CHAPTER 4
RESULTS AND DISCUSSION

4.1 Raw Rice Husk Analysis

4.1.1 Ultimate analyses and calorific values

Table 4.1 shows the result of the ultimate analysis and the calorific values of rice husks.

Table 4.1 Ultimate analyses (wt%, daf.) and calorific values of rice husk

Biomass	Ultimate analysis (%, daf.)				HHV (MJ/kg, db.)
	C	H	N	O	
Rice Husk	45.9	6.2	0.6	47.3	15.65

4.1.2 Proximate Analysis

Figure 4.1 shows the TGA curve of rice husks. The TGA curve is used to analyze the changing in weight fraction of rice husks when it is heated. Rice husks were heated in nitrogen atmosphere from room temperature to 900°C at the rate of 10°C/min. The weight lost during this step refers to the weight fraction of volatile matter in rice husks. When the temperature reached 900°C, the nitrogen stream was switched to air stream in order to combust the carbon. The weight loss at this step refers to the weight fraction of fixed carbon. Weight fraction of ash was identified from the weight fraction of remaining solid which is noncombustible.

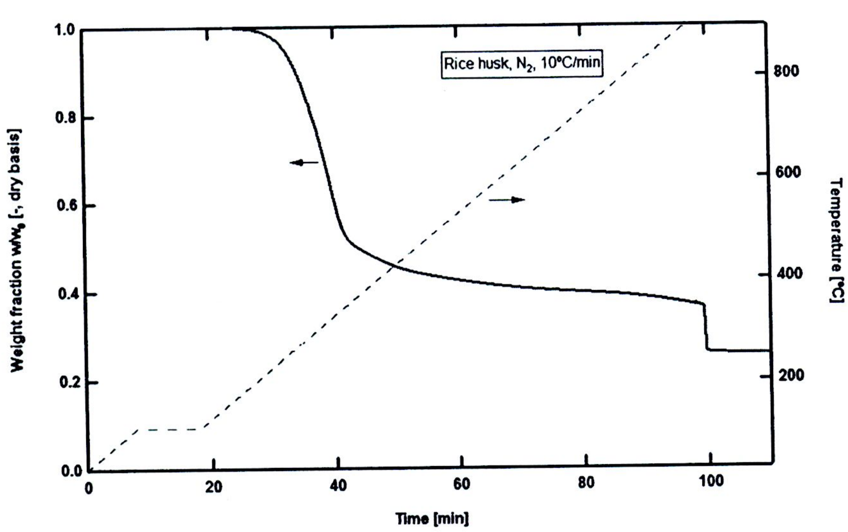


Figure 4.1 TGA curve of rice husk

Table 4.2 shows the proximate analysis results of the TGA curve analysis. Volatile matter is the majority content of rice husk which has the percentage of volatile matter (63.7%, db.), fixed carbon (10.5%, db.) and ash (25.8 %, db.).

Table 4.2 Proximate analyses (%, db.) of rice husk

Biomass	Proximate analysis (%, db.)		
	Volatile matter	Fixed carbon	Ash
Rice Husk	63.7	10.5	25.8

4.1.3 Ash Analysis

Table 4.3 shows the components of the rice husk ash. The main component is silicon (94.81% as SiO_2), followed by potassium (2.99% as K_2O) and calcium (0.74% as CaO).

Table 4.3 The component of the rice husk ash

Biomass	Components (wt%)							
	SiO_2	K_2O	CaO	P_2O_5	Fe_2O_3	MgO	Cr_2O_3	ZnO
Rice Husk	94.81	2.99	0.74	0.51	0.26	0.15	0.05	0.02

4.2 Minimum Fluidization Velocity

This experiment was tested to find the minimum fluidization velocity. The bed material used in all experiments was silica sand with a particle density of 2433 kg/m^3 and average particle size of 250 microns. Ambient air was applied to the reactor under the air distributor plate to investigate the relation between pressure drop across the bed, ΔP and superficial velocity at a static bed height of 20, 30, 40 cm. The results from the experiment could be used to set a minimum superficial air velocity which was used to calculate the size of the reactor.

At the beginning of the experiment, the fluidization column was charged with bed up to a static bed height of 20, 30, 40 cm, then started the pump. During the experiments, the air superficial velocity was increased from zero up to a maximum value. Figures 4.2 – 4.4 show the relationship between pressure drop across the bed at different superficial velocities at a static bed height of 20, 30, 40 cm, respectively. At the first stage, ΔP increased significantly with increasing superficial velocity. Until a fluidization velocity was reached, the point of velocity which ΔP started constant and continued to rather constant afterward.

In the experiment, the minimum fluidization velocity of a static bed height of 20, 30, 40 cm were about 0.11, 0.12, 0.12 m/s, respectively. The value was found to be mostly close to the empirical correlation proposed by the research of fluidization correlations for coal gasification materials: minimum fluidization velocity and bed expansion ratio (Babu *et al.*, 1978). While the value relatively high when compared to other research (Rozainee *et al.*, 2008). The observed U_{mf} was then used to set the lower limit of the air feed rate.

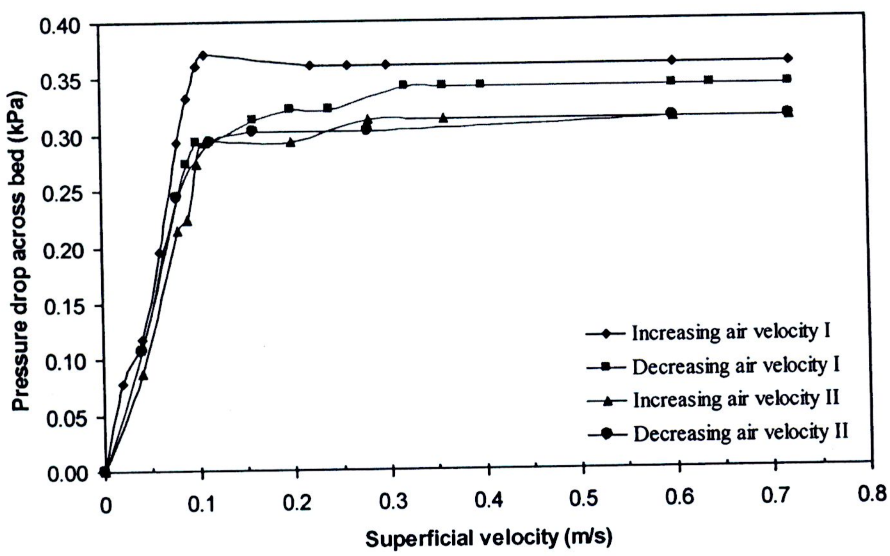


Figure 4.2 Relationship between pressure drop across bed and superficial velocities of a static bed height 20 cm

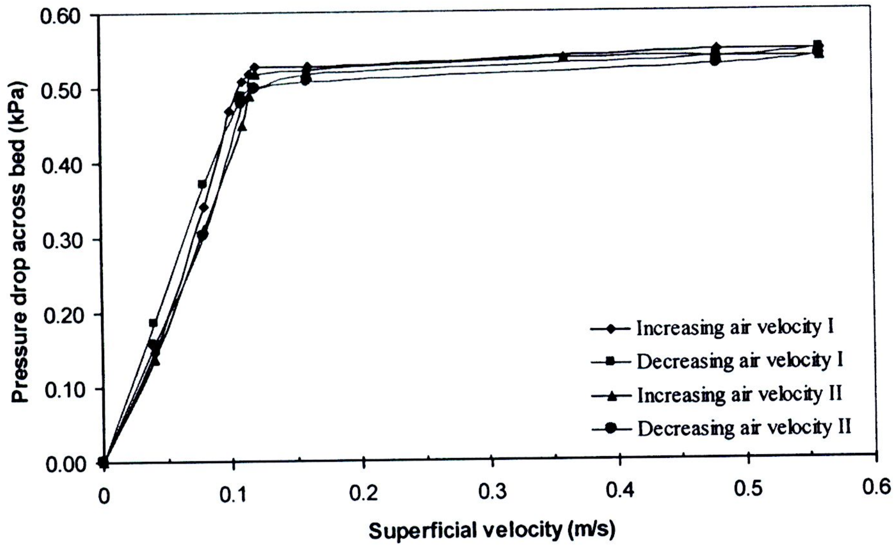


Figure 4.3 Relationship between pressure drop across bed and superficial velocities of a static bed height 30 cm

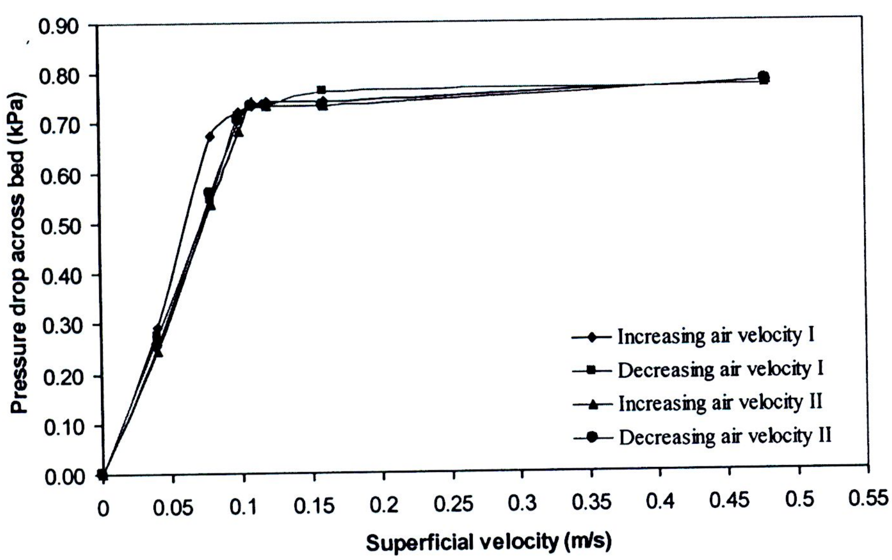


Figure 4.4 Relationship between pressure drop across bed and superficial velocities of a static bed height 40 cm

4.3 Two-Stage Gasification System Equipment

4.3.1 Biomass Feeding System

From studying the characteristics of rice husks, the material class code of rice husks is C-25 (Appendix C) that a cross-sectional loading area is 45%. A cross-sectional loading area from the material class matches with the information from the Materials Handling Handbook that factor of trough loading of rice husk is 0.45. Mass flow rate of rice husk is 0.3 kg/min from calculation. Diameter of shaft and speed of screw, which were defined, are 0.020 m and 150 rpm, respectively. A defined pitch equals to diameter of screw. From equation (3.1), diameter of screw is 0.0404 m or rounded-off to 0.041 m. According to clearance between the screw and the radius of the inside of the screw trough is around 12 – 15 mm, thus the diameter of the inside of the screw trough is 0.067 m. As screw trough is designed as pipe, the selected internal diameter of pipe is equal or higher than 0.067 m. The selected pipe is NPS 2½ with SCH 5 which the outside diameter is 0.07302 m, the internal diameter is 0.0688 m, and the wall thickness is 2.108 mm. The drawing of screw conveyor is shown in Figure 4.5.

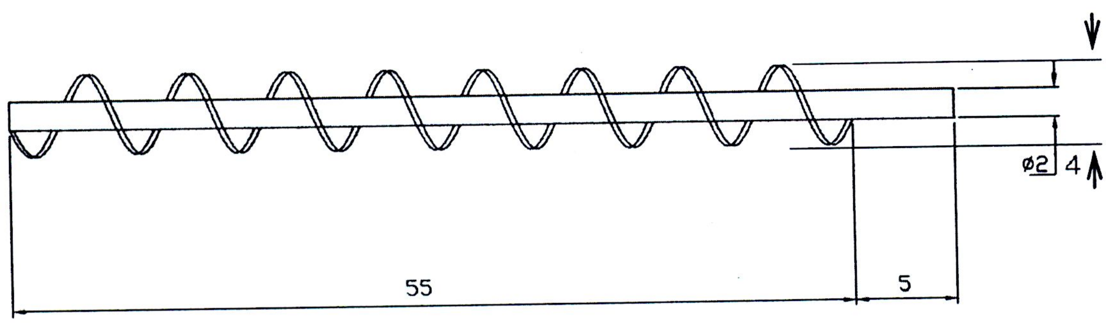


Figure 4.5 The drawing of the designed screw (unit:cm)

4.3.2 Air Supplied System

Primary Air Supplied

The primary air supplied, which was designed as a conical spouted bed type, consisting of an inlet air nozzle located at the center of the bottom cone, a funnel-shape structure through which the air passes and is distributed uniformly into the reactor, as shown in Figure 4.6. Spouted bed is a technique to make the fluid and particles contact each other. The principle is similar to a fluidized bed. The difference between fluidized bed and spouted bed lies in the dynamic behavior of the solid particles (Salam and Bhattacharya, 2006). In a fluidized bed, air is passed through a uniform distributor plate to float the particles which move up and down in groups. In a spouted bed, air enters the bed through a small orifice at the center of a conical or flat base, instead of a uniform distributor, resulting in a systematic cyclic pattern of solid movement inside the bed as shown in Figure 4.7.

The advantages of a spouted bed are:

1. Fluid and bed materials, which velocity is lower than minimum fluidization, contact with each other extensively.
2. The bottom of the spouted bed chamber is cone, it reduces the occurrence of airless, accumulated particles and no flow, which is found in the case of fluidized bed that bottom of chamber is flat.

In order to get good bed stability, operating conditions must be established so that a feeding gas velocity above 1.7 times the minimum spouting velocity (V_{ms}) are obtained (Bilbao *et al.*, 1987).

The diameter of the tuyer orifice was calculated from the ratio of a pilot plant which used a conical spouted bed instead of a distributor plate to solve the problem of unmixing of rice husk and silica sand. The parameters considered necessary for cone design are the air velocity at nozzle and the tuyer orifice diameter, which are 5.4-6.8 m/s and 0.02 m, respectively.

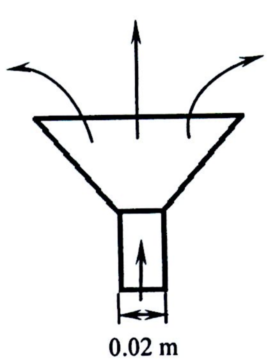


Figure 4.6 A conical spouted bed for pyrolysis stage

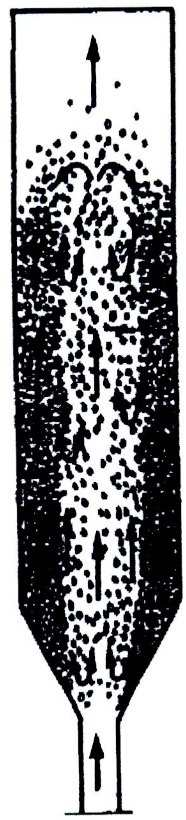


Figure 4.7 Diagramatic representation of the particles movement in a classical spouted bed (Bilbao *et al.*, 1987)

Secondary Air Supplied

Air is supplied from the air tuyers placed radially around the circumference of the partial oxidation region. The air tuyers are designed as two layers where each layer alternates between a serration, as shown in Figure 4.8, to protect the black spot. The design of diameter tuyer orifice, firstly determine air volumetric flow rate and define the number of tuyer orifices, then the diameter tuyer orifice is calculated by dividing of air volumetric flow rate by the number of tuyer orifices. The considered necessary parameters for the air tuyer design can be seen in Table 4.4.

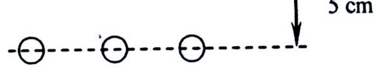


Figure 4.8 Alignment of air tuyers for char gasifier stage

Table 4.4 Design parameters for char gasifier stage

Parameter	Value
Tuyer orifice velocity ($m.s^{-1}$)	20
Number of tuyer orifices	6
Tuyer orifice diameter (m)	0.004

4.3.3 Gasifier

1) Pyrolysis Stage

The pyrolysis stage was designed as a bubbling fluidized bed. The TGA analysis shows that volatile matter start to leave from rice husk at 200°C and the devolatilization of rice husk completes at 600°C. The effect of temperature in terms of total product yield with a maximum is typically 500-520°C for most forms of woody biomass (Bridgwater *et al.*, 1999). The optimum temperature for pyrolysis seems to be within the range of 420-540°C (Ji-lu, 2007). To remove volatiles remaining in the char as much as possible, the pyrolysis temperature was chosen around 600°C.

Fluidization Velocity

The experiment of minimum fluidization velocity shows that the minimum fluidization velocity of the 250 μm of bed material was about 0.12 m/s. The value of terminal velocity from calculation was 1.89 m/s. Consequently, the range of fluidizing velocity is between 0.12 and 1.89 m/s. The optimum fluidizing velocity was approximately 3.3 U_{mf} for mean particle diameter 342 μm of sand as the mixing of rice with the bed was good with a high degree of penetration into the sand bed (Rozainee *et al.*, 2008). The fluidization velocity for this stage was designed for the range of fluidization velocity from 1 U_{mf} to 4 U_{mf}.

The Internal Diameter

This lab-scale two-stage gasification system was designed to facilitate varying essential parameters on the quality of the produced gas. In the experiment of effects of the ER at the first stage on the HHV of the producer gas (Wang *et al.*, 2007), the ER varied from 0.16 to 0.26. When the ER was higher than 0.21, the temperature of the combustion zone was higher than 1000 °C. To avoid ash melting problem, this stage temperature has to be controlled below 1100 °C. Therefore, the internal diameter was designed for the range of ER from 0.16 to 0.25 which is in the range from literature.

Table 4.5 The internal diameter range of ER between 0.16 and 0.25

		$1U_{mf}$	$2U_{mf}$	$3\ U_{mf}$	$4U_{mf}$
ER=0.16	Fix u	0.120 m/s	0.24 m/s	0.36 m/s	0.48 m/s
	A = V/v	0.016 m^2	0.008 m^2	0.005 m^2	0.004 m^2
	r	0.070 m	0.050 m	0.041 m	0.035 M
	D	**0.141 m**	**0.099 m**	**0.081 m**	**0.070 m**
ER=0.20	Fix u	0.120 m/s	0.24 m/s	0.36 m/s	0.48 m/s
	A = V/v	0.019 m^2	0.010 m^2	0.006 m^2	0.005 m^2
	r	0.079 m	0.056 m	0.045 m	0.039 M
	D	**0.157 m**	**0.111 m**	**0.091 m**	**0.079 M**
ER=0.25	Fix u	0.120 m/s	0.240 m/s	0.360 m/s	0.480 m/s
	A = V/v	0.024 m^2	0.012 m^2	0.008 m^2	0.006 m^2
	r	0.088 m	0.062 m	0.051 m	0.044 M
	D	**0.176 m**	**0.124 m**	**0.101 m**	**0.088 M**

To support variable parameters, the internal diameter of the reactor was designed of two parts which is still sustaining the condition of the bubbling fluidized bed of each ER. The internal diameter of the lower part was designed as 0.09 m. The internal diameter of the upper part was designed as 0.13 m as shown in Table 4.5. When these values were compared with the pilot plant, with the internal diameters of the lower part and upper part being 0.42 m and 0.60 m, respectively found that the ratio of lower and upper part are relevant which the lower and upper part ratios of the pilot plant are 0.7 and 0.69, respectively.

Reactor Height

The study of effect of pyrolysis reaction time on pyrolysis weight loss (Wannapeera, 2006) found the char yield at 700°C at 1 s after injection to be as low as 32% of rice husk. This implies that major weight loss could occur during the fast heating period. For longer reaction time, a small additional release of volatile continued by the decrease of char yields at a much slower rate. The lower yield of pyrolysis char product was observed at the longer holding time, which should be attributed to the progressive gasification and/or thermal cracking of the pyrolysis product (Tsai *et al.*, 2006). Therefore the design of reactor height, the reaction time was determined as 3 min which is the longer holding time for combustion. A reactor height was mainly calculated from volume of rice husk and sand. Calculation of the height is shown in the following steps:

- Calculate the volume of rice husks at 3 min of the reaction time. (0.0075 m^3)
- Define the height of silica sand part as 0.30 m
- Calculate the volume of lower part of reactor which consists of rice husks and sand
- The rest of rice husks volume is calculated for the upper part

Table 4.6 The height of the pyrolysis stage reactor

Lower part	
Internal diameter	0.09 m
Height	0.5 m
$V_{additional\ rice\ husk}$	0.001272 m^3
Upper part	
$V_{the\ rest\ of\ rice\ husk}$	0.006811 m^3
Internal diameter	0.13 m
Height	0.51 m
Height of reactor	**1.01** m

Table 4.6 shows the calculated height of the reactor, and the heights of the lower and the upper parts are 0.5 m and 0.51 m, respectively. To ease production, the height of the upper part was designed as 0.5 m. Therefore the overall height of the reactor is 1 m.

2) Char Gasifier Stage

The char gasifier stage was designed as a downdraft gasifier. The high-temperature gasification chamber above 800°C rapidly cracks the primary pyrolysis oils to olefins and aromatic compounds. These compounds continue to react in the absence of oxygen to make polynuclear aromatic compounds (PNAs), and eventually, soot, which is a final gas very low tar content. The high temperature also promotes reaction with char, which in turn rapidly quenches the gas to 800°C. Therefore, the time available for tar cracking in a bed of hot charcoal is very short. For this reason, a bed of hot char may not be very effective in tar cracking as was originally believed (Reed and Das 1988).

The parameters that have to be considered are internal diameter and height of reaction chamber.

The Internal Diameter

The diameter of the throat was calculated from the volume of gas produced per unit cross-sectional area at the throat per hour that the flow rate of producer gas was calculated from the percentage of produced gas composition of the pilot plant. The flow rate of producer gas was calculated by dividing the energy output (50kW) by heating value of producer gas in units of MJ per m^3 which was converted from heating value in units of Joule per mol. The heating value in units of Joule per mol was calculated by the multiplication of mol fraction of combustible gases, which was converted from the percentage of producer gas composition, with its heating value which heating value of CH_4, CO and H_2 is 802600, 283000 and 241800 J/mol, respectively (Perry, 1997). At 50 kW thermal of producer gas, the volumetric flow rate of produced gas is around 40 Nm^3/h. A maximum hearth load value for an Imbert-style gasifier, which the diameter is measured at the throat or at the air entry level, is about 0.9 $Nm^3/cm^2\cdot h$ (Reed and Das, 1988). For this study, hearth load was determined as 70% of the maximum value of hearth load which was 0.63 $Nm^3/cm^2\cdot h$.

The diameter of reactor was calculated from a function of the amount of the fuel mass flow rate to the specific gasification rate (SGR) of rice husk. Fixed carbon, ash and some unburned fuel were determined for fuel mass flow rate because volatile matter is released in the first stage around 63.7 wt% from proximate analysis. Since there is some unburned fuel in the first stage, fuel mass flow rate was assumed around 50 wt% of rice husk feed (9.7 kg/h) which take the time of accumulation of char at the height of 30 cm around 12 min. In the case of fuel mass flow rate is less than 50 wt%, it will take over. On the other hand, it takes less than if fuel mass flow rate is rather than 50 wt%. SGR is in the range of 110 to 210 $kg/m^2\cdot hr$ as revealed by the results of several test on rice husk gas stoves (Kythavone). Also, an optimal value of the specific gasification rate exists in the vicinity of 200 $kg/m^2\cdot hr$ that an experimental determination of the optimum specific gasification rate for static bed rice husk gas producers was conducted for reactor diameters of 16-30 cm. Cold-gas efficiency was observed to increase as specific gasification rate increased from 100 to about 200 $kg/m^2\cdot h$, and then begin to decline as gasification rate was increased further (Tiangco *et al.*, 1996). SGR value was selected for this study was 200 $kg/m^2\cdot h$.

From the information above, the calculated diameter of throat was 0.089 m rounded-off to 0.10 m. The calculated diameter of reactor was 0.24 m. When these values were compared with the pilot plant, which the internal diameter of throat and reactor is 0.92 m and 1.94 m, respectively found that the ratio of the internal diameter of throat and reactor are relevant which the internal diameter ratio of throat and reactor are 0.42 and 0.47, respectively.

Reactor Height

The important parameters to design the height of the char gasifier are the char bed height and the height between beneath the throat and char bed for reducing the temperature from approximately 1000 - 1100°C to less than 800°C to prevent ash melting. With the simulations to investigate the influence of char bed height of rubber wood with an average particle diameter of 3.3 cm on molar percentage of gasification products, calorific value of product gas (Sharma A.Kr., 2008), cold gasification efficiency, exit gas temperature and endothermic heat absorption, the char decreases exponentially with increasing char bed length. The minimum char bed height at which all char get consumed completely is approximately 0.25 m. According to the relation of temperature and distance from grate (GØbel B., 2007), temperature start to decrease as low as 750 °C at distance of 0.42 m. Therefore height beneath the throat should be higher than 0.67 m.

From the information above, the height of the char gasifier was determined to be 1.4 m, and was divided into two parts: 0.70 m for above and beneath the throat. Table 4.7 shows the internal diameter of char gasifier stage. Figure 4.9 shows a drawing of the designed reactor.

Table 4.7 The internal diameter of the char gasifier stage reactor

$\text{Diameter}_{\text{throat}}$	0.075 m
$\text{Diameter}_{\text{reactor}}$	0.240 m

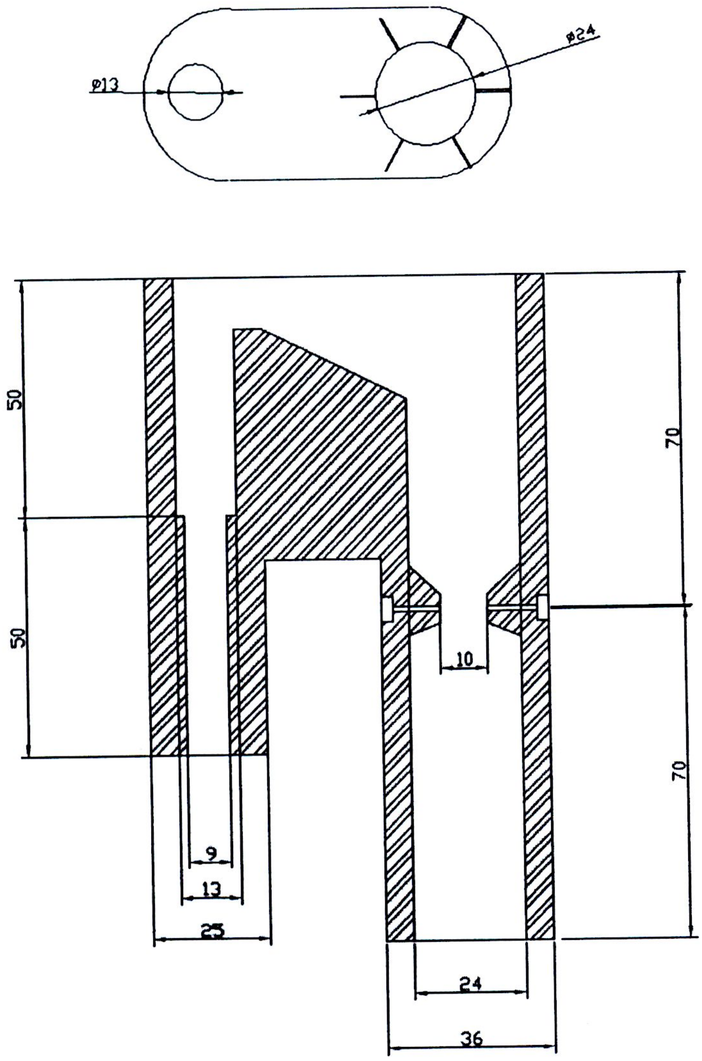


Figure 4.9 The drawing of the designed reactor (unit:cm)

4.3.4 Ash removal device

The design of screw conveyor for ash removal is determined similarly to the design of the screw conveyor for rice husk feeding, but the mass flow rate of ash is smaller than for rice husk feeding. Due to the small size of designed rice husk feeding screw, it is difficult to construct screw conveyor for ash removal is smaller, therefore solve this problem by decreasing the speed of the screw.

The calculation of the screw conveyor for ash removal, the speed of the screw was determined to calculate the same size of the screw conveyor for rice husk feeding. The speed of the screw is defined as 60 rpm. Rice husk ash from proximate analysis was 25.8

wt% but there is some unburned fuel in the process. Therefore, the rice husk ash was defined as 30 wt% for calculation of the screw conveyor with mass flow rate of rice husk ash being around 0.16 kg/min. The diameter of the shaft, screw speed and pitch are set equal to the screw conveyor for biomass feeding system. From Equation (3.1), diameter of screw is 0.04 m. When diameter of screw include the clearance between the screw and the radius of the inside of the screw trough, the diameter of the inside of the screw trough is 0.067 m which is close to the size of screw for feeding system.

4.3.5 Gas Cooling Device

The first shell and tube heat heat exchanger was designed as a one-shell and two-tube pass because two passes provides slightly better heat transfer than a single pass and the limitation of the lab-scale should not be too high. Front and rear-end head are designed for use with fixed tubesheet and is removable for mechanical cleaning. Shell is designed as rectangular shape with the dimensions: L 0.34 m x W 0.28 m x H 1.6 m. The outside and inside diameter of tube are 1" and 0.93", respectively from tube dimensional data table. Selection of the tube size is considered from easier size to clean. The total of used tube is 56 tubes, which is divided into 28 tubes for each pass. Tube layout is designed as square pitch for ease of manufacture. Baffle spacing is designed as 0.1360 m with the baffle cut 25%. Pitch size is 0.0318 m.

The size estimation and rating of the design of shell and tube heat exchanger results in tube length of the heat exchanger between the tube sheets being 1.6 m with the pressure drop of both shell and tube within the acceptable range. The machine drawing of the shell and tube heat exchanger is shown in Figures 4.10 and 4.11.

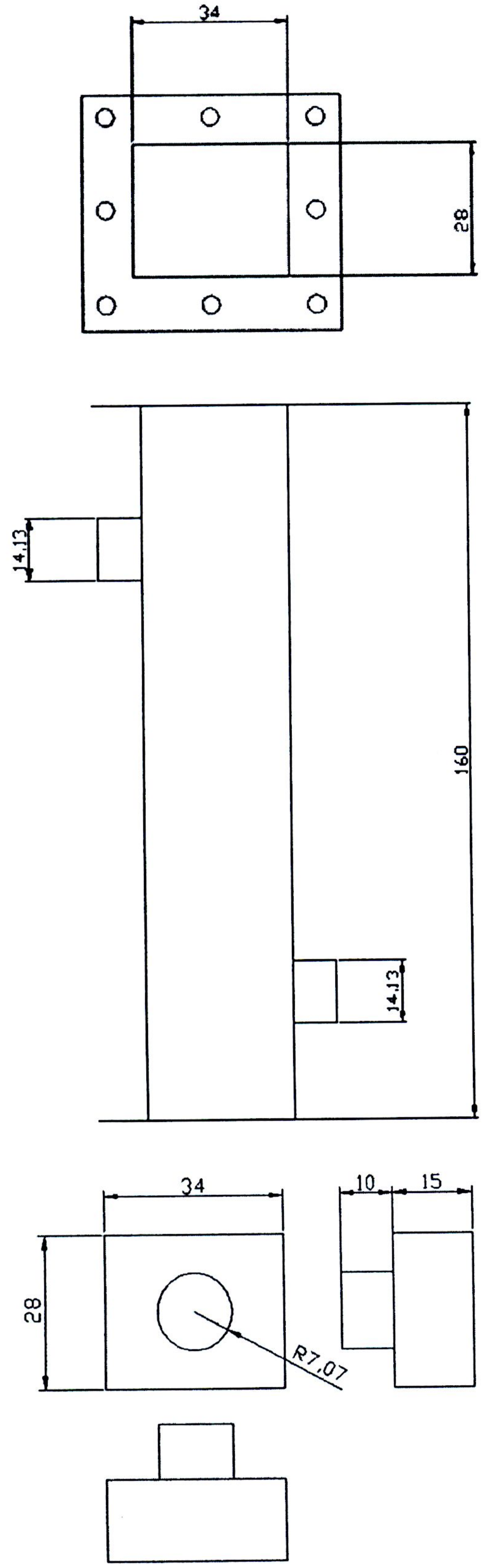


Figure 4.10 The drawing of the designed shell and tube heat exchanger (unit:cm)

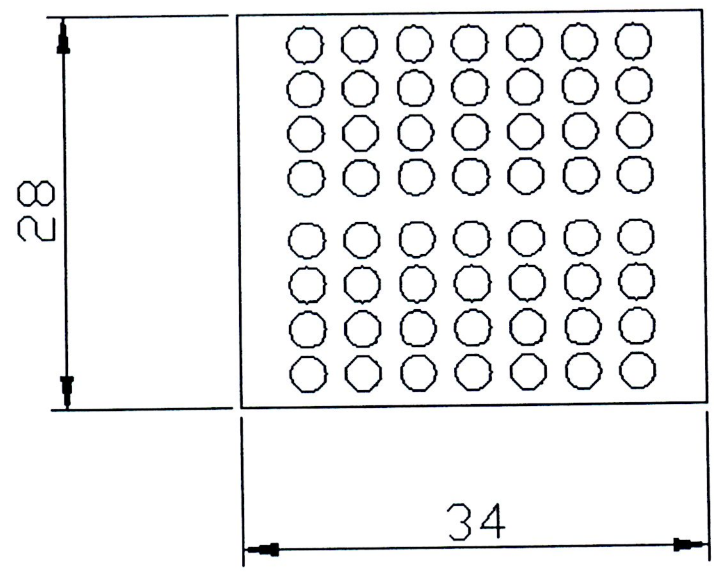
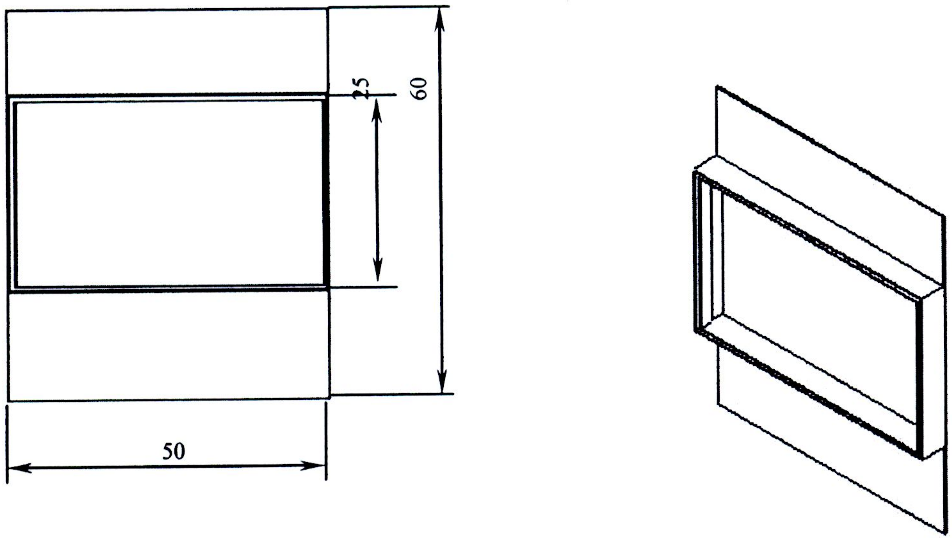
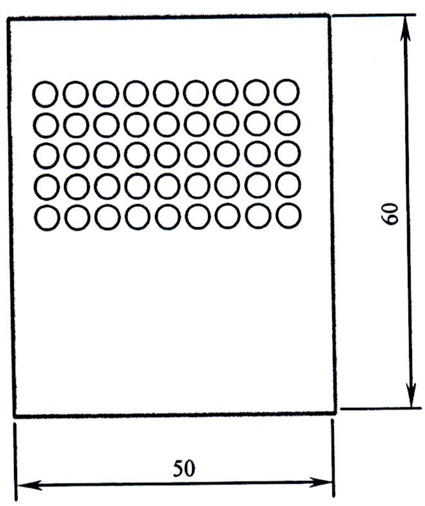


Figure 4.11 The drawing of the designed tube sheet (unit:cm)

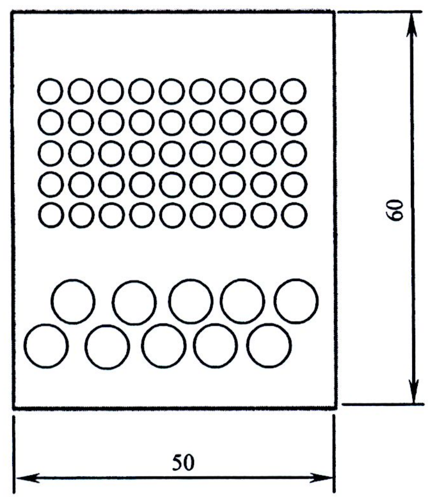
The second shell and tube heat exchanger was designed as a one-shell and two-tube pass because this heat exchanger was designed for flowing out of the condensed tar, which was designed following Figure 3.10. This heat exchanger was designed as rectangular shape with the dimensions: L 0.6 m x W 0.6 m x H 1.6 m. Which the upper part is heat transfer from hot producer gas to cold producer gas was designed as rectangular shape with the dimensions: L 0.6 m x W 0.5 m x H 0.6 m. The lower part height of heat transfer from air/water to hot producer gas was designed as 1 m that the upper tube sheet was designed smaller than lower tube sheet. The outside and inside diameter of small tube are 1.5" and 1.334", respectively from tube dimensional data table. The outside and inside diameter of big tube are 2.875" and 2.469", respectively from tube dimensional data table. The total of used small tube is 45 tubes with 0.0476 m of pitch size and 0.0095 m of clearances. The total of used big tube is 10 tubes with 0.09128 m of pitch size and 0.01825 m of clearances. Tube layout is designed as square pitch for ease of manufacture. The machine drawing of condensed tar shell and tube heat exchanger is shown in Figure 4.12.



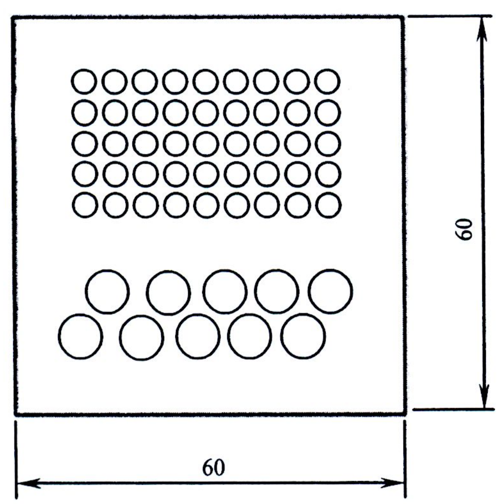
A. The cover of producer gas input



B. Tube sheet for producer gas input



C. The upper tube sheet of heat transfer from air/water to hot producer gas



D. The lower tube sheet of heat transfer from air/water to hot producer gas

Figure 4.12 The drawing of the designed condensed tar shell and tube heat exchanger (unit:cm)

4.3.6 Gas Cleaning Device

Cyclone

A 50 kW thermal requires 40 Nm^3/h of producer gas, which corresponds to a gas energy output of 180 MJ/h. Due to the cyclone being next to the char gasifier, the cyclone inlet temperature is close to the char gasifier temperature. The cyclone inlet temperature is determined as 600°C that the volume of gas at the cyclone inlet temperature of 600°C is 127.86 m^3/h. Then the calculated body diameter is 0.120 m. The cyclone pressure drop is 1.216 kPa and particle cut size is around 21 μm. The designed dimension of design cyclone as shown in Figure 4.13 are:

D_c	0.120 m
$B_c = D_c/4$	0.030 m
$D_e = D_c/2$	0.060 m
$H_c = D_c/2$	0.060 m
$L_c = 2D_c$	0.240 m
$S_c = D_c/8$	0.015 m
$Z_c = 2D_c$	0.240 m
$J_c = D_c/4$	0.030 m

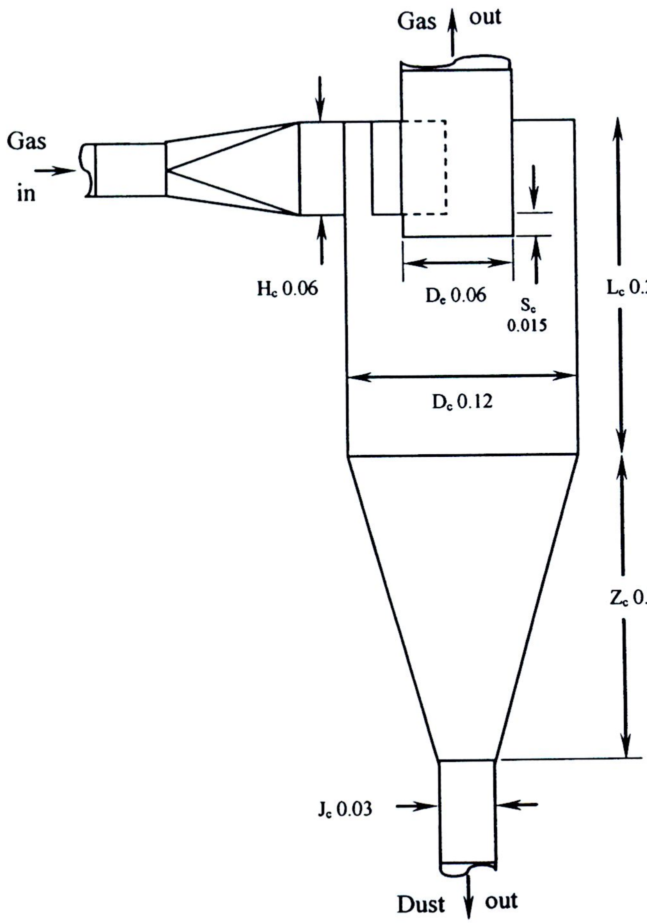


Figure 4.13 The drawing of the designed cyclone (unit:m)

Baghouse filter

Cotton was selected for filter material because of its availability and low cost. The inlet temperature of baghouse filter is determined around 70°C which is lower than the maximum temperature of cotton (82°C). After determine the inlet temperature, the number of bags required in the baghouse is calculated shown in Table 4.8. In the calculation of the number of bags required, bag diameter and height is first defined as 15 and 50 cm, respectively. These value is determined from the baghouse filter at the pilot plant. Because the number of bags is not an integer, so the number of bags is determined to be 4 bags. The design of the baghouse filter is shown in Figure 4.14.

Table 4.8 The calculation of bag required in the baghouse

Volumetric air flow rate (Nm^3/h)	40
Volumetric air flow rate@200°C (m^3/h)	50.25
The total cloth area (cm^2)	6979.22
The amount of fabric required per bag (cm^2/bag)	2356.19
The number of bags required in the baghouse	2.96

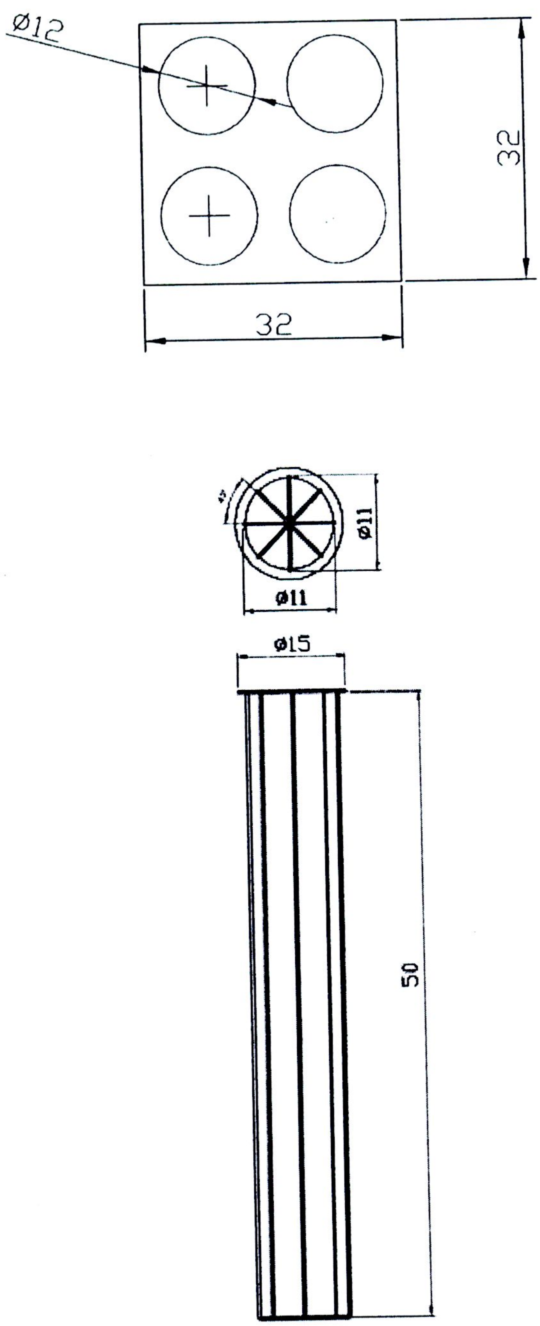


Figure 4.14 The design of the baghouse filter (unit:cm)