DESIGN, CONSTRUCTION AND EXPERIMENTAL TESTING OF MULTI-AIR-STAGE WOOD GASIFICATION

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ID: 55300700501

A THESIS SUBMITTED AS A PART OF THE REQUIREMENTS
FOR THE DEGREE OF MASTER OF ENGINEERING
IN ENERGY TECHNOLOGY AND MANAGEMENT

THE JOINT GRADUATE SCHOOL OF ENERGY AND ENVIRONMENT
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ABSTRACT

This study focuses on gasification system design and construction of a wood biomass in a multi-air-stage downdraft gasifier. The gasification system consists of a multi-air-stage downdraft gasifier, equipped with a cyclone, a heat exchanger, a bag filter, a blower and a flare tower. The design calculation was developed based on the required gasification process power output and air equivalence ratio (ER). To find out the dimension for each part, the measurement refers to the essential parameter of each equipment, the theoretical and the experimental data obtainable from several literatures. The gasifier had 0.15 m in internal diameter, 0.1 m in refractory wall thickness and 2.1 m in height. Separated into hopper, gasifier and ash chamber, it has three stages of air injection. Cyclone diameter was 0.127 m with 0.064 m in inlet height and 0.031 m in inlet width. A shell and tube heat exchanger design as one-shell and two-tube pass, 0.46 m in diameter, 2 m in length, and use 24 tubes in total. There was four requires bag for bag filter. A one horse power induced draft blower was used to absorb the synthetic gas and at the end this system finalized with a 2 m long flare tower to burn the gas. The outcome design was used for the experimental testing of multistage wood gasification.

Experimental testing was conducted on the outcome design of multi-air-stage wood gasification. Measuring the temperature distribution along the gasifier wall; the CO, CO_2, CH_4, and H_2 gas concentration was analyzed; the heating value of producer gas, tar analysis, and gasification efficiency were calculated. It showed that the multi air-stage gasifier followed the right pattern of downdraft gasifier. The average hot gas and cold gas efficiencies of this design are 11.79% and 28.05% for wood chips and 17.09% and 34.06% for wood chip-wood pellet mixture respectively.

Key words: Design, Biomass, Gasification, Downdraft, Multistage.

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CHAPTER 1
INTRODUCTION

1.1. Rationale/Problem Statement

Data published in 2009 by The International Energy Agency (IEA) illustrated that in 2030 Indonesia will be responsible for about 36% of the overall increase in ASEAN's energy demand, followed by Thailand (18%), Malaysia (11%) and Philippines (9%) consecutively. The major sources of energy are shown in Figure 1.1.

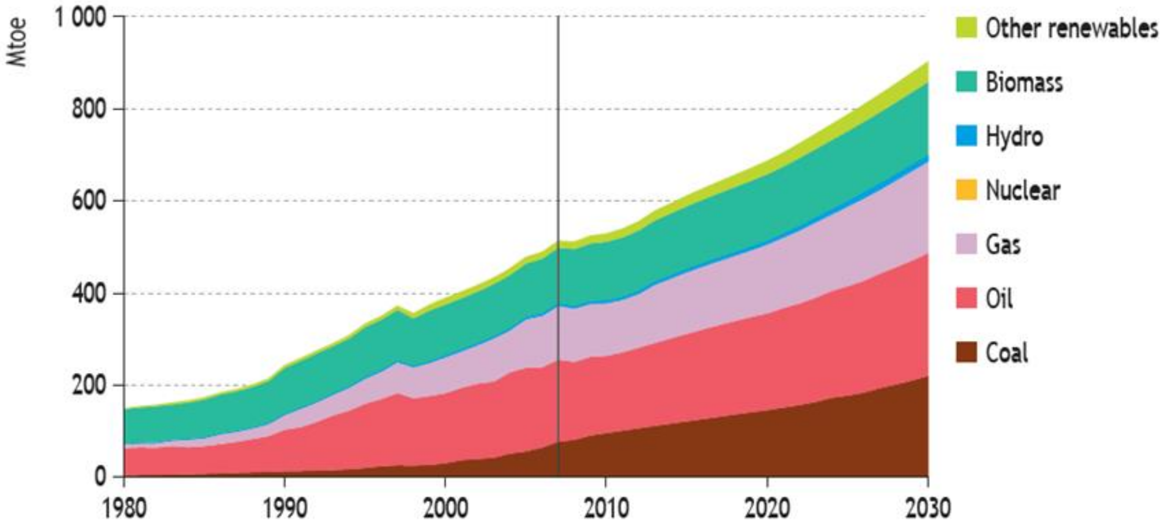


Figure 1.1 ASEAN Primary Energy Demand by Fuel (IEA, 2009)

Such increases were caused by a long running oil crisis, starting from The Arab oil embargo in 1973, then another oil crisis in 1979 due to The Iranian revolution. Further, the Iraq-Iran tanker war was at 1981-1988, then Gulf war at 1990 which would end with global energy crisis at 2009. The world is now looking for alternate new energy resources, which is renewable energy. Renewable energy has many available forms, such as wind, hydro, solar, nuclear, and biomass. At last few decades, biomass was considered only as an agricultural residue, until people invented gasification technology.

Gasification increases the value of noncommercial feedstock into marketable fuel by converting them using thermo-chemical conversion into a producer gas consisting of CO and H_2 as the major contents (Pereira, 2012). Other advantages of gasification are the producer gas easily transported by pipeline, as for chemical synthesis and being used in

Internal Combustion (IC) engine. Then gasification technology can make more efficient and clean combustion processes (Knoef, 2005).

Biomass is likely to be more suitable than fossil fuels because of its low price, lower greenhouse gas emissions and huge amount of resources. In the contrary, biomass has a serious problem in tar contents, affecting stack and fouling in the rector design and reduce producer gas quality. Even many reactor designs have been developed to analyzethe gasifier performance, however the fixed one model could not be found to handle all the type of biomass.

1.2. Literature Review

1.2.1 Biomass Gasification

Dogru et al., (2002a) studied the gasification of hazelnut shells in a downdraft gasifier using air as the gasifying agent. The results showed the optimum operation of the gasifier was found between 1.44 and 1.47 Nm^3/kg of air fuel ratios at the values of 4.06 and 4.48 kg/h of wet feed rate which produces the producer gas with a good GCV of about 5 MJ/m^3 at a volumetric flow of 8-9 Nm^3/h product gas. At this optimum, low tar and char were produced at a ratio of 0.005 and 0.051 of the feed. Maximum temperatures in drying, pyrolysis and throat zones were determined as 125, 566 and 1206^oC but the throat temperature fell about 1020^oC at the optimum. With hazelnut shells there was no sign of bridging or ash fusion at the optimum throat temperature of about 1000-1050^oC. It was concluded that hazelnut shells could be easily gasified in a downdraft gasifier to produce good quality gas with minimum polluting by-product.

Dogru et al., (2002b) studied the gasification of sewage sludge using a throated downdraft gasifier. The combustible gases, produced from sewage sludge, contain as volume by volume percentages: 8.89-11.17% H_2, 1.26-2.09% CH_4, 6.28-10.77% CO, 0.62-0.95% C_2H_2 and 0.15-0.27% C_2H_6. The quality of product gas was 4 MJ/m^3. The most productive region of the gasification of sewage sludge could be selected around 514.60 Nm^3/m^2h ± 1.50% of specific gasification rate. The optimum efficiency of cold gas was estimated between 63.54% and 65.52% around 514.60 Nm^3/m^2h ± 1.50% of specific gasification rate at which the highest ratio of H/C was estimated as 0.198 kgH_2/kgC.

Na et al., (2003) studied the gasification of combustible waste mixed with plastic and cellulosic in a fixed bed gasifier at 1100-1450^oC. The results showed that as the bed height was increased, the formation of H_2 and CO increased, while the CO_2 decreased by the char-CO_2 reaction and plastic cracking. The cold gas efficiency was around 61% and the heating values of product gases were in the range of 2800-3200 kcal/Nm^3.

Erlich et al., (2011) studied the downdraft gasification of pellets made of sugar-cane bagasse from sugar/alcohol production and empty fruit bunches (EFB) from palm oil production with an air gasifying medium. The result showed gasification of wood pellets provides a richer producer gas while EFB pellets give a poorer one with higher contents of non-combustible compounds.

Ahmed et al., (2011a) studied the waste rubber in high temperature steam gasification in a semi-batch reactor. Rubber gasification yielded more energy at the 900^oC as compared to hard wood and wood chips. But, less syngas and less hydrogen were obtained from rubber than hard wood and wood chips sample.

Ahmed et al., (2011b) studied the gasification of polyethylene (PE) and woodchip (WC) mixtures in a semi-batch reactor, using high temperature steam as the gasifying agent at a reactor temperature of 900^oC. Superior results in terms of syngas yield, hydrogen yield, total hydrocarbon yield, energy yield and apparent thermal efficiency have been shown from PE-WC blends at PE percentage of approximately 60-80% as compared to the expected weighed average yields from the individual components. Small amounts of biomass (WC) material to the PE sample resulted in higher energy yield than that obtained from 100% PE sample. These results revealed the importance of input feed composition to a gasifier.

Thanapal et al., (2012) studied a fixed bed gasification of dairy biomass with an enriched air mixture. Gasifying medium used enriched oxygen varying from 24% to 28% oxygen on volume basis. The results showed peak temperature and carbon dioxide production increased with corresponding decrease in carbon monoxide with increase in oxygen concentration in the incoming gasification medium. HHV of the gases decreased as well as oxygen concentration.

Chiang et al., (2012) investigated the characteristic of 26 varieties of biomass produced from forestry, agriculture, municipality, and industry in Taiwan to test their applicability to thermal conversion technology. The experimental results indicated that the volatile matter content of the all tested biomass was 60% and above. It meant that the higher carbon conversion rate will occur in the thermal conversion processes of all tested biomass. Lower heating value (LHV) of MSW and non-hazardous industrial sludge was lower than other tested biomass that was 1000 to 1800 kcal/kg. On the other hand, the LHV of other tested biomass and their derived fuels was similar to the tested coal. However, the energy density of wood and agricultural waste were smaller than coal, because of their bulk density were low.

Naqvi et al., (2012) studied the synthetic natural gas (SNG) production from dry black liquor gasification (DBLG) using oxygen-blown pressurized circulating fluidized bed (CFB) black liquor gasification. The results indicated a large potential of SNG production (about 162 MW) from black liquor, the cold gas energy efficiency was about 58%, then about 700 ktonnes of CO_2 abatement per year.

Ahmed et al., (2012) studied the energy recovery from the pyrolysis and the gasification of mangroves. The mangrove pyrolysis yielded more syngas than food waste pyrolysis (13.5 g) and paper pyrolysis (15.1 g) at the same experimental conditions. Hydrogen yield from mangrove pyrolysis was 0.43 g at 900^oC. It was 50 % higher than food waste pyrolysis (0.2 g) and paper pyrolysis (0.24 g). Higher syngas and hydrogen yield from mangrove was mainly attribute by high volatile content in the sample. Gasification yielded more syngas and energy than pyrolysis. Syngas and energy yield from gasification was almost double than pyrolysis at 900^oC. Hydrogen yield from gasification was more than five times than pyrolysis at 900^oC.

Nipattummakul et al., (2012a) studied oil palm trunks for steam gasification in a batch gasifier. Reactor temperature was fixed at 800^oC. The result showed that oil palm trunk waste yielded more syngas, energy and hydrogen than mangrove wood, paper and food waste.

Ouadi et al., (2012) studied the gasification of paper industry waste using a fixed bed downdraft as gasifier. Paper industry waste consisted of pulper rejecting, de-inking sludge and wood chips. The results showed that 80 wt% of a brown paper mill’s rejects blend with 20 wt% wood chips could be successful gasified. The producer gas composition was 16.24% H_2, 23.34% CO, 12.71% CO_2, 5.21% CH_4 and 42.49% N_2 with a higher heating value of 7.3 MJ/Nm^3.

1.2.2 Gasifying Medium

Lv et al., (2007) studied the characteristics of hydrogen production from pine wood block by using different gasifying agents in a downdraft gasifier. The results indicated both of biomass air gasification and biomass oxygen/stem gasification improves hydrogen yield depending on the volume of downdraft gasifier, and also nearly doubles the heating value of fuel gas. For biomass oxygen/steam gasification, the content of H_2 and CO are 63.27-72.56%, then for biomass air gasification was 52.19-63.31%.

Ahmed et al., (2009a) studied the characteristics of cardboard and paper gasification using CO_2 as the gasifying agent. The result showed that H_2 mole fraction peaks at early stages of the process which was attributed mainly to the sample pyrolysis. However, CO_2 mole fraction kept increasing with time as a result of sample decay with time in the reactor. On the other hand, CO mole fraction decreased monotonically with time and its formation was attributed to the gasification of char with CO_2 to produce CO.

Umeki et al., (2010) studied a high temperature steam gasification process to generate hydrogen-rich fuel gas from woody biomass. The results showed both of the steam temperature and the molar ratio of steam to carbon (S/C ratio) affected the reaction temperature which strongly affects the gasified gas composition through the water-gas shift reaction. The H_2 fraction in the produced gas was 35-55 vol%. The tar concentration in the produced gas from the high temperature steam gasification process was higher than that from the oxygen-blown gasification process. The highest cold gas efficiency was 60.4 % which the ideal cold ga efficiency of the whole system with heat recovery process was 71%.

Nipattumakul et al., (2010b) studied high temperature steam gasification of wastewater sludge. Steam gasification yielded more syngas, energy and higher apparent thermal efficiency as compared to that obtained from pyrolysis at the same temperature of 1173 K. peak value of syngas yield, energy yield, and hydrogen yield was obtained at S/C ratio of 5.62.

Thanapal et al., (2012) studied fixed bed gasification of dairy biomass using a different gasifying medium. The gas mixture with different steam fuel ratio has low heat value due to large amount of diluents nitrogen. The gases produced using carbon dioxide and oxygen mixture has a higher HHV when compared to air and enriched air gasification.

Hernadez et al., (2012) studied the effect of adding steam to the air gasifying agent in biomass entrained flow gasification. Dealcoholised march of grape used as the fuel. The steam content in the gasifying medium varied from 0% (air gasification) to 100% (steam gasification). The optimal range of air-steam mixture content is 40-70% mol steam. Temperature increase as well as CO and H_2 content in the product gas for air gasification. On the other hand air-steam gasification leads to boost the H_2 production at higher temperatures, as well as the CH_4 content.

Nipattummakul et al., (2012a) investigated the steam gasification of oil palm trunk waste using a batch type gasifier at 800°C. Results showed approximately over 50% of the total syngas generated was obtained during the first 5 minutes of the process An increase in steam flow rate accelerated the gasification reactions and resulted in reduced gasification time. Variation in steam flow rate slightly affected the apparent thermal efficiency and was found to be very high. It concluded steam gasification as one of the most efficient approach for waste to clean energy conversion.

1.2.3 Biomass Gasifier

Ueki et al., (2011) studied pellets of black pine wood gasification under both updraft and downdraft gasifier conditions using air as the oxidizing agent. Lower heating value of the syngas under updraft and downdraft condition were 4.8 and 3.8 MJ/Nm^3. it was easier to control the height of the packed bed under the downdraft condition than the updraft condition. Under the updraft condition, a bridging phenomenon occurred. Temperature fluctuations under downdraft conditions are larger than updraft conditions, causing unstable producer gas composition. Tar generation under the downdraft condition was lower than the updraft condition. This is because tar passes through a partial combustion zone or higher temperature zone in the downdraft gasifier.

Zhang et al., (2012) studied the gasification of municipal solid waste in the plasma gasification melting (PGM) process. A series of tests were conducted to study the performance of PGM gasification. The results imply LHV of the syngas varied from 6 to 7 MJ/Nm^3. For air gasification, the syngas yield increased significantly with increasing equivalence ratio (ER), whereas the LHV decreased slightly. Feeding high-temperature steam into the PGM reactor greatly increased syngas yield, with even more LHV. It also reduced the air demand for gasification. The energy efficiency of air and steam gasification was much higher than that of air gasification. The cold-gas efficiency (CGE) of PGM air and steam gasification can reach approximately 60%. Tar formation represents the main energy loss for the PGM reactor.

Prabu et al., (2012) studied underground coal-air gasification-based solid oxide fuel cell systems (SOFCs) using air as the gasifying medium. The syngas produced from underground coal gasification (UCG) process using dry air contains more steam, tar and higher hydrocarbon compared to the conventional gasifier. A UCG system is integrated with the SOFC system taking advantage of the high temperature exhaust from the latter to reform the syngas for producing hydrogen.

1.2.4 Stages of Gasifier

Bhattacharya et al., (1999) studied two-stage wood gasification, using Yang wood and Para wood as the raw materials. An increase in the secondary air flow of the two-stage gasifier resulted in decrease of the tar content and the CO_2 and H_2 concentrations while that of CO increased. The tar content of the product gas was in the range 19-34mg/Nm^3 for a charcoal gasifier couple to a two-stage wood gasifier. With a floating-drum gas-storage system and a 3.5 retention time, the tar content was reduced to 9.24 mg/Nm^3, which is 85% less than that obtained by using the a two-stage wood gasifier alone.

Jaojaruek et al., (2011) studied the gasification of Eucalyptus wood on three different downdraft gasification approaches, which were single stage, conventional two-stage, and an innovative two-stage air and premix air/gas supply. The innovative two-stage gasifier has two nozzle locations, one for air supply at combustion zone and the other at the pyrolysis zone to supply the premixed gas. The premixed gas came from the producer gas which is partially bypassed to mix with air. The result showed that the producer gas quality generated by the innovative two-stage approach was better than the conventional two-stage. The higher heating value (HHV) increased from 5.4 to 6.5 MJ/Nm^3. Moreover, tar content in producer gas reduced to less than 45 mg/Nm^3 compared to that of conventional one.

Martinez et al., (2011) studied the gasification of Eucalyptus wood in a moving bed downdraft gasifier with two-air supply stages. The result showed an appositive effect of the secondary stage over the reduction of the CH_4 concentration, which had influents with the decreases of the tar content in the producer gas composition. After that, the biomass devolatilization in the pyrolysis zone also gave much lighter compounds which were more easily cracked when the gas stream passed through the combustion zone.

Ma et al., (2012) investigated the Camphor wood chips gasification using a double air stage downdraft gasifier. Wood chips converted by polygeneration gasification proceeds into three phases, which is gas, solid and liquid. The result showed that the secondary air supply yielded temperature as high as 900^{o}C in the oxidation zone. It is greatly helped to crack the tar and improved the quality of the producer gas. The flow rate of producer gas was 500 Nm^3/h, charcoal 60 kg/h, and the extract 65 kg/h. The producer gas compositions were 16.6$_{vol}$% CO, 2.3$_{vol}$% CH_4, 16.1$_{vol}$% H_2, 13.8$_{vol}$% CO_2 and 0.4$_{vol}$% O_2 with an average of LHV is 4.7 MJ/Nm^3.

1.3. Objectives

The objectives of this study are:

1. To design and construct a 50 kWth multi air-stage downdraft wood gasification system.
2. To test the 50 kWth multi air-stage downdraft wood gasification system.
3. To analyze the producer gas composition, tar content, gas conversion efficiency and heating value of producer gas in a multi-air-stage gasifier.

1.4. Scopes of Research Work

1. The raw material was Eucalyptus wood chips, which was supplied by Asia Char Coal. Ltd.
2. The gasifying medium was pre-heated air.
3. A newly multi-air-stage downdraft gasifier was designed and constructed in collaboration with Thai Steam Service & Supply Co. Ltd.
4. A series of purification equipment, such as Cyclone, Heat Exchanger and Bag Filter were designed and constructed in collaboration with Susuke Metal Work Company.
5. A new workshop and control room was designed and constructed in collaboration with T-Con Engineering & Construction Co. Ltd.
6. The gasifier temperature distribution, composition of producer gas, tar content and gas conversion efficiency were measured.

CHAPTER 2
THEORIES

2.1. Gasification

Gasification converts fossil or non-fossil fuels (solid, liquid, or gaseous) into useful gases and chemicals. It requires a medium for reaction, which can be gas or supercritical water. Gaseous media include air, oxygen, subcritical steam, or a mixture of these. There are three major motivations for such a conversion (Basu, 2010) :

- To increase the heating value of the fuel by rejecting noncombustible components, such as nitrogen and water.
- To remove sulfur and nitrogen such that when burnt, the gasified fuel does not release them into the atmosphere.
- To reduce the carbon-to-hydrogen (C/H) mass ratio in the fuel.

Gasification is also called “staged combustion”, since usually the gas is produced with the intent to burn it later. This raises the question why firstly gasifying and then burning the gas over direct combustion of the biomass. There are several essential advantages (Reed, 2000) :

- Possibility to transport in pipelines.
- Ease of control and continuous operation.
- Clean combustion of produced gas since impurities can be removed from the producer gas, and the volume of producer gas is much smaller than that of flue gas.
- Efficient and clean combustion since the exact required air can be mixed from optimum combustion.
- Producer gas can be used in IC engines or turbines with higher efficiencies over steam devices.
- Producer gas, particularly syngas, can be used for chemical synthesis, such as for fertilizers and transportation fuels.

From a chemical point of view, the process of biomass gasification is quite complex. It includes a number of steps (Knoef, 2005):

- Thermal decomposition to gas, condensable vapors and char.

- Subsequent thermal cracking of vapors to gas and char.
- Gasification of char by steam or carbon dioxide.
- Partial oxidation of combustible gas, vapors and char.

Combustion, occurring in the oxidation zone, is described by the following heterogeneous chemical reactions (Ahmed, 2009b) :

Complete oxidation of solid carbon $C + O_2 \rightarrow CO_2$ $\Delta H = -394$ kJ/mol (1)
Partial oxidation of solid carbon $C + ½\ O_2 \rightarrow CO$ $\Delta H = -111$ kJ/mol (2)
Oxidation of hydrogen $H_2 + ½\ O_2 \rightarrow H_2O$ $\Delta H = -242$ kJ/mol (3)
Oxidation of carbon monoxide $CO + ½\ O_2 \rightarrow CO_2$ $\Delta H = -283$ kJ/mol (4)

These reactions are exothermic and provide – by autothermal gasification – the heat necessary for the endothermic reaction in the drying, pyrolysis and reduction zones. Although Reactions (3) and (4) produce heat, which is beneficial to the gasification process, they are not desired because they reduce the heating value (Higman, 2003).

The devolatilisation step (pyrolysis) is slightly endothermic and, for temperatures above 500^{o}C, 75 to 90 wt% volatile matter is produced in the form of steam plus gaseous and condensable hydrocarbons. The water vapour is introduced with the air and production by drying and pyrolysis of the biomass reacts with the hot carbon according to the following heterogeneous reactions (Knoef, 2005):

Water gas Reaction $C + H_2O \rightarrow CO + H_2$ $\Delta H = +131$ kJ/mol (5)

The most important reduction reactions are the water gas reaction (5) and the following Boudoard Reaction (Knoef, 2005):

Boudouard Reaction $C + CO_2 \rightarrow 2\ CO$ $\Delta H = +172$ kJ/mol (6)

These heterogeneous endothermic reactions increase the gas volume of CO and H_2 at higher temperatures and low pressures. Besides these reactions, several other reduction reactions take place, of which the most important ones worthwhile mentioning are the water gas shift reaction (7) and the methanation reaction (8) (Sarasuk, 2010):

Water gas shift reaction $CO + H_2O \rightarrow CO_2 + H_2$ $\Delta H = -41$ kJ/mol (7)

Methanation reaction $C + 2\ H_2 \rightarrow CH_4$ $\Delta H = -75$ kJ/mol (8)

The final product gas contains carbon monoxide, hydrogen and methane as the desired components, next to steam and carbon dioxide. Its precise composition is determined by the water gas shift equilibrium, which is a homogeneous gas phase reaction between steam plus carbon monoxide (7).

Other homogeneous gas reactions also take place in the heterogeneous reactions in the following way (Basu, 2006; Kaewluan, 2009):

Steam methane reforming $CH_4 + H_2O \rightarrow CO + 3\ H_2$ $\Delta H = +206$ kJ/mol (9)

CO_2 reforming reaction $CH_4 + CO_2 \rightarrow 2\ CO + 2\ H_2$ $\Delta H = +247$ kJ/mol (10)

Despite a favorable equilibrium constant at 900^{o}C, methane reformation to CO and H_2 does not occur at atmospheric pressure due to the slow rate of reaction.

Apart from decomposition reactions, char gasification includes several heterogeneous gas-solid reactions where solid carbon is converted to CO or H_2 by steam and carbon dioxide. The gas-solid reactions are slow compared to the gas phase reaction, and may limit the overall rate of the gasification process and the carbon conversion efficiency.

2.2. Gasifying Medium

Gasifying agents react with solid carbon and heavier hydrocarbon to convert them into low-molecular-weight gases, such as CO and H_2. The specification of the gasifying medium is based on the selection of steam, oxygen, and/or air and their proportions (Basu, 2010).

These parameters could influence the gasifying medium chosen, as follows :

- The desired heating value of the product gas. Table 2.1 gives typical ranges of heating value for different media.
- Hydrogen can be maximized with steam, but if it is not a priority, oxygen or air is a better option, as it reduces the energy used in generating steam and the energy lost through unutilized steam.

- If nitrogen in the product gas is not acceptable, air cannot be chosen.
- Capital cost is lower for air, followed by steam. A much larger investment is needed for an oxygen plant, which also consumes a large amount of auxiliary power.

Table 2.1 LHV of Product Gas Ranges and Choice of Gasifying Medium

Gasification Medium	Range of Heating Value of Product Gas (MJ/Nm^3)
Air Gasification	4 – 7
Steam Gasification	10 – 18
Oxygen Gasification	12 – 28

Steam gasification favors steam reforming reactions, while the air gasification promotes combustion reaction. The yield of H_2 and hydrocarbon with steam gasification are higher than those with air injectors, whereas CO and CO_2 are lower, since the extent of combustion of char and volatiles is reduced by replacing air with steam (Ahmed, 2009b).

Generally, gasification used ambient air at 300K, but in fact, conversion efficiency increases as the inlet air temperature increases, because hot air provides additional enthalpy necessary for the reaction, thereby decreasing the air/fuel ratio (Jayah, 2003).

2.3. Biomass

Biomass refers to any organic materials that are derived from plants or animals (Loppinet-Serani et al., 2008). A generally accepted definition is difficult to find. However, the one used by the United Nations Framework Convention on Climate Change (UNFCCC, 2005) is non-fossilized and biodegradable organic materials originating from plants, animals and micro-organisms. These also include products, by-products, residues and wastes from agriculture, forestry and related industries as well as the non-fossilized and bio-degradable organic fractions of industrial and municipal wastes.

Biomass is very versatile feedstock in its morphology and physical characteristics. It can be quite wet or dry, very dense or fluffy, high or low ash containing, small in shape or large, homogeneous or inhomogeneous, etc. This makes the use of biomass fuels in dedicated gasifier reactors quite difficult, and in most cases some pretreatment of the biomass is needed (Knoef, 2005).

The ideal biomass feedstock for thermal conversion, whether it be combustion, gasification, or a combination of both, is one that contains low or zero levels of elements, such as nitrogen, sulfur, or chlorine. This can form undesirable pollutants and acids which cause corrosion, no mineral elements and form inorganic ash and particulate. Biomass is likely similar to some coals with respect to total ash content, but due to its biomass diversity, several species and types have relatively low ash and also low sulfur contents. The compositions of wood compared to those of other potential biomass feed stocks make woody biomass a preferred feedstock for thermal gasification as shown in Table 2.2 (Klass, 1998). If wood chips are given sufficient sun exposure, the moisture content can be reduced to 15%, giving a conversion efficiency of 56% (Jayah, 2003).

Table 2.2 Typical Proximate Analyses and High Heating Values of Representative Biomass, Coal and Peat

Category	Name	Type	Moisture Range (wt %)	Organic matter (dry wt %)	Ash (dry wt %)	High Heating Value (MJ/dry kg)
Wastes	Cattle Manure	Feedlot	20 – 70	7.5	23.5	13.4
	Activated biosolid	Sewage	90 – 97	76.5	23.5	18.3
	Primary biosolid	Sewage	90 – 98	73.5	26.5	19.9
	Refuse-derived fuel (RFD)	Municipal	15 – 30	86.1	13.9	12.7
	Sawdust	Woody	15 – 60	99.0	1.0	20.5
Herbaceous	Cassava	Tropical	20 – 60	96.1	3.9	17.5
	Euphorbia lathyris	Warm season	20 – 60	92.7	7.3	19.0
	Kentucky bluegrass	Cool season	10 – 70	86.5	13.5	18.7
	Sweet sorghum	Warm season	20 – 70	91.0	9.0	17.6
	Switchgrass	Warm season	30 – 70	89.9	10.1	18.0

Category	Name	Type	Moisture Range (wt %)	Organic matter (dry wt %)	Ash (dry wt %)	High Heating Value (MJ/dry kg)
Aquatic	Giant brown kelp	Marine	85 – 97	54.2	45.8	10.3
	Water hyacinth	Fresh water	85 – 97	77.3	22.7	16.0
Woody	Black alder	Hardwood	30 – 60	99.0	1.0	20.1
	Cottonwood	Hardwood	30 – 60	98.9	1.1	19.5
	Eucalyptus	Hardwood	30 – 60	97.6	2.4	18.7
	Hybrid poplar	Hardwood	30 – 60	99.0	1.0	19.5
	Loblolly pine	Softwood	30 – 60	99.5	0.5	20.3
	Redwood	Hardwood	30 – 60	99.8	0.2	21.0
	Sycamore	Hardwood	30 – 60	98.9	1.1	19.4
Derivatives	Paper		3 – 13	94.0	6.0	17.6
	Pine bark	Softwood	5 – 30	97.1	2.9	20.4
	Rice straw		5 – 15	80.8	19.2	15.2
	Redwood char		2 – 6	95.9	4.1	30.5
Coal	Illionois bituminous	Soft	5 – 10	91.3	8.7	28.3
	North Dakota lignite	Soft	5 – 15	89.6	10.4	14.0
Peat	Reed sedge	Young coal	70 – 90	92.3	7.7	20.8

Source: (Klass, 1998)

2.4. Downdraft Gasifier

Basically, biomass gasifiers can be categorized into several reactor design groups : a descending bed of biomass, often referred to as a moving or fixed bed, with countercurrent gas flow (updraft); a descending bed of biomass with concurrent gas flow (downdraft); a descending bed of biomass with cross flow of gas; a fluidized bed of biomass with rising gas; an entrained-flow circulating bed of biomass; and tumbling beds. All of them have their own specifications as described in Table 2.4.

Table 2.3 Operating Conditions of Various Gasifiers.

	Downdraft	**Updraft**	**BFB**	**CFB**	**EF**
T°C	700-1200	700-900	< 900	< 900	~ 1450
Tars	Low	High	Moderate	Moderate	Very low
Control	Easy	Very easy	Moderate	Moderate	Complex
Scale (MW_{th})	< 5	< 20	10-100	20-??	> 100
Feedstock	Very critical	Critical	Less critical	Less critical	Only fine

Source : (BTG, 2002)

Gasifiers can be classified into different ways according to the gasification agent, air-blown gasifiers, oxygen gasifiers and steam gasifiers. According to heat for gasification, there are autothermal or direct gasifiers (heat is provided by partial combustion of the biomass), allothermal or indirect gasifiers (heat is supplied from an external source through heat exchanger or indirect process, i.e. separation of gasification and combustion zones). According to pressure in the gasifier, they are atmospheric and pressurized gasifier. According to the design the reactor, they are fixed bed, fluidized bed and entrained flow (Rauch, 2003).

In a downdraft gasifier, biomass is fed at the top and the air intake is also at the top or from the sides. The gas leaves at the bottom of the reactor, so the fuel and the gas move in the same direction. The same zones can be distinguished as in the updraft gasifier, although the order is somewhat different. Adding air to the char zone is an excellent approach for achieving low tar gas.

The main advantage of a downdraft gasifier is the production of a gas with low tar content, which is nearly suitable for engine applications, as shown in Table 2.5. The drawbacks of this gasifier are limited in scalling-up design, the high amounts of ash and dust particles in the gas due to the fact that the gas has to pass the oxidation zone where small ash particles are entrained. This leads also to a relative high temperature of the leaving gases resulting in lower gasification efficiency.

Table 2.4 Some Characteristic of Fixed Bed Gasifiers

	Updraft	**Downdraft**	**Cross draft**	**Open Core**
Fuel (wood)			**(charcoal)**	
- Moist content (% wet basis)	43 (max 60)	12 (max 25)	10 – 20	7 – 15 (max 15)
- Ash content (% dry basis)	1.4 (max 25)	0.5 (max 6)	0.5 -10	1 – 2 (max 20)
- Size (mm)	5 – 100	20 – 100	5 – 20	1 – 5
Gas exit temp (℃)	200 – 400	700	1250	250 – 500
Tars (g/Nm^3)	30 – 150	0.015 – 0.5	0.01 – 0.1	2 – 10
Sensitivity to load fluctuation	Not sensitive	Sensitive	Sensitive	Not sensitive
Turn down ratio	5 – 10	3 – 4	2 – 3	5 – 10
HG full load (%)	90 – 95	85 – 90	75 – 90	70 – 80
CG full load (%)	40 – 60	65 – 75	70 – 85	35 – 50
Producer gas LHV(kJ/Nm^3)	5.0 – 6.0	4.5 – 5.0	4.0 – 4.5	5.5 – 6.0

Source : (Knoef, 2005)

2.5 Gasification Efficiency

Gasifier performance is measured in terms of both quality and quantity of gas produced. Gasification efficiency is expressed in terms of quantity. It is the amount of biomass converted into gas. Otherwise, the product quality is measured in terms of heating value as well as amount of desired product. The gasification efficiency is expresses as cold-gas efficiency and hot-gas efficiency (Basu, 2010).

Cold-gas efficiency is the energy input over the potential energy output, expressed as :

$$\eta_{cg} = \frac{Q_g M_g}{LHV_f M_f} \qquad \text{(Eq. 1)}$$

Where η_{cg} = Cold-gas Efficiency (%)

Q_g = LHV of product gas (MJ/kg)

M_g = Mass of product gas (kg)

LHV_g = LHV of the solid fuel (MJ/kg)

M_f = Mass of solid fuel (kg)

Hot-gas efficiency is taking the sensible heat of the hot gas into account, expressed as :

$$\eta_{hg} = \frac{Q_g M_g + M_g C_p (T_f - T_o)}{LHV_f M_f} \qquad \text{(Eq. 2)}$$

Where η_{hg} = Hot-gas Efficiency (%)

Q_g = LHV of product gas (MJ/kg)

M_g = Mass of product gas (kg)

C_p = Specific Heat of gas (kJ/kmol.K)

T_f = The gas temperature at the gasifier exit (K)

T_o = The temperature of fuel entering the gasifier (K)

LHV_g = LHV of the solid fuel (MJ/kg)

M_f = Mass of solid fuel (kg)

2.6 Tar Production

Tar is a complex mixture of condensable hydrocarbons, including oxygen-content, 1-to-5 ring aromatic, and complex polyaromatic hydrocarbons (Devi et al., 2003). The previous researcher defined tar as an all organic contaminant with a molecular weight larger than 78. It is larger than benzene (Neeft et al., 1999). There is no universal accepted definition of tar. The International Energy agency (IEA) Bioenergy Agreement, the U. S. Department of Energy (DOE), and the DGXVII of the European Commission agreed to identify all components of product gas having a molecular weight higher than benzene as tar (Knoef, 2005). Tar is a thick, black, highly viscous liquid that condenses in the low-temperature zones of gasifiers, clogging the gas passage and leading to system disruptions (Basu, 2010).

Nevertheless, tar is an unavoidable by-product of the thermal conversion process. The presence of tar in the product gas from gasification can potentially decide the usefulness of the gas. That way, there is a limitation of tar presence in gas product as shown in Table 2.5.

Table 2.5 Upper Limits of Biomass Gas Tar and Particulate

Application	Particulate (mg/Nm^3)	Tar (mg/Nm^3)
Direct combustion	No limit specified	No limit specified
Syngas production	0.02	0.1
Gas turbine	0.1 – 120	0.05 – 5
IC engine	30	50 – 100
Pipeline transport		50 – 500 for compressor
Fuel cells		< 1.0

Source : Milne et al., 1998

Several options are available for tar reduction depending on where tar removal takes place. They are primary tar reduction (in-situ), which avoid tar formation inside the gasifier while secondary tar reduction (post-gasification) is strip the product gas of the tar already product on the outside of gasifier (Basu, 2010; Devi et al., 2003).

In gasification process with air as the gasifying medium, both the yield and concentration of tar in the product gas decrease with an increase in the ER. Above an equivalence ratio of 0.27 phenols are nearly all converted and less tar is formed (Kinoshita et al.,1994). At a higher ER, the fraction of PHP, benzene, naphthalene, and other 3- and 4-ring aromatic increase. While higher ER reduces the tar, it reduces the quality of the gas as well (Basu, 2010).

CHAPTER 3
METHODOLOGY

3.1 Raw Materials

The raw materials for this experiment were Eucalyptus wood chips, Eucalyptus wood pellets, and Eucalyptus wood blocks. Moisture dried was supplied by a solar dryer. The ultimate and proximate analyses will be measured with ASTM standard as shown in Table 3.1.

Figure 3.1 Raw Material Sample

The size of the fuel wood chips is an important parameter that affects the char conversion rate and the conversion efficiency. Smaller particles produce more CH_4, CO and less CO_2 than the larger one. The smaller particle causes a rapid raise of particle temperature and made the final temperature more uniform and close to the surrounding gas temperature. On the other hand, the larger particle is slowly heated, when it reaches the surrounding temperature, the local temperature overshooting can occur (Kaewluan, 2009).

The size of the chips used mainly depends on the diameter of the gasifier. Walawender used 4.7 cm chips with a reactor diameter of 60 cm (Walawender, 1985), while Milligan used 0.6 cm chips with a reactor diameter of 7.5 cm (Milligan, 1994). On the other hands, Jayah mentioned the optimum gasification zone length is defined as the length at which the efficiency is 99.5% of the ultimate conversion efficiency, it concluded for a 5 cm chip size the optimum length of the gasification zone is approximately 33 cm (Jayah, 2003).

Table 3.1 Standard Method for Biomass Compositional Analysis

Biomass Constituent	Standard Methods
Ultimate Analysis	
Carbon	ASTM E-777
Hydrogen	ASTM E-777
Nitrogen	ASTM E-778
Oxygen	By difference
Ash	ASTM D-1102
Moisture	ASTM E-871
Proximate Analysis	
Volatile matter	ASTM E-872
Ash	ASTM D-1102
Moisture	ASTM E-871
Fixed carbon	By difference
HHV	ASTM E711-87

3.1.1 Moisture Content Analysis

The oven drying method is the most acceptable method for determining moisture content. Following ASTM standard E-871, the sample was place and heated up to 105-110^{o}C in the oven (Memmert, VO500). The weights of sample after and before drying were used to determine the percentage of moisture content.

Figure 3.2 The Oven (Memmert, VO500)

3.1.2 Proximate Analysis

Consisting of volatile matter, fixed carbon and ash content, the proximate analysis was determined by using a Thermo-Gravimetric Analyzer (PerkinElmer, TGA Pyris1) following ASTM standard procedure. Sample was heated from 30°C to 110°C at 10°C/min, then held the temperature for 10 minute at 110°C. After that, sample was heated again from 110°C to 900°C at 10°C/min, next held it for 20 minutes at 900°C. Volatile matter determines by the different weight between 110°C to 900°C, while the fixed carbon occur during combustion process at 900°C and the remaining weight represent the ash percentage.

Figure 3.3 Thermo-Gravimetric Analyzer (PerkinElmer, TGA Phyris 1)

3.1.3 Ultimate Analysis

The percentage of chemical elements, such as C, H, N, S, O were analysed using the Organic Elemental Analysis (Thermo Finnigan, Flash EA 1112 series). The percentages of chemical elements are important to determine the theoretical air required for complete combustion and the amount of air required for gasification (partial combustion) as well as for determining the higher heating value of biomass.

Figure 3.4 Organic Elemental Analysis (Thermo Finnigan, Flash EA 1112 Series).

3.1.4 Calorific Value Analysis

The ASTM E711-87 standard method was used to determine the sample calorific value. The apparatus used in calorific value analysis is the Bomb Calorimeter (LECO AC-350). In the bomb calorimeter, the sample is placed in a high pressure oxygen environment called a bomb. The bomb is surrounded by water and the sample is ignited. The temperature of the water is measured by an electronic thermometer. During analysis, the stirrer is rotated to control the uniform temperature.

Figure 3.5 The Bomb Calorimeter (LECO AC-350)

3.1.5 Ash Analysis

Ash represents the amount of minerals in a sample. Following the ASTM standard D-1102, a horizontal furnace (Carbolite Furnace) was used. Sample was burned at 850^{o}C for 8 hours. The weights of sample after and before burning were used to determine the percentage of ash.

Figure 3.6 Horizontal Furnace (Carbolite Furnace)

3.2 Gasification System

In order to study the gasification process in a multi-air-stage gasifier, a newly downdraft gasifier was constructed and equipped with purifier equipment, such as cyclone, heat exchanger and bag filter, to make a complete gasification system.

3.2.1 Triple Air Stage Downdraft Gasifier

The downdraft gasifier with a triple stage of air supply was designed and built by the Thai Steam Service & Supply. Co. Ltd (Bangkok, Thailand), having an internal diameter of 0.15 m. The equipment was utilized to maintain wood and other carbonaceous material to produce fuel gas. The complete design calculation can be seen at Appendix C.

The gasifier is built with an internal coating of refractory material. The top section is a hopper which can hold 1.5 hours supply. Near its bottom end, three tubes were located at equal intervals around the perimeter and used for the first stage air supply. At a distance of 0.3 m and 0.6 m above its tube, other three tubes were installed for the secondary and tertiary air supply. At the bottom of the first stage, a gate attached as the ash separator. The cleaning systems consist of cyclone and heat exchanger.

The drying process occurs in the upper part of the gasifier where devolatilization of lighter compounds takes place. Follow by pyrolysis zone, where the organic devolatilization is favored in order to produce charcoal and pyrolytic gases. The purpose of the second and tertiary stage of air supply is to reduce the tar present in the produced gas, as well as to increase the oxidation and reduction phase. The producer gas stream flows by the bottom of the reactor where the mixture of unconverted carbon and ash, promotes its cleaning.

3.2.2 Cyclones

Cyclones are usually designed with a geometric standard of the dimensions ratio based on the body diameter (Dc). The cyclone standard dimension and design calculation can be seen at Appendix D. The whole design process of cyclone can be concluded with several steps, such as calculate the gas flow rate through the cyclone; calculate the body diameter; calculate the dimension refers to cyclone standard dimension; and predict the cyclone pressure drop.

3.2.3 Heat Exchangers

Heat exchangers are used to transfer the thermal energy between two fluids in thermal contact at differing temperatures. The design process of heat exchanger are determine the temperature inlet and outlet of the fluids; calculate air mass flow rate; estimate the size; and rating of shell and tube heat exchanger, as shown in Appendix E.

3.2.4 Bag Filters

Bag filters are used to capture fine dust particles from combustion gases. The design processes of bag filter is as follow, determine the inlet temperature; selected the appropriate filter material; calculate the total cloth area; calculate the amount of fabric required per bag; and calculate the number of bags required in the bag house. The design calculation of bag house filter located at Appendix F.

3.3 Experimental Planning

The controllable variables studied for the assessment of the gasification process are the total air flow and the stages variation between the primary, secondary and tertiary stages. During the tests, the following variables were measured:

3.3.1 Sampling

Sampling was utilized to obtain a sufficient number of data points per run. Temperature will be recorded continuously using Data Logger. This will allow temporal monitoring and recording of the variation in gasifier zone. Producer gas will be collected at the end of the process while tar sampling will be collected at the output line of Heat Exchanger.

3.3.2 Air Flow Controller

Air flow rates used for the gasifying agents were controlled with Flow meter controllers. The flow rates are measured just before the heat exchange.

3.3.3 Gasification Conditions

Reactor temperature	: 1000^{o}C
Raw material	: 14 kg eucalyptus wood/h
Air flow rates	: 6 m^3/h

3.3.4 Gasifier temperature

The temperature measurements were carried out using a K-type thermocouple. The maximum temperature that can be measured by thermocouple is 1200^{o}C with accuracy 0.1^{o}C and 1 second of responsibility. Seven thermocouples were projected along the internal gasifier wall in order to avoid possible problems with the flow of the biomass as it is consumed. One thermocouple for each zone reactors was arranged to register the temperatures at different reactor points.

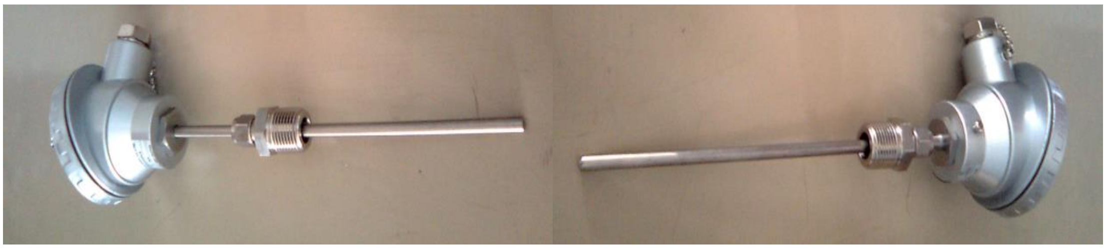


Figure 3.7 Thermocouple Type T and Type K

Figure 3.8 Data Logger (YOKOGAWA, DC 100)

3.3.5 Gas composition measurements

Gas chromatography were used to measure the composition of the producer gas. It is to measure volume percentages of CO, CO_2, H_2 and CH_4. Further, to analyze gas compositions, gas will be sampled through the gas analyzers after the condition of steady state operation is achieved. The concentration is then measured at the end of the process.

Figure 3.9 Gas Chromatograph (GC-14B and GC-2014, SHIMADZU)

3.3.6 Tar Sampling Process

A series of impinger bottles consisting of six impinger bottles were used as the tar sampling equipment after heat exchange. The first impinge bottle acts as a moisture collector, after the moisture collector the gas is passed through a series of four impingers with isopropanol solvent and the final impinger which is empty to trap moisture. Heated and cold baths were used. Cooling liquid made of a mixture of salt, water, and crunch ice. Standard glass impingers, 100 ml or 250 ml volume, with an inner tube diameter of 4 mm are recommended (Good et al, 2005). Tar sampled is fond out the quantities by evaporator in JGSEE laboratory.

Figure 3.10 Tar Sampling (Impinger Bottle)

Figure 3.11 Vacuum Evaporator (BUCHI)

3.4 Experimental Procedures

Firstly, the biomass was dried by solar drying and analyzed by ultimate and proximate analyses. Secondly, the 14 kg biomass is weighted and loaded to hopper at line 1. After being ignited with diesel oil, air is flowed through pipe 5, which is pre-heated from the exchange of heat with the producer gas at heat exchanger and the temperature of air measured with thermocouple (T9). Air flow is separated in 3 lines which is pipe 7, 9 and 11 according to the experimental planning, each pipe has a valve (V-2, V-3, V-4) and Flowmeter (F-2, F-3, F-4) to control and measure the air flow. Air is blown by blower (B-1) at 2, then controlled by a valve (V-1) and measured by a flow meter (F1) before flowing into a heat exchanger (HE-1). As long as the experiment is in progress, the temperature along the reactor wall for each zone will be measured with some thermocouples T1, T2, T3, T4, T5, T6 and T7.

The producer gas exited from the reactor through pipe 13 purified by the cyclone while the ash was removed from line 12. The producer gas will be separated from impurities by using centrifugal force at cyclone (C-1), then exits at 15 while the impurities discharge at 14. The temperature of producer gas is measured using thermocouple (T8) between cyclone and heat exchanger (HE-1). Heat exchange between producer gas and air occur at heat exchanger. Producer gas comes in at 17, then out at 21, while air flows in at 4 to out at 5. Tar sampling occurs at 20. The producer gas passed the bag filter (BF-1), clean gas flowed at 22. The clean producer gas sucked out by blower (B-2) at 23. The sampling for the analyzing process took part at 24. It ended by burning the producer gas at flare tower (FT-1). The whole procedures are shown in Figures 3.12 and 3.13. Gasifier efficiency can be calculated based on Equations 1 and 2 in Section 2.6 (p.16).

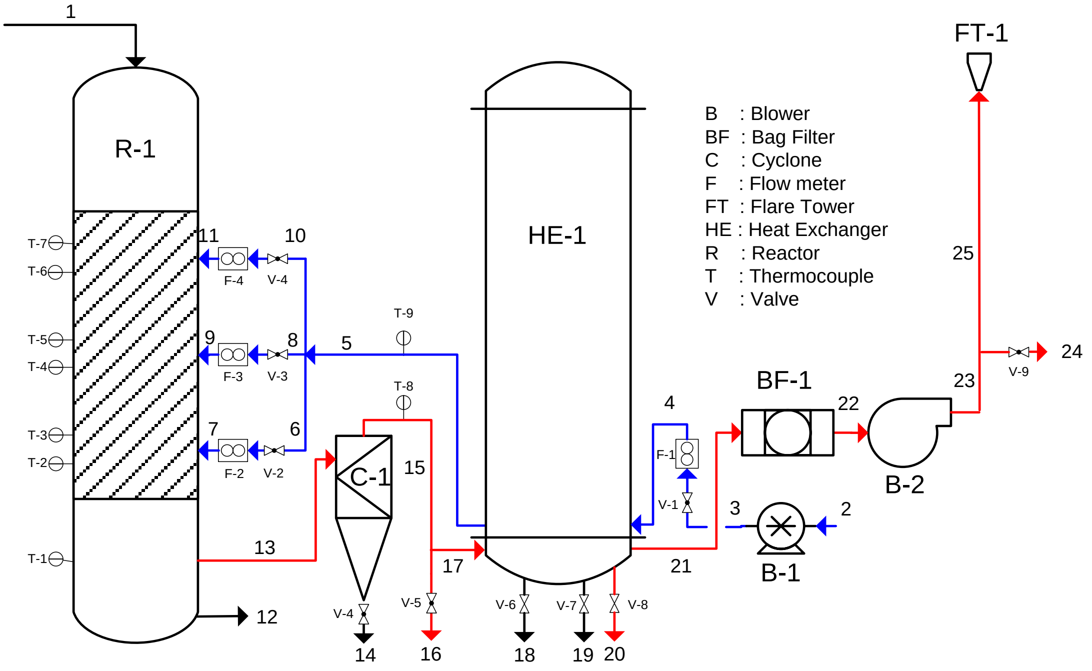


Figure 3.12 Schematic Diagram of a 50 kWth Downdraft Gasification

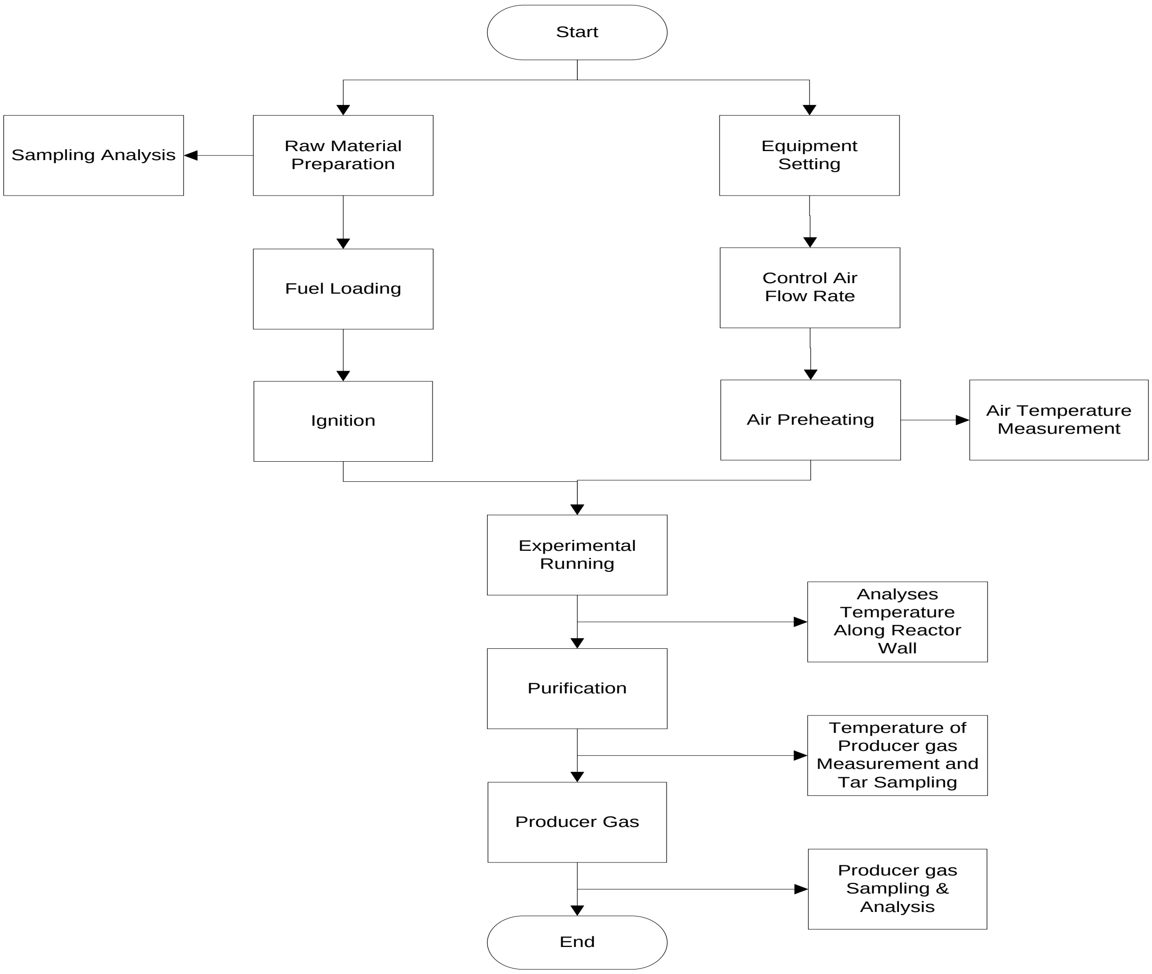


Figure 3.13 Experimental Procedure Diagram

CHAPTER 4
RESULTS AND DISCUSSION

4.1 Raw Material Properties

Raw materials selected for this study are Eucalyptus Wood Chips, Eucalyptus Wood Block and Eucalyptus Wood Pellets. The properties of these materials are shown in Table 4.1.

Table 4.1 Raw Material Properties (Dry Basis)

Composition	Wood Chips	Wood Block	Wood Pellet
Ultimate Analysis (wt%)			
Carbon	45.91	45.18	37.78
Hydrogen	5.74	5.38	4.73
Nitrogen	0.44	1.80	0.51
Oxygen	47.35	46.98	32.95
Proximate Analysis (wt%)			
Ash	0.56	0.66	24.03
Volatile	78.84	83.21	68.42
Fixed Carbon	20.60	16.13	7.55
HHV (kcal/kg)	4221.84	4195.32	3443.47

4.2 Multi Air--Stage Gasification System Equipment

4.2.1 Gasifier

The gasifier was constructed by Thai Steam. Co. Ltd. It has 150 mm internal diameter with 100 mm refractory wall thickness consisting ceramic and castable then wrapping by paper tape. The total height of gasifier was 2100 mm and separated into 3 zones. 600 mm for hopper, 1200 mm for gasification zone and 300 mm for ash chamber. The gasification zone separated into three stages air injection, 300 mm length for each stage. The hopper capacity is 24.383 kg., which can distribute 2.4 kg biomass to gasification zone for second.

The gasification zone has 9 nozzles for air injection, with 3 nozzles for each stage. There are an odd number of nozzles for each stage, so the hot air from one nozzle does not impinge with another nozzle in the opposite direction (Reed and Das, 1988; Knoef, 2005). The ignition holes are added at the oxidation and pyrolysis zones.

The gasifier was equipped with ladder and service platform to help researchers to load the fuel into the hopper. The detail design showed at Appendix C. The gasifier also has 7 holes for installing the thermocouple, two thermocouples for each stage and the rest for the reduction zone. The thermocouples are type K due to its endurance on high temperature until 1200 ^{o}C.

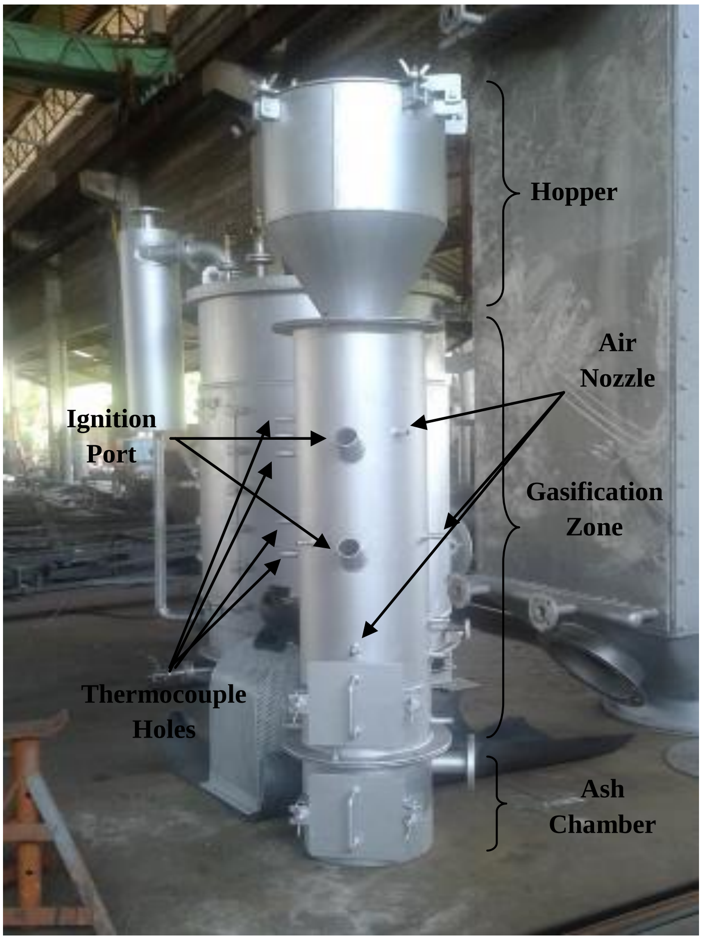


Figure 4.1 Three Air-Stage Biomass Gasifiers

Figure 4.2 The Ladder and Service Platform

4.2.2 Cyclone

Based on the calculations, a 50 kW thermal gasifier needed 36 Nm^3/h of producer gas and correlated with a volume 128.26 m^3/h. The inlet gas temperature at cyclone should be the same as the gasifier output temperature, so it determined at 700^oC. The calculated body diameter is 0.127 m. There are two different ways to design cyclone, first the cyclone inlet pipe width (B_c) should be equal to the gasifier output pipe diameter, or set the cyclone inlet velocity equal to the pipe velocity. The cyclone design dimension is shown in Figure 4.3.

The recommended minimum gas velocity for conveying medium density dust is 15 m/s, and for heavy dust (metal turning) is 25 m/s (Reed and Das, 1988). A 2 inch inside diameter pipe, will provide 17.57 m/s which is well above the minimum gas velocity. The pressure drop is 150.78 Pa and the particle cut point is 2.8 µm. The design calculation can be seen in Appendix D.

Figure 4.3 The Cyclone

4.2.3 Heat Exchanger

The inlet temperature was estimated to be 700^{o}C which was nearly the same as that of the cyclone output stream. Because almost all the tars start to condense around 250^{o}C, therefore the producer gas outlet temperature defined as 300^{o}C. In order to prevent the heat lost from a heat exchanger and to increase the conversion factor of gasification process, the heat lost from heat exchanger was used to preheat the air as gasifying medium. The highest temperature possible for outlet air temperature is 350^{o}C. The ambient temperature was assumed as the input of air temperature.

The shell and tube heat exchanger was designed as a one-shell and two-tube pass, because two passes provide slightly better heat transfer than does a single pass. Shell is designed as a tube with 360 mm inside diameter, 460 mm outside diameter and 2000 mm length. The inside and outside diameter of tube are 0.87 inch and 1 inch. The total used tube is 86 tubes which separated into 43 for each pass. Tube layout is designed as triangular pitch. Baffle spacing is designed as 3.05 inch with baffle cut 25%. Pitch size is 1.25 inch. The tube length estimation is 7 ft which provided a proper pressure drop for both shell and tube. The design calculation for the heat exchanger is shown in Appendix E.

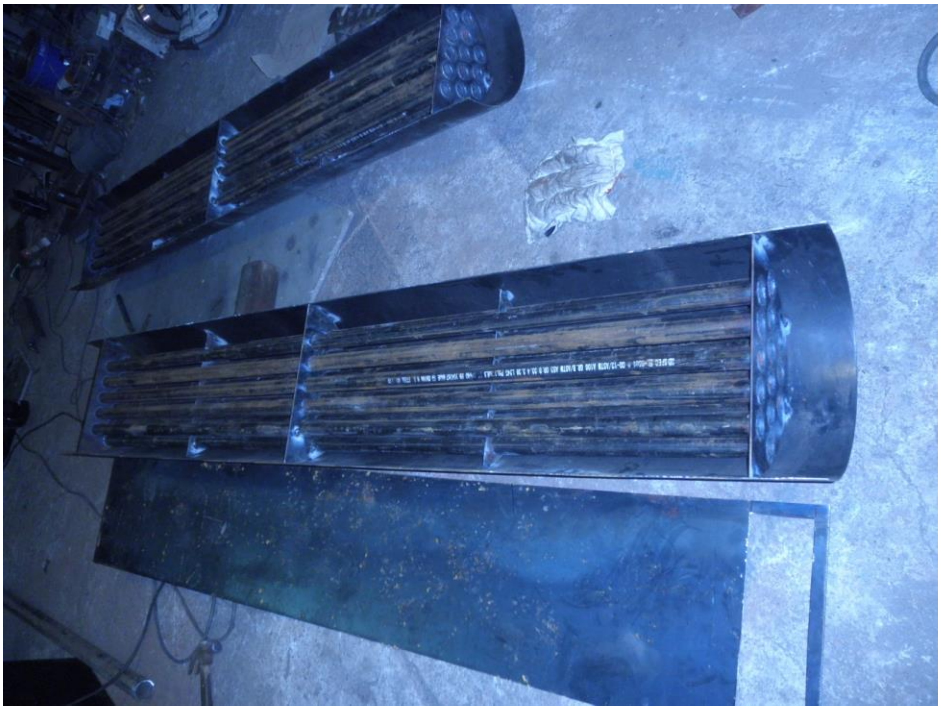


Figure 4.4 The Heat Exchanger

4.2.4 Baghouse Filter

The fiber glass was selected as filter material due to its resistance at high temperature. The inlet temperature determined as 350°C same as the outlet temperature of heat exchanger. The bag diameter and height first defined as 15 and 50 cm respectively. The number of bags is determined to be 4 bags. The design of the bag house filter is shown in Appendix F.

Figure 4.5 The Bag House Filter

4.2.5 Blower

Air gasifiers can operate in two ways: by forcing air through the fuel (pressurized) and by drawing the air through the fuel (suction). However, the suction gasifiers used widely rather than pressurized gasifier due to the suction of the engine generally use to move air through the gasifier and clean up train (Reed and Das, 1988). The 1Hp induced draft blower used to pull the producer gas through the system at negative pressure. The design calculation can be seen in Appendix G.

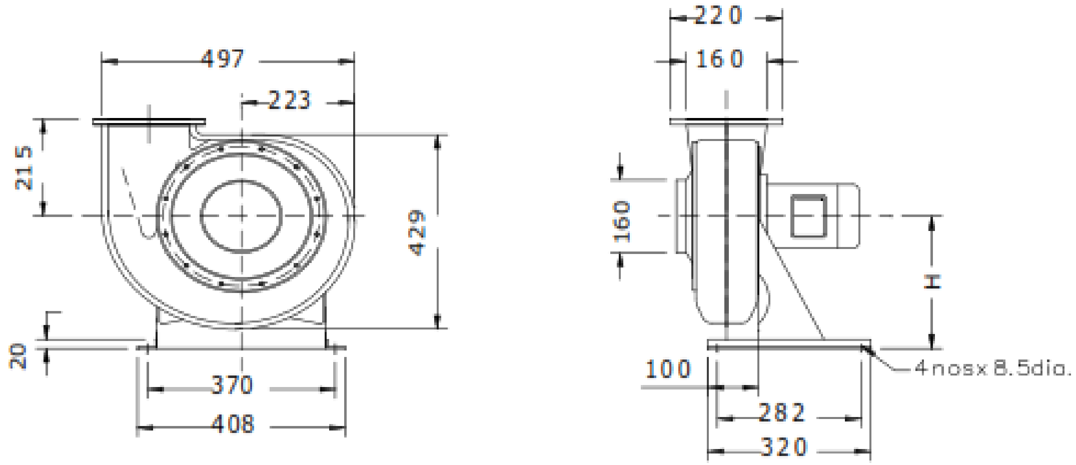


Figure 4.6 The Blower Design

4.2.6 Flare Tower

At the end of the process, the fuel gas will burn to reduce the toxic gas. A 2- meter length flare tower is set at the end of the final equipment. This tower has an automatic ignition port.

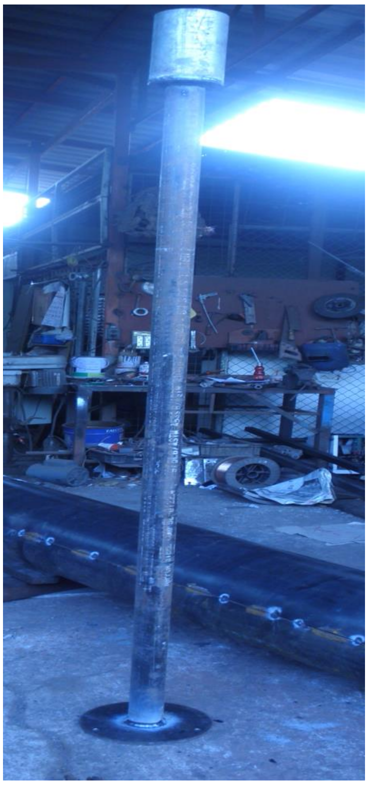


Figure 4.7 Flare Tower

4.3 The Construction Process

The construction process was started in July 2013 and completed in February 2014. It took four months to construct the equipment and another four months to prepare the site. Meanwhile, several modifications were conducted to adjust the operating condition based on the experimental planning.

Figure 4.8 The Concrete

Figure 4.9 The Workshop Installation

Figure 4.10 The Equipment Installation

Figure 4.11 The Roof and Electricity Installation

Figure 4.12 The Piping Installation

Figure 4.13 The Control Room Installation

Figure 4.14 The Insulated Piping System and Cyclone

Figure 4.15 The Modified Hopper

Figure 4.16 Adjustable Blower

4.4 The Experimental Results

4.4.1 Temperature Distribution on Multi-Air-Stage Gasification

The temperature distribution along the gasifier wall can indicate the zone of the reactions, which is important for indicating the position and length of each reaction zone, such as drying, pyrolysis, oxidation and reduction. The temperatures of gasifier were collected by using data logger through experimental period from beginning till the end of experiment. The experimental result of temperature distribution on various gasifier air-stage are shown in Figure 4.17.

The results show that temperature fluctuated during the experiment, while the gasifier took 20 minutes to reach an almost stable condition. The most noticeable temperature fluctuation happened on T1, T2, T3 and T8. It was on the first stage oxidation zone and output product. As it happened on the oxidation zone, then it can assume as the effect of the bridging fuels. However, the temperature fluctuation has already studied (Zainal, 2002; Nipattummakul, 2006; Kerdsuwan, 2010). It concluded that the bridging happened due to the raw materials rough surface, so that it would not move down smoothly.

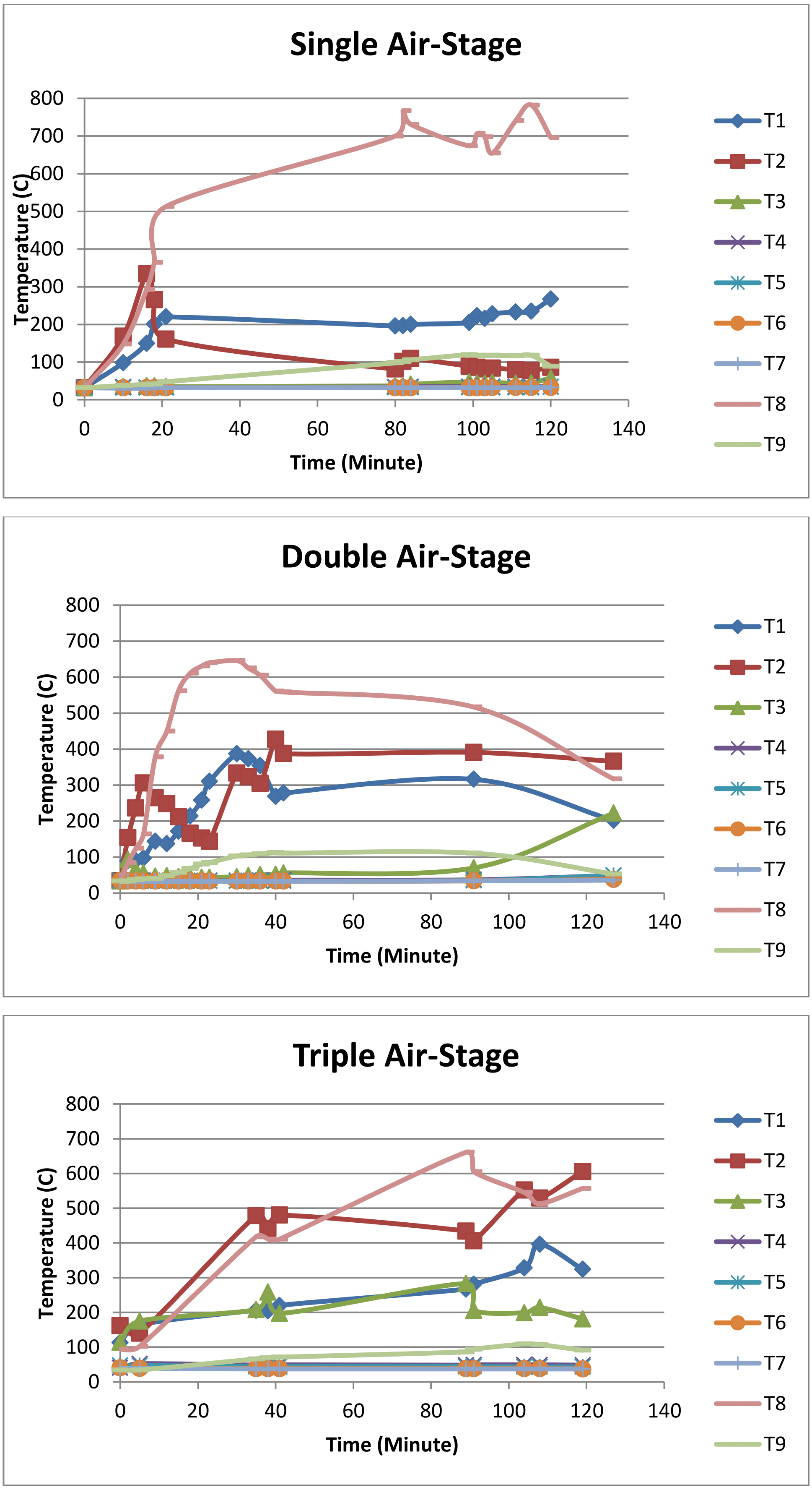


Figure 4.17 Temperature Distribution on Wood Chip gasification

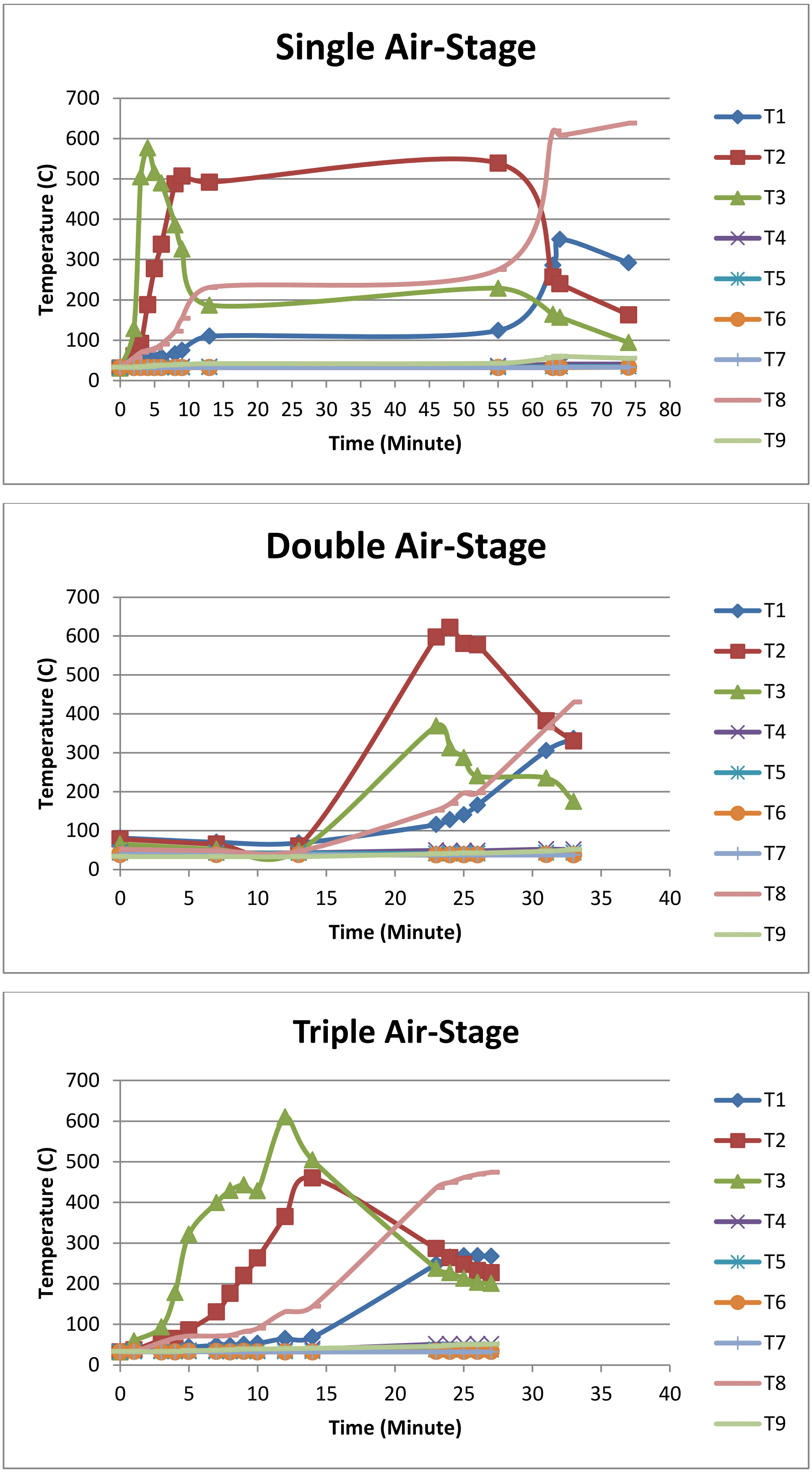


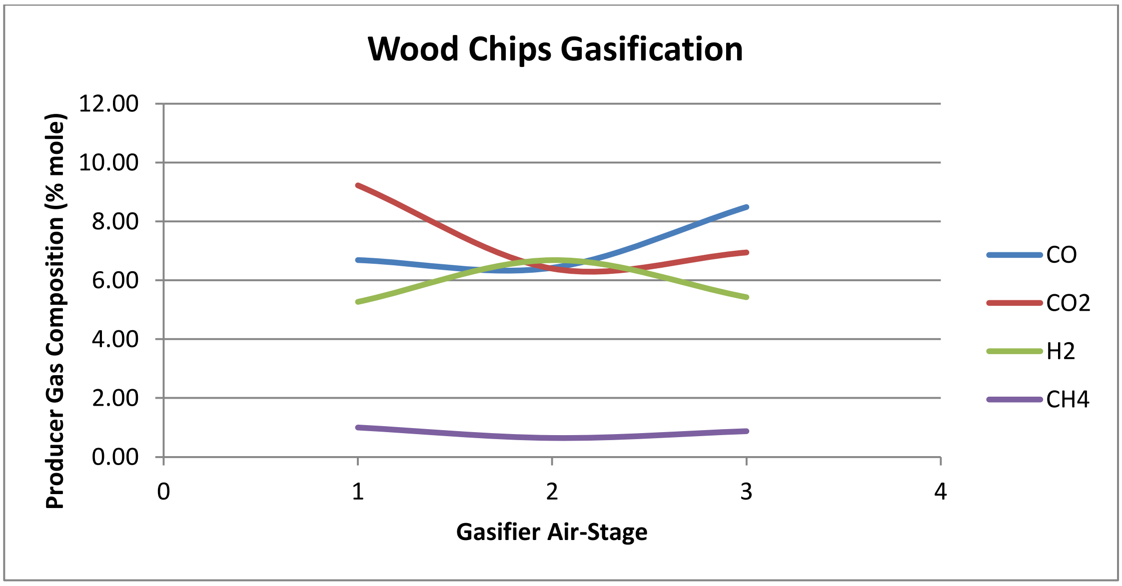
Figure 4.18 Temperature Distribution of Wood Chip and Wood Pellet Gasification

During the bridging, a large cavity formed in the oxidation zone affected by the temperature rise. As the experiment went on, the bed was consumed instead of the fuel above the oxidation zone. At this state, it became a direct combustion not gasification anymore. The small fuel will flow through the cyclone and burn leads the temperature increase at cyclone and creates fire.

Wood chips and wood blocks cause the bridging, while wood pellet remained a sticky ash. A mixing raw material between wood pellets as the bed and wood chip as the fuel was made. Figure 4.18 shows that the T8 remained low at the beginning while the wood chips were burning. The fire began to consume the wood pellets and the cycle began just as in the wood chips gasification.

4.4.2 Producer Gas Composition

Dividing the air-stages into three levels can help to increase the oxidation zone. Meanwhile, setting the air nozzle on the several different angles can help to distribute air evenly influence the temperature distribution along the gasifier. Furthermore, the temperature can affect the composition of producer gas. H_2 and CO concentration will increase as well as the temperature due to the effect of carbon conversion efficiency. A high amount of carbon was converted to H_2 and CO through Boudouard and Water Gas reaction. Meanwhile, CH_4 and CO also can be converted to H_2 through Methane Steam Reforming and Water Gas Shift reaction. However, the maximum value for wood gasification is CO : 11-24 volume%, while H_2 : 11-14 volume % (Zainal et al., 2002)



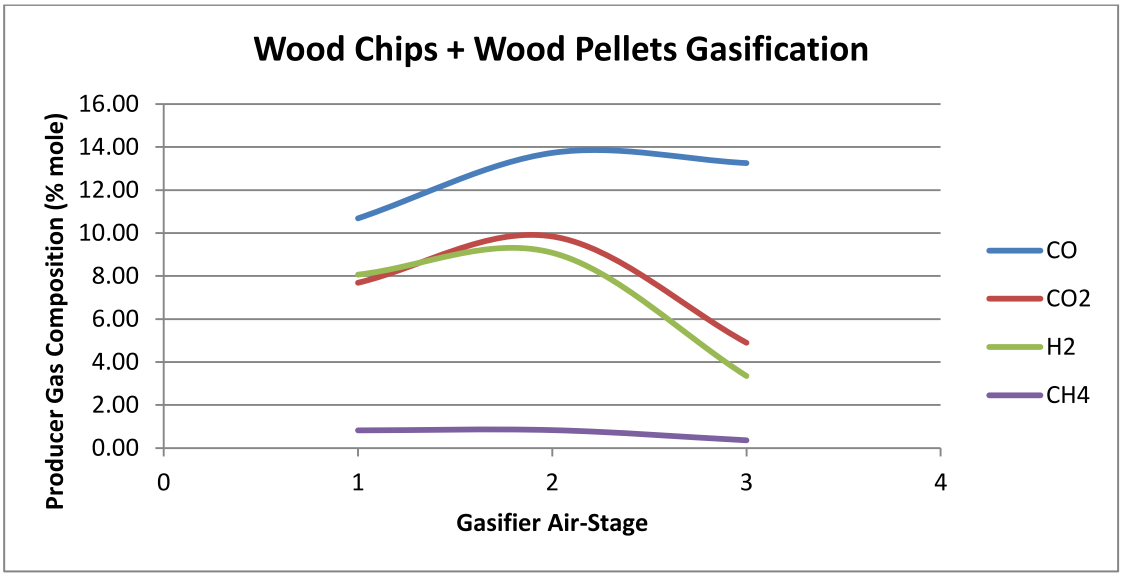


Figure 4.19 Producer Gas Composition on Wood Chips and Wood Chips + Wood Pellets Gasification

4.4.3 Gas Conversion Efficiency

The performance of the gasification system can be shown by the cold and hot gas efficiencies. As shown in Figure 4.20, the efficiency was increase along with the air-stages. The raw material properties indeed affect the producer gas properties as well as the gas efficiency, but due to the rapid increase temperature at cyclone, the gasification process should turn-off immediately. It causes some of the biomass unburn, while the conversion efficiency calculation still base on the mass input basis. At the end it made the carbon conversion became low. The maximum efficiency conditions of wood gasification are 66.7±6.7% for cold gas efficiency and 72.7±6.9% for hot gas efficiency (Patil et al., 2011). However, this experiment resulted 11.81% and 25.36% for cold and hot gas efficiency on wood chip gasification, 17.35% and 31.11% for cold and hot gas efficiency on wood chip + wood pellet mixture gasification.

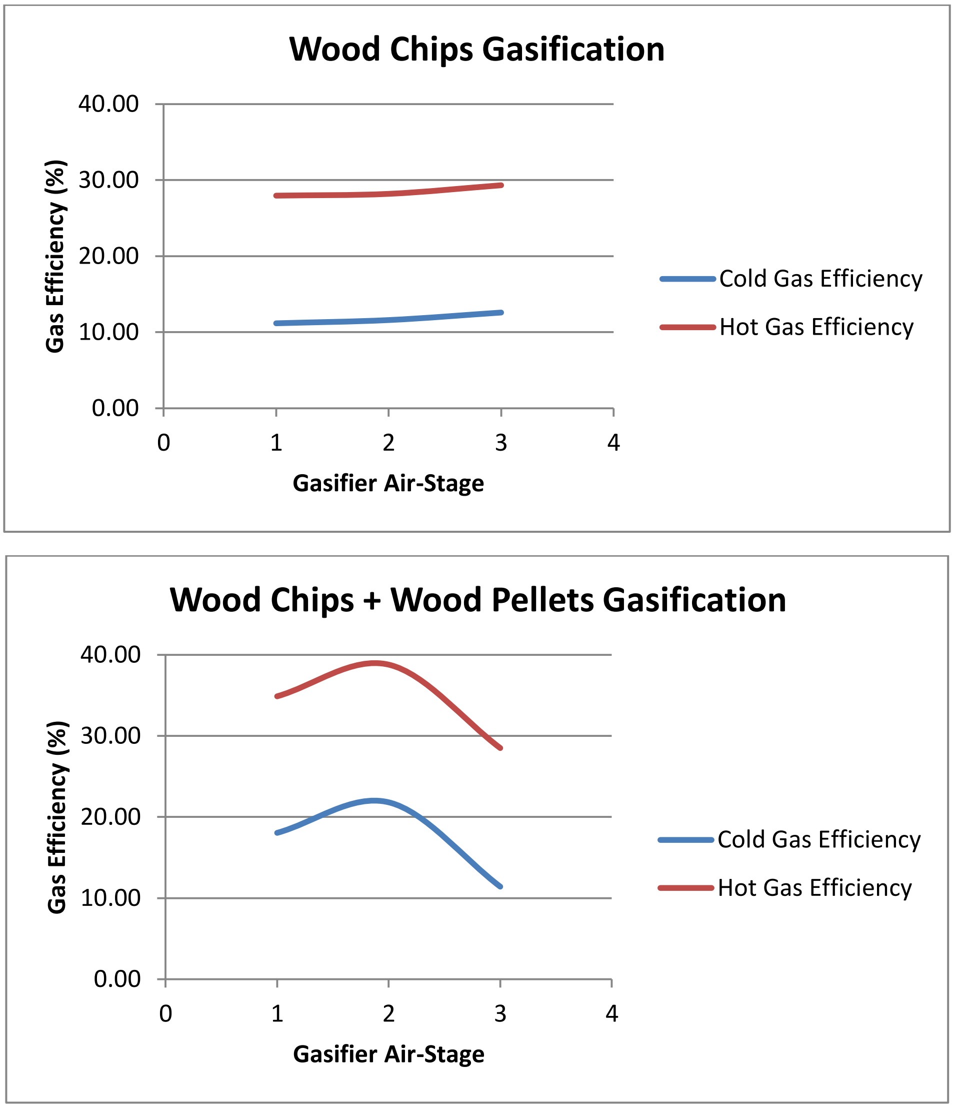


Figure 4.20 Gas Conversion Efficiency on Wood Chips and Wood Chips + Wood Pellets Gasification

4.4.4 Tar Content

Dogru et al., (2002) reported that, the non-throat downdraft gasifier with the air gasifying medium has a higher tar concentration than does the throat reactor. The reason is because the oxidation reaction in the throat section of the gasifier, is more efficient than the non-throat case due to good mixing and the high velocity of the pyrolysis gas with the air agent. However, the multi air-stage was invented to expand the oxidation zone on gasifier. Even though, it was non-throat downdraft gasifier, it has wider oxidation zone than conventional throat gasifier. Jayah (2002), stated that the optimum length for gasification zone on conventional gasifier was 30 cm, while the multi air-stage gasifier has tripled it.

The tar sample was sampled at the end of the heat exchanger while the temperature was still hot, but not too extremely hot to harm the tar sampling unit according to the guidelines given by Good et al., (2005).

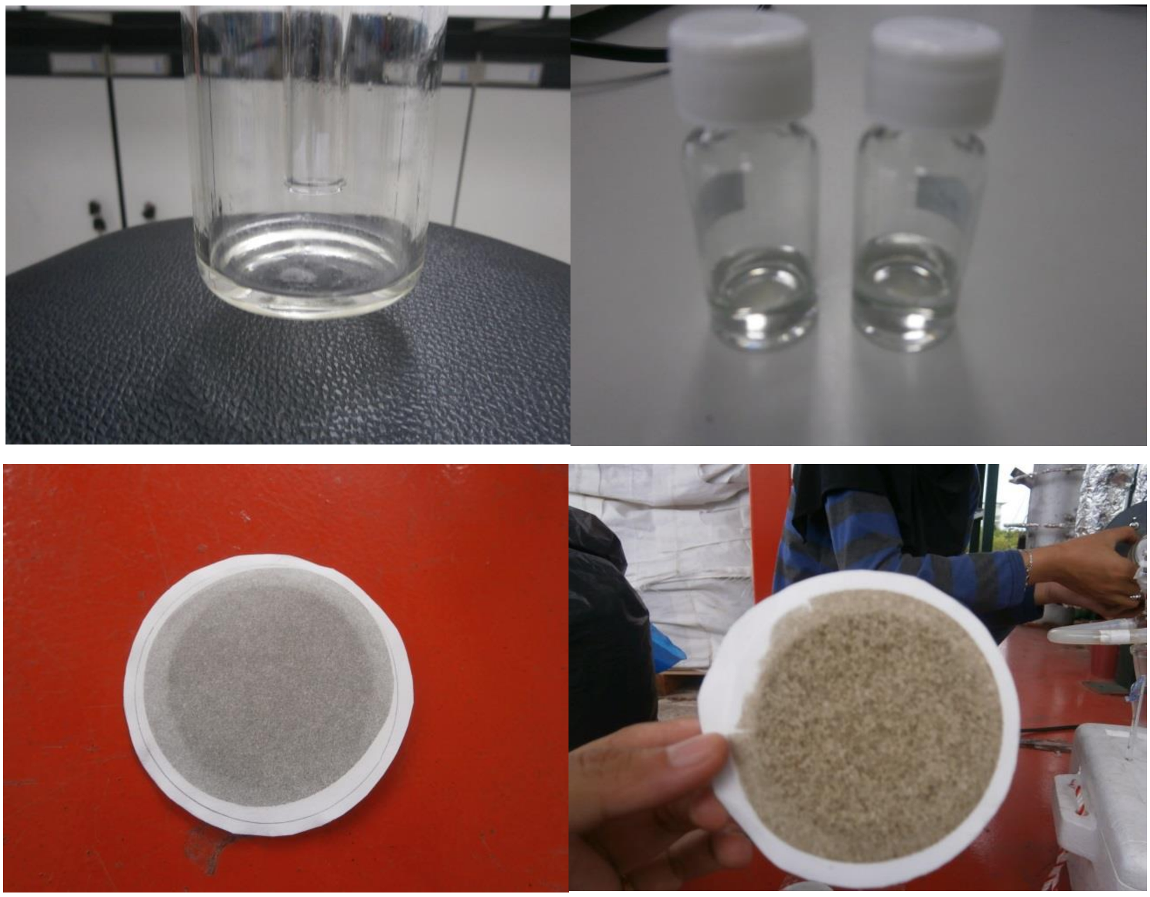


Figure 4.21 Tar Samples

There was no major difference in the tar sample between wood chips and wood pellet-wood chip mixture. It was as clear as the isopropanol, without a single drop of tar being found. However, the filter showed the different result on both raw materials. The wood pellet-wood chip mixture had a darker color due to the high ash content in the wood pellets.

4.5 Design and Actual Data Comparison

Table 4.2 Design and Actual Data Comparison

Parameter	Design	Actual
Power Output	50 kWth	17.44 kWth
Product Gas Volume Flow Rate	0.036 m^3/s	0.037 m^3/s
Biomass Feed Rate	14.18 kg/h	15 kg/h
LHV of Producer Gas	2.119 MJ/kg	1.732 MJ/kg
LHV of Biomass	18.14 MJ/kg	16.505 MJ/kg
Air Feed Rate	450 L/min	100 L/min
Equivalence Ratio	0.4	0.14
Gasifier Efficiency	70 %	25.36 %

The design was calculated to generate 50 kWth, but the actual data showed that it only generated 17.44 kWth. There are some influencing factors, such as the lack of air gasifying media and un-burnt biomass. The lack of air created the low temperature (600^{o}C) for gasification, it can not reach the optimum gasification temperature, which is 1000^{o}C. Un-burn biomass reduce the efficiency of gasifier until 25.36%, because efficiency was calculated based on the amount of biomass feed rate. Unfortunately, due to safety reasons, the gasification process should turn off before all of the biomass was burnt. At the end, it left some biomass un-burnt inside the gasifier.

CHAPTER 5

CONCLUSIONS AND RECOMMENDATIONS FOR FUTURE STUDIES

5.1 Conclusions

1. Wood chips and wood blocks have better properties than do wood pellets, while their shape causes the bridging inside the gasifier.
2. The Gasifier has 150 mm internal diameter with 100 mm refractory wall thickness. The total height of gasifier was 2100 mm and separated into 3 zones. 600 mm for hopper, 1200 mm for gasification zone and 300 mm for ash chamber. The gasification zone was separated into three stages of air injection, 300 mm for each stage.
3. The cyclone diameter (D_c) was 0.127 m, with 0.0637 m inlet height (H_c) and 0.0318 m inlet width (B_c). The other dimensions were calculated based on the high efficiency cyclone performance template as follows:

   $$D_e = \frac{Dc}{2} = 0.0637\ m$$

   $$L_c = 2Dc = 0.2547\ m$$

   $$S_c = \frac{Dc}{8} = 0.0159\ m$$

   $$Z_c = 2Dc = 0.2547\ m$$

   $$J_c = \frac{Dc}{4} = 0.0318\ m$$

4. A 1-2 shell and tube heat exchanger was designed with 2 m length. The shell’s outside and inside diameter were 0.46 m and 0.36 m. The tube pipe had 0.025 m outside diameter and 0.022 m inside diameter, and used 86 tubes.
5. Fiber Glass was selected as the filter material. The bag’s diameter and height were designed at 15 and 50 cm, and 4 bags were required.
6. The multi-air-stage gasification system developed for this study had followed the right patterns in terms of temperature distribution, producer gas composition and gasifier efficiency.
7. The gasification test using wood chips and a mixture of wood chips and wood pellets showed average hot gas and cold gas efficiencies of 12.59% and 27.96% for wood chips and 18.05% and 34.88% for wood chips and wood pellets mixtures.

5.2 Recommendations for Future Studies

A new modification should be made in order to increase the gasifiers performance. An ash shaking unit would resolve the bridging case. Manometers should be installed in between the equipment to measure the pressure drop in order to fulfil the appropriate blower capacity as the air supply source. Another thermocouple may install after heat exchanger to ensure that the temperature of the tar sampling point is higher than that of the tar condensation point.

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APPENDIX A

Table A1. Typical Gas Properties of Various Wood-Fuelled Gasifiers at Atmospheric Pressure

Compound	Symbol	Gas (vol%)	Dry Gas (vol%)
Typical Gas Composition			
Carbon monoxide	CO	21.0	22.1
Carbon dioxide	CO_2	9.7	10.2
Hydrogen	H_2	14.5	15.2
Water (v)	H_2O	4.8	-
Methane	CH_4	1.6	1.7
Nitrogen	N_2	48.4	50.8
Typical Heat Output			
Generator gas (wet basis)		5506 kJ/Nm^3	
Generator gas (dry basis)		5800 kJ/Nm^3	
Air Ratio Required for Gasification		2.38 kg wood/kg air	
Air Ratio Required for Gas Combustion		1.15 kg wood/kg air	
Typical Contaminant Level			
Particles		0.1 – 1 g/m^3	
Tar (generic definition)		0.5 – 2 g /m^3	
Typical Tar Composition (wt %)			
Benzene		37.9	
Toluene		14.3	
Other 1-ring aromatic hydrocarbons		13.9	
Naphthalene		9.6	
Other 2-ring aromatic hydrocarbons		7.8	
3-ring aromatic hydrocarbons		3.6	
4-ring aromatic hydrocarbons		0.8	
Phenolic compounds		4.6	
Heterocyclic compounds		6.5	
Others		1.0	

Source : Reed, 1988 and Milne et al., 1998.

APPENDIX B

BIOMASS AND AIR FEED RATES

1. Producer gas flow rate (V_g)

$$V_g = \frac{Q}{LHV_g} \, Nm^3/s \qquad \text{B.1}$$

Where V_g = Volume flow rate of the product gas (Nm^3/s)

Q = Gasifier required power output (MWth)

LHVg = Lower heating value of the product gas (MJ/Nm^3)

For an air-blow gasifier, LHVg can start with a value of 5 MJ/Nm^3 (Quaak et al., 1999).

Q = 50 kWth

= 0.05 MWth

= 0.05 MJ/s

$$V_g = \frac{Q}{LHV_g} \, Nm^3/s$$

$$Vg = \frac{0.05 \, MJ/s}{5 \, MJ/Nm^3}$$

V_g = 0.01 Nm^3/s

= 0.036 m^3/s at 1 atm and 700°C

2. Fuel feed rate (M_f)

$$M_f = \frac{Q}{LHV_{bm}\eta_{gef}} \qquad \text{B.2}$$

Where M_f = Biomass feed rate (kg/s)

Q = Gasifier required power output (MWth)

LHV_{bm} = Lower heating value of biomass (MJ/kg)

η_{gef} = Gasifier efficiency

LHV_{bm} = 18140 kJ/kg (Martinez, 2011)

= 18.14 MJ/kg

η_{gef} = 70 % (Ciferno and Marano, 2002)

= 0.7

$$M_f = \frac{Q}{LHV_{bm}\eta_{gef}}$$

$$M_f = \frac{0.05\ MJ/s}{18.14\frac{MJ}{kg}x\ 0.7}$$

M_f = 0.0039 kg/s

= 14.18 kg/h

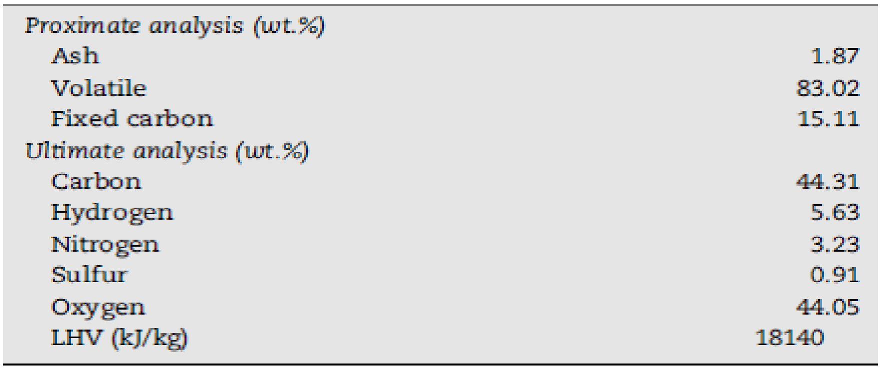
3. Flow rates of gasifying mediums (Ma)

Table B.1 Oxygen Calculations

Element	wt	kmol	Rx	O_2 Required
C	44.31	3.69	$C + O_2 \rightarrow CO_2$	3.69
H	5.63	2.82	$H_2 + ½\ O_2 \rightarrow H_2O$	1.41
O	44.05	2.75	$O_2 \rightarrow O_2$	-1.38
N	3.23	0.23	$½\ N_2 + O_2 \rightarrow NO_2$	0.23
		Total		3.95

Stoichiometric O_2 required is 3.95 kmol

Table B.2 Proximate and Ultimate Analyses of Eucalyptus sp



Proximate analysis (wt.%)	
Ash	1.87
Volatile	83.02
Fixed carbon	15.11
Ultimate analysis (wt.%)	
Carbon	44.31
Hydrogen	5.63
Nitrogen	3.23
Sulfur	0.91
Oxygen	44.05
LHV (kJ/kg)	18140

Stoichiometric air required :

$m_{dry\ air}$ = wt O_2 + wt N_2 B.3

= (3.95 kmol x 32 kg O_2/kmol) + (3.95 kmol x 3.76 x 28 kg N_2/kmol)

= 126.4 kg O_2 + 415.86 kg N_2

= 543.04 kg air / 100 kg biomass$_{(dry\ basis)}$

= 5.43 kg air / kg biomass$_{(dry\ basis)}$

ER = 0.25 (Reed and Das, 1988)

$$ER = \frac{(wt\ of\ air/wt\ of\ fuel)_{actual}}{(wt\ of\ air/wt\ of\ fuel)_{stoiciometri}} \quad \text{B.4}$$

$$0.25 = \frac{(wt\ of\ air/14.18\ kg/h)}{5.43\ kg\ air/kg\ biomass}$$

wt of air $_{(actual)}$ = 19.24 kg/h dry air

$m_{moisture}$ = $m_{dry\ air}$ x ω B.5

Where $m_{moisture}$ = Mass of moisture (kg/h)

$m_{dry\ air}$ = Weight of air$_{(actual)}$ (kg/h)

ω = Humidity (kg/kg)

= 0.016 kg/kg for Thailand

$m_{moisture}$ = $m_{dry\ air}$ x ω

= 19.24 kg/h x 0.016 kg/kg

= 0.31 kg/h

m_{air} = $m_{dry\ air}$ + $m_{moisture}$ B.6

= 19.24 kg/h + 0.31 kg/h

= 19.55 kg/h

V air = m_{air} / ρ air B.7

Where V air = Volume of air (m^3/h)

m_{air} = mass of air (kg/h)

ρ_{air} = Air density (kg/m^3)

= 1.157 kg/m^3 at 30°C

V air = m_{air} / ρ air

= 19.55 kg/h / 1.157kg/m^3

= 16.90 m^3/h

Table B.3 Equivalence Ratios

ER	Mass of Air kg/h	Volume of Air m^3/h
0.25	19.55	16.90
0.3	23.46	20.28
0.35	27.37	23.66
0.4	31.28	27.04

$$Q = v.A \qquad \text{B.8}$$

Where Q : Air Flow Rate (m^3/s)

v : Air velocity (m/s)

A : Surface Area of nozzle (m^2)

Q = 16.90 m^3/h

= 0.0047 m^3/s

A = ¼ .π. D^2

= ¼. 3.14. $(0.0095\ m)^2$

= 7.128×10^{-5} m^2

v = Q/A

= 0.0047 m^3/s / 7.128×10^{-5} m^2

= 65.852 m/s

For three nozzles at the first stage, the velocity becomes :

V_n = v / amount of nozzle B.9

= 65.852 m/s / 3 nozzle

= 21.951 m/s for each nozzle

Table B.4 Air Velocity

Stage	**Amount of Nozzle**	**Air Velocity (m/s)**			
		ER = 0.25	**ER = 0.3**	**ER = 0.35**	**ER = 0.4**
1	**3**	21.951	26.341	30.731	35.121
2	**6**	10.975	13.170	15.365	17.561
3	**9**	7.317	8.780	10.244	11.707

APPENDIX C
GASIFIER DESIGN

I. HOPPER

1. Hopper Volume (V)

$$V_H = V_C + V_T$$

Where V_H = Hopper Volume (m^3)
V_C = Cone Volume (m^3)
V_T = Tube Volume (m^3)

$$V_C = \frac{1}{3}\pi . t(R^2 + R.r + r^2) \tag*{C.1}$$

Where V_C = Cone Volume (m^3)
T = Height, 0.285 m
R = Upper Radius, 0.24 m
r = Bottom Radius, 0.075 m

$$\begin{aligned} V_C &= \frac{1}{3}\pi . t(R^2 + R.r + r^2) \\ &= \frac{1}{3}\ x\ \frac{22}{7} x\ 0.285\ m\ x\ (0.24\ m^2 +\ 0.24\ m\ x\ 0.075\ m +\ 0.075\ m^2) \\ &= 0.024\ m^3 \end{aligned}$$

$$V_T = \pi . R^2 . t \tag*{C.2}$$

Where V_T = Tube Volume (m^3)
R = Tube Radius, 0.24 m
t = Tube Height, 0.315 m

$$\begin{aligned} V_T &= \pi . R^2 . t \\ &= \frac{22}{7}\ x\ 0.24\ m^2 x\ 0.315\ m \\ &= 0.057\ m^3 \end{aligned}$$

$$V_H = V_C + V_T \tag{C.3}$$

$$= 0.024\ m^3 + 0.057 m^3$$

$$= 0.081\ m^3$$

$$m = \rho\ x\ V_H \tag{C.4}$$

Where m = Hopper Capacity (kg)

ρ = Biomass density, 300 kg/m^3

V$_H$ = Hopper Volume (m^3)

$$m = \rho\ x\ V_H$$

$$= 300\ \frac{kg}{m^3}\ x\ 0.081\ m^3$$

$$= 24.383\ kg$$

2. Mass Flow Rate (m)

$$m = \rho . A \sqrt{\frac{B.g}{2(1+C)\tan\theta}} \tag{C.5}$$

Where m = Mass Flow Rate (m)

ρ = Biomass Density, 300 kg/m^3

g = Acceleration due to gravity, 9.81 m/s^2

θ = Semi-included angle of the hopper, 60^o

B = Outlet Diameter, 0.15 m

A = Surface Area, 0.018 m^2

Table C.1 Hopper Parameter Design

Parameter	Conical Outlet	Symmetric Slot
B	Outlet diameter, D	Slot width, W
A	π/4. D^2	Width x Breadth
C	1.0	0

$$m = \rho.A\sqrt{\frac{B.g}{2(1+C)\tan\theta}}$$

$$= 300\frac{kg}{m^2}\,x\;0.018\,m^2\sqrt{\frac{0.15\,m\,x\,9.81\,\frac{m}{s^2}}{2(1+1.0)\tan 60^o}}$$

$$= 2.444\,kg/s$$

3. Hopper Pressure (Po)

$$P_o = \frac{\rho.D.g}{4.k_f.K}\left(1 - exp\left(-\frac{4.H.k_f.K}{D}\right)\right) \qquad \text{C.6}$$

Where P_o = Hopper vertical pressure (Pa)

ρ = Biomass Density, 300 kg/m^3

D = Diameter, 0.48 m

g = Acceleration due to gravity, 9.81 m/s^2

k_f = Wall solid friction coefficient, 0.37

K = Janssen coefficient, 0.4

H = Height, 0.6 m

$$P_o = \frac{\rho.D.g}{4.k_f.K}\left(1 - exp\left(-\frac{4.H.k_f.K}{D}\right)\right)$$

$$= \frac{300\frac{kg}{m^3}\,x\;0.48\,m\,x\;9.81\frac{m}{s^2}}{4\,x\;0.37\,x\;0.4}\left(1 - exp\left(-\frac{4\,x\;0.6\,m\,x\;0.37\,x\;0.4}{0.48\,m}\right)\right)$$

$$= 1247.719\,Pa$$

Since the lateral pressure (P_w) is proportional to the vertical pressure (P_o), then

$$P_w = KP_o$$

$$= 0.4\,x\;1247.719\,Pa$$

$$= 499.088\,Pa$$

II. GASIFICATION ZONE

1. Gasifier Temperature

Thermal gasification of ligno-cellulosic biomass prefers a minimum gasification temperature in the range 800 – 900^{o}C. The exit gas temperature of a downdraft gasifier is about 700^{o}C. Peak gasifier temperature of a downdraft biomass gasifier at the throat is 1000^{o}C (Knoef, 2005).

2. Specific Gasification Rate (SGR)

$$D = \left(\frac{1.27\ x\ M_f}{SGR}\right)^{0.5} \qquad \text{C.7}$$

Where D = Diameter of gasifier (m)
M_f = Fuel consumption reaction (kg/h)
SGR = Specific gasification rate (kg/m^2.hr)

Assumption D is 0.15 m (Reed & Dans, 1988),

SGR = Biomass fed rate / hearted cross sectional area C.8
= M_f / A
= M_f / πr^2
= 14.18 kg/h / 3.14 x (0.075 m)^2
= 801.844 kg/m^2.h

$$D = \left(\frac{1.27\ x\ M_f}{SGR}\right)^{0.5}$$

$$D = \left(\frac{1.27\ x\ 14.18\ kg/h}{801.844\ kg/m^2.h}\right)^{0.5}$$

D = 0.149 m

3. Time to consume biomass (T)

$$T = \frac{\rho\ x\ V_r}{M_f} \qquad \text{C.9}$$

Where T = Time require to consume biomass (hr)
ρ = Biomass density (kg/m^3)
V_r = Volume of the reactor (m^3)
M_f = Fuel feed rate (kg/h)

Assuming the reactor height is 1.2 m, then

$V_r = \frac{1}{4} \pi d^2 . T$ C.10

$V_r = \frac{1}{4}\ 3.14\ (0.15\ m)^2 .\ 1.2\ m$

$V_r = 0.021\ m^3$

$\rho = 330\ kg/m^3$ (Reed & Das, 1988)

$$T = \frac{\rho\ x\ V_r}{M_f}$$

$$T = \frac{330 \frac{kg}{m^3}\ x\ 0.021\ m^3}{14.18\ kg/h}$$

T = 0.493 hr

T = 29.60 minute

4. Reactor Height (H)

$$H = \frac{SGR\ x\ T}{\rho}$$ C.11

Where H = Length of the reactor (m)

SGR = Specific gasification rate ($kg/m^2.h$)

T = Time require to consume biomass (hr)

ρ = Biomass density (kg/m^3)

$$H = \frac{SGR\ x\ T}{\rho}$$

$$H = \frac{801.844 \frac{kg}{m^2.h}\ x\ 0.493\ h}{330\ kg/m^3}$$

H = 1.198 m

III. LADDER

The gasifier also had a ladder and a service platform to help the researcher climb up to feed the gasifier. It had 6 steps with 30 cm elevation for each step.

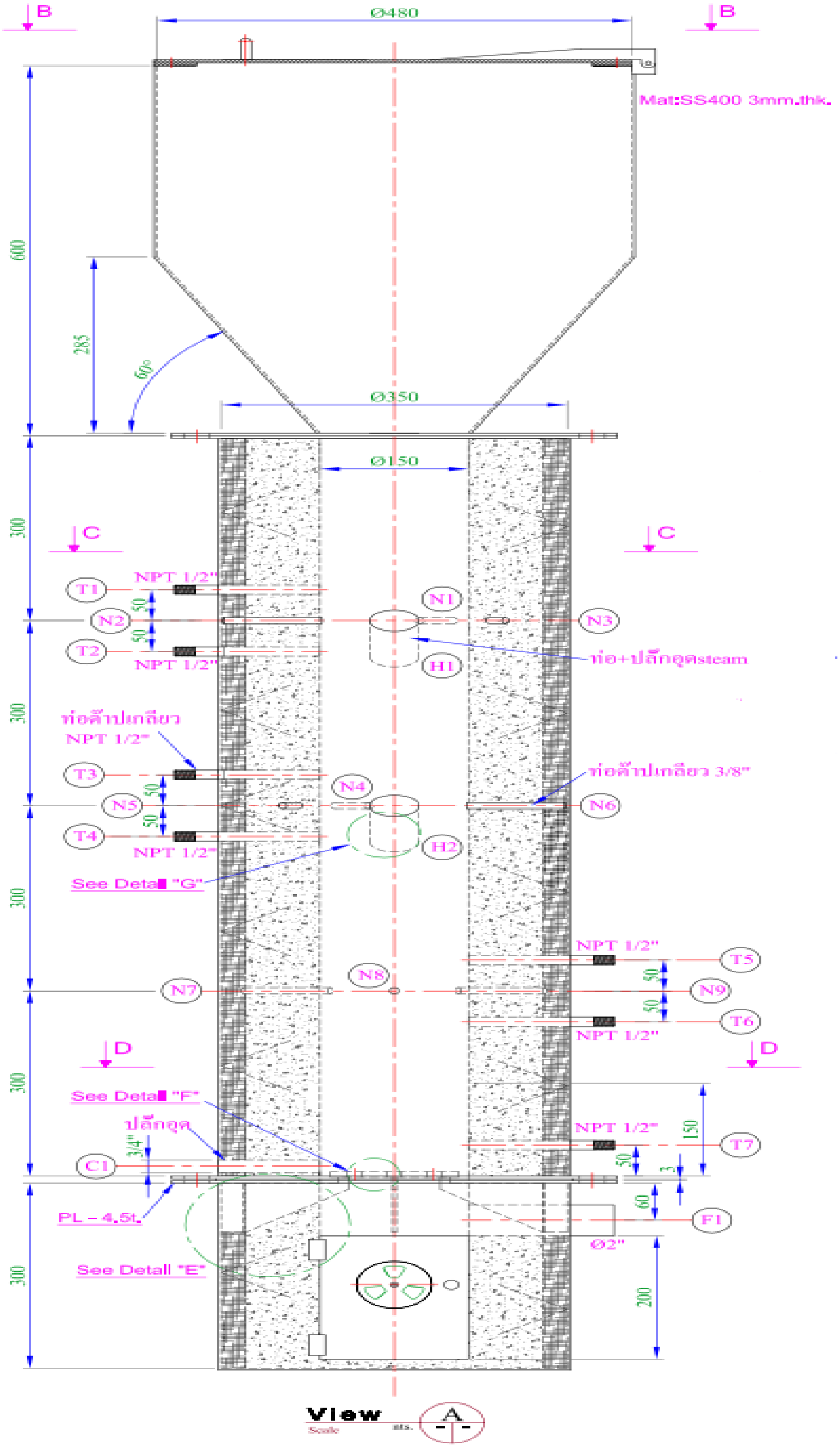


Figure C.1 The Gasifier Design Details (Side View)

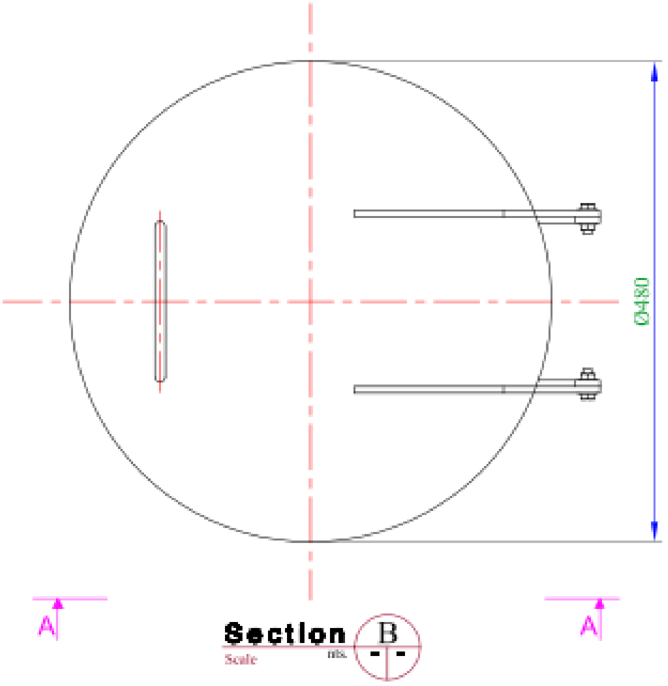


Figure C.2 The Gasifier Design Details (Upper view)

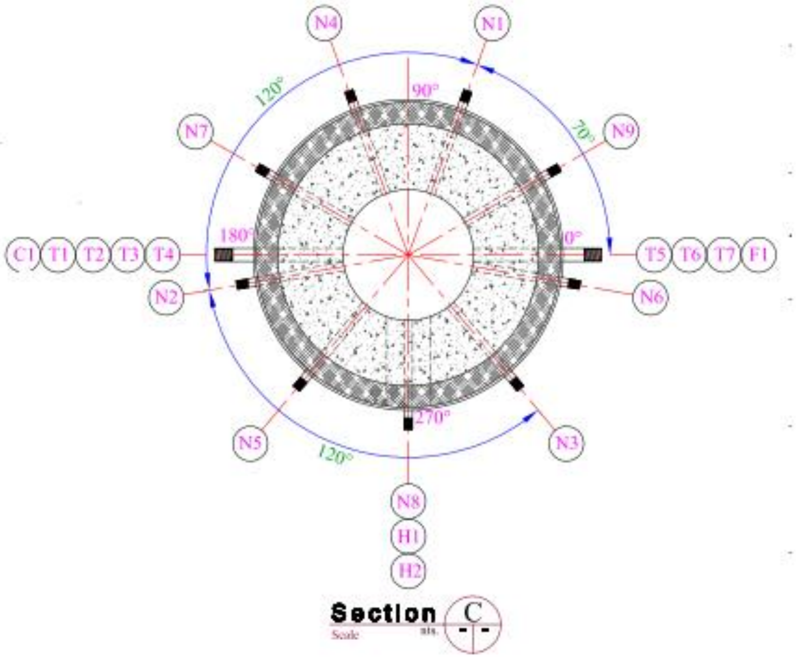


Figure C.3 The Position of Thermocouple and Air Nozzle (C : Shake Grate; H : Ignition Port; N : Air Nozzle; T : Thermocouple)

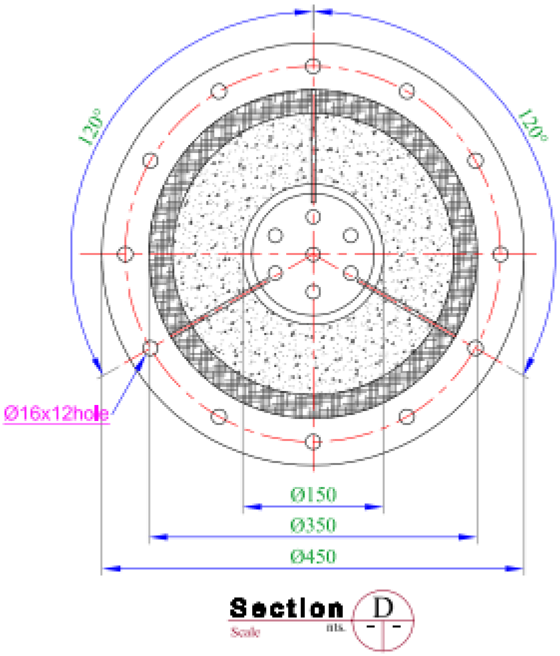


Figure C.4 The Detailed Design of Gasifier Grate

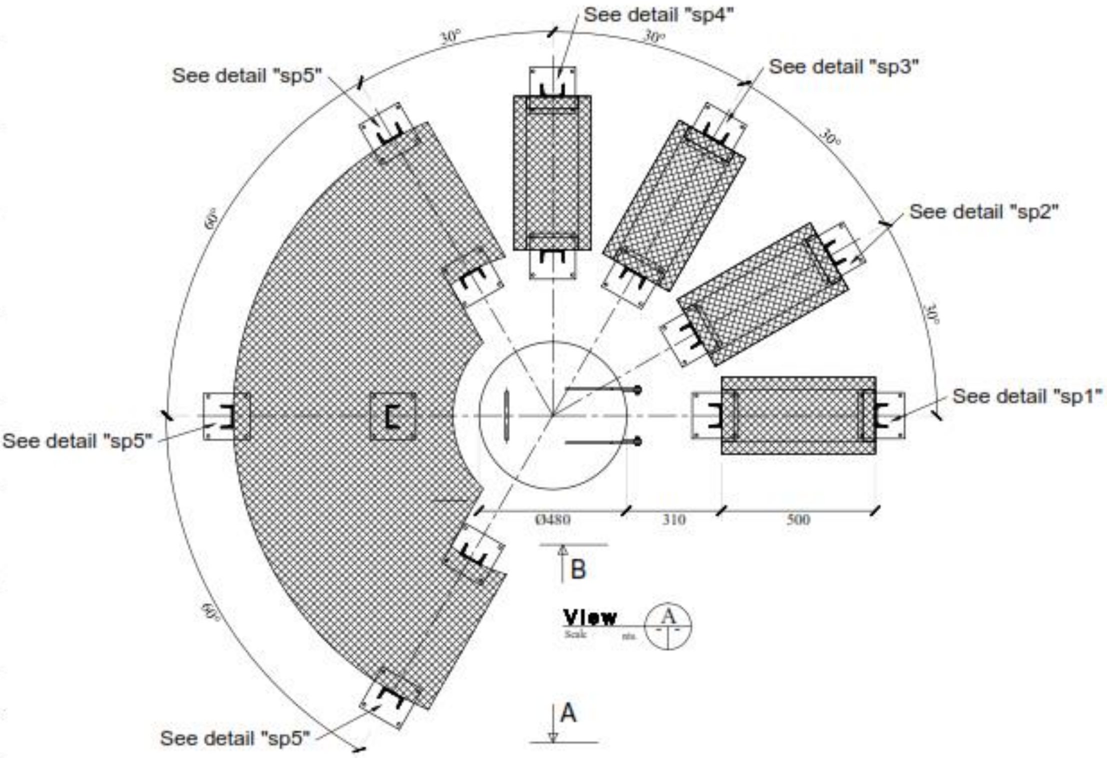


Figure C.5 The Ladder and Service Platform, Upper View

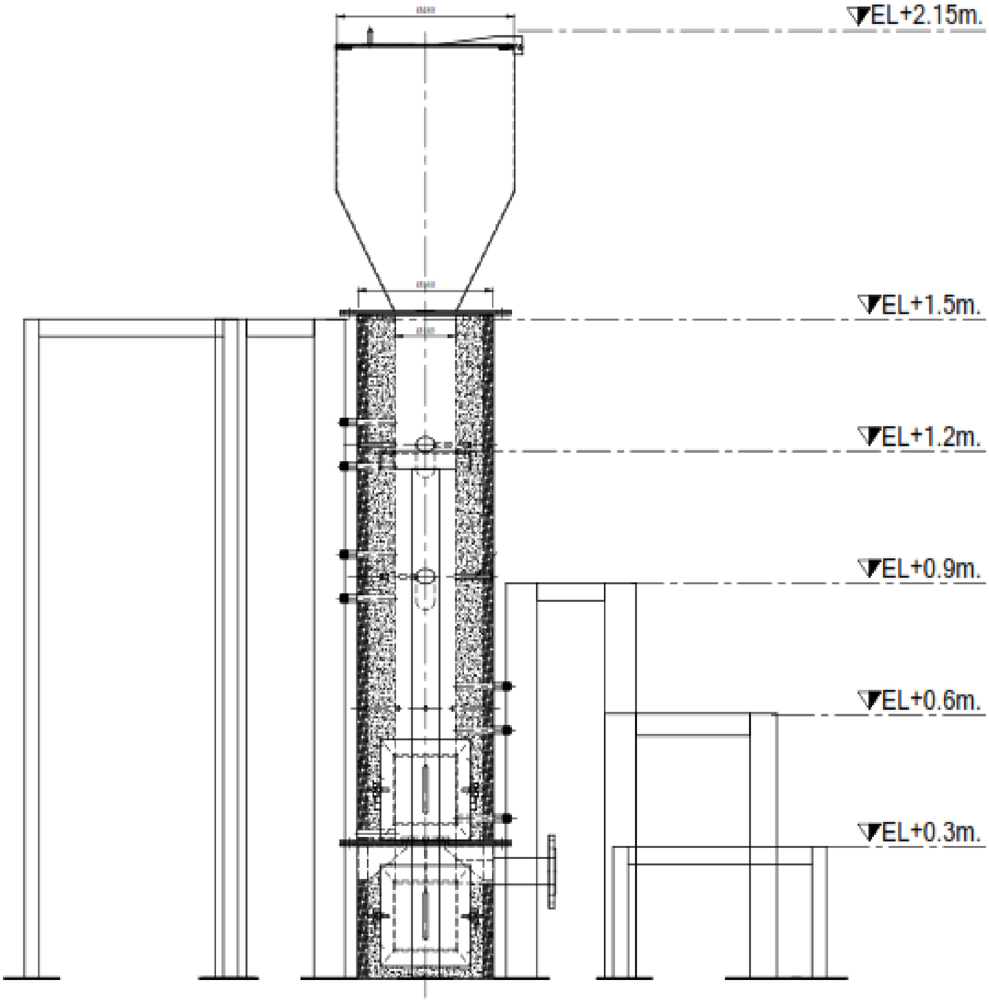


Figure C.6 The Ladder and Service Platform Detailed Design

APPENDIX D
CYCLONE DIMENSIONS

1. The Body Diameter (Dc)

Q = 128.26 m^3/h

= 0.0356 m^3/s

D = 2 inch

= 0.0508 m

$$A = \tfrac{1}{4}\pi . d^2 \tag{D. 1}$$

Where A = Surface Area of Nozzle (m^2)

d = Diameter (m)

$$A = \frac{1}{4}\pi . d^2$$

$$= \frac{1}{4} . \left(\frac{22}{7}\right) . (0.0508\ m)^2$$

$$= 0.0020\ m^2$$

$$v = \tfrac{Q}{A} \tag{D. 2}$$

Where v = Velocity (m/s)

Q = Flow Rate (m^3/s)

A = Surface Area (m^2)

$$v = \frac{Q}{A}$$

$$= \frac{0.0356 \frac{m^3}{s}}{0.020\ m^2}$$

$$= 17.5706\ m/s$$

According to the proportion of the standard cyclone dimensions, the cyclone inlet area (A) is equal to the cyclone inlet height (H_c) multiplied by the cyclone inlet width (B_c), and substituted in to Equation D.2. $A = H_c x B_c$; $H_c = {}^{D_c}/_2$ and $B_c = {}^{D_c}/_4$

$$v = \frac{Q}{A}$$

$$v = \frac{Q}{H_c x B_c}$$

$$v = \frac{Q}{{}^{D_c}/_2 \, x \, {}^{D_c}/_4}$$

$$v = \frac{8xQ}{D_c^2}$$

$$D_c^2 = \frac{8xQ}{v}$$

$$D_c = \sqrt{\frac{8 \, x \, Q}{v}} \qquad \text{D.3}$$

$$= \sqrt{\frac{8 \, x \, 0.0356 \; m^3/s}{17.5706 \, m/s}}$$

$$= 0.1274 \, m$$

2. The Dimensions

The proportion for a high-efficiency cyclone separator based (Perry, 1973) are :

$$B_c = {}^{D_c}/_4 = {}^{0.1274 \, m}/_4 = 0.0318 \, m$$

$$D_e = {}^{D_c}/_2 = {}^{0.1274 \, m}/_2 = 0.0637 \, m$$

$$H_c = {}^{D_c}/_2 = {}^{0.1274 \, m}/_2 = 0.0637 \, m$$

$$L_c = 2D_c = 2 \, x \, 0.1274 \, m = 0.2547 \, m$$

$$S_c = {}^{D_c}/_8 = {}^{0.1274 \, m}/_8 = 0.0159 \, m$$

$$Z_c = 2D_c = 2 \, x \, 0.1274 \, m = 0.2547 \, m$$

$$J_c = {}^{D_c}/_{4} = {}^{0.1274\,m}/_{4} = 0.0318\,m$$

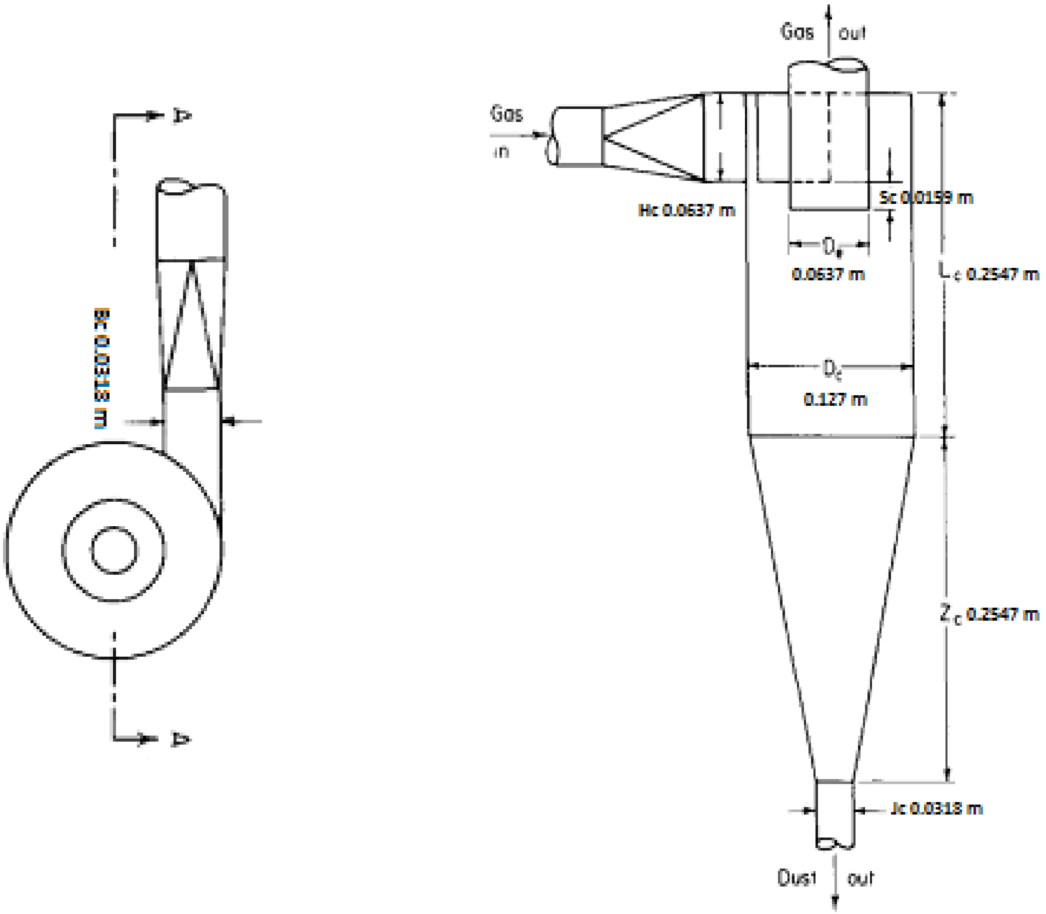


Figure D.1 The Cyclone Dimensions

3. The Pressure Drop (ΔP)

$$H_v = K\frac{H_c B_c}{D_e^2} \qquad \text{D.4}$$

Where H_v = The number of velocity heads

K = An empirical constant with a value of 16 for a tangential inlet cyclone and 7.5 for one with an inlet vane

H_c, B_c, D_e = Dimensions of Cyclone

$$H_v = K\frac{H_c B_c}{D_e^2}$$

$$= 16\ x\ \frac{0.0637\ m\ x\ 0.0318\ m}{(0.0637\ m)^2}$$

$$= 8$$

$$\Delta P = \frac{1}{2}\rho_g V_g^2 H_v \quad \text{D.5}$$

Where ΔP = Pressure Drop (N/m^2)

ρ_g = Gas Density (kg/m^3)

V_g = Gas Velocity (m/s)

$$\Delta P = \frac{1}{2}\rho_g V_g^2 H_v$$

$$= \frac{1}{2}\ 0.1221\frac{kg}{m^3}x\ (17.5706\frac{m}{s})^2\ x\ 8$$

$$= 150.7817\ \frac{kg}{m.s^2}$$

$$= 150.7817\ N/m^2$$

$$= 150.78\ Pa$$

4. The Number of Effective Turns (N_e)

$$N_e = \frac{1}{H_c}\left[L_c + \frac{Z_c}{2}\right] \quad \text{D.6}$$

$$= \frac{1}{0.0637\ m}\left[0.2547\ m + \frac{0.2547\ m}{2}\right]$$

$$= 6$$

5. Particle Cut Point (d_{50})

$$d_{50} = \left[\frac{9\mu B_c}{2\pi N_e V_i(\rho_p - \rho_g)}\right]^{1/2} \quad \text{D.7}$$

Where π = Viscosity of producer gas (kg/m.s)

N_e = The number of effective turns

V_i = Cyclone inlet velocity (m/s)

ρ_p = Density of Solid (kg/m^3)

ρ_g = Density of Producer gas (kg/m^3)

$$d_{50} = \left[\frac{9\mu B_c}{2\pi N_e V_i(\rho_p - \rho_g)}\right]^{\frac{1}{2}}$$

$$= \left[\frac{9\ x\ \left(365\ x10^{-7}\ \frac{kg}{m.s}\right)x\ 0.0318\ m}{2\ x\frac{22}{7}x6x\ 17.5706\frac{m}{s}x\ (2000 - 0.1221\frac{kg}{m^3})}\right]^{1/2}$$

= 2.8 μm

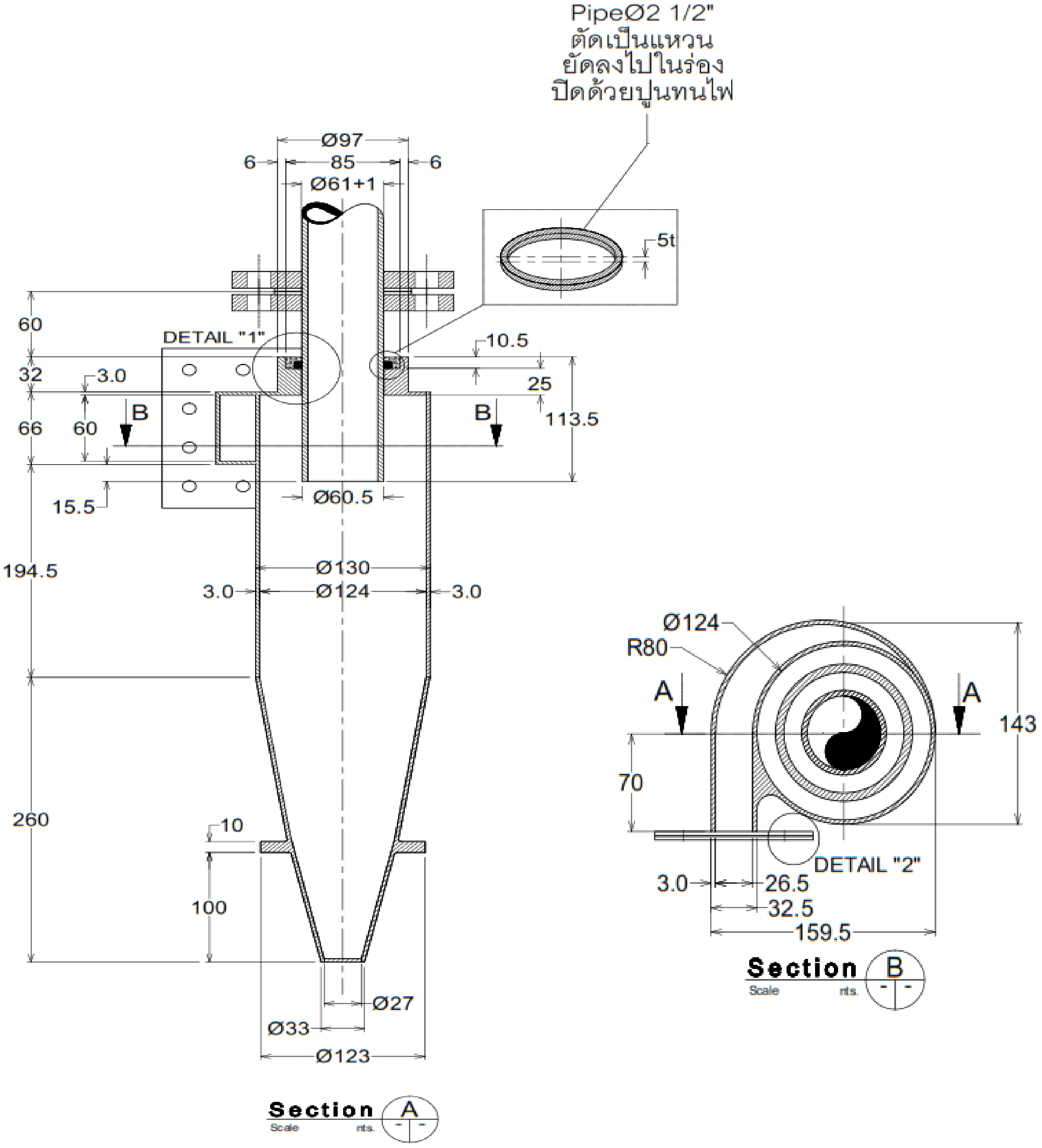


Figure D.2 The Cyclone Design Details (unit: mm)

APPENDIX E
HEAT EXCHANGER DESIGN

Hot Fluid = Producer Gas

W = 15.66 kg/h = 34.525 lb/h

T_1 = 700°C = 1292°F

T_2 = 400°C = 752°F

Cold Fluid = Air

w = 31.28 kg/h = 68.969 lb/h

t_1 = 30°C = 86°F

t_2 = 400°C = 752°F

1. Heat Load (Q)

Table E.1 Energy Balance

Component	INPUT			OUTPUT		
	n (kmol/h)	Cp (kJ/kmol.K)	Q_{in} (kJ/h)	n (kmol/h)	Cp (kJ/kmol.K)	Q_{out} (kJ/h)
CO	0.117	32.486	2575.459	0.117	30.986	1364.744
H_2	1.135	31.019	23771.810	1.135	29.999	12772.285
CH_4	0.016	69.061	730.020	0.016	54.661	321.002
CO_2	0.035	53.510	1246.960	0.035	49.148	636.285
H_2O	0.042	39.936	1125.719	0.042	37.179	582.230
N_2	0.271	31.297	5718.653	0.271	30.037	3049.125
Total	1.615		35168.622	1.615		18725.671

$$Q = Q_{in} - Q_{out} \qquad \text{E.1}$$

= 35168.622 kJ/h – 18725.671 kJ/h

= 16442.950 kJ/h

= 4567.486 Watt

$$W_{air} = Q/Cp.T \qquad \text{E.2}$$

$$= \frac{16442.950\ kJ/h}{29.23 \frac{kJ}{kg.C}\ x\ (30^oC - 25^oC)}$$

= 112.507 kg/h

= 248.034 lb/h

2. Log Mean Temperature Difference (LMTD)

Table E.2 Temperature Difference

Hot Fluid (^oF)		Cold Fluid (^oF)	Difference
1292	High Temp	752	540
752	Low Temp	86	666
540	Difference	666	

LMTD $= \frac{\Delta t_2 - \Delta t_1}{\ln(\Delta t_2 / \Delta t_1)}$ E.3

$$= \frac{666^oF - 540^oF}{\ln(666^oF - 540^oF)}$$

$= 600.799$

$R = \frac{T_1 - T_2}{t_2 - t_1}$ E.4

$$= \frac{1292^oF - 752^oF}{752^oF - 86^oF}$$

$= 0.811$

$S = \frac{t_2 - t_1}{T_1 - t_1}$ E.5

$$= \frac{752^oF - 86^oF}{1292^oF - 86^oF}$$

$= 0.552$

F_T = 0.81 Fig. E.1

Δt = LMTD x F_T E.6

= 600.799 x 0.81

= 486.648^{o}F

3. Caloric Temperature (Tc and tc)

Tc $= 0.5(T_1 + T_2)$ E.7

$= 0.5\ (1292^{o}F + 752^{o}F)$

$= 1022^{o}F$

tc $= 0.5(t_1 + t_2)$ E.8

$= 0.5\ (86^{o}F + 752^{o}F)$

$= 419^{o}F$

There are many limiting parameters to be considered for the design of shells and tube heat exchangers.

- Tube pass
- Tube diameter base on template.
- Tube length ID/OD > 15
- Tube layout 1/5 > ID/L > 1/15
- Tube pitch and clearance 1.5 > Pt/OD > 1.25
- Tube number
- Baffle 0.6 > B/ID > 0.4

$$A = \frac{Q}{U_D\ x\ \Delta t} \qquad \text{E.9}$$

$$= \frac{20507.158\ Btu/h}{70\ \frac{Btu}{h.ft^2.°\mathrm{F}}\ x\ 421.716\ °\mathrm{F}}$$

$$= 0.528\ ft^2$$

Since the producer gas and air have viscosities less than 0.5 cP, there should be between 40 – 75 Btu/hr.ft^2.^{o}C. Assuming U_D = 70 Btu/hr.ft^2.^{o}F

Table E.3 Approximate Overall Design Coefficients

Hot Fluid	Cold Fluid	Overall U_D
Water	Water	250 – 500
Aqueous solution	Aqueous solution	250 – 500
Light organic	Light organic	40 – 75
Medium organic	Medium organic	20 – 60
Heavy organic	Heavy organic	10 – 40
Heavy organic	Heavy organic	30 – 60
Light organic	Heavy organic	10 – 40

*Light organic are fluids with viscosities of less than 0.5 cP, *Medium organic have viscosities of 0.5 to 1.0 cP, *Heavy organic have viscosities above 1 cP.
(Source : Kern, 1950)

To calculate the tube number, we should estimate the trial tube length based on this equation:

$$Nt = \frac{A}{\pi.OD.L} \qquad \text{E.10}$$

Table E.4 Trial Tube Length

L	Nt calculation	Nt table	ID Shell	L/ID Shell
7	6.67	66	13.25	6.34
7	6.67	86	15.25	5.51
7	6.67	118	17.25	4.87

Table E.5 Shell and Tube Properties

	TUBE	SHELL
Type	1 – 2 Exchanger	
Outside Diameter (OD)	1 inch	
Birmingham Wire Gauge (BWG)	16	
Surface Area (a")	0.2618 ft^2/ft	
Inside Diameter (ID)	0.87 inch	15.25 inch
Tube length (L)	7 feet	
Pitch (Pt)	1.25 inch – triangular pitch	
Pass (Np)	2	
Baffle (B)	3.05 inch	
Tube Amount (Nt)	86	
	PROPERTIES	
Fluid	Producer Gas	Air
Density (ρ)	0.1443 kg/m^3	0.718 kg/m^3
Heat Capacity (Cp)	31.42 kJ/kg.C 304.42 kJ/kg.K	30.32 kJ/kg.C 303.32 kJ/kg.K
Viscosity (μ)	0.02649 cP 26.49 x 10^{-6} N.s/m^2	0.02677 cP 26.77 x 10^{-6} N.s/m^2
Thermal Conductivity (k)	0.2103 W/m.K	0.03843 W/m.K

4. Heat Transfer Area at Shell

a. Equivalent Diameter (De)

$$De = \frac{4\left(\frac{Pt^2 - \pi.OD^2}{4}\right)}{\pi.OD} \qquad E.11$$

$$= \frac{4\left(\frac{1.25\ inch^2 - 3.14\ x\ 1\ inch^2}{4}\right)}{3.14\ x\ 1\ inch}$$

$$= 0.989\ inch$$

$$= 0.025\ m$$

b. Clearance between Tubes (C)

$$C = Pt - OD \qquad E.12$$

$$= 1.25\ inch - 1\ inch$$

$$= 0{,}25\ inch$$

$$= 0.00635\ m$$

c. Heat Transfer Area (Ac)

$$Ac,s = \frac{ID\ shell\ x\ C\ x\ B}{Pt} \tag{E.13}$$

$$= \frac{15.25\ inch\ x\ 0.25\ inch\ x\ 3.05\ inch}{1.25\ inch}$$

$$= 9.3025\ inch^2$$

$$= 600.1973\ x\ 10^{-5}\ m^2$$

d. Velocity (v)

$$v = \frac{w}{\rho.Ac,s} \tag{E.14}$$

$$= \frac{112.507\frac{kg}{h}}{0.718\frac{kg}{m^3}x\ 600.1973\ x\ 10^{-5}m^2}$$

$$= 26107.246\ m/h$$

$$= 7.252\ m/s$$

e. Reynold Number (Re)

$$Re = \frac{\rho\ x\ De\ x\ v}{\mu} \tag{E.15}$$

$$= \frac{0.718\frac{kg}{m^3}x\ 0.025\ m\ x\ 7.252\frac{m}{s}}{26.77\ x\ 10^{-6}\ N.s/m^2}$$

$$= 4884.345$$

f. Tube Wall Temperature (Tw)

$$Tw = \frac{1}{2}\ (Tc + tc) \tag{E.16}$$

$$= \frac{1}{2}\ (1022°\text{F} + 419°\text{F})$$

$$= 720.5\ °\text{F}$$

$$= 382.5\ ℃$$

At a tube wall temperature of 382.5^{o}F, the viscosity was 0.03348 cP or 33.48x10^{-6} N.s/m^{2}

g. Prandtl Number (Pr)

$$\mathrm{Pr} = \frac{Cp.\mu}{k} \qquad \text{E.17}$$

$$= \frac{303.32\,\frac{kJ}{kg.K}\ x\ 26.77x10^{-6}\,\frac{N.s}{m^2}}{0.03843\,\frac{W}{m.K}}$$

$$= 0.2113\ \frac{kJ.m}{W.s}$$

h. Factor for Heat Transfer (Nu)

$$Nu = 1.86\ \left(\frac{Re.Pr.OD}{L}\right)^{1/3}\left(\frac{\mu}{\mu w}\right)^{0.14} \qquad \text{E.18}$$

$$= 1.86\left(\frac{4884.345\ x\ 0.2113\ \frac{kJ.m}{W.s}\ x\ 1\ inch\ x\ 0.0254\,\frac{m}{inch}}{7\ ft\ x\ 0.3048\,\frac{m}{ft}}\right)^{\frac{1}{3}}\left(\frac{26.77x10^{-6}\,\frac{N.s}{m^2}}{33.48x10^{-6}\,\frac{N.s}{m^2}}\right)^{0.14}$$

$$= 10.315$$

i. Heat-Transfer Coefficient (ho)

$$ho = \frac{Nu.k}{De} \qquad \text{E.19}$$

$$= \frac{10.315\ x\ 0.03843\,\frac{W}{m.K}}{0.025\ m}$$

$$= 0.011\,\frac{W}{m^2.K}$$

4. Heat Transfer Area at Tube

a. Heat Transfer Area (Ac)

$$Ac,t = \frac{\pi.ID^2.Nt}{4.Np} \qquad \text{E.20}$$

$$= \frac{3.14\ x\ 0.87inch^2\ x\ 86}{4\ x\ 2\ pass}$$

$$= 25.572\ inch^2$$

$$= 16.499\ x\ 10^{-3}\ m^2$$

b. Velocity (v)

$$v = \frac{W}{\rho.Ac,t} \qquad \text{E.21}$$

$$= \frac{15.66\ \frac{kg}{h}}{0.1443\frac{kg}{m^3}x\ 16.499\ x\ 10^{-3}m^2}$$

$$= 6577.554\ m/h$$

$$= 1.827\ m/s$$

c. Reynold Number (Re)

$$Re = \frac{\rho\ x\ ID\ x\ v}{\mu} \qquad \text{E.22}$$

$$= \frac{0.1443\frac{kg}{m^3}x\ 0.87\ inch\ x\ 1.827\frac{m}{s}\ x\ 0.0254\frac{m}{inch}}{26.49\ x\ 10^{-6}\ N.s/m^2}$$

$$= 219.938$$

d. Prandtl Number (Pr)

$$\mathrm{Pr} = \frac{Cp.\mu}{k} \qquad \text{E.23}$$

$$= \frac{304.42\frac{kJ}{kg.K}\ x\ 26.49x10^{-6}\frac{N.s}{m^2}}{0.2103\frac{W}{m.K}}$$

$$= 0.0383\ \frac{kJ.m}{W.s}$$

e. Factor for Heat Transfer (Nu)

$$Nu = 1.86\ \left(\frac{Re.Pr.ID}{L}\right)^{1/3}\left(\frac{\mu}{\mu w}\right)^{0.14} \qquad \text{E.24}$$

$$= 1.86\left(\frac{219.938\ x\ 0.0383\ \frac{kJ.m}{W.s}\ x\ 0.87\ inch\ x\ 0.0254\frac{m}{inch}}{7\ ft\ x\ 0.3048\frac{m}{ft}}\right)^{\frac{1}{3}}\left(\frac{26.49x10^{-6}\frac{N.s}{m^2}}{33.48x10^{-6}\frac{N.s}{m^2}}\right)^{0.14}$$

$$= 0.7986$$

f. Heat-Transfer Coefficient (hi)

$$hi = \frac{Nu.k}{De} \qquad \text{E.25}$$

$$= \frac{0.7986\ x\ 0.2103\,\frac{W}{m.K}}{0.025\ m}$$

$$= 6.688\,\frac{W}{m^2.K}$$

5. Overall Heat Transfer coefficient

$$U = \frac{1}{\frac{1}{ho}+\frac{1}{hi}} \qquad \text{E.26}$$

$$= \frac{1}{\frac{1}{0.011\,\frac{W}{m^2.K}} + \frac{1}{6.688\,\frac{W}{m^2.K}}}$$

$$= 0.011\,\frac{W}{m^2.K}$$

6. Pressure Drop at Shell

a. Friction Factor (f)

$$f = \exp(0.576 - 0.19 \ln Re) \qquad \text{E.27}$$

$$= \exp(0.576 - 0.19 \ln 884.3448)$$

$$= 0.354$$

b. The Viscosity Ratio (φs)

$$\varphi s = \left(\frac{\mu}{\mu w}\right)^{0.14} \qquad \text{E.28}$$

$$= \left(\frac{26.77x10^{-6}\,\frac{N.s}{m^2}}{33.48x10^{-6}\,\frac{N.s}{m^2}}\right)^{0.14}$$

$$= 0.969$$

c. Number of Shell-Side Baffles (Nb)

$$Nb = \frac{L}{B} - 1 \qquad \text{E.29}$$

$$= \frac{7\ ft\ x\ 0.3048\frac{m}{ft}}{3.05\ inch\ x\ 0.0254\frac{m}{inch}} - 1$$

$$= 26.541$$

d. Shell Pressure drop (ΔPs)

$$\Delta Ps = \frac{f.Vs^2.\rho.ID.(Nb+1)}{2.De.\varphi s} \qquad \text{E.30}$$

$$= \frac{0.354\ x\ 7.252\frac{m^2}{s}\ x\ 0.718\frac{kg}{m^3}\ x\ 15.25\ inch\ x\ 0.0254\frac{m}{inch}\ x\ (26.541-1)}{2\ x\ 0.025\ m\ x\ 0.969}$$

$$= 2931.613\ Pa$$

$$= 0.425\ psi$$

6’. Pressure Drop at Tube

a. Friction Factor (f)

$$f = 0.0035 + \frac{0.265}{Re^{0.42}} \qquad \text{E.31}$$

$$= 0.0035 + \frac{0.265}{219.938^{0.42}}$$

$$= 0.031$$

b. Tube Pressure Drop (ΔPt)

$$\Delta Pt = \left(\frac{4.f.L.Np}{ID} + 4.Np\right)\frac{f.vt^2}{2} \qquad \text{E.32}$$

$$= \left(\frac{4\ x\ 0.031\ x\ 7\ ft\ x\ 0.3048\frac{m}{ft}x\ 2}{0.87\ inch\ x\ 0.0254\frac{m}{inch}}\ x\ 4\ x\ 2\right)\frac{0.031\ x\ 1.827\frac{m^2}{s}}{2}$$

$$= 0.011\ psi$$

7. Correction Tube Length

a. Heat Transfer Area (A)

$$A = \frac{Q}{U\ x\ LMTD\ x\ F} \qquad \text{E.33}$$

$$= \frac{4567.486\ Watt}{0.011\ \frac{W}{m^2.K}\ x\ 333.777K\ x\ 0.81}$$

$$= 1538.856\ m^2$$

b. Correction Tube Length (Lt)

$$Lt = \frac{A}{\pi.OD.Nt} \qquad \text{E.34}$$

$$= \frac{1538.856\ m^2}{3.14\ x\ 1inch\ x\ 0.0254 \frac{m}{inch}\ x\ 86}$$

$$= 224.355\ m$$

$$= 736.065\ ft$$

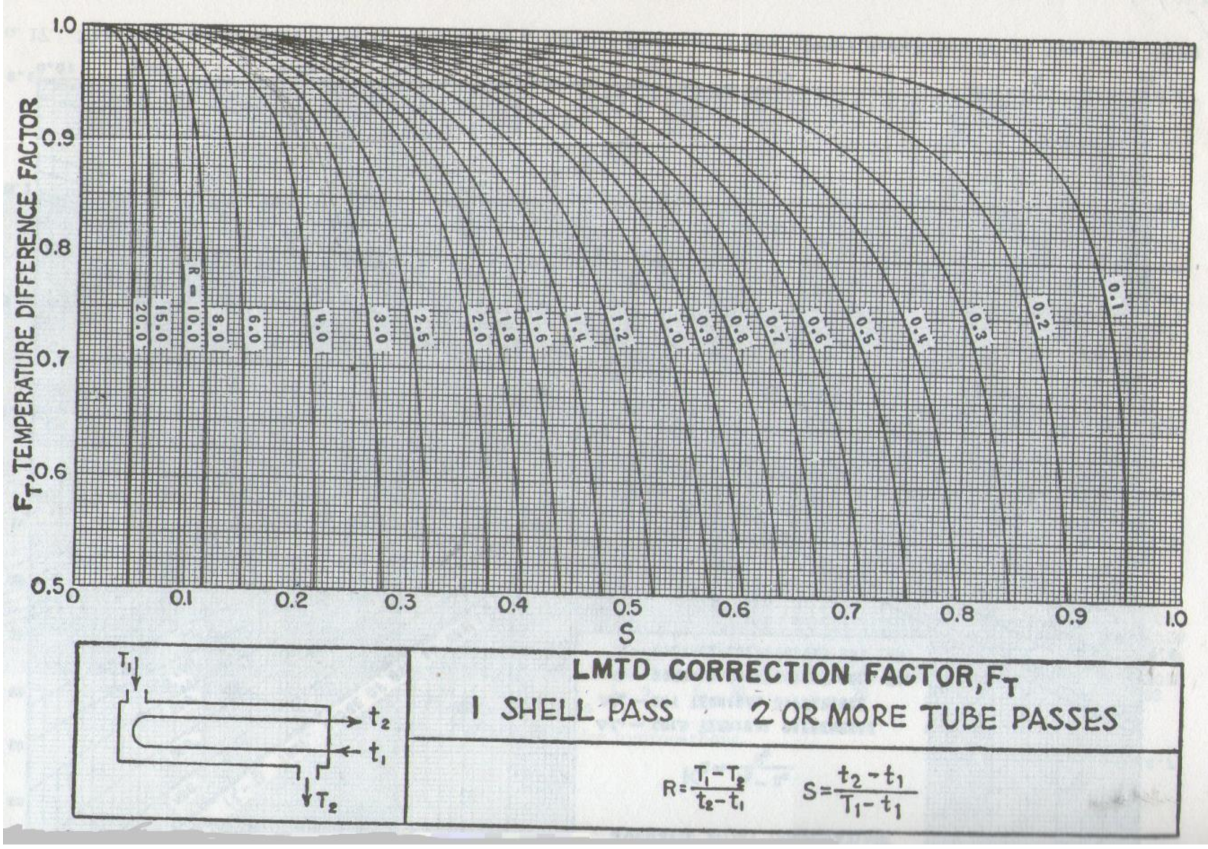
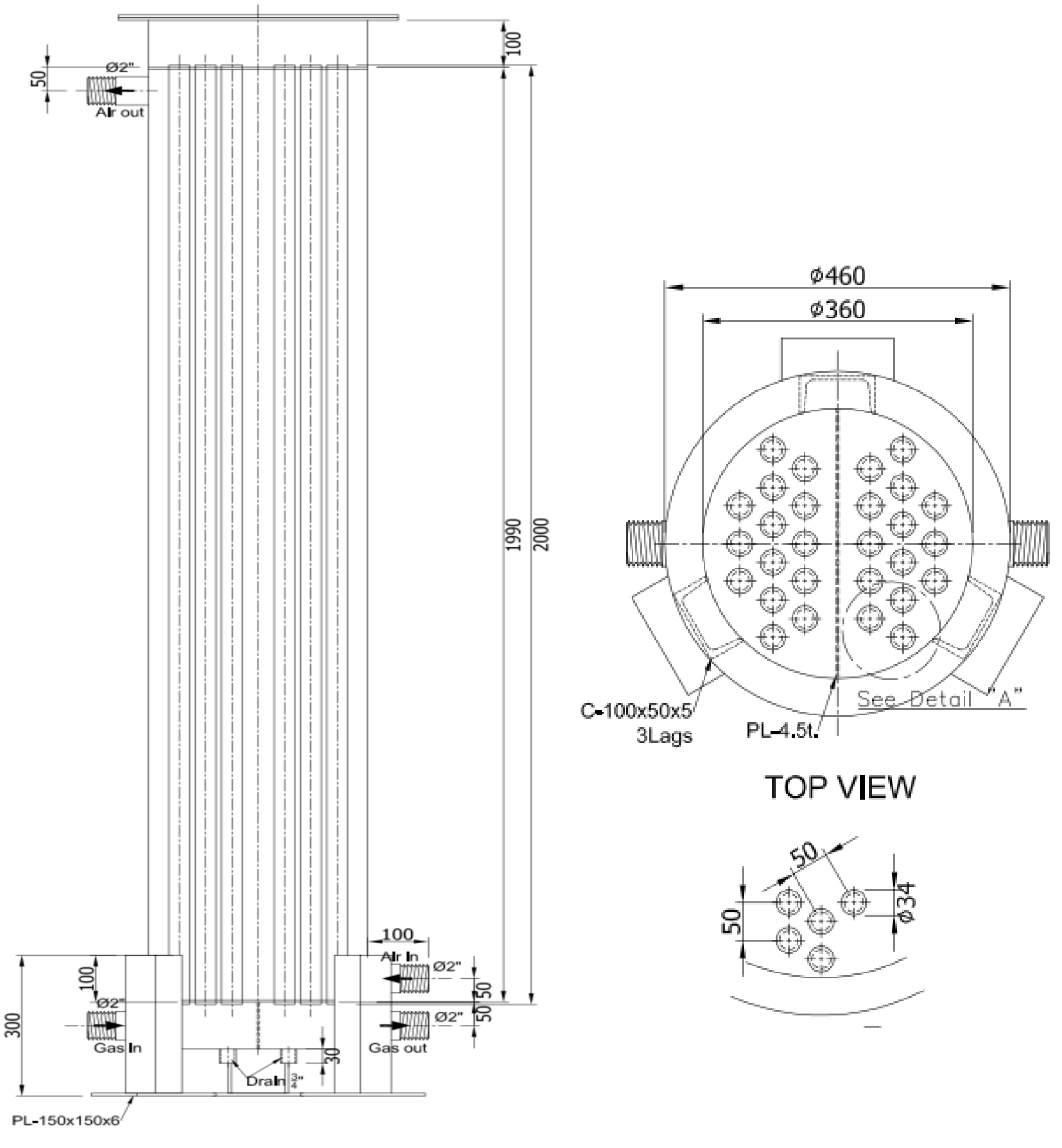


Figure E.1 LMTD Correction Factor (F_T) (Kern, 1950)



(a) (b)

Figure E.2 The Heat Exchanger Design Details (a) Side View, (b) Upper View (unit:mm)

APPENDIX F

BAGHOUSE FILTER DESIGN

1. Total Cloth Area (Ac)

$$Ac = \frac{Q}{vf} \qquad \text{F.1}$$

Where Ac = Total Cloth area (cm^2)

Q = Volumetric air flow rate (cm^3/s)

vf = Filtration Velocity (cm/s)

Q = 36 Nm^3/h

= 88.718 m^3/h at 400°C

= 24643.969 cm^3/s

The filtration velocity and air-to-cloth (A/C) ratio can be used interchangeably. Air-to cloth ratio is a measure of the amount of gas driven through each square foot of fabric in the bag house. The A/C ratio (filtering velocity) is a very important factor used in the bag house filter design. Improper ratios can contribute to inefficient operation.

Table F.1 Typical A/C or Filtration Velocity

Cleaning Mechanisms	Air-to-cloth Ratio		Filtration Velocity	
	$(cm^3/s)/cm^2$	$(ft^3/min)/ft^2$	cm/s	ft/min
Shaking	1 to 3 : 1	2 to 6 : 1	1 to 3 : 1	2 to 6 : 1
Reverse Air	0.5 to 2 : 1	1 to 4 : 1	0.5 to 2 : 1	1 to 4 : 1
Pulse-jet	1 to 7.5 : 1	2 to 15 : 1	1 to 7.5 : 1	2 to 15 : 1

$$Ac = \frac{Q}{vf}$$

$$= \frac{24643.969\ cm^3/s}{3\ cm/s}$$

$$= 8214.656\ cm^2$$

2. Amount of Fabric Required per Bag (Ab)

$$Ab = \pi . d . h \qquad \text{F.2}$$

Where Ab = Amount of fabric required per bag

d = Fabric bag diameter (cm)

h = Fabric bag height (cm)

If bag diameter and height is determined to be 15 cm and 50 cm, then

$$Ab = \pi . d . h$$

$$= 3.14 \ x \ 15 \ cm \ x \ 50 \ cm$$

$$= 2355 \ cm^2$$

3. Number of Bags Required

$$Number \ of \ Bags = \frac{Ac}{Ab} \qquad \text{F.3}$$

$$Number \ of \ Bags = \frac{24643.969 \ cm^2}{2355 \ cm^2}$$

$$= 3.48 \ Bags$$

$$= 4 \ Bags$$

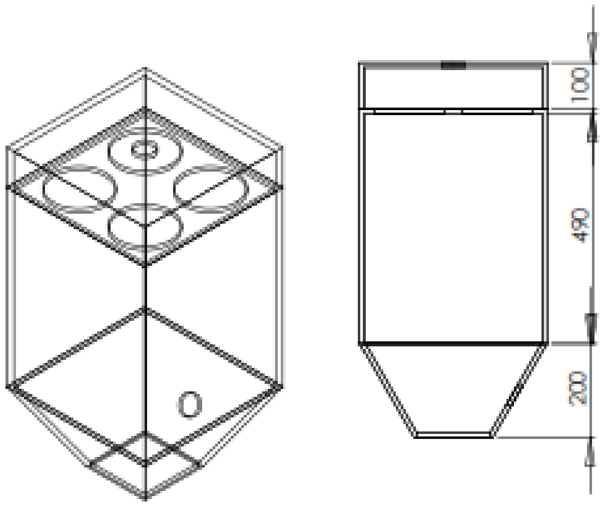


Figure F.1 Bag House Filter Design (Side View)

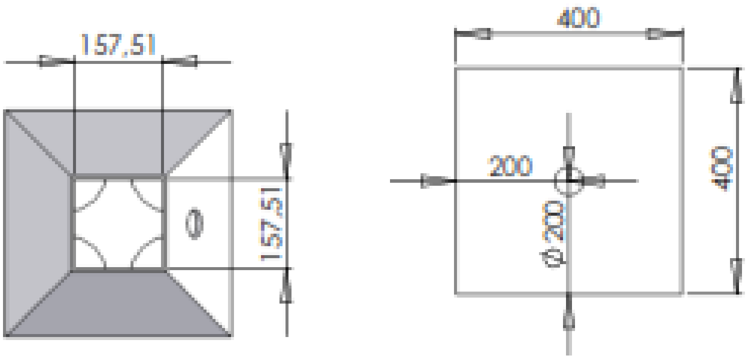


Figure F.2 Bag House Filter Design (Bottom and Top Views)

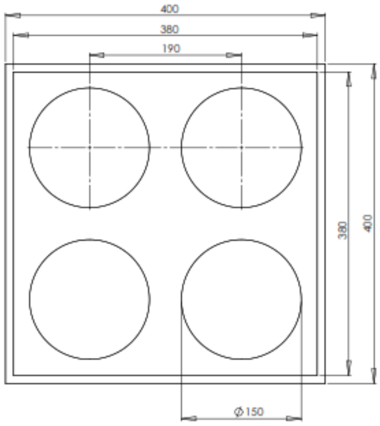


Figure F.3 Bag Filter Design

APPENDIX G
BLOWER DESIGN

1. Actual Volumetric Flow Rate (Vact)

$$V_{act} = (1 + Safety\ Factor)\ x\ Q \qquad \text{G.1}$$

Where V_{act} = Actual Volumetric Flow Rate (ft^3/s)

Q = Flow rate (ft^3/s)

Q = 128.26 m^3/h

= 75.502 ft^3/m

= 1.258 ft^3/s

Safety Factor = 20 %

= 0.2

$$V_{act} = (1 + Safety\ Factor)\ x\ Q$$
$$= (1 + 0.2) x\ 1.258\ ft^3/s$$
$$= 1.51\ ft^3/s$$

2. Optimum Pipe Diameter (Di_{opt}) (Peter, p 256)

$$Di_{opt} = 3.9\ x\ V_{act}^{0.45}\ x\ \rho^{0.19} \qquad \text{G.2}$$

Where Di_{opt} = Optimum Diameter (in)

V_{act} = Actual Volumetric Flow Rate (ft^3/s)

ρ = Gas Density (lb/ft^3)

ρ = 0.2073 kg/m^3

= 12.942 x 10^{-3} lb/ft^3

$$Di_{opt} = 3.9\ x\ V_{act}^{0.45}\ x\ \rho^{0.19}$$
$$= 3.9\ x\ 1.51\ {ft^3/s}^{0.45}\ x\ 12.942\ x\ 10^{-3}\ {lb/ft^3}^{0.19}$$
$$= 2.055\ in$$
$$= 2\ in$$

From Kern Table 11, the piping dimension are :

Inside Diameter (ID)	= 2.067 in
Outside Diameter (OD)	= 2.38 in
Nominal Pipe Size (IPS)	= 2 in
Schedule Number (SN)	= 40
Surface Area (A)	= 0.522 ft^2

3. Gas Velocity (v)

$$v = \frac{Q}{A} \tag{G.3}$$

Where v = Gas Velocity (ft/s)

Q = Flow Rate (ft^3/s)

A = Surface Area (ft^2)

$$v = \frac{Q}{A}$$

$$= \frac{1.258\ ft^3/s}{0.522\, ft^2}$$

$$= 2.411\ {}^{ft}/_{s}$$

4. Power Required (Hp)

$$Hp = 1.57x10^{-4}\ x\ Q\ x\ P \tag{G.4}$$

Where Hp = Power Required (Hp)

Q = Flow Rate (ft^3/s)

P = Pressure (in H_2O)

P = 1 atm

= 407.091 in H_2O

$$Hp = 1.57x10^{-4}\ x\ Q\ x\ P$$

$$= 1.57x10^{-4}\ x\ 1.258\ \ ft^3/s\ \ x\ 407.091\ in\ H_2O$$

$$= 0.080\ Hp$$

Assuming Motor Efficiency is 24%, then the sharp power input is

$$Ws = \frac{Hp}{Motor\ Efficiency} \qquad \text{G.5}$$

$$= \frac{0.080\ Hp}{0.24}$$

$$= 0.335\ Hp$$

$$= 1\ Hp$$

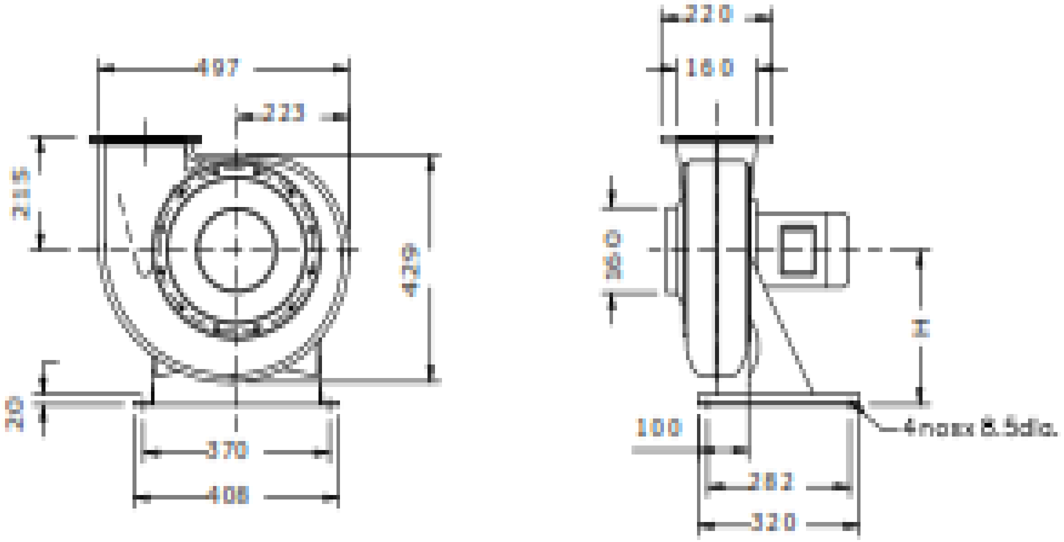


Figure G.1 Blower Design

APPENDIX H

GAS CONVERSION EFFICIENCY

1. Biomass and Producer Gas Composition

Table H.1 Raw Material and Producer Gas Composition

Biomass Composition	
Element	**Wt%**
C	45.909
O	47.338
H	5.757
N	0.442
Ash	0.555
HHV (kcal/kg)	4221.840
LHV (kcal/kg)	3700.349

Producer Gas Composition	
Element	**Vol%**
CO	6.69
CO_2	9.23
CH_4	1.00
H_2	5.27
N_2	77.81

2. The Dry Air Supply

Air supply Volume : 100 L/min

: 6 m^3/h

Air Density (ρ) : 1.157 kg/m^3 at 30°C

Air Humidity (ω) : 0.016 kg/kg for Thailand

$$m_{air} = V_{air}\ x\ \rho$$

$$= 6\ {m^3}/{h}\ x\ 1.66\ {kg}/{m^3}$$

$$= 9.960\ {kg}/{h}$$

$$m_{air} = m_{dry\ air} + m_{moisture}$$

$$= m_{dry\ air} + (m_{dry\ air}\ x\ \omega)$$

$$= m_{dry\ air} + 0.016\ m_{dry\ air}\ {kg}/{kg}$$

$$= 1.016\ m_{dry\ air}\ {kg}/{kg}$$

$$m_{dry\ air} = \frac{m_{air}}{1.1016\ {}^{kg}/_{kg}}$$

$$m_{dry\ air} = \frac{9.960\ {}^{kg}/_{h}}{1.016\ {}^{kg}/_{kg}}$$

$$= 9.803\ kg/h$$

$m_{\text{dry air}}$ for each kg of biomass $= \frac{9.803 kg/h}{14.18\ kg/h}$

= 0.691 kg dry air/kg biomass

3. **Nitrogen Flow**

The mass fraction of N_2 and O_2 in the air is 75.47 % and 23.2%.

The N_2 supply from air = Mass fraction of N_2 x $m_{\text{dry air}}$

= 0.7547 x 0.691 kg dry air/kg biomass

= 0.522 kg N_2/kg biomass

Total N_2 supplied by the feed air and the biomass, which carry 0.442% N_2 is :

= 0.522 kg N_2/kg biomass + 0.00442 kg N_2/kg biomass

= 0.526 kg N_2/kg biomass

= 0.019 kmol N_2/kg biomass

In a gas mixture, volume % = mole %

Producer gas contains 77.18% by volume of N_2. The amount of producer gas per kg of biomass is

= 0.019 kmol N_2/kg biomass / 0.7718

= 0.024 kmol gas/kg biomass

4. **Oxygen Flow**

The O_2 supply from air = Mass fraction of O_2 x $m_{\text{dry air}}$

= 0.232 x 0.691 kg dry air/kg biomass

= 0.160 kg O_2/kg biomass

The O_2 supply from moisture air feed is

$= m_{\text{moisture air}}$

$= m_{\text{dry air}} \times \omega$

= ((9.803 kg dry air/h x 0.016 kg/kg)/ 14.18 kg biomass/h) x (16/18)

= 0.010

= 0.010 kg O_2/ kg biomass

Total O_2 supply from air, moisture and 47.338% O_2 from biomass is

= 0.160 kg O_2/kg biomass + 0.010 kg O_2/kg biomass + 0.473 kg O_2/kg biomass

= 0.644 kg O_2/kg biomass

5. Hydrogen Balance

The H_2 supply from biomass and moisture in the air is

= 0.0576 kg H_2/kg biomass

+ ((9.803 kg dry air/h x 0.016 kg/kg)/ 14.18 kg biomass/h) x (2/18)

= 0.059 kg H_2/kg biomass

The H_2 leaving with CH_4 in producer gas, where 1 mole CH_4 contributes 2 mole H_2, is

= (0.527 + (2 x 0.01)) x 0.024 kmol gas/kg biomass x 2 kg H_2/kmol

= 0.004 kg H_2/kg biomass

The H_2 in producer gas is

= H_2 in – H_2 out

= 0.059 kg H_2/kg biomass – 0.004 kg H_2/kg biomass

= 0.055 kg H_2/kg biomass

The $H_2O_{(l)}$ in producer gas associated with this hydrogen is

= 0.055 kg H_2/kg biomass x (18/2) kg moisture/kg H_2

= 0.498 kg moisture/kg biomass

6. Carbon Balance

The C output in producer gas consisting of CO, CO_2 and CH_2 is

= (0.0669 + 0.0923 + 0.01) x 0.024 kmol gas/kg biomass x 12 kg C/kmol

= 0.049 kg C/kg biomass

The C input found from biomass composition 0.459 kg C/kg biomass

The Carbon conversion efficiency = (The C output / The C input) x 100%

= 10.681 %

7. Energy Balance

The Heating values for several gas constituents are :

CO : 12.63 MJ/nm^3

H_2 : 12.74 MJ/nm^3

CH_4 : 39.82 MJ/nm^3

The volume of 1 kmol of any gas is 22.4 nm^3.

Total Heating value of product gas is,

= (12.63 x 0.0669 + 12.74 x 0.0527 + 39.82 x 0.01) MJ/nm^3

x 0.024 kmol gas/kg biomass x 22.4 nm^3/kmol

= 1.732 MJ/kg biomass

Total energy input = The heating value of biomass

= 3942.050 kcal/kg

= 16504.57 kJ/kg

= 16.505 MJ/kg

8. Cold Gas Efficiency

$$\eta_{cg} = \frac{Q_g M_g}{LHV_f M_f}$$

$$= \frac{Total\ Energy\ Output}{Total\ Energy\ Input}\ x\ 100\%$$

$$= \frac{1.732\ MJ/kg}{16.505\ MJ/kg}$$

$$= 10.49\ \%$$

9. Enthalpy

The producer gas exits the gasifier at 600°C (873 K). The enthalpy of producer gas is the total enthalpy of its components. The specific heats of individual components are calculated using the relations from Table H.2 (Basu, 2010).

Table H.2 Specific Heat of Gases

Gases	Specific Heat Relation at Temperature T (K)	Specific Heat (kJ/kmol.K)
CO	27.62 + 0.005T	31.985
CO_2	$43.28 + 0.0114T – 818363/T^2$	52.158
H_2	27.71 + 0.0034T	30.678
N_2	27.21 + 0.0042T	30.877
CH_4	22.35 + 0.048T	64.254

The enthalpy of CO in the product gas that contains 6.69% CO above the ambient temperature, 30°C or 303°K, is

Q = n.Cp.ΔT

= (0.0669 x 0.024 kmol gas/kg biomass) x 31.985 kJ/kmol.K x (873-303) K x 10^{-3} MJ/kJ

= 0.0297 MJ/kg biomass

Table H.3 Enthalpy Calculation

Component	Vol%	Mol (kmol/kg)	Specific Heat (kJ/kmol.°K)	ΔT (°K)	Enthalpy (MJ/kg)
CO	6.69	0.0016	31.985	570	0.0297
CO_2	9.23	0.0022	52.158	570	0.0669
H_2	5.27	0.0013	30.678	570	0.0225
N_2	77.81	0.0188	30.877	570	0.3336
CH_4	1.00	0.0002	64.254	570	0.0089

The amount of steam in the producer gas was calculated as 0.498 kg moisture/kg biomass. The steam enthalpies at 873K and 303 K are 3705 kJ/kg and 125.76 kJ/kg, respectively, so the enthalpy in water is

= 0.498 kg moisture/kg biomass x (3705 – 125.76 kJ/kg) x 0.001 MJ/kJ
= 1.781 MJ/kg biomass

The total enthalpy of producer gas at 873 K is

= (0.0297 + 0.0669 + 0.0225 + 0.3336 + 0.0089 + 1.781) MJ/kg biomass
= 2.243 MJ/kg biomass

Total thermal energy = Heating Value + Enthalpy
= 1.732 MJ/kg biomass + 2.243 MJ/kg biomass
= 3.974 MJ/kg biomass

10. Hot Gas Efficiency

$$\eta_{hg} = \frac{Q_g M_g + M_g C_p (T_f - T_o)}{LHV_f M_f}$$

$$= \frac{Total\ Thermal\ Energy}{Total\ Energy\ Input}\ x\ 100\%$$

$$= \frac{3.974\ MJ/kg}{16.505\ MJ/kg}\ x\ 100\%$$

$$= 24.08\ \%$$

Table H.4 Effect of Air-stage on Gas Conversion Efficiency

Stages	**Wood Chips**			**Wood Pellets + Wood Chips**		
	Carbon	Cold Gas	Hot Gas	Carbon	Cold Gas	Hot Gas
1	10.68	10.49	24.08	14.41	17.35	31.11
2	8.29	10.70	24.31	20.04	20.97	34.87
3	10.24	11.81	25.36	9.60	10.97	24.98

APPENDIX I

BIOMASS AND AIR FEED RATE ACTUAL

1. Power Output (kWth) Based on Producer Gas

$$V_g = \frac{Q}{LHV_g} \, Nm^3/s \tag{I.1}$$

Where V_g = Volume flow rate of the product gas (Nm^3/s)

Q = Gasifier required power output (MWth)

$LHVg$ = Lower heating value of the product gas (MJ/Nm^3)

LHVg = 1.732 MJ/kg

= 13.063 MJ/m^3

= 4.085 MJ/Nm^3

Vg = 0.024 kmol gas/kg biomass

= 0.0373 m^3/s at 15 kg/h biomass

= 0.0117 Nm^3/s

$$Q = Vg \; x \; LHVg$$

Q = 0.0117 Nm^3/s x 4.085 MJ/Nm^3

= 0.0477 MJ/s at 1 atm and 700°C

= 0.0477 MWth

= 47.69 kWth

2. Power Output (kWth) Based on Biomass Feed Rate

$$M_f = \frac{Q}{LHV_{bm}\eta_{gef}} \tag{I.2}$$

Where M_f = Biomass feed rate, 0.00417 kg/s

Q = Gasifier required power output (MWth)

LHV_{bm} = Lower heating value of biomass, 16.5046 MJ/kg

η_{gef} = Gasifier efficiency, 25.36%

$$
\begin{aligned}
Q &= M_f \cdot LHV_{bm} \cdot \eta_{gef} \\
&= 0.00417 \text{ kg/s} \times 16.5046 \text{ MJ/kg} \times 0.2536 \\
&= 0.0174 \text{ MJ/s} \\
&= 17.4 \text{ kWth}
\end{aligned}
$$

To generate 50 kWth, the biomass feed rate should be

$$M_f = \frac{Q}{LHV_{bm}\eta_{gef}}$$

$$M_f = \frac{0.05\ MJ/s}{16.5046\ \frac{MJ}{kg}\ x\ 0.2536}$$

$$
\begin{aligned}
M_f &= 0.0119 \text{ kg/s} \\
&= 43.01 \text{ kg/h}
\end{aligned}
$$

3. Equivalence Ratio (ER)

Table I.1 Proximate and Ultimate Analyses of Eucalyptus sp

Biomass Composition	
Element	**Wt%**
C	45.909
O	47.338
H	5.757
N	0.442
Moisture	44.760
Ash	0.555
HHV (kcal/kg)	4221.840
LHV (kcal/kg)	3700.349

Table I.2 Oxygen Calculations

Element	**wt %**	**kmol**	**Rx**	**O_2 Required**
C	45.909	2.67	$C + O_2 \rightarrow CO_2$	1.22
H	5.757	8.00	$H_2 + ½\ O_2 \rightarrow H_2O$	0.46
O	47.338	-1.00	$O_2 \rightarrow O_2$	-0.47
N	0.442	2.29	$½\ N_2 + O_2 \rightarrow NO_2$	0.01
		Total		1.22

Stoichiometric O_2 required is 1.22 kg O_2/kg biomass

The total O_2 required for 15 kg Eucalyptus wood chip is

= 15 kg biomass/h x 1.22 kg O_2/kg biomass

= 18.323 kg O_2/h

= 18.323 kg O_2/h / 32 kg O_2/kmol

= 0.573 kmol/h

Stoichiometric air required :

$m_{\text{dry air}}$ = wt O_2 + wt N_2 I.3

= (0.573 kmol/h x 32 kg O_2/kmol) + (0.573 kmol/h x 3.25 x 28 kg N_2/kmol)

= 18.323 kg O_2/h + 52.154 kg N_2/h

= 70.477 kg air/h

Actual air supply (V) = 100 L/min

= 6000 L/h

= 6 m^3/h

Air Density (ρ) = 1.66 kg/m^3 at 30°C

Humidity (ω) = 0.016 kg/kg for Thailand (Sarasuk, 2010)

Actual air mass = V x ρ

= 6 m^3/h x 1.66 kg/m^3

= 9.960 kg/h

m_{air} = $m_{\text{dry air}}$ + m_{moisture}

= $m_{\text{dry air}}$ + ($m_{\text{dry air}}$ x ω)

= $m_{\text{dry air}}$ + 0.016 $m_{\text{dry air}}$ kg/kg

= 1.016 $m_{\text{dry air}}$ kg/kg

$m_{\text{dry air}}$ = m_{air} / 1.016 kg/kg

$$= \frac{9.960\ kg/h}{1.016\ kg/kg}$$

= 9.803 kg/h

$$ER = \frac{(wt\ of\ air/wt\ of\ fuel)_{actual}}{(wt\ of\ air/wt\ of\ fuel)_{stoiciometri}}$$

$$ER = \frac{9.803\ ^{kg}/_{h}}{70.477\ ^{kg}/_{h}}$$

ER = 0.139

4. **Air Velocity (v)**

$Q = v.A$ I.4

Where Q = Air Flow Rate (m^3/s)

v = Air velocity (m/s)

A = Surface Area of nozzle (m^2)

Q = 6 m^3/h

= 0.002 m^3/s

A = ¼ .π. D^2

= ¼. 3.14. $(0.013\ m)^2$

= $12.673 \times 10^{-5}\ m^2$

v = Q/A

= 0.002 m^3/s / $12.673 \times 10^{-5}\ m^2$

= 13.152 m/s

For three nozzles at first stage, the velocity becomes :

V_n = v / amount of nozzle 1.5

= 13.152 m/s / 3 nozzle

= 4.384 m/s for each nozzle

Table I.3 Air Velocity

Stage	Amount of Nozzle	Air Velocity (m/s)
1	**3**	4.384
2	**6**	2.192
3	**9**	1.461

APPENDIX J

EXPERIMENTAL DATA

Table J.1 Temperature Distribution of Single Air-Stage Wood Chips Gasification

Time	T1	T2	T3	T4	T5	T6	T7	T8	T9
0	32	32	32	32	32	32	32	32	32
10	98	169	35	34	32	32	31	148	38
16	150	334	38	33	32	32	31	293	41
18	201	266	37	33	32	32	32	365	42
21	220	161	35	33	32	32	32	513	48
80	196	82	37	34	33	32	32	700	98
82	197	102	37	34	33	32	32	767	102
84	200	110	41	35	33	32	32	731	106
99	205	89	48	35	33	32	32	674	120
101	223	86	45	36	33	32	32	706	118
103	216	82	44	36	33	32	32	698	117
105	228	84	46	35	33	33	32	655	118
111	233	80	44	35	33	33	32	741	117
115	235	78	45	35	33	33	32	782	118
120	267	86	58	35	34	33	33	696	89

Table J.2 Temperature Distribution of Double Air-Stage Wood Chips Gasification

Time	T1	T2	T3	T4	T5	T6	T7	T8	T9
0	33	34	33	34	34	34	34	33	33
2	89	154	93	34	33	33	33	84	33
4	90	236	65	33	33	33	33	124	37
6	97	306	56	34	33	33	33	164	39
9	144	265	47	34	33	33	33	378	41
12	137	248	48	34	33	33	33	450	48
15	171	212	45	34	33	33	33	562	57
18	214	166	45	34	33	33	33	611	68
21	258	153	44	34	33	33	33	631	80
23	310	144	44	34	33	33	33	640	85
30	387	333	45	34	33	33	33	646	103
33	373	323	48	34	33	33	33	625	106
36	354	304	51	35	33	33	33	605	109
40	268	427	52	35	33	33	33	561	112
42	278	388	56	36	33	33	33	559	111
91	316	391	70	37	35	34	34	517	111
127	202	366	222	49	49	38	36	317	53

Table J.3 Temperature Distribution of Triple Air-Stage Wood Chips Gasification

Time	T1	T2	T3	T4	T5	T6	T7	T8	T9
0	113	162	114	40	47	40	38	94	34
5	164	140	175	52	47	39	38	102	34
35	206	479	208	48	44	38	37	418	66
38	204	442	258	48	45	38	37	411	69
41	220	480	197	49	44	38	37	411	71
89	267	434	283	48	44	38	37	661	86
91	281	406	206	49	44	38	37	605	94
104	328	552	199	49	43	38	37	545	109
108	396	529	214	49	43	38	37	513	107
119	324	606	181	48	43	37	37	557	91

Table J.4 Temperature Distribution of Single Air-Stage Wood Chips + Wood Pellet Gasification

Time	T1	T2	T3	T4	T5	T6	T7	T8	T9
0	32	31	31	32	32	32	32	32	33
2	39	61	129	33	32	32	32	57	33
3	54	92	505	33	32	32	32	70	34
4	57	188	577	33	32	32	32	75	35
5	55	277	516	33	32	32	32	80	37
6	58	338	490	34	32	32	32	90	38
8	66	488	386	34	32	32	32	122	39
9	75	507	326	34	32	32	32	154	40
13	110	492	187	34	33	32	32	231	42
55	124	539	229	36	33	32	32	275	43
63	286	257	164	41	35	32	32	619	57
64	350	240	157	41	35	32	33	608	60
74	292	163	94	41	35	33	33	638	55

Table J.5 Temperature Distribution of Double Air-Stage Wood Chips + Wood Pellet Gasification

Time	T1	T2	T3	T4	T5	T6	T7	T8	T9
0	81	78	65	44	43	38	39	52	33
7	70	65	52	43	42	38	38	48	33
13	68	60	48	43	42	38	38	47	33
23	115	597	369	49	41	37	37	152	40
24	128	622	312	48	41	37	37	169	41
25	140	581	287	48	41	37	37	197	42
26	165	577	240	48	41	37	37	197	42
31	305	382	235	52	43	38	37	363	47
33	336	330	175	51	43	37	37	430	51

Table J.6 Temperature Distribution of Triple Air-Stage Wood Chips + Wood Pellet Gasification

Time	T1	T2	T3	T4	T5	T6	T7	T8	T9
0	33	32	32	33	32	32	32	33	33
1	33	37	59	33	33	33	32	33	33
3	36	61	93	35	33	32	32	56	33
4	41	65	178	35	33	32	32	66	33
5	44	86	321	35	33	33	33	71	35
7	48	130	399	36	33	33	32	71	36
8	47	176	429	36	33	32	32	73	37
9	50	220	443	36	33	33	32	82	39
10	53	263	428	36	33	32	32	90	39
12	65	365	610	37	33	32	32	131	40
14	68	460	504	39	34	32	32	144	41
23	248	286	237	51	38	33	32	436	46
24	265	264	227	51	38	33	32	449	48
25	268	247	213	51	38	33	32	461	50
26	268	232	203	51	38	33	32	469	50
27	267	227	200	51	38	33	32	474	51

Table J.7 Producer Gas Composition of Wood Chips Gasification

Stage	CO	CO2	H2	CH4
1	6.69	9.23	5.27	1.00
2	6.43	6.40	6.68	0.64
3	8.49	6.95	5.43	0.87

Table J.8 Producer Gas Composition of Wood Chips + Wood Pellet Gasification

Stage	CO	CO2	H2	CH4
1	10.68	7.69	8.06	0.82
2	13.73	9.84	9.08	0.83
3	13.25	4.90	3.35	0.36