

**HYDROGEN-RICH GAS PRODUCTION FROM BIOMASS AIR/STEAM
GASIFICATION BY USING HEAT RECOVERY IN
A DOWNDRAFT GASIFIER**



**A Thesis Submitted to the Graduate School of Naresuan University
in Partial Fulfillment of the Requirements
for the Doctor of Philosophy Degree in Renewable Energy
June 2015**

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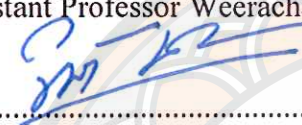
Thesis entitled “Hydrogen-Rich Gas Production From Biomass Air/Steam Gasification
By Using Heat Recovery In A Downdraft Gasifier”


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has been approved by the Graduate School as partial fulfillment of the requirements
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Title HYDROGEN-RICH GAS PRODUCTION FROM BIOMASS
AIR/STEAM GASIFICATION BY USING HEAT
RECOVERY IN A DOWNDRAFT GASIFIER

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ABSTRACT

This experimental study aims to increase the concentrations of hydrogen gas in the process of biomass gasification with self – produced steam by using waste sensible heat of product gas from a fixed bed downdraft gasifier. The original gasifier systems can produce a thermal power of 50 kW_{th}. The temperature of a reactor is held at 920°C, and the biomass rate of 14 kg/h. The flow rate of the product gas is 32 Nm³/h, and the temperature of the flow from the gasifier to the heat exchanger is 350°C. For the original system, the energy balance indicated that the waste heat from the product gas is 3.41 kW_{th}, which can be used by the heat exchanger to produce steam. A small size of a shell and tube counter-flow heat exchanger with the heating surface area of 0.068 m² is designed. The effectiveness of the heat exchanger is 0.58. The steam (120°C and 1.2 bar) can be produced at 2.6 kg/h (maximum). The steam produced and the ambient air at 30°C are mixed, the mixture flows into the gasifier through the 6-hot pipes. Each pipe is 40 cm in length and 6/8 inch in diameter. Then, a mixture of the steam and the air is heated up to 580°C, and then flows into the gasifier as an oxidizing agent. To conclude, it has been found that the product gas is contained with 19% of hydrogen (8% higher). The heating value of the product gas is in a range of 4.70 to 5.20 MJ/Nm³ (26% higher concentration than air gasification). The equivalent ratio, ER, is 0.45. Furthermore, the economic analysis implied that the

energy cost of the gasifier with a heat exchanger would be lower than that of the original one. The payback period is 3-year and 1-month, which would be 4-month shorter than the original period.



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LIST OF SYMBOLS

Abbreviations

A	=	area (m ²)
C _p	=	specific heat capacity at constant pressure (kJ/kg °C)
\dot{m}	=	mass flow rate (kg/s)
P	=	pressure (Pa)
\dot{Q}	=	heat transfer rate, kW
T	=	temperature (°C)
U	=	Overall heat transfer coefficient (W/m ² °C)
\dot{V}	=	volumetric flow rate (m ³ /s)
k	=	thermal conductivity (W/m°C)
HHV	=	high heating value (MJ/Nm ³)
LHV	=	low heating value (MJ/Nm ³)
H	=	weight fraction of hydrogen
C	=	weight fraction of carbon
N	=	weight fraction of nitrogen
O	=	weight fraction of oxygen
S	=	weight fraction of sulphur
ER	=	equivalence ratio
BWG	=	Birmingham wire gauge
y	=	molar fraction
h	=	specific enthalpy, kJ/kg
HE	=	heat exchanger
GC	=	Gas chromatography
kW _{th}	=	kilowatt-hour (thermal)

Greek Symbols

LIST OF SYMBOLS (CONT.)

Subscripts

CO	=	carbon monoxide
H ₂	=	hydrogen
CH ₄	=	methane
CO ₂	=	carbon dioxide
a	=	air
e	=	exit
o	=	outlet
i	=	inlet
s	=	surface or steam
h	=	hot
c	=	cold
th	=	thermal power
fg	=	difference in property between saturated liquid and saturated vapor

CHAPTER I

INTRODUCTION

Rationale of the study

At present, there are concerns about the depletion of fossil fuel reservation and air pollution caused by continuously increasing energy demands. These problems have led many researchers to search for new energy sources in replacement of fossil fuels. As an alternative to fossil fuels in the future, biomass is a renewable fuel, and has significant environmental benefits, for instance, a near zero CO₂ emissions for the reduction in global warming [1].

Thailand is an agriculture-based country. There are plenty of biomasses in the country area. Wood chip is one of the interesting biomasses because it is easy to cultivate, it is one of the fastest-growing plant, and it could be harvested in 2-3 years.

Biomass energy conversion system can be achieved through several thermochemical processes, namely 1) combustion 2) pyrolysis and 3) gasification [2]. Among these processes, gasification process for hydrogen production is one of the most promising methods. It is widely known that biomass gasification offers a great potential to produce fuel gas that can be used to the synthesis of gas applications. However, the production of gas from this process usually contains unacceptable levels of tar and low heating value. Tar can cause operational problems by blocking gas cooler and filter elements. Most producer gas applications also require the removal of dust and tar in order for the gas to be used. Tar can be effectively minimized in raw producer gas by steam gasifying agent [3, 4].

The gasification process requires some gasifying agent that provides oxygen for the formation of CO from solid carbon in the fuel [5]. The gasifying agents can be air, oxygen, steam or CO₂ [6, 7]. The CO₂ is produced during pyrolysis and early oxidation processes and is generally not externally added. The most common agent is air because of its extensive availability at no cost. Air, though cheap, is not a perfect agent because of its nitrogen content. The product gas from air gasification produces a low heating value gas (4 -7 MJ/Nm³). Oxygen gasification produces

a higher heating value ($10-18 \text{ MJ/Nm}^3$) but has a drawback of a relatively high cost associated with oxygen. Steam could be another alternative as well. It also generates a medium calorific value gas ($10-14 \text{ MJ/Nm}^3$) [8]. The main advantage is that it increases the hydrogen content in the product gas.

Therefore, the addition of steam in gasification process should be made to obtain a higher H_2 content and to remove tar [9]. Many researchers have been extensively studied and proved that steam is useful and effective in decreasing tar, and improving gas quality in the process of biomass gasification.

In this study, a downdraft gasifier was utilized to investigate gas production from the gasification of wood chip in a downdraft fixed bed gasifier with air and steam in the reactor. The purpose of this study is to evaluate the hydrogen production from gasification of wood chip, as well as to explore the effects of some operating parameters, such as a steam flow rate on the product gas composition, and a heating value of a product gas.

Statement of problem

Due to the oil crisis, energy shortage and environmental pollution. As a result, many countries are increasingly turning to renewable energy. Biomass is a renewable energy that can be easily provided, it is cheap and it does not cause global warming.

At present, the role of fossil fuels as an energy source are possible, although in the future it will be less supplied. The problem of the inefficient use is also important. Due to the loss of energy and the pollution in the environment, the development in technology is essential and must be done continuously. You will need to use energy more efficiently, but you also need to find new sources of energy as well. You will also need to use technology appropriately and efficiently to produce energy, whether in the form of electricity, the heat if the liquid fuel, natural gas and synthetic, etc.

In recent years, the use of hydrogen as a fuel for energy production are few despite the fact that the combustion and exhausted gases are happen to be clean, and friendly to the environment.

Biomass gasification technology changes biomass into fuel gas with low calorific value, especially when using only air as oxidizing agents. The composition of the gas product obtained consists of hydrogen at low concentrations, making a combustible gas to be a low-calorific value [10, 11].

The use of steam in the gasification process is one way to increase the concentration of hydrogen in the product gas [12]. The sensible heat in product gas at high temperature will heat the water to vaporize it. With this method, it will not use an external heat source. The reason for this is the source of waste heat recovery of synthesis gas to produce hydrogen at high concentrations.

Hydrogen as a clean energy carrier is believed to be the most promising source to replace fossil fuel. Biomass gasification with the presence of steam offers a feasible, sustainable, and environmental-friendly option as well as a favorable alternative for higher hydrogen yields and for large-scale hydrogen production which will be able to satisfy the need of hydrogen in the future [13].

Biomass is a renewable energy that is readily available and cheap [14]. It does not cause global warming. Gasification technology changes biomass into fuel gas with low a calorific value; especially, when the air is only used as the oxidizing agents. The composition of the product gas, consisting of hydrogen at low concentrations, makes the combustible gas to have a low calorific value [13].

Air, as the oxidizing agent, is easy to find with no cost. Unfortunately, the air in the atmosphere consists of nitrogen in excessive amounts. The product gas from the air gasification has a low calorific value (4.1-4.5 MJ/Nm³), while oxygen gasification produces a high heating value gas (10-18 MJ/Nm³)[8]. Furthermore, the production costs are high. The use of steam, as the oxidizing agent, is an attractive alternative. Steam is cheap and easy to be generated. In addition, the heating value of the product gas is moderate (5-10 MJ/Nm³) and the amount of hydrogen in the product gas can be increased [15].

As the oxidizing agent, steam from the external source has a high cost. Therefore, the sensible heat of the product gas can be used to produce steam without an external heat source. (A self-produced system in the same reactor). The recovery of waste heat in the product gas can increase the concentrations of hydrogen gas[16].

The original gasifier system used in this study is $50 \text{ kW}_{\text{th}}$ and it used air as an oxidizing agent. The study found that the system producing hydrogen was low. Therefore, a small heat exchanger is designed to work with the gasifier system.

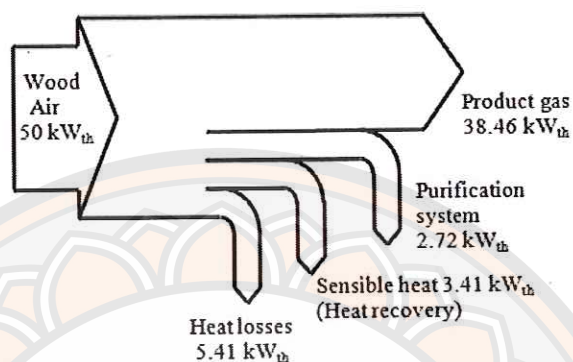


Figure 1 Sankey diagram for energy balance of a $50 \text{ kW}_{\text{th}}$ downdraft gasifier

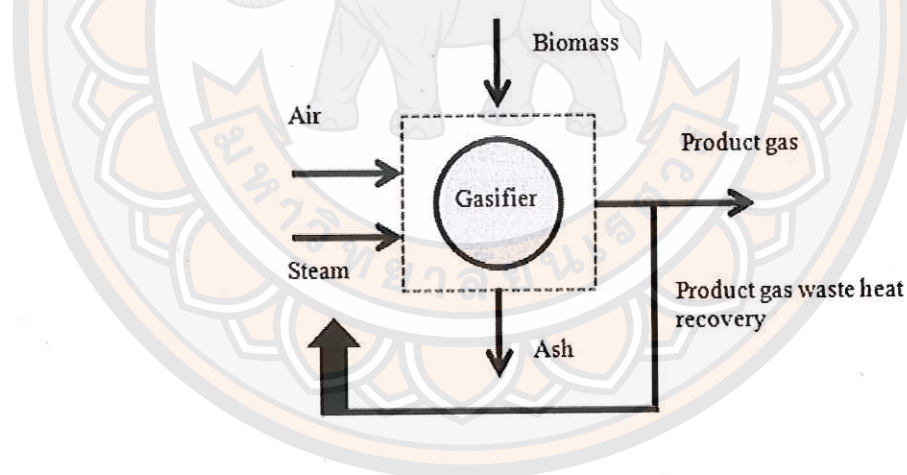


Figure 2 A scheme of the steam generator using waste heat from the same reactor

According to the figure 1, the energy balance of the gasifier system indicated that waste heat is $3.41 \text{ kW}_{\text{th}}$ or 6.8% of total energy input. It can be reused by using the heat exchanger.

Purpose of the study

1. To design and evaluate the system efficiency of the hydrogen-rich gas production from biomass air/steam gasification by using heat recovery in a downdraft gasifier.
2. To compare the economic evaluation between biomass air gasification and biomass air/steam gasification by using heat recovery in a downdraft gasifier.

Scopes of the study

This study started with a technical performance evaluation.

1. The technical performance of biomass air/steam, gasification would be to evaluate a 50 kW_{th} biomass downdraft gasifier, combined with counter flow heat exchanger system and gas cleaning system. Wood chip ranging from 2 cm x 4 cm x 4 cm. is used for feeding in the system and the moisture content should not more than 25%.
2. The parameters of economics study are cost of energy (COE), payback period (PB), net present value (NPV) and internal rate of return (IRR).

Expected Benefits

1. This prototype can be scaled up for both thermal and electrical applications.
2. Originality of this research can be a basic knowledge for further development and other researches of biomass gasification.
3. High quality of producer gas will increase the system efficiency and affect the reduction in the initial, maintenance and feedstock cost.

CHAPTER II

REVIEW OF RELATED LITERATURE AND RESEARCH

Introduction

In this chapter, experimental studies for the biomass air-steam gasification are reviewed, including the method to analyze the gasification path. Various studies were performed to obtain the air-steam gasification product composition, and to compute the energy and the system efficiencies.

Solid biomass fuel reacts with a limited supply of air to completely convert all the carbonaceous raw material into the fuel gas [17]. Thus, thermo chemical characteristics of biomass plays a major role in the selection of the design and performance of the gasification system. This combustible gas is composed of hydrogen, H₂ carbon monoxide, CO methane, CH₄ and a very small amount of heavy hydrocarbons.

Gasification is a high temperature chemical process that uses heat, air, pressure, steam, and often oxygen to convert any carbonaceous raw material into product gas or synthesis gas. Syngas, for short, is primarily composed of the flammable gases carbon monoxide (CO), hydrogen (H₂), methane (CH₄) and a very small amount of heavy hydrocarbons [18].

The syngas has a variety of usage and can be further converted to nothing but hydrogen and CO₂ by adding steam and reacting over a catalyst in a water-gas-shift reactor. When hydrogen is burned, it creates nothing but heat and water, resulting in the ability to create power or electricity with no carbon dioxide in the exhaust gases. Most importantly, hydrogen-enriched syngas can be used to make gasoline and diesel fuel by F-T process [17].

Gasification is a partial oxidation process. The term partial oxidation is a relative term, which simply means that less oxygen is used in the gasification than it would be required for the combustion of the same amount of fuel. Gasification typically uses only 25 to 40 percent of the theoretical oxidant (either pure oxygen or air) to generate enough heat to gasify the remaining unoxidized fuel, and to product

gas [19]. The major combustible products of gasification are carbon monoxide (CO), and hydrogen (H₂), with only a minor amount of the carbon that are completely oxidized to carbon dioxide (CO₂) and water. The heat released by partial oxidation provides most of the energy needed to break up the chemical bonds in the feedstock, to drive other endothermic gasification reactions, and to increase the temperature of the final gasification products [20].

Gasification chemistry

The thermochemical reactions of gasification can be progressed up to different extents depending on the gasification conditions (for instance, temperature and pressure) and the feedstock used. Combustion reactions take place in a gasification process, but, in comparison with conventional combustion, which uses a stoichiometric excess of oxidant, gasification typically uses one-fifth to one-third of the theoretical oxidant. This only partially oxidizes the carbon feedstock. As a "partial oxidation" process, the major combustible products of gasification are carbon monoxide (CO) and hydrogen, with only a minor portion of carbon that are completely oxidized to carbon dioxide (CO₂). The heat produced by the partial oxidation provides most of the energy required to drive the endothermic gasification reactions [21].

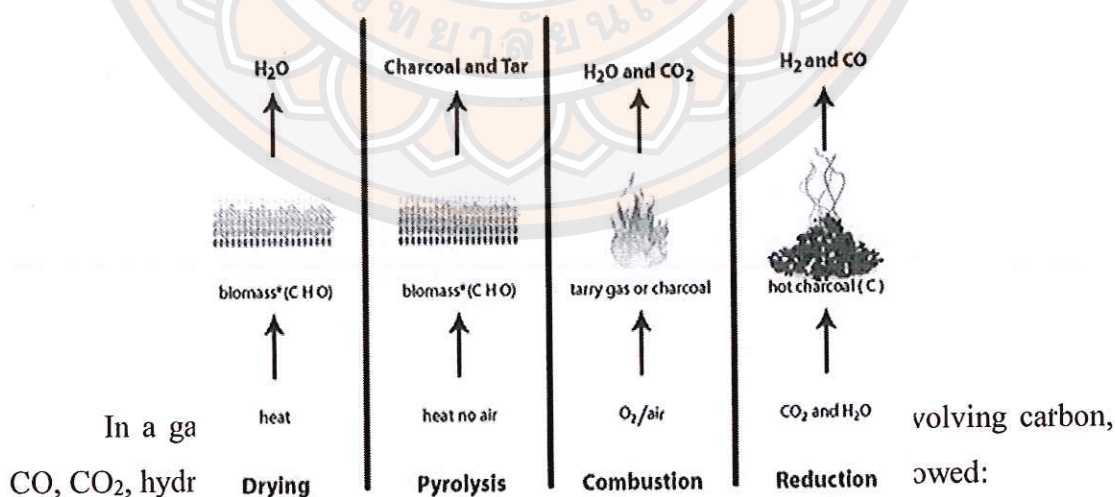


Figure 3 The four distinct processes take place in gasifier [22]

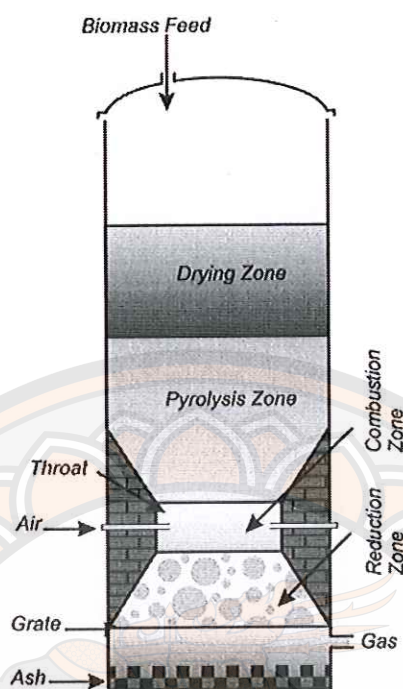


Figure 4 Cut-away view of gasifier [23]

Gasification Process

Basically four distinguishable stages occur inside a gasifier.

1. Drying

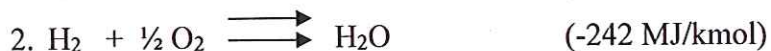
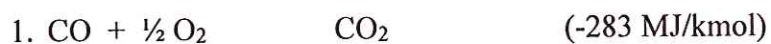
Fuel is loaded at the top of the gasifier and the drying of this biomass fuel is taking place at the most top section of the gasifier with the aid of a heat transferred from the lower part of the gasifier, resulting in water vaporization together with water vapor formed at combustion zone, partly leads to the production of hydrogen and the remaining goes with product gas.

2. Pyrolysis

Dry biomass then undergoes an endothermic reaction called pyrolysis, which decomposes the biomass fuel releasing its volatile materials in liquid and gaseous forms. The remaining is called a char.

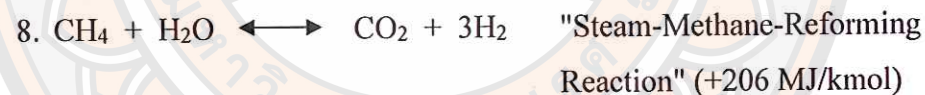
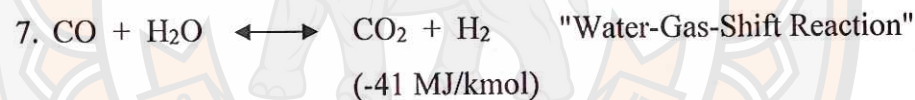
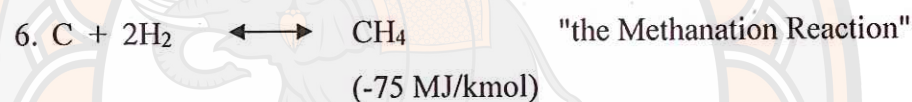
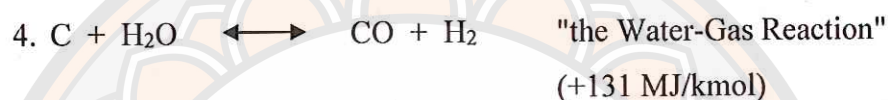
3. Combustion/Oxidation

At the level where oxygen is introduced to the gasifier, highly exothermic oxidation reactions happened.



3. Reduction

Oxidation products undergo several reduction processes converting sensible heat of the gases and charcoal in to chemical energy of the product gas as followed.



With the above, the three heterogeneous reactions (reactions 4 to 6) can be reduced into two homogeneous gas phase reactions of water-gas-shift and steam methane-reforming (reactions 7 and 8), which collectively play a key role in determining the final equilibrium product gas composition [13].

Table 1 Variation of gas composition with different gasification agents. [26]

Component Vol%, dry	Oxidizing agents		
	Air	Steam	Enriched air (80%O ₂)
H ₂	9-20	35-50	9-17
CO	10-20	25-47	40-50
CH ₄	1-8	14-25	< 1
CO ₂	10-20	9-15	19-25
N ₂	40-55	2-3	15-30
Net calorific value (MJ/Nm ³ , dry)	4-6.5	12-17	7-9

Types of Gasifier

1. Fixed bed gasifier

In this type of gasifier, air and gas pass up or down through a bed of solid fuel. These are the simplest type of gasifier and hence is suitable for small scale applications. According to the pathway of air and gas, fixed bed gasifier is further divided in to three categories.

1.1 Updraft gasifier

Updraft gasifier is the simplest type of gasification reactor. The process takes place in a reactor. The biomass wood chip is loaded from the top and then they move downwards. The reactive gas (air, steam, oxygen) comes from the bottom and goes upward in the opposite direction to the direction of the descending feedstock. The feedstock material reacts with the gasification agent producing the product gas, which leaves the gasifier through the top. The temperature in the combustion zone may reach 1200°C. The disadvantages of the updraft gasifiers are the relatively low efficiency and the higher tar production [23].

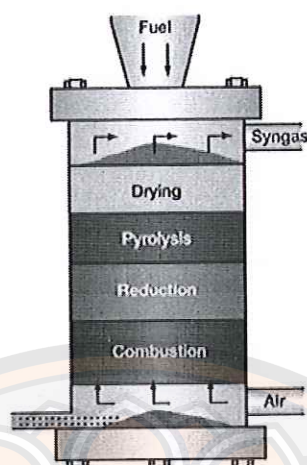


Figure 5 Schematic diagram of an updraft gasifier. [23]

However, there are several advantages of this type including the simplicity, the low gas exit temperature due to internal heat exchange and higher efficiency.

1.2 Downdraft gasifier

The biomass wood chip is loaded from the top and moves downwards. The reactive gas (air, steam, oxygen) comes from the bottom and goes downward in the same direction as that of the descending feedstock. The feedstock material reacts with the gasification agent producing the product gas, which leaves the gasifier through the bottom. In its way downwards the feedstock passes in series of several zones: the drying zone, the pyrolysis zone where the organic substance are converted into product gas and char (gasification) zone where the oxidizing gases are reduced in the reaction with carbon, the combustion (oxidizing) zone where the residual carbon is oxidized by the oxygen from the gasification agent. The temperature in the combustion zone may reach 1200°C. The advantages of downdraft gasifiers are the low tar [23].

According to the design, the pyrolysis products are passed through the glowing bed of charcoal and the tar is cracked in to gaseous products including CO₂, CO, H₂ and CH₄. Hence, this type of gasifier is suitable for highly volatile fuels such as wood, for the production of gas with low tar content. Therefore, it is very much suitable for power generation and internal combustion engine applications which

require clean gas. However, several drawbacks of this type of systems can be identified as the limitation to operation with low-density fuels due to flow problems and the excessive pressure drop, the slogging of ash and the lower efficiency compared to updraft type

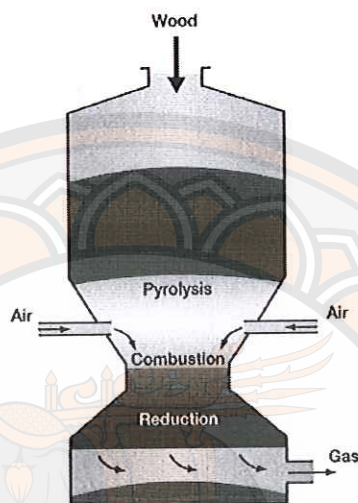


Figure 6 Schematic diagram of a downdraft gasifier [23]

due to lack of internal heat exchange. Ashes result from gasification are then removed by a rotating grate at the bottom and the product gas is obtained at a position depending on the type of the design. According to the literature average composition of product gas is given in Table 2.

Table 2 Gas composition of a typical downdraft gasifier. [23]

Component Composition	%
H ₂	12 - 20
CO	17 - 22
CH ₄	2 - 3
CO ₂	9 - 15
N ₂	50 - 54
Calorific value	5 - 5.9 MJ/Nm ³

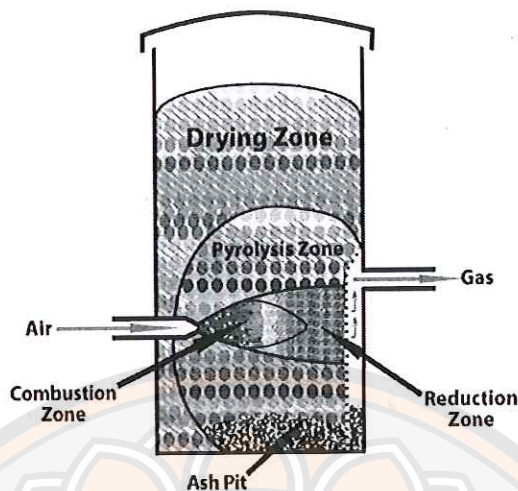


Figure 7 Schematic diagram of a cross-draft gasifier [23]

1.3 Cross-draft gasifier

The cross-draft gasifier is shown in Figure 7. Air enters at high velocity through a water-cooled nozzle mounted on one side of the firebox, induces substantial circulation and flows across the bed of fuel and char. The gas is produced in the horizontal direction in front of the nozzle and passes through a vertical grate into the hot gas port on the opposite side. This produces very high temperature in a very small volume and results in the production of very low tar gas. However, the cross draft gasifier is not commonly used [24].

2. Moving bed gasifier

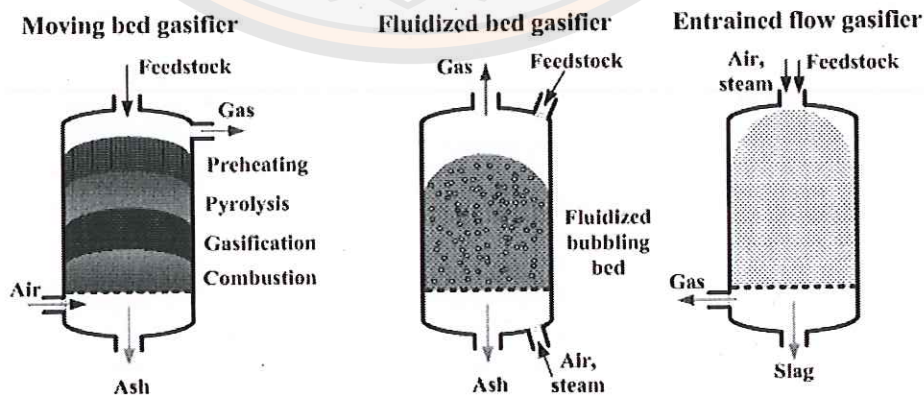


Figure 8 Different designs of moving bed gasifiers [24]

2.1 Fixed (moving) bed gasifier is the simplest type of gasification reactor. The process takes place in a shaft-type reactor. The coarse pieces of tires are loaded from the top and move downwards. The reactive gas (air, steam, oxygen) comes from the bottom and goes upwards in the opposite direction to that of the descending feedstock. The feedstock material reacts with the gasification agent producing the product gas, which leaves the gasifier through the top. In its way downwards the feedstock passes in series of several zones: the drying zone, the pyrolysis zone where the organic substance are converted into carbon black, the reduction (gasification) zone where the oxidizing gases are reduced in the reaction with carbon, the combustion (oxidizing) zone where the residual carbon is oxidized by the oxygen of the gasification agent. The temperature in the combustion zone may reach 1200°C. The disadvantages of fixed bed gasifiers are the relatively low efficiency and the possibility of the oxidant to break through forming channels in the feedstock and reacting with the gaseous product in form of explosion [23].

2.2 Fluidized bed gasifier does not have different zones. The chemical reactions occur in the isothermal dispersion of the fine feedstock particles in the gas. The feedstock particles are mixed with the gases introduced from the bottom of the reactor. The mixture is in a liquid-like form. The gaseous product goes upwards passing through the fluidized bed and leaves the gasifier through the top. Some of the ash particles are taken by gas and they are separated away from it in a cyclone or filters. The temperature of the fluidized bed process is lower than that of the fixed bed: 700-900°C [25].

2.3 Entrained flow gasifier operates with the feedstock particles fed through the top of the reactor. The gasification agent (oxidizing gas) also enters the gasifier through the top. Entrained flow process occurs at high temperature, therefore, it takes very short time (within few seconds). High temperature also causes the ash to melt forming a slag, which is removed from the bottom of the gasifier. The gaseous product is clean and it needs no additionally cleaned [25].

Gas Cleaning and Cooling

Gas cooling prior to engine application is very much essential to improve the volumetric efficiency and also to condense the tarry liquids. Direct wet scrubbing and indirect water-cooling are the two possible methods that are commonly used. Heat

exchanger can also be used to preheat the incoming air while cooling down the product gas. Gas cleaning systems basically include cyclone separators, biomass filters and/or bag filters for dust removal and water separators for possible water drops [26].

Equivalence ratio (ER)

Equivalence ratio is the ratio of actual air-fuel ratio to the stoichiometric air-fuel ratio. The theoretical gasification occurs between ER values of 0.19-0.43 [27]. The theoretical optimum point for gasification is near 0.25 ER. Below 0.25, the char remains and some energy losses through char. At higher ER, some gas is burned and the temperature inside the gasifier increases. At ER = 0.25 all the char is converted into product gas giving the highest energy of the product gas. Studies have been reported on how the performance of gasifier varies with equivalent ratio (Zainal, et al., [49] Pratik, et al., [54] and Ummadisingu, et al., [28]). In these studies they have studied the change of gas calorific value, cold gas efficiency and gas production rate with equivalent ratio. The calorific value was found to be increasing with the ER, but tends to reduce after a certain critical value. Cold gas efficiency varies in the same pattern, giving the maximum at maximum calorific value. On the other hand, gas production rate per unit weight of biomass was found to be increasing with the ER. The experimental observations of Sharma, 2011 concludes that any factor resulting in higher reaction temperatures due to energeticity of gasification reactions (increase of air/fuel ratio) or operating conditions (increase of gas flow rate) gives better gasifier performance. Equivalent Ratio reflects the combined effect of air flow rate and fuel flow rate. This is defined as the ratio of operating air-fuel ratio to Stoichiometric air-fuel ratio [29].

$$ER = \frac{\text{Operating or actual } \left(\frac{A}{F}\right)_o}{\text{Stoichiometric } \left(\frac{A}{F}\right)_s} \quad [\text{Eq.1}]$$

$$\left(\frac{A}{F}\right)_o = \frac{\text{Mass flow rate of air}}{\text{Fuel wood consumption rate}} = \left(\frac{A}{G}\right) \left(\frac{G}{F}\right) * \text{Density of air} \quad [\text{Eq.2}]$$

Stoichiometric air-fuel ratio is taken as 6.36 kg of air per kg of wood [24].

Biomass Fuels

Charcoal, wood, wood residues, agricultural residues and peat are some biomass fuels commonly used for gasification. Chemical, physical and morphological property differences of these fuels demand different gasification technologies or gasifier designs in order to have a smooth functioning of the system. The most important fuel properties can be identified as followed for stable and efficient operation of a gasifier with low pressure drop and production of high quality gas [24].



Figure 9 Various types of biomass feedstock.

1. Moisture content

High moisture content of fuel reduces the thermal efficiency of gasifier since some heat is wasted in order to drive off the moisture which is otherwise used in the reduction phase in converting thermal energy in to chemical energy or heating value of gas.

2. Volatile matter content

High volatile matter content of fuel demands special design of gasifier or cleaning system in order to remove tars from product gas when used in engine applications.

3. Ash content

Melting or agglomeration of ash results in slagging or clinker formation. This adds much labour work and also excessive tar formation or even blocking the gasifier with the risk of explosion. The use of moving grates has added the advantage of the ability to the operation with fuels having high ash content without slogging problem.

4. Bulk density

Fuels with high bulk density contain high energy content per unit volume and also require less space in fuel hopper. When the bulk density of fuel is low, it is

difficult to flow under gravity and this could result in low heating value of gas. To overcome limitations of the fuel properties above fuel properties, suitable pretreatment of fuel is desired. Generally, pretreatment involves mechanical chipping for size reduction, screening to ensure that uniform size distribution, drying for moisture removal and densification for low bulk density fuels.

Gasification Agents

Gasification agent is the means of supplying oxygen in to the gasifier.

1. Air gasification

Most common method of gasification is using air as gasification agent. This method is straight forward and very simple, requiring less capital and operating cost. However, the presence of inert nitrogen in air dilutes the gas, hence lowers the calorific value per unit volume of gas.

2. Oxygen gasification

Oxygen gasification can be achieved by removing nitrogen from air prior to supplying to the gasifier. This involves some additional cost, but it could avoid gas dilution problem previously mentioned and could result in medium level of energy content of gas per unit volume.

3. Steam gasification

This is a highly endothermic process. The heat needed should be supplied by external heat source or by partial oxidation of fuel. Partial oxidation of fuel is achieved by mixing steam with air or oxygen. This method produces gas with higher energy content when compared to previous methods.

4. High temperature air/steam gasification

This novel method, with the increase of physical enthalpy of gasification agent, ensures economical and environmental benefits over all methods and attracts more attention nowadays. The average product gas composition (vol. %) with different gasification agents are given in Table 3. [26].

Gasification Efficiency

The gasification efficiency is an important parameter determining the actual operation, as well as the economic feasibility of using a gasifier system. A useful definition of the gasification efficiency if the gas is used for direct burning or thermal applications is [26]:

$$\eta_g = \frac{\text{Heating value of gas} \times \text{gas flow rate}}{\text{Heating value of fuel wood} \times \text{fuel consumption rate}} \quad [\text{Eq.3}]$$

or

$$\eta_{th} = \frac{(H_g \times Q_g) + (Q_g \times \rho_g \times C_p \times T)}{H_s \times M_s} \times 100 \% \quad [\text{Eq.4}]$$

In which:

η_{th}	=	gasification efficiency (%) (thermal)
H_g	=	heating value of the fuel gas (kJ/Nm ³)
Q_g	=	volume flow rate of fuel gas (Nm ³ /s)
H_s	=	lower heating value of gasifier fuel (kJ/kg)
M_s	=	gasifier solid fuel consumption (kg/s)
ρ_g	=	density of the fuel gas (kg/Nm ³)
C_p	=	specific heat of the fuel gas (kJ/kg ^o K)
T	=	fuel gas temperature at the gasifier outlet (^o K).

Routes for hydrogen rich gas from biomass

As the demand of energy is growing faster due to society up gradation and rapid industrialization, renewable energy sources hold a promise for the future energy resources in developed countries and third world developing countries. In contrast to conventional energy sources, non-conventional energy sources like wind, sunlight, water, and biomass have been used as renewable energy sources since ancient times. Worldwide research and energy policy are focusing towards the hydrogen economy.

Hydrogen is considered, as a forecast, to become a major source of energy in the future. Hydrogen production plays an important role in the development of economy. As biomass is abundant, clean and renewable, production of hydrogen from biomass is one of the promising approaches [30].

Hydrogen rich gas produced by gasification technology as followed ;

1. Using air-steam or steam gasification
2. Use suitable catalyst gasification
3. Pure oxygen gasification
4. Pure oxygen with steam gasification

In this research, using steam as a catalyst for more hydrogen. Because water is cheap and easy to find and the heat exchanger system is not complicated. Although it is less effective than other methods, it is no less interesting. Research indicates that the system is more efficient than conventional syngas production for about 10%.

Heat Exchanger Design

The heat exchanger is a thermal heat transfer device that exchanges the thermal energy from one source and transfers it to another at different temperatures. In most heat exchanger designs, the fluids or gases used to transfer the heat are separated and do not mix [31].

The Heat Exchanger is a passive corrugated mass of metal which transfers the heat from one working fluid to another. The primary thermal fluid absorbs heat from a heat source and then circulates through the heat exchanger where the heat is removed from the fluid (either water or gas) and is transferred to a secondary fluid, again either water or a gas that circulates and dissipates the heat (heat sink) into home or the atmosphere.

There are many heat exchanger designs available to choose from the following: tubular, double-pipe, flat plate, spiral, and coil designs. The selection of a heat exchanger depends on many factors. Most heat exchangers are classified according to their construction, the heat transfer process and their surface compactness. That is the amount of surface area for the heat to dissipate from or transfer to, comparing to the physical size.

Types of Heat Exchanger

1. Shell and Tube Heat Exchanger
2. Double Pipe or Hairpin Exchanger
3. Flat Plate and Fin Exchangers
4. Radiators and Solar Exchangers
5. Spiral Heat Exchangers
6. Air Coolers, Chillers and Condensers

Shell and Tube Heat Exchanger

Shell and tube heat exchanger is one of the most common types of exchangers used in heat transfer. It is typically used in applications when a process requires large amounts of fluid to be heated or cooled. As a result of their design, shell and tube heat exchangers offers a large surface area, thus resulting in high heat transfer efficiency.

There are many different types or designs of shell and tube heat exchangers to meet various process requirements in almost every industry. Shell and Tube heat exchangers can provide reliable heat transfer by utilizing multiple passes of one or both fluids. One (1), two (2) and four (4) pass models are usually the standard, however multi-pass custom models are available [31].

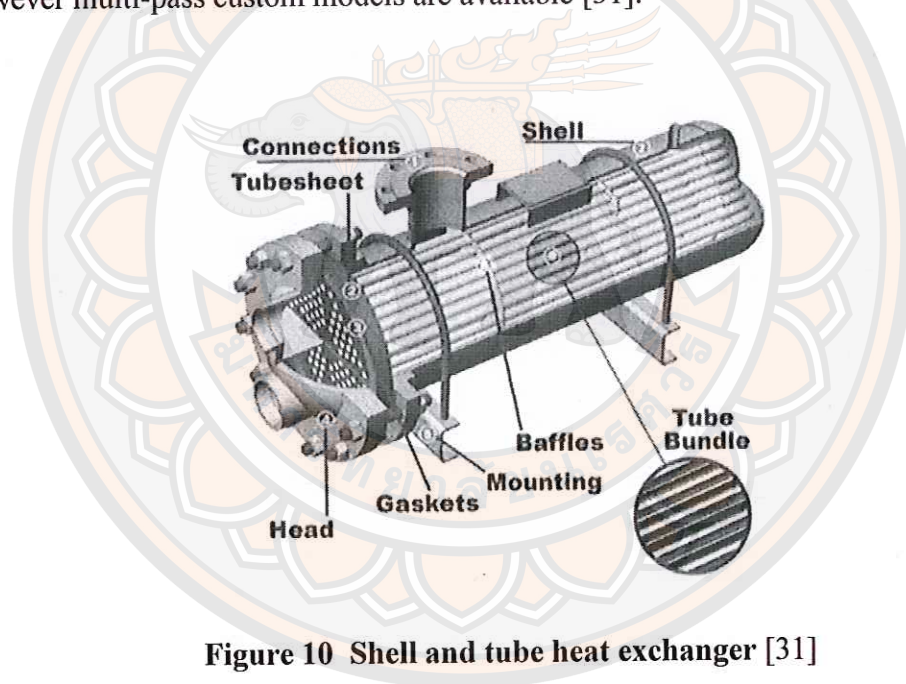
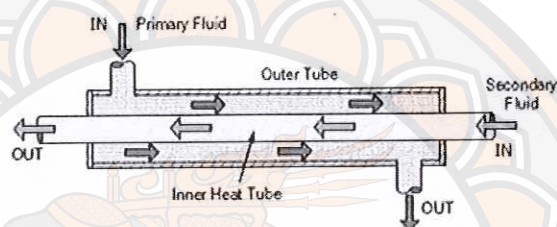


Figure 10 Shell and tube heat exchanger [31]

Tubular heat exchanger is by far the simplest design. The primary fluid circulates through straight or concentric pipes in the form of U-shaped tube design. These primary tubes are encased within an outer sealed tube that circulates the secondary fluid. Their usual application is in small heat water-to-water transfer applications. The advantage of this type of design is the flexibility since tubular heat exchangers can be added or removed as required. Also any number of heat exchangers can be connected together in series or parallel combinations.



Tubular heat pipe configurations of inner tubes



Figure 11 Shell-and-tube type with bare or finned tubes, vertical or horizontal, parallel and counter flow [31]

While being very simple and basic, the effectiveness of this type of single pass heat exchanger design can be increased by having the directional flow of the secondary fluid in the opposite direction to the primary flow to improve heat absorption and efficiency. If both the primary and secondary fluids flow in the same direction, then this is called "parallel-flow". If the primary and secondary fluids flow in the opposite direction, then this is called "counter-flow". Also the inner heat pipe can be either a single bare tube, fitted with fins to increase the surface area, or as a multi-tube design as shown.

Shell and tube heat exchangers consist of series of tubes. One set of these tubes contains the fluid that must be either heated or cooled. The second fluid runs over the tubes that are being heated or cooled so that it can either provide the heat or absorb the heat required.

Classification of Heat Exchangers by Flow Configuration

There are three basic flow configurations:

1. Counter flow
2. Cocurrent or parallel flow
3. Cross flow

The most common arrangements for flow paths within a heat exchanger are counter-flow and parallel flow. A counter-flow heat exchanger is the one, in which the direction of the flow of one of the working fluids is oppose to the direction of the flow of the other fluid. In a parallel flow exchanger, both fluids in the heat exchanger flow in the same direction [32].

Figure 12. illustrates an idealized counterflow exchanger in which the two fluids flow parallel to each other but in opposite directions. This type of flow arrangement allows the largest change in temperature of both fluids and is therefore the most efficient (where efficiency is the amount of actual heat transferred compared with the theoretical maximum amount of heat that can be transferred).

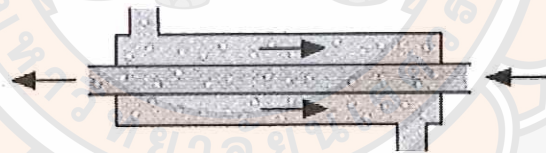


Figure 12 Counter-flow [33]

In cocurrent flow or parallel heat exchangers, the streams flow parallel to each other and in the same direction as shown in Figure 13. This is less efficient than the countercurrent flow but does provide more uniform wall temperatures.

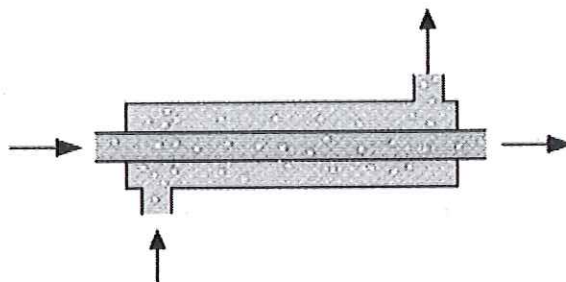


Figure 13 Parallel flow [33]

Cross flow heat exchangers are intermediate in efficiency between countercurrent flow and parallel flow exchangers. In these units, the streams flow at right angles to each other as shown in Figure 14.

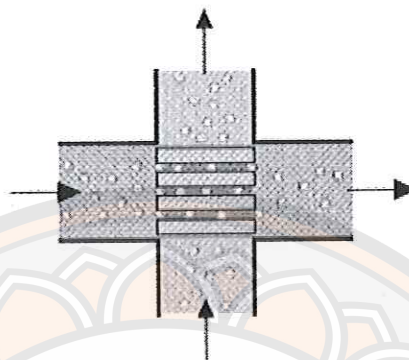
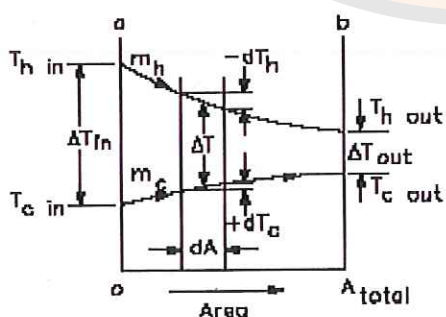


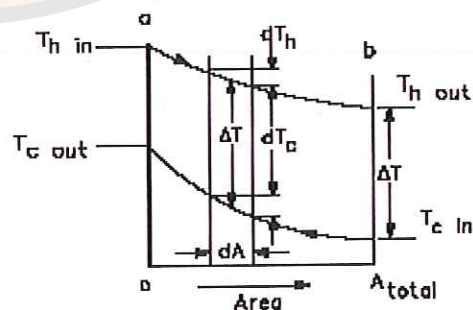
Figure 14 Cross-flow [33]

For efficiency, heat exchangers are designed to maximize the surface area of the wall between the two fluids, while minimizing the resistance to fluid flow through the exchanger. The exchanger's performance can also be affected by the addition of fins or corrugations in one or both directions, which increase surface area and may channel fluid flow or induce turbulence [59].

The driving temperature across the heat transfer surface varies with position, but an appropriate mean temperature can be defined. In the most simple systems this is the "log mean temperature difference" (LMTD). Sometimes direct knowledge of the LMTD is not available and the NTU method is used. The temperature profiles of the two heat exchangers indicate two major disadvantages in the parallel-flow design.



Temperature distribution in parallel flow heat exchanger



Temperature distribution in counter-flow heat exchanger

Figure 15 Heat exchanger temperature profiles [33]

thermal stresses. The opposing expansion and contraction of the construction materials due to diverse fluid temperatures can lead to eventual material failure. Second, the temperature of the cold fluid exiting the heat exchanger never exceeds the lowest temperature of the hot fluid. This relationship is a distinct disadvantage if the design purpose is to raise the temperature of the cold fluid.

The design of a parallel flow heat exchanger is advantageous when two fluids are required to be brought to at nearly the same temperature.

The counter-flow heat exchanger has three significant advantages over the parallel flow design. First, the more uniform temperature difference between the two fluids minimizes the thermal stresses throughout the exchanger. Second, the outlet temperature of the cold fluid can approach the highest temperature of the hot fluid (the inlet temperature). Third, the more uniform temperature difference produces a more uniform rate of heat transfer throughout the heat exchanger.

Whether parallel or counter-flow, heat transfer within the heat exchanger involves both conduction and convection. One fluid (hot) convectively transfers heat to the tube wall where the conduction takes place across the tube to the opposite wall. The heat is then convectively transferred to the second fluid. Since this process takes place over the entire length of the exchanger, the temperature of the fluids as they flow through the exchanger is not generally constant, but varies over the entire length, as indicated in Figure 15. The rate of heat transfer varies along with the length of the exchanger tubes because its value depends on the temperature difference between the hot and the cold fluid at the point being viewed [33].

Heat Exchanger Fundamentals: Calculation

Heat exchange is a natural phenomenon occurring throughout our environment. It drives the weather cycles and energy exchange between ecosystems. Harnessing its utility through accurate control of heat exchange has been a focus of our industry for over a century.

Below are parametric thermodynamic equations that define the nature of heat exchange and the performance of a heat exchanger for any given application. Once these thermal parameters are determined that they can be used to calculate heat exchanger performance in order to select the most suitable product based on the specific application.

The heat transfer principal in heat exchangers is based on a colder fluid gaining heat from a relatively hotter fluid separated by, and flowing over, a heat conductive material. The theoretical heat transfer is expressed by the following formula:

$$\dot{Q}_t = \dot{m} C_p \Delta T \quad [\text{Eq.5}]$$

where

- \dot{Q}_t = Total heat load, watt
- \dot{m} = Mass flow rate of fluid, kg/s
- C_p = Specific heat of fluid at constant pressure, $\text{W/m}^2 \cdot ^\circ\text{C}$
- ΔT = Change in temperature of the fluid, $^\circ\text{C}$

This formula provides the theoretical heat yield to or from a given fluid undergoing a temperature change, ΔT at a mass flow rate, \dot{m} with the fluid's specific heat property, C_p .

The theoretical heat yield of a fluid gives the amount of heat that is needed to be transferred into or from a fluid. The practical heat transfer is a function of the physical geometry of the heat exchanger, its material composition, and the fluid condition.

The general form of the equation defining the maximum potential heat transfer through a heat exchanger is expressed by the formula:

$$\dot{Q}_p = U \times A \times \text{LMTD} \quad [\text{Eq.6}]$$

where

$$\begin{aligned} \dot{Q}_p &= \text{Maximum potential heat transfer, watt} \\ U &= \text{Overall heat transfer coefficient, W/m}^2\cdot\text{K} \\ A &= \text{Surface area, m}^2 \\ \text{LMTD} &= \text{Logarithmic mean temperature difference, }^\circ\text{C} \end{aligned}$$

The practical heat transfer control is determined by the molecular thermodynamic interactions between the fluids flowing through the heat exchanger and the geometry of the heat exchanger itself. The overall U value is calculated by an equation specific to the geometric configuration of a heat exchanger. The overall U value is calculated over the total surface area A of the heat exchanger, across which the fluids exchange heat [34].

The log mean difference of the inlet and outlet temperatures (LMTD) of the hot and cold fluids for a counter flow exchanger is expressed by the formula:

$$\text{LMTD} = \frac{[(T_{hi} - T_{co}) - (T_{ho} - T_{ci})]}{\ln\left[\frac{(T_{hi} - T_{co})}{(T_{ho} - T_{ci})}\right]} \quad [\text{Eq.7}]$$

where

$$\begin{aligned} T_{hi} &= \text{Inlet temperature of hot fluid, }^\circ\text{C} \\ T_{co} &= \text{Outlet temperature of cold fluid, }^\circ\text{C} \\ T_{ho} &= \text{Outlet temperature of hot fluid, }^\circ\text{C} \\ T_{ci} &= \text{Inlet temperature of cold fluid, }^\circ\text{C} \end{aligned}$$

The overall heat transfer coefficient

The overall heat transfer coefficient for a wall or a heat exchanger can be calculated as [34]:

$$1/UA = 1/h_1A_1 + dx_w / kA + 1/h_2A_2 \quad [\text{Eq.8}]$$

where

- U = the overall heat transfer coefficient (W/m²K)
- A = the contact area for each fluid side (m²)
- k = the thermal conductivity of the material (W/mK)
- h = the individual convection heat transfer coefficient for each fluid (W/m²K)
- dx_w = the wall thickness (m)

Heat exchanger effectiveness, ϵ

Heat exchanger effectiveness is the ratio of the actual heat transfer to the maximum heat that can be possibly transferred [34]:

$$\epsilon = \frac{\text{Actual heat transfer rate in any heat exchanger}}{\text{Maximum possible heat transfer rate in a counter-flow heat exchanger}} \quad [\text{Eq.9}]$$

or

$$\epsilon = \frac{q}{q_{max}} \quad [\text{Eq.10}]$$

Waste Heat Recovery Unit

A waste heat recovery unit (WHRU) is a heat exchanger that recovers the heat from a hot gas stream while transferring it to a working medium, typically water or oils. The hot gas stream can be the exhaust gas from a gas turbine or a diesel engine or a waste gas from industry or refinery. Heat Recovery Steam Generation, HRSG, Figure 16.

\dot{Q}_{HE} = Rate of heat transfer from hot gas to water by heat exchanger

$\dot{m}_{water} C\Delta T$ = Rate of sensible heat addition of water

$\dot{m}_{steam} h_{fg}$ = Rate of latent heat addition of vaporization of saturated steam

Energy balance

$$\dot{Q}_{HE} = \overset{\text{Sensible heat}}{\dot{m}_{water} C\Delta T} + \overset{\text{Latent heat}}{\dot{m}_{steam} h_{fg}} \quad [\text{Eq.11}]$$

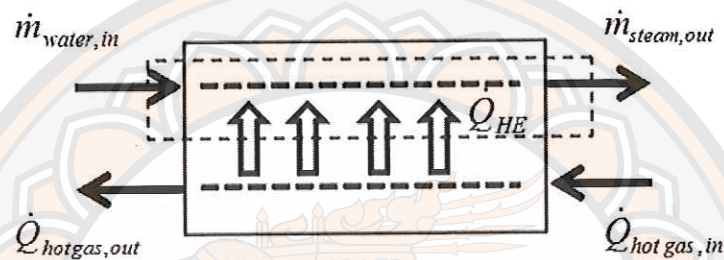


Figure 16 Heat recovery steam generation, HRSG.

Air/steam gasification by using waste heat

The gasification process requires a gasifying agent that provides oxygen for the formation of CO from solid fuel. The gasifying, or oxidizing, agents are air, oxygen, steam and CO₂. The most common agent is air because of its availability at zero cost. Air, though cheap, is not a perfect agent because of its nitrogen content. The product gas from air gasification has a generally low heating value of 4-7 MJ/Nm³. Oxygen gasification produces a higher heating value (10-18 MJ/Nm³) but has a drawback due to the high production cost of oxygen. Steam is another alternative. It also generates a medium calorific value gas (10-14 MJ/Nm³) and moreover increases the hydrogen content of the product gas. The heat required for the reaction has to be transferred either by partial combustion in the same reactor—mixing H₂O with oxygen/air.

Steps for design and calculation of the heat exchanger.

1. Assume tube diameter, D and tube length, L
2. Assume material of construction for the tubes and shell, thermal conductivity
3. From energy balance analysis ; assume four known temperature and find one of the shell or tube side flow rate. Use the heat duty equation

$$\dot{Q} = \dot{m}_c C_{pc} (T_{co} - T_{ci}) = \dot{m}_h C_{ph} (T_{ho} - T_{hi}) \quad [\text{Eq.12}]$$

where subscripts c and h refer to cold and hot streams. Then obtain the heat duty, \dot{Q}

4. Based on the type of flow (counter-flow), calculate Log Mean Temperature Difference, LMTD.

$$\text{LMTD} = \frac{[(T_{hi} - T_{co}) - (T_{ho} - T_{ci})]}{\ln \left[\frac{(T_{hi} - T_{co})}{(T_{ho} - T_{ci})} \right]} \quad [\text{Eq.13}]$$

5. Based of the exchanger configuration obtain the temperature correction factor. ; assume F_t

6. Calculate the mean temperature difference $\Delta T_m = F_t \times \text{LMTD}$ using

7. Calculate the overall heat transfer coefficient using equation below ;

$$\frac{1}{UA} = \frac{1}{h_i A_i} + \frac{dx_w}{kA} + \frac{1}{h_o A_o} \quad [\text{Eq.14}]$$

8. Calculate the provisional area,

$$A = \frac{\dot{Q}}{U \cdot \Delta T_m} \quad [\text{Eq.15}]$$

9. Assume fouling factor.
10. Determine the pressure drop.
11. Find the number of tube.

HHV and LHV heating value of product gas

The heating value of the product gas was evaluated in terms of higher heating value, HHV and lower heating value, LHV, at standard temperature and pressure and can be determined by considering the volumetric percentage of the gas constitutes (CO, H₂ and CH₄), as shown below [35].

Heating value (LHV) of product gas is determined from the chemical composition of the gas and LHV of individual components.

$$(\text{LHV})_{\text{Gas}} = \sum \text{volume\% of component} \times \text{LHV of the component} \quad [\text{Eq.16}]$$

$$\text{HHV} = (\text{H}_2\% \times 30.52 + \text{CO}\% \times 30.18 + \text{CH}_4\% \times 95.0) \times 4.2 \quad [\text{Eq.17}]$$

$$\text{LHV} = (\text{H}_2\% \times 25.70 + \text{CO}\% \times 30.00 + \text{CH}_4\% \times 85.5) \times 4.2 \quad [\text{Eq.18}]$$

$$\begin{aligned} \text{LHV (MJ/Nm}^3) &= (1/1000)(\text{CO} \times 126.36 + \text{H}_2 \times 107.98 \\ &\quad + \text{CH}_4 \times 358.18 + \text{C}_2\text{H}_2 \times 56.002 \\ &\quad + \text{C}_2\text{H}_4 \times 59.036 + \text{C}_2\text{H}_6 \times 63.772) \end{aligned} \quad [\text{Eq.19}]$$

Economic evaluation

1. Net present value, NPV

Net present value (NPV) is a formula used to determine the present value of an investment by the discounted sum of all cash flows received in the project. The formula for the discounted sum of all cash flows can be rewritten as

$$\text{NPV} = -C_0 + \sum_{i=1}^T \frac{C_i}{(1+r)^i} \quad [\text{Eq.20}]$$

where

-C₀ = Initial investment.

C = Cash flow.

r = Discount rate.

T = Time period.

If the NPV of a project is positive, then the project will provide a net financial benefit for the company. Such an approach can be used when evaluating the energy management options, which will involve the initial capital expense and produced subsequent energy savings.

The positive NPV means a positive surplus indicating that the financial position of the investor will be undertaking the project. Obviously, a negative NPV would indicate a financial loss and that the project is rejected. The NPV of Zero would mean that the present value of all the benefits over the useful lifetime is equal to the present value of all the costs [36].

2. Cost of energy, COE

The cost of energy is one of the major data for a thermal power plant because it can show the price of the generated thermal energy. Expressed in Baht per kWh, it is the annualized cost of the thermal power plant over the annual thermal energy delivered to the consumers.

$$\text{COE} = \frac{\text{Annualized cost}}{\text{Annual energy production}} \quad [\text{Eq.21}]$$

3. Payback period

The Payback Period (PB) takes into account the time taken to payback the capital investment "C" on the basis of the equal annual payments, the annual payments being the estimated savings. The simple payback period is given by the equation 22.

$$T = C / S \quad [\text{Eq.22}]$$

where

C = Initial or capital investment

S = Estimated savings or periodic cash flow

T = Payback period in years

The payback period method is widely used in financial decision making because it is easy to understand and to use. It also emphasizes a rapid recovery of the initial investment, which may be a very important objective in various situations. However, all cash flows that would commonly occur after the computed payback period are neglected. Thus the payback period method makes no allowance for the projects with long gestation periods. The payback period formula is used to determine the length of time it will take to recoup the initial amount invested on a project or investment.

4. Internal rate of return, IRR

The internal rate of return (IRR) is a widely accepted discounted measure of investment worth and is used as an index of profitability for the appraisal of the project [36]. IRR is defined as the interest rate in which the NPV of an investment is zero.

$$\text{NPV(IRR)} = \sum_{j=0}^n \frac{B_j - C_j}{(1 + \text{IRR})^j} = 0 \quad [\text{Eq.23}]$$

IRR is widely used in the appraisal of projects because the IRR on a project is its expected rate of return. Furthermore, it employs a percentage rate of return as the decision variable, which suits the banking community. For situations in which IRR exceeds the cost of the funds used to finance the project, a surplus would remain after paying the capital.

In economic terms, the IRR represents the percentage rate of interest earned on the unrecovered balance of an investment. The unrecovered balance of an investment is the portion of initial investment that remains to be recovered after interest payments have been added and receipts have been deducted up to the desired point of time.

Literatures Review

There were many researches that involved in an application of hydrogen rich gas production. The detail of some researches will be described as the following.

C. Mandl, I. Obernberger, and F. Biedermann. [37] presented a one-dimensional steady state mathematical model for the simulation of a small scale fixed-bed gasifier. The model is based on a set of differential equations describing the entire gasification process of softwood pellets and is solved by a two-step iterative method. The main features of the model are homogeneous and heterogeneous combustion and gasification reactions, one-step global pyrolysis kinetics and drying, heat and mass transfer in the solid and gas phases as well as in between phases, heat loss, particle movement and shrinkage within the bed. The pyrolysis model has been improved by partially cracking primary tar into lighter gases according to experimental data. The model is used to simulate a laboratory scale of a fixed-bed updraft gasifier. Good agreement is achieved between the prediction and the measurements for the axial temperature profiles and the composition of the product gas. Moreover, results are presented for different air to fuel ratios and varying power inputs. The gasification process is improved by increasing the power input of the gasifier as a result of higher temperatures. Furthermore, a higher air to fuel ratio could lower the efficiency of the gasification process.

J.J. Hernández, G. Aranda, J. Barba, and J.M. Mendoza. [38] studied the effect of the addition of steam to air as gasifying agent in biomass entrained flow gasification, a promising technology due to its commercial availability, high efficiency and high potential for the production of biofuels and chemicals from biomass. Dealcoholised marc of grape (an abundant residue in the southern regions of Europe with interesting physical and thermochemical properties) has been used as fuel in all tests. Firstly, the steam content of the gasifying mixture has varied from 0% (air gasification) to 100% (steam gasification) by keeping the molar fuel/gasifying agent ratio (F) constant, in order to determine the role of air and steam in the gasification process. On the other hand, the effect of the steam/biomass (S/B) ratio has been determined both for steam and air-steam gasification. Lastly, the effect of the operating temperature has been compared with both for air and air-steam gasification. The performance of the gasification process has been evaluated through some

characteristic parameters, such as the product gas composition and heating value, the gas yield, the cold gas efficiency, and the product gas ratios and production. The results obtained show that there is an optimal range in the steam content of the gasifying agent (found in air–steam mixtures containing 40–70% mol steam) for which a trade-off between gas quality, gas production, and cold gas efficiency is reached. In general, the addition of steam proved to be positive for the process performance, not only because of the lower dilution of gas in N_2 from air, but also because of the promotion of the steam reforming and the WGS reactions. An increase in the operating temperature has different effects depending on the gasifying agent used. Thus, a higher temperature increases the CO and H_2 content in the product gas for air gasification, whereas air–steam gasification leads to a boost in the H_2 production at higher temperatures, as well as an increase in the CH_4 content.

Kentaro Umeki, Kouichi Yamamoto, Tomoaki Namioka, and Kunio Yoshikawa. [39] studied a high temperature steam gasification process to generate hydrogen-rich fuel gas from woody biomass. In this study, the performance of the gasification system, which employs only high temperature steam exceeding 1200 K as the gasifying agent, was evaluated in a 1.2 ton/day-scale demonstrative plant. A numerical analysis was also carried out to analyze the experimental results. Both 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. The H_2 fraction in the produced gas was 35–55 vol.% at the outlet of the gasifier. Under the experimental conditions, S/C ratio had a significant effect on the gas composition through the dominant reaction, water–gas shift reaction. The tar concentration in the produced gas from the high temperature steam gasification process was higher than that from the oxygen-blown gasification processes. The highest cold gas efficiency was 60.4%. However, the gross cold gas efficiency was 35%, when considering the heat supplied by high temperature steam. The ideal cold gas efficiency of the whole system with heat recovery processes was 71%.

Madhukar, R. Mahishia, and Goswami, D.Y. [40] predicted the chemical composition of the products of biomass gasification. The effects of temperature, pressure, steam biomass ratio (SBR) and equivalence ratio (ER) on the equilibrium hydrogen yield were studied. Gibbs energy minimization approach was used to

determine the product gas composition. Wood (designated by $\text{CH}_{1.5}\text{O}_{0.7}$) was used as the model biomass compound and Stanjan software was used. Gasifier, the most critical component of any biomass gasification system, was modeled as an equilibrium reactor and the energy consumption and thermodynamic efficiency were determined. The first law analysis of the gasifier showed that the optimum conditions for hydrogen production occurred at a gasification temperature of 1000 K, SBR of 3, ER of 0.1. Finally, equilibrium calculations were compared with experimental data from literature, which showing that for high gas residence times and high gasification temperatures, there is a close match of equilibrium results with experimental ones.

P. Plis, and R.K. Wilk. [41] investigated the process of air biomass gasification in a fixed bed gasifier. Theoretical equilibrium calculations and experimental investigation of the composition of syngas were carried out and compared with findings of other researchers. The influence of the excess air ratio and the parameters of biomass on the composition of syngas were investigated. A theoretical model is proposed, based on the equilibrium and thermodynamic balance of the gasification zone. The experimental investigation was carried out at a setup that consists of a gasifier connected by a pipe with a water boiler fired with coal ($50 \text{ kW}_{\text{th}}$). Syngas obtained in the gasifier is supplied into the coal-firing zone of the boiler, and co-combusted with coal. The moisture content in biomass and excess air ratio of the gasification process are the crucial parameters, determining the composition of syngas. Another important parameter is the kind of applied biomass. Despite similar compositions and dimensions of the two investigated feedstocks (wood pellets and oats husk pellets), compositions of syngas obtained in the case of these fuels were different. On the basis of tests it may be stated that oats husk pellets are not a suitable fuel for the purpose of gasification.

Felipe Centeno, Khamid Mahkamov, Electo, E. Silva Lora, and Rubenildo, V. Andrade. [42] predicted a steady state performance of a biomass downdraft gasifier/spark ignition engine power system is described. A mathematical model of the integrated system consists of two parts: the fixed bed downdraft gasifier and the spark ignition internal combustion engine models. For calculations the gasifier is split into three zones, namely drying - pyrolysis, oxidation and reduction sections. The gasifier's mathematical model consists of three separate sub-models, each describes the

processes in the corresponding zone. The process taking place in the reduction zone has been described by using chemical kinetic principles in order to avoid the introduction of assumptions related to the achievement of the thermo-chemical equilibrium state during gasifier's operation. The model is capable to accurately predict molar concentrations of different species in syngas (CO_2 , CO , H_2O , H_2 , CH_4 and N_2) and the temperature profile in the gasifier along its height. This information then can be used to choose the size of the reactor and material selection. The engine's model is based on the fuel-air thermodynamic cycle for spark ignition engines and such model takes into account the composition of syngas used as fuel. The engine's model also takes into account the effects of heat losses in the cycle through the walls of the cylinders and due to the gas blown by. Finally, the influence of dissociation processes during the combustion and the residual gases remaining in the cylinders at the beginning of the compression stroke is accounted for the computations of the engine's performance. The numerical results obtained using the proposed model are in a good agreement with data produced with the use of other theoretical models and experimental data published in open literature and with experimental data obtained in these investigations. The proposed model is applicable for the modeling integrated downdraft gasifier/engine biomass energy systems and can be used for the more accurate adjustment of the designed parameters of the gasifier and the engine in order to provide the higher overall efficiency of the system.

Juan, F. Pérez, Andrés Melgar, and Pedro Nel Benjumea. [43] studied the effect of operating and designing parameters, mainly reactor geometry, air superficial velocity, biomass moisture content, particle size and biomass type (pine bark and sewage sludge), on the performance of the gasification/combustion process of waste biomass in the fixed bed downdraft reactors. This experimental approach allows the optimization of the dynamic behavior of the thermochemical process. Emphasis was put on interactions between the mentioned parameters with output or response variables, such as biomass consumption rate, fuel/air equivalence ratio, product gas heating value and composition, tars concentration, and flame front velocity. The effect of the operating and designing parameters on response variables was studied by the means of multifactorial experimental designs. The statistical R^2 obtained through the data fitting of the multifactorial experimental designs indicated that the factors are

taken into account explained between 80.8% and 98.8% of the variability of the response variables. For the particular experiments carried out, the optimal gasification conditions were obtained with the following set of inlet conditions: air superficial velocity of 0.06 m/s, biomass particle size between 2 and 6 mm, and biomass moisture content of 10.62%. The corresponding response variables were: flame front velocity = 11 mm/min, fuel air equivalence ratio = 3.2, biomass consumption rate = 125 kg/h/m², lower heating value of the product gas = 2965.6 kJ/N m³, tar concentration = 7.73 g/N m³, and an average composition of the product gas of 8.0% H₂, 13.0% CO, 1.4% CH₄, 14.9% CO₂, and 62.7% N₂.

Yaning Zhang, Bingxi Li, Hongtao Li, and Bo Zhang. [44] presented that the steam gasification and partial oxidation (with oxygen) of biomass produce higher heating value gases than air gasification, thereby being two important technologies to make use of biomass energy. However, the dominance may be leveled off if the heat exergy for generating steam or heating oxygen is taken into consideration. Based on experimental data from literature, the exergy values and efficiencies of product gas, tar and unreacted carbon (including char and coke) from biomass steam gasification and partial oxidation (with oxygen) are studied in this paper. The exergy for generating steam are much higher than that for heating oxygen, but both of these are much lower than the exergy input by biomass, the exergy efficiencies of product gas, tar and unreacted carbon are therefore mainly determined by the exergy values of product gas, tar and unreacted carbon, respectively. When gasification temperature increases from 800 °C to 1200 °C, the exergy efficiencies of product gas, tar and unreacted carbon from steam gasification are in ranges of 49.31–58.48%, 0–16.15% and 5.17–9.53%, respectively, being higher than the corresponding ones from partial oxidation, which are in a range of 35.45–43.49%, 0–8.03% and 4.77–8.76%, respectively. Higher gasification temperature leads to higher exergy efficiencies of product gas and lower exergy efficiencies of tar, proper high gasification temperature therefore can be considered to improve the processes of biomass steam gasification and partial oxidation.

Nimit Nipattummakul, Islam Ahmed, Somrat Kerdsuwan, and Ashwani K. Gupta. [45] presented the effect of steam to carbon ratio on syngas properties with specific focus on the amounts of syngas yield, syngas composition, hydrogen yield,

energy yield, and apparent thermal efficiency. The apparent thermal efficiency is similar to cold gas efficiency used in industry and was determined from the ratio of energy in syngas to energy in the solid sewage sludge feedstock. A laboratory scale of a semi-batch type gasifier was used to determine the evolutionary behavior of the syngas properties using calibrated experiments and diagnostic facilities. Results showed an optimum steam to carbon ratio of 5.62 for the range of conditions examined here for syngas yield, hydrogen yield, energy yield and energy ratio of syngas to sewage sludge fuel. The results show that steam gasification provided 25% increase in energy yield when compared to pyrolysis at the same temperature.

V. Skoulou, A. Zabaniotou, G. Stavropoulos, and G. Sakelaropoulos. [46] presented a laboratory fixed-bed gasification of olive kernels and olive tree cuttings. Gasification took place with air, in a temperature range of 750–950 °C, for various air equivalence ratios (0.14–0.42) and under atmospheric pressure. In each run, the main components of the gas phase were CO, CO₂, H₂ and CH₄. Experimental results showed that gasification with air at high temperatures (950 °C) favoured gas yields. Syngas production increased with reactor temperature, while CO₂, CH₄, light hydrocarbons and tar followed the opposite trend. An increase in the air equivalence ratio decreased product gas production and lowered the product gas heating value, while favouring tar destruction. It was found that gas from olive tree cuttings at 950 °C and with air equivalence ratio of 0.42 had a higher LHV (9.41 MJ/Nm³) in comparison to olive kernels (8.60 MJ/Nm³). Olive kernels produced more char with a higher content of fixed carbon (16.39 w/w%) than olive tree cuttings. Thus, they might be considered as an attractive source for carbonaceous material production.

Siyi Luo, Yangmin Zhou, and Chuijie Yi. [47] presented the catalytic steam gasification of municipal solid waste for syngas production in a lab-scale fixed-bed reactor. The influence of the reactor temperature, steam to carbon ratio (S/C) and catalyst type (NiO/g-Al₂O₃ or calcined dolomite) on the gas yield, gas composition, H₂/CO molar ratio and carbon conversion efficiency were investigated. The results indicated that increasing reactor temperature resulted in greater gas production in the initial pyrolysis and improved endothermic reactions (gasification of char, catalytic cracking and reforming of tar), which resulted in the increase of syngas yield. Compared with MSW catalytic pyrolysis, the introduction of steam leads to more tar

and char participating in steam gasification, which resulted in a rapid increase of syngas yield and carbon conversion efficiency. NiO/g-Al₂O₃ catalyst revealed a better catalytic performance for the cracking of tar than that of calcined dolomite. The highest H₂ content (54.22%) and gas yield (1.75 Nm³/kg) were achieved at 900 °C, S/C being 2.41 with NiO/g-Al₂O₃ as catalyst.

Juan Daniel Martínez, Khamid Mahkamov, Rubenildo V. Andrade, and Electo E. Silva Lora. [48] presented the biomass downdraft reactors, coupled with reciprocating internal combustion engines (RICEs), they are a viable technology for small-scale heat and power generation. This paper contains information gathered from a review of published papers on the effects of the particle size and the moisture content of biomass feedstock and the air/fuel equivalence ratio used in the gasification process with regard to the quality of the product gas. Additionally, data on the parameters of product gas, such as its energy density, flame speed, knock tendency, auto-ignition delay period and the typical spark ignition timing, are systematised. Finally, information on the typical performance of various diesel and spark ignition RICEs fuelled with product gas is presented.

Z.A. Zainal, Ali Rifau, G.A. Quadir, and K.N. Seetharamu. [49] presented an experimental investigation of a downdraft biomass gasifier that it is carried out using furniture wood and wood chips. The effect of equivalence ratio on the gas composition, calorific value and the gas production rate is presented. The calorific value of the product gas increases with equivalence ratio initially, attains a peak and then decreases with the increase in equivalence ratio. The gas flow rate per unit weight of the fuel increases linearly with equivalence ratio. It is also observed that the complete conversion of carbon to gaseous fuel has not taken place even for the optimum equivalence ratio.

I.-S. Antonopoulos, A. Karagiannidis, A. Gkouletsos, and G. Perkoulidis. [50] studied a non-stoichiometric model for a downdraft gasifier that it was developed in order to simulate the overall gasification process. Mass and energy balances of the gasifier were calculated and the composition of produced syngas was predicted. The capacity of the modeled gasifier was assumed to be 0.5 MW, with an Equivalence Ratio (EQ) of 0.45. The model incorporates the chemical reactions and species involved, while it starts by selecting all species containing C, H, and O, or any other

dominant elements. Olive wood, miscanthus and cardoon were tested in the formulated model for a temperature range of 800–1200 °C, in order to examine the syngas composition and the moisture impact on the supplied fuel. Model results were then used in order to design an olive wood gasification reactor.

Nimit Nipattummakul, Islam I. Ahmed, Somrat Kerdsuwan, and Ashwani K. Gupta. [51] studied a high temperature steam gasification to obtain high percentage of hydrogen in the syngas from low-grade fuels. Sewage sludge is considered to be a renewable fuel because it is sustainable and it has a good potential for energy recovery. In this investigation, sewage sludge samples were gasified at various temperatures to determine the evolutionary behavior of syngas characteristics and other properties of the syngas produced. The syngas characteristics were evaluated in terms of syngas yield, hydrogen production, syngas chemical analysis, and efficiency of energy conversion. In addition to gasification experiments, pyrolysis experiments were conducted to evaluate the performance of gasification over pyrolysis. The increase in reactor temperature resulted in an increase in the generation of hydrogen. Hydrogen yield at 1000 °C was found to be $0.076 \text{ g}_{\text{gas}} \text{ g}^{-1}_{\text{sample}}$. Steam, as the gasifying agent, increased the hydrogen yield three times when compared to air gasification. Sewage sludge gasification results were compared with other samples, such as, paper, food wastes, and plastics. The time duration for sewage sludge gasification was longer when compared to other samples. On the other hand, sewage sludge yielded more hydrogen than that from paper and food wastes.

Avdhesh Kr. Sharma. [52] presented an experimental study on a suction gasifier (downdraft) arrangement operating on kiker wood or *Acacia nilotica* (L). Studies were conducted to investigate the influence of fluid flow rate on pressure drop through the gasifier system for ambient isothermal airflow and ignited mode, pumping power, and air-fuel ratio, gas composition and gasification efficiency. Results of pressure drop, temperature profile, gas composition or calorific value are found to be sensitive with the fluid flow rate. Ignited gasifier gives much higher pressure drop when being compared against newly charged gasifier bed with isothermal ambient airflow. Higher reaction temperatures in gasifier tends to enhance gasifier performance, while, overall pressure drop and pumping power through the system

increases. Both ash accumulated gasifier bed and sand bed filters with tar laden quartz particles show much higher pressure drops.

Catharina Erlich, and Torsten H. Fransson. [53] studied the impact of the char bed properties such as char bed porosity and pressure drop on the gasification performance as well as the impact of fuel particle size and composition on the gasification process in one and the same gasifier. In addition, there is very little gasification data available in literature of “before disregarded” fuels, such as sugar cane bagasse from sugar/alcohol production and empty fruit bunch (EFB) from the palm-oil production. By pelletizing these residues, it is possible to introduce them into downdraft gasification technology, which has been done in this study. The results show that one and the same reactor can be used for a variety of fuels in pellet form, but at varying air-fuel ratios, temperature levels, gas compositions and lower heating values. Gasification of wood pellets results in a richer product gas while EFB pellets give a poorer one with higher contents of non-combustible compounds. In this gasification study, there is almost linear relation between the air-fuel ratio and the cold-gas efficiency for the studied fuels: Higher air-fuel ratios result in better efficiency. The pressure drop in the char bed is higher for more reactive fuels, which in turn is caused by low porosity char beds.

Pratik N. Sheth, and B.V. Babu. [54] studied biomass gasification, conversion of solid carbonaceous fuel into combustible gas by partial combustion, it is a prominent technology for the production of hydrogen from biomass. The concentration of hydrogen in the gas generated from gasification depends mainly upon moisture content, type and composition of biomass, operating conditions and configuration of the biomass gasifier. The potential of the production of hydrogen from wood waste by applying downdraft gasification technology is investigated. An experimental study is carried out using an Imbert downdraft biomass gasifier covering a wide range of operating parameters. The product gas generated in the downdraft gasifier is analyzed using a gas chromatograph (NUCON 5765) with thermal conductivity detector (TCD). The effects of air flow rate and moisture content on the quality of product gas are studied by performing experiments. The performance of the biomass gasifier is evaluated in terms of equivalence ratio, composition of product gas, and rate of hydrogen production.

Avdhesh, Kr. Sharma. [55] presented the experimental study on 75 kWth, downdraft (biomass) gasifier system that it has been carried out to obtain temperature profile, gas composition, calorific value and trends for pressure drop across the porous gasifier bed, cooling–cleaning train and across the system as a whole in both firing and well as non-firing mode. Some issues related to re-fabrication of damaged components/parts have been discussed in order to avoid any kind of leakage. In firing mode, the pressure drop across the porous bed, cooling–cleaning train, bed temperature profile, gas composition and gas calorific value are found to be sensitive to the gas flow rate. The rise in the bed temperature due to chemical reactions strongly influences the pressure drop through the porous gasifier bed. In non-firing mode, the extinguished gasifier bed arrangement (progressively decreasing particle size distribution) gives a much higher resistance to the flow when compared to a freshly charged gasifier bed (uniformly distributed particle size). The influence of ash deposition in fired-gasifier bed and tar deposition in sand filters is also examined on the pressure drop through them. The experimental data generated in this article may be useful for the validation of any simulation codes for gasifiers and the pressure drop characteristics may be useful towards the coupling of a gasifier to the gas engine for motive power generation or decentralized electrification applications.

V. Skoulou, A. Swiderski, W. Yang, and A. Zabaniotou. [56] presented the exploitation of olive kernel for bioenergy production, with respect to the green house gases (GHGs) mitigation, it is the main aim of this work. In this study, olive kernels were used as a solid biofuel, and high temperature steam gasification (HTSG) was investigated, in the fixed bed unit at KTH Sweden, with regard to hydrogen maximization in the produced gasification gas. Experiments were carried out in a temperature range of 750–1050 °C, with the steam as the gasifying agent. The behavior of the olive kernels, under residence times from 120 up to 960 s, has been studied. At 1050°C, a medium to high calorific value gas was obtained ($LHV_{\text{gas}} = 13.62 \text{ MJ/Nm}^3$), while an acquired H_2/CO molar ratio equal to four proved that olive kernel HTSG gasification could be an effective technology for a hydrogen-rich gas production ($\sim 40\%v/v \text{ H}_2$ in the produced gasification gas at 1050 °C). The produced char contained 79%ww of fixed carbon, low chlorine and sulphur content, which enables it to the further reuse of energetic purposes. Tar content in the produced gas at

750°C was 124.07 g/Nm³, while a 1050°C at 79.64% reduction was observed and reached the value of 25.26 g/Nm³.

Gong Cheng, Qian Li, Fangjie Qi, Bo Xiao, Shiming Liu, Zhiquan Hu, and Piwen He. [57] studied an allothermal biomass gasification system using biomass micron fuel (BMF) as external heat source was developed. In this system, heat supplied to gasifier was generated from the combustion of BMF. Biomass feedstock was gasified with steam and then tar in the produced gas was decomposed in a catalytic bed with NiO/C-Al₂O₃ catalyst. Finally, the product gas was employed as a substitute for civil fuel gas. An overall energy analysis of the system was also investigated. The results showed that the lower heating value of the product gas reached more than 12 MJ/Nm³. The combusted BMF accounted for 26.8% of the total energy input. Allothermal gasification based on the substituted BMF for conventional energy was an efficient and economical technology to obtain bioenergy.

P. Plis, and R.K. Wilk. [58] investigated the process of air biomass gasification in a fixed bed gasifier. Theoretical equilibrium calculations and experimental investigation of the composition of syngas were carried out and compared with the findings of other researchers. The influence of the excess air ratio (l) and the parameters of biomass on the composition of syngas were investigated. A theoretical model is proposed, based on the equilibrium and thermodynamic balance of the gasification zone. The experimental investigation was carried out at a setup that consists of a gasifier connected by a pipe with a water boiler fired with coal (50 kW_{th}). Syngas obtained in the gasifier is supplied into the coal-firing zone of the boiler, and co-combusted with coal. The moisture content in biomass and excess air ratio of the gasification process are the crucial parameters, determining the composition of syngas. Another important parameter is the kind of the applied biomass. Despite similar compositions and dimensions of the two investigated feedstocks (wood pellets and oats husk pellets), compositions of syngas obtained in the case of these fuels were different. On the basis of tests, it may be stated that oats husk pellets are not a suitable fuel for the purpose of gasification.

Gerardo Gordillo, Kalyan Annamalai. [59] presented experimental results obtained from adiabatic, fixed bed gasification of DB using air and steam as oxidizers. The effect of equivalence ratio (ER) and steam to fuel ratio (S/F ratio) on temperature

profile, gas composition (CO, CO₂, H₂, N₂, CH₄, and C₂H₆), gross heating value (HHV) and energy conversion efficiency (ECE) are discussed. The results show that the peak temperature, ECE, and CO decrease and H₂ and CO₂ increase with an increase in ER; the increase in S:F at same ER increases H₂, CO₂, CH₄, HHV, and ECE, and decreases CO.

Woei Saw, Hamish McKinnon, Ian Gilmour, and Shusheng Pang [60]. A studied the on steam gasification of mixtures of wood pellets and biosolids (dried sewage sludge). The gasification experiments were conducted in a fast circulating dual fluidised bed gasifier. In the experiments, the gasification temperature was set at 720 C and the fuel-feeding rate to the gasifier was 15.5 kg/h (as received). The biosolids was blended with wood pellets, with biosolids to wood pellet mass proportion being controlled at 0% (or 100% wood pellets), 10%, 20%, 40%, 60%, 80% and 100%. According to the results, the syngas produced from the biosolids had higher H₂ content (28%) when compared with that from pure wood (23%), with the H₂ content increasing with biosolids fuel loadings. In addition, the ratio of H₂/CO increased from 0.6 to 0.9 as the biosolids loading increasing from 0% to 100%. However, the syngas yield and the cold gas efficiency of the gasification dramatically decreased by 53% and 43%, respectively, at 100% biosolids loading compared with 100% pure wood loading. Nevertheless, the increase from 10% to 20% loading of biosolids in the fuel did not diminish the yields and the cold gas efficiency. For the gasification of pure biosolids, the gas concentrations of H₂ and CO in this study were found to be 40% higher than that being produced by other researchers using air, O₂ or CO₂/N₂ as the gasification agent.

Ligang Wei, Shaoping Xu, Li Zhang, Changhou Liu, Hui Zhu, Shuqin Liu [61]. studied the characteristics of steam gasification of two kinds of biomass (legume straw and pine sawdust) at high heating rate (in the order of 1000°Cs⁻¹) in a gas–solid concurrent downflow free-fall reactor were investigated. The effects of steam/biomass (S/B) mass ratio (0.0–1.0 g/g) and reactor temperature (750–850°C) on the product yields and the compositions of product gas were determined. The experimental results show that the gas yields and the content of H₂ in the gas increase with reactor temperature, while the yields of tar, char and the content of CO and CH₄ in the product gas decrease. The presence of steam increases the gas yields and

reduces the tar and char yields. It is inferred that an in situ steam reforming reaction of tar takes place even in a short gas residence time in the free-fall reactor. Water-gas shift reaction determines greatly the gas compositions and H₂ production at higher temperature. The effects of different natural occurring catalysts, limestone, olivine and dolomite, were also investigated in the same unit. Dolomite reveals a comparably good performance in terms of catalytic activity of tar destruction and the consequential increase in the production of gases in thus a short vapor-catalyst contact time. The tars produced were analyzed by FTIR spectrometry. The results verify that the presence of steam and catalyst favors the tar decomposition.

Nimit Nipattummakul, Islam I. Ahmed, Somrat Kerdsuwan, Ashwani K. Gupta. [62] presented the results obtained from the steam-assisted gasification of oil palm trunk waste that are presented. A batch type gasifier has been used to examine the syngas characteristics from the gasification of palm trunk waste using steam as the gasifying agent. Reactor temperature was fixed at 800 °C. The results show initial high values of syngas flow rate, which is attributed to rapid devolatilization of the sample. Approximately over 50% of the total syngas generated was obtained during the first five minutes of the process. An increase in steam flow rate accelerated the gasification reactions and resulted in reduced gasification time. The effect of steam flow rate on the apparent thermal efficiency has also been investigated. Variation in steam flow rate slightly affected the apparent thermal efficiency and was found to be very high. Properties of the syngas obtained from the gasification of oil palm trunk waste have been compared to other samples under the similar operating conditions. Oil palm trunk waste yielded more syngas, energy and hydrogen than that from other types of biomass, such as mangrove wood, paper and food waste.

Yanwen Guan, Siyi Luo, Shiming Liu, Bo Xiao, Lei Cai [63]. presented a two-region municipal solid waste (MSW) steam catalytic gasification process. The gasifier was composed of two individual reactors: one is the gasification reactor and the other is the catalytic reactor. The MSW was initially gasified and the produced tar was gasified in the gasification reactor, and further, the tar that is not gasified entered the catalytic reactor together with the fuel gas and was catalytically decomposed to be fuel gas. The influences of the catalysts, steam and temperature on the content of tar, dry gas yield and composition, and carbon conversion efficiency were studied. The

results indicated that under the optimum operating conditions, the dry gas yield can be up to 1.97 Nm³/kg MSW and the tar in the product can be completely eliminated. The concentration of hydrogen, carbon monoxide and methane in the fuel gas produced was 50.8%, 9.32% and 13.3%, respectively.

Manuel Campoy, Alberto Gómez-Barea, Fernando B. Vidal, Pedro Ollero. [64] studied the effect of oxygen concentration in the gasification agent by enriched-air-steam biomass gasification tests in a bubbling fluidised-bed gasification (FBG) plant. The oxygen content in the enriched air varied from 21% (v/v, i.e. air) to 40% (v/v), aiming at simulating FBG where enriched air is produced by membranes. The stoichiometric ratio (ratio of actual to stoichiometric oxygen flow rates) and steam-to-biomass ratio (ratio of steam to biomass, dry and ash-free, flow rates) were varied from 0.24 to 0.38 and from 0 to 0.63, respectively. The tests were conducted under simulated adiabatic and autothermal conditions, to reproduce the behavior of larger industrial FBG. The temperature of the inlet gasification mixture was fixed consistently at 400 °C for all tests, a value that can be achieved by energy recovery from the off-gas in large FBG without tar condensation. It was shown that the enrichment of air from 21 to 40% v/v made it possible to increase the gasification efficiency from 54% to 68% and the lower heating value of the gas from 5 to 9.3 MJ/Nm³, while reaching a maximum carbon conversion at 97%. The best conditions were found at intermediate values of steam-to-biomass ratio, specifically within the range 0.25–0.35. The enriched-air-steam gasification concept explored in this work seems to be an interesting option for the improvement of the standalone direct air-blown FBG because it considerably improves the process efficiency while maintaining the costs to be relatively low when compared to the oxygen-steam gasification.

Abrar Inayat, Murni M. Ahmad, M.I. Abdul Mutalib, Suzana Yusup. [65] studied the biomass steam gasification with in-situ carbon dioxide captures using CaO exhibits good prospects for the production of hydrogen rich gas. The present work focuses on the process modeling for hydrogen production from oil palm empty fruit bunch (EFB) using MATLAB for parametric study. The model incorporates the reaction kinetics calculations of the steam gasification of EFB (C_{3.4}H_{4.1}O_{3.3}) with in-situ CO₂ capture, as well as mass and energy balances calculations. The developed model is used to investigate the effect of the temperature and steam/biomass ratio on

the hydrogen purity, yield and efficiency. Based on the results, hydrogen purity at more than 76.1 vol.% can be achieved. The maximum hydrogen yield predicted at the outlet of the gasifier is 102.6 g/kg of EFB. It is found that the increment in temperature and steam/biomass ratio promotes the hydrogen production. However, it is also predicted that the efficiency decreases when using more steam. Due to the on-going empirical work, the results are compared with published literatures on different systems. The comparison shows that the results are in an agreement to some extent due to the different basis.

G. Schuster, G. Loer, K. Weigl, and H. Hofbauer. [66] developed a model for steam gasification of biomass by applying thermodynamic equilibrium calculations. With this model, the simulation of a decentralized combined heat and power station based on a dual fluidized-bed steam gasifier was carried out. Fuel composition (ultimate analysis and moisture content) and the operating parameters, temperature and amount of gasification agent, were varied over a wide range. Their influences on amount, composition, and heating value of product gas and process efficiencies were evaluated. It was shown that the accuracy of an equilibrium model for the gas composition is sufficient for the thermodynamic considerations. Net electric efficiency of about 20% can be expected with a rather simple process. Sensitivity analysis showed that gasification temperature and fuel oxygen content were the most significant parameters to determine the chemical efficiency of the gasification.

Feng Yan, Si-yi Luo, Zhi-quan Hu, Bo Xiao, and Gong Cheng. [67] presented the steam gasification experiments of biomass char that were carried out in a fixed-bed reactor. The experiments were completed at the bed temperature of 600–850°C, a steam flow rate of 0–0.357 g/min/g of biomass char, and a reaction time of 15 min. The aim of this study is to determine the effects of bed temperature and steam flow rate on syngas yield and its compositions. The results showed that the both high gasification temperature and the introduction of proper steam led to higher yield of dry gas and higher carbon conversion efficiency. However, excessive steam could reduce gas yield and carbon conversion efficiency. The maximum dry gas yield was obtained at the gasification temperature of 850 °C and steam flow rate of 0.165 g/min/g biomass char.

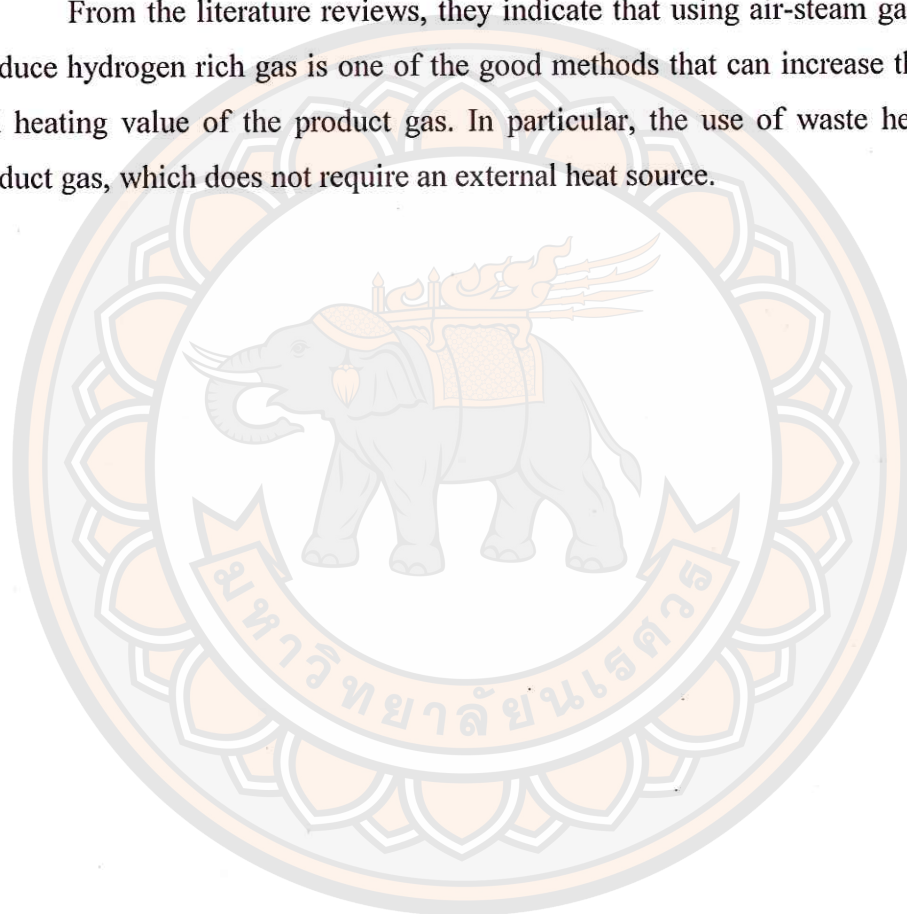
Murni M. Ahmad, Abrar Inayat, Suzana Yusup and Khalik M. Sabil. [68] presented a study on the process development of hydrogen production via gasification of Empty fruit bunch (EFB) with in-situ adsorption of CO₂ based on equilibrium modeling approach. The process flowsheet simulation is performed using iCON, PETRONAS process simulation software. This work investigates the influence of the temperature within the range of 600 to 1000°C and steam/biomass ratio between 0.1 to 1.0 on the hydrogen yield and product gas composition. The importance of different reactions involved in the system is also discussed. Using the simulation, the optimal operating conditions are predicted to be at 800°C and steam/biomass ratio of 0.6. Hydrogen yield of 149g kg⁻¹ of EFB can be obtained at 1,000°C. The preliminary economic potential per annum of the oxygen-steam gasification system coupled with in-situ CO₂ adsorption is RM 6.64 x 10⁶ or approximately USD 2 x 10⁶.

Pratik N. Sheth, and B.V. Babu. [69] studied a process of conversion of solid carbonaceous fuel into combustible gas by partial combustion that is known as gasification. The resulting gas, known as product gas, is more versatile in its use than the original solid biomass. In the present study, a downdraft biomass gasifier is used to carry out the gasification experiments with the waste generated while making furniture in the carpentry section of the institute's workshop. Dalbergia sisoo, generally known as sesame wood or rose wood, is mainly used in the furniture and the wastage of the same is used as a biomass material in the present gasification studies. The effects of air flow rate and moisture content on biomass consumption rate and quality of the product gas generated are studied by performing experiments. The performance of the biomass gasifier system is evaluated in terms of equivalence ratio, product gas composition, calorific value of the product gas, gas production rate, zone temperatures and cold gas efficiency. Material balance is carried out to examine the reliability of the results generated. The experimental results are compared with those reported in the literature.

Niladri Sekhar Barman, Sudip Ghosh, and Sudipta De. [42, 70] presented a model for fixed bed downdraft biomass gasifiers considering tar as one of the gasification products. A representative tar composition along with its mole fractions, as available in the literature, was used as an input parameter within the model. The study used an equilibrium approach for the applicable gasification reactions and also considered possible deviations from equilibrium to further upgrade the equilibrium

model to validate a range of reported experimental results. Heat balance was applied to predict the gasification temperature and the predicted values were compared with the reported results in literature. A comparative study was made with some reference models available in the literature and also with experimental results reported in the literature. Finally, a predicted variation of performance of the gasifier by this validated model for different air–fuel ratio and moisture content was also discussed.

From the literature reviews, they indicate that using air-steam gasification to produce hydrogen rich gas is one of the good methods that can increase the hydrogen and heating value of the product gas. In particular, the use of waste heat from the product gas, which does not require an external heat source.



CHAPTER III

RESEARCH METHODOLOGY

The research methodology of this study is presented in the Figure 17. as followed.

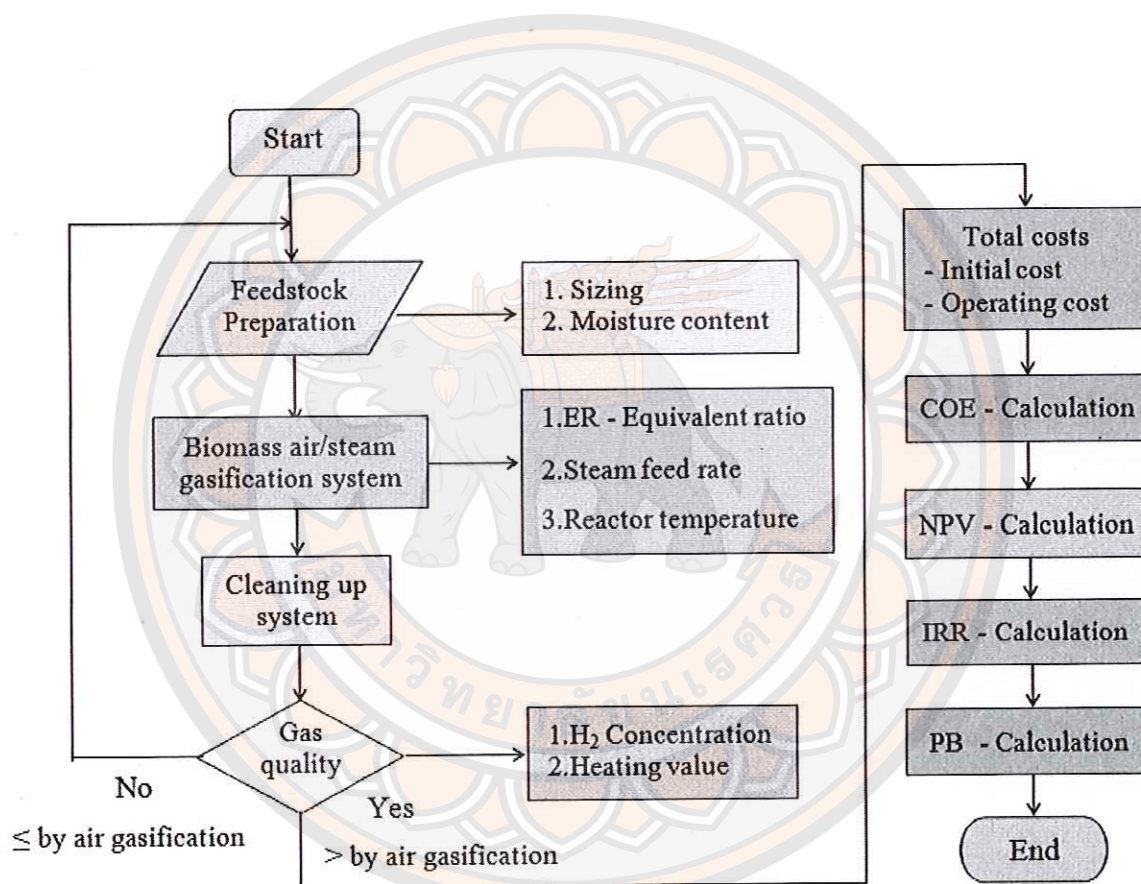


Figure 17 A research methodology flow chart of air/steam gasification system

Air/steam gasification by using waste heat

The use of air/steam mixture, as oxidizing agent, will result in more efficient production of product gas, if the process of steam production does not require an external heat source. In this study, steam is generated by the heat taken from the hot product gas. The 6.8% of total energy (waste heat) is derived from biomass feedstock,

and it can produce enough steam for this system. Consequently, a small heat exchanger is designed for the steam production, and it can be installed outside the gasifier where the temperature of the reactor would not be affected as shown in Figure 17.

The heat exchanger installed close to the throat of the reactor will result in a reduction of the reactor temperature, which is not good for the gasification system. The appropriate distance from the heat exchanger to the reactor is 90 cm. because the reactor temperature is not affected by the heat exchanger.

The subject matter in this article is the evaluation of the effects of operating parameters on the experimental results of the gasifier. Also, the operational performances of the air-steam gasification system are reported.

Experimental Setup

The experiment setup of the air-steam gasification system comprises of these integrated units:

1. Downdraft fixed bed gasifier
2. Gas cleaning system
3. Counter-flow heat exchanger
4. Monitoring and measuring devices
5. After burner

Figure 18. shows a schematic diagram of the gasification test facility system. It consists of a temperature sensor, the flow rates are regulated by means of two rotameters, a pump feeds water into a steel steam generator that produces slightly superheated steam ($\sim 110-120^{\circ}\text{C}$) whose temperature is monitored by K-type thermocouple readout prior to entering the reactor. The gasification process is run with excess H_2O to ensure that the biomass is the limiting agent in the steam gasification reactions.

The experimental fixed bed gasifier facility utilizes air-steam gasification technology. The facility provides steam-gasifying agent to the reactor at any desired temperature. To increase the amount of steam in the reactor, additional water is added to the counter flow heat exchanger by using a control valve. The product gas composition was measured by using a micro - GC.

The diagram of steam gasification system is shown in the following figure.

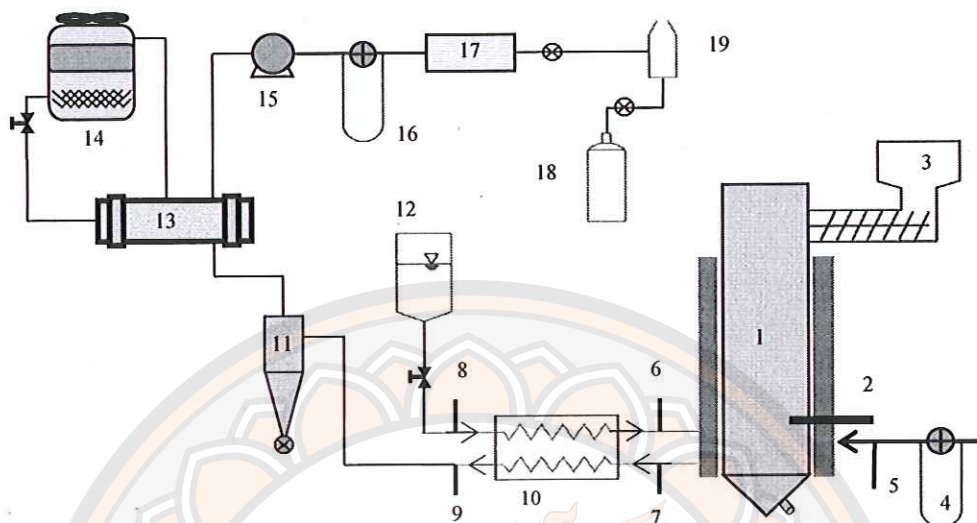


Figure 18 Schematic diagram of biomass air/steam gasification setup

Note: (1) Downdraft gasifier (2) Reactor Temp, T_2 (3) Feed Hopper (4) Air flow meter (5) Air inlet temp, T_5 (6) Steam temp, T_6 (7) Hot gas temp, T_7 (8) Water inlet temp, T_8 (9) Hot gas outlet, T_9 Thermocouple (10) Heat exchanger (11) Cyclone (12) Water tank (13) Condenser (14) Cooling tower (15) Root blower (16) Gas flow meter (17) Gas analyzer (18) LPG tank (19) Burner

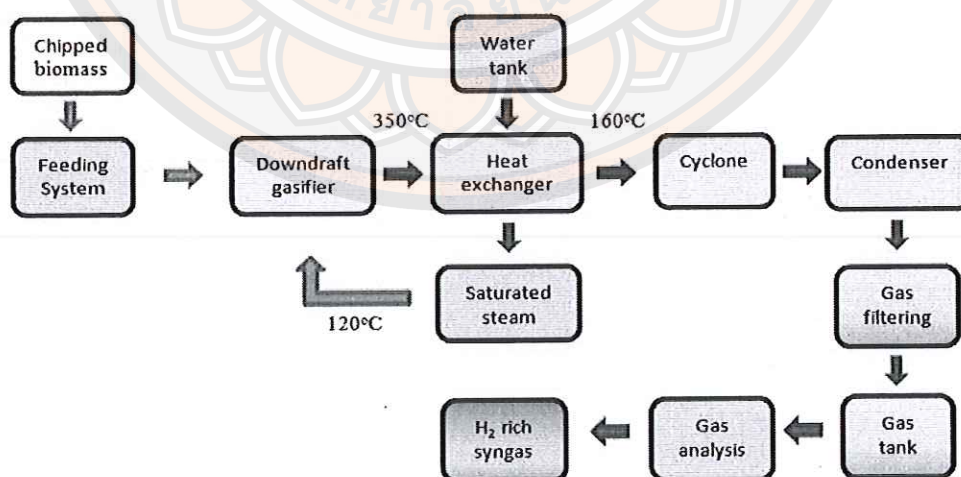


Figure 19 Process flow diagram for experimental facility to produce hydrogen rich syngas

Downdraft Fixed Bed Gasifier

To determine the efficiencies of the steam gasification system, an experimental facility was required that it could be able to operate continuously and produce repeatable and reliable gasification. The designed gasifier was capable of converting 14 kg of biomass per hour ($50 \text{ kW}_{\text{th}}$).

Figure 19. shows the process flow diagram for the gasification system designed to meet the aforementioned criteria. The process begins with wood chip being inserted into the gasifier through a feeding system. Gasification will take place inside a gasifier by partial combustion. After the gasification process has occurred, the hot product gas enters a shell and tube counter flow heat exchanger. Saturated steam is generated in a heat exchanger (boiler) by sensible heat from hot product gas and entering the oxidation zone. The product gas is filtered with large particles by a cyclone before being cooled in a cooler. The cooled gas is then condensed before being filtered again to remove any liquid and entrained ash. The cooled and filtered product gas is finally analyzed to determine the conversion of energy to product gas and hydrogen rich gas.

A down draft gasifier of $50 \text{ kW}_{\text{th}}$ capacity unit and a shell and tube counter-flow heat exchanger were used in the study. Experimental trials were carried out using steam as oxidizing agent. For each flow rate of steam, 4 experimental trails were carried out with different settings of steam supply in order to measure hydrogen increases. The air supply was fixed throughout the experiment. Wood chips size approximately $25 \times 25 \times 25 \text{ mm}^3$ was used as the fuel.

Air/steam gasification test facility (Figure 18.) is equipped with the following main equipment:

1. Continuous type, downdraft fixed bed gasifier.
2. Air root blower supplying the air to the gasifier.
3. Heat exchanger to heat steam in the lower range of temperatures that is up to $100\text{-}120^\circ\text{C}$, 1.2 bar. The heat exchanger is equipped with a water tank unit.
4. After burning combustion chamber to burn completely the product fuel gas.
5. Set of temperatures to control gasifier operation.

6. Micro gas chromatograph (GC).

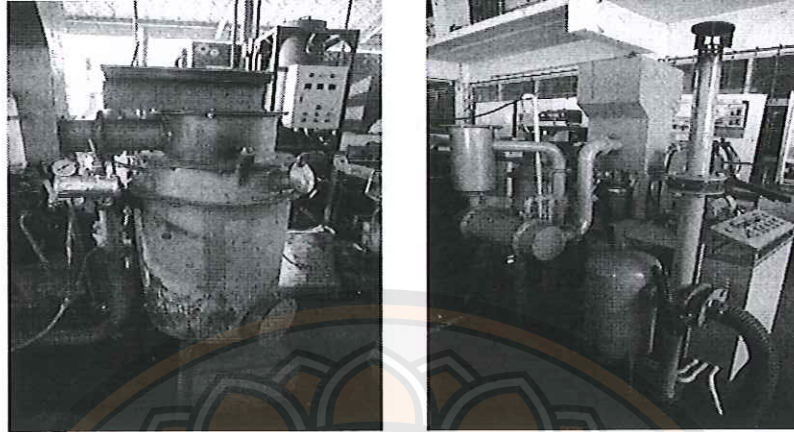


Figure 20 Experiment setup: a downdraft gasifier system

Shell and tube counter flow heat exchanger

In this study, a heat exchanger with one shell (65 mm.ID, 74 mm.OD) and seven tube (12.5 mm ID, 17.0 mm OD, $k = 20 \text{ W/m.K}$) passes is to be designed as shown in Figure 3. Hot product gas flowing in tubes heats water. The inlet and outlet temperature of the feedwater and the inlet temperature of hot product gas are 120°C and 65°C , respectively. Water flows through the shell at the rate of 2.6 kg/hr and its temperature increases from 28°C to 120°C .

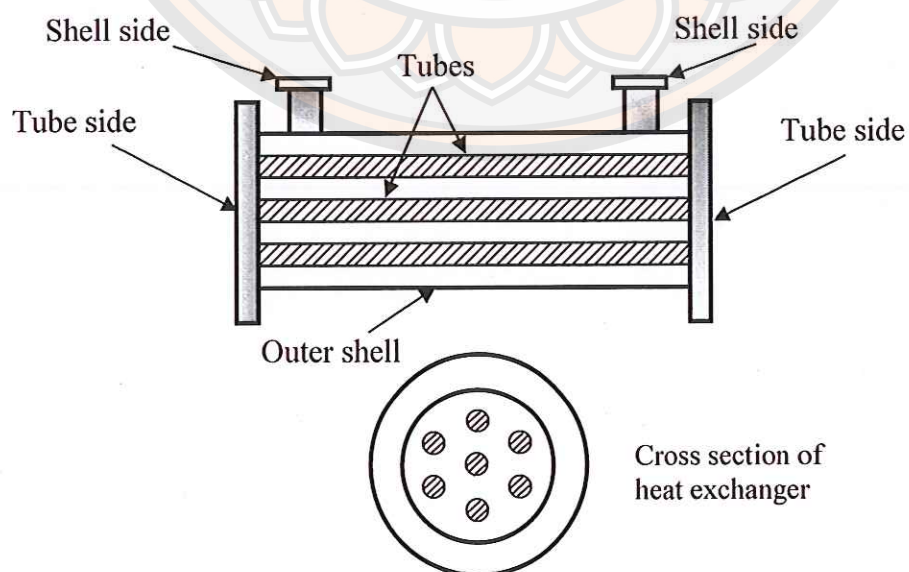


Figure 21 Side view and cross section of the prototype

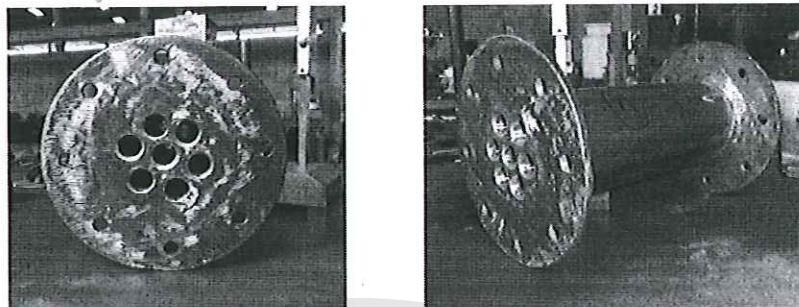


Figure 22 A prototype of shell and tube counter flow heat exchanger

Installation of the heat exchanger, whether or not the temperature affects the oxidation zone of the gasifier. The gasifier and boiler installed adjacent to the reduction zone. Result in a lower temperature in the oxidation zone where the boiler is installed outside the gasifier. So, in order not to affect the oxidation zone, boiler was installed immediately outside the gasifier to install the heat exchanger for waste heat recovery to produce steam that will have no effect on the temperature in the oxidation zone. Hence, to avoid such problems heat exchanger was installed outside the gasifier as shown in Figure 23.

Because the sensible heat that comes with synthesis gas flowing out of the downdraft gasifier high. Heat of the hot product gas can be used to produce steam and can be use as oxydizing agent. Considering the energy balance in a small-fired tube boiler, which mounted at gasifier gas outlet as shown in Figure 23.

Hot gas flows into the boiler at temperature of 350°C and flows out of the boiler at 160°C. Water at 28°C flows into the boiler is heated until it becomes steam at 120° C, 1.2 bar and flows into gasifier at the rate of 2.6 kg/h. The waste heat of the product (synthesis) gas is a load of gas purification system, so bringing waste heat recovery will allow the system to run more efficiently without affecting the oxidation zone. The use of steam to produce product gas to make hydrogen synthesis gas has soared. The heat will be higher as well.

Because the sensible heat that comes with the product gas flowing out of the downdraft gasifier high. Heat of hot product gas can be used to produce steam and use

as oxidizing agent. Considering the energy balance in a small-fired tube boiler which mounted at gasifier gas outlet as shown in Figure 23.

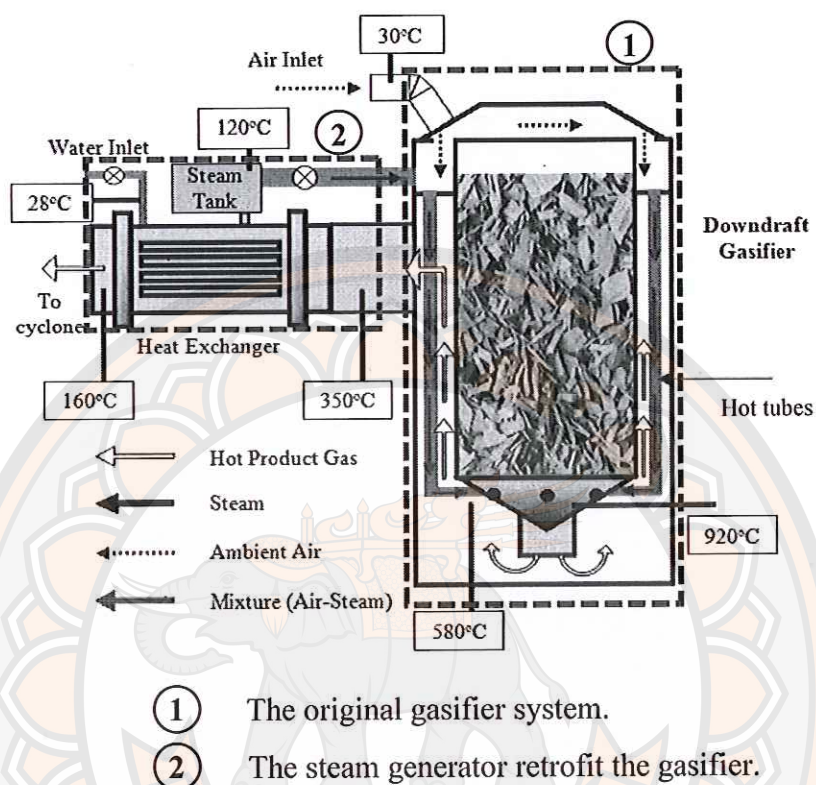


Figure 23 Schematic diagram of the gasifier system, which used in this research

The size of the original gasifier is $50 \text{ kW}_{\text{th}}$ and the position of the boiler is far from the reduction zone. This will not affect the temperature of the oxidation zone. Product gas flow from the reduction zone at a temperature of 580°C , and flows into heat exchanger at a temperature of about 380°C . The flow meter measures the flow rate of the product gas was $32 \text{ m}^3/\text{h}$. Heat from product gas transferred to the water until the water changes into steam at 120°C , 1.2 bar. Water flows into heat exchanger at 28°C , at the flow rate of 0.1 to 2.6 kg/h. The data showed that heat exchanger could produce steam at a rate of 0.1 to 2.6 kg/h of steam at a temperature of 120°C and pressure 1.2 bar.

Methodology of experiments

1. Mixture of air and steam were used as an oxidizing agent.
2. Wood chip with size ranging from 2 cm to 4 cm is used as feedstock.
3. The feedstock is loaded into the bed material and the mass of charge depending on the feedstock used.
4. A hot low calorific value gas is directly blown through the heat exchanger.
5. Fuel gas composition with help of micro Gas Chromatograph (GC) connected to the gas-sampling probe.

The following gas components of the fuel gas are measured: hydrogen (H_2), carbon monoxide (CO), carbon dioxide (CO_2), nitrogen (N_2), oxygen (O_2), methane (CH_4), ethane (C_2H_6), ethylene (C_2H_4), acetylene (C_2H_2) and higher hydrocarbons (C_xH_y). The composition of fuel gas was determined from gas samples taken at the gasifier outlet.

Experimental procedure

The air-steam gasifier system was operated and tested in the following step:

1. Before starting, all the parts of the gasifier were properly tightened.
2. Then the gasifier was loaded with biomass by screw feeder.
3. Start up the gasifier includes all operations required until reaching a steady state, whereby the pre-set air and steam temperature is stable.
4. The primary air supply was full at the starting of gasifier and then maintained around 45% of the stoichiometric condition to ensure the partial oxidation of the biomass with the help of a blower-control, gate valve and air flow meter.
5. After 5 minutes, the producer gas came out through the heat exchanger and burner. The producer was ignited at the burner.
6. Put water flows into the heat exchanger and flows out in steam. After water vapor with air flow into the gasifier.
7. A burning flame was observed and continued to burn for 30 minutes.
8. Shutdown procedure includes all the actions to safely seal the gasifier.
9. Shut down of all electric appliances is the last procedure following the cooling of fuel gas.

10. After the gasifier cooled down, the bottom section of the gasifier was opened so that all ash were carefully removed and monitored.

During the testing of the gasifier system the following measurements were taken.

1. The air and producer gas volumetric flow rate were measured by air flow rate instrument.

2. Average biomass consumption was measured using digital weight scale.

3. The temperature in the oxidation zone and gas outlet were measured with the help of a K-type thermocouple. Throughout the testing, the temperature of the oxidation zone was found up to 920°C. The temperature of producer gas at the gasifier exit was 350°C

4. Steam production rate measurements and measure the temperatures of the heat exchanger.

5. Gas analyzer is used to measure the composition of producer gas.

Proximate and elementary analysis of the wood chip is presented in Table 3.

Table 3 Proximate and ultimate analysis of wood chip

Proximate analysis	Method	Units
Moisture content	ASTM D 7582-10	25 %
Volatile matter	ASTM D 7582-10	54 %
Ash	ASTM D 7582-10	3 %
Fixed carbon	ASTM D 7582-10	18 %
Net calorific value	ASTM D 7582-10	3,500 MJ/kg
Carbon	ASTM E 777	49 %
Hydrogen	ASTM E 777	4 %
Nitrogen	ASTM E 778	0.1 %
Oxygen	ASTM E 870	25 %
Sulphur	ASTM E 775	0.1 %

Raw Data Evaluation

Based on the measured flow rates, temperatures and product gas composition, the following values were calculated:

1. Biomass consumption, [kg/h]
2. Air flow rate and product gas flow rate, [Nm³/h]
3. Gasification efficiency, η
4. High heating value, HHV of product gas, [MJ/Nm³]
5. Mass and energy balance

Table 4 Shows the operation data of the fixed bed downdraft gasification

Number of test	Temperature of air [°C]	Air flow rate [Nm ³ /h]	Mass of steam in feed gas [kg/h]	Mass of fuel [kg/h]	Fuel type
1	30	14.13	0.7	14.0	Wood chip
2	30	14.13	1.1	14.0	Wood chip
3	30	14.13	1.5	14.0	Wood chip
4	30	14.13	1.8	14.0	Wood chip
5	30	14.13	2.2	14.0	Wood chip
6	30	14.13	2.6	14.0	Wood chip

Process assumptions

There are a few assumptions considered in the model process:

1. The gasifier operates under steady flow-steady state conditions, adiabatic and atmospheric pressure [36].
2. Potential and kinetic energies are negligible.
3. The reactions occur isothermally and at constant volume [37]. The gasifier is isothermal and at equilibrium condition.

4. Residential time is sufficient to reach the gasifier the equilibrium mode.
5. Perfect mixing and uniform temperature in the gasifier [26].
6. Gases except H₂, CO, CO₂ and CH₄ are considered to be dilute.
7. The product gases are at the gasifier outlet temperature.
8. Air is composed of 21% oxygen and 79% nitrogen.
9. All gases behave ideally. The gases obey the ideal gas relations.
10. Tar formation in the process is negligible [40].
11. Ash residue behind gasification process is negligible.

Data collection

Data, which will be collected during the gasifying experiments are as followed.

Table 5 Gasifying experiment data

No.	Parameters	Equipment/Method	Result
1.	Air flow rate	U-tube manometer	14.13 m ³ /h
2.	Product gas flow rate	U-tube manometer	32 Nm ³ /h
3.	Gasifying temperature	Ceramic thermocouple	920°C
4.	Air inlet temperature	Type K thermocouple	30°C
5.	Water inlet temperature	Type K thermocouple	28°C
6.	Steam temperature	Type K thermocouple	120°C
7.	Hot gas temperature	Type K thermocouple	350°C
8.	Product gas compositions	GC-Gas analyzer	See page No.67
9.	HHV of product gas	GC-Gas analyzer	5,200 kJ/Nm ³
10.	Wood chip composition	Proximate/Ultime	See page No.54
11.	LHV of wood chip	Bomb calorimeter	12,500 kJ/kg

Feedstock; wood chip is examined. For each sample moisture content was determined gravimetrically by oven-drying method. Lower heating value (LHV) at a constant volume was measured using adiabatic oxygen bomb calorimeter. Proximate and ultimate analysis were determined by a chemical analysis laboratory using

standard analytical method. The proximate and ultimate fuel analysis of the samples is shown in Table 4.

Sensible heat of product gas

The sensible heat of product gas from the downdraft gasifier can be calculated.

$$\dot{Q} = \dot{m} C_p T \quad [\text{Eq.24}]$$

where

$$\begin{aligned} \dot{Q} &= \text{Heat rate, kW}_{\text{th}} \\ \dot{m} &= \text{Product gas mass flow rate, kg/s} \\ C_p &= \text{Specific heat capacity, kJ/kg}^\circ\text{C} \\ T &= \text{Temperature of product gas, }^\circ\text{C} \end{aligned}$$

It was found that the product gas flowing out of the downdraft gasifier with temperatures around 350°C at a flow rate of 32 Nm³ / hr. The sensible heat is equal to 3.41 kW_{th}. This sensible heat is a waste heat and can be reused again by boiling water into steam. Therefore, in this study, the heat exchanger is used to extract heat from the product gas for steam production.

CHAPTER IV

RESULTS AND DISCUSSION

Results

The results of the experiments are shown in Figure 24. The cold and hot fluid temperatures of shell and tube heat exchanger are shown in Figure 25. It is outstanding, considering the fact that the product gas composition over time seems to be constant over the experimental time as shown in Figure 26. Also, it can be seen that the steam catalyst can continuously be activated over the testing period for gas production. The bar graph in Figure 27. shows a comparison of the product gas composition resulting from two-different gasifying agents which are air and air-steam. It is clearly seen that the hydrogen content is 19% higher than of that without steam (11%). It can be summarized that the use of waste heat from the product gas increases hydrogen content without an external heat source.

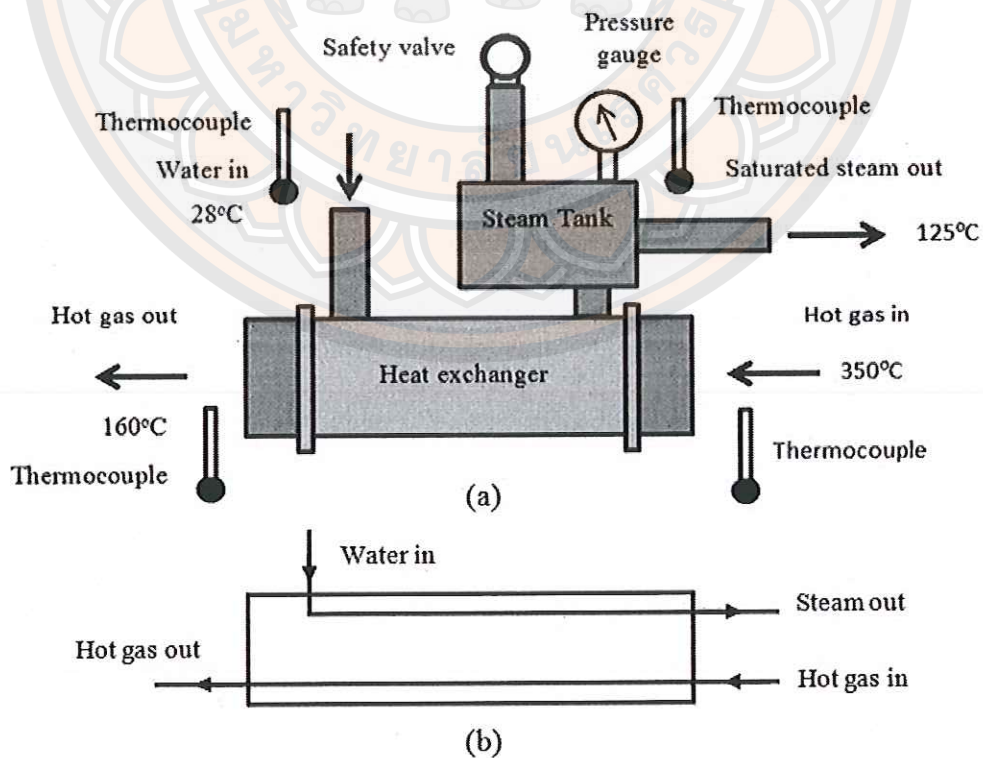


Figure 24 Schematic of saturated steam generation process

1. Shell and tube counter-flow heat exchanger: Temperature distribution

Table 6 Heat exchanger test results

Position NO.	Distance from the entrance, mm.	Temperature, °C	
		Cold fluid	Hot fluid
1	0	120	350
2	2	119	349
3	4	110	322
4	6	95	290
5	8	84	270
6	10	75	260
7	12	62	245
8	14	54	220
9	16	32	190
10	18	30	171
11	20	28	160

In the heat exchanger, water flows through the shell at the mass rate of 2.6 kg/h and its temperature increases from $T_{ci} = 28^\circ\text{C}$ to $T_{co} = 120^\circ\text{C}$ by the help of hot product gas with a flow rate of $32 \text{ Nm}^3/\text{h}$. The inlet and outlet temperature of the hot product gas are 350°C and 160°C , respectively. The outlet water vapor at temperature of 120°C , 1.2 bar flows to the gasifier as an oxidizing agent.

The variation of temperatures for both the hot product gas and water in a tube and shell heat exchanger is shown in Figure 26. It is noted that the hot product gas and cold water enter the heat exchanger from opposite ends. The hot gas temperature is decreased, the cold water will flow out at a higher temperature. The temperature difference at each end is very satisfactory.

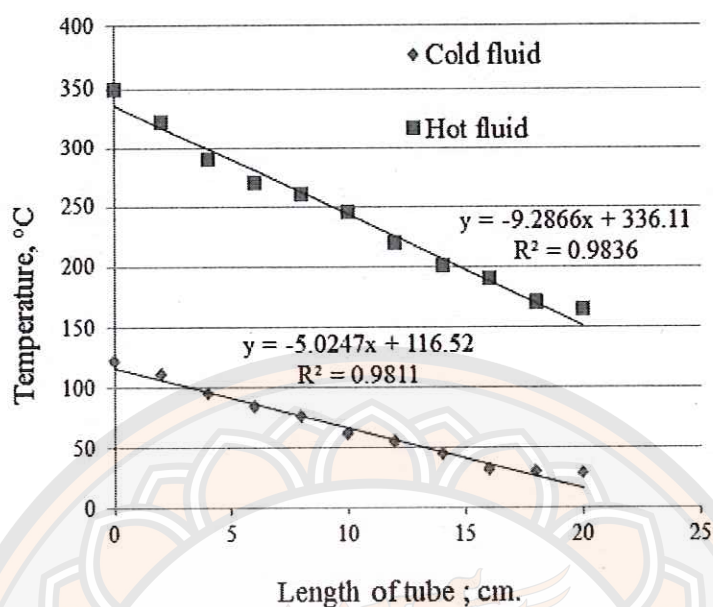


Figure 25 Counter-flow temperatures distribution

Table 7 Summary of heat exchanger test results

Test No.	Pressure, bar	Temp., °C	Steam production, kg/h	Steam/Sensible heat, (kg/h)/(kW _{th})	q, kW _{th}	q _{max} , kW _{th}	Effectiveness, \mathcal{E}
1	1.10	104	1.50	0.708	1.116	2.118	0.527
2	1.40	109	1.90	0.865	1.232	2.196	0.561
3	1.50	113	2.10	0.915	1.305	2.294	0.569
4	1.65	115	2.30	0.922	1.447	2.538	0.570
5	1.75	116	2.50	0.896	1.574	2.733	0.576
6	1.90	118	2.52	0.861	1.692	2.928	0.578
7	2.00	120	2.60	0.833	1.811	3.123	0.580

The experimental shows a result that when the flow rate of the product gas increases, the rate of steam generation increases and the effectiveness will gradually increase as well. The effectiveness with a maximum value of 0.58 is considered satisfactory because the heat exchanger is designed to save costs and it is easy to use.

Table 8 Summary of heat exchanger test results

Test No.	Waste heat flow rate, m ³ /h	Steam production, kg/h	Steam/Sensible heat, (kg/h)/(kW _{th})	Effectiveness, ϵ
1	22.0	1.50	0.708	0.527
2	22.5	1.90	0.865	0.561
3	23.5	2.10	0.915	0.569
4	26.0	2.30	0.922	0.570
5	28.0	2.50	0.896	0.576
6	30.0	2.52	0.861	0.578
7	32.0	2.60	0.833	0.580

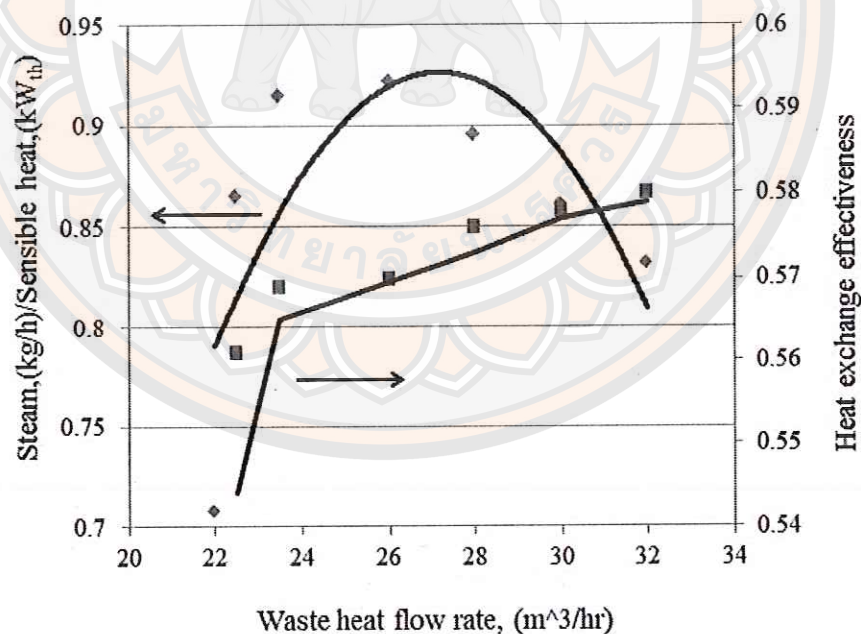


Figure 26 Heat exchanger effectiveness versus waste heat flow rate and Steam/Sensible heat versus waste heat flow rate

Figure 26 shows the heat exchanger effectiveness increases with the increase of waste heat. The effectiveness ranged from 0.52 to 0.58 when the waste heat flow rate was increased from 25 to 32 m³/h. The heat exchanger effectiveness has a maximum value of 0.58. While the maximum of steam-sensible heat ratio is 0.925. This heat exchanger can produce steam for a maximum value when the waste heat flow rate of 27 m³/h, the heat exchanger is appropriate for this gasifier.

2. Composition of product gas

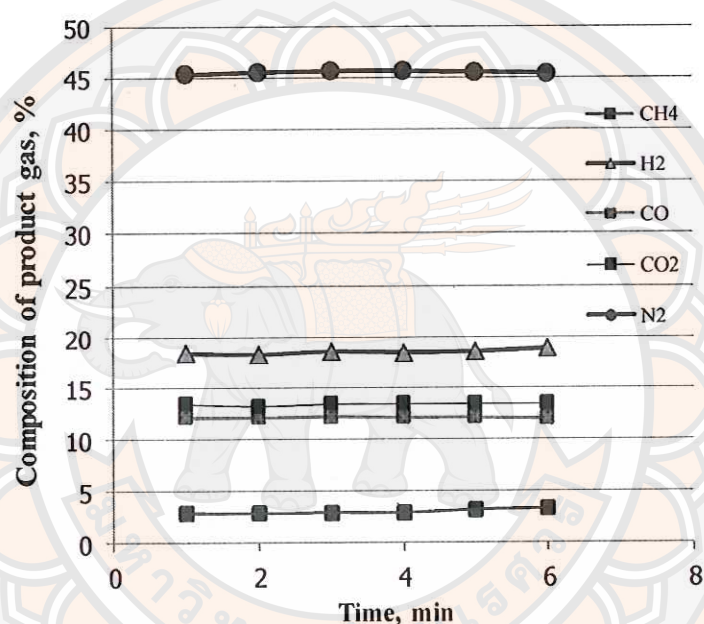


Figure 27 The gas composition versus time at 920°C and ER = 0.45

Figure 27 shows the composition of the product gas resulting from the air-steam gasification at 920°C and steam flow rate of 2.6 kg/h (ER = 0.45). Each value of the gas concentration is quite constant throughout the experiments.

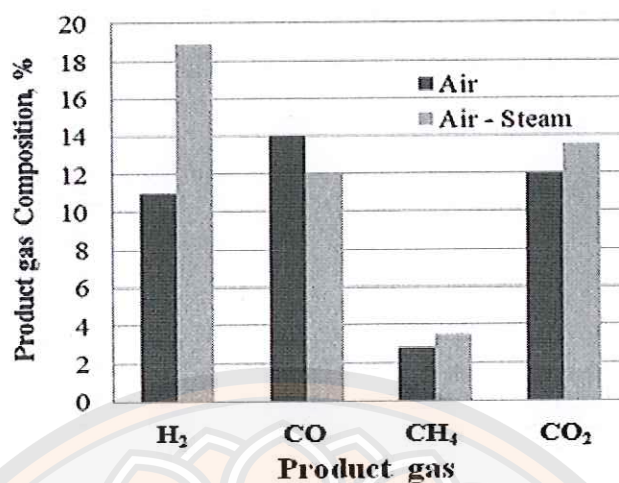


Figure 28 Comparison on gas composition between the use of air and air-steam as the gasifying agents. (ER = 0.45) at steam flow rate of 2.6 kg/h

Figure 28 shows a comparison of the product gas compositions between air and air-steam gasifications. It is noted that H₂ and CH₄ are increased at the rate of 8% and 2%, respectively, while CO is slightly decreased.

3. Composition of product gas and high heating value (HHV)

Table 9 Air and air-steam gasification test results (ER = 0.45, 920°C)

Time , min	Steam feeding , kg/h	Gas compositions, %					HHV, kJ/Nm ³	
		H ₂	CO	CH ₄	CO ₂	N ₂	Air	Air-steam
0	-	11.0	14.0	2.60	12.0	57.0	4,310	-
1	0.70	11.5	14.8	2.89	12.3	45.2		4,667
5	1.05	12.2	15.2	2.83	12.4	45.6		4,780
10	1.40	13.5	15.5	2.91	12.6	45.8		4,910
15	2.10	15.4	14.1	2.90	12.7	45.7		5,110
20	2.45	17.5	12.6	3.20	13.0	45.6		5,160
25	2.60	19.0	12.0	3.30	13.4	45.5		5,200

Table 9 shows the combination of experimental datas such as time, the feed rate of steam into gasifier, gas composition of product gas as well as the high heating value of the air gasification and air-steam gasification. The experiments will be fixed the ER = 0.45 and gasifying temperature at 920°C throughout the experiment.

The high heating value of the product gas is up to 5,200 kJ/Nm³ at steam feeding of 2.6 kg/h as shown in Figure 29.

The gas composition of each gas component increases, the only exception is for carbonmonoxide that it decreases. Because the steam is injected into the reactor while the water is broken down into hydrogen and oxygen. Oxygen then combines with carbon monoxide into carbon dioxide, carbon monoxide is decreased as shown in Figure 29.

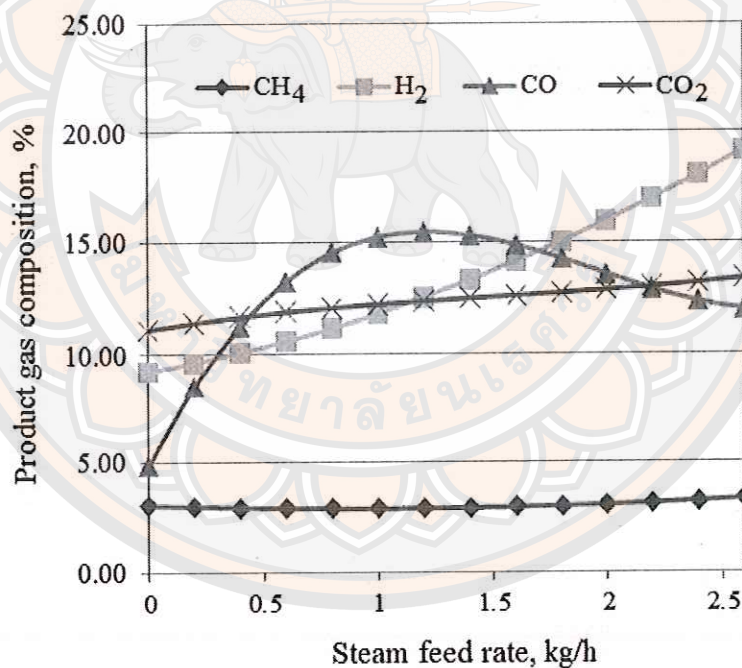


Figure 29 Effect of the steam on the product gas composition (ER = 0.45)

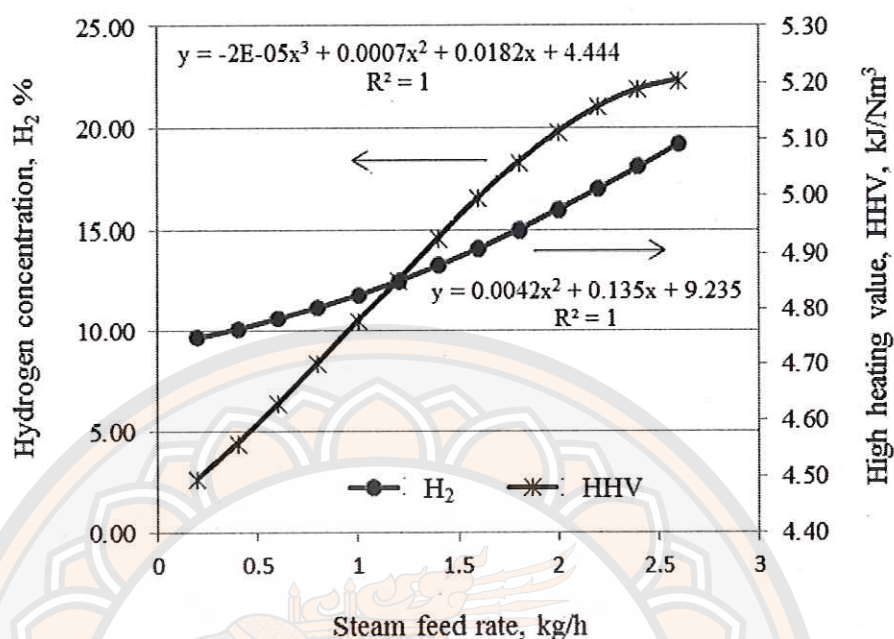


Figure 30 Effect of the steam on hydrogen concentration and high heating value (HHV) (ER = 0.45)

Figure 29 shows the influence of the steam on the dry gas yielded and the variation of the product gas composition at the outlet of the gasifier as a function of the steam flow rate. An increase in the steam flow rate results in a change in the product gas composition. At the steam flow rate greater than 1.46 kg/h, the molar fraction of CO starts to decrease, but H₂ and CH₄ are increased while steam flow rate is increased.

In addition, the HHV is also increased by the steam flow rate as shown in Figure 30. The experiments cannot increase the steam flow rate beyond 2.6 kg/h because waste heat is limited. It can be summarized that H₂ production of the air-steam gasification has been increased by 8% when compared to that of the air gasification. While the steam flow rate is changed from 0.7 to 2.6 kg/h, the HHV seems to be steadily increased.

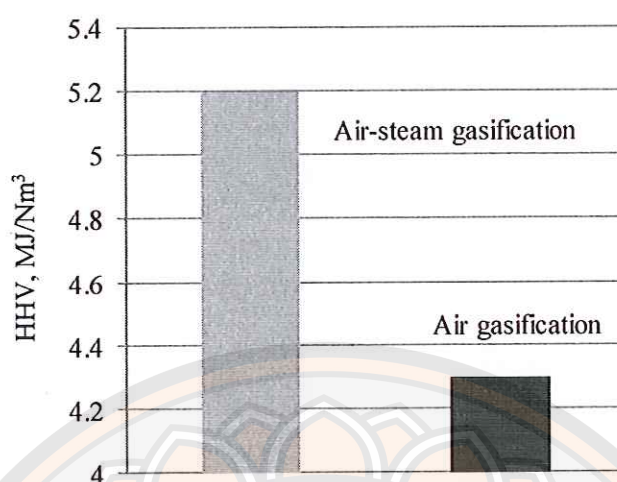


Figure 31 Comparison on high heating value (HHV) of the product gas between air gasification and air-steam gasification at steam flow rate of 2.6 kg/h (ER = 0.45)

Figure 31 shows the comparison on high heating value (HHV) of the product gas between the use of air and air-steam as the gasifying agents. The HHV of the product gas resulting from the air-steam gasification is at 4.7 to 5.2 MJ/Nm³, which is 16% higher than that resulting from the air gasification.

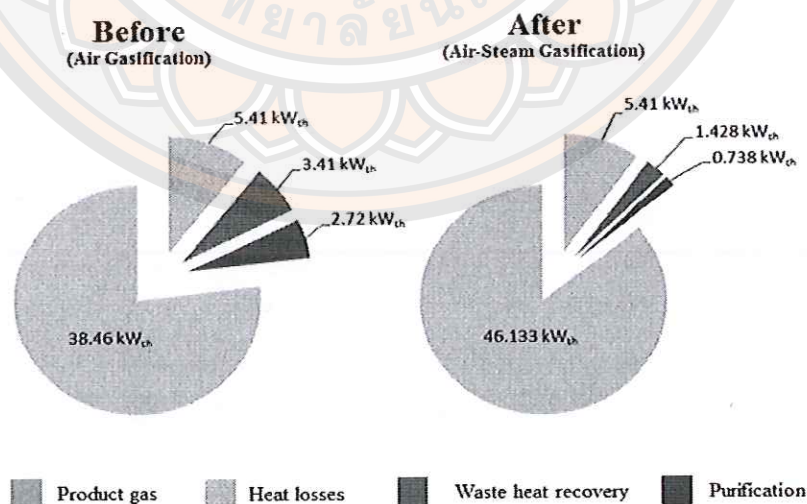


Figure 32 Comparison on energy balance of the system between the use Of air and air-steam as the gasifying agents. (ER=0.45) at Steam flow rate of 2.8 kg/h

Figure 32 after installing a heat exchanger to the old system. The system can produce steam as oxidizing agent effectively. The production of energy increased by approximately 20%. That is, the system can use the waste heat as well.

Table 10 Technical data of gasifier and heat exchanger system

Downdraft gasifier	50 kW _{th}
Biomass consumption	14 kg/h
Waste heat energy	3.41 kW _{th}
Heat exchanger	Shell and tube counter-flow type
- Heating surface	0.054 m ²
Steam-biomass ratio	0.2
Steam-sensible heat ratio	0.922 (kg/h)/(kW _{th})

Table 11 A summary of the performance of using the air-steam as an oxidizing agent

	Oxidizing agent	
	Air	Air-steam
Maximum steam flow rate [kg/h]	0	2.6
Reactor temperature [°C]	920	920
Biomass consumption [kg/h]	14	14
Product gas flow rate [Nm ³ /h]	32	32
Specific producer gas yield [Nm ³ /kg biomass]	2.28	2.28
H ₂ [%]	11	19
CO [%]	14	12
CH ₄ [%]	2.6	3.3
Producer gas HHV [kJ/Nm ³]	4,310	5,200
Cold gas efficiency, CGE [%]	73.87	87.75
Equivalent ratio, ER	0.45	0.45
The effectiveness of the heat exchanger, $\varepsilon = 0.58$		

Discussion

From this study it can be concluded as followed ;

In this experiment the biomass-feeding rate is 14 kg/h. whereas the steam flow rate has been increased from 0.7 to 2.60 kg/h, the equivalent ratio is fixed at 0.45 and the gasifying temperature is also fixed at 920°C.

The product gas composition graph shows the percentage of CO decreases, but CO₂ and CH₄ slightly increase, while H₂ percentage increases from 11% to 19%. and N₂ almost remains constant (Figure 30).

The product gas composition that has shown HHV value ranging 4,310 kJ/Nm³ (Figure 32.) for air gasification. In case of air-steam gasification, it can be said that an increase in the steam flow rate resulted in an increase in the percentage of the HHV value ranging 5,200 kJ/Nm³. The addition of Steam into the gasifier increases the amount of hydrogen molecules inside the gasifier, and this increase may have increased the H₂, CO, CH₄ volume fraction in the product gas. The increase in the percentage fraction of H₂, CO, and CH₄ also enhances the HHV of production gas.

The economic analysis

The economic analysis of the integration of heat exchanger devices into the original biomass gasifier systems will be discussed here. To compare the investment value between the original biomass gasifier system, the heat exchanger for steam generation system and the integrated system combining the heat exchanger into the original gasifier system, these economic indicators are needed to be considered as followed ;

1. Payback period (PB).
2. Net present value (NPV).
3. Internal rate of return (IRR).
4. Cost of energy (COE).

1. A payback period (PB) can be calculated from initial investment and cash flow to the investor. In this research, only a formula for continuous and stable cash flow will be considered, as followed;

$$\text{Payback period (PB)} = \frac{\text{Initial investment}}{\text{Cash flow per period}} \quad [\text{Eq.25}]$$

The benefits of the PB analysis are that, investors can calculate whether it is economically reasonable to invest in a particular project and how long the payback period that project would take. The PB analysis can indicate whether the project has high successful rate or not. However, the PB analysis also has disadvantages, because the value of money can change over time, there is a risk that cash flow to the investor in the future will decrease and it would extend a payback period. Moreover, there is no any criteria that can indicate whether the investment will increase the value of the project or not.

2. The Net Present Value (NPV) is a method to evaluate the desirability of the investment in a particular project. It can be defined as followed ;

$$NPV = \frac{\sum_{t=1}^n B_t - \sum_{t=1}^n C_t}{(1+i)^t} - TIC \quad [\text{Eq.26}]$$

where,

- B_t = The expected benefit at the end of year n.
 C_t = The expected cost at the end of year n.
 TIC = The total initial investment.
 i = The discount rate or the minimum annual return required to finance the project.
 t = The time of the cash flow. (in years)
 n = The total time of evaluation. (in years)

3. The Internal rate of return (IRR) is another method to measure and compare the profitability of investments. The IRR is defined as the rate of the return that makes the NPV of all cash flows both negatively and positively from a particular investment equals to zero.

$$NPV = \sum_{n=1}^N \frac{B_n - C_n}{(1+r)^n} - TIC = 0 \quad [\text{Eq.27}]$$

where

r = The internal rate of return (IRR).

The economic analysis of the original biomass gasifier system, the thermoelectric power generation system and the integrated system combining the thermoelectric power generation into the original gasifier system are shown in the Table 6-8. The economic life of the project, the expected period of time during in which an asset is useful to the owner, is 15 years with a discount rate of 8%.

4. Cost of energy (COE).

This cost is the most comprehensive measure of the thermal power generation system, which is the cost of energy (COE). This measure incorporates all elements of costs, i.e. installed capital cost, cost of operation and maintenance, cost of major overhauls and subsystem replacements.

$$\text{COE} = \frac{\text{ICC} + \text{FCR} + \text{O\&M} + \text{LRC}}{\text{Total energy generation (kWth)per year}} \quad [\text{Eq.28}]$$

where

- ICC = Installed capital cost
- FCR = Annual fixed charge rate
- O&M = Operation and maintenance cost
- LRC = Levelised replacement cost

Table 12 Assumptions for the economic analysis

No.	Assumptions	
1	The integrated thermal system life	15 years
2	Interest rate	7 %
3	Operation and maintenance cost	10% of the initial investment
4	Biomass feedstock	Wood chip
	- Wood chip price	1,200 THB/ton
5	Proximate analysis	
	Moisture content	25 %
	Ash	3 %
	Volatile	54 %
	Fixed carbon	18 %
	Heating value	12,800 kJ/kg
6	Ultimate analysis	
	C	49 %
	H	4 %
	O	25 %
	N	0.2 %
	S	0.0 %
7	Biomass feed capacity	14 kg/h
8	Price of LPG	25 THB/kg
9	Gasifier design parameters	
	Gasification temperature	920°C
	Equivalent ratio (ER)	0.45
	Steam to biomass ratio (SB ratio)	0.20

Table 13 The economic analysis of the original biomass gasifier system

Capital investment	The original biomass gasifier system (THB)	Detail
1) Initial investment	650,000	Thermal Power Output = 50 kWth
2) Biomass fuel cost	42,000	Wood chips price = 1,200 THB/Ton
3) Operation and maintenance cost	65,000	10% of the initial investment
4) Benefit (per year)	225,000	Price of LPG = 25 THB/kg

Table 14 The economic analysis of the heat exchanger/steam generation system

Capital investment	Heat exchanger /steam generation system (THB)	Detail
1) Initial investment	35,000	Shell and tube heat exchanger 1 set , steam capacity 2.6 kg/h (max.)
2) Fuel cost	-	
3) Operation and maintenance cost	3,500	10% of the initial investment
4) Benefit (per year)	6,609	Price of wood chip = 1.2 HB/kg

The economic life of the biomass air-steam gasifier system is 15 years and has 8% discount rate. The results of the comparison between the original biomass gasifier system and, the heat exchanger/steam generation gasifier system are shown as followed;

Table 15 The economic analysis of the integrated system combining the heat exchanger/steam generation into the original gasifier system

Capital investment	Integrated system (THB)	Detail
1) Initial investment	685,000	Volume = 1 set Heat exchanger /steam generation system
2) Fuel cost	50,400	
3) Operation and maintenance cost	68,500	10% of the initial investment
4) Benefit (per year)	265,000	Price of LPG = 25 THB/kg

Table 16 The comparison of the economic analysis between each system

Economic Indicator	The original biomass gasifier system	The integrated system
Net Present Value (NPV)	1,168,865 THB	1,464,341 THB
Cost of energy (COE)	6.60 THB/kW _{th} h	5.80 THB/kW _{th} h
Internal Rate of Return (IRR)	29.72 %	32.96 %
Payback period (PB)	3 years 5 months	3 years 1 months

According to the results shown in Table 15, the cost of energy of the integrated system is lower than the original biomass gasifier system and the net present value is positive. In addition, the pay back period of the integrated system is less than the original biomass gasifier system as well. Thus, the integrated system is a good choice for investment.

CHAPTER V

CONCLUSION AND RECOMMENDATION

Conclusion

The use of energy from fossil would only cause more environmental pollution. Fuel gas produced by biomass gasification technology has a benefit of low heating value and hydrogen quantity is low as well. Therefore, using steam to solve such problems by using a heat exchanger to extract waste heat to produce steam by itself is an interesting way. The waste heat from the hot gas is used to design the heat exchanger. Heat exchanger changes water into steam and is reheated by the hot gas again until the temperature reaches 580°C and then flows into the reactor. From the experimental results, the results are satisfactory, although hydrogen and high heating value is not very high.

The majors conclusions are as follows:

1. The amount of HHV of product gas varied from 4,310 to 5,200 kJ/Nm³. From the experiment, in which 14 kg/h biomass feeding rate and 2.6 kg/h of steam flow rate is maintained to give the maximum HHV of 5,200 kJ/Nm³. It is observed that the increase in steam flow rate increases the HHV of the product gas.
2. The results of the experiment show that the waste heat can cause high heating value up to 5,200 kJ / Nm³ and hydrogen content up to 19%, but not high, because there has been a limit.
3. The exhaust gas temperature of the gasifier was around 350 °C and 3.41 kW_{th} waste heat is available for energy efficient like steam generation.
4. The economic analysis shows that the system has a payback period of 11 months and IRR is 16.09%. Therefore it is appropriate to encourage the installation of heat exchanger to be part of the system.

Recommendation

1. Study and install a fire tube heat exchanger inside the furnace due to the higher temperature of the syngas and to reduce heat loss to the environment.
2. Determine the effect of gasifying temperature when the heat exchanger is mounted inside the gasifier.
3. In addition to the use of steam, it should be combined with other systems, such as air preheat, to make the system more efficient.





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APPENDIX

มหาวิทยาลัยรัตนนคร

APPENDIX A THE CALCULATION OF TOTAL HEAT LOSSES FROM THE GASIFIER

The calculation of total heat losses from the gasifier

Total heat losses from the gasifier = Top loss + side loss + bottom loss + ash loss

$$\begin{aligned} \text{Top loss} &= hA(T_s - T_\infty) + \epsilon\sigma A(T_s^4 - T_\infty^4) \\ &= (10.45)(0.237)(90 - 30) + (0.8)(5.67 \times 10^{-8})(0.237)[(363)^4 - (303)^4] \end{aligned}$$

$$148.60 + 96.00$$

$$= 244.60 \text{ watts}$$

$$\begin{aligned} \text{Bottom losses} &= hA(T_s - T_\infty) + \epsilon\sigma A(T_s^4 - T_\infty^4) \\ &= (10.45)(0.197)(110 - 30) + (0.8)(5.67 \times 10^{-8})(0.197)[(383)^4 - (303)^4] \\ &= 164.69 + 116.94 \end{aligned}$$

$$= 281.6 \text{ watts}$$

$$\begin{aligned} \text{Side loss} &= hA(T_s - T_\infty) + \epsilon\sigma A(T_s^4 - T_\infty^4) \\ &= (10.45)(1.727)(170 - 30) + (0.8)(5.67 \times 10^{-8})(1.727)[(443)^4 - (303)^4] \end{aligned}$$

$$2,526.6 + 2356.37$$

$$= 4,882.96 \text{ watts}$$

$$\text{Thus} = 5,409.16 \text{ watts}$$

$$= 5.409 \text{ kW}_{th}$$

- Notes
- h = convective heat transfer coefficient, W/m^2K
 $= 10.45 - V + 10V^{1/2}$, V = wind velocity, $m/s \approx 0$
 $= 10.45 \text{ W/m}^2K$
 - ϵ = emissivity of surface = 0.8
 - T_∞ = ambient temperature, $^\circ C$
 $= 30^\circ C$
 - T_s = surface temperature, $^\circ C$

APPENDIX B THE CALCULATION OF HEAT EXCHANGER DESIGN

Preliminary Test : Sensible heat of the product gas

The sensible heat of the product gas from the downdraft gasifier can be calculated as followed.

$$\dot{Q} = \dot{m} C_p T \quad [\text{Eq.22}]$$

where

\dot{Q} = Heat rate, kW_{th}

\dot{m} = Product gas mass flow rate, kg/s

C_p = Specific heat capacity, kJ/kg°C

T = Temperature of product gas, °C

It was found that the product gas was flowing out from the downdraft gasifier with the temperatures around 350°C at a flow rate of 32 Nm³/h. The sensible heat equals to 3.41 kW_{th}. This sensible heat is a waste heat and can be reused by boiling water into steam. Therefore, in this study, the heat exchanger is used to extract the heat from the product gas for the steam production.

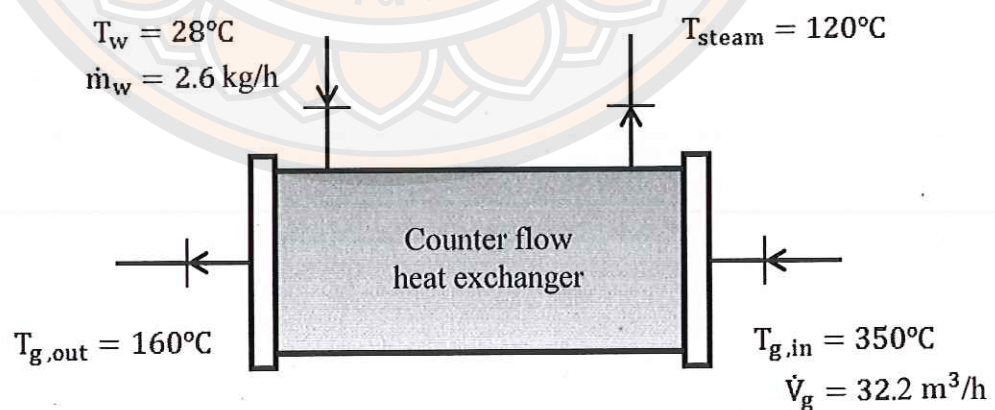


Figure 33 Energy and mass balance of the counter flow heat exchanger.

$$\dot{Q} = \dot{m} C_p T$$

$$\begin{aligned}\dot{Q}_{th} &= \left(\frac{32}{3,600}\right) \left(\frac{\text{kg}}{\text{hr}}\right) \left(\frac{\text{hr}}{\text{s}}\right) (1.004) \left(\frac{\text{kJ}}{\text{kg}^\circ\text{C}}\right) (350^\circ\text{C}) \\ &= 3.23 \text{ kwatts}\end{aligned}$$

$$\text{Assumed effectiveness} = 0.65$$

$$\begin{aligned}\dot{Q}_{th,actual} &= (3.23)(0.65) \\ &= 2.099 \text{ kwatts}\end{aligned}$$

Energy balance

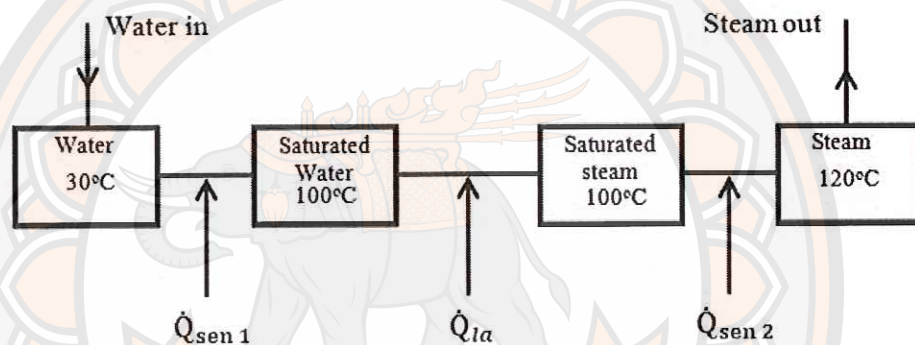


Figure 34 Energy balance of water vaporization.

$$\begin{aligned}\dot{Q}_{sen,1} &= \dot{m} C_p \Delta T \\ &= \left(\frac{2.6}{3,600}\right) \left(\frac{\text{kg}}{\text{h}}\right) \left(\frac{\text{h}}{\text{s}}\right) \left(4.18 \frac{\text{kJ}}{\text{kg} \cdot \text{K}}\right) (70 \text{ K}) \\ &= 227.5 \text{ watts}\end{aligned}$$

$$\begin{aligned}\dot{Q}_{la} &= \dot{m} L \\ &= \left(\frac{2.6}{3,600}\right) \left(\frac{\text{kg}}{\text{h}}\right) \left(\frac{\text{h}}{\text{s}}\right) \left(2,250 \frac{\text{kJ}}{\text{kg}}\right) \\ &= 1,750 \text{ watts}\end{aligned}$$

$$\begin{aligned}\dot{Q}_{sen,2} &= \dot{m} C_p \Delta T \\ &= \left(\frac{2.6}{3,600}\right) \left(\frac{\text{kg}}{\text{h}}\right) \left(\frac{\text{h}}{\text{s}}\right) \left(4.18 \frac{\text{kJ}}{\text{kg} \cdot \text{K}}\right) (20 \text{ K}) \\ &= 65 \text{ watts}\end{aligned}$$

$$\dot{Q}_{\text{Total}} = 2,042.5 \text{ watts.}$$

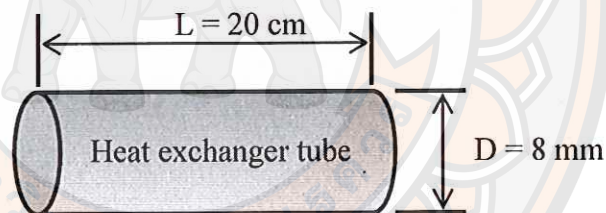
$$T_{\text{exit}} = 170^\circ\text{C (From experiment)}$$

The LMTD method

$$\begin{aligned} \Delta T_m &= \frac{(160 - 120) - (350 - 28)}{\ln \left[\frac{(160 - 120)}{(350 - 28)} \right]} \\ &= 145.47^\circ\text{C} \end{aligned}$$

$$\begin{aligned} A &= \frac{(\dot{m}C_p)_c \Delta T_c}{U \Delta T_m} \\ &= \frac{\left(2.6 \frac{\text{kg}}{\text{h}}\right) \left(4.18 \frac{\text{kJ}}{\text{kg}}\right) (90 \text{ K})}{\left(72 \frac{\text{W}}{\text{m}^2 \cdot \text{K}}\right) (145.47^\circ\text{C})} \end{aligned}$$

$$\begin{aligned} \text{The surface area, } A &= 0.035 \text{ m}^2 \\ &= 350 \text{ cm}^2 \end{aligned}$$



$$\begin{aligned} A &= \pi(DL) = \pi(0.008 \text{ m})(0.20 \text{ m}) \\ &= 50.24 \text{ cm}^2 \end{aligned}$$

$$\begin{aligned} \text{The number of tubes} &= \frac{350}{50.24} \\ &= 6.996 \\ &= 7.0 \end{aligned}$$

Assumptions

1. Negligible heat transfers between heat exchanger and surroundings and negligible kinetic and potential energy changes.
2. Tube internal flow and thermal conditions are fully developed.
3. Negligible thermal resistance of tube material and fouling effects
4. Constant properties.

Schematic diagram :

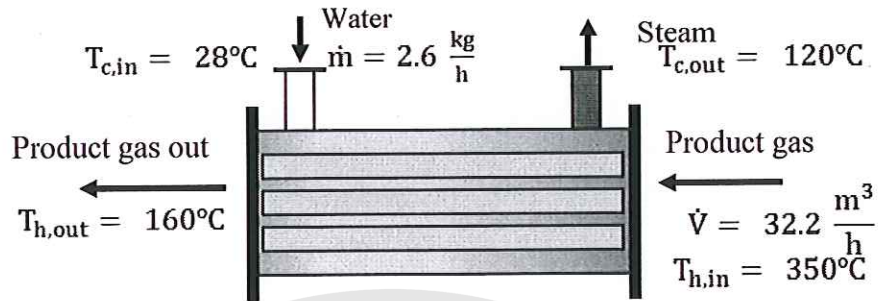


Figure 35 Counter flow heat exchanger : One shell with one tube pass

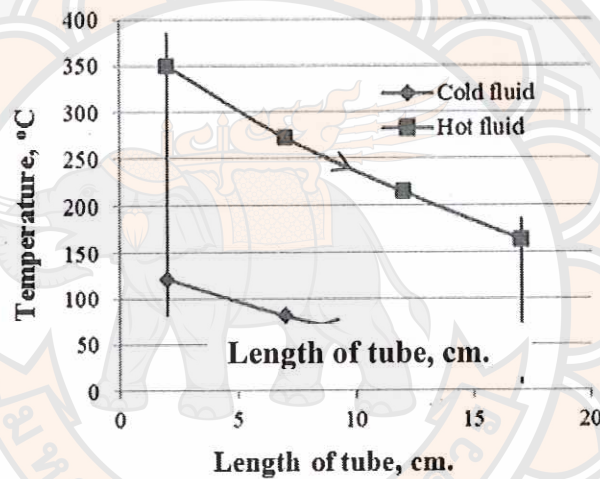


Figure 36 Counter flow heat exchanger temperatures distribution

Properties : Water ($T_c = 30^\circ\text{C}$) : $C_{p,c} = 4.197 \text{ kJ/kg.K}$

Product gas : $C_{p,h} = 1.004 \text{ kJ/kg.K}$.

Using the LMTD method,

$$\dot{Q} = UA \Delta T_{LM}$$

where $A = N \times L \times \pi D$

The overall heat transfer coefficient, U for a wall or heat exchanger can be calculated as:

$$\frac{1}{UA} = \frac{1}{h_i A_i} + \frac{dx_w}{kA} + \frac{1}{h_o A_o} \quad [\text{Eq.8}]$$

where

U = the overall heat transfer coefficient, ($\text{W}/\text{m}^2\text{K}$)

A = the contact area for each fluid side (m^2)

K = the thermal conductivity of the material ($\text{W}/\text{m.K}$)

h = the individual convection heat transfer coefficient for each fluid

($\text{W}/\text{m}^2\text{K}$)

dx_w = the wall thickness (m)

The calculation of HHV and LHV heating value of the product gas

Heating value (HV) of the product gas is determined from the chemical composition of the gas and HV of individual components.

Table 17 Gas components and concentration of the product gas

Gas Components	Concentration, %
CO	12
H ₂	19
CH ₄	3.5

From [Eq.16]

$$\begin{aligned} \text{HHV} &= (\text{H}_2\% \times 30.52 + \text{CO}\% \times 30.18 + \text{CH}_4\% \times 95.0) \times 4.2 \\ &= (19 \times 30.52 + 12 \times 30.18 + 3.5 \times 95.0) \times 4.2 \\ &= 5,353 \text{ kJ}/\text{Nm}^3 \end{aligned}$$

From [Eq.17]

$$\begin{aligned} \text{LHV} &= (\text{H}_2\% \times 25.70 + \text{CO}\% \times 30.00 + \text{CH}_4\% \times 85.5) \times 4.2 \\ &= (19 \times 25.70 + 12 \times 30.00 + 3.5 \times 85.5) \times 4.2 \\ &= 4,819 \text{ kJ}/\text{Nm}^3 \end{aligned}$$

Table 18 A comparison of heating value of the product gas in various ways

Heating value	Eq.[16]	Eq.[17]	GC.
HHV. (MJ/Nm ³)	5,353		5,191
LHV. (MJ/Nm ³)		4,819	4,731



Figure 37 Gas analyzer used to measure the properties of the product gas

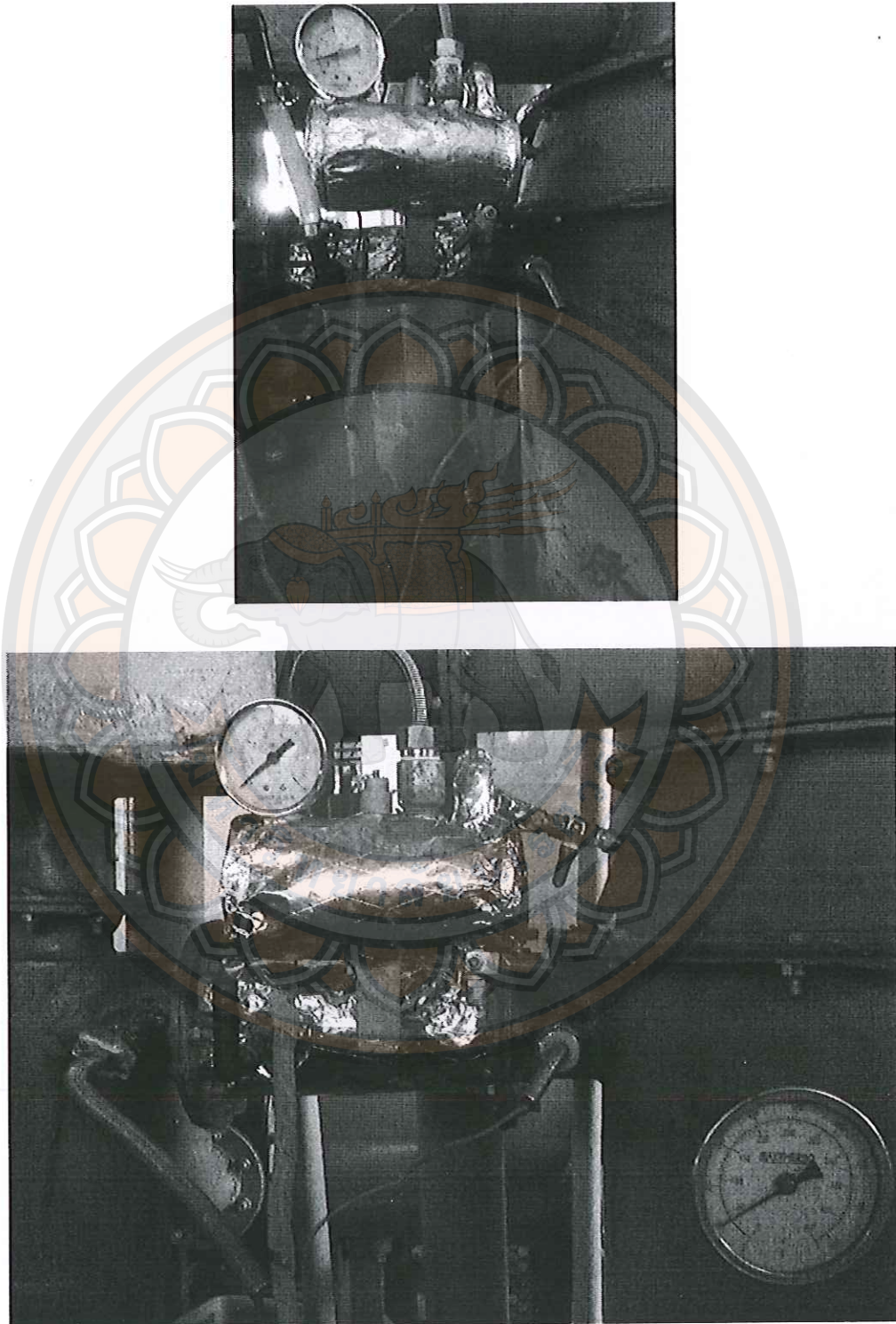


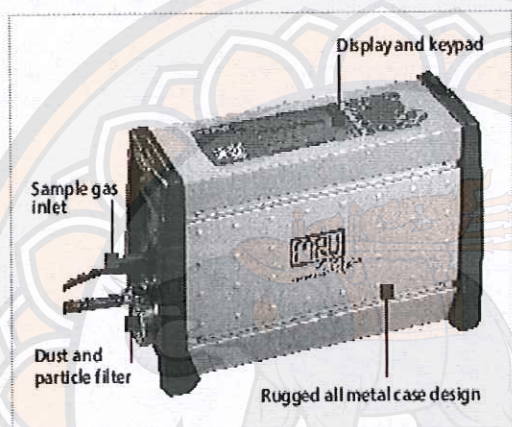
Figure 38 Steam generation by a small shell and tube heat exchanger

VARIO plus

Mobile process gas analyzer

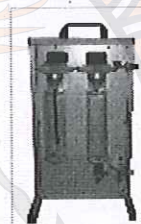
Electrochemical sensors and IR-Modules are combined in the **VARIOplus** Industrial. The multi component analyzer is the solution where-ever efficient solutions are requested. This unit delivers precise measuring results using a NDIR-module for CO + CO₂ + CH₄, a TCD sensor for H₂ and electro-chemical sensor for O₂.

Applications:
BIOMASS GASIFICATION
COAL GASIFICATION
STEEL INDUSTRY
STEAM REFORMING



Equipment configuration

Modern, state of the art all metal enclosure with robust and shock proof plastic side panels. Microprocessor controlled and monitored processes. Flow rate monitoring, zero calibration, RS485 Data transfer, analog output, integrated electrical gas cooler and automatic condensate peristaltic pump



Sample gas conditioning unit to separate tar and dirt removal from sample gas

Technical specifications

Measured components	measuring range:	accuracy:	measuring cell:
Oxygen O ₂	Ox 21 %	±0,2 Vol.-% abs.	electrochemical long-life
3-gas infrared bench	min. meas. range:	max. meas. range:	linearity error:
Carbon monoxide CO	Ox 3 %	Ox 100 %	3 % of full scale
Carbon dioxide CO ₂	Ox 3 %	Ox 100 %	3 % of full scale
Hydrocarbons CH ₄ (Methane)	Ox 3 %	Ox 100 %	3 % of full scale
Thermal conductivity detector	min. meas. range:	max. meas. range:	linearity error:
Hydrogen H ₂	Ox 1 %	Ox 100 %	2 % of full scale
Repeatability	1 % of smallest measuring range		
Calibration	by software, calibration gases, instrument air or clean ambient air for auto zero		
Operating- /storage temperature	+ 5 °C x +40 °C, max. 90 % rh, non condensing / -20 °C x +50 °C		
Ambient conditions	not for use in aggressive, corrosive or very high dust atmosphere hazardous areas use only with special equipment (on request)		
Data transfer	8-channel analog output 4... 20 mA, RS 485 digital (modbus RTU)		
Mains	110 ... 230 Vac / 50 ... 60 Hz / 100 W		
Dimensions / Weight / Protection class	530 x 490 x 310 mm (W x H x D) / 7 kg, IP 21		

Data subject to change without notice.

Dealer:



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Figure 39 Technical specifications of mobile gas analyzer