a type of bacterial converts saccarides into methane, thermal gasification is the other route to convert solid carbonaceous fuels into flammable gas mixtures comprising of carbon monoxide and hydrogen along with

some amount of carbon dioxide and nitrogen. Gaseous fuel thus obtained can be used as fuel for power engines and other thermal applications, as reducing medium or as raw material for some chemical and electro-chemical industries. The heat content (enthalpy) of the produced gas (producer gas) may be used in agricultural sector for par boiling of paddy, drying of grains etc. and in industrial sectors requiring process heat, as a retrofit for oil fired boilers, in ovens, kilns and dryers. With the aim of generating power from it, the produced gas may be connected to gas-fired engines or gas turbines to achieve significantly higher electrical efficiencies. The carbon monoxide and hydrogen rich gas can also be used in metallurgical applications for reduction purposes as in sponge iron kiln.

It is known from the literature that thermal biomass gasification is done in one of the three different ways. They are fixed bed, fluidized bed, and entrained flow techniques. Different technologies have been developed to process different shapes and sizes of biomass.

Fixed bed gasifier:

Fixed bed gasifier is handy for elongated stiff lumpy chips type of solid biomass. It can tolerate wide range of sizes and shapes. There have been enormous R&D efforts made in the country and abroad to develop fixed bed gasification technology based on lumpy material. Three different chemical reactions i.e. combustion of char (oxidation), reduction and pyrolysis occur sequentially in distinct zones along the height o the gasifier while biomass is fed from top or bottom and oxygen/air is injected in combustion zone. The produced gas is withdrawn after pyrolysis zone. Relatively low temperature pyrolysis occurs resulting in high amount o tar, oil and phenols in the final product. Efficient cleaning and scrubbing of gas is an essential step and costly part of the process. Up draft, downdraft and cross-draft gasifiers are few among the state of advancement producing different grade of product gas in respect to tar, phenol and oil content. Chips and bigger particles are generally taken as feedstock in fixed beds.

Fluidized bed gasifier:

Fluidized bed technology is another one applicable on small grains of biomass with uniform size. Fluidization of a bed of particle occurs when a gas (or liquid) passes through it at a rate such that particles are suspended, being maintained in a state of continuous random motion. In gasifiers the fluidization is normally achieved by a gasifying medium. Fluidized beds are characterized by their high gas-to-solid heat and mass transfer rates, so that the mixing of solids within the bed approaches that of an ideally mixed reactor, with a resultant uniformity of solids composition and temperature. This makes selective removal of char and ash different, and the bed often has to operate at high ash concentrations to avoid excessive carbon loss in the ash. The heat and mass transfer characteristics of the bed can be modified by the choice of the bed geometry and the positioning of nay internal baffles, heat transfer elements, and so on. The particle size distribution of the feedstock must be carefully regulated to satisfy constraints of low entrainment losses and complete solids fluidization.

Attempts have been made to develop different types of fluidized bed such as twin-fluidized bed, circulating – fluidized bed and sand-aided fluidized bed gasifiers to produce producer gas from biomass. Tar and char in the product are not very low.

Entrained flow gasifier:

In entrained flow gasifiers, the fuel is carried in suspension irrespective of gangue or combustible matter in the flow of the gasifying medium. Each discrete particle is separate from other in the flowing gas stream. The mixing of solids in the reactor approaches that of an ideal plug flow reactor. High velocity leads to high throughput and reaction time is of only a few seconds. Solids get equal share of oxygen without local over-heating as back mixing is minimized. It is expected that cracking may be less prominent allowing partial oxidation to continue as it moves along the reactor.

Dimensions of entrained-flow reactors are determined by particle size, reactivity of the fuel, reaction temperature and gas phase velocity. Product gases generally contain little or no tars, oils or methane, the volatile matter being flashed off and rapidly gasified as the fuel enters the high temperature zone. Given the powdery grains of wider size range as raw material for gasification, entrained flow gasifier is the first option that comes for consideration. The dusty and powdery material like saw dust, coir-dust, paddy husk etc. is difficult to treat in fixed bed reactors due to technical problems like local agglomeration and channelling, thought research is going on to treat powdery bio-mass in fixed bed. An appropriate technology that can efficiently use the last category of biomass is either fluidized bed or entrained flow type. The presences of wide range of particle size in naturally occurring powdery biomass make the entrained flow technique more acceptable than the other one. This is because, one of the requirements that fluid bed demands is relatively closer range of sizes than that present in common powdery biomass.

1.3 Characteristics of Entrained Flow Gasification

Many processes converting biomass into a gaseous fuel have been derived from established coal gasification processes. Though entrained flow gasification is an established technology for coal, its application for biomass has not been sufficiently exploited. The application of entrained flow concept to biomass will have different consideration because of the kinetics and thermodynamics of the conversion reaction. It will also depend on the size of the gasifier for commercial use. Entrained flow gasifier for coal is mostly operated at temperatures in the order of 1600^0C or above. Through injection of oxygen and steam or air with high preheat. Because of the high temperature operating conditions ash is handled in molten form. Tar, oils and phenols are not produced as they are thermally cracked or hydrogenated to lower molecular weight compounds. Some of the well-known and

commercially used entrained flow coal gasifiers are Koppers-Totzok gasifier, Texaco gasifier, Babcock and Wilcox gasifier, Dupont gasifier etc. The entrained flow gasifiers are some time associated with swirling motion of carrier medium so as to increase the mixing intensity as well as residence time of fuel in the gasifier. The cyclone type gasifier, which is a special case of entrained flow system, operates with high intensity vortex flow (high swirl flow). Though there have been reports in literature regarding development of cyclone type coal gasifier, there has not been much of commercial application of the same.

Entrained flow gasification of biomass:

Cyclone type biomass gasifiers have been developed in small scale and tested for it's performance for various types of biomass such as rice husk, sawdust etc. Such cyclone gasifiers are mostly of reverse flow type with limited L/D ratio. Due to less residence time the char is not completely gasified and is collected at the bottom to be re-circulated. These limitations can be overcome by proper design of entrained flow biomass gasifier where the height of the gasifier takes care of required residence time for complete gasification. Modeling of entrained flow gasifiers for biomass has been recently reported in literature, which may be considered useful in designing such system. The characteristics of entrained flow gasification for biomass may be summarized as follows:

- The fuel (biomass) reacts as it is carried under full suspension in the flow of the gasifying medium. In Chemical engineering term, it is a plug flow reactor with intimate radial mixing with minimum axial mixing.
- Along the height of the reactor completely different reaction zones can be designed to be present and different reactions like partial oxidation and shift reaction/steam reformation can be carried independently.

- Due to intimate mixing between the fuel and the medium (air/steam) high intensity reactions can be achieved and gasification of char is maximized.
- Since gasification if planned to takes place at high temperature in the range of 1000^0C with low equivalent ratio, methane (CH_4) format ion may be made less, tar formation eliminated and percentage of H_2 in the gas high.
- Capable of handling wide range of particle size of biomass.

1.5 Objective and Scope of the work

The main aim of this project was to study the existing 30 KW_e capacity entrained flow gasification system at Institute of Minerals & Materials Technology, Bhubaneswar and evaluate its performance with use of the powdery biomass feedstock (coconut coir dust). Entrained flow method will be adopted to ensure full suspension of powdery biomass in a vigorous radial mixing condition in the reactor, which will result in high intensity reaction and gasification. It is intended to operate the designed reactor at a temperature of ~ 1000^0C to take the advantage of kinetics of the process and produce low tar and char formation. The endothermic steam inducted reactions is planned to couple with the gasification reaction to conserve the sensible process heat in the form of more hydrogen and carbon monoxide. The process includes activities like biomass pre-treatment including collection, drying storage, dense flow feeding, and thermal conservation into producer gas followed by conversion of CO to H_2 and CO_2 by shift reaction to add value to this. Keeping in view of the current status of research the following objectives are set in the scope of the present research work:

- Model calculations for different equivalence ratio.
- Analysis of data & interpretation.

2.0. LITERATURE REVIEW ON ENTRAINED FLOW GASIFICATION SYSTEMS.

From process point of view gasification of biomass and gasification of coal are very much similar. Coal gasification technologies using fixed bed, fluidized bed, entrained flow or molten bath systems have been established worldwide. Hence it has not been unusual to utilize some of the design concepts of coal gasifiers to biomass gasifiers, though the scale of operation for the later in comparatively much smaller. In the above context, it has been felt pertinent to study the design aspects of entrained flow coal gasifiers for its possible use for biomass gasification. A few conceptual designs of entrained flow biomass gasifiers are also reported in literature. Some of them have been implemented in the terms of developing experimental prototypes. This literature study would enumerate the design aspects of such gasifiers for coal and biomass.

2.1 Entrained flow coal gasifiers:

A number of studies have been conduced worldwide on coal gasification systems and designs of many different types have been developed and commercialized. They include fixed-bed gasifiers, Fluidized-bed gasifiers and Entrained flow gasifiers. Entrained flow gasifiers use pulverized coal, about 70% of which are smaller than 200 mesh, and have high feedstock flexibility. The coal particles are entrained in the steam-oxygen feed and the recycled gas stream and gasified at residence times of a few seconds, after which the product gas is separated from the ash. The low residence times can offer potentially higher throughputs at elevated pressures. Entrained flow gasifiers can be operated at low temperatures to maintain the ash as a dry solid, or at temperatures well above the ash fusion point in the slagging mode so that the ash is removed as a molten liquid. Operation at higher temperatures results in little or no tars and oils in the product gas. Some of the developed coal gasification systems are enumerated below.

Koppers-Tortize gasifier [1]: This gasifier has been designed as refractory line horizontal cylindrical vessel with conical ends. A homogenous mixture of oxygen and pulverized coal is introduced through coaxial burners at each end. Steam is introduced around these burners to shroud the high temperature reactor zone and protect the burners and refractories from excessive temperatures. The product gas leaves through a water-cooled central vertical outlet. Depending on the temperature achieved, the ash is separated in dry form or as molten slag. This gasifier has been in commercial use at different capacities and operated at atmosphere pressure.

Babcok & Wilcox – du pont gasifier [1]: This atmospheric pressure operated gasifier is a refractory lined vessel having a circular cross section and have primary and secondary reaction zone. Temperatures in the primary zone are high enough to melt the coal as into slag which is tapped at the bottom. Coal is conveyed into the unit by steam at a temperature of 325^0F. Reactants are injected downward with flow reversal occurring in the primary zone.

Ruhrgas Vortex gas producer [1]: This unit is a refractory-lined cylindrical vessel with water cooled walls in the cyclone section at the base. The cyclone chamber has an internal diameter of 6.5 ft and is 7.5 ft high. Above the cyclone chamber the diameter of the shaft increases to 9.5 ft and the shaft is 75 ft high. Fuel is injected with preheated air (1300^0F) into the cyclone chamber through 6 tangential nozzles. The gasifying capacity of the units is 100 tonnes/day using bituminous coal. The unit is also operated with partial injection of oxygen. The operation in general is at atmospheric pressure.

Bureau of Mines Gasifier [1]: This was a pilot scale gasifier made as a refractory lined cylindrical vessel having an internal diameter of 2.5 ft and an effective internal volume of 86 cft. The reactants are fed near the bottom, tangential to a 2 ft diameter circle. Steam was preheated to a temperature ranging from 2600 to 3300^0F in a pebble stove, fired with natural

gas and when necessary oxygen enriched air. The capacity of the gasifier is 200-450 lbs/hr with strongly caking coal and ash is separated in molten form.

Institute of Gas Technology Gasifier [1]: This pilot scale gasifier operated with pulverized coal at a pressure of 7 atmospheres. It is a down flow slagging system having a coal feed rate up to 1450 lbs/hr. with internal diameter of 3 ft and effective length of 6 ft. The gasifier is a refractory-lined non cooled reactor into which the pulverized coal and super heated steam are introduced tangentially at the upper cylindrical section or "cycloniser". Oxygen is introduced axially through multiple jets immediately below the outlet of the cycloniser chamber. The hot gasification product passes through a slag cat pot where liquid ash is separated and the gas then tangentially enters a heat exchanger where fly ash and unreacted carbon are separated.

Texaco gasifier [1]: This is a high-pressure suspension gasifier, which is commercialized and operates at a pressure of 400-500 psi. Instead of dry coal, it uses coal slurry of 40-60 % wt concentration. Preheated oxygen is fed in the gasifier through water cooled nozzles. Molten slag flows through a constricted opening at the bottom.

Lousiana Gasifier [2]: The gasifier was of vertical design with 4.5 ft internal diameter of around 20 ft height. Coal + Oxygen and superheated steam are injected tangentially into the gasifier at the bottom zone at a 3.5 ft diameter circle. The coal and oxygen are injected b a 2 inch dia stainless pipe and steam by 4 inch diameter pipe. The product gas is taken out through the narrowed top outlet of 1.5 ft dia. The gasifying capacity for the unit was in the range of 1600-1800 lbs/hr. Since the performance of this unit from process point of view was not satisfactory it was modified with injection of oxygen + coal and steam by two sets of inclined nozzles. The performance improved with this modification and ash was handled in molten form.

Cyclone Reactors: As seen from afore mentioned gasifiers, many of them are operated with unit flow vortex/cyclonic motion of fuel and carrier medium. A part from this, gasifiers have also been designed as reverse flow cyclones with limited length to diameter ratios. Yaggi and Kunni [3] demonstrated the feasibility of conducting coal gasification experiments in a lab scale reactor of vertical reverse flow design. The diameter of the reactor was 30 cm, length 40 cm and capacity 45 kg/hr. Coal was gasified with steam and oxygen and a small amount of primary air yielding carbon efficiency of 80-98%. Ash retention was 70-80% in molten form.

Barnhart and Laurendean [4] studied coal gasification in a small laboratory scale cyclone reactor of reverse flow vertical type having diameter of 15 cm. length 25 cm and capacity 3-12 kg/hr. Air and coal entered the chamber through four equally spaced sets of tangential concentric stainless steel tubes. Coal and primary air entered through inner tubes of 6.0 mm dia, while secondary air entered through the outer tubes of 12.7 mm dia. On gasification performance the reactor achieved a H_2 content of around 9% and CO of 14%.

ABB combustion Engineering/IGC gasifier [5]: Developed jointly by ABB Combustion Engineering USA and IGC Association, Japan, the CE/IGC gasifier is an air blown entrained flow two state slagging gasifier. It is divided into two sections by a throat on its wall. The coal is pulverized to 70% less than 70 microns and dried to a moisture content of 4% and then pneumatically conveyed to the gasifier, Around 40% of the coal is fed to the first stage, along with all the gasification air and recycled char. The state is sometimes referred to as the “combustor” through it operates under overall reducing conditions. The remaining coal is fed to the second stage or “reducer”. The 1st stage operates at a temperature of 1400-1600^0 C so as to melt the ash. Coal and char gasification takes place along the length of the reducer until the temperature has fallen to a point (950 -1050^0C) where the gasification

reaction for coal becomes too slow for further reaction. The char laden outlet gas is cooled and the char is separated for recycling.

VEW gasifier [5]: Developed by Vereinigte Electrizitawerke Westfalen AG(VEW), this gasifier is of vertical down flow design. It operates at 20 bar pressure and uses a single burner. The air supply is pre heated to 600^0C using a gas burner. The main air supply to gasifer may produce axial or swirl flow of reactants within the gasifier vessel. The reactor is refractory brick-lied and the operating temperature is 1600^0C. The reactor exit temperature is around 1400^0C and hot gas, char and slag are cooled in a radiant gas cooler to 800^0C which is used for evaporation and to superheat steam. The char cools before entry to the convective cooler. A cyclone at the base of the convective cooler removes most of the char particles from the gas stream.

KELLOGG Transport Gasifier[5]: For this gasifier, the coal and solvent are dried and ground to a mean size of 50 microns and the pneumatically conveyed to the gasifier. The refractory lined gasifier is divided into two sections: the lower mixing zone where recycled hot solids, coal, lime stone, steam and air are introduced and the riser section. In the mixing section the fresh coal devitalises and the volatiles are burnt consuming all the supplied oxygen in air. The riser section provides extra residence time for gasification and sulphur capture. The gasifier exit conditions are nominally 788^0C and 17.9 bar and the upward gas velocity is 6.5 m/sec. Hot solids are separated from the fuel gas in cyclones and passed either to a stand pipe for recycle to the gasifier or directly to a char combustor.

Modelling and other studies on entrained flow coal gasifers:

Though a number of entrained flow coal gasifiers have been studied in pilot scale and some of them have been commercially used, literatures on the modelling aspect of such gasifiers are quite limited. Govind and Shah[6] have reported a mathematical model to

simulate the Texaco down flow gasifier which used coal slurry. This model describes the physical and chemical processes occurring in an entrained flow gasifier. Azuhata et al [7] studied the carbon conversion in a small atmospheric pressure entrained flow gasifier of 20 cm diameter 173 cm long. Coal was entrained and injected in a stream of argon which served a as tracer gas. The inlet oxygen was heated to 300^0K and added to the primary stream before injection. Steam was preheated by an electric heater to 590^0k and injected as a secondary stream.

Lee et al [8] studied the characteristics of entrained flow coal gasification in a drop tube reactor {Dia= 5 cam, Length =100 cm}. The effects of reaction temperature, oxygen/coal and steam/coal ratios and residence time on coal gasification were investigated. Liu et al [9] had taken up the modelling of a pressurized entrained flow coal gasifier of 10 cm diameter, 2.5 m length and gasification capacity 258-478 kg/hr, and studied the effect of reaction kinetics and char structure in the reactor. Vamvuka et al [10,11] developed an one dimensional steady state model for bituminous coal gasification under entrained horizontal flow, based on mass and energy balances, heterogeneous reaction rates and homogeneous gas phase equilibrium. Chen et al [12, 13, 14] developed a comprehensive three dimensional model for an entrained flow gasifier of the same design as that of CE/IGC gasifier, mentioned earlier, The two stage gasifier has a diameter of 1.2m and over all height of 13m and is designed to operate with swirl flow in the lower combustion zone. A numerical study on the coal gasification characteristics in a slurry feed type entrained flow gasifier was conducted by Choi et al [15].

2.2. Entrained Flow Biomass Gasifiers

Unlike Entrained Flow coal gasification literature on entrained flow biomass gasification is found to be very limited. Only a few investigations on design and performance tests of entrained flow biomass gasifiers, which also included cyclone type gasifiers have been reported. There are also a couple of papers on CFD modelling of entrained flow gasifiers for biomass.

Fletcher et al [16] conducted 3-dimensional computational fluid dynamics (CFD) simulation using CFX_4 model for design and optimization of entrained flow biomass gasification systems. In order to investigate the optimal design, six different gasifier geometries were considered and the effect of number and location of inlets on the flow field within the reactor were studied. The total height of the gasifier was 2.7 m and the dia of the reaction chamber was 0.3 m. Air was injected through the primary inlets at 250^0C with an average speed of 26m/sec. and through the secondary inlets at an average velocity of 20m/sec. A total mass of 0.028 kg/sec of biomass with a uniform size distribution by volume over the range 200-800μm was injected at the primary air inlets and a total mass of 0.0025kg/sec of char with a uniform distribution over the range 10-150μm was injected at the secondary inlets. BY providing a number of inlets at different locations the geometries of the reactors were changed and the optimal design of the gasifier for trouble free operation could be obtained with an excellent recirculation of char inside the reactor.

Fletcher et al [17] also developed a detailed Computational Fluid Dynamics (CFD) to simulate the flow and reaction in an entrained flow biomass gasifier. The CFX package as a powerful tool was used in gasifier design and analysis. The total height of the gasifier was 7.5 m and the diameter 0.8m. The biomass fuel and air were injected via the lower pair of tangential inlets and superheated steam via the upper pair. Biomass particulate is modelled via a Lagrangian approach as it enters the gasifier, releases its volatiles and finally undergoes

gasification. Transport equations are solved for the concentration of CH_4, H_2, CO, CO_2, H_2O are modelled. The model provided detailed information on residence time, gas composition and temperature at the outlet and allowed different scenarios to be examined in an efficient manner.

Joseph et al [18] reported developed of a pilot scale entrained flow biomass gasifier of 400 kW capacities. As per their design, the gasifier consists of two chambers separated by a baffle. The fuel is blown through a high temperature fan tangentially into the base of the reactor using the medium of pre heated air and steam. Pyrolysis and oxidation occurs in this chamber. The char and the oxidized gases are then swirled into the next chamber where reaction takes place to form more CO, H_2 and CH_4. Not all the char is reacted in the second chamber. The unreacted char is subsequently captured by a hot cyclone. The separated char is re-injected into the gasifier above the baffle using a steam injector. The fuel feed rate of the gasifier was 75-100 kg/hr, while the Equivalent Ratio was around 0.25. The authors further reported that based on the pilot scale results, a 3 MW entrained flow biomass gasifier was designed, where the gasifier was 0.8 m in diameter and 9 m tall.

3.0. ANALYSIS OF BIOMASS GASIFICATION

The chemistry of biomass gasification is very similar to that of coal gasification in the sense that thermal decomposition of both solids occurs to yield a mixture of essentially the same gases. The thermodynamic equilibrium concentrations of special gases in the mixture depend on the abundance of carbon, hydrogen and oxygen, the temperature, and the pressure. Biomass is gasified at lower temperatures than coal because its main constituents, the high oxygen cellulosic and hemi cellulosic have higher reactivity than the oxygen-deficient, carbonaceous materials in coal. The additions of co-reactants to the biomass system, such as oxygen and steam, can results in large changes in reaction rates, product gas compositions and yields, and selectivities as in coal conversion.

Biomass feed stocks contain a high proportion of volatile material, 70 to 90% for wood for wood compared 30 to 45% for typical coals. A relatively large fraction of most biomass feed stocks can be volatilized rapidly at low to moderate temperatures, and the organic volatiles can be rapidly converted to gaseous products. The chars formed on pyrolytic gasification of most biomass feed stocks have high reactivity and gasify rapidly.

Undesirable emissions and by-products from the thermal gasification of biomass can include particulates, alkali and heavy metals, oils, tars and aqueous condensates. Various hot-gas cleanup methods were developed that will permit biomass gasification to supply suitable fuel gas for advanced power cycles.

The biomass sample used for this study is coir dust collected from M/s Konark Coir Industries, Sakhigopal, Puri. The PSD was found to be d_{50} = 0.6mm. The moisture content (wet basis) was measured in a moisture analyzer and recorded as 10.23 %. All the calculations were based on the biomass sample with 10% MC.

3.1. Composition of Biomass

To determine the behaviour of chemical reactions of biomass species, the composition of biomass needs to be known. The table -1 below gives the different components of various biomass species.

Table-1: Composition of some biomass species:

Species	Total Ash	Solvent soluble	Water Soluble	Lignin	Hemi-celluloses	Cellulose
Soft wood	0.4	2.0	-	27.8	24.0	41.0
Hard wood	0.3	3.1	-	19.5	35.0	39.0
Wheat Straw	6.6	3.7	7.4	16.7	28.2	39.9
Rice Straw	16.1	4.6	13.3	11.9	24.5	30.2
Bagasse	2.3	8.4	10.0	18.5	29.0	33.6

(Source: Biomass Conversion processes for Energy and Fuels by Samir S. sofer and Oskar R. Zaborsky , Page – 106)

The percentage composition coconut husk (Dry weight basis) was found out to be 29.23% Lignin, 23.97% Cellulose, 18.50% hemi cellulose and remaining portions are soluble in water and solvent (Source: Coir, it's Extraction Properties and Uses, CSIR New Delhi, Page-18)

As cellulose and hemicelluloses come from the same monomer i.e. glucose and we, here treat both of them as cellulose. Though chemical composition of coir dust is different from that of rice husk and wood, to arrive at a compromise so that the design parameters of the gasifier will accommodate all the species of biomass, we take the composition of biomass (coir dust) as 80% cellulose to 20% lignin.

For, the design we need to arrive at an empirical formula for the biomass, which is, as earlier shown, a mixture of Cellulose ($-C_6H_{10}O_5-$) and Lignin ($-C_9H_{10}O_2-$). Cellulose is a co polymer of D-Glucose while Lignin is a giant polymer of Phenyl-propane. As at high

temperature both the components will completely gasify to give the same kind of products we can assume, both the components to react in a similar fashion.

Estimation of Empirical Formula for the Biomass:

For, 80% Cellulose (-$C_6H_{10}O_5$-) and 20% Lignin (-$C_9H_{10}O_2$-), the total number of Carbon, Hydrogen and Oxygen atoms will be as follows:

Number of Carbon atom = 0.8*6 +0.2*9 = 6.6

Number of Hydrogen atom = 0.8*10 + 0.2*10 = 10

Number of Oxygen atom = 0.8*5 +0.2*2 = 4.4

Empirical formula of mixture of Cellulose and Lignin i.e. formula for Biomass is $\mathbf{C_{6.6}H_{10}O_{4.4}}$

Molecular weight of the Empirical Formula is

=12*6.6 + 1*10+16*4.4

=159.6

3.2 Aspect of Biomass Gasification at Higher Temperature

The composition of gas at different temperatures as given in literature is given in table-2 as follows

Table -2 : Composition of gas at different temperatures

Composition of the gas	Temperature, ^{0}C			
	500	**600**	**800**	**900**
CO_2, mol%	44.8	31.8	20.6	18.3
CO, mol%	33.5	30.5	34.1	35.3
CH_4, mol%	12.4	15.9	13.7	10.5
H_2, mol%	5.56	16.6	28.6	32.5
Ethane, mo1%	3.03	3.06	0.077	1.07
Ethylene, mol%	0.45	2.18	2.24	2.43
High Heat Value (HHV), MJ/m^3	12.3	15.8	15.4	15.1
Gas yield, m^3/kg	0.114	0.166	0.216	0.202

The above table on gas composition at different temperatures from Pyrolysis of biomass reveals that at higher temperature, the concentration of CO and H_2 increases. As among the components of the product gas only CO and H_2 has higher calorific value, it is desired to have more of these two gases in the product. Also Entrained Flow gasification allows the reactions to take place at higher temperature i.e. at around 1000 ^{0}C and more, a composition of the product gas is expected to be better in this process.

3.3 Thermo Chemical Reactions

When biomass is fed t the reactor, a number of reactions occur simultaneously. The moisture in the biomass gets devolatalized as it enters the reactor. After devolatalization, the fuel particles tend to gasify in the reactor in a sub-stoichiometric condition i.e. amount of air supplied to the reactor will be less than that required for complete combustion. In the gasification or partial oxidation reaction, the product gas formed mainly consists of carbon monoxide and hydrogen along with carbon dioxide, ethane etc. As carbon monoxide and carbon dioxide are low calorific gases, Shift action is allowed to occur in the reactor so that carbon monoxide can be converted to hydrogen by reacting it with steam.

The process of biomass gasification is broadly divided into three steps.

Pyrolysis

Partial Oxidation

Steam Reforming

The three steps can be accomplished in a single reactor or may be segregated to individual separate steps, though pyrolysis is an inherent part of this gasification process.

(a) Pyrolysis:

It consists of the slow, irreversible, thermal degradation of the organic components in biomass, in the absence of oxygen, Controlling the pyrolysis temperature, heating rate and residence time, can control the components of pyrolysis. The primary products of biomass pyrolysis under conventional pyrolysis conditions are gas, oil, char, and water. As the reaction temperature increases, gas yield increases. Water is released as the biomass dries in the gasifier and is also a product of the chemical reactions that occur, even with bone-dry biomass. This water is expected to participate in the process along with any added feed water or steam.

b) Partial Oxidation:

This is the process where reaction is intended to occur at sub-stoichiometric condition i.e. amount of oxygen is less than that is required for complete combustion. Here partially oxidized products are formed. Both air and oxygen have been utilized for this type of system. When the oxygen is supplied by air, low energy gases are formed that contain higher concentrations of hydrogen, carbon monoxide, and carbon dioxide than medium energy gases. When pure oxygen is used, gases with higher energy values can be obtained Completed.

The heating value of the gas is dependent upon the amount of air supplied and is essentially independent of the type of feed organics. The greater the deficiency of air needed to achieve complete combustion, the higher the fuel value of the product gas.

The char particles formed in the Pyrolysis Reaction are expected to completely gasify because of the high temperature attained in the partial oxidation reaction zone. Also removal of tar is another benefit obtained due to high temperature partial oxidation. Absence of char

and tar is one of the basic requirements in the product gas. Produce gas free from char and tar can be easily used in various pumps and engines.

c) Steam Reforming:

It is the process of conversion of hydrocarbon gases and vaporized organic compounds to hydrogen containing gases such as synthesis gas, a mixture of Carbon monoxide and hydrogen.

Under idealized conditions, the primary products of biomass gasification by pyrolysis, partial oxidation, or steam reforming are essentially the same. The carbon oxides and hydrogen are formed. Methane and light hydrocarbon gases are also formed under certain conditions. The energy content of the product gas from biomass gasification can be varied. Low energy gases formed when there is direct contact of biomass feedstock and air. This is due to dilution of product gas with nitrogen from air during the gasification process. Where as high energy content product gas can be obtained when nitrogen is not allowed to mix with the producer gas, by allowing only oxygen to react with the biomass.

3.4 Stoichiometry

As the exact governing reaction is not easy to determine, various hypothetical reactions are considered, taking Empirical formula as a representative composition for the biomass. Depending upon the composition of the product gas a number of reactions were formulated.

The reactions below sited are obtained from Donald L. Klass, Page-275. The empirical formula is taken here as a representative unit, in spite of cellulose.

Partial Oxidation:

1. $C_{6.6}H_{10}O_{4.4} + 1.1\ O_2 = 6.6\ CO + 5\ H_2$ (1)

2. $C_{6.6}H_{10}O_{4.4} + 2.2\ O_2 = 4.4\ CO + 2.2\ CO_2 + 5\ H_2$ (2)

3. $C_{6.6}H_{10}O_{4.4} + 1.6\ O_2 = 6.6\ CO + 4H_2 + H_2O$ (3)

4. $C_{6.6}H_{10}O_{4.4} + 1.8\ O_2 = 4\ CO + 2\ CO_2 + 5\ H_2 + 0.6\ C_S$ (4)

Reactions to be dealt in the design are:

(1) Thermal Gasification:

$C_{6.6}H_{10}O_{4.4} + 1.8\ O_2 = 4\ CO + 2\ CO_2 + 5\ H_2 + 0.6\ C_S$

(2) Steam Gasification:

$C_{6.6}H_{10}O_{4.4} + 2.2\ H_2O = 6.6\ CO + 7.2\ H_2$

(3) Char Oxidation (water gas reaction)

$Cs + H_2O = CO + H_2$

(4) Shift Reaction:

$CO + H_2O = CO_2 + H_2$

3.5. Estimation of adiabatic Flame Temperature by Analytical Method

The flame temperature can be calculated by following two methods.

I) By using the established thermodynamic data available in different sources.

II) By assuming constant specific heat of different components and estimating enthalpies of the components by $C_p\Delta T$ and heat of formation at thermodynamic standard condition.

I) As it is insulated the total input energy = total output energy.

By energy balance:

$$\Sigma n \{E_1 + P_1v_1/J + U_1^2 / 2\rho J + Z_1\} + \Delta H = \Sigma m\{E_2 + P_2v_2/J + U_2^2/2\rho J + Z_2\} + W$$

Or $$H_1 + \Delta H = H_2 \text{ (Neglecting U and Z changes)}$$

Where:

H_1 and H_2 = Enthalpy of reactants and products.

ΔH = Heat Addition = 0

W = Shaft Work = 0

Or $H_1 = H_2$

So by using enthalpy data (H_1, H_2) from literature the adiabatic flame temperature can be estimated by graphical interpolation.

II) This equation can be modified as follows

$$\Sigma n \{E_1 + P_1v_1/J\} = \Sigma m\{E_2 + P_2v_2/J\}$$

$$\Sigma nH_1 = \Sigma mH_2$$

Or $$\Sigma n (H_{f1}+C\rho_1\Delta T_1) = \Sigma m(H_{f2}+C\rho_2\Delta T_2)$$

$\Sigma n\ H_{f2} - \Sigma m\ H_{f1}$ = Heat of combustion at standard state.

Where, $C\rho_1$ & $C\rho_2$ are the heat capacities of reactant and product gases.

H_{f1} & H_{f2} are standard heat of formation at standard state.

ΔT_1 & ΔT_2 are inlet and outlet temperature difference respectively of reactant and product with respect to the standard state of 25^0C.

So, from the knowledge of either heat of combustion of the whole reaction or heat of formation of reactants and products along with specific heat data, the adiabatic flame temperature can be estimated. In the present work second method was used. Here average heat capacities values were taken into consideration.

3.6. Equivalent Ratio

Equivalent Ratio is the ratio of the amount of air used for partial combustion to the Stoichiometric amount of air required for full combustion. Complete combustion is the process of reacting all the biomass with surplus amount of oxygen so that all the reactant gets converted into carbon dioxide and water.

For partial combustion and complete combustion of biomass the following equations have been considered

Partial Combustion : $C_{6.6}H_{10}O_{4.4} + 1.8\ O_2 = 4\ CO + 2\ CO_2 + 5\ H_2 + 0.6\ C_S$

Complete Combustion: $C_{6.6}H_{10}O_{4.4} + 6.9\ O_2 = 6.6\ CO_2 + 5\ H_2\ O$

For complete combustion 6.9 moles of oxygen is required, whereas for gasification 1.8 moles of oxygen is required. Therefore the equivalence ratio is given by Equivalent ratio

$$= 1.8/6.9 = 0.26$$

3.7. Shift Reaction

$$\mathbf{CO + H_2O = CO_2 + H_2}$$

The shift reaction is to produce Hydrogen from producer gas, this is an endothermic reaction so the sensible of the partial oxidation can be preserved as chemical energy in the form of Hydrogen by introducing steam. The total heat value of room temperature increases through the calorific value of the product gas decreases. Moreover the producer gases with high Hydrogen content become a chemical raw material for alcohol and other chemicals. That's why the producer gas having H_2: CO more than 2.5 is called Syn gas. The present work envisages producing the Hydrogen rich gas, which may have versatile end uses.

4. SOLUTIONS & EXPERIMENTAL SETUP

4.1 Solutions

The biomass gasifier has been conceived as an entrained flow type where a number of reactions occur simultaneously as the biomass is fed into the reactor. The chemical formula of the biomass has been derived as $C_{6.6}H_{10}O_{4.4}$ is explained in section 3.3. The various partial gasification reactions that will occur in the reactor as are also presented in section 3.4. Equation (4) in section 3.4 has been considered to be occurring in the reaction, which is given by

$$C_{6.6}H_{10}O_{4.4} + 1.8\ O_2 = 4\ CO + 2\ CO_2 + 5\ H_2 + 0.6\ C_S$$

4.1.1. Material balance and Estimation of Temperature in the reactor during thermal gasification and steam gasification

Due to inherent moisture content of the biomass, the predominant reactions in the gasifier are both partial oxidation (Thermal gasification) and steam gasification. The temperature attended in the reactor at different equivalence ratios have been computed and the estimated species concentration of the product gases are given in Annexure-I. With a moisture content of 10 % in the biomass, the calculated value of temperature achieved in the gasifier was in the range of 900 ^{0}C at *ER=0.26*. It was also indicated that with increase in equivalence ratio, the temperature inside the gasifier reactor increased. Therefore the species concentration of producer gas after partial oxidation and steam gasification were computed for coir dust and is given in Table -3.

Table-3: Species concentration of product gases after partial oxidation and steam gasification

Species	Moles	Mole % age by volume	Wt%
CO	4.26	24.53	29.46
CO_2	1.80	10.36	18.47
H_2	5.22	30.05	2.70
N_2	6.09	35.06	49.37

4.1.2. Material balance and estimation of temperature in the reactor after water gas reaction.

After the thermal gasification and steam gasification in the reactor, it is assumed that the char formed will be treated with steam to enable the water gas reaction to occur in the upper part of the reactor. The prevailing reaction in the upper chamber has been assumed as follows

$$\mathbf{4.26\ CO + 1.8\ CO_2 + 5.22\ H_2 + (0.54\ C_s + 0.54\ H_2O)}$$

$$\mathbf{= (0.54\ CO + 0.54\ H_2) + 4.26\ CO + 1.8\ CO_2 + 5.22\ H_2}$$

$$\mathbf{= 4.8\ CO + 1.8\ CO_2 + 5.76\ H_2}$$

By balancing the enthalpies the temperature in the upper chamber of the rector was computed to be 775 ^{0}C and the estimated species concentration of the product gases is given in table-4. The calculations for temperature and species concentration at different equivalence ratio are given in Annexure – II.

Table-4: Species concentration of product gases after water gas reaction

Species	Moles	Mole % age by volume	Wt%
CO	4.80	26.02	32.00
CO_2	1.80	9.76	17.81
H_2	5.76	31.22	2.60
N_2	6.09	33.01	47.59

4.1.3. Material balance and estimation of temperature in the reactor after shift reaction.

Steam is introduced in the upper chamber of the gasification unit to promote overall process via the endothermic steam carbon reactions to form hydrogen and carbon monoxide. The equation that prevails after shift reaction is

$$\mathbf{4.8\ CO + 1.8\ CO_2 + 5.76\ H_2 + 2.4\ H_2O = 4.53\ CO + 2.07 CO_2 + 6.03\ H_2 + 2.13\ H_2O}$$

By balancing the enthalpies the temperature in the upper chamber of the reactor after shift reaction was computed to be around 907 ^{0}C and the estimated species concentration of the product gases is given in table–5. The calculations for temperature and species concentration are given Annexure-III. It was also calculated that 11 kg/hr steam may be injected to enhance hydrogen concentration in the product gas.

Table-5: Concentration of product gas after shift reaction with 2.4 moles of steam

Species	Moles	Mole % age by volume
CO	4.53	21.73
CO_2	2.07	9.93
H_2	6.03	28.92
H_2O	2.13	10.22
N_2	6.09	29.21

It appears that the effect of steam is not very much predominant and it is observed that though H_2 amount is increasing, un-reacted steam is left so much so, the increase in volume percentage of H_2 is not significant. The original biomass feedstock may be pre-dried to less than 10% moisture content to achieve more hydrogen by introducing superheated steam at the upper chamber of gasifier.

4.2. Fuel Feed Rate

For the existing 30 KWe entrained flow gasifier the fuel feed rate was considered as 40 Kg/hr. From literature survey, it is found that for 1KW gasifier, the biomass feed rate would be approximately 1.3-1.4 kg/hr. Hence for a 30 KWe capacity gasifier unit, the biomass fuel feed rate would be = 30 x 1.3 = 39 kg.

Therefore, biomass fuel feed rate was taken as 40 kg/hr.

4.3 Calorific value of product gas

The heat content of CO and H_2 are considered for computing the calorific value of producer gas

Heat content of CO

= 4.53 x 281

=1273 KJ

Heat content of H_2

= 6.03 x 241

=1453 KJ

Total energy available per mole of fuel = 1273 KJ + 1453 KJ

= 2726 KJ

Total nos. of moles in the product gas is = 20.85 moles

Volume occupied by these gases are = 20.85 x 22.4 liters

= 467 liters

=0.467 m^3

So, 0.467 m^3 of product gas has 2726 KJ of energy

Therefore, the Calorific Value of the gas = 2726 KJ/0.467 m^3 = 5.84 MJ/Nm^3

Similarly, the calorific value of product gas is computed at different equivalence ratios and shown in table-6.

Table-6: Calorific Value of product gas after shift reaction.

Equivalence Ratio	Calorific Value (MJ/Nm^3)
0.20	6.52
0.26	5.84
0.30	5.45
0.35	4.86
0.40	3.97

4.4 Volume of the Product gas

Our desired equation is

$$\mathbf{4.8CO + 1.8CO_2 + 5.76H_2 + (1.62 * 3.76)N_2 + 2.4H_2O}$$

$$\rightarrow \mathbf{4.53CO + 2.07CO_2 + 6.03H_2 + (1.62 * 3.76)N_2 + 2.13H_2O}$$

Assuming that the rate of formation of products is constant and the fuel is burnt immediately. 1 mole of substance weighs 22.4 liters at NTP, so as 907^0C, volume of 1 mole of a gas is given by

$V_2 = V_1T_2/T_1$

= 22.4 x (907 + 273)/(273+25)

=88.7 liters

<u>Volume of Products</u>

Composition of the product gas is given by the following chemical equation

4.53 CO + 2.07 CO_2 + 6.03 H_2 + 2.13 H_2O + 6.09 N_2

1 mole of biomass produces

CO : 4.53 moles

CO_2 : 2.07 moles

H_2 : 6.03 moles

H_2O : 2.13 moles

N_2 : 6.09 moles

Total : 20.85 moles

So the total no. of moles of product gas = 20.85 moles. At 907^0C the volume occupied by the product gas.

=20.85 x 88.7 litres

=1849.4 litres

So total volume of the products

=40x 6.27x 1849.4

=463.829 m^3/hr

So volume of the product gases = 0.129 m^3/sec

Similarly, the rate of gas generation of product gas is computed at different equivalence ratios and shown in table-7.

Table-7: Rate of gas generation.

Equivalence Ratio	Rate of gas generation (m^3/sec)
0.20	0.089
0.26	0.129
0.30	0.137
0.35	0.169
0.40	0.187

4.5. Fuel & Air requirement

The governing equation for gasification is

$$C_{6.6}H_{10}O_{4.4} + 1.62(O_2 + 3.76N_2) + 0.22H_2O \rightarrow 4.26CO + 1.8CO_2 + 5.22H_2 + 0.54C_S + (3.76 * 1.62)N_2$$

1 mole of fuel requires 1.62 moles of oxygen

1.62 moles of oxygen is present in 1.62 / 0.23 = 7.04 moles of air

1 mole of air weights 28.9 grams

7.04 moles of air will weight = 28.9 x 7.04= 203.456 gms.

So 1 mole of cellulose require 203 grams of air

1 kg (6.27 moles) of fuel will require = 6.27 x 203 = 1272.81 grams of air

So the air requirement for feeding powdery biomass at a rate of 40 hg/hour will be = 40 x 1273 = 50920 grams = 50.92 kg.

Similarly, the air requirement for gasification was calculated for different equivalence ratios.

Table-8: Air requirement for gasification at different equivalence ratio.

Equivalence Ratio	Air requirement for gasification (Kg/hr)
0.20	39
0.26	51
0.30	59
0.35	68
0.40	78

4.6 Instrumentation

The gasifier unit was provided with a number of instruments for measuring the primary air mass flow rate and secondary preheated air flow rate to the gasifier, temperatures at various points such as in the gasifier inlet & outlet, gasifier chamber etc.

An air flow meter procured from Krone-Marshal with a flow rate range of 14.1 kg/hr-141 kg/hr was fitted on the compressed air pipe line near the biomass air jet feeder to measure the total air flow into the gasifier. 4 nos. of Thermocouples (PT-RT/RH 13%, 0.45 dia shealth, KER-710, Recrystalized Alumina of 12 mm OD and 8 mm ID, 600 mm long with standard flanges) with digital temperature indicators (Model: MC 2838-301-000, Temperature range: 0-1600^{0}C, PID function) were provided at different elevation of the gasifier to measure the temperatures.

4.7 Gasifier Plant Layout and Accessories

The general arrangement of the gasification system is given in fig.1. The gasification system consists of the following sub units

i. Gasifier
ii. Biomass feeding system
iii. Oil Ignition System

iv. Steam generator

v. Air Compressors and blower

vi. Producer gas Sampling system

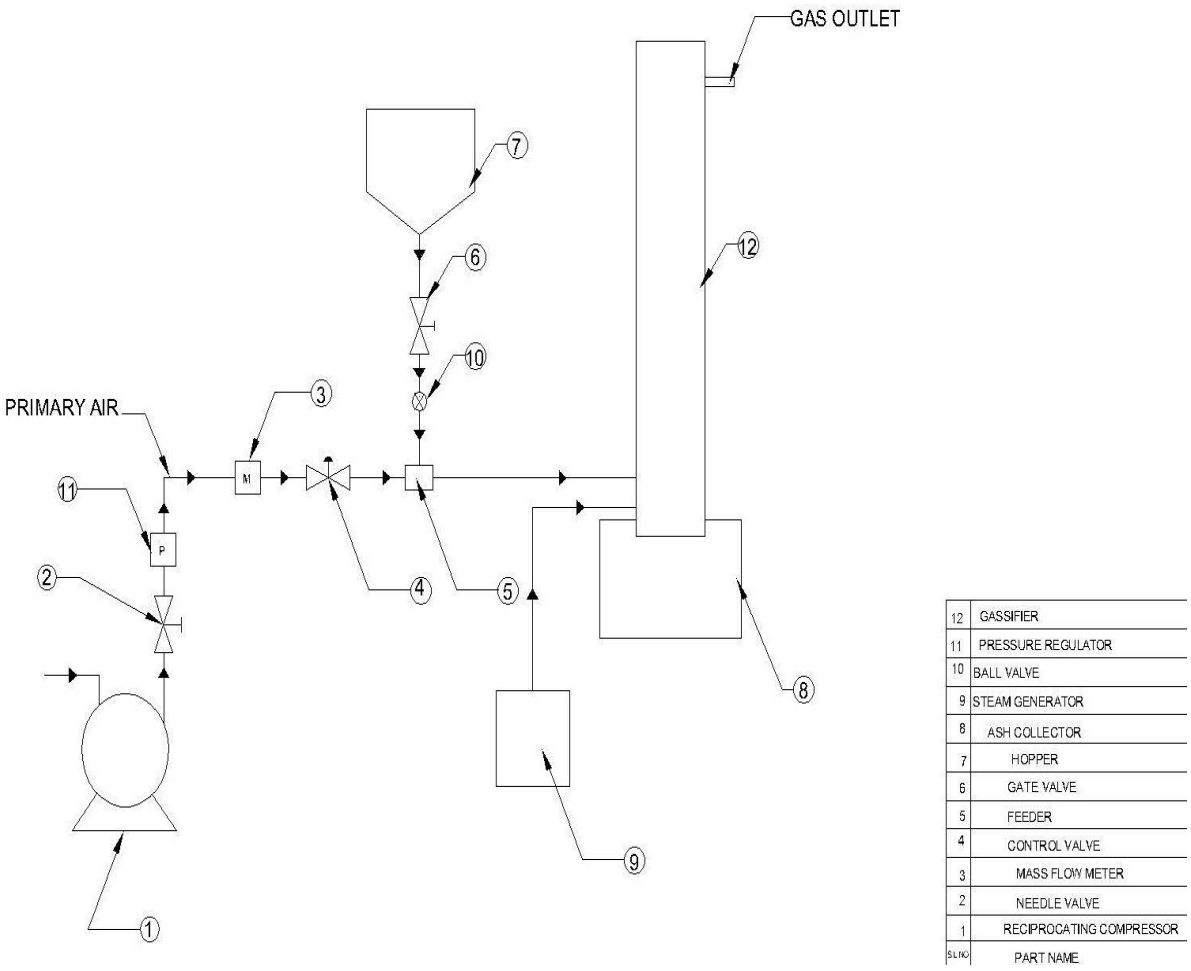

Fig. 1 Layout of gasifier plant

4.8 Gasifier

The gasifier consists of an outer mild steel shell of 5 mm thick and the inside of which is lined with high alumina (70%) refractory castables of 100 mm thick. It is expected that these refractory materials with typical qualification temperatures of 1000 – 1500^{0}C. will serve the purpose for adequate insulation with minimum heat loss. Some of the views of the 30 KW entrained flow biomass gasification system is shown in figure.

The various dimensions of the gasifier are given below.

Dimensions

Length : 2.5 m

Diameter: 0.25 m

Internal lining: 0.11 m of high Alumina refractory

4.9 Biomass feeding system

Handling of wide variety of powdery biomass and injecting to the gasifier has been a challenging job worldwide. Therefore a suitable feeding system capable of handling powdery biomass fuels of widely differing densities – from 70 kg/m^3 for sugar cane to 350 kg/m^3 for the rice husks or saw dusts should be designed. The feeding system in the present design consists of a fuel fuel bin of 0.2 m^3 temporary storage capacity, capacity, air jet biomass feeder with injection nozzle fitted at the outlet of biomass fuel bin. The powdery biomass is sucked into the intermixing chamber by the suction created at the nozzle of the air jet biomass feeder and is blown into the gasifier along with the transporting primary air.

The design air velocity and air flow rate in the air jet biomass feeder have been evaluated as 25 m/sec and 22 kg/hr. respectively. The design velocity of the higher order has been arrived in order to maintain a strong swirl and high intensity of vortex flow in the gasifier. One novelty of the design in air jet biomass feeder is that back flow of biomass can be prevented by supplying additional amount of air jets through an extra port provided near the aeration chamber in the feeder. To avoid sticking of biomass near the walls of the fuel bin and to ensure continuous supply of biomass in the biomass fuel pipe a vibrating mechanism has been incorporated with eccentric attachment as shown in fig.2.

Fig. 2 Biomass feeding system

4.10 Oil Ignition System

Before injecting biomass into the gasifier reactor, it is required to pre-heat the gasifier to a temperature of 500-600^{0}C. to sustain flame and gasify the biomass instantaneously for continuous operation. It will also enhance the gasification reaction rate and for the purpose a diesel oil firing system was proposed. Accordingly an oil ignition system comprising of a oil tank of 0.3 m^3 capacity, fuel-air injection with atomizing nozzles, fuel pump system with 2 oil pumps (one for operation and other standby, with 1 hp, 3 phase induction motors, flow rate: 30 litres/hr, each), a 3 hp blower with air regulating valve to supply air for atomization and combustion etc. was provided. The set up is shown in fig. 3.

Fig. 3 Oil Ignition System

4.11 Steam Generator

From the material balance and estimation of temperature in the reactor after shift reaction, as discussed previously in section 4.1.3, it was seen that nearly 11 kg/hr. steam should be injected into the gasifier to enhance hydrogen production in the producer gas.

The details of steam requirement calculation are given in Annexure-III. Accordingly the steam generator with output capacity of 20 kg/hr, at a pressure of 3 kg/cm^3 was used shown in fig. 4.

Fig. 4 Steam Generator

4.12 Air Compressors and Blowers

The primary air for transport of powdery biomass and secondary air requirements were done by means of compressed air from 2 nos. of existing compressors. (One in operation and other as standby).

Specifications of Compressors

- Make: Kirloskar
- Normal Pressure: 12 kg/cm^2
- Maximum Pressure: 16 kg/cm^2
- FAD/DIS : 500 liters/minute
- Tank Capacity: 500 litres
- Motor HP: 5
- Unit RPM: 750

The atomization air required for oil ignition system during preheating of the system, combustion air requirement and dilution air requirement were done by means of 3 nos of

blowers. These blowers of different capacities (3 hp, 5 hp and 10 hp) available in the lab were used in the gasification unit.

4.13 Producer gas sampling system

In order to collect the sample gas from the gasifier for determining the gas composition of the producer gas by ORSAT apparatus shown in fig.6, the existing gas sampler was incorporated in the gasifier.

The gas sampler shown in fig.5 consists of a 20 liter capacity SS tank air vent valve and a sample gas outlet valve and was placed over a MS angle stand near the gas sampling point of the gasifier. The gas port of the gasifier at sampling point is connected to the SS tank near the base by means of SS pipe and a SS ball valve. This pipe extends into the tank and is made perforated inside the tank. Another piping outlet with a valve is provided near the base of the tank and is connected to a 20 litre capacity PVC water tank through flexible rubber pipe. A wire rope and pulley arrangement has been provided for up and down adjustment of the water tank, so that gas is collected in the gas sample tank by water displacement method. As the gas is passed from bottom of the gas sample tank through the water bed, it gets cooled and is collected at the top of the tank and then the sample gas is taken to ORSAT apparatus though a rubber piping arrangement.

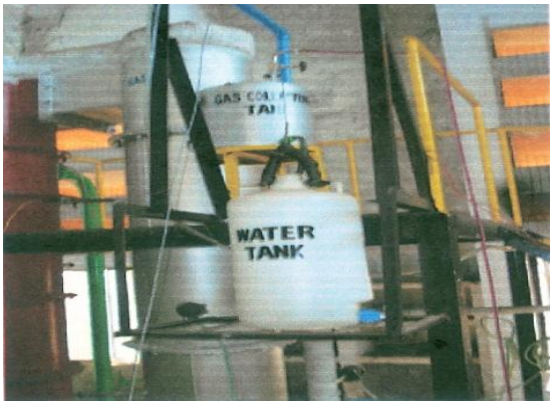

Fig. 5. Producer gas sampling system

Fig. 6. ORSAT apparatus

4.14. Experimental studies with Gasifier unit

To start with, the gasifier unit was preheated to a temperature of 500-600^{0}C using the oil ignition system. The gasifier achieved the pre-heat temperature in about 35 minute and then a number of tests with the coir dusts were conducted. 3 nos. of Thermocouples (PT-RT) with standard flanges were inserted at 3 different elevations of the gasifier (Level-1: just above the oil ignition point and near the biomass feeding point, Level-2; well below the steam injection point, Level-3: above the steam injection point) with digital temperature indicators (Temperature range: 0-1600^{0}C). Each run of the gasifier with the biomass was conducted for duration of about 2 hours.

For each experimental run of the gasifier, gas sampling was carried out four using the gas sampler connected at the gas sampling point of the gasifier and the gas composition of the producer gas was determined using ORSAT apparatus.

5. RESULTS & DISCUSSION

5.1 Effect of equivalence ratio on Adiabatic flame temperature.

Gas outlet temperature was measured through the high temperature thermocouple with indicator. The results are shown in table-9. Fig.7 shows the theoretical and experimental adiabatic flame temperature with change in equivalence ratio. The difference in theoretical and experimental adiabatic flame temperature is difficult to interpret as the equations are solved with some assumptions. Besides the instruments used & their accuracy, other losses also play a major role in the deviation. Trend shows gradual increase of error percentage (2-24 %) with higher temperature. From the figure it was observed that the adiabatic flame temperature increased with increase in equivalence ratio. This attributes to availability of higher oxygen with increase in equivalence ratio.

Table-9: Comparison of Adiabatic Flame Temperature.

Equivalence ratio	Analytical adiabatic flame temperature (^{0}C)	Experimental adiabatic flame temperature (^{0}C)
0.20	586	575
0.26	907	790
0.30	965	890
0.35	1175	975
0.40	1380	1050

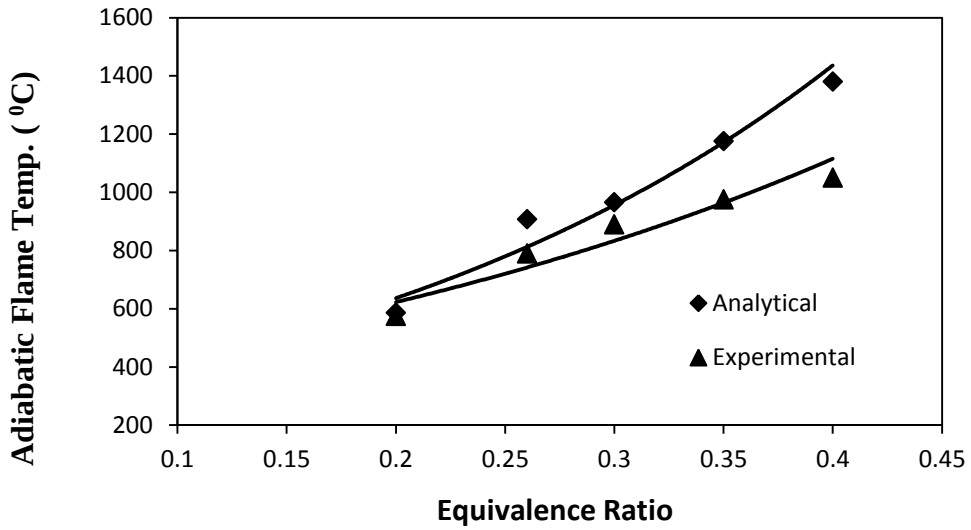

Fig.7. Effect of equivalence ratio on adiabatic flame temperature

5.2 Effect of equivalence ratio on Gas Composition

It is observed from fig.8 that CO & H_2 decrease with increase in equivalence ratio. Whereas, CO_2 increases with increase in equivalence ratio. With increase in equivalence ratio temperature inside the reactor increases to 850-1000 ^{0}C, sufficient char may not be available for the endothermic reaction of CO_2 with char. Therefore, the concentration of CO decreases in the product gas. Similarly at very high equivalence ratios, the hydrogen production decreases due to exothermic and reversible behaviour of water gas shift reactions. This also attributes to decrease in calorific values of the product gas at higher equivalence ratios.

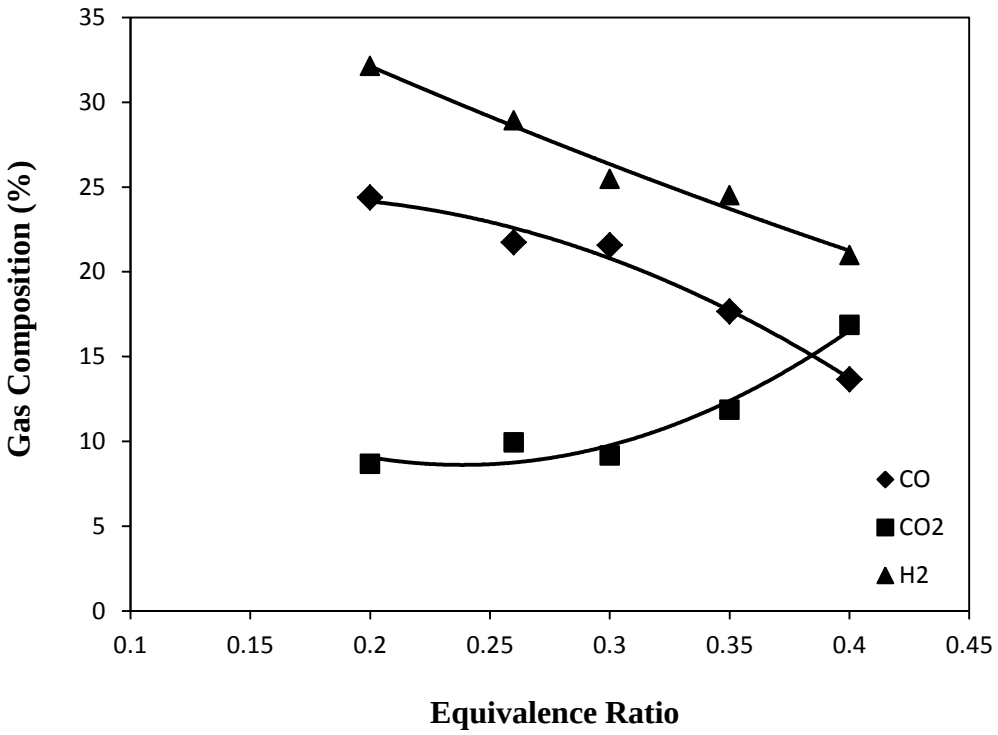

Fig.8. Effect of equivalence ratio on product gas composition

Table-10: Comparison of product gas component.

Species	ER=0.2		ER=0.26		ER=0.30		ER=0.35		ER=0.40	
	A	E	A	E	A	E	A	E	A	E
CO (Vol %)	24.38	20.00	21.73	18.00	21.57	18.00	17.65	14.00	13.65	10.00
CO_2 (Vol %)	8.67	15.00	9.93	16.00	9.17	16.00	11.87	19.00	16.87	22.00
H_2 (Vol %)	32.13	30.00	28.92	26.00	25.47	22.00	24.51	21.00	20.99	16.00

5.3 Effect of equivalence ratio on Calorific value & Rate of gas production

It is observed from fig.9 that the calorific value of product gas decreases with increase in equivalence ratio. This is due to the reduction of CO & H_2 at higher equivalence ratios explained in section 5.2. As the heat content of CO & H_2 in the product gas plays a major role in total calorific value of the gas. Availability of less CO & H_2 moles leads to low calorific value of the gas at higher equivalence ratios.

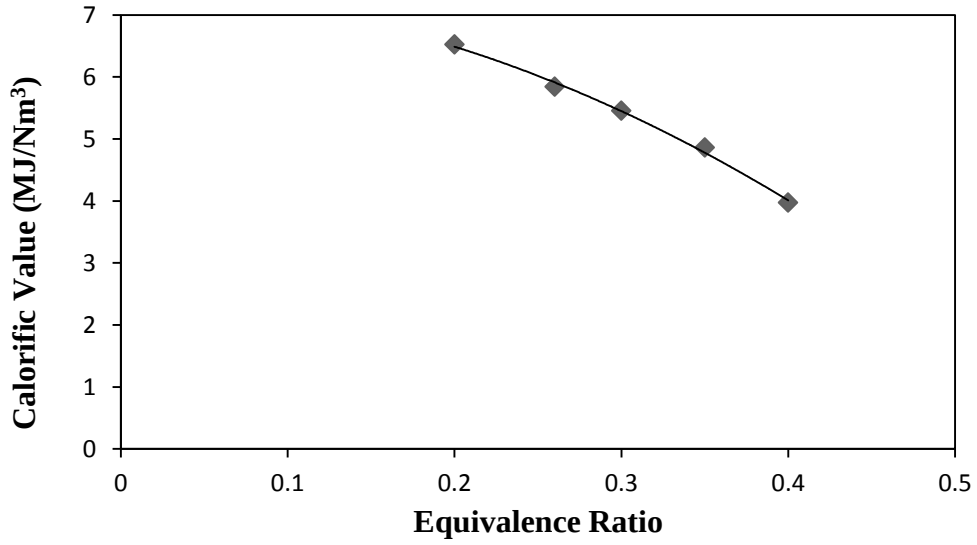

Fig.9. Effect of equivalence ratio on calorific value

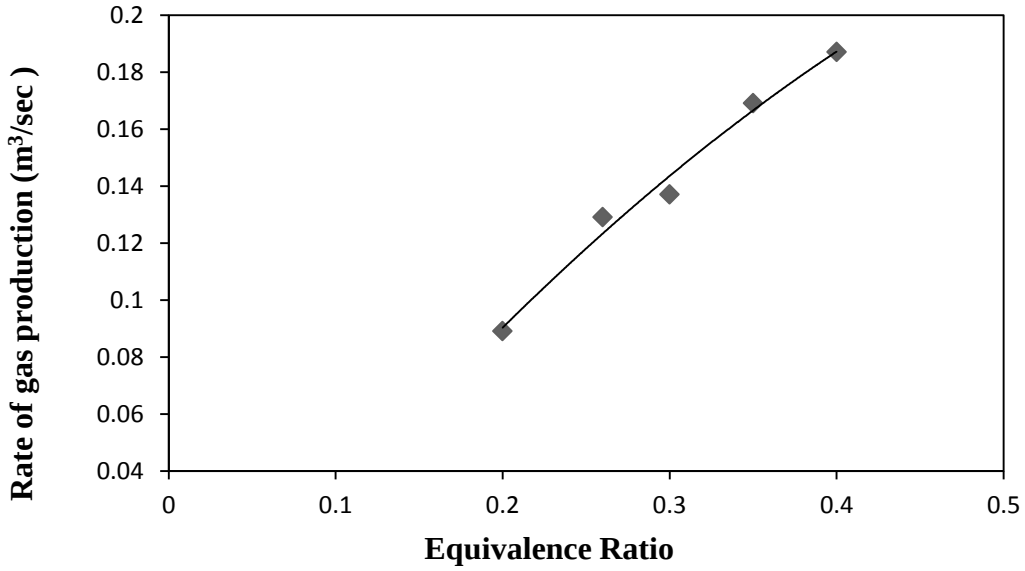

Fig.10. Effect of equivalence ratio on rate of gas production

From fig.10 it is observed that the rate of gas production increased with increase in equivalence ratio. This is due to the availability of more air to react with the feedstock fuel. Mostly production of CO_2 & N_2 increases at higher equivalence ratios. This results in high rate of gas production.

6. CONCLUSION

The conversion of biomass wastes from low to medium calorific gas is now a practical and economic proposition, which can be carried out using an entrained flow gasifier. The experimental studies conducted on the Gasifier developed at CSIR-IMMT, Bhubaneswar indicates that it is possible to gasify efficiently the coir dusts with wide range of particle sizes in entrained mode. The injection of steam into the gasifier enhances the energy value of the producer gas in producing more hydrogen. The biomass along with primary air enters the gasifier in a secant manner to attain high intensity swirl inside the reactor and a temperature of 790-1050 ^{0}C is achieved (ER: 0.26-0.40) to take the advantages of the kinetics of the process and produces low tar and char formation. The volume percentage of CO and H_2 in the producer gas while gasifying with coir dust at equivalence ratio 0.26 was found to be 21.73% and 28.92% respectively. The concentration of CO & H_2 decreases with increase in equivalence ratios. By varying Equivalence Ratio (ER), temperature of the partial oxidation can be varied as per requirement. The computed calorific value of the producer gas from the analytical & experimental data was fond to be 5.84 MJ/Nm^3& 5.07 MJ/Nm^3 respectively.

The adoption of the technology will benefit many micro & medium industries such as ceramic and glass industries, metallurgical industries, food processing industries etc. for thermal applications. Also the coir dust can be effectively used in an organized manner ensuring extra profit to the rural coir board industries and it will create an eco friendly environment.

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Annexure-I

Material Balance and Estimation of Temperature in the reactor during thermal gasification and steam gasification

Moisture Content of biomass sample (coir dust) is 10 %. Therefore, 90% will be partial gasification type & 10% will be of steam gasification type.

(For Equivalence Ratio=0.2).

Partial Oxidation Reaction.

$$0.9 * [C_{6.6}H_{10}O_{4.4} + 1.38(O_2 + 3.76N_2) \rightarrow 3.98CO + 1.59CO_2 + 5H_2 + 1.03C_S + (1.38 * 3.76)N_2]$$

Steam Gasification Reaction.

$$0.1 * [C_{6.6}H_{10}O_{4.4} + 2.2H_2O \rightarrow 6.6CO + 7.2H_2]$$

Combined equation:

$$C_{6.6}H_{10}O_{4.4} + 1.242(O_2 + 3.76N_2) + 0.22H_2O \rightarrow 4.242CO + 1.431CO_2 + 5.22H_2 + 0.927C_S + (3.76 * 1.242)N_2$$

The equation for conservation of enthalpies of the reactants and products (constant pressure adiabatic process) can be written as:

$$\Delta H_{fR} + M_R \text{ x } C_{pR} \text{ x } \Delta T_R = \Delta H_{fP} + M_P \text{ x } C_{pP} \text{ x } \Delta T_P$$

Enthalpy of reactants:

= [(-720*1*1000/4.2) + (0.5*159.6*(240-25)) + (-0.22*286*1000/4.2) + (0.22*18*75*1) + (0.22*18*538) + (0.22*18*0.46*140) + (1.242*32*0.23*(240-25)) + (1.242*3.76*28*0.25*(240-25))]

=-158741.66 cal.

Enthalpy of Products:

= [(-4.242*111*1000/4.2) + (4.242*28*0.29*(T-25)) + (-1.431*394*1000/4.2) + (1.431*44*0.3*(T-25)) + (5.22*2*3.5*(T-25)) + (0.927*12*0.71/4.2*(T-25)) + (1.242*3.76*28*0.29*(T-25))]

= [-246351.43 + 125.23(T-25)] cal.

Upon solving T comes out to be 724.59 ≅ 725℃

(For Equivalence Ratio=0.26).

Partial Oxidation Reaction.

$$0.9 * [C_{6.6}H_{10}O_{4.4} + 1.8(O_2 + 3.76N_2) \rightarrow 4CO + 2CO_2 + 5H_2 + 0.6C_S + (1.8 * 3.76)N_2]$$

Steam Gasification Reaction.

$$0.1 * [C_{6.6}H_{10}O_{4.4} + 2.2H_2O \rightarrow 6.6CO + 7.2H_2]$$

Combined equation:

$$C_{6.6}H_{10}O_{4.4} + 1.62(O_2 + 3.76N_2) + 0.22H_2O \rightarrow 4.26CO + 1.8CO_2 + 5.22H_2 + 0.54C_S + (3.76 * 1.62)N_2$$

Enthalpy of reactants:

= [(-720*1*1000/4.2) + (0.5*159.6*(240-25)) + (-0.22*286*1000/4.2) + (0.22*18*75*1) + (0.22*18*538) + (0.22*18*0.46*140) + (1.62*32*0.23*(240-25)) + (1.62*3.76*28*0.25*(240-25))]

=-152994.25 cal.

Enthalpy of Products:

= [(-4.26*111*1000/4.2) + (4.26*28*0.29*(T-25)) + (-1.8*394*1000/4.2) + (1.8*44*0.3*(T-25)) + (5.22*2*3.63*(T-25)) + (0.54*12*0.71/4.2* (T-25)) + (1.62*3.76*28*0.29*(T-25))]

= [-281442.85 + 146.81(T-25)] cal.

Upon solving T comes out to be 899.62 ≅ 900 ^{0}C

(For Equivalence Ratio=0.3).

Partial Oxidation Reaction.

$$\mathbf{0.9 * [C_{6.6}H_{10}O_{4.4} + 1.8(O_2 + 3.76N_2) \rightarrow 4CO + 2CO_2 + 5H_2 + 0.6C_S + (3.76 * 1.8)N_2]}$$

Steam Gasification Reaction.

$$\mathbf{0.1 * [C_{6.6}H_{10}O_{4.4} + 2.2H_2O \rightarrow 6.6CO + 7H_2]}$$

Combined equation:

$$\mathbf{C_{6.6}H_{10}O_{4.4} + 1.863(O_2 + 3.76N_2) + 0.22H_2O \rightarrow 4.413CO + 1.9215CO_2 + 5.2H_2 + 0.1755C_S + (3.76 * 1.863)N_2}$$

Enthalpy of reactants:

= [(-720*1*1000/4.2) + (0.5*159.6*(240-25)) + (-0.22*286*1000/4.2) + (0.22*18*75*1) + (0.22*18*538) + (0.22*18*0.46*140) + (1.863*32*0.23*(240-25)) + (1.863*3.76*28*0.25*(240-25))]

=-153423.66cal.

Enthalpy of Products:

= [(-4.413*111*1000/4.2) + (4.413*28*0.29*(T-25)) + (-1.9215*394*1000/4.2) + (1.9215*44*0.3*(T-25)) + (5.2*2*3.63*(T-25)) + (0.1755*12*0.71/4.2*(T-25) + (1.863*3.76*28*0.29*(T-25))]

= [-296884.29 + 148.02(T-25)] cal.

On solving T comes out to be 994.6 ≅ 995℃ .

(For Equivalence Ratio=0.35).

Partial Oxidation Reaction.

$$\mathbf{0.9 * [C_{6.6}H_{10}O_{4.4} + 2.42(O_2 + 3.76N_2) \rightarrow 3.92CO + 2.66CO_2 + 5H_2 + 0.02C_S + (3.76 * 2.42)N_2]}$$

Steam Gasification Reaction.

$$\mathbf{0.1 * [C_{6.6}H_{10}O_{4.4} + 2.2H_2O \rightarrow 6.6CO + 7H_2]}$$

Combined equation:

$$\mathbf{C_{6.6}H_{10}O_{4.4} + 2.178(O_2 + 3.76N_2) + 0.22H_2O \rightarrow 4.188CO + 2.394CO_2 + 5.2H_2 + 0.018C_S + (3.76 * 2.178)N_2}$$

Enthalpy of reactants:

= [(-720*1*1000/4.2) + (0.5*159.6*(240-25)) + (-0.22*286*1000/4.2) + (0.22*18*75*1) + (0.22*18*538) + (0.22*18*0.46*140) + (2.178*32*0.23*(240-25)) + (2.178*3.76*28*0.25*(240-25))]

=-150929.43cal.

Enthalpy of Products:

= [(-4.188*111*1000/4.2) + (4.188*28*0.29*(T-25)) + (-2.394*394*1000/4.2) + (2.394*44*0.3*(T-25)) + (5.2*2*3.65*(T-25)) + (0.018*12*0.71/4.2* (T-25) + (2.178*3.76*28*0.29*(T-25))]

= [-335262.86 + 160.92(T-25)] cal.

On solving T comes out to be 1170.25 ≅ 1170 ^{0}C

(For Equivalence Ratio=0.4).

Partial Oxidation Reaction.

$$0.9 * [C_{6.6}H_{10}O_{4.4} + 2.76(O_2 + 3.76N_2) \rightarrow 3.28CO + 3.32CO_2 + 5H_2 + (3.76 * 2.42)N_2]$$

Steam Gasification Reaction.

$$0.1 * [C_{6.6}H_{10}O_{4.4} + 2.2H_2O \rightarrow 6.6CO + 7H_2]$$

Combined equation:

$$C_{6.6}H_{10}O_{4.4} + 2.484(O_2 + 3.76N_2) + 0.22H_2O \rightarrow 3.612CO + 2.988CO_2 + 5.2H_2 + (3.76 * 2.484)N_2$$

Enthalpy of reactants:

= [(-720*1*1000/4.2) + (0.5*159.6*(240-25)) + (-0.22*286*1000/4.2) + (0.22*18*75*1) + (0.22*18*538) + (0.22*18*0.46*140) + (2.484*32*0.23*(240-25)) + (2.484*3.76*28*0.25*(240-25))]

=-150929.43cal.

Enthalpy of Products:

= [(-3.612*111*1000/4.2) + (3.612*28*0.3*(T-25)) + (-2.988*394*1000/4.2) + (2.988*44*0.3*(T-25)) + (5.2*2*3.7*(T-25)) + (2.484*3.76*28*0.29*(T-25))]

= [-375762.86 + 173.64(T-25)] cal.

On solving T comes out to be $\cong$ 1330 ^{0}C.

Annexure-II

Material Balance and Estimation of Temperature in the reactor after Water gas reaction.

(For Equivalence Ratio=0.2).

$$4.242\mathrm{CO} + 1.431\mathrm{CO_2} + 5.22\mathrm{H_2} + 0.927\mathrm{C_S} + (3.76 * 1.242)\mathrm{N_2} + 0.927H_2O$$
$$\rightarrow\ 5.169CO + 1.431CO_2 + 6.147H_2 + (1.242 * 3.76)N_2$$

Enthalpy of reactants:

= [(-4.242*111*1000/4.2) + (4.242*28*0.29*(725-25)) + (-1.431*394*1000/4.2) + (1.431*44*0.3*(725-25)) + (5.22*2*3.5*(725-25)) + (0.927*12*0.71/4.2*(725-25)) + (1.242*3.76*28*0.29*(725-25))]

=-204928.486 cal.

Enthalpy of products:

= [(-5.169*111*1000/4.2) + (5.169*28*0.29*(T-25)) + (-1.431*394*1000/4.2) + (1.431*44*0.3*(T-25)) + (6.147*2*3.5*(T-25)) + (1.242*3.76*28*0.29*(T-25))]

= [-270712.83 + 137.6(T-25)] cal.

Upon solving T comes out to be 504.8 ≅ 505℃.

(For Equivalence Ratio=0.26).

$$4.26\mathrm{CO} + 1.8\mathrm{CO_2} + 5.22\mathrm{H_2} + 0.54\mathrm{C_S} + (3.76 * 1.62)\mathrm{N_2} + 0.54H_2O$$
$$\rightarrow 4.8CO + 1.8CO_2 + 5.76H_2 + (1.62 * 3.76)N_2$$

Enthalpy of reactants:

= [(-4.26*111*1000/4.2) + (4.26*28*0.29*(900-25)) + (-1.8*394*1000/4.2) + (1.8*44*0.3*(900-25)) + (5.22*2*3.5*(900-25)) + (0.54*12*0.71/4.2*(900-25)) + (1.62*3.76*28*0.29*(900-25))]

=-181923.038 cal.

Enthalpy of products:

= [(-4.8*111*1000/4.2) + (4.8*28*0.3*(T-25)) + (-1.8*394*1000/4.2) + (1.8*44*0.29*(T-25)) + (5.76*2*3.49*(T-25)) + (1.62*3.76*28*0.29*(T-25))]

= [-295714.28 + 152.4(T-25)] cal.

Upon solving T comes out to be 774.6 ≅ 775℃.

(For Equivalence Ratio=0.3).

$$4.413\mathrm{CO} + 1.9215\mathrm{CO_2} + 5.2\mathrm{H_2} + 0.1755\mathrm{C_S} + (3.76 * 1.863)\mathrm{N_2} + 0.1755H_2O$$
$$\rightarrow 4.5885CO + 1.9215CO_2 + 5.3755H_2 + (1.863 * 3.76)N_2$$

Enthalpy of reactants:

= [(-4.413*111*1000/4.2) + (4.413*28*0.29*(995-25)) + (-1.9215*394*1000/4.2) + (1.9215*44*0.3*(995-25)) + (5.2*2*3.63*(995-25)) + (0.1755*12*0.71/4.2* (995-25)) + (1.863*3.76*28*0.29*(995-25))]

=-155294.92 cal.

Enthalpy of products:

= [(-4.5885*111*1000/4.2) + (4.5885*28*0.3*(T-25)) + (-1.9215*394*1000/4.2) + (1.9215*44*0.29*(T-25)) + (5.3755*2*3.63*(T-25)) + (1.863*3.76*28*0.29*(T-25))]

= [-301522.5 + 158.2(T-25)] cal.

Upon solving T comes out to be 960.41 ≅ 960℃

(For Equivalence Ratio=0.35).

$$4.188\mathrm{CO} + 2.394\mathrm{CO_2} + 5.2\mathrm{H_2} + 0.018\mathrm{C_S} + (3.76 * 2.178)\mathrm{N_2} + 0.018H_2O$$
$$\rightarrow 4.206CO + 2.394CO_2 + 5.218H_2 + (2.178 * 3.76)N_2$$

Enthalpy of reactants:

= [(-4.188*111*1000/4.2) + (4.188*28*0.3*(1170-25)) + (-2.394*394*1000/4.2) + (2.394*44*0.31*(1170-25)) + (5.2*2*3.7*(1170-25)) + (0.018*12*0.71/4.2* (1170-25)) + (2.178*3.76*28*0.29*(1170-25))]

=-145909.93 cal.

Enthalpy of products:

= [(-4.206*111*1000/4.2) + (4.206*28*0.3*(T-25)) + (-2.394*394*1000/4.2) + (2.394*44*0.29*(T-25)) + (5.218*2*3.7*(T-25)) + (2.178*3.76*28*0.29*(T-25))]

= [-335738.57 + 166.21(T-25)] cal.

Upon solving T comes out to be 1165℃.

Annexure-III

Material Balance and Estimation of Temperature in the reactor after shift reaction.

$$\mathbf{CO + H_2O = CO_2 + H_2}$$

$$K = \frac{[P_{CO_2}][P_{H_2}]}{[P_{CO}][P_{H_2O}]} = 1$$

The equilibrium constant for shift reaction at such high temperatures is assumed to be 1. Studies show that it goes towards 1 at higher temperatures (~800 ^{0}C). From calculations it was found that steam should be added as half the mole of CO.

(For Equivalence Ratio=0.2).

Let x moles of CO is converted to produce x moles of H_2 , then after reaction new concentration of gas will be:

CO : 5.169- x

CO_2 : 6.147+ x

H_2 : 1.431+x

H_2O : 2.5845-x

N_2 : 4.67

$$K = \frac{(6.147 + x)(1.431 + x)}{(5.169 - x)(2.5845 - x)} = 1$$

Upon solving x comes out to be 0.3.

Hence the equation that prevails after shift reaction is

$$\mathbf{5.169CO + 1.431CO_2 + 6.147H_2 + (1.242 * 3.76)N_2 + 2.585H_2O}$$
$$\mathbf{\rightarrow 4.869CO + 1.731CO_2 + 6.417H_2 + (1.242 * 3.76)N_2 + 2.285H_2O}$$

Enthalpy of reactants:

= [(-5.169*111*1000/4.2) + (5.169*28*0.29*(505-25)) + (-1.431*394*1000/4.2) + (1.431*44*0.3*(505-25)) + (6.147*2*3.5*(505-25)) + (1.242*3.76*28*0.29*(505-25)) + (-2.585*286*1000/4.2) + (2.585*18*75) + (2.585*18*538) + (2.585*18*0.57*480)]

= -341687.52 cal.

Enthalpy of products:

= [(-4.869*111*1000/4.2) + (4.869*28*0.29*(T-25)) + (-1.731*394*1000/4.2) + (1.731*44*0.3*(T-25)) + (6.417*2*3.5*(T-25)) + (1.242*3.76*28*0.29*(T-25)) + (-2.285*286*1000/4.2) + (2.285*18*75) + (2.285*18*538) + (2.285*18*0.57*(T-100)]

=-265852.31 + 135.3(T-25)

Upon solving T comes out to be 585.5≅ 586℃.

Concentration of product gas after shift reaction at a temperature of 586℃.

Species	**Moles**	**Mole % age by volume**
CO	4.869	24.38
CO_2	1.731	8.67
H_2	6.417	32.13
H_2O	2.285	11.44
N_2	4.67	23.38

(For Equivalence Ratio=0.26).

Let x moles of CO is converted to produce x moles of H_2 , then after reaction new concentration of gas will be:

CO : 4.8- x

CO_2 : 5.76+ x

H_2 : 1.8+x

H_2O : 2.4-x

N_2 : 6.09

$$K = \frac{(5.76 + x)(1.8 + x)}{(4.8 - x)(2.4 - x)} = 1$$

Upon solving x comes out to be 0.268.

Hence the equation that prevails after shift reaction is

$$\mathbf{4.8CO + 1.8CO_2 + 5.76H_2 + (1.62 * 3.76)N_2 + 2.4H_2O} \\ \rightarrow \mathbf{4.53CO + 2.07CO_2 + 6.03H_2 + (1.62 * 3.76)N_2 + 2.13H_2O}$$

Enthalpy of reactants:

= [(-4.8*111*1000/4.2) + (4.8*28*0.29*(505-25)) + (-1.8*394*1000/4.2) + (1.8*44*0.3*(775-25)) + (5.76*2*3.5*(775-25)) + (1.62*3.76*28*0.29*(775-25)) + (-2.4*286*1000/4.2) + (2.4*18*75) + (2.4*18*538) + (2.4*18*0.57*480)]

= -299806 cal.

Enthalpy of products:

= [(-4.53*111*1000/4.2) + (4.53*28*0.3*(T-25)) + (-2.07*394*1000/4.2) + (2.07*44*0.3*(T-25)) + (6.03*2*3.5*(T-25)) + (1.62*3.76*28*0.29*(T-25)) + (-2.13*286*1000/4.2) + (2.13*18*75) + (2.13*18*538) + (2.13*18*0.57*(T-100)]

=-454129.17 + 174.94(T-25)

Upon solving T comes out to be 906.84≅ 907 ℃

Concentration of product gas after shift reaction at a temperature of 907℃.

Species	**Moles**	**Mole % age by volume**
CO	4.53	21.73
CO_2	2.07	9.93
H_2	6.03	28.92
H_2O	2.13	10.22
N_2	6.09	29.21

(For Equivalence Ratio=0.3).

Let x moles of CO is converted to produce x moles of H_2 , then after reaction new concentration of gas will be:

CO : 4.5885- x

CO_2 : 5.3755+ x

H_2 : 1.9215+x

H_2O : 2.2943-x

N_2 : 7.00

$$K = \frac{(5.3755 + x)(1.9215 + x)}{(4.5885 - x)(2.2943 - x)} = 1$$

Upon solving x comes out to be 0.02.

Hence the equation that prevails after shift reaction is

$$\mathbf{4.5885}\boldsymbol{CO} + \mathbf{1.9215}\boldsymbol{CO_2} + \mathbf{5.3755}\boldsymbol{H_2} + (\mathbf{1.863} * \mathbf{3.76})\boldsymbol{N_2} + \mathbf{2.2943}\boldsymbol{H_2O} \rightarrow \mathbf{4.5685}\boldsymbol{CO} + \mathbf{1.9415}\boldsymbol{CO_2} + \mathbf{5.3955}\boldsymbol{H_2} + (\mathbf{1.863} * \mathbf{3.76})\boldsymbol{N_2} + \mathbf{2.2743}\boldsymbol{H_2O}$$

Enthalpy of reactants:

= [(-4.5885*111*1000/4.2) + (4.5885*28*0.29*(960-25)) + (-1.9215*394*1000/4.2) + (1.9215*44*0.3*(960-25)) + (5.3755*2*3.63*(960-25)) + (1.863*3.76*28*0.29*(960-25)) + (-2.2943*286*1000/4.2) + (2.2943*18*75) + (2.943*18*538) + (2.943*18*0.57*860)]

= -267932.03 cal.

Enthalpy of products:

= [(-4.5685*111*1000/4.2) + (4.5685*28*0.3*(T-25)) + (-1.9415*394*1000/4.2) + (1.9415*44*0.3*(T-25)) + (5.3955*2*3.63*(T-25)) + (1.863*3.76*28*0.29*(T-25)) + (-2.2743*286*1000/4.2) + (2.2743*18*75) + (2.2743*18*538) + (2.2743*18*0.57*(T-100)]

=-432644.5 + 175.39(T-25)

Upon solving T comes out to be 964.7≅ 965℃.

Concentration of product gas after shift reaction at a temperature of 965℃.

Species	**Moles**	**Mole % age by volume**
CO	4.5685	21.57
CO_2	1.9415	9.17
H_2	5.3955	25.48
H_2O	2.2743	10.74
N_2	7.000	33.05

(For Equivalence Ratio=0.35).

Let x moles of CO is converted to produce x moles of H_2 , then after reaction new concentration of gas will be:

CO : 4.206- x

CO_2 : 5.218+ x

H_2 : 2.394+x

H_2O : 2.103-x

N_2 : 8.19

$$K = \frac{(5.218 + x)(2.394 + x)}{(4.206 - x)(2.103 - x)} = 1$$

Upon solving x comes out to be 0.02.

Hence the equation that prevails after shift reaction is

$$\mathbf{4.206CO + 2.394CO_2 + 5,218H_2 + (2.178 * 3.76)N_2 + 2.103H_2O} \\ \rightarrow \mathbf{3.946CO + 2.654CO_2 + 5.478H_2 + (2.178 * 3.76)N_2 + 1.843H_2O}$$

Enthalpy of reactants:

= [(-4.206*111*1000/4.2) + (4.206*28*0.3*(1165-25)) + (-2.394*394*1000/4.2) + (2.394*44*0.31*(1165-25)) + (5.218*2*3.7*(1165-25)) + (2.178*3.76*28*0.29*(1165-25)) + (-2.103*286*1000/4.2) + (2.103*18*75) + (2.103*18*538) + (2.103*18*0.57*(1165-100)]

= -243273.63 cal.

Enthalpy of products:

= [(-3.946*111*1000/4.2) + (3.946*28*0.3*(T-25)) + (-2.654*394*1000/4.2) + (2.654*44*0.31*(T-25)) + (5.478*2*3.7*(T-25)) + (2.178*3.76*28*0.29*(T-25)) + (-1.843*286*1000/4.2) + (1.843*18*75) + (1.843*18*538) + (1.843*18*0.57*(T-100)]

=-459839.73 + 188.42(T-25)

Upon solving T comes out to be 1174.5≅ 1175℃.

Concentration of product gas after shift reaction at a temperature of 1175 ℃.

Species	**Moles**	**Mole % age by volume**
CO	3.946	17.66
CO_2	2.654	11.87
H_2	5.478	24.51
H_2O	2.083	9.32
N_2	8.19	36.64

(For Equivalence Ratio=0.4).

Let x moles of CO is converted to produce x moles of H_2 , then after reaction new concentration of gas will be:

CO 3.612- x

CO_2 : 5.2+ x

H_2 : 2.988+x

H_2O : 1.806-x

N_2 : 9.34

$$K = \frac{(5.2 + x)(2.988 + x)}{(3.612 - x)(1.806 - x)} = 1$$

Upon solving x comes out to be 0.66.

Hence the equation that prevails after shift reaction is

$$3.612CO + 2.988CO_2 + 5{,}2H_2 + (2.484 * 3.76)N_2 + 1.806H_2O \\ \rightarrow 2.952CO + 3.648CO_2 + 4.54H_2 + (2.484 * 3.76)N_2 + 1.146H_2O$$

Enthalpy of reactants:

= [(-3.612*111*1000/4.2) + (3.612*28*0.3*(1330-25)) + (-2.988*394*1000/4.2) + (2.988*44*0.31*(1330-25)) + (5.2*2*3.7*(1330-25)) + (2.484*3.76*28*0.29*(1330-25)) + (-1.806*286*1000/4.2) + (1.806*18*75) + (1.806*18*538) + (1.806*18*0.57*(1330-100)]

= -210513.05 cal

Enthalpy of products:

= [(-2.952*111*1000/4.2) + (2.952*28*0.3*(T-25)) + (-3.648*394*1000/4.2) + (3.648*44*0.32*(T-25)) + (4.54*2*3.9*(T-25)) + (2.484*3.76*28*0.3*(T-25)) + (-1.146*286*1000/4.2) + (1.146*18*75) + (1.146*18*538) + (1.146*18*0.64*(T-100)]

=-486616.03 + 203.27(T-25)

Upon solving T comes out to be 1380℃.

Concentration of product gas after shift reaction at a temperature of 1380 ℃.

Species	**Moles**	**Mole % age by volume**
CO	2.952	13.65
CO_2	3.648	16.87
H_2	4.54	20.99
H_2O	1.146	5.30
N_2	9.34	43.19