DESIGN OF HIGH TEMPERATURE STEAM AND AIR GASIFICATION PLANT USING NON-WOODY BIOMASS WASTE

BY

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DECLARATION

I hereby declare that this project work is a record of a research work that was undertaken and written by me. It has not been presented before for any degree or diploma or certificate at any university or institution. Information derived from personal communications, published and unpublished work were duly referenced in the text.

Tyger precious Gashon

08/03/020
Date

CERTIFICATION

This is to certify that **"The Design of High Temperature Steam and Air Gasification Plant Using Non-Woody Biomass"** Waste was solely carried out by Mr. Tyger Gashon Precious (2004/18560EH) under the supervision of Mrs. M.D Yahya. Proper mention has been made in the reference section of the authors whose materials were found useful in one way or the other in the course of this project work.

	05 - 03 - 2010
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EXTERNAL EXAMINER	**Date**

DEDICATION

This project is dedicated to God almighty for seeing me through this years of my academic endeavours glory be to his name. amen

ACKNOWLEDGEMENT

All praises, gratitude and appreciation go to almighty God for seeing me through all the vigorous difficulties of this programme.

I am indebted to my able supervisor Engr. Umar for his patience, understanding encouragement and for giving this project work its proper shape and form. To the Head of Department Chemical Engineering, Engr. J.O. Okafor , I say a big thank you for your fatherly role in the department and to the entire staff of the department thank you for being there for us always.

My sincere thanks go to my friends who has one way or the other contributed to my success throughout my university programme, thanks goes to, Desmond Okpe, Ibekie Henry, Ariyo Mohammed, Chima Owunchekwa, Bekee Nuka Holy, Chidi Anunobi, segun.

I wish to thank my parent for their financial support to made me successful in this citadel institution my prayer is that God in his infinite mercy should guide them, protect and to shower his blessing over them (amen). I will never forget the contribution of my Parents Chief Gashon Tyger and Mrs. Eunice Brown and all my engineering friends.

ABSTRACT

The project is aimed at designing a plant for the production of at least 28.92149 kg/hr synthesis gas. In this design the material balance, equipment design, energy balance and economic analysis are carried out manually. The various units involved in the process plant are as follows Gassifier, Preheater, Catalytic Reformer, Mechanical filter, Scrubber, Boiler, Cyclone, Filtration tank.

Standard cost estimation was carried out. The total capital investment was determined to be $3858186.67 dollars, working capital cost were $1664292.3 dollars the net profit of $1433961.15 dollars it was revealed that the project is economically viable with pay back period of 2.307 years and rate of return of 43.551 %

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CHAPTER ONE

1.0 INTRODUCTION

Energy is an important input for economic development. Since exhaustible energy sources in the country are limited, there is an urgent need to focus attention on development of renewable energy sources and use of energy efficient technologies. The exploitation and development of various forms of energy and making energy available at affordable rates is one of our major thrust areas.

Utilization of renewable energy sources such as biomass offers environmental benefits While providing sustainable power generation for utilities and industry. Many countries of the world agreed on the policy of CO_2 reduction to the atmosphere, or at least keeping them at current levels in which heat and power generation in future will have to be CO_2 neutral, or at least CO_2 lean, to meet environmental constraints and agreed commissions. A way to expeditiously mitigate greenhouse gas emissions is switching fuels from fossil fuels to renewable energy e.g. biomass fuels as its utilization gives a zero net production of greenhouse gases. The utilization also offers markets for dedicated energy crops such as switchgrass, while providing needed energy for power generation, thereby offsetting or reducing the need for fossil fuels. In areas where fossil fuels are scarce and biomass fuels are available, biomass utilization offers alternative low cost methods for power generation.

The use of biomass as a feedstock for gasification systems is becoming increasingly economically attractive. The syngas produced by gasification should ideally be used locally, utilizing the fuel to produce power, using the fuel to offset natural gas in heating applications or to convert the syngas into a liquid fuel.

There are many types of gasifiers with varying schemes for both reactor design and reaction media in a wide range of operating conditions. The suitability for a particular usage, i.e. the fuel gas quality, is determined by the gas composition and the level of contamination by particulates, alkali compounds, nitrogen-containing components, sulphur and tars. Typically the counter current gasifiers produce more tar than fluidized bed and co-current gasifiers, the last being the cleanest technology of the three.

The fuel gas quality is influenced by the gasifier type and also by the composition of the reactants. The gasifying agent provides oxygen for the formation of CO from solid carbon in the fuel. The gasifying agent can be air, pure oxygen, steam, CO_2 or a mixture of these gases. The CO_2 is produced during pyrolysis and early oxidation processes and is generally not externally added. The most common agent is air because of its extensive availability at no cost but is not a perfect agent because of its nitrogen content. The product gas from air gasification produces a low heating value gas [4-7MJ/Nm^3].

Oxygen gasification produces a higher heating value [10-18MJ/Nm^3] but has a disadvantage of relatively high cost. Steam is another alternative and it generates a medium calorific value gas [10-14MJ/Nm^3]. The key advantage is that it increases the hydrogen content of the product gas. The presence of steam is important in case of further catalytic upgrading of the product gas. Steam gasification is however a highly endothermic reaction and requires a temperature in excess of 800^{o}c to take place if no catalyst is used.

Quality gas is generally positively influenced by a high reaction temperature. To reach higher gasifier bed temperatures [above 1000^{o}c], the heat required for the reaction has to be transferred either by partial combustion of fuel in the same reactor [mixing of H_2O with oxygen/air] or by indirect heating, the procedure normally associated with adjustment of the gasifier design or configuration. Gasification produces hydrogen, carbon monoxide, methane, light hydrocarbons and several other undesired compounds, such as organic aerosols, tars, that may include sodium, potassium and chlorine compounds, ammonia, and hydrocyanic acid. Particularly cumbersome is tar and its significant yield, varying from 1 to 180g/Nm^3, depending on the reactor type, feed gas, fuel type.

Gasification of biomass, rather than combustion, offers efficiency, environmental, and operational advantages including the ability to use the producer gas in fuel cells. Gasification process usually takes place at high temperature from 800 to1800C. However, with recent development in high temperature air combustion and gasification technology, it is now possible to have air preheats much in excess of 1100C. Temperature, gasification agent and biomass waste composition affects the quality of product gas composition and tars produced during gasification. Hydrogen or hydrogen-rich gas produced from biomass could be readily used in most of the present natural gas or petroleum derived hydrogen energy conversion devices and also in advanced systems such as fuel cells.

1.1BIOMASS ENERGY

Non woody fuel has several environmental advantages over fossil fuel. The main advantage is the non-woody is a renewable biomass resource, offering a sustainable, dependable supply. Other advantages include the fact that it produces a rich hydrogen gas. The gasification characteristics of all non woody biomass waste examined was similar and depends on the biomass fuel composition. It is not a threat to acid rain pollution, and particulate emissions are controlled.

The principle economic advantage of non woody biomass energy is that it is usually significantly less expensive than competing fossil fuels. Public institutions such as schools, hospitals, prisons, and municipality-owned heating projects, are prime targets for using non-woody biomass energy.

Before building or remodeling a facility to utilize non–woody biomass for energy, potential users should evaluate the local market for the available supply of the feedstock. Transportation cost may limit the benefits of burning non-woody fuels. New and existing technology for using non-woody fuel effectively can be a combination of non-woody combustion, gasification, cogeneration and cofiring, depending on the fuel application. (Techline 2008)

1.2 AIR/STEAM GASIFICATION

In the steam air gasification process, we are concerned with systems in which the oxidant is introduced directly into the gasifier either with air or as relatively pure oxygen. Such systems are said to be directly heated or autothermal. The intended use of the product gas and the feedstock properties determine whether the oxygen is supplied as air or in the pure form, and the relative rates of steam, oxygen, and feedstock fed to the gasifier. Considering the limiting case in which no steam is used and the oxygen supplied in air is sufficient to form carbon monoxide. Then

$$C + \frac{1}{2}(O_2 + 4N_2) = CO + 2N_2 \quad \Delta H^{o}_{1000} = -111.9 KJ/mol$$

This gas is called producer gas, contains about 33percent by volume of carbon monoxide, and has a calorific value of about 5.7MJ/m^3. If in addition, steam is supplied to the gasifier then the product gas, sometimes called a "semi-water gas," will contain hydrogen and have a gross calorific value of about 7MJ/m^3.

Such gases are suitable for burning on site, for example to produce electricity in gas turbines, or in so-called "combined cycle plants" in which the hot exhaust gases from the gas turbine are used to raise steam for additional electricity generation. If the gas is intended for chemical synthesis such as fischer-tropsch, methanol, or substitute natural gas (SNG) production, then the presence of nitrogen is undesirable. This is because the presence of inerts increases the volume of gas to be processed or transported and hence increases processing and transmission cost. Although air separation to produce pure oxygen is expensive, and can infact amount to 25 to 35 percent of the cost of the product gas in SNG production. The process is currently the chosen method of overcoming the nitrogen barrier. The use of pure oxygen feed to the gasifier does, however, have advantages which are listed below other, other than the absence of nitrogen in the product and subsequent decrease in the size of downstream processing equipment.

- Smaller quantity of gas produced means a correspondingly smaller sensible heat loss from the gasifier.
- The gasifier can be operated economically at elevated pressure as the volume of oxygen to be compressed is only one-fifth the equivalent volume of air.
- Elevated pressure operation in turn means smaller equipment size and increased methane production.

The semi-water gas mentioned above becomes a water or blue water gas if nitrogen is not present, and would typically have a calorific value in the neighborhood of 13MJ/m^3.

A gas consisting mainly of carbon monoxide and hydrogen will be called a "synthesis gas," as it may be processed to yield a variety of synthetic fuels and chemicals.

$$C + H_2O\,(g) = CO + H_2 \qquad \Delta H^{o}_{1000} = +\ 135.9KJ/mol$$

The reaction is highly endothermic and so require a heat supply. Although the required heat is generally obtained by burning carbon in oxygen, the transfer of heat to the gasifier may be by indirect means via electricity or a heat transfer medium. From the equation above, the heat required for the gasification of carbon with steam to form carbon monoxide and hydrogen at 1000k is 135.9KJ/mol.

1.3 THE OBJECTIVES AND SCOPE OF THE DESIGN

The project is aimed at:

- Designing a gasifier plant to utilize non-woody biomass waste using high temperature steam and air.
- Carry out material and energy balance for the entire process.
- Carry out the equipment design of the major equipment and to carry out detailed design of the major unit in the process.
- Carry out the operability study, cost and economic evaluation of the entire plant.

1.4 METHODOLOGY

All the calculation aspect in this project will be calculated manually.

CHAPTER TWO

2.0.0 LITERATURE REVIEW

A literature search on biomass gasification technology was done to determine the current Status of biomass gasification commercialization, identify near-commercial processes and collect reliable gasification data. More than 40 sources, including a number of web sites, provided data on biomass gasification technologies. The goal was not to select a 'superior' technology, but rather to collect, organize, verify and assess biomass gasification process data. Such data can be used in future studies to determine the best match of an available biomass gasification technology to a process application of interest, such as chemical synthesis, fuel production, or combined heat and power (CHP) generation (Anil K. Rajvanshi)

2.0.1 CURRENT STATUS OF GASIFICATION TECHNOLOGY

Excellent survey of current status of gasification technology has been carried out by Foley and Barnard. They have reviewed the status in both developed and developing countries.

However there is confusion regarding the number of manufacturers of gasification equipment. Quite a number of these manufacturers have just produced few units, which are still in experimental stages. There are therefore, close to 64 manufacturers all over the world. In U.S. alone there are 27 manufacturers and about 13 Universities and USDA research stations working on various aspects of biomass gasification.

The world's largest gasification manufacturing facility is Gasifier and Equipment Manufacturing Corporation (GEMCOR) in Philippines. They produce about 3000 units/year ranging in size from 10-250 kW. Besides they have recently started producing gasifiers for direct heat applications. Their primary applications have been for irrigation pumps and power generating sets. To date about 1000 units have been installed within Philippines running on charcoal, wood chips and briquettes. Brazil is another country where large scale gasification manufacturing program has been undertaken. About 650 units of various sizes and applications have been installed.

In both the Brazilian and Philippines program the gasifiers are mostly charcoal powered. In this a strict quality control of the fuel has to be maintained. Thus the companies involved in gasifier manufacturing also supply the quality fuel. Inadequate fuel quality is the biggest problem in running these gasifiers.

In Europe there are many manufacturers especially in Sweden, France, West Germany and Netherlands who are engaged in manufacturing gasification systems for stationery applications. Most of market for these European manufacturers has been in developing countries. The U.S. and North American manufacturing activities have been summarized by Goss. In the research area the most active program in gasification is at University of California, Davis and University of Florida, Gainesville. Many systems in the range of 10-100 kW have been developed at Davis. U.S. also is ahead of the rest of world in direct heat application gasifiers.

Both fluidized and fixed bed gasifiers have been developed for this purpose.

In other countries of Asia and Africa the work is being carried out in research institution and few prototypes have been made and tested . Interestingly enough no mention of Japan is there in any world wide gasification literature. However if the gasification technology does pick up it will be only a matter of time before Japan flexes its economic muscle and mass produces the gasifiers at cheaper rates.

Most of the gasifiers (upto 100 kW range) being sold by different manufacturers show a leveling off price of $ 380/KWe for plant prices and about $ 150 KWe for basic gasifier price . This leveling off comes at about 100 kW system. However, for small systems the prices are extremely high. Thus a 10 KWe gasifier plant costs about $ 840/KWe while the basic gasifier is $ 350/KWe. To this must be added the transportation costs (especially for shipment to developing countries). These prices therefore can make the gasifiers uneconomic. This explains the big gasifier manufacturing push being give in countries like Philippines, Brazil etc.

Unfortunately with all the activities going around the world the impact of gasification technology till to date on the economy has been negligible and far smaller than that of other renewable energy namely Solar. However gasification is a recently rediscovered technology and most of the development is still on learning curve.

2.1.0 THEORY OF GASIFICATION

The production of generator gas (producer gas) called gasification, is partial combustion of solid fuel (biomass) and takes place at temperatures of about 10000C. The reactor is called a gasifier.

The combustion products from complete combustion of biomass generally contain nitrogen, water vapor, carbon dioxide and surplus of oxygen. However in gasification where there is a surplus of solid fuel (incomplete combustion) the products of combustion are (Figure below) combustible gases like Carbon monoxide (CO), Hydrogen (H_2) and traces of Methane and non-useful products like tar and dust. The production of these gases is by reaction of water vapor and carbon dioxide through a glowing layer of charcoal. Thus the key to gasifier design is to create conditions such that a) biomass is reduced to charcoal and, b) charcoal is converted at suitable temperature to produce CO and H_2(Anil et a.,1986)

2.1.1 BIOMASS FUEL

Almost any carbonaceous or biomass fuel can be gasified under experimental or laboratory conditions. However the real test for a good gasifier is not whether a combustible gas can be generated by burning a biomass fuel with 20-40% stoichiometric air but that a reliable gas producer can be made which can also be economically attractive to the customer. Towards this goal the fuel characteristics have to be evaluated and fuel processing done.

Many gasifier manufacturers claim that a gasifier is available which can gasify any fuel. There is no such thing as a universal gasifier. A gasifier is very fuel specific and it is tailored around a fuel rather than the other way round. Thus a gasifier fuel can be classified as good or bad according to the following parameters :

1) Energy content of the fuel
2) Bulk density
3) Moisture content
4) Dust content
5) Tar content
6) Ash and slagging characteristic

A. Energy content and Bulk Density of fuel

The higher the energy content and bulk density of fuel, the similar is the gasifier volume since for one charge one can get power for longer time.

B. Moisture content

In most fuels there is very little choice in moisture content since it is determined by the type of fuel, its origin and treatment. It is desirable to use fuel with low moisture content because heat loss due to its evaporation before gasification is considerable and the heat budget of the gasification reaction is impaired. For example, for fuel at 250C and raw gas exit temperature from gasifier at 3000C, 2875 KJ/kg moisture must be supplied by fuel to heat and evaporate moisture.

Besides impairing the gasifier heat budget, high moisture content also puts load on cooling and filtering equipment by increasing the pressure drop across these units because of condensing liquid.

carbon dioxide, acetic acid and water are given off. The real pyrolysis, which Thus in order to reduce the moisture content of fuel some pretreatment of fuel is required.

Generally a desirable moisture content for fuel should be less than 20%.

C. Dust content

All gasifier fuels produce dust. This dust is a nuisance since it can clog the internal combustion engine and hence has to be removed. The gasifier design should be such that it should not produce more than 2-6 g/m3 of dust. Figure 7 shows dust produced as a function of gas production for wood generators used during World War.

The higher the dust produced, more load is put on filters necessitating their frequent flushing and increased maintenance.

2.2 GASIFICATION CHEMISTRY

The following major reactions take place in combustion and reduction zone.

1. Combustion zone

The combustible substance of a solid fuel is usually composed of elements carbon, hydrogen and oxygen. In complete combustion carbon dioxide is obtained from carbon in fuel and water is obtained from the hydrogen, usually as steam. The combustion reaction is exothermic and yields a theoretical oxidation temperature of 14500^0C[6]. The main reactions, therefore, are:

$C + O_2 = CO_2$ (+ 393 MJ/kg mole) (1)

$2H_2 + O_2 = 2H_2O$ (- 242 MJ/kg mole) (2)

2. Reaction zone

The products of partial combustion (water, carbon dioxide and uncombusted partially cracked pyrolysis products) now pass through a red-hot charcoal bed where the following reduction reactions take place.

$C + CO_2 = 2CO$ (- 164.9 MJ/kg mole) (3)

$C + H_2O = CO + H_2$ (- 122.6 MJ/kg mole) (4)

$CO + H_2O = CO + H_2$ (+ 42 MJ/kg mole) (5)

$C + 2H_2 = CH_4$ (+ 75 MJ/kg mole) (6)

$CO_2 + H_2 = CO + H_2O$ (- 42.3 MJ/kg mole)

Reactions (3) and (4) are main reduction reactions and being endothermic have the capability of reducing gas temperature. Consequently the temperatures in the reduction zone are normally 800-10000C. Lower the reduction zone temperature (~ 700-8000C), lower is the calorific value of gas(Anil et al., 1986)

3. Pyrolysis zone

Wood pyrolysis is an intricate process that is still not completely understood. The products depend upon temperature, pressure, residence time and heat losses. However following general remarks can be made about them. Upto the temperature of 2000^0C only water is driven off. Between 200 to (650 to 830 °C). The higher feed gas temperature led to increase in concentration of CO (20.1 to 26.8 %) and H2 (6.6 to 12.7 %) in producer gas. The critical feed gas temperature (at which the yield of gaseous products was maximum) was between ignition temperature of biomass and the melting temperature of the ash in the biomass. 2800^0C takes place between 280 to 5000^oC, produces large quantities of tar and gases containing carbon dioxide. Besides light tars, some methyl alcohol is also formed. Between 500 to 7000^0C the gas production is small and contains hydrogen.

Thus it is easy to see that updraft gasifier will produce much more tar than downdraft (one). In downdraft gasifier the tars have to go through combustion and reduction zone and are partially broken down. Since majority of fuels like wood and biomass residue do have large quantities of tar, downdraft gasifier is preferred over others. Indeed majority of gasifiers, both in World War II and presently are of downdraft type.

Finally in the drying zone the main process is of drying of wood. Wood entering the gasifier has moisture content of 10-30%. Various experiments on different gasifiers in different conditions have shown that on an average the condensate formed is 6-10% of the weight of gasified wood. Some organic acids also come out during the drying process. These acids give rise to corrosion of gasifiers(H. Boerrigter, et al., 2007).

2.3 GASIFICATION TECHNOLOGY AND DESIGN

2.3.1 FIXED BED GASIFIER SYSTEMS

There are many types of fixed bed gasifiers with varying schemes for both reactor design and reaction media. The fixed bed gasifier can be classified according to the ways in which the gasifying agent enters the gasifier i.e. updraft, downdraft, crossdraft and two stage gasifier. The downdraft gasifiers are Imbert type (gasifier with throat) and open core type (throatless). The gasifying media may be air, steam, oxygen or a mixture of these and the producer gas may be used in thermal (heat gasifiers) or engine (power gasifiers) applications. The composition of producer gas and the level of contamination vary with the biomass, type of gasifier and operating conditions (Bridgwater, et al.,1999)

2.3.2 Updraft Gasification Systems

2.3.2.1 Principle

The updraft gasifier is the oldest and simplest form of fixed bed gasifier. It can handled.Biomass fuels with high ash (up to 15 %) and high moisture content (up to 50 %). It is more robust than other fixed bed gasifiers because it is less sensitive to variations in size and quality of biomass. In an updraft (counter current) gasifier, biomass fuel enters from the top of the reaction chamber and the gasifying media or agent (air, O2 or mixture) enters from the bottom of the unit from below a grate. The fuel flows down slowly through the drying, pyrolysis, gasification and combustion zones. The ash is removed from the bottom. The updraft gasifier has high thermal efficiency as the sensible heat of the producer gas is recovered by direct heat exchange with the entering feedstock, which is dried, and pyrolysed before entering the gasification zone. The producer gas exits at low temperature (80-300 °C) and contains an abundance of oils and tar (10-20 %) since the products of the pyrolysis and drying zone are directly drawn into it without decomposition. The dust content in the producer gas is low due to low gas velocities and filtering effect of feed in drying and pyrolysis zones(sangeeta et al.,).

2.3.2.2 High Temperature Agent Gasification (HiTAG) in Updraft Gasifiers

The high temperature (>1000 °C) agent gasification in updraft gasifiers permits the gasification of a wide range of feedstock including low rank biomass fuels and waste such as sludge. The higher temperatures of the preheated feed gas lead to a lower yield of tar, higher production of producer gas rich in H_2 content. Carlos (2005) investigated the high temperature air/steam gasification process for gasification of bark, charcoal, woodchips and wood pellets in a conventional batch type countercurrent updraft gasifier. Preheated air, steam and air/steam mixture were used as gasifying media. Preheating of air up to 830 °C kept the temperature in the oxidation zone of gasifier above 1000 °C, which promoted the thermal cracking of tar in one step in the gasifier. The lower heating value (LHV) of producer gas, cold gas efficiency and specific gas production rate (SGPR) increased from 4.6 to 7.3 MJ/Nm3, 36.1 to 45.2 % and 744.9 to 916.6 kg m-2 h-1 respectively as the temperature of the preheated feed gas was increased from 350 to 830 °C. With preheated air of 830 °C the LHV (7.3 MJ/Nm3) of producer gas was well above that reported for downdraft gasifiers. When the feed gas used was mixture of steam and air, increase in the steam fraction with respect to the high temperature air favoured the water gas shift equilibrium and steam reforming of tars and hydrocarbons which led to increase in fraction of H_2 (10.4 to 29.9%) and decrease in CO (29.4 to 18.4%) in the producer gas. However use of steam lowered the temperature of gasification and the cold gas efficiency (43 to 37%). of wood pellets in a batch type updraft fixed bed gasifier (Fig.2.1). The gasifier was a vertical,cylindrical reactor consisting of wind box and gasifier bed. The feedstock bed was supported by a bed of ceramic balls placed on perforated disk inside the reactor. The gasification process became faster with increase in temperature of feed gas(sangeeta et al.,).

2.3.2.3 Gasification of Non-Woody Biomass in Updraft Gasifiers

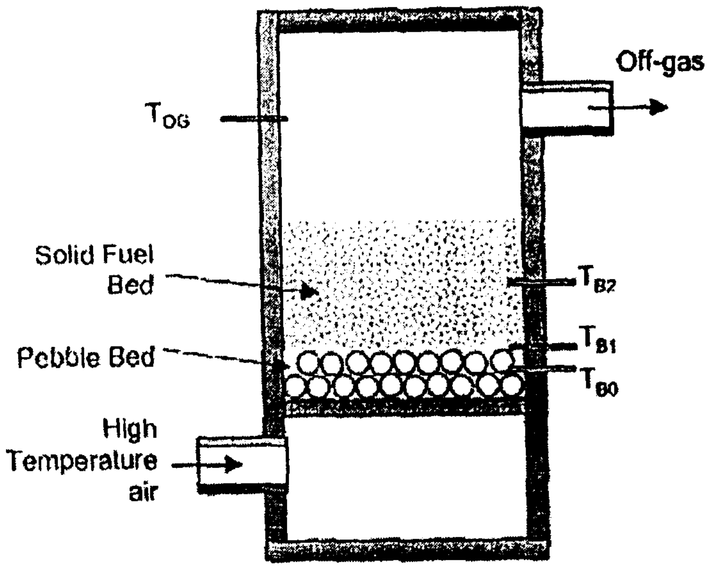


Fig.2.1 Updraft gasifier for high temperature air gasification

The updraft gasifier was generally used for gasification of conventional biomass fuel like wood, wood chips, bark etc. However studied the gasification of nonconventional fuels like low-density refuse derived fuel (RDF) pellets in an updraft gasifier with an inclined grate. The grate inclination helped in crushing of large clinkers. An airpartitioner in the gasifier ensured uniform distribution of air in the combustion zone. The high heating value (HHV) and energy content of producer gas was 5.58 MJ/Nm3 and 12.2 MJ/kg respectively. The tar content from RDF pellets was 45 % less than that in gas generated from wood chips. The specific gasification rate of RDF pellets was within the range reported for updraft gasifiers with fixed grate i.e. SGR 100-200 kg m-2 h-1. The cold gas efficiency obtained with RDF was 73 %, which was over 8 % higher than that obtained with wood chips. gasified combustible waste pellets in an updraft fixed bed gasifier lined with alumina refractory (Fig. 2) using O_2 as the gasifying medium. The composition of H_2 and CO in the producer gas was in the range 30-40 % and 15-30 % respectively depending on oxygen/waste ratio. As the bed height was increased the H_2 and CO content in producer gas increased whereas CO_2 content reduced. Gasification was difficult above a bed height of 700 mm because of pressure drop of O2. The cold gas efficiency was around 61 % at O_2/waste ratio of 0.4-0.5 and the heating value of producer gas was 11.72-13.40 MJ/Nm3, which was more than heating value of producer gas generated by air gasification. The gas was cleaned of the particulate matter by a cyclone. O_2 gasification was reported to produce a medium heating,which is suitable for limited pipeline distribution and as synthesis gas for conversion to methane and methanol(sangeeta et al.,).

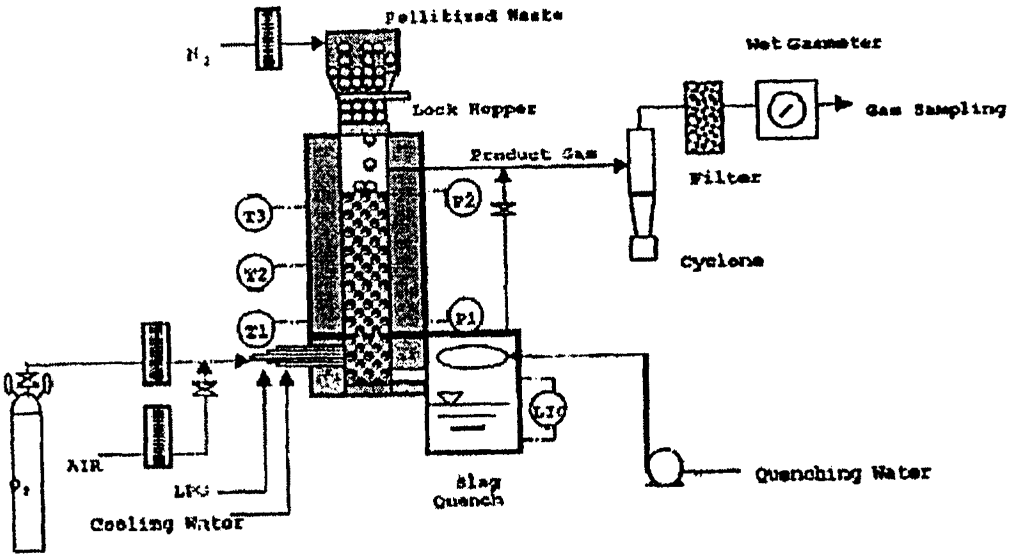


Fig. 2.2 . Updraft gasifier using O2 as gasifying medium

2.3.2.4 Application of Updraft Gasifier Systems

The producer gas from updraft gasifier has high amount of tar and is therefore mostly used in thermal applications i.e. close coupled steam boilers and crop dryers. the performance of an updraft gasifier (Fig.2.2) used in conjunction with a combustor for grain drying. Maize cobs of varying moisture content (9-46 %) were gasified in the gasifier with primary air as the gasifying medium. The producer gas was then completely combusted with secondary air and the exhaust gas mixed with the ambient air and used directly for drying the grain. The total particulate emission was proportional to the second power of the gasification rate (amount of corncobs gasified per unit grate area per unit time), which in turn was dependent on the corncob moisture content and primary airflow rate(sangeeta et al.,).

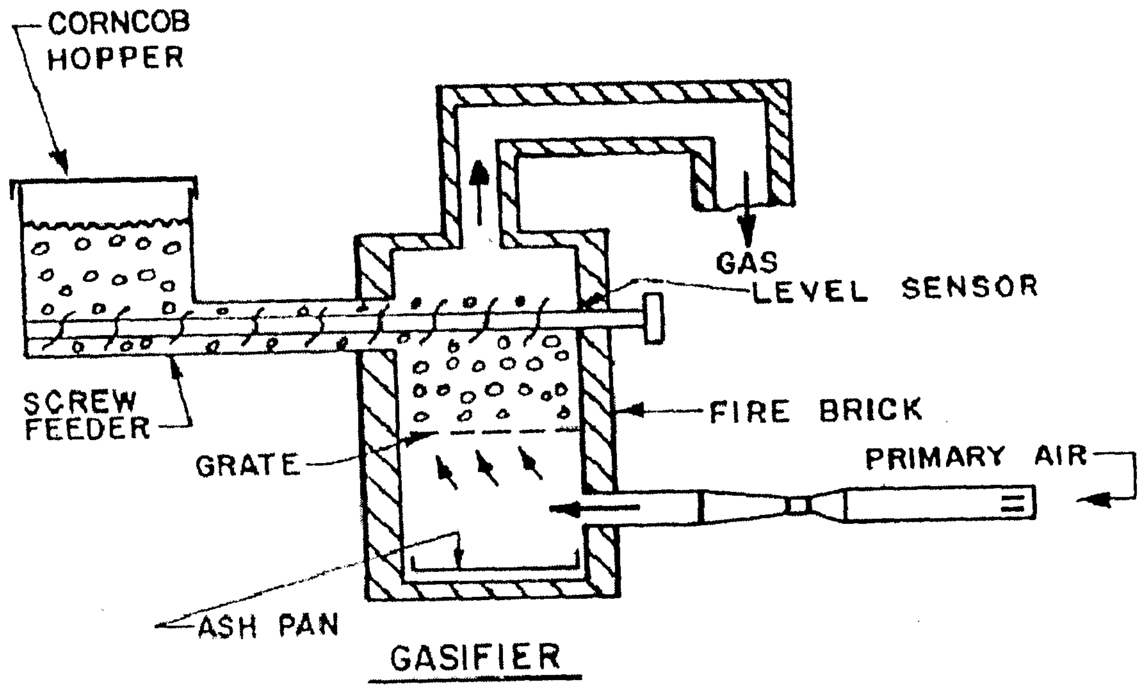


Fig. 2.3 Updraft gasifier for corn drying

VTT (2002) reported commercial use of an updraft fixed bed gasifier of 8 MW capacity (BIONEER) for thermal applications. The gasifier consisted of a refractory lined vessel with a rotating cone shaped grate. The temperature of the combustion zone was controlled by humidification of gasification air. The gasifier was used to gasify a variety of feedstock (wood chips, forest wastes, peat, straw etc) and the gas was used for burning in a boiler to generate hot water. For successful operation the biomass fuel had moisture content less than 50 %, ash less than 10 % of dry matter with minimum softening point of ash above 1190 °C.

Above 50 % moisture content the producer gas contained lot of tar aerosols. The producer gas generated from gasification of wood chips of 41 % moisture content consisted of 30 % CO, 11 % H_2, 3 % CH_4, 7 % CO_2, and 49 % N_2 with HHV of 6.2 MJ/Nm3. The tar content of dry producer gas was in the range 50–100 g/Nm3. The gasifier proved to be economically feasible for small heating systems. reviewed the gasification technologies being applied in China and surmised that the updraft gasifiers of capacity 2–30 kW were used for gasification of tree barks and timber blocks generating producer gas of lower heating value 4.1–5.3 MJ/Nm3 at efficiency of 70-75 %. The producer gas was being used as boiler fuel commercially(sangeeta et al.,)..

The high amount of tar and pyrolysis products in the producer gas of updraft gasifier makes it highly unsuitable in engines and gas turbines, that the updraft gasification between 2 – 20 MW was suitable for supplying fuel gas to reciprocating engines at efficiency of 30-35 % only, if the gas phase tars were cracked into smaller components, which do not condense at lower temperatures. This can be done by thermal cracking i.e. by raising temperature of gasifier over 1200 °C(sangeeta et al.,).

2.3.3 Imbert Downdraft Gasification System

2.3.3.1 Principle

The Imbert downdraft gasifier or throated downdraft gasifier features a cocurrent flow of gases and solids through a descending packed bed, which is supported across a constriction or throat. The biomass fuel enters through the hopper and flows down, gets dried and pyrolysed before being partially combusted by the gasifying media (air) entering at the nozzles. The throat allows maximum mixing of gases in high temperature region, which aids tar cracking.below the constriction n or 'throat' the combustion gases along with tar pass through the hot char and are reduced to primarily CO and H_2. The gasifier can handle uniformly sized biomass fuels having moisture content and ash content less than 20 % and 5 % respectively(sangeeta et al.,).

The throated (Imbert) downdraft gasifier is generally used for gasification of woody biomass of uniform sizes and shapes (blocks) as they flow smoothly through the constricted hearth.

The producer gas from downdraft gasifier has lesser tar-oils (<1 %), higher temperature (around 700 °C) and more particulate matter than that from an updraft gasifier. The gasifier has lower overall efficiency since a high amount of heat content is carried over by the hot gas (Clarke, 1981; Reed and Das, 1988). The physical limitations of biomass particle size limit the capacity of the throated downdraft gasifiers to 500 Kw(sangeeta et al.,).

2.3.3.2 Modifications in Throated Downdraft Gasifier System

The throat of the downdraft gasifier presents a hazard to low-density biomass fuels. The biomass in form of twigs, sticks or bars does not flow down smoothly causing bridging and channelling leading to increased tar production. The throated downdraft gasifier operation is very sensitive to feedstock size and quality. Modifications have been made in the gasifier to successfully gasify the low and medium density fuels like stalks, shells etc(sangeeta et al.,)..

Introduced a rotating grate and a double conical hopper in a downdraft gasifier for gasification of coir dust, cotton stalk and wheat straw. The fuel had lower density and higher ash content compared to that of wood and was densified into briquettes. The double conical hopper in the gasifier (Fig.2.4) created space for the fuel briquettes to expand when they were heated up to avoid the bridging problem in the pyrolysis zone. A rotating grate fed out the slag particles before their agglomeration in the gasifier. The distance between the choke plate and the grate was also increased to increase the volume of the reduction zone to get better distribution of slag in the char bed. The gasification system consisted of a gasifier with a cyclone for coarse dust removal and a glass fibre-fabric for removal of fine material. The cold gas efficiency was 72-77 %. The tar content was below 1 g/Nm3. Wheat straw pellets were also tested but due to small particle size, streams of cold gas passed between the throat and air nozzles resulting in incomplete tar cracking and in producer gas with high tar content(sangeeta et al.,).

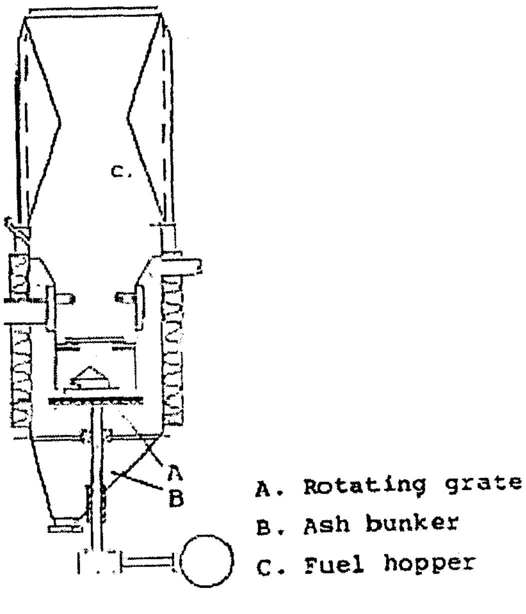


Fig. 2.4 Downdraft gasifier modified for non-woody biomass

Investigated the gasification of chips of rubber wood of varying moisture content (12.5-18.5 %) and chip size (3.3–5.5 cm) in an 80 kW downdraft throated gasifier, which was double walled with an air gap in between. During the operation the gasifier was shaken to avoid bridging of fuel in throat. Small chips underwent faster char conversion increasing the conversion efficiency and therefore required smaller gasification zone length.

Higher throat angle decreased temperature, reaction rate and conversion efficiency whereas smaller angles required a longer gasification zone length to reach optimum efficiency. For wood chip of moisture content 15 %, the optimum chip size was 5 cm, gasification zone length 22 – 33 cm, throat angle 610 and conversion efficiency 56 %. The tar, water vapour and ash content were 7-9 % of the producer gas output.

Downdraft gasifiers lined with ceramic material had greater life and lesser cost than downdraft gasifiers made of austenitic steel for biomass gasifier system below 500 kW. The gasifier made of steel failed at the throat and air nozzle region due to high temperature oxidation and corrosion mostly within 1500 hours of operation whereas the gasifier lined with ceramic material could withstand thermal shock.

The composition of ceramic material was alumina (50-70 %), kaolin (16-20 %), feldspar (10- 15 %) and talc (3-7 %). reduced the grate spacing and increased the grate height so that it does not block with char in downdraft throated gasifier used for gasification of wood chips of coppice willow and poplar. The fine particles of the wood chips blocked the flow of air reducing the temperature in some areas of the throat. A fuel agitator was used to stir up the wood chip to allow free fuel flow but it was ineffective. The bridging of the fine fuel particles was not solved successfully so the fuel had to be removed of these particles prior to gasification(sangeeta et al.,).

2.3.4 Gasification of Non-Woody Biomass in Throated Downdraft Gasifier

Also used the throated downdraft gasifier to gasify residues of moisture content 11.23% (d.b.).as the feed rate of BDLW increased from 2.09–5.04 kg/h, 0.08–0.21 kg/h of ash, 0.04–0.06 kg/h of tar and 0.60–1.39 kg/h of condensate were obtained the flow rates of wet gas with the increase of the fuel feed rate. The combustible gases varied between 29 and 33 % of the total wet gas produced. The most productive region for the gasification of leather residues was between 486.39 and 584.36 Nm3 m-2 h-1 of the specific gasification rate (SGR). The optimum efficiency of cold gas was 46.50 % around 486.39 Nm3 m-2 h-1 of the SGR. The large briquette sizes of the leather residues occasionally formed bridging in the throat zone of the gasifier.

Gasified sewage sludge in a throated downdraft gasifier of 10 kW

capacity and the producer gas generated had 19-23 % combustible gases and LCV of 2.55-3.2MJ/Nm3.Investigated the gasification of hazelnut shells of moisture content 12 % (d.b.) in a 5 kW throated downdraft gasifier. The optimum operation of the gasifier was between 1.44 and 1.47 Nm3/kg of air fuel ratios and 4.06 and 4.48 kg/h of wet feed rate. The high quality of producer gas of HHV (4.75 –5.15 MJ/m3) was due to high throat temperature (1015-1206 °C) and low char (0.201-0.228 kg/h) and low tar (0.023-0.025 kg/h) content. There was no sign of bridging or ash fusion at optimum throat temperature of 1000-1050 °C since the ash content in hazel nut shells was quite low i.e. 0.77 %. At higher flow rates of air (>1.5 Nm3/kg), the HHV of producer gas showed a significant decrease while tar output increased as more air resulted in more combustion. The producer gas was cleaned and cooled by passing through a packed bed scrubber and a dry filter with wood or charcoal(sangeeta et al.,)..

2.3.5 Applications of Throated Downdraft Gasifier

The producer gas from downdraft gasifier is used for thermal application like grain drying or as fuel in boiler. Used a downdraft-throated gasifier (830 MJ/h) in conjunction with a combustor for grain drying. The throat of the gasifier was made up of V shaped channels framed by triangular ducts, which were subdivided to separate primary air from secondary air. The gasifier was insulated with firebricks lined by mild steel shell. The air was preheated as it traveled through the ducts. The maize cobs having 8.1, 23.2 and 32 % moisture content were gasified with air. As the primary airflow rate increased the gasification rate increased linearly whereas as moisture content of biomass increased the gasification rate decreased. The turn down ratio also decreased with the increase in moisture content of biomass and was 3:1 at 8.1% moisture content. The particulate emission to energy input ratio was 134 mg/MJ, which was above the EPA standard for that from high capacity coal fired boilers.

The downdraft gasifier with throat is known to generate best quality producer gas with minimum tar for engines, the downdraft gasifiers with throat were commercially available in China (Fig. 5). These could operate on multi-fuels (wood, corncobs, hard nut shells, sawdust and hard coal) and the producer gas had 20-28 % CO and over 12 % H_2 content and was suitable to drive diesel engines. Used the throated downdraft gasifier for gasification of wood chips of coppice willow and poplar. The producer gas was passed through a twin cyclone, a pipe and fin cooler and a sawdust and foam filter to remove particulate matter and tar. It was mixed with air in a venturi before using in an engine to generate 30 kW of electricity(sangeeta et al.,)..

A 10 kW downdraft throated gasifier was tested for various biomass fuels after their pretreatment (size reduction and/or briquetting). The producer gas was used as a supplementary fuel in diesel engine pump set. The producer gas from the gasification of acacia auriculiformais, wood chips, fuel wood and sorghum stems replaced diesel in compression ignition (CI) engines by 70-76 %, 62.25 %, 22.5 % and 58.60 %

respectively. In downdraft throated gasifier, tamarindus and maize cobs were gasified and producer gas replaced 62.27 % and 48-52 % of diesel respectively(sangeeta et al.,).

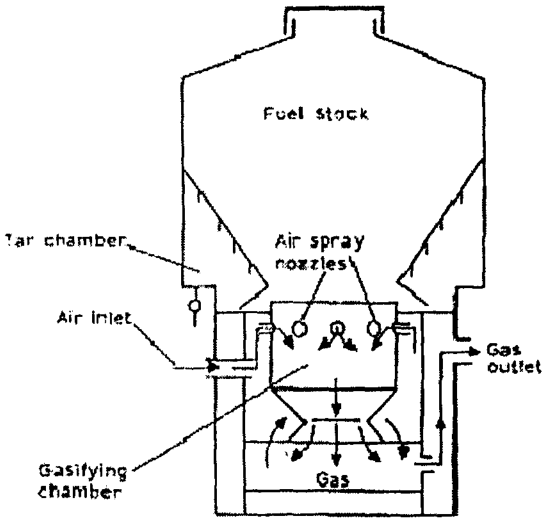


Fig. 2.5 Multifuel downdraft gasifier

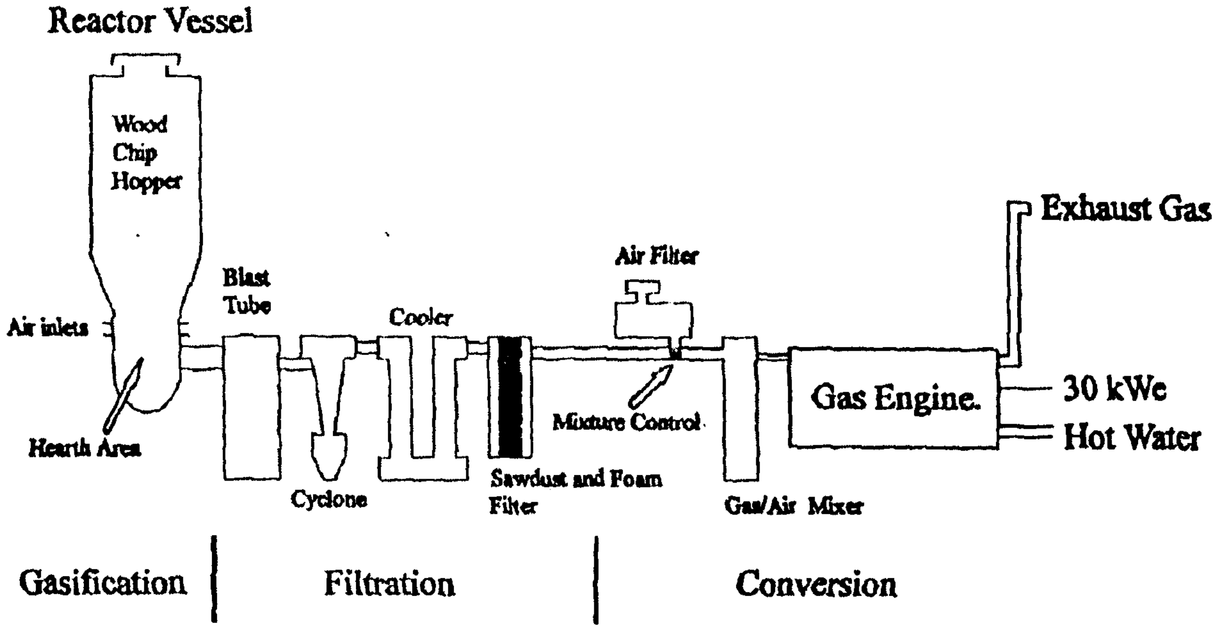


Fig. 2.6.Downdraft gasification system for engine application

Downdraft gasifier was equipped with a water sprayed gas cooling system, two-stage gas cleaning system, blower, an engine (165 hp) coupled to a 125 kVA alternator and pump sets for circulating water in the gas cooling tower and for removal of ash from the reactor. The average consumption of fuel wood and diesel per kWh of electricity generated was 0.822 kg and 0.135 l. The diesel replacement by producer gas was 59 %.

The overall efficiency of the Gosaba plant was 19 %.the use of downdraft gasifiers with efficiency of 75 % in China for gasification of straw for domestic cooking. The output of the gasifiers ranged from 60- 200 kW and the heating value of producer gas was 3.8-4.6 MJ/m3.

2.3.6 Throatless Downdraft Gasification Systems

2.3.6.1 Principle

The throatless (stratified or open top) downdraft gasifier was developed to overcome the problem of bridging and channeling in throated (Imbert) downdraft gasifiers. The gasifier consists of a cylindrical vessel with a hearth at the bottom. During operation the air and biomass move downwards through the four zones in the reactor. The open top ensures uniform access of air and permits fuel to be fed easily and uniformly, which keeps the local temperatures in control. The hot producer gas generated is drawn below the grate and up through the annulus of the reactor, where a part of the heat of the gas is transferred to the cold fuel entering the reactor, improving the thermal efficiency of the system (Reed and Das).The pyrolysis components are cracked in the oxidation zone, as gas traverses a long uniformly arranged bed of hot char without any low temperature zones, therefore the tar generated is low 0.05 kg tar/kg gas. The open top throatless gasifier is suitable for small sized biomass having high ash content up to 20 % (Stassen and Knoef, 1993, Jain 2000b and Sims 2003, Tiwari *et. al.* 2006). The gasifier is easy to construct and has good scale up properties(sangeeta et al.,).

2.3.6.2 Modifications in Design of Throatless Gasifiers

In order to improve the performance of throatless (stratified or open top) gasifier and use it for power and heat applications, changes were made in the design by allowing a varying air distribution in the gasifier, improving insulation of gasifier and re-circulating the gas within the gasifier developed an open top gasifier consisting of a vertical tubular reactor with an open top and a water seal at the bottom The lower two-thirds of the reactor was lined with a ceramic material to prevent high temperature corrosion. The upper part of the reactor was made of stainless steel with an annular jacket around it. The producer gas was drawn from below the grate and taken through an insulated pipe (re-circulating duct) to the upper annulus of the reactor where part of the sensible heat of the gas was transferred to the cold wood chips inside the reactor improving the thermal efficiency of the system. The entire reactor surface along with the re-circulating duct was insulated with aluminosilicate blankets, an open top downdraft reburn reactor, a cylindrical vessel made of mild steel, with an inner lining of ceramic. Air nozzles were provided around the combustion zone. Uniform air distribution across the section was established by locating these nozzles at two different heights. The dual air entry – from top and the nozzles – favoured a high residence time for gases at elevated temperatures, thus eliminating the tar.

The top of reactor helped in loading of fuel. The fuel bed was supported on an ash extraction screw. The screw was operated based on the ash content of biomass and / or the pressure drop across the reactor. Two discharge outlets were provided for ash extraction. Wander (2004) developed an open top stratified gasifier with internal gas recirculation, which could burn a part of the gas produced to raise the gasification temperature. The gasifier (Fig. 7) consisted of a cylinder with a cast iron grate fixed to a rotating shaft. Rods were fixed on the shaft to mix sawdust in the reduction zone and to extract the ash. In the center of the gasifier a device like a venturi aspirated part of the gases produced in it to be burnt in a chamber. The gasifier was used for gasification of 12 kg/h of sawdust of moisture content (9- 11 %). With no recirculation of gases, the air / sawdust ratio had to be kept above 1.5 to get cold gas efficiency of 60 %, whereas with re-circulation of gases, the ratio was lower i.e. 1.1 – 1.4. Gas re-circulation raised the gasification reaction temperature, helped in burning a part of the tar and improved efficiency of gasifier. Strong mixing of sawdust was required to avoid bridging and channeling. The gasifier was used with a cyclone to remove particulate matter from producer gas(sangeeta et al.,).

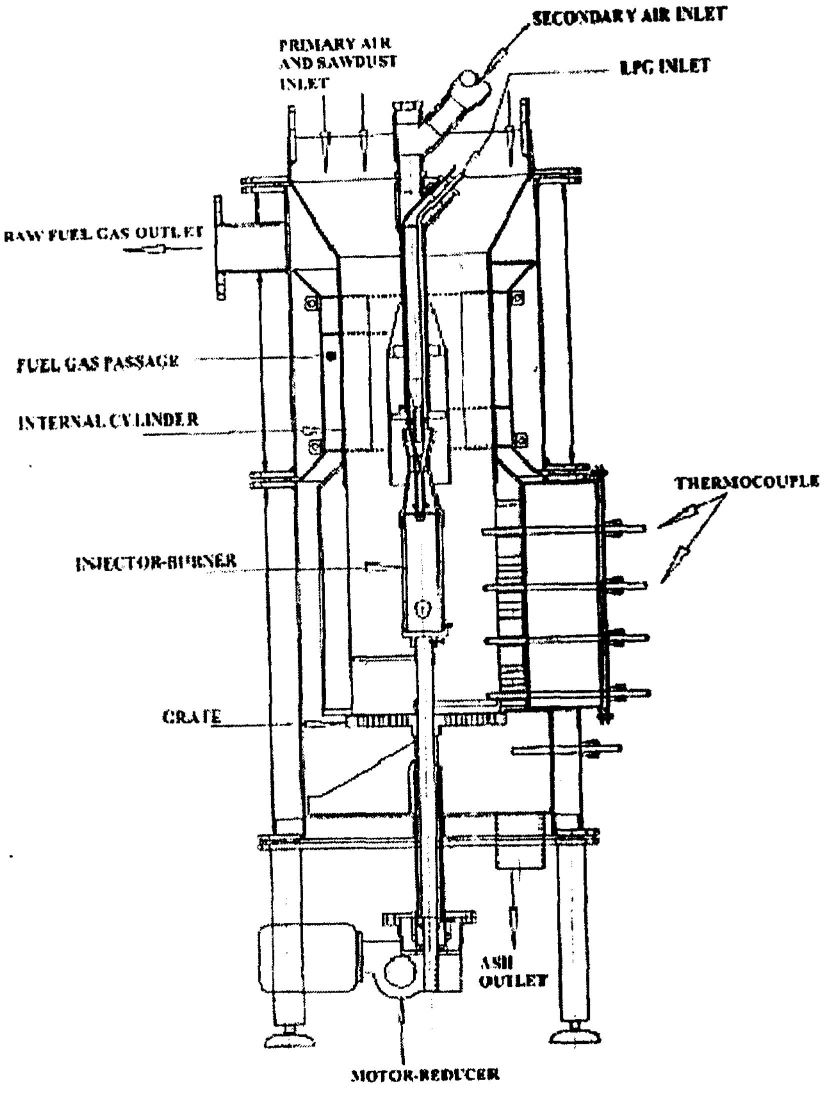


Fig. 2.7 Stratified gasifier with gas re-circulation

In another trial on the open core gasifier with gas re-circulation conducted by Altafin (2003) the feed rate of sawdust, temperature and pressure in the gasifier was 11.34 kg/h, 800 °C and 0.93 bar respectively. The moisture content of sawdust was varied from 0 – 30 %. It was observed that as the moisture content of the sawdust increased the air/sawdust ratio had to be increased to keep the temperature constant. The LHV of the producer gas reduced with the increasing moisture content of sawdust. Maximum cold gas efficiency (68 %) was achieved with sawdust having moisture content of 20 %.

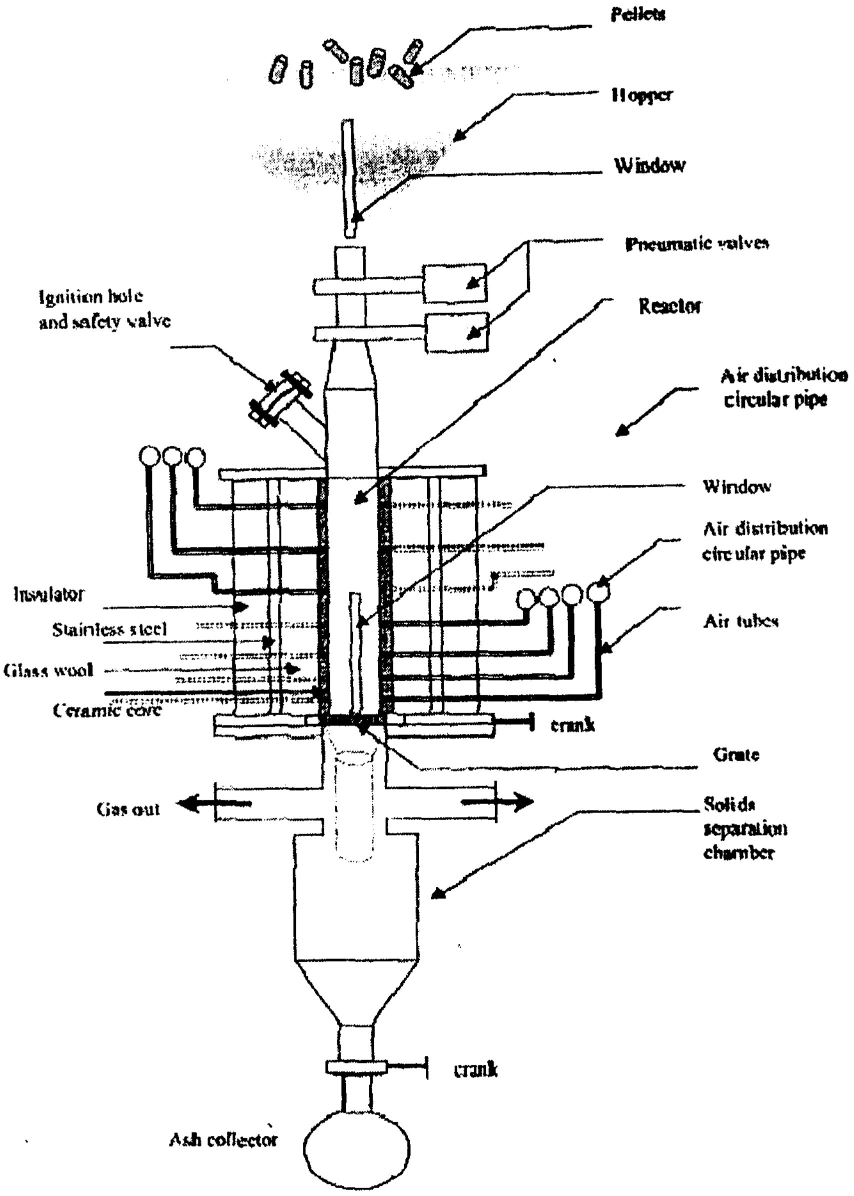


Fig. 2.8 Stratified downdraft gasifier

Barrio,, (2001) used a small-scale (30 kW) stratified downdraft gasifier (Fig. 8) to gasify wood pellets at a feed rate 5 kg/h. The design of the gasifier allowed for variation in the point of air injection along the length of gasifier. The grate was a perforated plate with a crank, which could be shaken manually. The equivalence ratio was lower (0.3) when air was taken in from the top (80 %) and sides (20 %) of the gasifier. It was 0.4-0.45 when 100 % air was taken in from the top (traditional open core). The gasifier produced 12 Nm^3/h of producer gas with a calorific value of 5 MJ/Nm3 and CO and H2 content of 20% each(sangeeta et al.,).

2.3.7 Optimization of Operating Conditions of Throatless Gasifiers

The specific gasification rate (SGR) i.e. rate of fuel consumption per unit area of reactor area has been optimized for throatless gasifiers. Knowing the optimal value of SGR, the size of the reactor can be computed from the energy demand of the gasification system. Gasified rice hulls in an open core or static bed gasifier of varying reactor diameters from 16-30 cm. The specific gasification rate was varied in the range of 100-400 kgm-2h-1. As the specific gasification rate (SGR) increased the producer gas flow rate and reaction airflow ratio increased. The cold gas efficiency increased up to an optimum value of SGR and then decreased. The highest cold gas efficiency was between 50-60 % at SGR of 200 kgm-2h-1. It was observed that the cold gas efficiency and optimum value of SGR were independent of the size of the reactor thus indicating that SGR could be used satisfactorily to upscale the reactor size(sangeeta et al.,).

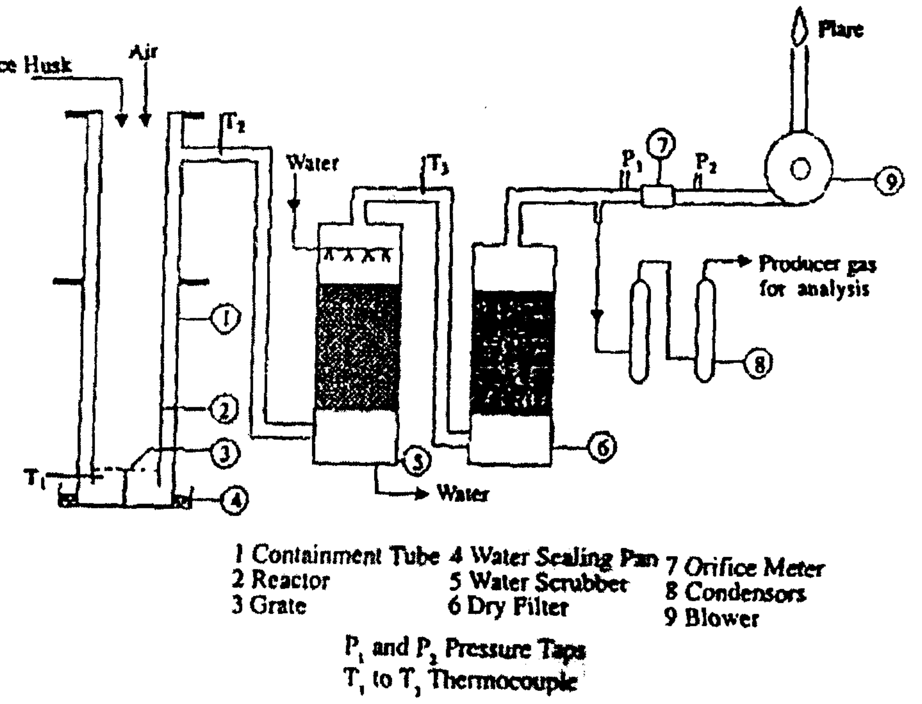


Fig. 2.9 Throatless gasifier for gasification of rice husk

Jain (2000) developed the design parameters for rice husk throatless gasifier for engine applications. The gasifier was made of two concentric cylinders of mild steel the inner one being reactor and the outer the containment tube. The grate was made of stainless steel wire mesh. The gas was passed through packed bed water scrubber and dry filter (Fig.2.9).

SGR was varied in the range 110-240 kgm-2h-1. As the SGR increased the equivalence ratio increased from 0.33 to 0.44. The cold gas efficiency increased from 53 to 69 % up to SGR of 190 kgm-2h-1. Further increase in SGR reduced the efficiency. The optimum value of SGR, at which the gasification efficiency was maximum, was 190 kgm-2h-1. Jain and Goss (2000) conducted gasification of rice husk in open core throatless batch fed gasifiers of four different diameters to develop the up scaling parameters for gas production. For all the reactors the gasification efficiency was the maximum (65 %) between SGR as 190-195 kgm-2h-1. The specific gas production rate (SGPR) was 410-429 m3h-1m-2 and equivalence ratio was 0.4. The LHV of producer gas was 4 MJNm-3. The optimum SGR and specific heat rate for 100 kg/h rice husk gasifier system was 200 kgm-2h-1 and 2.7 GJ m-2h-1 respectively (Jain, 2006).

Singh (2006) carried out the gasification of cashew nut shells in an open core downdraft gasifier of capacity 150,000 kcal/h at varying gas flow rates. The maximum gasification efficiency was 70 % at SGR of 167 kgm-2h-1. The producer gas flow rate was 130m^3/h and its LHV was 1081 kcal/m3. The producer gas was used for thermal applications like boilers through direct combustion or for heating air using heat exchangers.

2.3.8 Applications of Throatless (stratified or open top) Gasifiers

biomass gasification system for thermal applications at Nimbkar Agricultural Research Institute (NARI), India. The reactor was lined by high temperature resistant firebricks on the inner side. The gas conditioning system consisted of high temperature char/ash coarse settler and a high efficiency cyclone separator. The gasifier was operated on sugarcane leaves, bagasse and their mixture and the HHV of the gas was between 3.56-4.82 MJ/Nm3. The system also produced char, (with gross calorific value of 18.9 MJ/kgl) which was about 24 % by weight of the original fuel. The output was in the range of 288-1080 MJh-1. The temperature of the gas was greater than 300 °C and so there was no condensation of tars and particulate matter. The performance of the gasifier was excellent up to fuel moisture content of 15 %(sangeeta et al.,).

Dasappa, (2003) used throatless reactors of capacity 500 kg/hl and 300 kg/hl were used for low (100 °C) and high (600 °C) temperature applications respectively. The low temperature application was drying of marigold flowers for which the coconut shells were gasified and gas from the gasifier system replaced diesel fuel in the range of 125–150 l/hr.

The high temperature application was for a heat treatment furnace in which biomass gasifier replaced 2000 of diesel per day. The gas quality after cleaning and cooling through a cyclone and water scrubber was comparable to that which could be used in engines. Mukunda, (1994) reported that producer gas from throatless gasifier could be used for engine (power) applications as it contained low amount of tar. The gasifier was used for gasification of wood chips. The ratio of the cold gas flow rate to wood chips consumption rate was about 2.6. The composition of gas was 18 % H_2, 19 % CO, 1.25 % CH_4, 12 % CO_2 and N_2. The calorific value ranged from 4 MJ/kg to 4.4 MJ/kg. The amount of tar and particulate matter in producer gas were 100 mg/m3 and 700 mg/m3. They were reduced to 20 mg/m3 and 50 mg/m3 respectively after cleaning through a sand filled coarse filter and a fine filter. The total efficiency of the system measured in relation to the final electric energy was 27 % in a 100 kW compression ignition engine with 85 % diesel replacement(sangeeta et al.,)..

Sardar Patel Renewable Energy Research Institute (SPRERI), India developed a 3.5 kW throatless gasifier for gasification of paddy husk and the producer gas was used in a diesel engine pump set. The average diesel replacement and specific husk consumption was 61.1 % and 9.2 kg/h respectively. Punjab Agricultural University, India developed a 10 kW throatless downdraft paddy husk gasifier coupled to a diesel engine. The gasifier efficiency, tar content and diesel replacement were 67 %, 75 mg/Nm3 and 75 % respectively at 8 kW load (Jain, 1996).

Leung, (2004) reviewed the open core gasifiers (60-200 kW) being developed in China for electricity generation. These operated on rice husk and produce producer gas of lower heating value 3.8-4.6 MJ/m3 with efficiency of 50 %.

Dasappa(2004) tested the open top ceramic lined reactor for gasification of biomass with varying moisture contents up to 37 % and studied the variation of tar and particulate in the raw gas. The tar content in this reactor design was lower due the high quality insulating material used for reactor and air distribution between the nozzle and the reactor top. The cracking of the tars improves the overall gasification efficiency. With increase in moisture content, the tar level in the gas increased and particulate matter reduced. The tar content in the raw gas was in the range of 50–250 mg/Nm3 for fuel having moisture content less than 15 %, beyond which it increased to about 700 mg/Nm3. At a capacity of 75 kg/h the cold gas efficiency was around 75 %, whereas the large capacity gasifier system of 650 kg/h resulted in cold conversion efficiencies in the range of 85 %. The fraction of tar and particulate matter in the hot gas from open top reactor were observed to be lower than in the Imbert design. The gasification system consisted of the gasifier, high efficiency cyclone, and ejector scrubber to cool and clean gas before using it in the engine. Diesel savings of up to 70 % and an overall efficiency of 20 % were achieved(sangeeta et al.,).

2.3.9 Two Stage Gasifier System

2.3.9.1 Principle

In contrast to the updraft, throated and throatless downdraft gasifiers where pyrolysis and gasification zones occur in the same chamber, these zones were separated in the two stage gasifier which improved control of the process temperatures resulting in markedly lower tar production and a high energy efficiency.Experts have estimated that for internal combustion engine applications, the tar level in producer gas should be less than 30-50 mg/Nm3 (Milne and Evans, 1998) to avoid problems associated with tar condensation on critical engine parts. Stassen and Knoef (1993) while conceding that some engine fouling is inevitable when fuelled by producer gas suggested that up to 100 mg/Nm3 is acceptable and less than 50 mg/Nm3 is preferable. The tar content was reported to range from 2 g/Nm3 in a conventional downdraft gasification of wood to 58 g/Nm3 in conventional updraft gasifiers investigated that the two-stage gasifier was very effective in producingclean gas. The concept of this design was to separate the pyrolysis zone (first stage) from the reduction zone (second stage). The gasifier has two levels of air intakes, primary air supply at the top section and secondary air at the middle section of the gasifier. The high temperature achieved in the second stage due to the addition of a secondary air helps in reducing the tar level to a considerably lower value. The two-stage gasifier resulted in gas having tar content about 50 mg/m3, about 40 times less than a single-stage reactor under similar operating conditions. However most of the tars were formed during the warm-up period. This could be avoided by filling the gasifier with a bed of char, which almost totally eliminated tar formation during start-up in the reactor(sangeeta et al.,).

2.3.9.2 Modifications in Design of Two Stage Gasifier System

The modification in the design of two stage gasifier was done in such a way that char was produced inside the gasifier itself, thus avoiding input of external char. For this purpose, an extra air inlet above the original primary air supply was added. The tar content in producer gas reduced with increase in the flow rate of primary and secondary air. The lowest value of tar was 45 mg/Nm3 at primary and secondary air flow rates of 120 lmin-1. The gas composition also improved at higher airflow rate because of high temperatures in the gasifier.

A two-stage gasifier designed at the Technical University of Denmark was a combination of pyrolysis of the biomass feed with subsequent partial oxidation of the volatile products in presence of a char bed. Used the reactor (Fig.10) of 100 kW capacity for gasification of wood chips. The char and the volatile pyrolysis products from the pyrolysis unit entered the top of the gasification unit where the gases were mixed with the preheated steam and air starting partial combustion. The gases were allowed to pass through the bed of char resulting in significant tar reduction (15 mg/m3). Lower amounts of tar were attributed to partial combustion of the pyrolytic gases as well as the catalytic effect of the char bed. The gas cleaning system consisted of a hot cyclone, gas cooler, venturi scrubber and a gas filter(sangeeta et al.,).

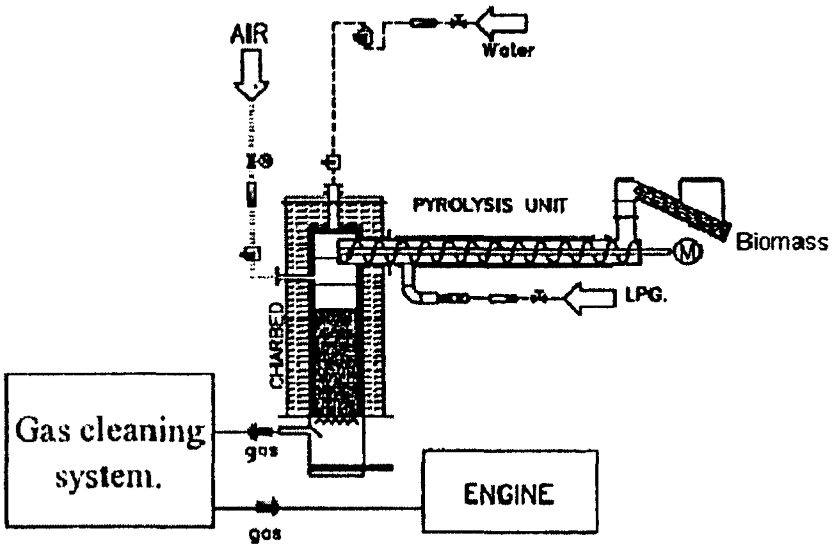


Fig. 2.10 Two-stage gasifier

A two-stage gasifier of 100 kW consisting of anexternally heated pyrolysis unit and a downdraft char gasifier for gasification of wood chips.

The pyrolysis occurred at 600 °C and gasification at 1000 °C to 800 °C. The gasifying media was preheated steam air mixture. The gas was cleaned by venturi scrubber, demister (a steel sponge) and a filter. Under biomass: steam ratio of 1:1, the particle load was about 300 g/Nm3, and under ratio 3:1, the particle load increased above 1000 g/Nm3. The pressure drop through char bed was less when fuelled by briquettes than when fuelled by wood chips. Proposed gasification of bagasse pellets of moisture content of about 11 % in a two-stage gasification reactor, which allowed for complete gasification avoiding the formation of carbon and tar. The reactor consisted of two stages, separated by the biomass feeder. Stage one was a reactor with a frit on the bottom side working as an updraft gasifier. Stage two was a fixed bed (high surface alumina or nickel on alumina) reforming reactor. The reactor was preheated and the feed was introduced in the reactor. It underwent immediate pyrolysis, and volatile component and char were formed. The char fell to the bottom of first stage, which reacted with the oxygen-steam (gasifying media) introduced through the frit. The gaseous mixture produced in stage one flowed up to stage two. In the second stage tar and heavy hydrocarbons were trapped in the bed, increased their residence time allowing for complete gasification and, if a reforming catalyst was used, also for complete reforming. The feed flow rate of bagasse was maintained constant at 0:80 g/min. The decrease in steam caused decrease in H_2, and increase in CO_2 content in producer gas. The increase in temperature of stage one caused increase in CO and in CH_4 to a lesser extent. The presence of Ni (reforming and water shift catalyst) in stage two increased the reaction rate and CO and H_2 in the producer gas(sangeeta et al.,).

2.3.10 Crossdraft Gasification System

In this gasifier the fuel moves downwards while air is introduced on one side of the reactor and producer gas at 800-900 °C removed from the opposite side at the same level. Ash is removed from bottom. The overall energy efficiency is low and the tar content in producer gas is high and is mostly used in close-coupled boilers. A number of these gasifiers have been decommissioned in Brazil and South-American countries, as the fuel of acceptable quality could not be produced on a sustainable basis. The tar cracking capability is limited therefore the gasifier is suitable only for low-tar fuels. Thus large-scale implementation of the cross draft gasifiers has not been done (Clarke et al., 1981)

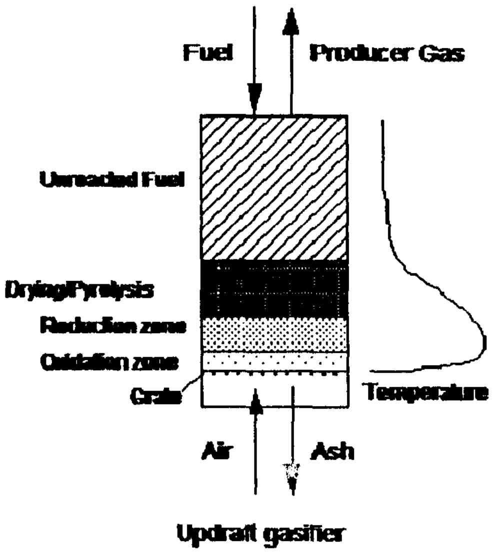


Figure 1 Schematic of fixed-bed updraft gasifier [5]

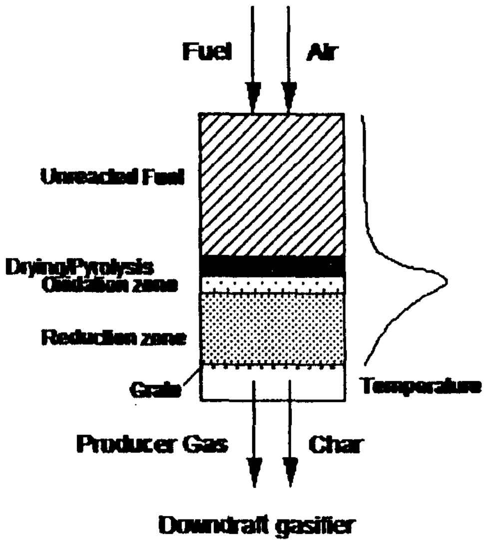


Figure 2 Schematic of fixed-bed downdraft gasifier [5]

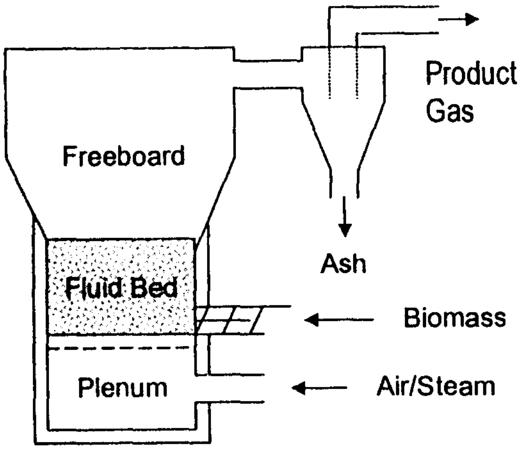


Figure 3 Bubbling bed reactor [5]

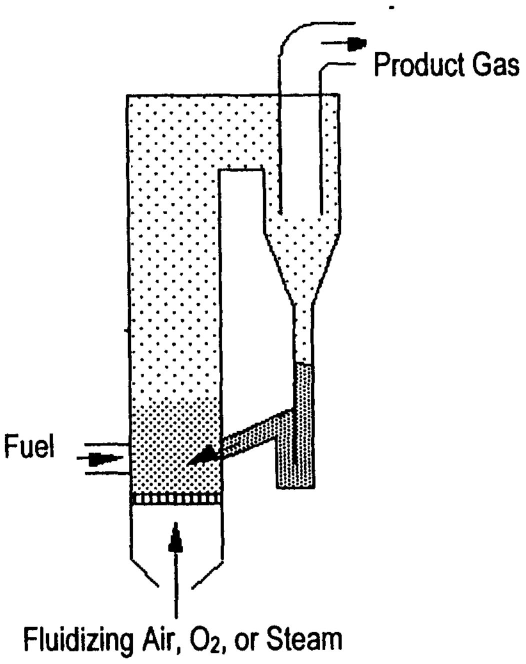


Figure 4 Circulating fluidized bed reactor [5]

FIGURE 2.11

2.3.11FLUDIZIED BED

Fluidized bed reactors contain a bed of relatively small particles of inorganic material (often sand or small diameter ceramic beads or gravel). The bed is 'fluidized' by blowing hot oxidant up from the bottom (individual particles are lifted by aerodynamic drag, and become suspended or entrained on the gas stream at velocities for which the drag force becomes equal to or exceeds the gravitational force or weight). When fluidized, the bed behaves much like a liquid. When the bed media is hot enough, biomass is injected either into the bed or onto the surface (bubbling beds) and can begin to combust or gasify depending on the amount of oxygen available.

Bubbling fluidized bed reactors have relatively slow velocity air, oxygen, or steam flow and therefore have lower particle entrainment in the gas leaving the reactor. The bed material is concentrated in the lower dense-bed region because the freeboard section above the bed has a larger diameter and lower gas velocity. The gas velocity in the freeboard section is too low to continue to suspend bed particles, which fall back into the bed region. The design is simple but has lower capacity and potentially less uniform reactor temperature distribution than circulating fluidized beds.

The circulating fluidized bed uses higher gas velocities but offers higher conversion rates and efficiencies. Instead of a freeboard section, the reactor diameter remains essentially constant, which keeps bed and fuel particles suspended. The bed material flows up with the fluidizing gas and is carried over into a cyclone which separates most of the particles from the gas stream which are re-injected (recirculate) into the lower part of the bed. Ideally, the fuel particles are small enough to completely react before carried over into the cyclone, but in practice large fuel particles recirculate with bed media until small and light enough to be carried out with the product gas exiting the cyclone or other separation device. Oxygen fired circulating fluidized bed gasifiers are candidates for the production of hydrogen and liquid fuels.

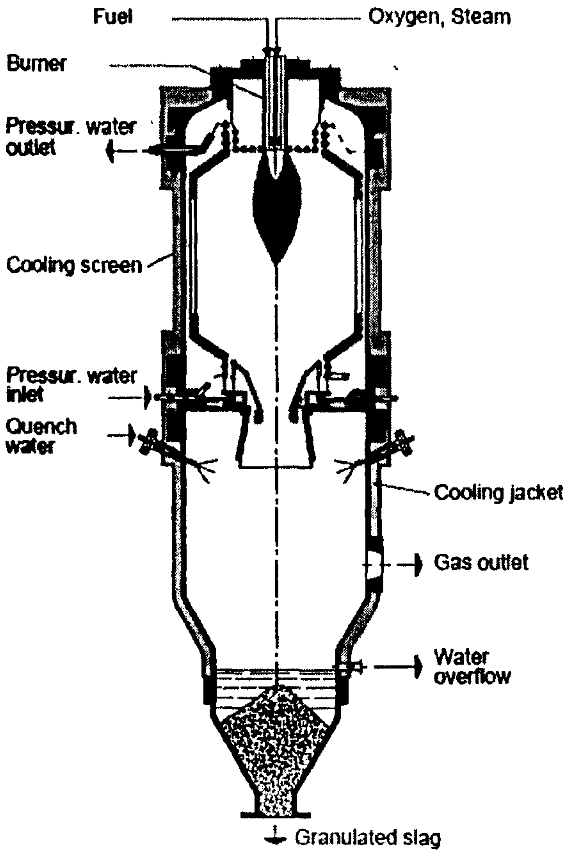


FIGURE 2.12.Schematic of an entrained flow gasifier [6]

2.3.12 ENTRAINED FLOW

Entrained flow gasifiers are used extensively by the petroleum to convert petroleum residues (e.g., petroleum coke) to useful products and energy. Most coal gasification is done with entrained flow systems.

Entrained flow gasifiers have high gas velocities and high material throughput. Consequently, time for reaction (residence time) is short which requires the feedstock to be of very small particle size, a liquid or liquid slurry. The systems are generally oxygen blown and can be pressurized or atmospheric. High temperature (>1250 °C) is generated from combustion in oxygen which melts the ash (sometimes called slagging gasifier) and requires reactor cooling. Little to no tar is formed as the feedstock is essentially completely converted to H_2, CO, CO_2, and H_2O.

2.4 Syngas Composition

Table 2.1 presents syngas compositions for a number of the biomass gasification technologies examined in this survey. These compositions were cited as being from existing commercial applications or based on large-scale process development units. A large number of parameters influence composition, including feedstock, pressure, temperature and oxidant. Quite a few biomass gasification studies failed to report the content of tar and other impurities in the syngas. At the operating temperatures reported, significant quantities of methane, higher hydrocarbons and tar can be expected. Due to the higher operating temperatures used in coal gasification, coalderived syngas contains essentially no methane or other hydrocarbons and tar. However, since coal usually contains sulfur and nitrogen, significant quantities of H2S and NH3 are

present in the raw syngas.

2.5 AIR AND STEAM GASIFICATION

The composition of gasification products is directly related to gasifying agent properties of temperature and chemical composition (oxygen, steam and other gases). The gasification characteristic of biomass fuels examined can be observed from graphical record of gasifying agent temperature and species concentration. The concentration of desired species, such as hydrogen and carbon monoxide, is significantly increased with increase in gasifying agent temperature. Fig. 2 shows the effect of gasifying agent temperature on the concentration of H_2 and CO from several different biomass wastes using air/steam mixture as the gasifying agent. The results show strong effect of gasifying agent temperature on the amounts (concentration) of gasification products produced. For all biomass examined, H2 and CO concentration diminishes at low temperature (540O^{0}C) gasification. Significant increase in gasification product yield for both H_2 and CO is observed with temperature increase from 540O^{0}C to 1000O^{0}C for all the gasification wastes. The small differences with steady increase in temperature could be due to some experimental uncertainty in complex waste-gasifying agent interactions. None the less these results show good agreement with the equilibrium calculations for cellulose presented by the authors. The equilibrium calculation of cellulose shows that concentration of species, such as hydrogen and carbon monoxide increases significantly with increase in gasification temperature up to a certain temperature.

Experimentally there appears to be a limiting temperature of about 1200 K after which H_2 concentration does not change considerably. However, the CO continues to increase with increase in gasifying agent temperature. This is attributed to the different energies required for the dissociation of these two species. The experimental trend of gasification products at higher gasification temperature (> 930OC) for rice husk, rice straw and corn cob showed increased production of both H_2 and CO, thus showing good agreement with the equilibrium calculations.

From the experimental results conducted here in this study, the results showed that the overall gasification characteristics of the biomass wastes is similar and depends on fundamental composition of the fuel. However, the specific characteristics of each biomass have some differences. From the experimental results it can be seen that cellulose has a limiting temperature (near to 930oC) beyond which the H_2 production decreases. This is attributed to the increased dissociation of H_2 with increase in temperatures so that the H atom produced is attracted by O atom to form OH radicals in the post reaction zone. This point requires further examination.

Gasification products produced from all three biomass wastes examined showed the same trend for the range of temperatures examined. The yield of H_2 and CO in the gasification products from corn cob is highest, followed by rice husk (low) and rice straw (lowest). These results show good general agreement with the amount of fixed carbon in biomass waste, see Table 1. Corn cob has the lowest amount of fixed carbon (5.3 %) while the rice husk and rice straw has 21.46 % and 21%, respectively(Woranuch, et al.,2006).

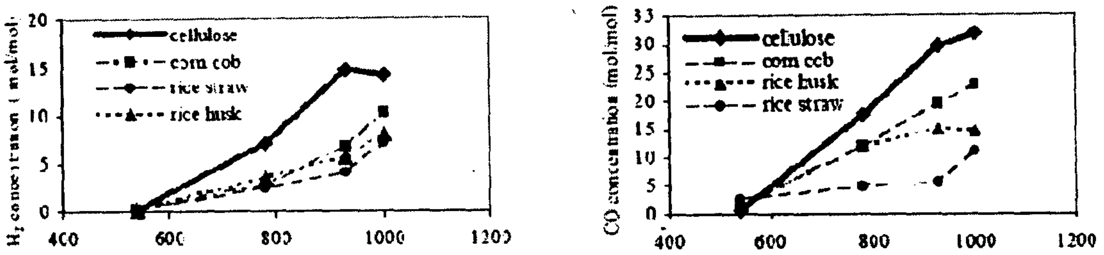
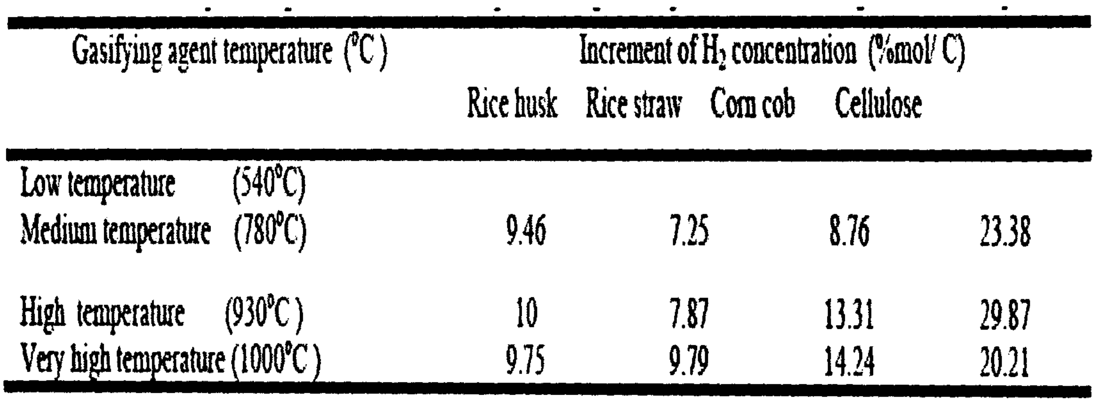


Fig. 2.13 Concentration of H_2 and CO at various gasification temperatures with air/steam mixture as the gasifying agent

The increase in percent H_2 concentration per unit degree increase in gasification temperature is now compared by using the lowest gasification agent temperature as the datum. The effect of gasification temperature for each step increase in temperature level is shown in Table 2. The gasifying agent temperature in this study is grouped into four categories based on the magnitude of temperature examined: low temperature (540°C), medium temperature (780°C), high temperature (930°C) and very high temperature (1000°C). The results show that H2 concentration continuously increases for all the biomass waste examined, including cellulose, when gasified at medium and high temperatures. In contrast very high gasifying agent temperature does not show significant differences in the amounts of H_2 produced for all the biomass examined. Thus H_2 production per unit degree for each level of temperature examined can be considered as a good indicator to determine the optimum temperature(Woranuch, et al.,2006)

TABLE.2.3 Increase in percentage of H_2 concentration per unit degree temperature increase in gasification temperature[3]



Gasifying agent temperature (°C)	Increment of H_2 concentration (%mol/°C) Rice husk	Rice straw	Corn cob	Cellulose
Low temperature (540°C)				
Medium temperature (780°C)	9.46	7.25	8.76	23.38
High temperature (930°C)	10	7.87	13.31	29.87
Very high temperature (1000°C)	9.75	9.79	14.24	20.21

CHAPTER THREE

3.0 MATERIAL BALANCES

3.1 Introduction

Material balances are the basis of process design. A material balance taken over the complete process will determine the quantities of raw materials required and products produced. Balances over individual process units set the process stream flows and compositions.

Good understanding of material balances calculations is essential in process design. Material balances are also useful tool for the study of plan operation and troubleshooting. They can be used to check performance against design; to extend the often-limited data available from the plant instrumentation; to check instrument calibration; and to locate sources of material losses (Sinnot, 1999).

3.2 Basis of Material Balances (Conservation of Mass)

The general conservation equation for any process system can be written as:

Material out = Material in + Generation – Consumption – Accumulation

For a steady-state process, the accumulation term will be zero. Except in nuclear processes, mass is neither generated nor consumed; but if a chemical reaction takes place, a particular chemical species may be formed or consumed in the process. If there is no chemical reaction, he steady-state balance reduces to

Material out = material in

A material balance equation can be written for each separately identifiable species present, elements, compounds or radicals; and for the total material.

3.3 Material Balance Calculations

The material flow of the feed stock is chosen such that the efficiency of the process is optimized. Given the dual feed base, CORN COB and RICE HUSK, the percentage (%) of each must be controlled and kept within a certain range due to the difference in their properties and composition. Data from Table 7 and 9, for corn cob and rice husk are as follows;

Table 7 abridge; slagging of agricultural residues in small laboratory downdraft gasifier

FUELS	ASH CONTENT	DEGREE OF SLAGGING
RICE HUSK	14.90	SEVERE SLAGGING
CORNCOB	1.50	NO SLAGGING

Table 9 abridge; proximate analysis

FUELS	MOISTURE %	PROXIMATE ANALYSIS		
		VOLATILES	F.C	ASH
RICE HUSK	8.2	64.19	21.40	14.38
CORN COB	8.24	87.90	5.30	6.79

Slagging posses a serious problems to current available gasifier type and technology, and thus must be guided against.

Table 7 shows that for the generality of agricultural waste, slagging is observed with ASH content above 5.8%. It is also observed that corn cob does not produce slag while rice husk produce very high quantity of slag.

The composition of the two materials at the feed point of the gasifier is chosen such that the overall ash content is kept below 5.8, so as to minimize and possibly eliminate slag formation.

With 70% corn cob and 30% rice husk, we obtain an ash content of.

$$0.7(1.5) + 0.3(14.9) = 5.52$$

This ash content is suitable as it presents no problem of slagging. This ratio will be employed in the proximate analysis of the combined feed.

Converting Table 7 to wet basis and making adjustment on the variable value of F.C for corn cob such that ash content for corn cob remains 1.5% wet basis so as to meet balance, we have

FUELS	MOISTURE %	PROXIMATE ANALYSIS (WET BASIS)		
		VOLATILES	F.C	ASH
RICE HUSK	7.58	59.31	18.22	14.90
CORN COB	7.61	81.22	9.77	1.5

With 70% corn cob and 30% rice husk, the combined feed will have a composition;

FUELS	MOISTURE %	PROXIMATE ANALYSIS (WET BASIS)		
		VOLATILES	F.C	ASH
RICE HUSK/CORN COB	7.60	74.65	12.31	5.52

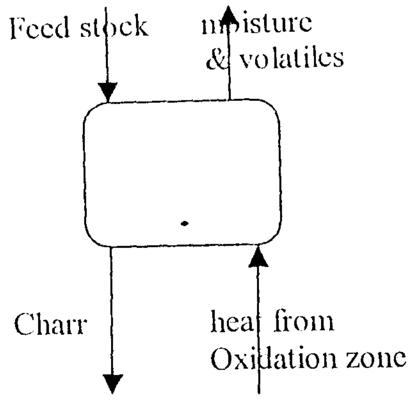
BASIS: 100kg/hr feed stock containing 70% corn cob and 30% rice husk

The type of gasifier chosen for this design is a downdraft gasifier is made of four (4) operating zones, namely; Drying zone, Pyrolysis zone, Oxidation (combustion) zone and the Reduction (gasification) zone.

The design chosen here is such that the drying and pyrolysis zones are in a compartment at the top of the column. The overall operation of the column, though continuous, may be viewed as semi-batch, due to the compartmentalization of the unit. The feeding at the top of the column (drying and pyrolysis zones) is done in such a way that a batch of feed is charged into the top of the column and allowed sufficient time to dry and pyrolyze before it is released into the lower section of the column (oxidation and reduction zones), via the use of retractable valves. The operation in the lower section of the column is essentially continuous.

Material Flow across the Gasifier

DRYING ZONE



The material balance for the drying and pyrolysis are taken together:

Assumption

1. All the moisture and volatile content is removed, either at the top or as a side stream close to the top.
2. No chemical reaction takes place, so that all the heat supplied is used to provide kinetic energy for vaporization of moisture and volatiles.

	kg/hr	kg/hr	kg/hr
MOISTURE	7.6		7.6
VOLATILES	74.65		74.65
CHARR	12.31	12.31	
ASH	5.52	5.52	
	100.08	17.83	82.25

The char in the table is C&Cγ where C represents carbon content of the char that is burnt using oxygen in the combustion zone to yield CO_2 and CO

Mass flow rate of Char into the combustion =12.31kg/hr

Mass of ash = 5.52 kg/hr

It must be noted that the ash is only formed after the oxidation and reduction process, it is however represented here as an independent and non-reacting component for the sake of convenience.

The molar flow rate of char (carbon) is given by;

$$F_{charr} := \frac{12.31\frac{kg}{hr}}{12.01\frac{kg}{mol \cdot 10^3}}$$

$$F_{charr} = 0.285 s^{-1} \cdot mol$$

The ratio of carbon(C) combusted in the oxidation zone to the carbon (Cγ) that goes down to the reduction zone is set by inspection at, 1:1.5

The flow rate of carbon(C) is given by;

$$F_c := \frac{1.0}{2.5} \cdot F_{charr}$$

$$F_c = 0.114 s^{-1} \cdot mol$$

The molar flow rate of carbon (Cγ) is given by;

$$F_{c\gamma} := \frac{1.5}{2.5} \cdot F_{charr}$$

$$F_{c\gamma} = 0.171 s^{-1} \cdot mol$$

$$\rho_c := 1642.06 \cdot 10^{-3} \frac{kg}{L} \qquad MM_c := 12.01 \frac{kg}{mole \cdot 10^3}$$

$$C_c := \frac{\rho_c}{MM_c}$$

$$C_c = 1.367 \times 10^5 m^{-3} \cdot mol$$

$$\rho_{c\gamma} := 1642.06 \cdot 10^{-3} \frac{kg}{L} \qquad MM_{c\gamma} := 12.01 \frac{kg}{mole \cdot 10^3}$$

$$C_{c\gamma} := \frac{\rho_{c\gamma}}{MM_{c\gamma}}$$

$$C_{c\gamma} = 1.367 \times 10^5 \frac{mol}{m^3}$$

$$\rho_{O2} := 0.21994 \cdot 10^{-3} \frac{kg}{L} \qquad MM_{O2} := 32 \frac{kg}{mole \cdot 10^3}$$

$$C_{O2} := \frac{\rho_{O2}}{MM_{O2}}$$

$$C_{O2} = 6.873 m^{-3} \cdot mol$$

$\rho_{CO2} := 0.30248 \cdot 10^{-3}\frac{kg}{L}$ $\qquad MM_{CO2} := 44.01\frac{kg}{mole \cdot 10^{3}}$

$$C_{CO2} := \frac{\rho_{CO2}}{MM_{CO2}}$$

$C_{CO2} = 6.873\,m^{-3} \cdot mol$

$\rho_{CO} := 0.19252 \cdot 10^{-3}\frac{kg}{L}$ $\qquad MM_{CO} := 28.01\frac{kg}{mole \cdot 10^{3}}$

$$C_{CO} := \frac{\rho_{CO}}{MM_{CO}}$$

$C_{CO} = 6.873\,m^{-3} \cdot mol$

$\rho_{N2} := 0.19253 \cdot 10^{-3}\frac{kg}{L}$ $\qquad MM_{N2} := 28.013\frac{kg}{mole \cdot 10^{3}}$

$$C_{N2} := \frac{\rho_{N2}}{MM_{N2}}$$

The ratio of carbon(C) combusted in the oxidation zone to the carbon ($C\gamma$), that goes down to the reduction zone is set by inspection at, 1:1.5

The reactions in the oxidation zone are

$\left(C + 1.5C^{\gamma} + O_2 + 3.76N_2 \rightarrow\right) CO_2 + 1.5C^{\gamma} + 3.76N_2$ ....

$\left(C + 0.5O_2 + 1.5C^{\gamma} + 1.88N_2 \rightarrow\right) CO + 1.5C^{\gamma} + 1.88N_2$ ....

The basic assumption here is that 80% of carbon(C) is involved in reaction 1, while the remaining 20% goes into reaction 2. Thus, for reaction 1;

$F_{c1} := 0.8 \cdot F_c$ $\qquad F_{c\gamma 1} := F_{c\gamma} \cdot 0.8$

$F_{c1} = 0.091\,s^{-1} \cdot mol$ $\qquad F_{c\gamma 1} = 0.137\,s^{-1} \cdot mol$

$\left(C + 1.5C^{\gamma} + O_2 + 3.76N_2 \rightarrow\right) CO_2 + 1.5C^{\gamma} + 3.76N_2$

From the reaction stoichiometry we obtain the required amount of oxygen required and subsequently the amount of air required.

The ratio of C to O_2 in the stoichiometry is 1:1, and for each 1 mole of O_2 in Air there is an associated 3.76 moles of N_2, thus;

$F_{O2.1} := F_{c1}$ $\quad F_{O2.1} = 0.091 mol s^{-1}$

$F_{N2.1} := 3.76 \cdot F_{O2.1}$ $\quad F_{N2.1} = 0.343 mol s^{-1}$

5% excess O_2 is supplied to optimize the combustion process thus

$F_{O2.1} := 1.05 \cdot F_{O2.1}$ $\quad F_{O2.1} = 0.096 mol s^{-1}$

$F_{N2.1} := 1.05 \cdot F_{N2.1}$ $\quad F_{N2.1} = 0.36 mol s^{-1}$

$$R \equiv 8.314 \frac{J}{mole \cdot K}$$

$$A := 12 \frac{L}{mole \cdot s}$$

$$E_a := 0.9 \cdot 10^4 \frac{J}{mole}$$

$$T := (1500 + 273)K$$

$$k_1 := A \cdot e^{\left(\frac{-E_a}{R \cdot T}\right)}$$

$$k_1 = 6.517 \times 10^{-3} m^3 s^{-1} mol^{-1}$$

Let CA_O be the initial concentration of carbon (C) in the feed stream

$$\rho_{N2} = 0.193 kg\, m^{-3}$$

Total moles of reactant = 1+1+1.5+3.76=7.26

$$C_{A0} := \frac{C_c}{7.26}$$

$$C_{A0} = 1.883 \times 10^4 mol\, m^{-3}$$

$$X_A := 0.99$$

$$\xi_A := \frac{6.26 - 7.26}{7.26}$$

$$C_A := C_{A0} \cdot \left(\frac{1 - X_A}{1 + \xi_A \cdot X_A}\right)$$

With equimolar concentration of reactant A and B (C and O_2) the simple rate expression is given as:

$$r_A := -k_1 \cdot C_A^{\,2}$$

$$\tau := C_{A0} \cdot \int_0^{X_A} \frac{1}{-r_A}\, dX_A$$

$$\tau = 60.168s$$

Reaction 2

$$\left(C + 0.5O_2 + 1.5C^{\gamma} + 1.88N_2 \rightarrow\right) CO + 1.5C^{\gamma} + 1.88N_2$$

From the reaction stoichiometry, the ratio of carbon (C) to oxygen is 2:1, and the amount of carbon(C) to be combusted is;

$$F_{c\gamma2} := F_{c\gamma} \cdot 0.2 \qquad F_{c2} := 0.2 \cdot F_c$$

$$F_{c\gamma2} = 0.034mols^{-1} \qquad F_{c2} = 0.023mols^{-1}$$

The amount of O_2 required is given-by;

$$F_{O2.2} := 0.5 \cdot F_{c2} \qquad F_{O2.2} = 0.011mols^{-1}$$

$$F_{N2.2} := 3.76 \cdot F_{O2.2} \qquad F_{N2.2} = 0.043mols^{-1}$$

5% excess O_2 is supplied to optimize the combustion process thus

$$F_{O2.2} := 1.05 \cdot F_{O2.2} \qquad F_{O2.2} = 0.012mols^{-1}$$

$$F_{N2.2} := 1.05 \cdot F_{N2.2} \qquad F_{N2.2} = 0.045mols^{-1}$$

$$A := 10 \frac{L^{0.5}}{mole^{0.5} \cdot s}$$

$$E_a := 1.1 \cdot 10^4 \frac{J}{mole}$$

$$T := (1500 + 273)K$$

Total moles of reactant=1+0.5+1.5+1.88 = 4.88

Let CA_0 be the initial concentration of carbon

$C_{A0} := \frac{C_c}{4.88}$ $\qquad C_{A0} = 2.802\times 10^4\,\mathrm{molm^{-3}}$

$X_A := 0.99$

$\xi_A := \frac{4.38 - 4.88}{4.88}$

$k_2 := A \cdot e^{\left(\frac{-E_a}{R\cdot T}\right)}$

$C_A := C_{A0} \cdot \left(\frac{1 - X_A}{1 + \xi_A \cdot X_A}\right)$

With molar concentration of reactant A and B (C and O_2) in the ratio 2:1, the simple rate expression is given as;

$r_A := -k_2 \cdot C_A^{\frac{3}{2}}$

$\tau := C_{A0} \cdot \int_0^{X_A} \frac{1}{-r_A}\, dX_A$

$\tau = 33.599\mathrm{s}$

The reaction in the combustion zone (oxidation zone) are very rapid and thus are not the rate determining step of the process flow.

The overall material flow resulting from the two reactions and based on 99% conversion specified is as follows;

Total component in flows;

$F_{ci} := F_{c1} + F_{c2}$ $\qquad F_{ci} = 0.114\mathrm{mols^{-1}}$

$F_{c\gamma i} := F_{c\gamma 1} + F_{c\gamma 2}$ $\qquad F_{c\gamma i} = 0.171\mathrm{mols^{-1}}$

$F_{O2i} := F_{O2.1} + F_{O2.2}$ $\qquad F_{O2i} = 0.108\mathrm{mols^{-1}}$

$F_{N2i} := F_{N2.1} + F_{N2.2}$ $\qquad F_{N2i} = 0.405\mathrm{mols^{-1}}$

Total component outflows

$F_{c.out} := 0.01 \cdot F_{ci}$ $F_{c.out} = 1.139 \times 10^{-3}\,mols^{-1}$

$F_{c\gamma out} := F_{c\gamma i}$ $F_{c\gamma out} = 0.171 mols^{-1}$

$F_{O2out} := (1.05 - 0.99) \cdot F_{O2i}$ $F_{O2out} = 6.457 \times 10^{-3}\,mols^{-1}$

$F_{N2out} := F_{N2i}$ $F_{N2out} = 0.405 mols^{-1}$

$F_{CO2out} := 0.99 \cdot F_{c1}$ $F_{CO2out} = 0.09 mols^{-1}$

$F_{COout} := 0.99 \cdot F_{c2}$

$F_{COout} = 0.023 mols^{-1}$

Component	INPUT Stream...		OUTPUT Stream...	
	mol/s	kg/s	mol/s	kg/s
Carbon(C)	0.114	1.36914	1.139*10^-3	0.01367939
Carbon(Cy)	0.171	2.05371	0.171	2.05371
Oxygen(O2)	0.108	3.456	6.457*10^-3	0.206624
Nitrigen(N2)	0.405	11.34527	0.405	11.34527
CO2			0.09	3.9609
CO			0.023	0.64423

Reduction Zone

$C + CO_2 <=> 2CO$ ....

$C + H2O <=> CO + H_2$ ....

$C + 2H_2 <=> CH_4$ ....

$CO_2 + H_2 <=> CO + H2O$ ....

Note the C, in each of the above (3-6) is Cγ, coming down red hot from the combustion zone

The first two equations are the main reactions taken place in the reduction zone and are both highly endothermic

Note that this is the rate determining step

The optimum temperature for maximum yield of the efficient gas is obtained from the figure 13 as around: 950^0C

$T := (950 + 273)K$

$C + CO_2 < \blacksquare = \blacksquare > 2CO$

A whole range of factors (particle size, diffusion of reactants and products in and out of the reaction zone, flow mechanism, e.t.c.) affect the reaction rate and thus the rate constant. Thus the rate expressions for the reactions taken place are unknown as is the rate constant. A rate simple expression based on the stoichiometric equation will be employed and the rate constant chosen at random to achieve the experimental gas yield. The reactor operation can be taken as a PFR or a CSTR as the actual operation is in between the operations, but a PFR gives a closer analogue of the actual process operation mechanism.

$A_1 := 14.0 \frac{L}{mole \cdot s}$ $\qquad$ $A_2 := 0.04 \frac{L}{mole \cdot s}$

Let k_1 and k_2 represent the rate constants for the forward and reverse reactions respectively

$E_{a1} := 8.0 \cdot 10^4 \frac{J}{mole}$ $\qquad$ $E_{a2} := 7.8 \cdot 10^4 \frac{J}{mole}$

Total moles of reactant= 1+1 = 2

Let CA_0 represent the initial number of moles of carbon. It must be noted that the Carbon in this reaction equation is a representative of all Carbon containing material in the biomass otherwise called CHAR coming down from the oxidation zone.

$$C_{A0} := \frac{C_{c\gamma}}{2.0}$$

$$X_A := 0.90$$

$$\xi_A := \frac{1.0 - 1.0}{1.0}$$

$k_1 := A_1 \cdot e^{\left(\frac{-E_{a1}}{R \cdot T}\right)}$ $\qquad$ $k_2 := A_2 \cdot e^{\left(\frac{-E_{a2}}{R \cdot T}\right)}$

$k_1 = 5.36\times 10^{-6}\,m^3 s^{-1} mol^{-1}$ $k_2 = 1.864\times 10^{-8}\,m^3 s^{-1} mol^{-1}$

$$C_A := C_{A0}\cdot\left(\frac{1 - X_A}{1 + \xi_A \cdot X_A}\right)$$

With equimolar concentration of reactant A and B (C and CO_2) the simple rate expression for the forward reaction is given as:

$$r_f = -k_1 \cdot C_A^{\,2}$$

With the formation of 2moles of product C (CO) for one mole of A and one mole B the simple rate expression foe the reaction is given as

$$r_r = k_2 \cdot C_C^{\,2}$$

The overall rate expression in terms of A is thus given as

$$r_A = -k_1 \cdot C_A^{\,2} + k_2 \cdot C_C^{\,2}$$

$$C_C := C_{A0} - C_A$$

$C_{A0} = 6.836\times 10^4\,molm^{-3}$ $C_A = 6.836\times 10^3\,molm^{-3}$

$$k_1 \cdot C_A^{\,2} = 250.505\,molm^{-3}s^{-1}$$

$$C_C = 6.153\times 10^4\,molm^{-3}$$

$$k_2 \cdot C_C^{\,2} = 70.576\,molm^{-3}s^{-1}$$

$$r_A := -k_1 \cdot C_A^{\,2} + k_2 \cdot C_C^{\,2}$$

$$\tau := C_{A0} \cdot \int_0^{X_A} \frac{1}{-r_A}\, dX_A$$

$$\tau = 341.946s$$

Molar flow rate of CO_2 from oxidation zone;

$$F_{CO2i} := F_{CO2out}$$

$$F_{CO2i} = 0.09mols^{-1}$$

Total molar flow rate of carbon into reduction zone

$$F_{ci} := F_{cyout} + F_{c.out}$$

$F_{ci} = 0.172mols^{-1}$

Amount of CO_2 converted in this reaction is equal to the amount of C converted=

$0.9 \cdot \left(0.09mols^{-1}\right) = 0.081mols^{-1}$

Amount of CO_2 Unconverted in this reaction is;

$0.1 \cdot \left(0.09mols^{-1}\right) = 9 \times 10^{-3} mols^{-1}$

Total Amount of C left Unreacted by this reaction

$0.172mols^{-1} - 0.081mols^{-1} = 0.091mols^{-1}$

Amount of CO produced in this reaction step =

$2 \cdot 0.9 \cdot 0.09\frac{mol}{s} = 0.162mols^{-1}$

$C + H_2O \Leftrightarrow CO + H_2$

$A_1 := 13.5\frac{L}{mole \cdot s}$

$A_2 := .03\frac{L}{mole \cdot s}$

$E_{a1} := 8.0 \cdot 10^4 \frac{J}{mole}$

$E_{a2} := 7.8 \cdot 10^4 \frac{J}{mole}$

Total moles of reactant= 1+1 = 2

Let CA_0 represent the initial number of moles of carbon. It must be noted that the Carbon in this reaction equation is a representative of all Carbon containing material in the biomass otherwise called CHAR coming down from the oxidation zone.

$C_{A0} := \frac{C_{c\gamma}}{2.0}$

$X_A := 0.90$

$\xi_A := \frac{2.0 - 2.0}{2.0}$

$k_1 := A_1 \cdot e^{\left(\frac{-E_{a1}}{R \cdot T}\right)}$

$$k_2 := A_2 \cdot e^{\left(\frac{-E_{a2}}{R \cdot T}\right)}$$

$$C_A := C_{A0} \cdot \left(\frac{1 - X_A}{1 + \xi_A \cdot X_A}\right)$$

For the reaction of one mole of Carbon with one mole of steam to yield one mole of Carbon monoxide and one mole of Hydrogen gas, the simple rate expression is as follows;

$$r_f = -k_1 \cdot C_A \cdot C_B$$

$$r_r = k_2 \cdot C_C \cdot C_D$$

Overall rate expression in terms of reactant A (Carbon)

$$r_A = -k_1 \cdot C_A \cdot C_B + k_2 \cdot C_C \cdot C_D$$

Where A, B, C and D represent C, H_2O, CO and H_2 respectively. With equimolar concentrations of reactants on one hand and product on the other hand, this expression may be written as follows;

$$r_A = -k_1 \cdot C_A^2 + k_2 \cdot C_C^2$$

$$r_A := -k_1 \cdot C_A^2 + k_2 \cdot C_C^2$$

$$k_1 \cdot C_A^2 = 241.558 \text{molm}^{-3} \text{s}^{-1}$$

$$k_2 \cdot C_C^2 = 52.932 \text{molm}^{-3} \text{s}^{-1}$$

$$C_C := C_{A0} - C_A$$

$$\tau := C_{A0} \cdot \int_0^{X_A} \frac{1}{-r_A} \, dX_A$$

$$\tau = 326.179 \text{s}$$

Total Amount of C left after reaction 3 = 0.091mols^{-1}

In this reaction (reaction 4), steam reacts with C to yield CO and H_2 the required mole ratio of Carbon to steam is 1:1, should all the carbon left be available for reaction with H_2O(steam), then 0.091mol/s H_2O is required.

Thus amount of H_2O supplied is 0.091mols^{-1}

With 90% conversion we have

Total Amount of Carbon left after this reaction (reaction 4) =

$$0.1 \cdot \left(0.091 mols^{-1}\right) = 9.1 \times 10^{-3} mols^{-1}$$

Amount of H_2O consumed by this reaction =

$$0.9 \cdot \left(0.091 mols^{-1}\right) = 0.082 mols^{-1}$$

Amount of H_2O left by this reaction =

$$0.1 \cdot \left(0.091 mols^{-1}\right) = 9.1 \times 10^{-3} mols^{-1}$$

Amount of H_2 produced in this step

$$0.9 \cdot \left(0.091 mols^{-1}\right) = 0.082 mols^{-1}$$

Amount of CO produced in this step

$$0.9 \cdot \left(0.091 mols^{-1}\right) = 0.082 mols^{-1}$$

$$C + 2H_2 < = > CH_4$$

$$A_1 := 11.20 \frac{L}{mole \cdot s} \qquad A_2 := 0.4 s^{-1}$$

$$E_{a1} := 6.5 \cdot 10^4 \frac{J}{mole} \qquad E_{a2} := 6.3 \cdot 10^4 \frac{J}{mole}$$

Total moles of reactant= 1+2 = 3

Let CA_O represent the initial number of moles of carbon. It must be noted that the Carbon in this reaction equation is a representative of all Carbon containing material in the biomass otherwise called CHAR coming down from the oxidation zone.

$$C_{A0} := \frac{C_{c\gamma}}{2.0}$$

$$X_A := 0.95$$

$$\xi_A := \frac{1.0 - 3.0}{3.0}$$

$$k_1 := A_1 \cdot e^{\left(\frac{-E_{a1}}{R \cdot T}\right)} \qquad k_2 := A_2 \cdot e^{\left(\frac{-E_{a2}}{R \cdot T}\right)}$$

$$k_1 = 5.36 \times 10^{-6} m^3 \cdot s^{-1} \cdot mol^{-1} \qquad k_2 = 8.151 \times 10^{-4} s^{-1}$$

$$C_A := C_{A0} \cdot \left(\frac{1 - X_A}{1 + \xi_A \cdot X_A} \right)$$

$$C_A = 9.322\times 10^{3} \mathrm{molm}^{-3}$$

For the reaction of one mole of Carbon with three moles of hydrogen to yield one mole of methane, the simple rate expression is as follows;

$$r_f = -k_1 \cdot C_A \cdot C_B$$

$$r_r = k_2 \cdot C_C$$

Overall rate expression in terms of reactant A (Carbon)

$$r_A = -k_1 \cdot C_A \cdot C_B + k_2 \cdot C_C$$

$$C_C := C_{A0} - C_A$$

$$C_C = 5.904\times 10^{4} \mathrm{molm}^{-3}$$

Where A, B and C represent C, H_2 and CH_4 respectively. With equimolar concentrations of reactants on one hand this expression may be written as follows;

$$r_A := -k_1 \cdot C_A{}^{2} + k_2 \cdot C_C$$

$$\tau := C_{A0} \cdot \int_0^{X_A} \frac{1}{-r_A} \, dX_A$$

$$\tau = 41.075\mathrm{s}$$

Total amount of Carbon left after reaction 4 is $9.1\times 10^{-3}\mathrm{mols}^{-1}$

Amount of H_2 required in equation 5 to convert this remaining C = $2 \cdot 9.1\times 10^{-3}\mathrm{mols}^{-1}$

Going by the 95% conversion stipulated for this reaction (reaction 5)

Amount of H_2 used up= $0.95 \cdot 2 \cdot \left(9.1\times 10^{-3}\mathrm{mols}^{-1}\right) = 0.017\mathrm{mols}^{-1}$

Amount of H_2 left after this stage is

$$0.082\frac{\mathrm{mol}}{\mathrm{s}} - 0.017\frac{\mathrm{mol}}{\mathrm{s}} = 0.065\mathrm{mols}^{-1}$$

Amount of C converted in this step

$$0.95 \cdot \left(9.1\times 10^{-3}\mathrm{mols}^{-1}\right) = 8.645\times 10^{-3}\mathrm{mols}^{-1}$$

Amount of C left after reaction 5

$$0.05 \cdot \left(9.1 \times 10^{-3} mols^{-1}\right) = 4.55 \times 10^{-4} mols^{-1}$$

Amount of CH_4 produced in this step

$$0.95 \cdot \left(9.1 \times 10^{-3} mols^{-1}\right) = 8.645 \times 10^{-3} mols^{-1}$$

$$\frac{V}{v} = \tau = \frac{volume\_\left(m^3\right)}{volumetric\_flow\_rate\_\left(\frac{m^3}{s}\right)}$$

$$CO_2 + H_2 <==> CO + H_2O$$

Total moles of reactant= 1+1 = 2

Let CA_0 represent the initial number of moles of CO_2, It must be noted that the in this reaction equation ,called water gas equilibrium, is a reaction that attempt to balance the concentration of CO and water vapor(H_2O) and the product of the concentration of CO_2 and H_2. The experimental values of Equilibrium constant (Kwe) as a function of temperature is tabulated in Table 1

$$K_{WE} = \frac{(CO) \cdot (H2O)}{(CO2) \cdot (H2)}$$

From the table it can be deduced that the value of Kwe at 950^{0}C is approximately 1.42 One thing that must be noted, however is that the extent to which this reaction occurs is small and depend on a number of factors, which are not entirely understood. It must be stated however that the operating condition chosen is suitable as it shifts the equilibrium to the left, thus favoring the production of CO and water vapor,

The basis for this reaction is the quantity of CO_2 left in reaction 3, which is =

$9 \times 10^{-3} mols^{-1}$

The required amount of H_2 for this reaction, going by the stoichiometry is also =

$9 \times 10^{-3} mols^{-1}$

The amount of CO and H_2O formed by this reaction is a function of the position of the equilibrium constant, Kwe, which in this is;

Kwe: = 1.42

Which in simple terms represent an increase in the amounts of CO and H_2O participating in this reaction step by a factor 1.42.

In actual sense this represent a decrease in the amount of all CO_2 left by the end of the above reactions by a factor(1/1.42) and the Equivalent decrease in the amount of H_2 and corresponding increase in the amount of CO and H_2O.

$$2 \cdot 1.42 = 2.84$$

$$K_{WE} = \frac{(C_i \cdot X)(C_i \cdot X)}{(C_i(1 - X))(C_i(1 - X))}$$

$$1.42 = \frac{X^2}{(1 - X^2)}$$

$$\text{polyroots}\left(\left(\begin{matrix} 1.42 \\ -2.84 \\ 0.42 \end{matrix}\right)\right) = \left(\begin{matrix} 0.544 \\ 6.218 \end{matrix}\right)$$

thus

$$X := 0.544$$

$$F_{CO2in} := 9 \cdot 10^{-3} \frac{mol}{s}$$

$$F_{CO2out} := F_{CO2in}(1 - X) \qquad F_{CO2out} = 4.104 \times 10^{-3} mols^{-1}$$

CO_2 converted =

$$F_{CO2converted} := F_{CO2in} \cdot X$$

$$F_{CO2converted} = 4.896 \times 10^{-3} mols^{-1}$$

$$F_{H2converted} := F_{CO2converted}$$

$$F_{H2converted} = 4.896 \times 10^{-3} mols^{-1}$$

$$F_{H2left} := (0.065 \cdot mol \cdot s^{-1}) - F_{H2converted}$$

$$F_{H2left} = 0.06 mols^{-1}$$

$$F_{COproduced} := F_{CO2converted}$$

$$F_{COproduced} = 4.896 \times 10^{-3} mols^{-1}$$

Total CO after the final step

$$\left(0.023+0.162+0.082+4.896\times 10^{-3}\right)\frac{mol}{s}=0.272mols^{-1}$$

$$F_{H2Oproduced} := F_{CO2converted}$$

$$F_{H2Oproduced} = 4.896\times 10^{-3}mols^{-1}$$

Total amount of water vapor after the final step

$$\left(9.1\cdot 10^{-3}+4.896\times 10^{-3}\right)\frac{mol}{s}=0.014mols^{-1}$$

Total amount of Carbon unconverted = $4.55\cdot 10^{-4}\frac{mol}{s}$

Total amount of CH4 produced is $8.645\cdot 10^{-3}\frac{mol}{s}$

Summary of Material balance across the reduction zone.

Component	mol/s	kg/s	mol/s	kg/s
C	0.172	0.002066	0.000455	5.46455E-06
CO2	0.09	0.003961	0.004104	0.000180617
CO	0.023	0.000644	0.272	0.007619
H2O	0.091	0.001639	0.014	0.000252
H2			0.06	0.000121
CH4			0.008645	0.000138666
OXYGEN	0.006457	0.000207	0.006457	0.000207
NITROGEN	0.405	0.011344	0.405	0.011344
Total	0.787457	0.01986	0.770661	0.019867

OVERALL MATERIAL BALANCE ACROSS THE GASIFIER

GASIFIER							
Component	STREAM100	STREAM101		STREAM102	STREAM103	STREAM104	
	kg/hr	kg/hr		kg/hr	kg/hr	kg/hr	
MOISTURE	7.6			7.6			
VOLATILES	74.65			74.65			
CARBON(CHARR)	12.31				0.009836	0.009836	
ASH	5.52				1.104	4.416	
OXYGEN		12.4416			0.7452		
NITROGEN		40.8384			40.8384		
STEAM		5.900076			0.90756		
CO					27.396		
CO2					0.650221		
H2					0.4356		
CH4					0.499198		
Total	100.08	59.18008	159.2601	82.25	72.58601	4.425836	159 2619

Material Balance across Hopper 1.

Note the material flow through this unit is determined by the Gasifier design/ flow rate;

HOPPER 1				
	INPUT		OUTPUT	
	STREAM1		STREAM2	
	kg/hr		kg/hr	
RICE HUSK	70		70	
HOPPER2				
	INPUT		OUTPUT	
	STREAM5		STTREAM6	
	kg/hr		kg/hr	
	30		30	
SREDDER				
	INPUT		OUTPUT	
	STREAM2		STREAM4	
	kg/hr		kg/hr	
	70		70	
PREHEATER				
	INPUT			OUTPUT
Component	STREAM7	STREAM8		STREAM9
	kg/hr	kg/hr		kg/hr
Oxygen	12.4416			12.4416
Nitrogen	40.8384			40.8384
Steam		5.900076		5.900076
Total	53.28	5.900076		59.18008
BOILER				
	INPUT		OUTPUT	
	STREAM 10		STREAM 11	
	kg/hr		kg/hr	
H2O(L)	6.460596			
H2O(G)			6.460596	

CYCLONE

The design of the cyclone is such that 50% of all entrained carbon and 50% of all entrained Ash is removed in the cyclone

	INPUT	OUTPUT	
COMPONENT	STREAM103	STREAM105	STREA106
	kg/hr	kg/hr	kg/hr
Carbon	0.009836	0.004918	0.004918
Ash	1.104	0.552	0.552
Oxygen	0.7452	0.7452	
Nitrogen	40.8384	40.8384	
Steam	0.90756	0.90756	
CO	27.396	27.396	
CO2	0.650221	0.650221	
H2	0.4356	0.4356	
CH4	0.499198	0.499198	
Total	72.58602	72.0291	0.556918

MECHANICAL FILTER

The unit is designed such that all the remaining Carbon and Ash entrained in the gas is removed in this unit.

	INPUT	OUTPUT	
	STREAM 105	STREAM 107	STREAM 108
COMPONENTS	kg/hr	kg/hr	kg/hr
Carbon	0.004918		0.004918
Ash	0.552		0.552
Oxygen	0.7452	0.7452	
Nitrogen	40.8384	40.8384	
Steam	0.90756	0.90756	
CO	27.396	27.396	
CO2	0.650221	0.650221	
H2	0.4356	0.4356	
CH4	0.499198	0.499198	
	72.0291	71.47218	0.556918

SCRUBBER / COOLER

In this unit all the water vapor is condensed to liquid water and all Carbon dioxide present in the gas stream is removed by $Ca(OH)_2$ dissolved in the water.

The solvent is either dilute or concentrated lime water solution, which for the purpose of analysis will be taken to be $Ca(OH)_2$

$H_2O\ (L) \longrightarrow H_2O\ (G)$

$Ca(OH)_2 + CO_2 + 3H_2O \longrightarrow CaCO_3 + H_2O + 3H_20$

$MM_{CO2} = 0.044 kg\, mol^{-1}$

$F_{CO2in} := 0.650221 \frac{kg}{hr}$

$F_{CO2in} := \frac{F_{CO2in}}{MM_{CO2}}$

$F_{CO2in} = 4.104 \times 10^{-3} mols^{-1}$

Lime_water_rep_Ca(OH)_2

$F_{lime\_water\_required} := F_{CO2in}$

$F_{lime\_water\_required} = 4.104 \times 10^{-3} mols^{-1}$

$F_{associated\_water\_suppled} := 3 \cdot F_{lime\_water\_required}$

$F_{associated\_water\_suppled} = 0.012 mols^{-1}$

$F_{output\_water} := 4 \cdot F_{CO2in}$

$F_{output\_water} = 0.016 mols^{-1}$

$F_{CaCO3out} := F_{CO2in}$

$F_{CaCO3out} = 4.104 \times 10^{-3} mols^{-1}$

SCRUBBER								
		INPUT				OUTPUT		
COMPONENT	STREAM107		STREAM109		STREAM110		STREAM111	
	mol/s	kg/hr	mol/s	kg/hr	mol/s	kg/hr	mol/s	kg/hr
Oxygen	0.006469	0.7452			0.006469	0.7452		
Nitrogen	0.404998	40.8384			0.404998	40.8384		
Steam	0.013998	0.90756					0.013998	0.90756
CO	0.271689	27.396			0.271689	27.396		
CO2	0.004104	0.650221						
H2	0.06002	0.4356			0.06002	0.4356		
CH4	0.008645	0.499198			0.008645	0.499198		
Ca(OH)2			0.004104	1.094783				
H2O			0.012	0.778032			0.016	1.037376
CaCO3							0.004104	1.47877
total	0.769922	71.47218	0.016104	1.872815	0.751821	69.9144	0.034102	3.423706
	total input	73.34499			totaloutput	73.3381		
		kg/hr				kg/hr		

FILTER

FILTER					
COMPONENT	STREAM111		STREAM112		STREAM113
	kg/hr		kg/hr		kg/hr
CaCO3	1.47877		1.47877		
Water	1.944936				1.944936
	3.423706		1.47877		1.944936

COOLING TOWER

In this unit hot water is allowed to fall through open air under gravity.

COOLING TOWER				
	INPUT		OUTPUT	
COMPONENT	STREAM113		STREAM114	
	kg/hr		kg/hr	
Water	1.944936		1.944936	

CATALYTIC REFORMER

The synthesis gas is the passed through the catalytic reformer, where the methane (CH_4) reacts with steam to form carbon monoxide (CO) and hydrogen (H_2). The reaction is endothermic and the forward reaction is favored by increased temperature. The introduction of a catalyst however reduces the activation energy of the process thus allowing the reacting to take place rapidly at temperatures 120^0C.

$CH_4 + H_2O \Leftrightarrow CO + 3H_2$..

The reaction above is an equilibrium reaction and the value of the equilibrium constant at different temperatures for a NON-catalyzed reaction is obtained from HYSYS simulator as follows;

371.1	1.024*10^-5
426.7	2.659*10^-4
482.2	4.338*10^-3
537.8	4.9*10^-2
593.3	0.4098
648.9	2.676
704.4	14.26
760	63.43
815.6	242.6

From the above table it is observed that the operation of this unit without a catalyst will require a very high temperature for high conversion. This temperature will be around 780^0C in which case, the product of concentration of the products will be about 100 times the product of the concentration of the reactants

$$Keq = \frac{(CO) \cdot (3H_2)^3}{(CH_4) \cdot (H2O)}$$

With the effective use of catalyst results show that the same conversion achieved at 780^0C in a NON-catalyzed reactor can be achieved at 120^0C in a catalyzed reactor. This value varies depending on the type of catalyst, the age of catalyst, the contacting patterns in the reactor, e.t.c.

The basic assumption here is that the catalyst so employed is able to achieve to above conversion at 120^0C

Let the initial molar concentration of CH_4 be Ci and its fractional conversion after a time t be X, if the stoichiometric quantity of H_2O is supplied the initial concentration of H_2O is also Ci and fractional conversion after any time t is X.

Thus the concentration of CH_4 and H_2O after any time t are both Ci (1-X) and corresponding concentrations of CO and H_2 are, Ci(X) and 3Ci(X) respectively.

$$At\_30^0C\_and\_1atm\_the\_density\_of\_CH4\_is\_0 \cdot 6463\frac{kg}{m^3}$$

$$\rho_{CH4} := 0.06463\frac{kg}{m^3}$$

$$MM_{CH4} := 16.04\frac{kg}{mol \cdot 10^3}$$

$$C_{CH4} := \frac{\rho_{CH4}}{MM_{CH4}}$$

$c_{CH4} = 4.029 \mathrm{mol\,m^{-3}}$

The initial concentration of CH_4 in the bulk gas entering the catalytic reformer is given by;

$$C_i := \frac{c_{CH4}}{\text{total\_no\_of\_moles\_of\_gas\_entering\_the\_reactor\_per\_mol\_CH4}}$$

NB; the total no. of moles of the gas entering this unit is the sum of the no of moles of gas from the preceding unit and that of the steam supplied

For each mole of CH_4 that goes through the catalytic reformer there is an associated;

(0.06/8.645*10^-3)= 6.94 moles of H_2

And

(0.272/8.645*10^-3)=31.46 moles of CO,

(0.006457/0.008645)=0.747 moles of O_2

(0.405/0.008645)= 46.848 moles of N_2

Entering, in Addition to 1mol of H_2O supplied to react with 1mol of CH_4.

Thus

Total no. of moles of gas entering the reactor per mol CH_4 =

$$(6.94 + 31.46 + 0.747 + 46.848 + 1 + 1) = 87.995$$

$C_i = 0.046 \mathrm{mol\,m^{-3}}$

$Keq := 100$

$$Keq = \frac{(C_i \cdot X)(3C_i \cdot X)^3}{(C_i(1 - X)\,C_i(1 - X))}$$

$$X^4 - 1231X^2 + 2462X - 1231 = 0$$

$$\text{polyroots}\left(\begin{pmatrix} -1231 \\ 2462 \\ -1231 \\ 0 \\ 1 \end{pmatrix}\right) = \begin{pmatrix} -36.059 \\ 0.973 \\ 1.03 \\ 34.055 \end{pmatrix}$$

Since X must be less than 1, then

$X := 0.973$

Thus the concentration of CH_4 at the exit of the catalytic reformer is

$C_{CH4out} := C_i \cdot (1 - X)$

$C_{CH4out} = 1.236\times 10^{-3} mol m^{-3}$

Note this is the concentration of CH_4 in the bulk gas exiting the catalytic reformer

$C_{H2Oout} := C_{CH4out}$

$C_{H2Oout} = 1.236\times 10^{-3} mol m^{-3}$

$C_{COproduced} := C_i \cdot X$

$C_{COproduced} = 0.045 mol m^{-3}$

$C_{H2produced} := 3 \cdot C_i \cdot X$

$C_{H2produced} = 0.134 mol m^{-3}$

The volumetric flow rate of the bulk fluid is obtained as follows;

Molar flow rate;

$F_{CH4in} := 8.645 \cdot 10^{-3} \frac{mol}{s}$

$F_{CH4out} := F_{CH4in} \cdot (1 - X)$

$F_{CH4out} = 2.334\times 10^{-4} mol s^{-1}$

$F_{H20in} := F_{CH4in}$

$F_{H20out} := F_{CH4out}$

$F_{H20out} = 2.334\times 10^{-4} mol s^{-1}$

$F_{H2produced} := 3F_{CH4in} \cdot X$

$F_{H2produced} = 0.025 mol s^{-1}$

$F_{COproduced} := F_{CH4in} \cdot X$

$F_{COproduced} = 8.412\times 10^{-3} mol s^{-1}$

$F_{H2in} := 0.06 \frac{mol}{s}$

$F_{H2out} := F_{H2produced} + F_{H2in}$

$F_{H2out} = 0.085 \mathrm{mol\,s^{-1}}$

$F_{COin} := 0.272 \frac{\mathrm{mol}}{\mathrm{s}}$

$F_{COout} := F_{COproduced} + F_{COin}$

$F_{COout} = 0.28 \mathrm{mol\,s^{-1}}$

$MM_{CH4} = 0.016 \mathrm{kg\,mol^{-1}}$

$MM_{H2} \equiv 2.016 \cdot 10^{-3} \frac{\mathrm{kg}}{\mathrm{mol}}$

$MM_{CO} = 0.028 \mathrm{kg\,mol^{-1}}$

$MM_{H2O} := 18.01 \cdot 10^{-3} \frac{\mathrm{kg}}{\mathrm{mol}}$

$F_{CH4in} = 8.645 \times 10^{-3} \mathrm{mol\,s^{-1}}$

$F_{CH4in} := F_{CH4in} \cdot MM_{CH4}$

$F_{CH4out} := F_{CH4out} \cdot MM_{CH4}$

$F_{CH4in} = 1.387 \times 10^{-4} \mathrm{kg\,s^{-1}}$

$F_{CH4out} = 3.744 \times 10^{-6} \mathrm{kg\,s^{-1}}$

$F_{H20in} = 8.645 \times 10^{-3} \mathrm{mol\,s^{-1}}$

$F_{H20in} := F_{H20in} \cdot MM_{H2O}$

$F_{H20out} := F_{H20out} \cdot MM_{H2O}$

$F_{H20out} = 4.204 \times 10^{-6} \mathrm{kg\,s^{-1}}$

$F_{H20in} = 1.557 \times 10^{-4} \mathrm{kg\,s^{-1}}$

$F_{H2in} = 0.06 \mathrm{mol\,s^{-1}}$

$F_{H2in} := F_{H2in} \cdot MM_{H2}$

$F_{H2out} := F_{H2out} \cdot MM_{H2}$

$F_{H2in} = 1.21 \times 10^{-4} \mathrm{kg\,s^{-1}}$

$F_{H2out} = 1.718\times 10^{-4}\,kg\,s^{-1}$

$F_{COin} = 0.272mol\,s^{-1}$

$F_{COin} := F_{COin}\cdot MM_{CO}$

$F_{COout} := F_{COout}\cdot MM_{CO}$

$F_{COin} = 7.619\times 10^{-3}\,kg\,s^{-1}$

$F_{COout} = 7.854\times 10^{-3}\,kg\,s^{-1}$

	INPUT		OUTPUT	
Component	mol/s	kg/hr	mol/s·	kg/hr
CO	0.272	27.4284	0.28	28.2744
H2	0.06	0.4356	0.085	0.61848
CH4	0.008645	0.49932	0.0002334	0.0134784
steam	0.008645	0.56052	0.0002334	0.0151344
Oxygen	0.006469		0.006469	
Nitrogen	0.404998		0.404998	
Total	0.760757	28.92384	0.776934	28.92149

CHAPTER FOUR

4.0 ENERGY BALANCE

Energy balances across unit operation equipments

The energy balances across the equipments is given as

$$\Delta H = \int_{Tref}^{T} nCp\,dT$$

Where Cp = Heat capacity which is given as

$$Cp = a + b\cdot T + c\cdot T^2 + d\cdot T^3$$

a, b, c, d = Heat constants

T = "Operating temperature"

Tref = "Reference temperature"

n = "amount in kmol"

Since $Cp = a + b\cdot T + c\cdot T^2 + d\cdot T^3$, the formula for the calculation of enthalpy across each unit operation equipment can then be written as

$$\Delta H := \int_{Tref}^{T} n\left(a + bT + cT^{\langle 2\rangle} + dT^{\langle 3\rangle}\right) dT$$

Energy balances across reactor

The energy balances across the reactor will be calculated using the formula

$$\Delta H := \int_{Tr}^{T} nCp\,dT + Hf$$

Similarly where Cp = Heat capacity which is given as

$$Cp = a + b\cdot T + c\cdot T^2 + d\cdot T^3$$

T = "Operating temperature"

Tref = "Reference temperature"

n = "amount in kmol"

H_f = "Heat of formation"

Heat capacity equation constants for the following

$$C := \begin{pmatrix} 11.18 & 1.095\cdot 10^{-2} & -4.891\cdot 10^{5} & 0 \\ 29.10 & 1.158\cdot 10^{-2} & -0.6076\cdot 10^{-5} & 1.311\cdot 10^{-9} \\ 29.000 & 0.2199\cdot 10^{-2} & 0.5723\cdot 10^{-5} & -2.871\cdot 10^{-9} \\ 33.46 & 0.6880\cdot 10^{-2} & 0.7604\cdot 10^{-5} & -3.593\cdot 10^{-9} \\ 28.95 & 0.4110\cdot 10^{-2} & 0.3548\cdot 10^{-5} & -2.220\cdot 10^{-9} \\ 36.1 & 4.233\cdot 10^{-2} & -2.887\cdot 10^{-5} & 7.464\cdot 10^{-9} \\ 28.84 & 0.00765\cdot 10^{-2} & 0.3288\cdot 10^{-5} & -0.8698\cdot 10^{-9} \\ 34.31 & 5.469\cdot 10^{-2} & 0.3661\cdot 10^{-5} & -11.0\cdot 10^{-9} \\ 18.2964 & 47.212\cdot 10^{-2} & -133.88\cdot 10^{-5} & 1314.2\cdot 10^{-9} \end{pmatrix} \qquad \begin{pmatrix} \text{Carbon} \\ \text{Oxygen} \\ \text{Nitrogen} \\ \text{Steam} \\ \text{CO} \\ \text{CO2} \\ \text{H2} \\ \text{CH4} \\ \text{Water} \end{pmatrix}$$

$$a := C^{\langle 0\rangle}\cdot\frac{1}{K} \qquad b := C^{\langle 1\rangle}\cdot\frac{1}{K^2} \qquad c := C^{\langle 2\rangle}\cdot\frac{1}{K^3} \qquad d := C^{\langle 3\rangle}\cdot\frac{1}{K^4}$$

$$\Delta H_f := \begin{pmatrix} 0 \\ 0 \\ 0 \\ -241826 \\ -110520 \\ -393510 \\ 0 \\ -74840 \\ -285840 \end{pmatrix}\frac{J}{mol} \qquad \begin{pmatrix} \text{Carbon} \\ \text{Oxygen} \\ \text{Nitrogen} \\ \text{Steam} \\ \text{CO} \\ \text{CO2} \\ \text{H2} \\ \text{CH4} \\ \text{Water} \end{pmatrix}$$

GASIFIER							
Component	STREAM100	STREAM101		STREAM102	STREAM103	STREAM104	
	kg/hr	kg/hr		kg/hr	kg/hr	kg/hr	
MOISTURE	7.6			7.6			
VOLATILES	74.65			74.65			
CARBON(CHARR)	12.31				0.009836	0.009836	
ASH	5.52				1.104	4.416	
OXYGEN		12.4416			0.7452		
NITROGEN		40.8384			40.8384		
STEAM		5.900076			0.90756		
CO					27.396		
CO2					0.650221		
H2					0.4356		
CH4					0.499198		
Total	100.08	59.18008	159.2601	82.25	72.58601	4.425836	159.2619

GASIFIER INPUT				OUTPUT		
COMPONENT	STREAM100 mol/s	STREAM101 mol/s		STREAM102 mol/s	STREAM103 mol/s	STREAM104 mol/s
MOISTURE	0.117219			0.117219		
VOLATILES	0.471168			0.471168		
CARBON	0.284716				0.000227	0.000227
ASH	0.127671				0.025534	0.102137
OXYGEN		0.108			0.006469	
NITROGEN		0.404998			0.404998	
STEAM		0.091			0.013998	
CO					0.271689	
CO2					0.004104	
H2					0.06002	
CH4					0.008645	

ENTHALPY OF INPUT

$T_r := (27 + 273)K$ $\qquad\qquad$ $T_{G\_agent} := (350 + 273)K$

$T_r = 300\,K$

$T_a := (30 + 273)K$ $\qquad\qquad$ $T_{G\_agent} = 623K$

$T_a = 303K$

Tr = Reference Temperature

Ta = Ambient Temperature

T_{G_agent} = Temperature of Gasification Agent

T_{top} = Temperature at the top of Gasifier

T_{bot} = Temperature at the bottom of Gasifier

Note: The thermodynamic properties of Ash will be taken as that of Carbon and that for Volatiles will be taken as that of CO_2. Also Note the heat of formation of the volatiles will be fixed at zero, with the assumption that no reaction takes in the volatilization process

$$H_{moisture} := 0.117219\frac{mol}{s}\left[\left[\int_{T_r}^{T_a}\left(a_8 + b_8\cdot T + c_8\cdot T^2 + d_8\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_8}\right]$$

Tr = 300K $\qquad\qquad$ Ta = 303K

$$\int_{Tr}^{Ta}\left(a_8 + b_8\cdot T + c_8\cdot T^2 + d_8\cdot T^3\right)dT$$

$$= \left(a_8 \cdot T + b_8 \cdot \frac{T^2}{2} + c_8 \cdot \frac{T^3}{3} + d_8 \cdot \frac{T^4}{4} \right) \Big/_{Tr}^{Ta}$$

$$H_{moisture} = 18.2964T + 47.212 \times 10^{-2} \frac{T^2}{2} + \{-133.88 \times 10^{-5}\} \frac{T^3}{3} + 1314.2 \times 10^{-9} \frac{T^4}{4} \Big/_{Tr}^{Ta}$$

$$H_{moisture} = -3.348 \times 10^{4} s^{-1} J$$

$$H_{volatiles} := 0.471168 \frac{mol}{s} \left[\left[\int_{T_r}^{T_a} \left(a_5 + b_5 \cdot T + c_5 \cdot T^2 + d_5 \cdot T^3 \right) dT \right] \frac{J}{mol} + \Delta H_{f_0} \right]$$

$$H_{volatiles} = 65.647 s^{-1} J$$

$$H_C := 0.284716 \frac{mol}{s} \left[\left[\int_{T_r}^{T_a} \left(a_0 + b_0 \cdot T + c_0 \cdot T^2 + d_0 \cdot T^3 \right) dT \right] \frac{J}{mol} + \Delta H_{f_0} \right]$$

$$H_C = -3.798 \times 10^{10} s^{-1} J$$

$$H_{ash} := 0.127671 \frac{mol}{s} \left[\left[\int_{T_r}^{T_a} \left(a_0 + b_0 \cdot T + c_0 \cdot T^2 + d_0 \cdot T^3 \right) dT \right] \frac{J}{mol} + \Delta H_{f_0} \right]$$

$$H_{ash} = -1.703 \times 10^{10} s^{-1} J$$

$$H_{O2} := 0.108 \frac{mol}{s} \left[\left[\int_{T_r}^{T_{G\_agent}} \left(a_1 + b_1 \cdot T + c_1 \cdot T^2 + d_1 \cdot T^3 \right) dT \right] \frac{J}{mol} + \Delta H_{f_1} \right]$$

$$H_{O2} = 1.16 \times 10^{3} s^{-1} J$$

$$H_{N2} := 0.404998 \frac{mol}{s} \left[\left[\int_{T_r}^{T_{G\_agent}} \left(a_2 + b_2 \cdot T + c_2 \cdot T^2 + d_2 \cdot T^3 \right) dT \right] \frac{J}{mol} + \Delta H_{f_2} \right]$$

$$H_{N2} = 4.051 \times 10^{3} s^{-1} J$$

$$H_{steam} := 0.091 \frac{mol}{s} \left[\left[\int_{T_r}^{T_{G\_agent}} \left(a_3 + b_3 \cdot T + c_3 \cdot T^2 + d_3 \cdot T^3 \right) dT \right] \frac{J}{mol} + \Delta H_{f_3} \right]$$

$$H_{steam} = -2.089 \times 10^{4} s^{-1} J$$

$$\Delta H_{in} := H_{moisture} + H_{volatiles} + H_C + H_{ash} + H_{O2} + H_{N2} + H_{steam}$$

$\Delta H_{in} = -5.501\times 10^{10}\,s^{-1}\,J$

OUPUT ENTHALPY

$T_{top} := (120 + 273)K \qquad T_{bot} := (950 + 273)K$

$$H_{moisture} := 0.117219\frac{mol}{s}\left[\left[\int_{T_r}^{T_{top}}\left(a_8 + b_8\cdot T + c_8\cdot T^2 + d_8\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_8}\right]$$

$H_{moisture} = -3.268\times 10^{4}\,s^{-1}\,J$

$$H_{volatiles} := 0.471168\frac{mol}{s}\left[\left[\int_{T_r}^{T_{top}}\left(a_5 + b_5\cdot T + c_5\cdot T^2 + d_5\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_0}\right]$$

$H_{volatiles} = 2.086\times 10^{3}\,s^{-1}\,J$

$$H_C := 0.0002272\frac{mol}{s}\left[\left[\int_{T_r}^{T_{bot}}\left(a_0 + b_0\cdot T + c_0\cdot T^2 + d_0\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_0}\right]$$

$H_C = -1.334\times 10^{11}\,s^{-1}\,J$

$$H_{ash} := (0.025534 + 0.102137)\frac{mol}{s}\left[\left[\int_{T_r}^{T_{bot}}\left(a_0 + b_0\cdot T + c_0\cdot T^2 + d_0\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_0}\right]$$

$H_{ash} = -3.751\times 10^{13}\,s^{-1}\,J$

$$H_{O2} := 0.006469\frac{mol}{s}\left[\left[\int_{T_r}^{T_{bot}}\left(a_1 + b_1\cdot T + c_1\cdot T^2 + d_1\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_1}\right]$$

$H_{O2} = 207.518\,s^{-1}\,J$

$$H_{N2} := 0.404998\frac{mol}{s}\left[\left[\int_{T_r}^{T_{bot}}\left(a_2 + b_2\cdot T + c_2\cdot T^2 + d_2\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_2}\right]$$

$H_{N2} = 1.221\times 10^{4}\,s^{-1}\,J$

$$H_{steam} := 0.013998\frac{mol}{s}\left[\left[\int_{T_r}^{T_{bot}}\left(a_3 + b_3\cdot T + c_3\cdot T^2 + d_3\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_3}\right]$$

$H_{steam} = -2.849 \times 10^{3}\,s^{-1}\,J$

$$H_{CO} := 0.271689\frac{mol}{s}\left[\left[\int_{T_r}^{T_{bot}}\left(a_4 + b_4 \cdot T + c_4 \cdot T^2 + d_4 \cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_4}\right]$$

$H_{CO} = -2.174 \times 10^{4}\,s^{-1}\,J$

$$H_{CO2} := 0.004104\frac{mol}{s}\left[\left[\int_{T_r}^{T_{bot}}\left(a_5 + b_5 \cdot T + c_5 \cdot T^2 + d_5 \cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_5}\right]$$

$H_{CO2} = -1.41 \times 10^{3}\,s^{-1}\,J$

$$H_{H2} := 0.06\frac{mol}{s}\left[\left[\int_{T_r}^{T_{bot}}\left(a_6 + b_6 \cdot T + c_6 \cdot T^2 + d_6 \cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_6}\right]$$

$H_{H2} = 1.69 \times 10^{3}\,s^{-1}\,J$

$$H_{CH4} := 0.008645\frac{mol}{s}\left[\left[\int_{T_r}^{T_{bot}}\left(a_7 + b_7 \cdot T + c_7 \cdot T^2 + d_7 \cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_7}\right]$$

$H_{CH4} = -74.89\,s^{-1}\,J$

$$\Delta H_{out} := H_{moisture} + H_{volatiles} + H_C + H_{ash} + H_{O2} + H_{N2} + H_{steam} + H_{CO} + H_{CO2} + H_{H2} + H_{CH4}$$

$\Delta H_{out} = -3.765 \times 10^{13}\,s^{-1}\,J$

$\Delta H_{Gasifier} := \Delta H_{out} - \Delta H_{in}$

$\Delta H_{Gasifier} = -3.759 \times 10^{13}\,s^{-1}\,J$

BOILER

BOILER						
	INPUT		OUTPUT			
COMPONENT	STREAM...		STREAM...			
	kg/hr	mol/s	kg/hr	mol/s		
	6.460596	0.099645	6.460596	0.099645		

$T_s := (100 + 273)K$

$T_s = 373K$

Ts = Temperature of steam

$$H_{feed\_water} := 0.099645\frac{mol}{s}\left[\left[\int_{T_r}^{T_a}\left(a_8 + b_8\cdot T + c_8\cdot T^2 + d_8\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_8}\right]$$

$H_{feed\_water} = -2.846\times\ 10^4\, s^{-1}\, J$

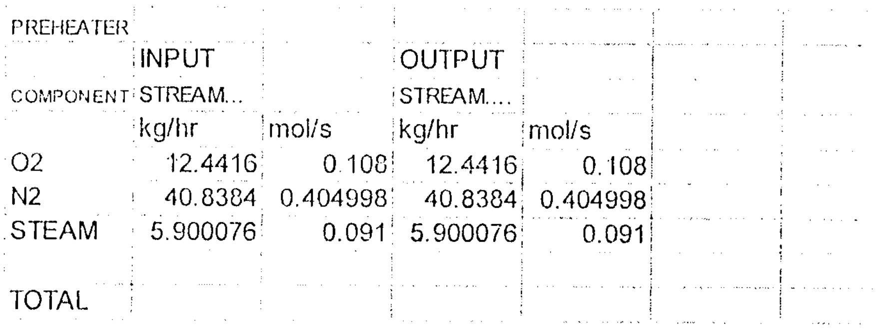
$$H_{steam} := 0.099645\frac{mol}{s}\left[\left[\int_{T_r}^{T_s}\left(a_3 + b_3\cdot T + c_3\cdot T^2 + d_3\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_3}\right]$$

$H_{steam} = -2.383\times\ 10^4\, s^{-1}\, J$

$boiler\_duty := H_{steam} - H_{feed\_water}$

$boiler\_duty\ = 4.629\times\ 10^3\, s^{-1}\, J$

PREHEATER



PREHEATER

	INPUT		OUTPUT	
COMPONENT	STREAM...		STREAM....	
	kg/hr	mol/s	kg/hr	mol/s
O2	12.4416	0.108	12.4416	0.108
N2	40.8384	0.404998	40.8384	0.404998
STEAM	5.900076	0.091	5.900076	0.091
TOTAL				

INPUT

The pre-heater is a heat exchanger operated in a counter current mode

The hot synthesis at 950^0C from the gasifier is used to heat the gasifying agent coming in at ambient conditions,

Total heat load of the syngas=n*Cp_ave*(Tbot - Tr)

Where Cp_ave is the average specific heat capacity of the syn-gas

n, is the total moles of the syn-gas.

Total enthalpy of syn-gas;

$$n \cdot Cp\_ave \cdot (Tbot - Tr) = H_{O2} + H_{N2} + H_{steam} + H_{CO} + H_{CO2} + H_{H2} + H_{CH4}$$

$$n := (0.006469 + 0.404998 + 0.01338 + 0.272 + 0.004104 + 0.06 + 0.008645) \frac{mol}{s}$$

$$\Delta T := T_{bot} - T_r$$

$$C_{p\_ave} = \frac{\left(H_{O2} + H_{N2} + H_{steam} + H_{CO} + H_{CO2} + H_{H2} + H_{CH4}\right)}{n \cdot \Delta T}$$

$$H_{O2} := 0.006469 \frac{mol}{s} \left[\left[\int_{T_r}^{T_{bot}} \left(a_1 + b_1 \cdot T + c_1 \cdot T^2 + d_1 \cdot T^3\right) dT\right] \frac{J}{mol}\right]$$

$$H_{O2} = 207.518 s^{-1} J$$

$$H_{N2} := 0.404998 \frac{mol}{s} \left[\left[\int_{T_r}^{T_{bot}} \left(a_2 + b_2 \cdot T + c_2 \cdot T^2 + d_2 \cdot T^3\right) dT\right] \frac{J}{mol}\right]$$

$$H_{N2} = 1.221 \times 10^4 s^{-1} J$$

$$H_{steam} := 0.013998 \frac{mol}{s} \left[\left[\int_{T_r}^{T_{bot}} \left(a_3 + b_3 \cdot T + c_3 \cdot T^2 + d_3 \cdot T^3\right) dT\right] \frac{J}{mol}\right]$$

$$H_{steam} = 535.916 s^{-1} J$$

$$H_{CO} := 0.271689 \frac{mol}{s} \left[\left[\int_{T_r}^{T_{bot}} \left(a_4 + b_4 \cdot T + c_4 \cdot T^2 + d_4 \cdot T^3\right) dT\right] \frac{J}{mol}\right]$$

$$H_{CO} = 8.288 \times 10^3 s^{-1} J$$

$$H_{CO2} := 0.004104 \frac{mol}{s} \left[\left[\int_{T_r}^{T_{bot}} \left(a_5 + b_5 \cdot T + c_5 \cdot T^2 + d_5 \cdot T^3\right) dT\right] \frac{J}{mol}\right]$$

$$H_{CO2} = 204.741 s^{-1} J$$

$$H_{H2} := 0.06 \frac{mol}{s} \left[\left[\int_{T_r}^{T_{bot}} \left(a_6 + b_6 \cdot T + c_6 \cdot T^2 + d_6 \cdot T^3\right) dT\right] \frac{J}{mol}\right]$$

$$H_{H2} = 1.69 \times 10^3 s^{-1} J$$

$$H_{CH4} := 0.008645\frac{mol}{s}\left[\left[\int_{T_r}^{T_{bot}}\left(a_7 + b_7 \cdot T + c_7 \cdot T^2 + d_7 \cdot T^3\right) dT\right]\frac{J}{mol}\right]$$

$$H_{CH4} = 572.101 s^{-1} J$$

$$C_{p\_ave} := \frac{\left(H_{O2} + H_{N2} + H_{steam} + H_{CO} + H_{CO2} + H_{H2} + H_{CH4}\right)}{n \cdot \Delta T}$$

$$C_{p\_ave} = 33.377\frac{J}{mol \cdot K}$$

Heat loss by Syngas = Heat gain by Gasifying agent

$$n \cdot C_{p\_ave} \cdot \left(T_{bot} - T_{out}\right) = n_{G\_agent} \cdot C_{p\_G\_agent} \cdot \left(T_{G\_agent} - T_{in}\right) \ldots\ldots\ldots i$$

$$n_{G\_agent} \cdot C_{p\_G\_agent} \cdot \left(T_{G\_agent} - T_a\right) = H_{O2} + H_{N2} + H_{steam}$$

$$H_{O2} := 0.108\frac{mol}{s}\left[\left[\int_{T_a}^{T_{G\_agent}}\left(a_1 + b_1 \cdot T + c_1 \cdot T^2 + d_1 \cdot T^3\right) dT\right]\frac{J}{mol}\right]$$

$$H_{O2} = 1.149 \times 10^3 s^{-1} J$$

$$H_{N2} := 0.404998\frac{mol}{s}\left[\left[\int_{T_a}^{T_{G\_agent}}\left(a_2 + b_2 \cdot T + c_2 \cdot T^2 + d_2 \cdot T^3\right) dT\right]\frac{J}{mol}\right]$$

$$H_{N2} = 4.014 \times 10^3 s^{-1} J$$

$$H_{steam} := 0.091\frac{mol}{s}\left[\left[\int_{T_s}^{T_{G\_agent}}\left(a_3 + b_3 \cdot T + c_3 \cdot T^2 + d_3 \cdot T^3\right) dT\right]\frac{J}{mol}\right]$$

$$H_{steam} = 872.234 s^{-1} J$$

Let

$$\Delta H_{G\_agent} = n_{G\_agent} \cdot C_{p\_G\_agent} \cdot \left(T_{G\_agent} - T_{in}\right)$$

$$\Delta H_{G\_agent} := H_{O2} + H_{N2} + H_{steam}$$

$$\Delta H_{G\_agent} = 6.036 \times 10^3 s^{-1} J$$

From equation i above

$$T_{out} := T_{bot} - \left(\frac{\Delta H_{G\_agent}}{n \cdot C_{p\_ave}}\right)$$

$T_{out} = 988.022K$

Mean temperature difference for counter flow operation

The average inlet temperature of the of the air and steam is around 70^{0}C due the higher heat capacity of steam, where the steam was at 100^{0}C and air was at ambient temperature of 30^{0}C

$T_{in\_ave} := (70 + 273)K$

$\theta_1 := \left(T_{bot} - T_{G\_agent}\right)$

$\theta_2 := \left(T_{out} - T_{in\_ave}\right)$

$$\theta_m := \frac{\left(\theta_1 - \theta_2\right)}{\ln\left(\frac{\theta_1}{\theta_2}\right)}$$

$\theta_m = 622.24K$

MECHANICAL FILTER

The basic assumption in this unit is that there is no temperature drop across the unit and since there is no reaction taken place in this unit, the change in enthalpy is zero.

That is; $\Delta H_{out} - \Delta H_{in} = 0$

SCRUBBER

SCRUBBER

COMPONENT	INPUT mol/s	OUTPUT mol/s
O2	0.006469	0.006469
N2	0.404998	0.404998
STEAM	0.013998	
CO	0.272	0.272
CO2	0.004104	
H2	0.06	0.06
CH4	0.008645	0.008645
Ca(OH)2	0.004104	
H2O	0.012	0.029998
CaCO3		0.004104

$$X := \begin{pmatrix} 89.5 & 0 & 0 & 0 \\ 82.34 & 4.975\cdot 10^{-2} & -12.87\cdot 10^{5} & 0 \end{pmatrix}$$

$$\begin{pmatrix} \mathrm{Ca(OH)2} \\ \mathrm{CaCO3} \end{pmatrix}$$

$$a_a := X^{\langle 0\rangle}\cdot\frac{1}{K}$$

$$b_b := X^{\langle 1\rangle}\cdot\frac{1}{K^2}$$

$$c_c := X^{\langle 2\rangle}\cdot\frac{1}{K^3}$$

$$d_d := X^{\langle 3\rangle}\cdot\frac{1}{K^4}$$

$$H_{f\_lime} := -986560\frac{J}{mol}$$

$$H_{f\_CaCO3} := -1206900\frac{J}{mol}$$

INPUT

$$H_{O2} := 0.006469\frac{mol}{s}\left[\left[\int_{T_r}^{(T_{out})}\left(a_1 + b_1\cdot T + c_1\cdot T^2 + d_1\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_1}\right]$$

$$H_{O2} = 152.432 s^{-1} J$$

$$H_{N2} := 0.404998\frac{mol}{s}\left[\left[\int_{T_r}^{(T_{out})}\left(a_2 + b_2\cdot T + c_2\cdot T^2 + d_2\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_2}\right]$$

$$H_{N2} = 8.925\times\ 10^{3}\,s^{-1}\,J$$

$$H_{steam} := 0.013998\frac{mol}{s}\left[\left[\int_{T_r}^{T_{out}}\left(a_3 + b_3\cdot T + c_3\cdot T^2 + d_3\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_3}\right]$$

$$H_{steam} = -2.999\times\ 10^{3}\,s^{-1}\,J$$

$$H_{CO} := 0.271689\frac{mol}{s}\left[\left[\int_{T_r}^{T_{out}}\left(a_4 + b_4\cdot T + c_4\cdot T^2 + d_4\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_4}\right]$$

$$H_{CO} = -2.396\times\ 10^{4}\,s^{-1}\,J$$

$$H_{CO2} := 0.004104\frac{mol}{s}\left[\left[\int_{T_r}^{T_{out}}\left(a_5 + b_5\cdot T + c_5\cdot T^2 + d_5\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_5}\right]$$

$$H_{CO2} = -1.466\times\ 10^{3}\,s^{-1}\,J$$

$$H_{H2} := 0.06\frac{mol}{s}\left[\left[\int_{T_r}^{T_{out}}\left(a_6 + b_6\cdot T + c_6\cdot T^2 + d_6\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_6}\right]$$

$$H_{H2} = 1.242\times\ 10^{3}\,s^{-1}\,J$$

$$H_{CH4} := 0.008645\frac{mol}{s}\left[\left[\int_{T_r}^{T_{out}}\left(a_7 + b_7\cdot T + c_7\cdot T^2 + d_7\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_7}\right]$$

$$H_{CH4} = -245.997s^{-1}\,J$$

$$H_{feed\_water} := 0.012\frac{mol}{s}\left[\left[\int_{T_r}^{T_a}\left(a_8 + b_8\cdot T + c_8\cdot T^2 + d_8\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_8}\right]$$

$$H_{feed\_water} = -3.427\times\ 10^{3}\,s^{-1}\,J$$

$$H_{lime} := 0.004104\frac{mol}{s}\left[\left[\int_{T_r}^{T_a}\left(a_{a_0} + b_{b_0}\cdot T + c_{c_0}\cdot T^2 + d_{d_0}\cdot T^3\right)dT\right]\frac{J}{mol} + H_{f\_lime}\right]$$

$$H_{lime} = -4.048\times\ 10^{3}\,s^{-1}\,J$$

$$\Delta H_{in} := H_{O2} + H_{N2} + H_{steam} + H_{CO} + H_{CO2} + H_{H2} + H_{CH4} + H_{feed\_water} + H_{lime}$$

$$\Delta H_{in} = -2.583\times\ 10^{4}\,s^{-1}\,J$$

OUTPUT

$$H_{O2} := 0.006469\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a+2K)}\left(a_1 + b_1\cdot T + c_1\cdot T^2 + d_1\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_1}\right]$$

$$H_{O2} = 1.038s^{-1}J$$

$$H_{N2} := 0.404998\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a+2K)}\left(a_2 + b_2\cdot T + c_2\cdot T^2 + d_2\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_2}\right]$$

$$H_{N2} = 60.971s^{-1}J$$

$$H_{CO} := 0.271689\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a+2K)}\left(a_4 + b_4\cdot T + c_4\cdot T^2 + d_4\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_4}\right]$$

$$H_{CO} = -2.999\times 10^4 s^{-1}J$$

$$H_{H2} := 0.06\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a+2K)}\left(a_6 + b_6\cdot T + c_6\cdot T^2 + d_6\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_6}\right]$$

$$H_{H2} = 8.742s^{-1}J$$

$$H_{CH4} := 0.008645\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a+2K)}\left(a_7 + b_7\cdot T + c_7\cdot T^2 + d_7\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_7}\right]$$

$$H_{CH4} = -644.792s^{-1}J$$

$$H_{out\_water} := 0.029998\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a+2K)}\left(a_8 + b_8\cdot T + c_8\cdot T^2 + d_8\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_8}\right]$$

$$H_{out\_water} = -8.563\times 10^3 s^{-1}J$$

$$H_{CaCO3} := 0.004104\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a+2K)}\left(a_{a_1} + b_{b_1}\cdot T + c_{c_1}\cdot T^2 + d_{d_1}\cdot T^3\right)dT\right]\frac{j}{mol} + H_{f\_CaCO3}\right]$$

$$H_{CaCO3} = -2.417\times 10^9 s^{-1}J$$

$$\Delta H_{out} := H_{O2} + H_{N2} + H_{CO} + H_{H2} + H_{CH4} + H_{out\_water} + H_{CaCO3}$$

$$\Delta H_{out} = -2.417\times 10^9 s^{-1}J$$

$$Heat\_duty := \Delta H_{out} - \Delta H_{in}$$

$$Heat\_duty = -2.417 \times 10^{9}\, s^{-1}\, J$$

CATALYTIC REFORMER

CATALYTIC REFORMER		
	INPUT	OUTPUT
COMPONENT	STREAM...	STREAM...
O2	0.006469	0.006469
N2	0.404998	0.404998
CO	0.272	0.28
H2	0.06	0.085
CH4	0.008645	0.000233
STEAM	0.008645	0.000233

$$H_{O2} := 0.006469\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a+2K)}\left(a_1 + b_1 \cdot T + c_1 \cdot T^2 + d_1 \cdot T^3\right) dT\right]\frac{J}{mol} + \Delta H_{f_1}\right]$$

$$H_{O2} = 1.038 s^{-1}\, J$$

$$H_{N2} := 0.404998\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a+2K)}\left(a_2 + b_2 \cdot T + c_2 \cdot T^2 + d_2 \cdot T^3\right) dT\right]\frac{J}{mol} + \Delta H_{f_2}\right]$$

$$H_{N2} = 60.971 s^{-1}\, J$$

$$H_{CO} := 0.272\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a+2K)}\left(a_4 + b_4 \cdot T + c_4 \cdot T^2 + d_4 \cdot T^3\right) dT\right]\frac{J}{mol} + \Delta H_{f_4}\right]$$

$$H_{CO} = -3.002 \times 10^{4}\, s^{-1}\, J$$

$$H_{H2} := 0.06\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a+2K)}\left(a_6 + b_6 \cdot T + c_6 \cdot T^2 + d_6 \cdot T^3\right) dT\right]\frac{J}{mol} + \Delta H_{f_6}\right]$$

$$H_{H2} = 8.742 s^{-1}\, J$$

$$H_{CH4} := 0.008645\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a+2K)}\left(a_7 + b_7 \cdot T + c_7 \cdot T^2 + d_7 \cdot T^3\right) dT\right]\frac{J}{mol} + \Delta H_{f_7}\right]$$

$$H_{CH4} = -644.792 s^{-1}\, J$$

$$H_{steam} := 0.008645\frac{mol}{s}\left[\left[\int_{T_r}^{T_s}\left(a_3 + b_3 \cdot T + c_3 \cdot T^2 + d_3 \cdot T^3\right) dT\right]\frac{J}{mol} + \Delta H_{f_3}\right]$$

$H_{steam} = -2.068\times 10^{3}\,s^{-1}\,J$

$\Delta H_{in} := H_{O2} + H_{N2} + H_{CO} + H_{H2} + H_{CH4} + H_{steam}$

$\Delta H_{in} = -3.266\times 10^{4}\,s^{-1}\,J$

$$H_{O2} := 0.006469\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a)}\left(a_1 + b_1\cdot T + c_1\cdot T^2 + d_1\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_1}\right]$$

$$H_{N2} := 0.404998\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a)}\left(a_2 + b_2\cdot T + c_2\cdot T^2 + d_2\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_2}\right]$$

$$H_{CO} := 0.28\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a)}\left(a_4 + b_4\cdot T + c_4\cdot T^2 + d_4\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_4}\right]$$

$H_{CO} = -3.092\times 10^{4}\,s^{-1}\,J$

$$H_{H2} := 0.085\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a)}\left(a_6 + b_6\cdot T + c_6\cdot T^2 + d_6\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_6}\right]$$

$H_{H2} = 7.43s^{-1}\,J$

$$H_{CH4} := 0.0002334\frac{mol}{s}\left[\left[\int_{T_r}^{(T_a)}\left(a_7 + b_7\cdot T + c_7\cdot T^2 + d_7\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_7}\right]$$

$H_{CH4} = -17.432s^{-1}\,J$

$$H_{steam} := 0.091\frac{mol}{s}\left[\left[\int_{T_r}^{T_a}\left(a_3 + b_3\cdot T + c_3\cdot T^2 + d_3\cdot T^3\right)dT\right]\frac{J}{mol} + \Delta H_{f_3}\right]$$

$\Delta H_{out} := H_{O2} + H_{N2} + H_{CO} + H_{H2} + H_{CH4} + H_{steam}$

$\Delta H_{out} = -5.289\times 10^{4}\,s^{-1}\,J$

$Heat\_duty := \Delta H_{out} - \Delta H_{in}$

$Heat\_duty = -2.023\times 10^{4}\,s^{-1}\,J$

CHAPTER FIVE

5.0 FLOW SHEET /FLOW DIAGRAM

5.1 Introduction

This chapter covers the preparation and presentation of the process flow- sheet. The flow-sheet is the key document in process design. It shows the arrangement of the equipment selected to carry out the process; the steam connections; stream flow-rates and compositions; and the operating conditions. It is a diagamatic model of the process.

The flow-sheet will be used by the specialist design groups as the basis for their design. This will include piping, instrumentation, and equipment design and plant layout. It will also be used by operating personnel for the preparation of operating manuals and operator training. During plant stat-up and subsequent operation, he flow-sheet forms the basis for comparison of operating performance with design.

The flow-sheet is drawn up from material balances made over the complete process and each individual unit. Energy balances are also made to determine the energy flows and the service requirements.

Manual flow-sheeting calculations can be tedious and tome consuming when the process is large or complex, and computer-aided flow-sheeting programs are being increasingly used to facilitate this stage of process design. Their use enables the designer to consider different processes, and more alternative processing schemes, in his search for the best process and optimum process conditions. Some of the proprietary flow-sheeting program is presented in detail and listed in the appendices.

5.2 Flow Sheet Presentation

As the process flow-sheet is the definitive document on the process, the presentation must be clear, comprehensive, accurate and complete. The various types of flow-sheet are discussed below:

(i) block flow-sheet

(ii) process flow-sheet (pictorial presentation)

(iii) engineering flow-sheet

5.2.1 Block Diagrams (Block Flow-Sheet)

A block diagram is the simplest form of presentation. Each block can represent a single piece of equipment or a complete stage in the process. They are useful for showing simple processes with complex processes. Their use is limited to showing the overall process broken down into its principal stages. In the block diagram, each block represents the equipment for a complete reaction stage, e.g. the reactor, separators and distillation columns.

Block diagrams are useful for representing a process in a simplified form I reports and textbooks, but have only a limited use as engineering documents.

The streams flow-rates and composition can be shown on the diagram adjacent to the stream lines, when only a small amount of information is to be shown, or tabulated separately,

The blocs can be of an shape, but it is usually convenient to use a mixture of squares and circles, drawn wit a template.

5.2.2 Pictorial Representation (Process Flow-Sheet)

On the detailed flow-sheets used for design and operation, the equipment is normally drawn in a stylised pictorial form. For tender documents or company brochures, actual scale drawings of the equipment are sometimes used, but it is more usual to use a simplified representation. The symbols given in British Standard, BS 1553 (1977) "Graphical Symbols for General Engineering" part 1, "Piping Systems and Plant' are recommended; though most design offices use their own standard symbols. The American National Standards Institute (ANSI) has also published a set of symbols for use on flow-sheets. Austin (1979) has compared the British Standard, ANSI, and some propriety flow-sheet symbols.

In Europe, the German standard organization has published a set of guide rules and symbols for flow-sheet presentation, DIN 28004 (1988). This is available in an English translation from the British Standards Institution.

.2.3 Engineering Flow Sheet

The first two types of flow-sheet does not show the real equipment and their specifications, thus a need for a flow-sheet that will show the actual equipment. This type of flow-sheet is called engineering flow-sheet. It gives the exact equipment and show the mass and energy transfer from one equipment to another.

5.3 Information To Be Included

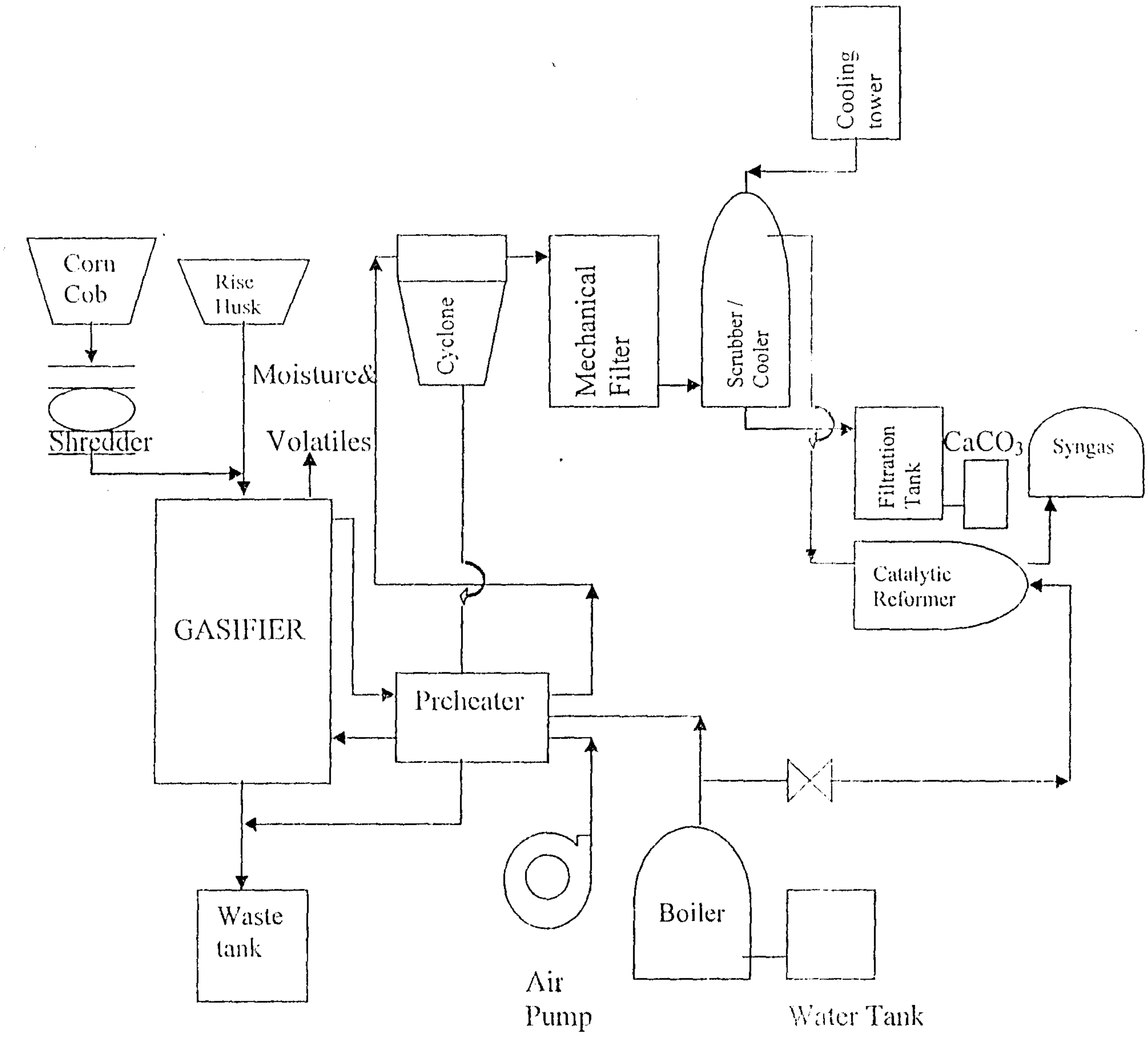
The amount of information shown on a flow-sheet will depend on the custom and practice of the particular design office. The list given bellow has therefore been divided into essential items and optional items. The essential items must always be shown; the optional items added to the usefulness of the flow-sheet but are not always included.

5.4 Essential Information

1. streams composition, either:
    - (i) The flow-rate of each individual component, kg/h, which is preferred or
    - (ii) The stream composition as a weight fraction.
2. Total stream flow-rate, kg/h.
3. Stream temperature, degrees Celsius preferred.
4. Nominal operating pressure (the required operating pressure).

5.5 Optional Information

1. molar percentages composition
2. physical property data, mean values for the stream, such as:
   - (i) density, g/m3
   - (ii) viscosity, mNs/m3
3. Stream name, a brief, one or two-word, description of the nature of the stream, for example "ACETONE COLUMN BOTTOMS".
4. Stream enthalpy, kg/h.



Pictorial representation of the gasifier plant

CHAPTER SIX

6.0 EQUIPMENT DESIGN

GASIFIER

The gasifier employed the downdraft type with double design. The reason for this is for suitability and capacity. the downdraft gasifier has been proven to be the most suitable for the feed stock choosen and double throat design has been choosen here to increase capacity.

Guidlines for Downdraft Gasifier Design

Dimensioning of this gasifier is closely related to the "heart load" concept. The heart load Bg is defined as the amount of producer gas reduced to normal (P,T0 conditions,divided by the surface area of the throat at the smallest circumference and is usually expressed in m3/cm2/hr

Alternatively the heart load can be expressed as the amount of dry fuel consumed divided by the surface area of the narrowest constriction (Bs), in which case hearth load is expressed in kg/cm2/hr. According to information proevided Bg may reach maximum value of about 0.9(Bs=0.36) in continuous operation(single throat) in a good 'imbert type" gasifier.

Table...... Suitable nozzles for wood gas generator operating with 4-cycle engines

Hogher values of Bg give rice to extreme pressure drop over the reduction zone.

Designing this type of gasifier now boils down to estimating the maximum amount of gas needed.

The Swedish academy of Engineering sciences also presents empirical data with regard to the height of the nozzles above the narrowest constriction. the daimeter of the nozzle opening ring as well as the suitable nozzles for different capacities

dt	dn	n
mm	mm	
70	10.5	3
80	9	5
90	10	5
100	11	5
120	12.7	5
130	13.5	5
150	15	5
170	14.3	7
190	16	7
220	18	7
270	22	7
300	24	7

dt= daimater of throat

dn=daimeter of nozzle

n= no of nozzles

Vanselaar (46) compares the design characteristic of similar gasifiers that were available during the second world war and came up with the following distinctions for "no throat", "single throat", and double throat"

The maximum allowable heart load Bs max= 0.03, 0.11, and 0.4 for no throat, siungle throat, and double throat respectivly

The nozzle inlet velocity = 30to35m/s

throat inclination= 45degree to 60 degree

the hearth daimeter at the inlet should be 10cm larger than the throat daiumater in the case of single throat and 20cm larger than the daimater of the smallest constriction in the case of double throat.

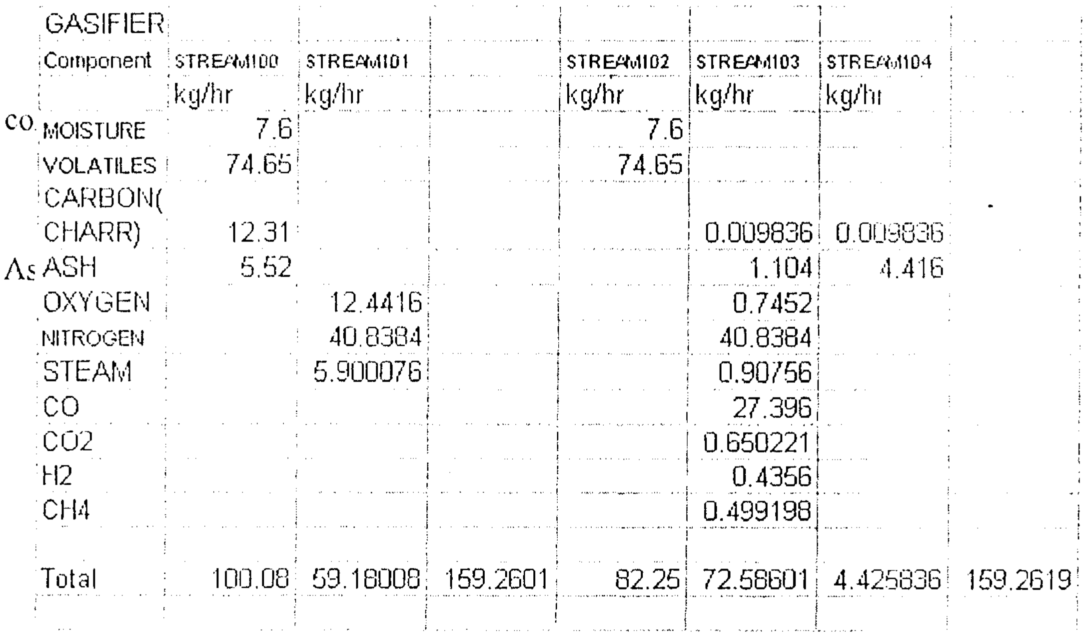
LOWER SECTION (oxidation and reduction zones)

Let the hearth load Bg be =0.8m3/cm2/hr

The height of the reductipon zone should be moer than 20cm the average height of the gasifier (The gasifier reviewed was 32cm)

CO

As



GASIFIER							
Component	STREAM100	STREAM101		STREAM102	STREAM103	STREAM104	
	kg/hr	kg/hr		kg/hr	kg/hr	kg/hr	
MOISTURE	7.6			7.6			
VOLATILES	74.65			74.65			
CARBON(CHARR)	12.31				0.009836	0.009836	
ASH	5.52				1.104	4.416	
OXYGEN		12.4416			0.7452		
NITROGEN		40.8384			40.8384		
STEAM		5.900076			0.90756		
CO					27.396		
CO2					0.650221		
H2					0.4356		
CH4					0.499198		
Total	100.08	59.18008	159.2601	82.25	72.58601	4.425836	159.2619

Note the residence time in the Gasifier if that of the limiting reaction that is the reaction that occurs at the slowest rate, and this has been determined in the MB as

$\tau := 341.946s$

Densities of the components on the bottom section of the gasifier at exit temperatuer of 950 degree celsuis.

Note the Gasifier operates essentially as an open system and thus parameters for the inlet stream(feed) ws employed in its sizing

$\rho_c := 1642.06\frac{kg}{m^3}$ $MM_c := 12.01\frac{kg}{mole \cdot 10^3}$

$\rho_{O2} := 0.21994\frac{kg}{m^3}$ $MM_{O2} := 32\frac{kg}{mole \cdot 10^3}$

$\rho_{N2} := 0.19253\frac{kg}{m^3}$ $MM_{N2} := 28.013\frac{kg}{mole \cdot 10^3}$

$\rho_{steam} := 0.17949\frac{kg}{m^3}$

volumetric flow rate of the component entering the bottom section of the gasifier

$$v_C := \frac{(12.31 + 5.52)\frac{kg}{hr}}{1642.06\frac{kg}{m^3}}$$

$v_C = 3.016\times 10^{-6}\, s^{-1}\cdot m^3$

$$v_{O2} := \frac{12.4416\frac{kg}{hr}}{\rho_{O2}}$$

$v_{O2} = 0.016 s^{-1}\cdot m^3$

$$v_{N2} := \frac{40.8384\frac{kg}{hr}}{\rho_{N2}}$$

$v_{N2} = 0.059 s^{-1}\cdot m^3$

$$v_{steam} := \frac{5.900076\frac{kg}{hr}}{\rho_{steam}}$$

$v_{steam} = 9.131\times 10^{-3}\, s^{-1}\cdot m^3$

Total volumetric flow rate through the bottom section

$v := v_C + v_{O2} + v_{N2} + v_{steam}$

$v = 0.084s^{-1} \cdot m^3$

Required volume of the lower section

$V := v \cdot \tau$

$V = 28.644m^3$

$B_s := 0.20 \frac{m^3}{cm^2 \cdot hr}$

Area or narrowest constriction

$A_{t1} := \frac{v}{B_s}$

$A_{t1} = 0.151m^2$

$A_{t1} = \frac{\pi \cdot D_{t1}^2}{4}$

$D_{t1} := \sqrt{\frac{4 \cdot A_{t1}}{\pi}}$

$D_{t1} = 0.438m$

Daimeter of the second contriction(considering the fact that this design is of a larger scale than the ones upon which we base our premise,we will scale up all associated parameters)

$D_{t2} := D_{t1} + 150cm$

$D_{t2} = 1.938\text{m}$

$$A_{t2} := \frac{\pi \cdot D_{t2}^{2}}{4}$$

$A_{t2} = 2.95\,\text{m}^2$

Daimeter of hearth inlet

$D_h := D_{t1} + 250\text{cm}$

$D_h = 2.938\text{m}$

$$A_h := \frac{\pi \cdot D_h^{2}}{4}$$

$A_h = 6.78\,\text{m}^2$

the angle of inclnation of each frustrum is put at 45 degree

Let the volume of the conical base(2 Frustrums one sitting on the other) labeled 'a' be 1/3 of the total vol of the bottom section, thus

$$V_a := V \cdot \frac{1}{3}$$

$V_a = 9.548\text{m}^3$

Let the volume and dimenssion of the two frustrums be the same, thus

$$V_{a\_top} = V_{a\_bot} = \frac{V_a}{2}$$

$$V_{a\_top} := \frac{V_a}{2}$$

volume of frustrum

$$V_{frustrum} = \frac{\pi \cdot h}{3} \cdot \left(r_2^{2} + r_2 \cdot r + r^{2}\right)$$

$h := 1$

$r := \frac{D_{t1}}{2}$ $\qquad$ $r_2 := \frac{D_{t2}}{2}$

$$\mathrm{root}\left[V_{a\_top} - \frac{\pi \cdot h}{3} \cdot \left(r_2^{\,2} + r_2 \cdot r + r^2\right), h\right] =$$

thus

$h := 3.80\ m$

The volume of the section lebeled 'b', the oxidation zone, this section consist of a vertical cylindrical section and an inverted frustrum.

let the volume of this section take another 1/3 of the volme of the lower section of the gasifier.

that is; $V\ \frac{1}{3} \cdot V$

Also let the cylindrical portion be 1/3 of the oxidation zone

$V_{cyld} := V_b \cdot \frac{1}{3}$

Volume of cylinder=

$V_{cyld} = \frac{\pi \cdot D_{t2}^{\,2} \cdot h_{cyld}}{4}$ $\qquad$ $h_{cyld} := 1$

$$\mathrm{root}\left(V_{cyld} - \frac{\pi \cdot D_{t2}^{\,2} \cdot h_{cyld}}{4}, h_{cyld}\right) =$$

$h_{cyld} := 1.079r$

for the inverted frustrum the following applies

$$V_f := \frac{2}{3} \cdot V_b \qquad r_h := \frac{D_h}{2}$$

$$\mathrm{root}\left[V_f - \frac{\pi \cdot h}{3} \cdot \left(r_h^{2} + r_h \cdot r_2 + r_2^{2}\right), h\right] = 1.345\,\mathrm{m}$$

$h := 1.345\mathrm{r}$

The remaining volume of the bottom section represent a transion region and could be empty at some stage in the process. The reason for this is that the upper section of the gasifier is operated as a batch process and is also separated from the lower section by a partition. the implication of this is that there is a time for feed entering the bottom section.

$V_h := \frac{}{3} \cdot V$

the following applies to this transition section;

$h_h := 1$

volume of a cylinder

$$\mathrm{root}\left(V_h - \frac{\pi \cdot D_h^{2} \cdot h_h}{4}, h_h\right) = \blacksquare$$

$h_h := 1.408\mathrm{r}$

The Upper section (Drying and Pyrolysis zones)

This section is essencially a vertical cylinder

Note , this section operates as an open system and due to the rapid volatilization, resulting from high temperatures, the volume of the unit is approximattely the volume of the the product after volatilization + the vome of moisture, thus

$$v_{top} := \frac{(7.6 + 12.31 + 5.52)\frac{\mathrm{kg}}{\mathrm{hr}}}{\rho_c}$$

$$v_{top} = 4.302 \times 10^{-6}\,\mathrm{s}^{-1} \cdot \mathrm{m}^{3}$$

$V_{top} := v_{top} \cdot \tau$

$V_{top} = 1.471 \times 10^{-3}\, m^3$ $\qquad\qquad h_T := 1$

$$\mathrm{root}\left(V_{top} - \frac{\pi \cdot D_h{}^2 \cdot h_T}{4}, h_T \right) = \blacksquare$$

$h_T := 2.17 \cdot 10^{-4} m$

Allowance between the Hearth and the outer wall is put at 10% of hearth daimeter, thus

$D_{wall} := D_h + 0.1 \cdot D_h$

$D_{wall} = 3.232m$

Below the reduction zone is a movable grate for removal of ash

PREHAETER DESIGN (Tube and Shell Heat Exchanger)

All the syn-gas from the gasifier is channeled to the preheater for heating the gasifing agent to a desire temperature.

From the Energy balance heat load is m*Cp*ΔT is the as

$$Heat\_load := 6.036 \times 10^3 \frac{J}{s}$$

But

$Heat\_load = U \cdot A \cdot \theta_m \cdot F$

$$U = \left(\frac{1}{h_i} + \frac{1}{h_o} \right)$$

$$h_i := 150 \frac{J}{s \cdot m^2 \cdot K}$$

$$h_o := 25 \frac{J}{s \cdot m^2 \cdot K}$$

$$U := \left(\frac{1}{h_i} + \frac{1}{h_o} \right)^{-1}$$

$$U = 21.429 \frac{J}{s \cdot m^2 \cdot K}$$

$$\theta_m := 622.2 \cdot K$$

The value of the correction factor for the mean temperature diffrence is obtained from the chart for 1 shell 2 tube pass. Ref;(R&C vol 1).

$$F := 0.85$$

$$A := 1$$

$$\mathrm{root}\left(Heat\_load - U \cdot A \cdot \theta_m \cdot F, A\right) = \blacksquare$$

Thus the surface area of the tube is

$$A := 0.533 m^2$$

Chosing the tube material as alluminum of daimeter 50mm (ref;R&C vol 1), the following applies

$$D_{tube} := 5cm$$

$$A = \pi \cdot D_{tube} \cdot L_{tube}$$

$$L_{tube} := \frac{A}{\pi \cdot D_{tube}}$$

$$L_{tube} = 3.393m$$

with 2 tube passes the span is approximately

$$span\_lenth := \frac{l_{tube}}{2}$$

$$span\_lenth = 1.697\,m$$

THE SHELL SIDE

if the residence time for both hot and cold fluid is the same the follwing is obtainable;
Note the cold fluid flows through the tube side and the hot fluid through the shell side

Volumetric flow rate of fluid of the cold fluid

$$\rho_{O2} := 0.21994\,\frac{kg}{m^3} \qquad \rho_{steam} := 0.17949\,\frac{kg}{m^3}$$

$$\rho_{N2} := 0.19253\,\frac{kg}{m^3}$$

$$v_{O2} := \frac{12.4410\,\frac{kg}{hr}}{\rho_{O2}}$$

$$v_{O2} = 0.016\,s^{-1}\cdot m^3$$

$$v_{N2} := \frac{40.84\,\frac{kg}{hr}}{\rho_{N2}}$$

$$v_{N2} = 0.059\,s^{-1}\cdot m^3$$

$$v_{steam} := \frac{5.900076\,\frac{kg}{hr}}{\rho_{steam}}$$

$$v_{steam} = 9.131\times 10^{-3}\,s^{-1}\cdot m^3$$

volumetric flow rate of tube side

$v_{tube} := v_{O2} + v_{N2} + v_{steam}$

$v_{tube} = 0.084\,s^{-1} \cdot m^{3}$

Volume of tube

$$V_{tube} := \frac{\pi \cdot D_{tube}^{\;2}}{4} \cdot L_{tube}$$

$V_{tube} = 6.663 \times 10^{-3}\ m^{3}$

Residence time of fluid in the tube

$$\tau := \frac{V_{tube}}{v_{tube}}$$

$\tau = 0.08\,s$

volumetric flow rate of fluid in the shell side

$\rho_{CO2} := 0.30248 \frac{kg}{m^{3}}$ $\rho_{CO} := 0.19252 \frac{kg}{m^{3}}$ $\rho_{H2} := 2.0086 \cdot 10^{\;2} \frac{kg}{m^{3}}$

$\rho_{CH4} := 0.15984 \frac{kg}{m^{3}}$

$$v_{O2} := \frac{0.7452\frac{kg}{hr}}{\rho_{O2}}$$

$$v_{N2} := \frac{40.84\frac{kg}{hr}}{\rho_{N2}}$$

$v_{O2} = 9.412\times 10^{-4}\, s^{-1}\cdot m^3$

$v_{N2} = 0.059\, s^{-1}\cdot m^3$

$$v_{steam} := \frac{0.90756\frac{kg}{hr}}{\rho_{steam}}$$

$$v_{CO} := \frac{27.396\frac{kg}{hr}}{\rho_{CO}}$$

$v_{steam} = 1.405\times 10^{-3}\, s^{-1}\cdot m^3$

$v_{CO} = 0.04\, s^{-1}\cdot m^3$

$$v_{CO2} := \frac{0.650221\frac{kg}{hr}}{\rho_{CO2}}$$

$$v_{H2} := \frac{0.4356\frac{kg}{hr}}{\rho_{H2}}$$

$v_{CO2} = 5.971\times 10^{-4}\, s^{-1}\cdot m^3$

$v_{H2} = 6.024\times 10^{-3}\, s^{-1}\cdot m^3$

$$v_{CH4} := \frac{0.499198\frac{kg}{hr}}{\rho_{CH4}}$$

$v_{CH4} = 8.675\times 10^{-4}\, s^{-1}\cdot m^3$

$v_{shell} := v_{O2} + v_{N2} + v_{steam} + v_{CO} + v_{CO2} + v_{H2} + v_{CH4}$

$v_{shell} = 0.108\, s^{-1}\cdot m^3$

Volume of shell

$V_{shell} := v_{shell}\cdot \tau$

$V_{shell} = 8.613\times 10^{-3}\, m^3$

some design allowance is necessary between the tube span and shell span(length), thus

$L_{shell} := 1.8m$

For a cylindrical shell design

$$V_{shell} = \frac{\pi \cdot D_{shell}^{2}}{4} \cdot L_{shell} \qquad\qquad D_{shell} := 1$$

$$\mathrm{root}\left(V_{shell} - \frac{\pi \cdot D_{shell}^{2}}{4} \cdot L_{shell}, D_{shell} \right) = \blacksquare$$

$D_{shell} := 0.079m$

CYCLONE

The residnece time of the gas in the cyclone is fixed at 2 seconds, and the volumetric flow rate of gas in the Cyclone is the same as that of the shell side of the Preheater.

$\tau := 2s$

$v_{cyclone} := v_{shell}$

Volume of Cyclone

$V_{cyclone} := v_{shell} \cdot \tau$

$V_{cyclone} = 0.217m^{3}$

Radial velocity of gas into the cyclone

$$v_{radial} = \frac{\left(\rho_{particle} - \rho_{gas}\right) \cdot r \cdot w^{2} \cdot d^{2}}{18 \cdot \mu}$$

EQUIPMENT DESIGN

CYCLONE

The residence time of the gas in the cyclone is fixed at 2 minutes, and the volumetric flow rate of gas in the cyclone is the same as that of the shell side of the pre heater.(Richardson and Coulson Volume 1)

$\tau := 2min$

$V_{cyclone} = V_{shell}$

Volume of cyclone

$V_{cyclone} = V_{Shell}^{\tau}$

$V_{cyclone} = 0.217m^3$ (from the pre heater)

Radial velocity of gas into the cyclone

$$V_{Radial} = \frac{(\rho_{particle} - \rho_{gas}).r.W^2.d^2}{18\mu}$$

Where:

$\rho_{particle}$ = Particle density,

ρ_{gas} = Average density of gas

r = radial distance

d = particle diameter

μ = gas viscosity [Synthesis Gas (CO)]

d = 1m (Anil K, 1986)

$W = 3\,\frac{rad}{s}$ *(Angular velocity, for a diameter of 1m of angular velocity ranges from 2 – 5 rads)*

R = 0.8m *(Radial distance, for 3, radial distance is 0.8m rads/s)*

μ = 0.00166poise [Synthesis gas (CO)]

$\rho_{partial} = \rho c$

$\rho_{gas} = \rho cc$

$$V_{radial} = \frac{(\rho_{particle} - \rho_{gas}).r.W^2.d^2}{18.\mu}$$ *(Richardson and Coulson Vol. 1, wikipedia.org)*

$\rho_{particle} = \rho_c$

$\rho_{gas} = \rho_{cc}$

$$\rho_{av(gas)} = \frac{(mm.co \times \rho.co) + (mm\,H_2 \times \rho H_2)}{mm.co + mm\,H_2}$$

$$= \frac{(28 \times 0.19252) + (2 \times 2.0086 \times 10^{-2})}{(28+2)}$$ *(Richardson and Coulson Vol. 1, wikipedia.org)*

$$= \frac{5.39056 + 6.04017}{30}$$

$$\rho_{av(gas)} = \frac{5.43073}{30} = 0.181$$

$$\rho_{av\,(particle)} = \frac{(mm.c \times \rho_c) + (mm._{Ash})}{(mm.c \mid mm.Ash)}$$

$$\rho_{Ash} = 2.044\text{g/cm}^2 = \begin{array}{c|c|c} 2.044 & \text{g} & (1\text{cm})^3 \\ \hline & \text{cm}^3 & (100\text{m})^3 \end{array}$$

$$\rho_{av(paritcle)} = \frac{(12 \times 1642.06 \times 10^{-3}) + (56.0983 \times 2.044 \times 10^{-4})}{12 + 56.0983}$$ *(Richardson and Coulson Vol. 1, wikipedia*

$$= \frac{197.0472 + 0.011466}{67}$$

$$= \frac{197.05867}{68.0983} = 3.3918\ \text{Kg/m}^3$$

$$V_{radial} = \frac{(\rho_{particle} - \rho_{gas}).r.W^2.d^2}{18.\mu}$$

$$= \frac{(3.3918 - 0.181)\,0.8.\,3^2\,(1)^2}{18.0 \times 0.00166}$$

$$= \frac{3.2108 \times 0.8 \times 9 \times 1}{0.2988}$$

$$V_{radiai} = 77.3686\text{s}^{-1}\ \text{m}$$

Oriface inlet diameter and surface area

$$A_{orifice} = \frac{V_{Cyclone}}{V_{radial}} \quad \frac{0.217}{77.3686} = 0.02805\text{m}^2$$

$$D_{orifice} = \sqrt{\frac{4A_{orifice}}{\pi}}$$

$$= \sqrt{\frac{4 \times 0.02805}{3.142}}$$

$$\sqrt{\frac{0.1122}{3.142}}$$

=

$= \sqrt{0.0367097}$

$D_{orifice} = 0.18897m$

= 0.189m

Cyclone diameter, top and base and cyclone height.

Assumption: the cyclone is in the form of a frustrum. Let, the top diameter be 40cm and base diameter be 15cm. This is to minimize material and cost of construction.

r_{top} = 20cm

ho = 10cm

r_{base} = 7.5cm

$$root\left[V_{cyclone} - \frac{1}{3}.\pi.\left(ho.r_{top}^2 - 0.3.ho.r_{base}^2\right), ho\right]\ [standard, wikipedia.org]$$

$$h = \sqrt{0.217 - \frac{1}{3}\pi(3.142)[0.1 \times (0.2)^2 - 0.3 \times 0.1(0.075)^2]0.1}$$

$$h = \sqrt{0.217 - 0.9426\,[0.004 - 0.00016875]0.1}$$

$$h = \sqrt{}(0.217 - 0.9426\,[0.00383125]0.1)$$

$$h = \sqrt{0.217 - 0.00361}$$

$$h = \sqrt{0.21339}$$

$$h = 0.4619m$$

h = (0.4619 - $\frac{1}{3}$ x 0.4619)m (because it is a cone, you need to subtract 1/3 of the height from the calculated

= 0.4619 -0.1386

h = 0.3233m

h_{design} = 1.3 h

=1.3 x 0.3233

= 0.42029m

h_{design} = 0.4203m

where :

$\rho_{particle}$= particle density, ρ_{gas}=ave density of gas, r=radial distance, d=particle diameter

μ=gas viscosity.

$d := 1mm$ $\quad w := 3\frac{rad}{s}$ $\quad r := 0.8m$ $\quad \mu := 0.0001667poise$

$\rho_{particle} := \rho_c$ $\quad \rho_{gas} := \rho_{CC}$

$$v_{radial} := \frac{\left(\rho_{particle} - \rho_{gas}\right) \cdot r \cdot w^2 \cdot d^2}{18 \cdot \mu}$$

$$v_{radial} = 39.515s^{-1} \cdot m$$

Oriface inlet Diameter and Surface area

$$A_{oriface} := \frac{v_{cyclone}}{v_{radial}}$$

$$A_{oriface} = 2.74 \times 10^{-3} m^2$$

$$D_{oriface} := \sqrt{4 \cdot \frac{A_{oriface}}{\pi}}$$

$$D_{oriface} = 0.059m$$

Cyclone Diameter, Top and Base and Cyclone Hieght.

Assumption : the cyclone is in the form of a fustrum

Let , The Top diameter be 40cm and Base diameter be 15cm

$r_{top} := 40cm$ $\quad$ guess $\quad h_0 := 10cm$

$r_{base} := 15cm$

$$root\left[V_{cyclone} - \frac{1}{3} \cdot \pi \cdot \left(h_0 \cdot r_{top}{}^2 - 0.3 \cdot h_0 \cdot r_{base}{}^2\right), h_0\right] = 1.349m$$

thus

$h := (1.349 - 0.3 \cdot 1.349)m$

$$h = 0.944m$$

$$h_{design} := 1.3 \cdot h$$

$$h_{design} = 1.228m$$

BOILER

$$m_{rate\_water} := 6.460596\frac{kg}{hr}$$

Note ; we shall employ the same residence time as in the gasifier for continuity purpose

$$\tau := 341.946s$$

$$m_{water} := m_{rate\_water} \cdot \tau$$

$$m_{water} = 0.614kg$$

$$\rho_{water} := 1000\frac{kg}{m^3}$$

$$V_{water} := \frac{m_{water}}{\rho_{water}}$$

$$V_{water} = 6.137 \times 10^{-4}\, m^3$$

The shape of the boiler is that of a horizontal cylinder

a vapor volume is to be provided to accomodate the steam produced before discharge, and this is put a half the liquid volume

$$V_{boiler} := V_{water} + 0.5 \cdot V_{water}$$

$$V_{boiler} = \pi \cdot r^2 \cdot L$$

Let , L=4r

$$r := \sqrt[3]{\frac{V_b}{4}}$$

$r = 0.042m$

$L := 4 \cdot r$

$L = 0.167m$

$L_{design} := L \cdot 1.3$

$L_{design} = 0.218m$

The boiler design is such that the heatting mechanism is in the form of a heating coil located inside
Bloiler power requirement;

and at the bottom of the Boiler.

$P := 4.629 \cdot 10^{3} W$ this is produced from an electrical power source.

Water level indicators'are attached to the equipment

Blow down Valves for removing solid particles that condense at the bottom of the boiler are

provided

SCRUBBER

Feed water check valves are are also provided.

Pressure drop;

The Scrubber is made of three sections; the converging section, the threat and the diverging section.

$P_{drop} = 10^{-6} \cdot v^{2} \cdot L$

v = gas_velocity

L = water_to_gas_volume_ratio

Liquid volume

$$m_{liq} := (1.094783 + 0.778032)\frac{kg}{hr} \cdot \tau$$

$$V_{liq} := \frac{m_{liq}}{\rho_{water}}$$

$V_{liq} = 1.779 \times 10^{-4} m^{3}$

$$v_{O2} := \frac{0.7452\frac{kg}{hr}}{\rho_{O2}}$$

$v_{O2} = 9.412 \times 10^{-4} s^{-1} \cdot m^{3}$

$$v_{N2} := \frac{40.84\frac{kg}{hr}}{\rho_{N2}}$$

$v_{N2} = 0.059 s^{-1} \cdot m^{3}$

$$v_{steam} := \frac{0.90756\frac{kg}{hr}}{\rho_{steam}}$$

$$v_{CO} := \frac{27.396\frac{kg}{hr}}{\rho_{CO}}$$

$$v_{steam} = 1.405\times 10^{-3}\, s^{-1}\cdot m^{3}$$

$$v_{CO} = 0.04\, s^{-1}\cdot m^{3}$$

$$v_{CO2} := \frac{0.650221\frac{kg}{hr}}{\rho_{CO2}}$$

$$v_{H2} := \frac{0.4356\frac{kg}{hr}}{\rho_{H2}}$$

$$v_{CO2} = 5.971\times 10^{-4}\, s^{-1}\cdot m^{3}$$

$$v_{H2} = 6.024\times 10^{-3}\, s^{-1}\cdot m^{3}$$

$$v_{CH4} := \frac{0.499198\frac{kg}{hr}}{\rho_{CH4}}$$

$$v_{CH4} = 8.675\times 10^{-4}\, s^{-1}\cdot m^{3}$$

$$v_{gas} := v_{O2} + v_{N2} + v_{steam} + v_{CO} + v_{CO2} + v_{H2} + v_{CH4}$$

$$v_{gas} = 0.108\, s^{-1}\cdot m^{3}$$

$$V_{gas} := v_{gas}\cdot \tau$$

$$V_{gas} = 37.028\, m^{3}$$

$$L := \frac{V_{liq}}{V_{gas}}$$

$$L = 4.804\times 10^{-6}$$

Gas Velocity

the radius of the throat is set for convinience at 0.5m

$$r_{throat} := 0.5m$$

$A_{throat} := \pi \cdot r_{throat}^{2}$

$A_{throat} = 0.785\,m^{2}$

$$vel_{throat} := \frac{v_{gas}}{A_{throat}}$$

$vel_{throat} = 0.138\,s^{-1} \cdot m$

$v := vel_{throat}$

$P_{drop} := 10^{-6} \cdot v^{2} \cdot L$

$P_{drop} = 9.132 \times 10^{-14}\,s^{-2} \cdot m^{2}$

Sizing

The gas is compressed in the scrubber 0.1 * its original volume

$V_{gas} := 0.1 \cdot V_{gas}$

$V_{scrubber} := V_{liq} + V_{gas}$

$V_{scrubber} = 3.703\,m^{3}$

Let the volume of theb three section be equal, thus

$$V_{diverging\_section} := \frac{1}{3} \cdot V_{scrubber}$$

$V_{diverging\_section} = 1.234\,m^{3}$

where the shape of this section is int the form of a fustrum

Let , The Top diameter be 60cm and Base diameter be 25cm

$r_{top} := 60cm$

$r_{base} := 25cm$

$$root\left[V_{diverging\_section} - \frac{1}{3} \cdot \pi \cdot \left(h_0 \cdot r_{top}^{2} - 0.3 \cdot h_0 \cdot r_{base}^{2}\right), h_0\right] = 3.454\,m$$

thus

$h := (3.454 - 0.3 \cdot 3.454)m$

$h = 2.418m$

$h_{design} := 1.3 \cdot h$

$h_{design} = 3.143m$

The throat takes the form of a vertical cylinder.

$V_{throat} := \frac{1}{3} \cdot V_{scrubber}$

Let the throat hieght be 4 * its radius

$r := \sqrt[3]{\frac{V_{throat}}{4 \cdot \pi}}$

$r = 0.461m$

$h_{throat} := 4 \cdot r$

$h_{throat} = 1.846m$

the dimension of the converging section is taken to be equal to that of diverging section

$V_{converging\_section} := \frac{1}{3} \cdot V_{scrubber}$

$\mathrm{root}\left[V_{converging\_section} - \frac{1}{3} \cdot \pi \cdot \left(h_0 \cdot r_{top}{}^2 - 0.3 \cdot h_0 \cdot r_{base}{}^2\right), h_0\right] = 3.454m$

thus

$h := (3.454 - 0.3 \cdot 3.454)m$

$h = 2.418m$

$h_{design} := 1.3 \cdot h$

$h_{design} = 3.143\,m$

Total hieght of the Scrubber

$h_{column} := h_{throat} + 2 \cdot h$

$h_{column} = 6.681\,m$

$h_{column\_design} := 1.3 \cdot h_{column}$

$h_{column\_design} = 8.686\,m$

Residence time

$Ca(OH)2+CO2\text{----->}CaCO3+H20$

$k := 0.199 \cdot 10^{2} \frac{m^{3}}{s \cdot mol}$

$\rho_{N2} = 0.193\,kg \cdot m^{-3}$

Let CAO be the initial concerntration of carbon (C) in the feed stream

Total moles of reactant = 1+1=2

$\rho_{CO2} = 0.302\,kg \cdot m^{-3}$

$MM_{CO2} := 44.01 \frac{gm}{mol}$

$\rho_{CO2\_mol} := \frac{\rho_{CO2}}{MM_{CO2}}$

$C_{A0} := \frac{\rho_{CO2\_mol}}{2}$

$C_{A0} = 3.436\,mol \cdot m^{-3}$

$X_A := 0.9$

$\xi_A := \frac{2-2}{2}$

$C_A := C_{A0} \cdot \left(\frac{1 - X_A}{1 + \xi_A \cdot X_A} \right)$

With equimolar concerntration of reactant A and B (CO2 and Ca(OH)2) the simple rate expression is given as:

$r_A := -k \cdot C_A^{\ 2}$ $\qquad C_{A0} = 3.436\,mol \cdot m^{-3}$

$$\tau := C_{A0} \cdot \int_0^{X_A} \frac{1}{-r_A}\, dX_A$$

$\tau = 144.766s$

MECHANICAL FILTER

$$v_c := \frac{0.004918\frac{kg}{hr}}{\rho_c} \qquad v_c = 8.319 \times 10^{-10}\, s^{-1} \cdot m^3$$

$$v_{ash} := \frac{0.552\frac{kg}{hr}}{\rho_c} \qquad v_{ash} = 9.338 \times 10^{-8}\, s^{-1} \cdot m^3$$

$$v_{O2} := \frac{0.7452\frac{kg}{hr}}{\rho_{O2}} \qquad v_{N2} := \frac{40.84\frac{kg}{hr}}{\rho_{N2}}$$

$$v_{O2} = 9.412 \times 10^{-4}\, s^{-1} \cdot m^3 \qquad v_{N2} = 0.059 s^{-1} \cdot m^3$$

$$v_{steam} := \frac{0.90756\frac{kg}{hr}}{\rho_{steam}} \qquad v_{CO} := \frac{27.396\frac{kg}{hr}}{\rho_{CO}}$$

$v_{steam} = 1.405 \times 10^{-3}\, s^{-1} \cdot m^3$

$v_{CO} = 0.04\, s^{-1} \cdot m^3$

$$v_{CO2} := \frac{0.650221 \frac{kg}{hr}}{\rho_{CO2}}$$

$$v_{H2} := \frac{0.4356 \frac{kg}{hr}}{\rho_{H2}}$$

$v_{CO2} = 5.971 \times 10^{-4}\, s^{-1} \cdot m^3$

$v_{H2} = 6.024 \times 10^{-3}\, s^{-1} \cdot m^3$

$$v_{CH4} := \frac{0.499198 \frac{kg}{hr}}{\rho_{CH4}}$$

$v_{CH4} = 8.675 \times 10^{-4}\, s^{-1} \cdot m^3$

$v_{gas} := v_c + v_{ash} + v_{O2} + v_{N2} + v_{steam} + v_{CO} + v_{CO2} + v_{H2} + v_{CH4}$

$v_{gas} = 0.108\, s^{-1} \cdot m^3$

$V_{gas} := v_{gas} \cdot \tau$

$V_{gas} = 15.676\, m^3$

Pressure Drop across Filter.

The inlet pressure into the filter is put at around 5atm

The estimated pressure drop from literature review is put at 20% of the inlet pressure.

outlet pressure

$0.8 \cdot 5atm = 4\, atm$

CATALYTIC REFORMER

COMPONENTS	INPUTS		OUTPUT	
	Mol/s	Kg/hr	Mol/s	Kg/hr
CH_4	0.008645	0.49932	0.0002334	0.0134784
Steam	0.008645	0.56052	0.0002334	0.0151344

Kmol=1000mol

Density of CH_4=0.06463Kg/m^3 MM$_{CH4}$=16.04Kg/kmol

Density of steam=0.17949Kg/m^3 MM$_{H2O}$=18.01Kg/kmol

$$Average\ density = \frac{(\rho_{CH_4} \times mmCH_4) + (\rho_{H_2O} \times mmH_2O)}{mmCH_4 + mmH_2O}$$

$$= \frac{(0.00463 \times 16.04) + (0.17949 \times 18.01)}{16.04 + 18.01}$$

$$= \frac{1.037 + 3.233}{34.05}$$

$$= 0.125\ kg/m^3$$

Volumetric flow rate

$$Volumetric\ flow\ rate(v) = \frac{M_{CH_4} + M_{H_2O}}{Average\ density}$$

$$= \frac{0.49932 + 0.56052}{0.124}$$

$$= 8.48\, m^3/s$$

Residence time of reaction in the catalytic reformer.

$$\tau = \frac{V_{CAO}}{F_{AO}}$$

$V = Volumetric\ flowrate$

$\tau = residue\ time$

$C_{AV} = Concentration$

$F_{AV} = molar\ flowrate$

$$\tau = \frac{8.48 \times 0.00241}{8.645 \times 10^{-3}}$$

$$= 2.4s$$

For a cylindrical design

$$V = \frac{\pi D^2 L}{4}$$

$$H = 2D$$

Geometric assumptions of a cylinder

$$L = H$$

$$H = 2D$$

$V = \frac{\pi D^2 2D}{4}$

$8.48 = \frac{\pi\ 2D^3}{4}$

$33.92 = \pi\ 2D^3$

$2D^3 = \frac{33.92}{\pi}$

$D^3 = \frac{10.797}{2}$

$D^3 = 5.399$

$D = \sqrt[3]{5.399}$

$D = 1.75\ m$

Since $H = 2D$

$H = 2 \times 1.75$

$= 3.5\ m$

CHAPTER SEVEN

7.0 OPTIMIZATION

7.1 OPTIMIZATION OF GASIFEIR PLANT

In optimizing, the first step is clearly to define the objective which is the criterion to be used to judge the performance of the system. In engineering design, the objective of optimizing any reactor or equipment must be an economical one. This is because for any chemical plant set up, the primary objective is to maximize profits.

The reactor can also be optimized based on the fact that in order to minimize cost of construction of the reactor, the length and diameter of the reactor must be kept at minimum.

For a tubular reactor, the total surface area of the reactor is:

$$A_{chamber} = \left(2 \cdot \pi \cdot r^2\right) + 2\pi \cdot r \cdot h_{chamber}$$

$$r = \frac{d_{chamber}}{2}$$

Where,

r = "radius of the chamber tube"

$d_{chamber}$ = "diameter of the chamber"

$h_{chamber}$ = "height of the chamber"

The formula now becomes

$$A_{chamber} = \left(2 \cdot \pi \cdot r^2\right) + 2\pi \cdot r \cdot h_{chamber}$$

$$A_{chamber} = \left[2 \cdot \pi \cdot \left(\frac{d_{chamber}}{2}\right)^2\right] + 2 \cdot \pi \cdot \frac{d_{chamber}}{2} \cdot h_{chamber}$$

$$A_{chamber} = \left(2 \cdot \pi \cdot \frac{d_{chamber}^{\;2}}{4}\right) + 2 \cdot \pi \cdot \frac{d_{chamber}}{2} \cdot h_{chamber}$$

$$A_{chamber} = \pi \cdot \frac{d_{chamber}^{\;2}}{2} + \pi \cdot d_{chamber} \cdot h_{chamber}$$

That is to say that the area of the reactor is a function of the diameter and length of the reactor.

$$A_{chamber} = f\left(d_{chamber}, h_{chamber}\right)$$

Where the objective function is:

$$A = \pi \cdot \frac{d_{chamber}^{2}}{2} + \pi \cdot d_{chamber} \cdot h_{chamber}$$

$d = d_{minimum}$ and $h = h_{minimum}$ are the constraints.

The equation of $A_{chamber}$ becomes

$$f(d_{chamber}, h_{chamber}) = \pi \cdot \frac{d_{chamber}^{2}}{2} + \pi \cdot d_{chamber} \cdot h_{chamber}$$

Note that the volume of the reactor is given as

$$V_{chamber} = \pi \cdot r^{2} \cdot h_{chamber}$$

$$r = \frac{d_{chamber}}{2}$$

$$V_{chamber} = \pi \cdot \left(\frac{d_{chamber}}{2}\right)^{2} \cdot h_{chamber}$$

$$V_{chamber} = \pi \cdot \frac{d_{chamber}^{2}}{4} \cdot h_{chamber}$$

$$h_{chamber} = \frac{4 \cdot V_{chamber}}{\pi \cdot d_{chamber}^{2}}$$

Substituting the expression of h into the equation of total area of the reactor which is a function of diameter, d, and height, h, it is obtained that

$$f(d_{chamber}, h_{chamber}) = \pi \cdot \frac{d_{chamber}^{2}}{2} + \pi \cdot d_{chamber} \cdot \frac{4 \cdot V_{chamber}}{\pi \cdot d_{chamber}^{2}}$$

$$f(d_{chamber}, h_{chamber}) = \pi \cdot \frac{d_{chamber}^{2}}{2} + \pi \cdot \frac{4 \cdot V_{chamber}}{\pi \cdot d_{chamber}}$$

Now, it can be observed that the term of h has disappeared. That is to say that the total surface area is now a function of only the diameter, d. The expression can be

$$f(d_{chamber}) = \pi \cdot \frac{d_{chamber}^{2}}{2} + \pi \cdot \frac{4 \cdot V_{chamber}}{\pi \cdot d_{chamber}}$$

In optimizing, it may be maximizing or minimizing. In this case the aim is to maximize buy to minimize the dimension of the reactor so that the construct can be less. So that the profit can be much.

Differentiating the above equation,

$$\frac{d}{dd_{chamber}} f(d_{chamber}) = \pi \cdot d_{chamber} - \frac{4}{d_{chamber}^{2}} \cdot V_{chamber}$$

To optimize, the differential will be equal to zero.

$$\frac{d}{dd_{chamber}} f(d_{chamber}) = \pi \cdot d_{chamber} - \frac{4}{d_{chamber}^{2}} \cdot V_{chamber} = 0$$

Taking the last two expressions,

$$\pi \cdot d_{chamber} - \frac{4}{d_{chamber}^{2}} \cdot V_{chamber} = 0$$

Making d the subject of formula

$$\pi \cdot d_{chamber} = \frac{4}{d_{chamber}^{2}} \cdot V_{chamber}$$

$$d_{chamber}^{3} = \frac{4 \cdot V_{chamber}}{\pi}$$

$$d_{chamber} = \sqrt[3]{\frac{4 \cdot V_{chamber}}{\pi}}$$

$d_{chamber}$ = "optimized diameter"

Note that:

$$h_{chamber} = \frac{4 \cdot V_{chamber}}{\pi \cdot d_{chamber}^{2}}$$

Substituting for d in this expression give

$$h_{chamber} = \frac{4 \cdot V_{chamber}}{\pi \cdot \left(\sqrt[3]{\frac{4 \cdot V_{chamber}}{\pi}}\right)^{2}}$$

The gasifier is considered as a cylindrical vessel and is optimized in two sections: the lower and upper section.

Lower chamber:

Numerically with Vlower chamber = 28.644m^3

$$d_{\text{lower chamber}} = \sqrt[3]{\frac{4V_{\text{lowerchamber}}}{\pi}} \qquad d_{\text{lower chamber}} = 3.316255077\text{m}$$

$$h_{\text{lower chamber}} = \frac{4 \cdot V_{\text{lower chamber}}}{\pi\left(\sqrt[3]{\frac{4V_{\text{lowerchamber}}}{\pi}}\right)^2} \qquad h_{\text{lower chamber}} = 3.316255077$$

This means that the optimum diameter and height of the lower chamber are the same.

Upper chamber:

Vupper chamber = 1.471 × 10^{-3} m^3

$$d_{\text{upper chamber}} = \sqrt[3]{\frac{4.V_{\text{upperchamber}}}{\pi}} \qquad d_{\text{upper chamber}} = 0.123265326$$

$$h = \frac{4 \cdot V_{\text{upper chamber}}}{\pi\left(\sqrt[3]{\frac{4V_{\text{upperchamber}}}{\pi}}\right)^2} \qquad h_{\text{upper chamber}} = 0.123265326$$

This means that the optimum diameter and height of the upper chamber are the same.

Total diameter of gasifier = $d_{\text{lower chamber}} + d_{\text{upper chamber}}$

= (3.316255077+ 0.123265326) m

= 3.439520404m

Total height of gasifier = $h_{\text{lower chamber}} + h_{\text{upper chamber}}$

= (3.316255077 + 0.123265326) m

= 3.439520404m

This means that the optimum diameter and height of the gasifier are the same.

7.2 CATALYTIC REFORMER

The catalytic reformer can be optimized using that fact that,inorder to minimize cost of construction of the catalytic reformer ,the legth and the diameter of the reactor must be kept at minimum.

Since the reactor is a tubular type , it is said to have a cylindrical shape.

That is, the total surface area of the catalytic reformer is given as

$Ar = 2\pi r^2 + 2\pi r$

And $r = \frac{D}{2}$

Where, $r = radius\ of\ the\ reformer$

$D = diameter\ of\ the\ reformer$

$H = heigth\ of\ the\ reformer$

$\pi = pie, a\ constant$

$So, the\ formula\ becomes$

$Ar = 2\pi r^2 + 2\pi rH$

$Ar = 2\pi \left(\frac{D}{2}\right)^2 + 2\pi \frac{D}{2} H$

$Ar = 2\pi \frac{D^2}{4} + 2\pi \frac{D}{2} H$

$Ar = \pi \frac{D^2}{2} + \pi\pi DH$

$Ar = f(D, H)$

Where

$Ar = \pi \frac{D^2}{2} + \pi DH\ is\ the\ objective\ function\ and\ D = minimum\ and\ H\ minimum\ are\ constants\ that\ are\ to\ be\ minimi$ zed.

So that the equation of Ar becomes

$f(D, H) = \pi \frac{D^2}{2} + \pi DH$

Noting that the volume of the tank is given as $V_r = \pi r^2 H$

With $r = \frac{D}{2}$

$V_r = \pi \left(\frac{D}{2}\right)^2 H$

$V_r = \pi \frac{D^2}{4} H$

Making H the subject of the formula in the above equation.

$$H = \frac{4V_r}{\pi . D^2}$$

Substitut

$$f(D,H) = \pi\frac{D^2}{2} + \pi D\frac{4vr}{\pi D^2}$$

$$f(D) = \pi\frac{D^2}{2} + \pi\frac{4vr}{\pi D}$$

Differentiating the equation

$$\frac{\delta}{\delta D}f(D) = \pi D\frac{4}{D^2}Vr$$

$$\frac{\delta}{\delta D}f(D) = \pi D\frac{4}{D^2}Vr = 0$$

Taking the last two expressions

$$\pi D = \frac{4}{D^2}Vr = 0$$

$$H = \frac{4Vr}{\pi D^2}$$

Substituting for D in this equation yields

$$H = \frac{4Vr}{\pi\left[\sqrt[3]{\frac{4Vr}{\pi}}\right]^5}$$

Simplifying

$$H = \frac{V_r^{1/3}}{\pi^{2/3}}\sqrt[3]{2^2}$$

Numerically

$$V_r = 0.0236\, m^3$$

$$D = \sqrt[3]{\frac{4V_r}{\pi}}$$

$$D = \sqrt[3]{0.03}$$

$$D = 0.311\, m$$

$$H = \frac{V_r^{1/3}}{\pi^{2/3}}\ \sqrt[3]{2^2}$$

$$H = \frac{0.287}{1.465} \times 1.587 = 0.311\, m$$

CHAPTER EIGHT

8.0 SAFETY MEASURES AND QUALITY CONTROL

In designing a plant, safety is one of the major criteria for selection of the best alternative along with economic viability. Safety becomes even more pertinent when the materials involved are hazardous because value should be placed on the operating personnel and the equipment handled. Safety measures are then recommended to fit into the plant design. The following recommendations are important in handling it.

8.1 General Recommendations

Written procedures detailing requirements for proper handling, safety equipment, first aid training, unloading and loading procedures are good practice and in many cases, required by law. They should include detailed instructions for handling and reporting spills. Considerable care should be taken when transferring the final product to maintain high product quality. These same careful practices will ensure the health and safety of workers. An above ground tank must be used whenever a large volume of syngas is to be stored at a customer or terminal site. Storage tanks should be large enough to contain a minimum of 150% of the normal delivery volume. The tank should be monitored closely during the initial filling with calcium hypochlorite to check for leaks not detected during water testing.

Storage tanks may be mounted horizontally or vertically. All local regulations concerning above ground storage tanks should be reviewed and all permits obtained before installing a bulk storage system. Bulk storage containers should be constructed of either mild, or stainless steel. Storage tanks should not be constructed of, nor contain, any non compatible plastic components. The storage tanks exterior should be cleaned, primed and painted with a white or aluminum colored paint to aid in keeping the tank and its contents cool.

A storage tank pads or saddles

Saddles used to support horizontal tanks may be constructed of reinforced concrete or steel. The design of the concrete pad or saddle foundation (if horizontal) should be based on at least the total weight of the tank filled with product.

B Dikes

All storage tanks should be diked to contain the tank contents in the event of a spill or tank rupture. They should be large enough to contain the tank's volume, and an additional appropriate volume as safety factor. (Containment volumes and diking requirements are often defined and mandated by individual states and localities. Regulations must be reviewed prior to construction.)

C Piping

Carbon or stainless steel is the material of choice for piping. Transfer pipes are typically 2-inch in diameter, but may be as large as 6-8 inches for barge and vessel deliveries. Fittings may be carbon steel or stainless steel, and may be threaded, flanged or welded. Threaded piping is only recommended for sizes less than 3/4" in diameter. It is essential that the pipe be threaded two full turns before applying threading compound or Teflon® tape to eliminate the possibility of contaminating the interior of the pipe. All pipes should be free of oils and any other contaminants prior to being placed in service.

D. Drain Lines

Storage tanks should be equipped with a flanged and valve outlet drain at the floor level to allow the tank complete drainage if necessary. This outlet should be located such that it can be tied into a pump.

E. Valves

Ball, gate, globe or plug valves are suitable for use with synthesis gas. Construction may be steel. Ball valves should have Monel balls and stems, and seats should be carbon filled PTFE, and globe valves should have 13% chrometrim and hard-faced Satellite seats. Valve stems may be packed with PTFE or graphite impregnated PTFE. Ball, gate or plug valves are generally preferred where full line flow is desired. Globe valves are used where throttling is necessary.

8.2 General Quality Control Measures

Quality simple means "fitness for use". But, according to the international standard organization (ISO), quality is defined as the totality of the characteristics of an entity that bears on its ability to satisfy stated and intended needs. It is more costly to exceed a specification than to meet it. Therefore, there is the need to get quality goal or target for effective quality control.

Maintaining product quality in accordance with acceptable standard has been a major role for industrial instrumentation since its inception decades ago with the ever growing interest in speeding up production, one becomes increasingly aware of the fact that rejects as well as acceptable products can be produced at very high rates. Below are some of the measures:

8.2.1 General safety measures

1. All materials of construction should be well selected on the basis of corrosion resistance and structural strength. This is to avoid the collapse of any equipment or structures.
2. All pipelines carrying flammable materials must be installed with flame traps.
3. Relief valves should be installed along all lines carrying gases and vapour.

4. Automatic controllers should be installed to control temperature, pressure, and flow rates of the material or equipment involved.
5. A badly maintained plant is a potential hazard. Ensure that a competent staff is responsible for recognizing maintenance and repairs on a planned basis.
6. Signs and placards warning of the hazardous materials should be placed all over the plant.
7. Foam fire extinguishers or those using carbon dioxide should be widely and easily available and ready for immediate use at all times.
8. Plant layout should be such that:
    - i. The storage facilities should be placed away from the plant.
    - ii. All electrical installation should be earthed and insulated and should be kept away from the processing unit.
    - iii. Exit and escape route: - the concept of quality control analysis has been greatly achieved by statistical quality control method. The general intent of SQC is that of sampling units and parts being produced and essentially determining trends in deviation from production as continuously (affordable and achievable) as possible. There are many important statistical tests, which can be used to determine the quality of products.

General Safety Rules

Relevant Instructions

a) Before attempting to operate the Gasifier plant, all relevant manufacturers' instructions and local regulations should be understood and implemented.

b) It is irresponsible and dangerous to misuse equipment or ignore instructions, regulations or warnings.

c) The specified maximum operating conditions must not be exceeded.

(Odigure, 1998)

d) It must be ensured that all staff must be fully aware of the potential hazards when the Gasifier plant is being operated.

e) Serious injury can result from touching apparently stationary equipment or rotating belt.

Operational Instructions

a) All materials of construction should be well selected on the basis of corrosion resistance and structural strength. This is to avoid the collapse of any equipment or structures.

b) All pipelines carrying flammable materials must be installed with flame traps.

c) Relief valves should be installed along all lines carrying gases and vapour.

d) Automatic controllers should be installed to control temperature, pressure, and flow rates of the material or equipment involved.

e) A badly maintained plant is a potential hazard. Ensure that a competent staff is responsible for recognizing maintenance and repairs on a planned basis.

f) Signs and placards warning of the hazardous materials should be placed all over the plant.

g) Foam fire extinguishers or those using carbon dioxide should be widely and easily available and ready for immediate use at all times.

Maintenance

a) A badly maintained Gasifier plant is a potential hazard. It must be ensured that competent members of staff is responsible for organizing maintenance and repairs on a planned basis.

b) Faulty equipment must not be permitted to be operated. Repairs must be carried out competently and the operation must be checked. (Odigure, 1998)

Using electricity

a) At least once a month, the electrical cables should be checked to ensure that they are operating normally.

b) Electricity is the commonest cause of accidents in the factory, it must be respected.

c) It must be ensured that electricity supply has been disconnected from the equipment before attempting repairs or adjustment.

d) It must be known that water and electricity are not compatible and can cause serious injury if they come into contact.

e) The plant must always be disconnected from electricity when not in use. (Odigure, 1998)

Avoiding fire or explosion

a) It must be ensured that the factory is provided with adequate fire extinguishers appropriate to the potential dangers.

b) It must be known that empty vessels having inflammable liquids can contain vapours and explode if ignited.

Handling poisons or toxic materials

a) Food must not be allowed to be brought into or consumed in the factory.

b) Smoking should not be allowed in the factory premises. Notices should be so displayed and enforced. (Odigure, 1998)

Avoiding cuts and burns

a) Care must be taken when handling sharp edged components. Undue force must not be exerted on glass or fragile items.

b) Hot surfaces cannot, in most cases, be totally shielded and can produce severe burns even when not "visibly hot". Common sense must be used always!

c) **Eye protection**

a) Facilities for eye irritation should always be available

Ear protection

a) Ear protectors must be worn when operating the plant e.g. while operating the dryer.

Guard and safety devices

a) Guards and safety devices must be installed on the plant to protect the operators. The equipment must not be operated with such devices removed.

b) Safety gauges, cut-out and other safety devices must be set to protect the equipment. Interference with these devices may create a potential hazard.

c) It is impossible to guard the operator against all contingencies. Common sense musc be used.

d) Before starting a machine, it must be ensured that the members of staff are aware of how it (the machine) should be stopped in an emergency. (Odigure, 1998)

First aid

a) It is essential that first aid equipment is available and that the supervisor knows how to use it.

b) A notice giving details of a proficient first aider should be prominently displayed. (Odigure, 1998)

8.2.2 Quality control measures

Maintaining product quality in accordance with acceptable standard has been a major role for industrial instrumentation since its inception decades ago with the ever growing interest in speeding up production, one becomes increasingly aware of the fact that rejects as well as acceptable products can be produced at very high rates.

8.3 Quality Assurance

Quality assurance is defined by ISO as all the panned and synthetic activities implemented within the quality system and demonstration is needed to provide adequate confidence on entity will fulfil requirement for quality. (Odigure; 1998)

8.3.1 Principles of Quality Assurance

The principle of quality assurance include the following: (Odigure, 1998)

i. Management involvements and objectives (management) involvement is very essential to ensure quality.

ii. Programming and planning

iii. Application of quality control principles.

iv. Design and specification control.

v. Purchasing control and vendor appraisal.

vi. Production control.

vii. Marketing and service quality functions.

viii. Proper documentation

ix. Non-conformance control.

x. Remedial action.

xi. Defect and failure analysis.

8.3.2 Quality Management

Quality management involves all activities of the overall management functions that determine the quality policy, objective and responsibilities and implement then by means, such as quality planning control assurance and improvement within the quality system. Responsibility of quality lies at all level of all. To successfully implement quality management, the organization structure, procedure process and resources are requisite. (Odigure, 1998)

CHAPHER NINE

9.0 PROCESS CONTROL AND INSTRUMENTATION

9.1 INTRODUCTION

In the design of an industrial plant, the methods, which will be used, for plänt operation and control help to determine many of design variables, for example, the extent of instrumentation can be a factor in choosing the type of process and setting the labor requirements. It should be remembered that maintenance work would be necessary to keep the installed equipment and facilities in good operating condition. The importance of such factors which are directly related to plant operation and control must be recognized and taken into proper account during the development of a design project. Processes may be controlled more precisely to give more uniform and higher – quality products by the application of automatic control, often leading to higher profits. Therefore, any process with an input and output which may be flow, pressure, liquid level, temperature, composition or any other inventory, environmental or quality variable that is to be held at a desired value must have some measure of control applied to it.

Change in output may occur:

1. Randomly as caused by changes in weather or raw material quality.
2. Diurnally with ambient temperature
3. Manually when operators change production rate.
4. Stepwise when equipment is switched in or out of service, or
5. Cyclically as a result of oscillations in other control loops.

Variation in any of the ways stated above would drive the output (controlled variable) further away from the set point (desired value) thus requiring a corresponding variable to bring it back (manipulative variable).

9.2 Instrumentation and Control Objectives

The key objectives adhered to in the specification of the instrumentation and control schemes are:

1. Safe plant operation
    - i. To keep the process variables within known safe operations limits.
    - ii. To defect dangerous situations as they develop and to provide alarms and automatic shut – down systems
    - iii. To provide interlocks and alarms to prevent dangerous operation procedures
2. Production rate. To achieve the design producer output
3. Product quality. To maintain the product composition within the specific quality standards
4. Cost. They operate at the lowest production cost, but not to the detriment of the product quality.

In the plant design some of the variables needed to be monitored and controlled are, the flow rate, temperature, pressure and composition.

- i. **Flow rates/feed ratio control**: a feed ratio controller is applied between the feedstock stream, the gasification agent feed stream and the steam and air stream with the aim of maintaining a complete conversion of biomass feed to synthesis gas.
- ii. **Temperature control reaction**: temperature is controlled by regulating the flow rate of water through the reactor jack (in the case of exothermic reactions) or the power of the heating coils (in the case of the endothermic reactions). This can also be controlled by heat transfer. Also, if the

reaction is exothermic. It produces heat, which tends to raise reaction temperature thereby increasing reaction rate and producing more heat. This positive feedback is countered by negative feedback in the cooling system, which removes more heat as reactor temperature rises. The temperature controller, which in turn operates the coolant value, to counter the rise or drop in temperature in the sector.

iii. **Pressure control**: pressure sensing is quite straightforward with the aid of pneumatic instrumentation such as Bourdon gauge, diaphragms and bellows. These sensors measure absolute pressure and pressure differences between two levels. Therefore, pressure control is achieved by manipulating the airflow rate in the equipments to avoid deviation from set point.

iv. **Composition control**: first requirement here is to establish proper stoichiometry of the reactants in propositions needed to satisfy the reaction chemistry and also the desired output product. This is achieved by setting input flow rates in ratio to one another, or a composition measurement (analyzer) can be used to trim the ratios to the right proportion

9.3 Control Mechanism

The pneumatic control hardware is recommended for this process it will be powered by instrument air supplies.

The control mechanism for this process consist of a sensor to detect the process variables; a transmitter to convert the sensor into an equivalent "signal" a controller that compares this process signal with a desired set point value and produces an appropriate controller output signal and a final control element (pneumatic activator) that changes the manipulated variable with the use of a mechanical action.

Types of control instruments

The control instruments are of four major categories

a) Temperature controllers (TC)

b) Pressure controller (PC)

c) Flow controller (FC)

d) Level controller (LC)

9.3.1 Control Sensors

The devices to be used for the on-line measurement of the process variables are:

1) Flow sensor: The orifice meter can be employed in the process since it is simple and of low cost.
2) Temperature sensor: The recommended temperature sensors are resistance thermometer detectors (RTDS) and Thermocouples.
3) Pressure sensors: Bourdon – Tube pressure gauge can be used.
4) Level sensor: float activated devices are sufficient.

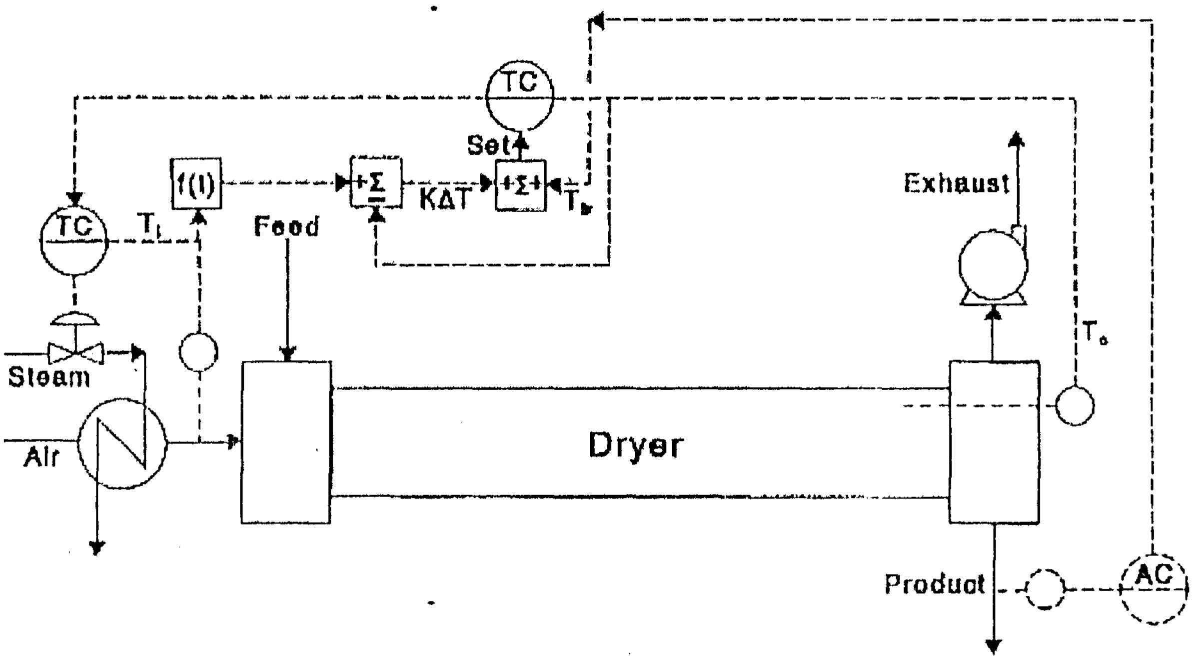
Alarms are to be employed to alert the process operator to a process that requires immediate action and attention. Instead of individually issuing point alarms, all alarms associated with a certain aspect of the process are to be simply wired to give a single trouble alarm.

9.3.2 Transmitters, Controllers and Control Valves

The transmitter is the interface between the process and its control system. The transmitter converts the sensors signal into a control signal. The pilot – acting controllers should be employed in the process. The pilot – acting controllers are capable of greater degree of sensitivity since they eliminate of the lags which would be inherent in self –

acting mechanism activated by the force of a large volume of fluid. The fluid control element is an automatic control which throttles the flow of the manipulated variable.

Figure 9.1: Control diagram for the rotary dryer (Unit 8).



CHAPTER TEN

10.0 ENVIRONMENTAL ACCEPTABILITY

10.1 Identification of Possible Pollutants

A Gasifier production plant requires a rational approach of assessing the methodology in order to identify systematically all possible pollutants. Suitability or acceptability of any process plant depends on its impact to the life, properties and environment.

Emissions arise from several sources within the Gasifier production plants. There are potential gaseous emissions from the transfer and storage of calcium hypochlorite in the factory, as well as potential leaks of solvent from pipes.

Vents are another source of emissions. Most gaseous emissions from the plant are basically leakage from pipes which contains slight carbon dioxide. This is due to incomplete combustion and reduction of activated carbon in the gasifier. Such emissions can the dangerous to the health of health workers.

Slag and tar from the waste tank can also constitute a major environmental hazard if not properly monitored.

During storage of synthesis gas, there might be breakdown which will result to the release of caobon monoxide and hydrogen gas. This is usually experienced if the storage vessel is placed in a location where it is exposed to excess sunlight.

10.1.1 Effect of Carbon Dioxide Emission on Human Health

10.1.1.1 Effects on Human Health

Carbon dioxide is a gas. Human health effect specific to the CO_2 gas may be caused by inhaling the gas and also the depletion of ozone layer which exposes the human race to excess heat thereby causing sun burns, unproductive vegetative environment etc. and they are thoroughly assessed in competent/pertinent international organization like WHO or in the EU risk assessment program.

10.1.1.2 Toxicokinetics, Metabolism and Distribution

The primary available chlorine species in aqueous solution are not different from those coming from hypochlorous acid at similar pH. In biological systems, characterised by pH values in the range of 6-8, the most abundant active chemical species is HClO, in equilibrium with ClO^-. The latter is predominant at alkaline pH values, while Cl_2 is mainly present at pH values below 4. Therefore, the studies performed with hypochlorite and its salts are used in this document. Limited data are available for the oral route only.

Inhalation

Exposure to sub-lethal doses of chlorine gas (from calcium hypochlorite) leads to obstructive disturbances of pulmonary ventilation which are usually reversible. Weill et al. (Weill et al., 1969) and Jones et al. (Jones et al.,1986) did not find abnormalities up to 6 years after accidental chlorine exposures that could not be attributed to other underlying lung diseases or smoking. Also Leroyer et al. [Leroyer et al., 1998] in their 4-year follow-up of 13 workers with accidental chlorine exposure showed complete recovery in three months for the individual who had decreased forced expiratory volume (FEV1) and two individuals with decreased PC20 (Leroyer et al., 1998). Hasan et al. (Hasan et al., 1983) found improvement in respiratory symptoms forced vital capacity (FVC) and FEV1 within 5 months. In this study, however, the bronchial hyper responsiveness was not assessed (Hasan et al., 1983). A syndrome, defined by Brooks et al. (Brooks et al., 1985) as "Reactive Airways Dysfunction Symptom". (RADS) has also been related to acute chlorine exposure, which is a sudden onset type of asthmatic illness following acute inhalation of high-dose irritant gases. In rare cases, persons may acquire a chronic bronchial hyper reactivity after acute or repeated contact with irritant gases like chlorine or sulfur dioxide. This depends on the individual disposition rather than on an allergic mechanisms. Chronic occupational exposure up to 0.01–1.4 ppm (0.03–4.2 mg/m3) of chlorine gas is reported to have produced no lesions in the exposed workers (Patil et al., 1970).

10.2 Suggestions on Treatments of Pollutants.

To make this process suitable and acceptable to the environment, there is need for these pollutants to be controlled. The design procedure employed gives optimum reduction in the number of pollutants, this is achieved by recycling most and all unreacted streams back to points of reaction. Some of the treatment procedures that can be employed include:

Ensuring that most, if not all of all the chlorine gas fed in to the chlorinators undergo reaction. For the very few that might escape from the vent, it could be passed through a of water that is standing at a relatively low temperature. This will produce a weak acid solution which can easily be disposed off.

Sludge from the impurity separator can only be harmlessly used as landfills if it is completely free of calcium chloride values. Thus it should be ensure that the impurity separator is always working at optimum.

As one of the possible sources of chlorine emission is from high temperature values in the dryer, the dryer temperature controller should be designed as a cascade controller. With a predetermined set point of the feed forward controller set at the operating temperature of the dryer. This is to correct any fluctuation in temperature before they even occur.

To ensure that there are no calcium chloride by-products serving as contaminants, an excess of sodium hypochlorite should be introduced in the mixing zone reactor, as this will ensure that all calcium chloride values are consumed.

CHAPTER ELEVEN

11.0 START-UP AND SHUT DOWN

11.1 Start-Up

1. Only responsible, well-trained and well-supervised employees should be entrusted with the starting-up process. A worker should be present during the entire time of plant operation. A written copy of the procedures should be readily available.

2. Fill the feedstock chamber with the biomass feedstock in the appropriate percentage and also the water tank with water so that water can flow into the boiler and steam can be produced. Once steam is produced, it should be channeled into the preheater along side with the process air so it can be preheated before normal operation.

3. Check all transfer lines and hoses to make sure they are dry and free of contamination. Inspect equipment for signs of deterioration or other conditions which might cause a leak. Do not continue if there is any question about the condition of any equipment.

4. Carefully vent the delivery vessel to zero gauge pressure and cautiously disconnect the air and discharge lines. Be prepared with a suitable container to collect any product remaining in the line. Have appropriate supplies to contain or clean up any possible leakage that may occur.

5. Close all valves and replace all caps and blind flanges on the delivery vessel. Make certain that the dome cover is fastened securely and electrical grounding

wires are removed. Flush hoses with water and wash down area where any vapors or leakage may have occurred.

6. Start cooling water to reactor jackets.
7. Adjust controls to set point.
8. Turn all valves to automatic position
9. Turn power on.

11.2 Shut-Down

1. Turn all valves to manual position
2. Close valve supply into all equipments.
3. Stop cooling water to all reactors.
4. Discharge reactors content.
5. Flush all line with air to inert the system

CHAPTER TWELVE

12.0 SITE FOR PLANT LOCATION

12.1 Site for Plant Location and Site Selection

The geographical location of the final plant can have strong influence on the success of the industrial venture. Considerable care must be exercised in selecting the plant site, and many different factors must be considered. Primarily the plant must be located where the minimum cost of production and distribution can be obtained but, other factors such as room for expansion and safe living conditions for plant operation as well as the surrounding community are also important. The location of the plant can also have a crucial effect on the profitability of a project. The choice of the final site should first be based on a complete survey of the advantages and disadvantages of various geographical areas and ultimately, on the advantages and disadvantages of the available real estate.
It is for these reasons that the plant would be located in Niger State, Nigeria.

12.2 Factors Considered for Plant Location

The various principal factors that must be considered while selecting a suitable plant site, are briefly discussed below. The factors to be considered are:

1. Raw material availability.
2. Location (with respect to the marketing area.)
3. Availability of suitable land.
4. Transport facilities.
5. Availability of labors.
6. Availability of utilities (Water, Electricity).

7. Environmental impact and effluent disposal.

8. Local community considerations.

9. Climate.

10. Political strategic considerations.

11. Taxations and legal restrictions

12.3 Justification of Selected Site

12.3.1 Raw materials availability

The source of raw materials is one of the most important factors influencing the selection of a plant site. This is particularly true for citing the Gasifier plant as large volumes of non woody biomass which is a primary raw material in the production of synthesis gas is found in the north. For these reason, the availability of raw materials would be high.

12.3.2 Market location

The location of markets or intermediate distribution centers affects the cost of product distribution and time required for shipping. Proximity to the major markets is an important consideration in the selection of the plant site, because the buyer usually find it advantageous to purchase from near-by sources.

12.3.3 Availability of suitable land

The characteristics of the land at the proposed plant site should be examined carefully. The topography of the tract of land structure must be considered, since either or both may have a pronounced effect on the construction costs. The cost of the land is important, as well as local building costs and living conditions. Future changes may make it desirable or necessary to expand the plant facilities. The land should be ideally flat, well drained and have load-bearing characteristics. Niger state has vast untapped land resources. The land has good drainage system which implies that the land is well drained and has load-bearing capabilities.

12.3.4 Transport

The transport of materials and products to and from plant will be an overriding consideration in site selection. If practicable, a site should be selected so that it is close to at least two major forms of transport: road, rail, waterway or a seaport. Road transport is being increasingly used, and is suitable for local distribution from a central warehouse. Rail transport (if efficient in Nigeria) will be cheaper for the long-distance transport. If possible the plant site should have access to all three types of transportation. There is usually need for convenient rail and air transportation facilities between the plant and the main company head quarters, and the effective transportation facilities for the plant personnel are necessary.

Niger state is at the northern part of Nigeria a little close to the federal capital territory. With this efficient transportation mechanism, (road), raw materials and finished products can reach the plants and the consumers.

12.3.5 Availability of labour

Labors will be needed for construction of the plant and its operation. Skilled construction workers will usually be brought in from outside the site, but there should be an adequate pool of unskilled labors available locally; and labors suitable for training to operate the plant. Skilled tradesmen will be needed for plant maintenance. Local trade union customs and restrictive practices will have to be considered when assessing the availability and suitability of the labors for recruitment and training.

Niger state has an abundant untapped labour force. Youths in their thousands wonder through the town jobless. This ensures that the labour forces for the plant are not only available but also cheap. This increases the profit margin for the plant.

12.3.6 Availability of Utilities

The word "utilities" is generally used for the ancillary services needed in the operation of any production process. These services will normally be supplied from a central facility and includes majorly, water and electricity, which is briefly described as follows:

(i) **Water:** - The water is required for large industrial as well as general purposes, starting with water for cooling, washing, steam generation and as a raw material in the water tank. The plant therefore must be located where a dependable water supply is available namely lakes, rivers, wells, seas. If the water supply shows seasonal fluctuations, it's desirable to construct a reservoir or to drill several standby wells. The temperature, mineral content, slit and sand content, bacteriological content, and cost for supply and purification treatment must also be considered when choosing a water supply. Demineralized water, from which all the minerals have been removed is used where pure water is needed for the process use, in boiler feed. Natural and forced draft cooling towers are generally used to provide the cooling water required on site.

As Niger state has abundant ground and surface water supply, the availability of water is on a great scale.

(ii) **Electricity:** Niger state is one of the most abundantly power supplied state. With a consistent supply of electricity from PHCN, the power consumption of the plant will be profitable as there would be reduced costing which would have otherwise been necessary if independent generating stations were employed for the plant.

12.3.7 Environmental impact and effluent disposal

Facilities must be provided for the effective disposal of the effluent without any public nuisance. In choosing a plant site, the permissible tolerance levels for various effluents should be considered and attention should be given to potential requirements for additional waste treatment facilities. As all industrial processes produce waste products, full consideration must be given to the difficulties and cost of their disposal. The disposal of toxic and harmful effluents will be covered by local regulations, and the appropriate authorities must be consulted during the initial site survey to determine the standards that must be met.

12.3.8 Local Community Considerations

The proposed plant must fit in with and be acceptable to the local community. Full consideration must be given to the safe location of the plant so that it does not impose a significant additional risk to the community.

A community that is not peaceful will therefore not be ideal for the plant. This is because any form of community unrest will disrupt normal plant operations. It is for this reason that Niger serves as a perfect location for the plant, as its people are peaceful and friendly to outsiders. Thus there will be

not issue of community rejection.

12.3.9 Climate

Adverse climatic conditions at site will increase costs. Extremes of low temperatures will require the provision of additional insulation and special heating for equipment and piping. Similarly, excessive humidity and hot temperatures pose serious problems and must be considered for selecting a site for the plant. Stronger structures will be needed at locations subject to high wind loads or earthquakes.

12.3.10 Political and strategic considerations

Capital grants, tax concessions, and other inducements are often given by governments to direct new investment to preferred locations; such as areas of high unemployment. The availability of such grants can be the overriding consideration in site selection.

Tax concessions, are not high in Niger state, as it is not really an industrially developed state. The government of Niger state due the high unemployment of its populace is in constant search for outside investors. This makes tax ratings for industries in Niger state low, thus it serves as a suitable site for locating the plant.

12.3.11 Taxation and legal restrictions

State and local tax rates on property income, unemployment insurance and similar items vary from one location to another. Similarly, local regulations on zoning, building codes, nuisance aspects and others facilities can have a major influence on the final choice of the plant site.

12.4 Plant Layout

After the flow process diagrams are completed and before detailed piping, structural and electrical design can begin, the layout of process units in a plant and the equipment within these process unit must be planned. This layout can play an important part in determining construction and manufacturing costs, and thus must be planned carefully with attention being given to future problems that may arise. Thus the economic construction and efficient operation of a process unit will depend on how well the plant and equipment specified on the process flow sheet is laid out. The principal factors that are considered are listed below:

1. Economic considerations: construction and operating costs.
2. Process requirements.
3. Convenience of operation.
4. Convenience of maintenance.
5. Health and Safety considerations.
6. Future plant expansion.
7. Modular construction.
8. Waste disposal requirements

12.4.1 Economic considerations: operating costs

The cost of construction can be minimized by adopting a layout that gives the shortest run of connecting pipe between equipment, and least amount of structural steel work. However, this will not necessarily be the best arrangement for operation and maintenance.

12.4.2 Process requirements

An example of the need to take into account process consideration is the need to elevate the base of columns to provide the necessary net positive suction head to a pump.

12.4.3 Convenience of operation

Equipment that needs to have frequent attention should be located convenient to the control room. Valves, sample points, and instruments should be located at convenient positions and heights. Sufficient working space and headroom must be provided to allow easy access to equipment.

12.4.4 Convenience of maintenance

Dryers need to be sited so that the tube bundles can be easily withdrawn for cleaning and tube replacement. Vessels that require frequent replacement of catalyst or packing should be located on the out side of buildings. Equipment that requires dismantling for maintenance, such as compressors and large pumps, should be places under cover.

12.4.5 Health and safety considerations

Blast walls may be needed to isolate potentially hazardous equipment, and confine the effects of an explosion. At least two escape routes for operators must be provided from each level in process buildings.

12.4.6 Future plant expansion

Equipment should be located so that it can be conveniently tied in with any future expansion of the process. Space should be left on pipe alleys for future needs, and service pipes over-sized to allow for future requirements.

12.4.7 Modular construction

In recent years there has been a move to assemble sections of plant at the plant manufacturer's site. These modules will include the equipment, structural steel, piping and instrumentation. The modules are then transported to the plant site, by road or sea.

The advantages of modular construction are:

1. Improved quality control.

2. Reduced construction cost.

3. Less need for skilled labors on site.

The disadvantages of modular construction are:

1. Higher design costs & more structural steel work.

2. More flanged constructions and possible problems with assembly on site.

12.4.8 Waste disposal requirements

In a plant layout, the permissible tolerance levels for various wastes should be considered and attention should be given to potential requirements for additional waste treatment facilities. As all industrial processes produce waste products, full consideration must be given to the difficulties and cost of their disposal. The disposal of toxic and harmful effluents will be covered by local regulations, and the appropriate authorities must be consulted during the initial site survey to determine the standards that must be met.

A detailed plant layout is drawn as shown in figure 9.1 below. Although this plant layout is just a reference plant layout. There may be changes in actual plant layout by the industrialist or whoever is setting up the plant. Below are the key words for the plant layout.

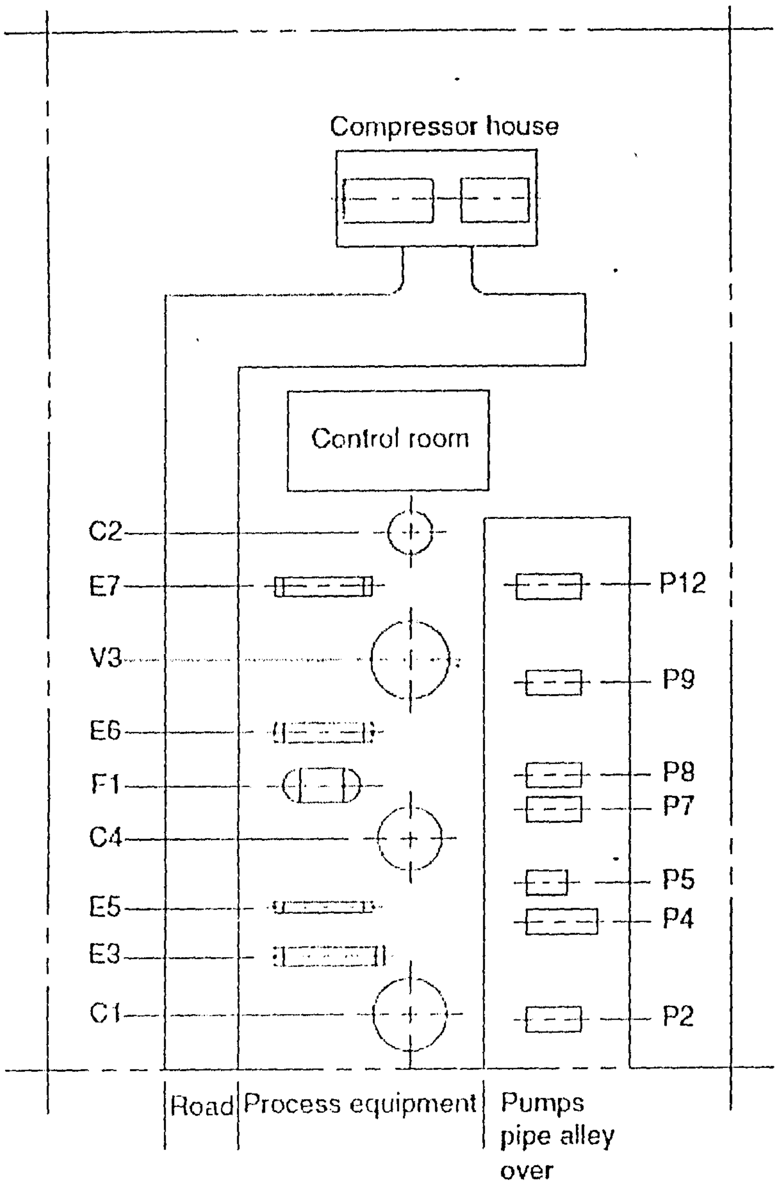


Fig 12.1: A typical plant layout

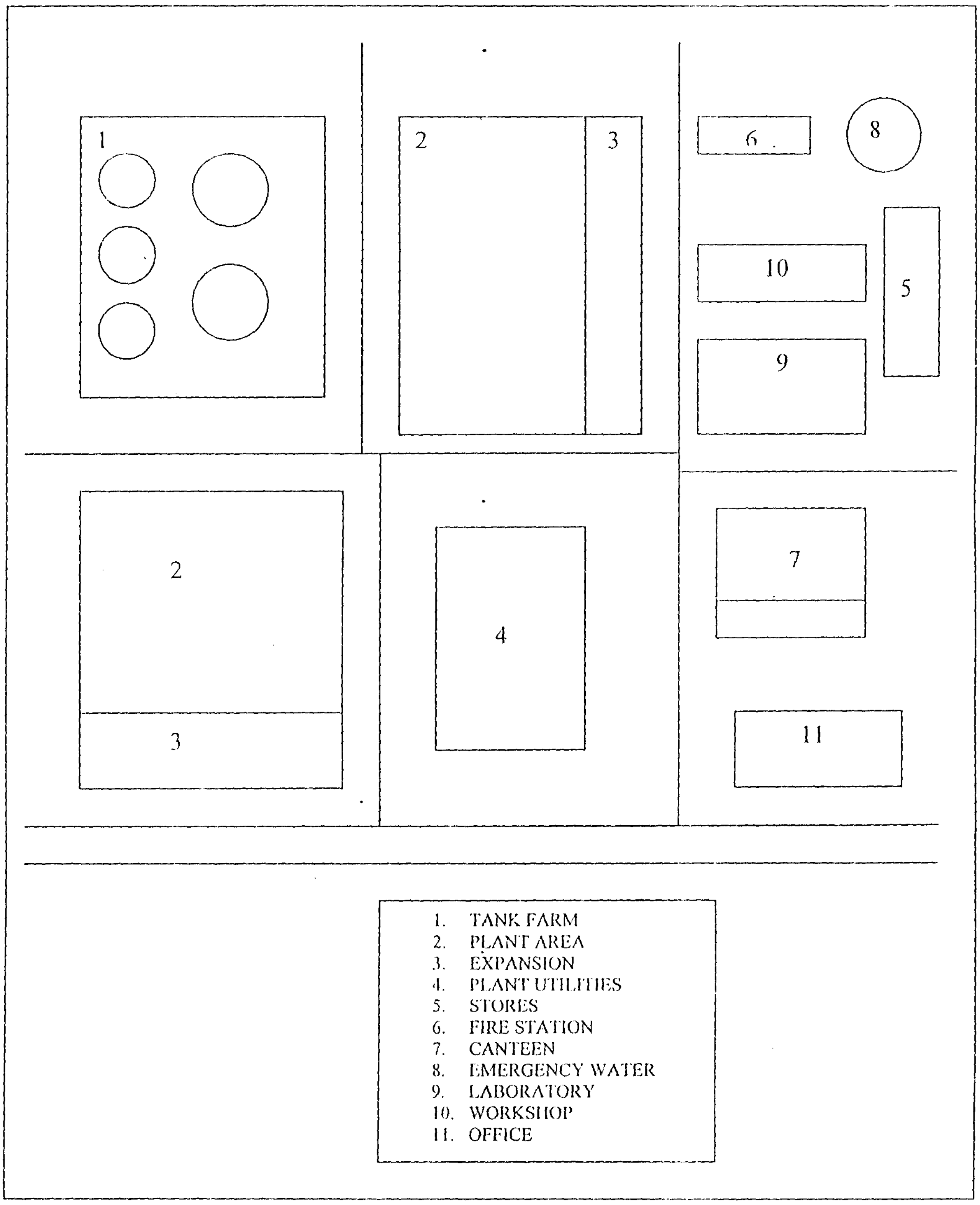


Figure 12.1: A Typical Plant Layout

CHAPTER THIRTEEN

13.0 ECONOMICS ANALYSIS

Chemical plants are built to make a profit, and an estimate of the investment required and the cost of production are needed before the profitability of a project can be assessed. For any industrial plant to be put into operation, huge sum of money must have been invested in it to purchase and install the necessary machinery and equipment. Land and service facilities must be obtained and the plant must be erected complete with the piping, controls, and service. In addition to all these cost, it is necessary to have money available for the payment of expenses involved in plant operation.

The capital requires to have the necessary manufacturing and plant facilities is called the fixed capital investment. That which is necessary for operation of the plant is termed the working capital. The sum of the fixed investment and the working capital is the total capital equipment.

The cost of the gasifier plant is tabulated below.

EQUIPMENT	SURFACE AREA(m^2)	UNIT COST ($/m2)	UNIT COST ($)
SHREDER	0.61	3000	1830
BOILER	0.058	6500	377
PREHEATER	0.45	6500	2925
GASIFIER	116.1	8000	928800
CYCLONE	3.09	4500	13905
MECH FILTER	40.94	3500	143290
SCRUBBER	30.66	4000	122640
FILTER	0.6	3500	2100
CATALYTIC REFORMER	47.06	5000	258830
PUMP	0.058	2500	145
Major Equip Cost			1474842
Mino eqip Cost=20%0f MEC			294968.4
Total Equip Purchase Cost(TCP)			1769810.4
Total Plant Direct Cost			
Euip Purchase Cost			1769810.4
Installation 10%TCP			176981.04
Process Piping 5%TCP			88490.52
Instrumentation/Control11%TCP			194679.14
Insulation %%TCP			88490.52
Electricals 6%TCP			106188.62
Buildings 25%TCP			442452.6
Land 6%TCP			106188.62
Auxillary facilities 40%TCP			707924.16
TPDC			3681205.62
TOTAL PLANT INDIRECT COST (TPIC)			
Engineering %%TCP			88490.52
Construction 5%TCP			88490.52
TPIC			176981.04
Fixed Capital Investment (FCI)			3858186.67

Estimation of Annual Production Cost		
Total Raw Material Cost		
Raw material cost consist of 30% of the average cost of fresh manure in Nigeria		
average cost of of Raw material(2009 prices)	$0.2 per kg	$/yr
Total Raw Material Cost(RMC) = 0.2*100*24*365/0.3 =		584000
Operating Labor Cost 10%FCI		385818.7
Cost of Maintenance 8%FCI		308654.9
Miscellaneous material 10%FCI		385818.7
Total Variable Cost(TVC)		1664292.3
Etimation of Utility Cost		
Utility cost (UC)5% TVC		83214.62
Total Operating Cost(TOC)		1747506.92
Total Revenue		
Selling Price of gas		$13.5/kg
Total Revenue=13.5*28.92*24*365=		$3420079.2/yr
Plant usefulness		5yrs
Salvage Value 10%FCI		$385,818.70
interest rate		8%

$$\text{Return\_on\_investment} = \frac{3420079.2 - 1747506.92}{3858186.67} 100\%$$

$$\text{Return\_on\_investment} = 43.351\%$$

$$\text{Pay\_back\_period} := \frac{1\text{yr}}{\text{Return\_on\_investme}}$$

$$\text{Pay\_back\_period} = 2.307\text{yr}$$

NPV (at 8% interest)

$$\text{NPV} = \frac{C}{(1+r)^n}$$

Where : C = amount

r = rate

n = no of years

$$\text{Annual\_profit} = 3420079.2 - 1747506.9$$

$$\text{Annual\_profit} = 1.673 \times 10^6$$

Present value for each year

$$PV = \frac{profit}{(1 + r)^{n}}$$

$r = interest\_rat$

$n = no\_of\_year$

year	amount	present value
1	1.67E+06	1548678.04
2	1.67E+06	1433961.15
3	1.67E+06	1327741.8
4	1.67E+06	1229390.56
5+salvage	2.06E+06	1400906.314
		6940677.864

Net present value

$$NPV := 6940678 -- 3858186.6$$

$$NPV = 3.082 \times 10^{6}$$

Internal Rate of Return IRR

$$3858186.67 = \frac{6940678}{(1 + IRR)^{5}}$$

$$IRR := 12.46\%$$

CHAPTER FOURTEEN

14.0 RECOMMENDATIONS TO THE INDUSTRIALIST

Having carried out the design of the Gasifier plant, the following recommendations are made to the industrialists to be noted during the procurement, construction, start-up and operating phases of the work:

i. The safety of workers, equipments and infrastructures should be highly evaluated during the design implementation stage of the design.

ii. The recycle design should be considered imperative, since it is useful in reducing the level of plant effluent, and it also reduces costs.

iii. Adequate data and technological parameters should be at the possession of the plant operators at all time to forestall any unwanted accident.

iv. Routine turn around plant maintenance should be of paramount importance in the design. An articulate and organised maintenance team should safeguard plant shut down and ensure equipment salvage value. This will also take care of schedule, slippage, cost over-run and possible re-work.

v. Personnel should undergo routine training about new work ethic and equipments to improve their knowledge of the plant operation and increase overall Gasifier plant productivity.

vi. Procurement of raw materials and equipments should be based on strict regulation of specification and maximum quality.

vii. The plant should not be operated above the design specification to avoid abnormal conditions and explosions.

viii. The implementation of this design work must be adequately supervised by the experts.

ix. The plant should be sited close to the source of raw materials.

x. Alternative sources of energy should be available at all times to avoid plant failure and possible sources of failure.

xi. The water and air around the plant should be monitored regularly to ensure compliance with the Environmental Protection Agency Standards.

REFERENCE

- Anil K. et al (1986), History of Biomass Gasification
- Anil K. et al (1986), Theory of Gasification of Biomass
- Bibao, R., Garcia, L., Salvador, M.L. and Aruzo, J. (1998): "Steam gasification of biomass in a fluidized bed: effect of a Ni-Al catalyst, Biomass for Energy and Industry.10th European Conf. and Tech. Exhibition" Hill book Company New York, USA.
- Gupta, A. K. (1946): "Thermal Characteristics of Gaseous Fuel Flames using High Temperature Air, ASME J. Engineering" .Cambridge, New York. pp1-3
- Jangsawang, W, Klimanek. A, and Gupta A. K. (2006): "Enhanced yield from wastes using high temperature steam gasification" Education Books Limited, London, England
- Rapagna, S., J and N. and Foscolo, P.U. (1998): "Utilization of suitable catalyst for the gasification of biomasses, Biomass" Wiley-interscience publication
- Rensflet, E., Blomkuist, G., Ekstrom, S., Espenas, B.G.& Linanki, L.(1987): "Basic gasification studies for development biomass medium -BTU gasification processes, Energy from Biomass and Wastes, IGT." Book Company Singapore
- Odigure, J. O. (1998), "Safety, Loss and Pollution Prevention in Chemical Process Industries", Jodigs and Associates, Nigeria.pp122-134