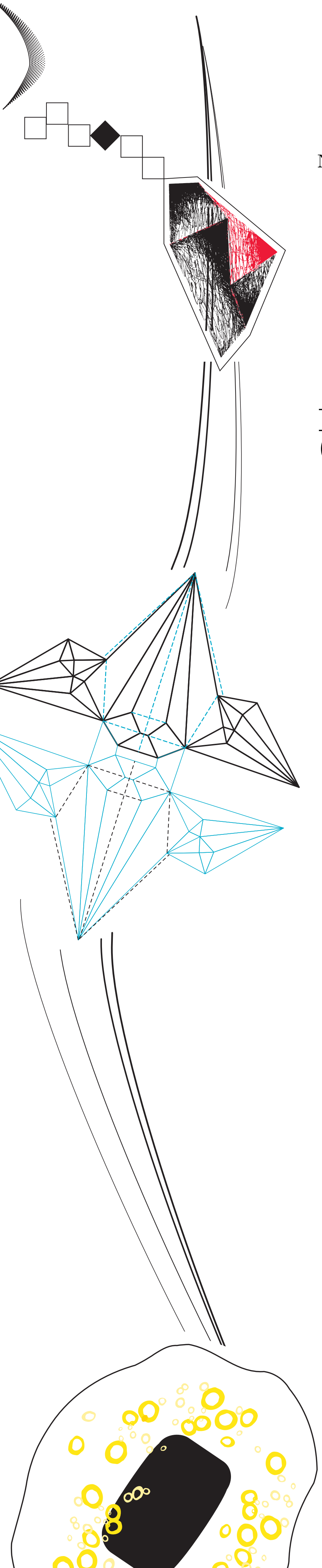

November 26, 2018

INTERNSHIP REPORT

BIOMASS TECHNOLOGY GROUP (BTG-BTL)

Chandrakiran Yada-	s1952412
Course name:	Internship
Course Number:	191199154
Supervisor 1:	Ruud Meulenbroek
Supervisor 2:	Elwin Gansekoele
Course Supervisor:	Ir.E.A Bramer

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INTRODUCTION

- In 2007, after delivering a pyrolysis plant to a customer in Malaysia and two test facilities, BTG established BTG BioLiquids B.V. (trade name BTG-BTL) in response to the large amount of requests they were getting from the market concerning their pyrolysis technology. BTG-BTL is dedicated to the worldwide commercial implementation of the fast pyrolysis technology developed by BTG.

  Empyro BV is jointly owned by BTG BioLiquids, Tree Power, the province of Overijssel and a private investor. BTG BioLiquids considers success of the Empyro project as the first leap towards full-scale fast pyrolysis technology commercialization.Empyro is the commercial pyrolysis production unit,which acts as the prototype for BTL's technology. Currently, a 5 t/hr plant has been designed, built and successfully taken into operation and a pipeline of projects and clients all over the world is lined up. Clients are investors, wood and agro-industry, waste and energy utilities, and oil companies.

  The empyro plant in Hengelo, takes 5 tons per hour of wood residue and converts this into pyrolysis oil, gas and char. In the empyro biomass plant, the flue gases are filtered using a bag-filter to maintain emissions under the limits. In order to select the best technology available for filter,the technical specifications are made. But before taking this step, it is imperative to know which flue-gases are there at the exit of the char combustor and the different scenarios that can effect the flue gas cleaning system. These typical scenarios are:

  -Heating up with burners using natural gas, propane and Light fuel oil.
  -Burners operation with biomass.
  -Normal Operating conditions.

  The goal of this assignment is to study the options available for the flue-gas cleaning such as using cyclones, Electrostatic precipitators, or a combination, to name a few. The cost analysis for the system is also considered and an economic comparison is made.Based on the technology and economical analysis, it can be decided which flue gas cleaning system is most suitable in connection to BTL's fast pyrolysis technology.

1.1 OVERVIEW OF THE EMPYRO PLANT

The pyrolysis process converts up to 70 wt % of the biomass feedstock into bio-oil and the remaining part into char and gas. Since 1993, BTG has played an active role in numerous projects on fast pyrolysis. BTG's unique and patented pyrolysis technology is characterised by an intense mixing without the need for an inert carrier gas. BTG-BTL's has taken BTG's patented RCR (Rotating Cone Reactor) fast pyrolysis technology and engineered it into a commercial industrial installation. The improved RCR design results in a remarkably small reactor, reduced system complexity and minimum down stream equipment size compared to competing pyrolysis technologies.

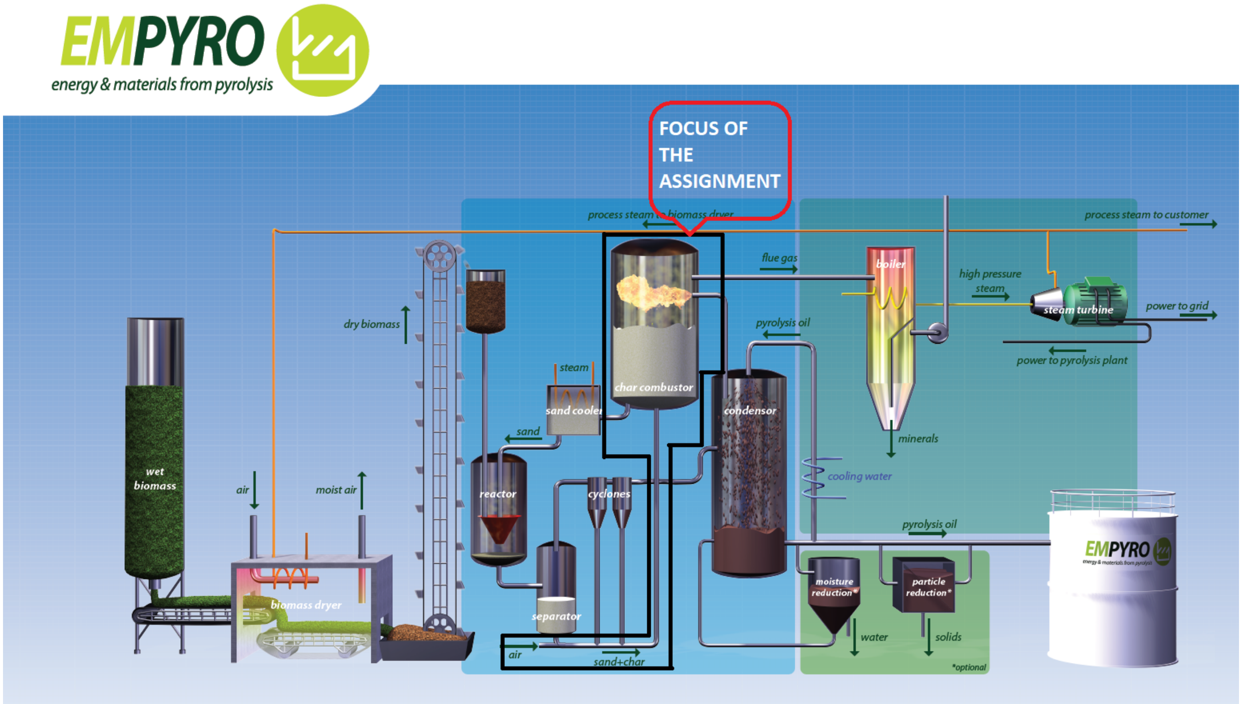


Figure 1.1: BTG-BTL Pyrolysis Plant

Dried biomass particles are fed into the pyrolysis reactor together with an excess flow of sand, which acts as a circulating heat carrier material. The biomass and sand are mixed within the pyrolysis reactor and converted into pyrolysis oil vapors, gas and char. The produced vapours and gasses pass through several cyclones before entering the condenser, in which the vapours are quenched by re-circulated oil. The sand and char are transported to a fluidized bed combustor, where air is added to combust the char. The non-condensable pyrolysis gasses enter the combustor from the condensor and are also combusted. The now reheated sand is then transported back to the reactor via a sand cooler to ensure a constant reactor sand feeding temperature. Excess heat from the sandcooler and from the hot combustor flue gasses is captured as high pressure steam.

The main advantages for BTG-BTL's technology in comparison to other pyrolysis technologies are:
-High biomass throughput per reactor volume resulting in compact reactor design.
-Absence of inert carrier gas resulting in minimum downstream equipment size.
-Maximum caloric value of pyrolysis gas.
-Very simple process: no gas recycle required.
-Straightforward to scale-up.
-Able to produce electricity or to produce a combination of electricity and steam for other applications.
-High flexibility for feedstocks: waste material, large particle size, etc.
-Low amounts of solids in the oil (down to 0.01 % wt).

The compact design of our modified rotating cone reactors make scaling-up straightforward to capacities larger than 5 t/h. Because of its simplicity of the rotating cone process, investment costs can be considerably lower in comparison to other pyrolysis technologies. CFB and fluid bed systems are more capital intensive, also because of the larger down-stream equipment, including ATEX and other safety issues.

Figure 1.2: BTG-BTL Pyrolysis Plant

BTL's standard design includes recovery of excess heat in the form of steam which can be used for industrial or local heating applications and electricity production. In general more electricity can be produced than required for the total plant. Enough steam is produced to dry biomass with a moisture content of up to 55 wt.% (wet basis) down to the required level. Depending on local conditions energy efficiencies of 85 – 90% can be achieved (based on biomass in and oil, heat, electricity out).

Because of the feed flexibility (related to combustor operation), BTL's technology can also handle biomasses with low ash melting temperatures such as palm derived EFB. BTL's technology can process particles with a thickness of up to 3 mm. Fluid bed technologies may use similar sized particles, while CFB technology must use smaller ones, as residence times are limited. Furthermore, due to a unique and patented cyclone designs used in BTL's plant and our high sand-to-biomass ratios, the produced pyrolysis oil is stable and has a very low solids content.

The empyro plant in Hengelo is a biomass based project, which also generates electricity using biomass. The plant also generates steam additionally. The char combustor generates flue gases and it also helps to produce steam by using saturated water passing through the pipes in the membrane walls. In order to simplify , the explanation here is focused on stages which effect the flue-gases and flue-gas filter.

1.2 Available Flue gas cleaning Technologies(literature Review)

Fluegas cleaning consists of removal of particulates,water soluble emissions,removal of NO_x and toxic materials. Various gas cleaning systems are classified in Fig. 4.Many gases cleaning system can be summarised as removal of particulates, removal of water soluble gases and pollutants.

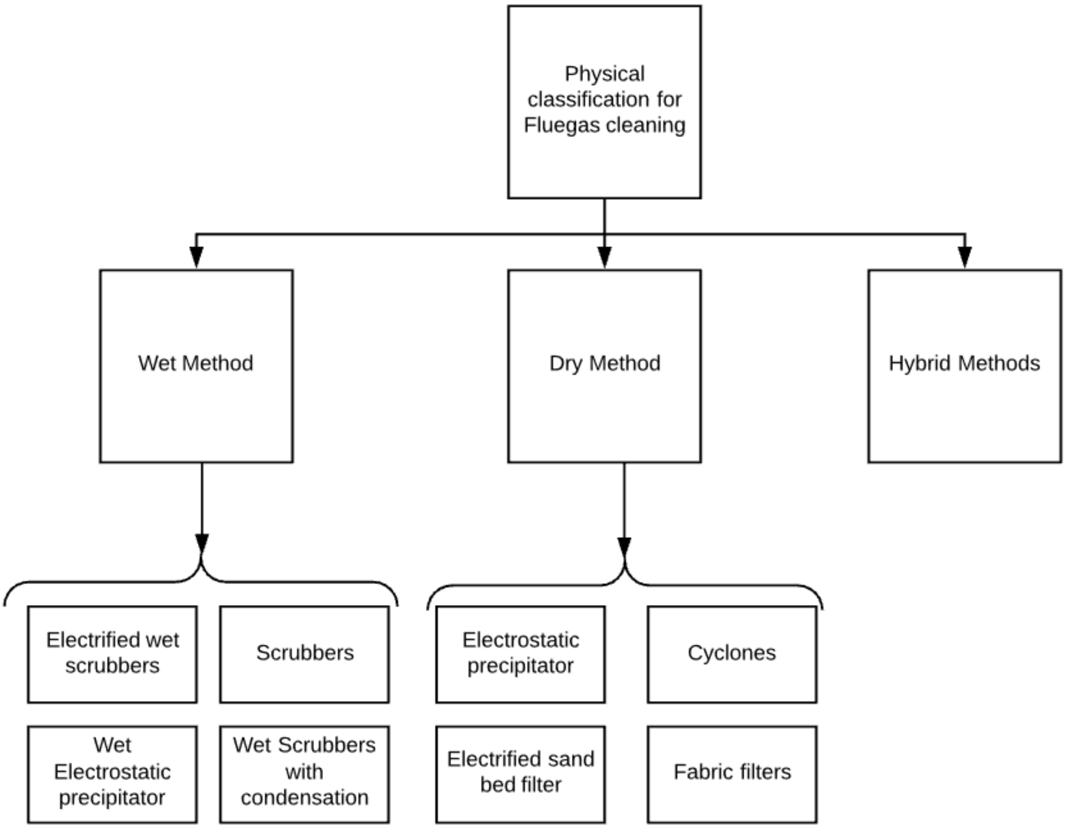


Figure 1.3: Cleaning systems

In this assignment the focus is on dry methods.Besides these methods,it is important to study about Lime injection process as well , as it will be seen that it is used to remove SO_x particulates from the flue gases when light fuel oil is used.

– **Electrostatic Precipitators:**
An ESP works because of electrostatic attraction .An ESP uses a high voltage electrostatic field to separate dust, fume or mist from a gas stream. The precipitator consists of vertical parallel plates (collecting plates/electrodes) forming gas passages 12 to 16 in. (30.5 to 40.6 cm) apart. Discharge electrodes are electrically isolated from the plates and suspended in rows between the gas passages. [1]
Every particle either has or can be given a charge - positive or negative. A high voltage system provides power to the discharge electrode to generate an electrical field. The particulate, entrained in the gas, is charged while passing through the electrical field. The particulate is then attracted to the grounded collector plate, and forms a dust layer on the plate.
Periodic rapping separates the accumulated dust layer from both the collector plates and discharge electrodes (in case of wet ESP by spraying it with a liquid). The dust layer released by the rapping collects in hoppers and is removed by material / ash handling system. At many places in this article, ash is used instead of particulate matter since major application of an ESP is for ash collection.

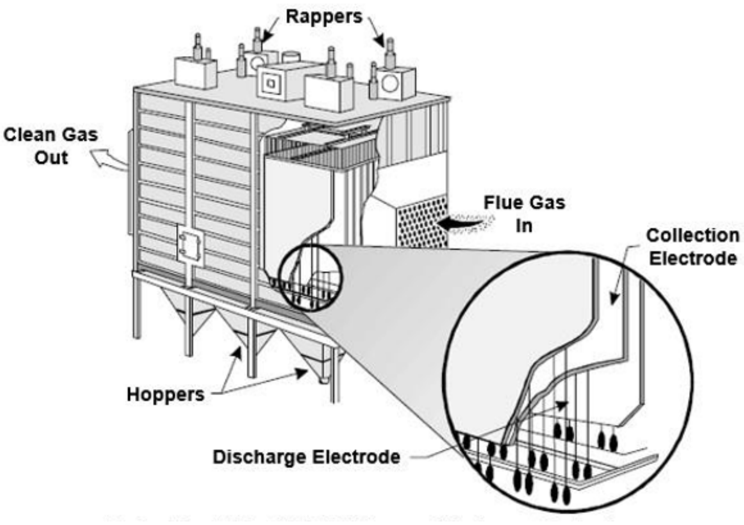


Figure 1.4: ESP

- **Fabric Filters:**
  Fabric filters are widely used as a control technology to remove particulates from the flue-gas streams of the industrial process or fuel- combustion units by passing the stream through a porous fabric.Fabric filters are also referred as bag houses which actually are the structures that houses the fabric filter system.These filters are highly efficient and the efficiency can reach upto 99 to 99.9 %.Because of their high efficiencies they are sometimes called as Best Available Control Technologies(BACT) for flue-gas cleaning.In this assignment,one advantage of using fabric which is based on the supplier information, can be used in combination with lime injection which is also known as Dry sorbent injection to filter SO_x from the emissions which is important in complexity wise, should be studied for sweden and also in finland case respectively since there are changes in the emission regulations in recent years.

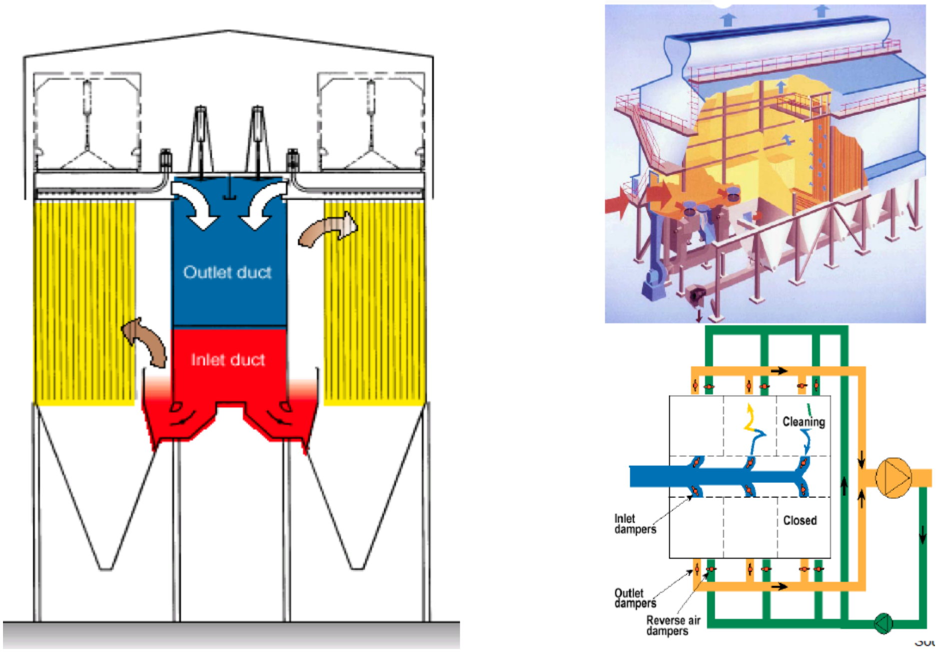


Figure 1.5: Fabric filter

- **Cyclones:**
  Cyclone separators have been used in countries like United states for about 100 yrs and are most widely used in all industrial gas-cleaning devices.This is because they are inexpensive to purchase,no

moving parts and can be constructed to withstand harsh operating conditions. Cyclones do not meet stringent particulate emission standards but they serve an important purpose that is low capital cost and their maintenance free operation. Cyclones are regarded as low efficiency devices but it varies greatly by the particle size and with cyclone design.Based on recent research , cyclones can achieve efficiencies greater than 98% for particles larger than 5μm.But the price to be paid as efficiencies increase so does the operating costs because of higher pressure drop.[2]

The operating principle of cyclones is very simple which is the use of centrifugal force created by the spinning gas stream to separate particles from gas.The crude gas enters tangentially on the top of the device. The cyclones shape and the tangential entry force the gas flow into a downward spiral.The centrifugal force and inertia cause the particles to move outward,collide with the outer wall and then slide downward to the bottom of the device.At the bottom of the cyclone,the gas reverses its downward spiral and moves upward in a smaller inner spiral.The 'clean' gas exits from the top through a vortex-finder tube and the particles exit from the bottom of the cyclone through a pipe sealed by a spring loaded flapper valve or a rotary valve.

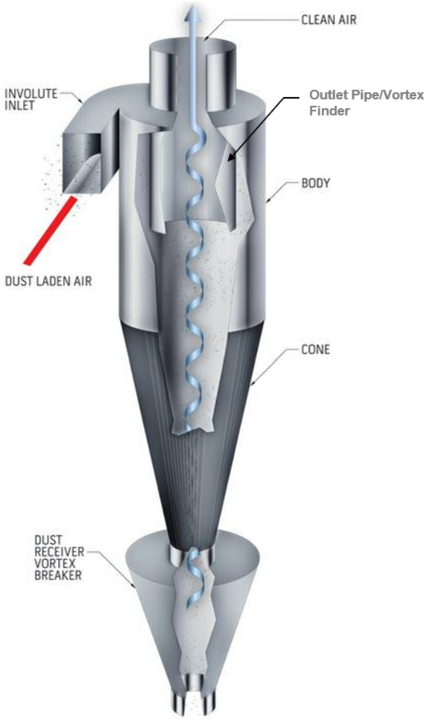


Figure 1.6: A typical cyclone

The centrifugal force is proportional to the square of the tangential velocity and inversely proportional to the radius of the curvature of the gas trajectory. Therefore , the efficiency of a cyclone increases as the diameter of the device is reduced.To achieve higher efficiencies dictates the use of smaller cyclones. However,the pressure drop through the cyclone increases rapidly as the tangential velocity increases. A way to maintain high efficiencies with a moderate pressure drop is to use large number of small cyclones placed in parallel.

Note 1: But in this assignment use of cyclone or cyclones in parallel before the fabric filter as a combination is not so influential according to the information stated from the supplier of BTG.The reason is that even though particulate matter is cleaned in the cyclone it does not reduce the size of the filter as the flow rate in Nm^3/hr is not varying a lot in different operational modes of the plant.Moreover, due to fine dust particles after the cyclone , the clogging of the filter bag inside the filter can occur easily which can reduce the efficiency of the filter and also leads to higher pressure drop than the rated pressure drop.(see the Appendix)

Note 2: Additionally using only the cyclone or multiple cyclones were also not beneficial because the lower efficiency in cleaning the crude gas does not satisfy the emission limits stated especially for sweden,finland and netherlands too.
As an overview, a comparison is made in the table below between different cleaning equipments and there possibilities in cleaning the flue gases. Please note that the data is from literature.In the end,the best suitable technology will be selected based on the design specifications of flue gases and emission of particulates in the clean gas.

Cleaning system	**Emission limits Dust in mg/m^3**	**Temperature allowed in Celsius**	**Energy consumption**	**Pressure drop in mbar**	**SO$_x$ emissions**
Cyclones	>500	>700	varies but its trivial	high	Not suitable
Bag filter	<<20	Important for bags because clogging is possible if less than dew point	high =1.5-3 KW/(m^3/s)	<=15-20	Not suitable without lime dosing
Bag filter with lime injection					>50% reduction
ESP	<= 20-50	<450 (corrosion if near to dewpoint)	low =0.2-1.5 KW/(m^3/s)	<=3	–

Table 1.1: Literature data

The bag filter plus lime injection is similar to the bag filter case but the inlet flue gases in the former can be pre-treated with lime before entering into the bags or can be sprayed onto the bags to remove specifically the sulphur oxides from the gases.In all the scenarios, the critical case is in the light fuel oil where we have the SO_x emissions. So the lime injection process could be necessary to be studied and also to be checked if it can meet the emission limits for the scandinavian countries.

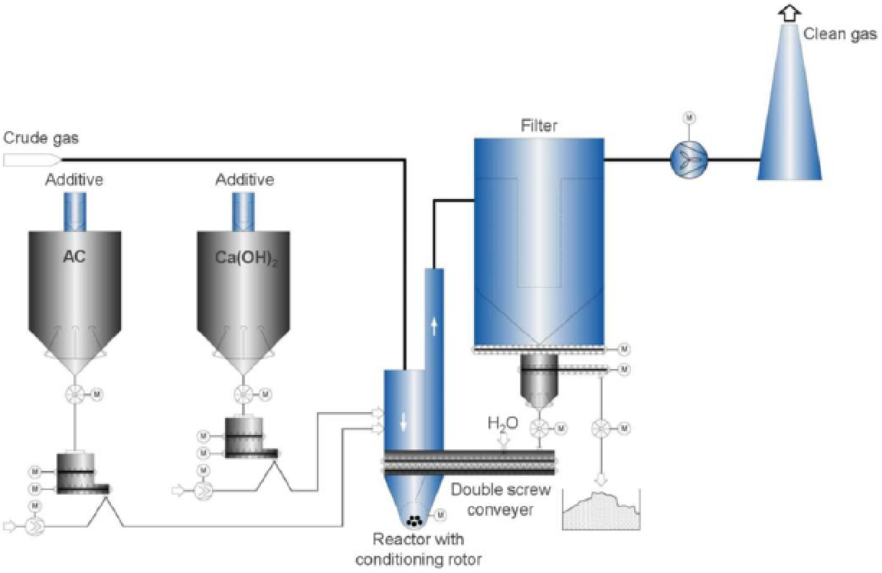


Figure 1.7: Pre-treatment of flue gases with lime

1.3 Design Specifications of Flue gases:

To select the best available technology,it is required to know first-hand about the flue gases that are going into the filter which defines the composition of the flue gases. But in order to have an insight of how these flue-gases are formed, a process flow diagram is made to understand more clearly about the char combustor as shown in the figure 1.1. The exit flue gas composition depends on the type of fuel used because,different fuel types have different reaction because of the difference in the chemical properties and also formula of the fuel. In this assignment , three different fuels are considered which is selected based on the client's location:

$$\begin{gathered} Propane - Finland \\ Methane - Netherlands \\ LightFuelOil - Sweden \end{gathered}$$

For simplicity,the natural gas is taken as methane for netherlands. It is safe to assume that the conclusion of this assignment will be based on the worst-case scenario for the filter selection , even though the design specifications are also made for methane and propane,the light fuel oil will be the critical case.

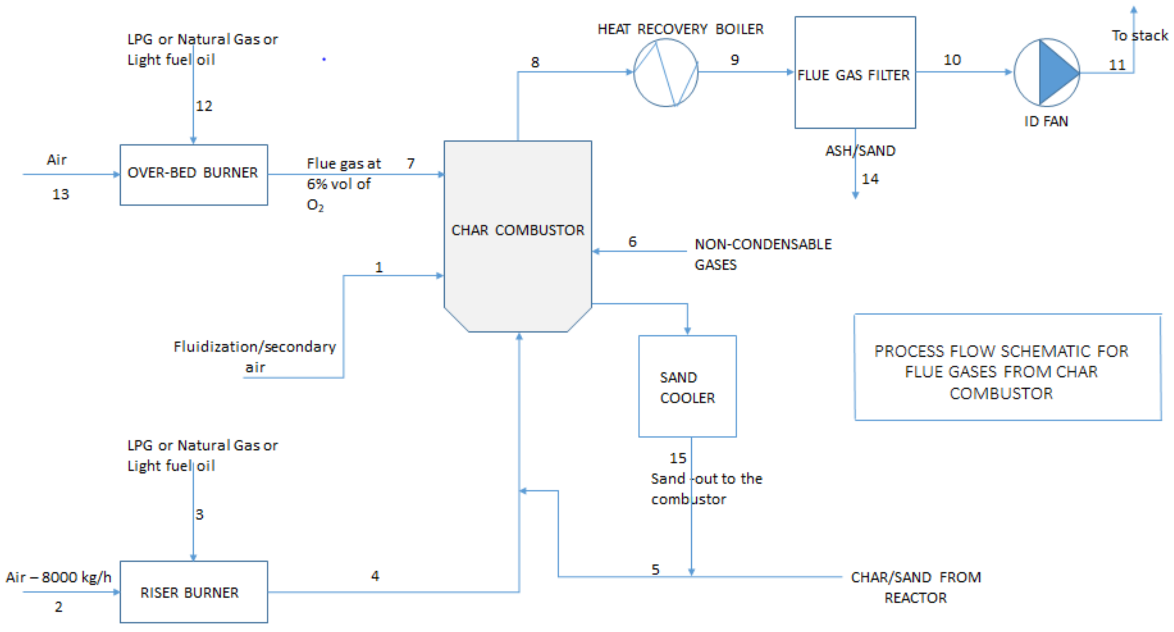


Figure 1.8: Flow Diagram

The dry biomass is combusted in a reactor with overflow of sand for heat circulation. The char which is the form of unburnt biomass is pumped back into the char combustor through the riser.In the char combustor,non-condensable gases and the char reacts with the surplus of oxygen. There are two burners which are used for the combustion process in the char combustor.

1. Riser Burner
2. Over-Bed Burner.

Consequently, the flue gases are formed as the products of these chemical reactions which are explained in the later chapters. The composition of the flue gas varies according to the type of fuel used in the burners and the mass flow rate of air varies based on the operational situation of the plant.Due to excessive collection of char during the start-up stage in the filter the possibility of burn-out has been solved by BTG using an intensive cleaning procedure for 10 minutes after the start-up.Hence, the assignment takes on three different scenarios namely:

- Heating up or start of the system
- Heating up with biomass and burners
- During the normal operation of the plant

The composition of the flue-gases at the exit of the char-combustor or the exit of the Heat Recovery Boiler is calculated using the mass balance of the char-combustor and also with the help of stoichiometric chemical reactions.Some of the parameters are taken as inputs from the current operation of the plant. These inputs are tabulated in each specific sections followed from here.

Besides this, the flue-gases are cooled down to a temperature of 240 °C before entering into the filter.An induced draft fan (ID fan) is used at the exit of the filter to maintain a pressure of 5mbar in the free board section of the char combustor and the Heat recovery boiler.The clean gases from the filter are then passed through the stack into the atmosphere.The ash or sand from the combustor is passed through a sand cooler with a heat transfer between saturated water from the steam drum and hot sand.The sand is then pumped back through the riser.At the bottom of the boiler we also have soot blowers which are used to blow out any soot collected on the furnace tubes of the boiler to maintain the efficiency of the boiler.In general, the whole process is explained in a process flow diagram which in figure 1.8.

This helps to understand the flue-gases formation in the empyro plant. But the effect on flue gases which is the current focus in the assignment depends on the type of fuel that is used in the burners. Therefore, these three fuels are considered in both the riser and over-bed burner individually and the composition of flue gases is calculated using mass balance for the char combustor with different scenarios.

1.4 Operational Modes and procedure of the Empyro Plant:

1. Heating Up of the system:
During the start-up of the entire plant, the fuel-burners are not working at their max-level and due to this there will be a time-lapse i.e 6 hrs for the fluidized bed combustor to reach its maximum temperature of $500°C$. The temperature of the sand cooler is also less in comparison to the normal working conditions which is explained in step 4. The volume flow of the air+fuel gas is 2500 N m3/hr during the start-up condition .The air flow+ sand mass flow is around 8000 kg/hr from the bottom of the fluidized bed.

2. 10 minutes after the start-up:
After 10 minutes of operation, the intensive cleaning procedure of the filter begins. This procedure is mandatory because of the un-burnt biomass in the form of char which is collected in the filter because of the low-temperatures in the combustor or reactor $<500°C$.

3. Temperature reached and Heat up with biomass and burners:
Once, the temperature which is higher than $500°C$ inside the combustor is reached,the mass flow of biomass begins to increase step-wise to 5000 kg/hr (this increment in mass flow is achieved slowly with time but not abruptly). Consequently, the air-fuel ratio in the burners is also scaled or increased eventually.

4. Normal Operation:
The normal operation of the entire pyrolysis plant is when the temperature inside the combustor is achieved $500°C$ and maintained seamlessly for 2.5 days without using the burners . At this condition, the amount of char produced or collected in the filter contains less moisture and there will be no further requirement of lime-injection (if used) before the filter. (Can be by-passed or shut-off).There is a burn-out procedure after 2.5 days which means that biomass is burnt completely and burners are switched back on and then the cycle repeats but at a higher temperature than the temperature as seen during the start-up.After 2.5 days the reactor and piping related to the condensor needs to be cleaned by cleaning

off the deposits. During the burn out the riser burner and the overbed burner are running on maximum capacity.

1.5 FLUE GAS CALCULATIONS:

a. **RISER BURNER**:

The riser burner is used for the purpose of heating the incoming air with propane and also char/sand is pumped to the char combustor in the flue-gas line.Each burner is analysed individually for the study of chemical reactions that occur and emissions that are out from the burner with input parameters tabulated below:

Table 1.2: Overview of Input parameters

Type of operation	**Heating up with burners**		**Heating up with burners and biomass**		**Normal operation**	
Input Parameters	**Air flow rate(in kg/h)**	**Temperature of flue gases(in ° C)**	**Air flow rate(in kg/h)**	**Temperature of flue gases(in ° C)**	**Air flow rate(in kg/h)**	**Temperature of flue gases(in ° C)**
Riser Bed Burner	8000	570	8000	570	6800	—

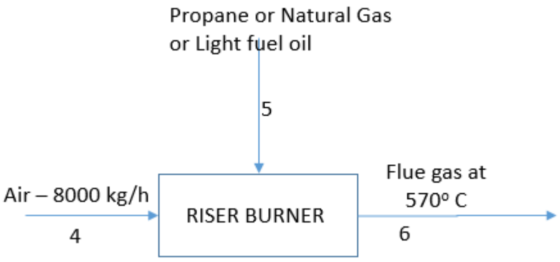


Figure 1.9: Riser burner

- Propane

Propane is used as the fuel for both the burners in the riser and overbed burner in finland.The air reacts with propane, which is at room temperature and due to chemical reaction , we have flue-gases as a consequence.The flue-gases contain the following components:

$$\begin{gathered} Carbondioxide - CO_2 \\ Nitrogen - N_2 \\ Water - H_2O \\ Oxygen - O_2 \end{gathered} \tag{1.1}$$

The stoichiometric reaction in the burner is given by:

$$C_3H_8 + 5O_2 \rightarrow 3CO_2 + 4H_2O \tag{1.2}$$

In the riser burner, it is known that the flue-gases exit at a temperature of 570 ° C after the reaction.The amount of air used is as given in table 1.1.The lower heating value of propane is considered to be 46.35 MJ/kg.In the riser burner,it is important to find the unknown- how much propane is required?.In order, to calculate the unknown variable i.e the mass flow rate of fuel, an initial guess is made and the energy balance is checked. This process is done iteratively until the energy balance is reached. The fundamental equation for the energy balance which is used in this process is given by the following equation.

$$LHV * m_{fuel} = h_{outlet} * m_{outlet} - h_{inlet} * m_{inlet} \tag{1.3}$$

In the above equation, the inlet conditions are considered at 25° C and outlet conditions at 570 ° C.But, the enthalpy is standardised by assuming the enthalpies at 25° to be zero.This is explained by an example for CO_2.

- Inlet Enthalpy of CO_2 at 25 $°C, \Delta H_{25°C}$: $H_{25°C}$-$H_{25°C}$= 0
- Outlet Enthalpy of CO_2 at 570 $°C, \Delta H_{570}$: $H_{570°C}$-$H_{25°C}$= 569.2 KJ/kg

All the enthalpies of the components are measured using a fluid property calculator[3] and calculated similarly.The values are tabulated below.

Table 1.3: Enthalpy Data

DATA TABLE			
Component	Enthalpy Outlet condition in kJ/kg at 570° C	Enthalpy Difference in kJ/kg	LHV in MJ/kg
CO_2	-8372	569.2	
H_2O	-12331	1092	
N_2	585.6	585.60	
O_2	540.61	540.62	
SO_2	401.46	402.06	
Propane	-871.89	1501.91	46.35
Methane	-2956.9	1710.4	50
Lightfuel oil			42

Using the data in 1.3, the values are substituted to check for the mass balance in 1.8. After many successive iterations, the mass flow rate is calculated to be 0.028515 kg/sec of propane.Consequently, the amount of flue-gases are also calculated for the amount of fuel used. This can be done by reaction balancing and the amount of flue gases in kmol/hr for the initial guess value.The results are tabulated below.

Component	**kg/sec**	**kmol/sec**	**kg/hr**	**kmol/hr**	**mol % or vol %**	**Nm^3/hr**
$\mathbf{CO_2}$	0.08	0.0019	307.95	6.99	2.48	154.17
$\mathbf{H_2O}$	0.04	0.0025	167.97	9.33	3.30	205.56
$\mathbf{N_2}$	1.70	0.0608	6136	219.14	77.69	4827.29
$\mathbf{O_2}$	0.41	0.0129	1490.71	46.58	16.51	1026.17

Table 1.4: Flue gases from riser burner-Propane as fuel

-Methane:
In this scenario, the fuel is methane in the burner.The lower heating value of methane is considered as 50 MJ/kg.As mentioned earlier,for simplicity it is assumed as the fuel used in the Netherlands. Methane reacts with incoming air and the flue gases are formed similar to the composition in 1.1. This is because of the following stoichiometric reaction.

$$CH_4 + 2O_2 \rightarrow 1CO_2 + 2H_2O \tag{1.4}$$

Since,the initial condition for the riser burner is known which is the amount of air that is used for the combustion. The only unknown quantity is the mass flow rate of fuel similar to the case in propane. Hence, the procedure is repeated in iterations and the final value for the methane fuel rate is found to be 0.02654 kg/sec. Inlet conditions of air is at temperature of 25 $°C$ with flow rate of 8000 kg/hr.The following table shows the final results of the product side after the energy is balanced on both sides in 1.8 by iterative process which was explained previously.

Component	**kg/sec**	**kmol/sec**	**kg/hr**	**kmol/hr**	**mol % or vol %**	**Nm^3/hr**
CO_2	0.073	0.00165	262.83	5.97	2.10	131.58
H_2O	0.0597	0.003319	215.04	11.94	4.21	263.16
N_2	1.7044	0.0608	6136	219.14	77.33	4827.29
O_2	0.411	0.0128	1481.69	46.30	16.34	1019.96

Table 1.5: Flue gases from riser burner-Methane as fuel

-Light fuel oil:

One of the important fuel types to be considered is the light fuel oil. The light fuel oil can be assumed as a dutch diesel or dutch fuel oil for this assignment. There are different types of light fuel oil with their own compositions.In this case,it is the following chemical composition.[4]

Composition of Light fuel Oil	**weight %**	**weight**	**Distribution ratio**
C content	85	7.083	1
H content	14	14	1.976
O content	4	0.25	0.0352
N content	0.03	0.0021	0.000303
S content	0.05	0.0015	0.000221

Table 1.6: Light Fuel Oil

As it can be observed that the light fuel oil contains sulphur which should be taken care of during cleaning via the bag filter.In order to do so , a special additive should be present which is called the lime or simply Lime injection process to condense the sulphur composition in the products side. The following chemical reaction occurs and as a consequence sulphur oxides are produced.

$$C_1H_{1.976}O_{0.0352}S_{0.000221}N_{0.000303} + 1.476O_2 \rightarrow 1CO_2 + 0.988H_2O + 0.000303NO_2 + 0.00022SO_2 \tag{1.5}$$

Therefore, sulphur dioxides are formed which should be controlled, as sweden has regulations on emission limits of SO_x which is less than 40 mg/Nm^3 at 6% volume of O_2.The NO_x emissions are quite negligible which when calculated was found to be around 0.0003 kg/sec. After some iterations the flow rate of light fuel oil was 0.0314 kg/sec.

Component	**kg/sec**	**kmol/sec**	**kg/hr**	**kmol/hr**	**mol % or vol %**	**Nm^3/hr**
CO_2	0.09	0.00215	341.70	7.76	2.76	171.06
H_2O	0.03	0.00213	138.14	7.67	2.72	169.05
N_2	1.70	0.0608	6136	219.14	77.88	4827.32
O_2	0.41	0.0129	1497.02	46.78	16.62	1030.51
SO_2	$3.05*10^{-5}$	$4.76*10^{-7}$	0.1096	0.0017	0	0.03

Table 1.7: Flue gases from riser burner-Light fuel oil

It can be observed that the SO_x emissions are in Kg/sec or kg/hr or Nm^3/hr.But, in general the SO_x emissions are measured in mg/Nm^3 at 6% volume of oxygen.Therefore, SO_x at 6% vol of O_2 is found to be as follows:

– Total flue gases flow rate at NTP conditions : 6198 Nm^3/hr

– SO_2 emissions in kg/hr : 0.10

- Percentage of O_2 for the existing SO_2 emissions : 16.62
- SO_2 at 6% vol of O_2 i s :
$$= \frac{0.10}{6198} * \frac{(21-6)}{(21-16.62)} * 10^6 = 60.67 mg/Nm^3$$

b. **OVERBED BURNER:**

Another burner that is used for the combustion is the overbed burner.This is located above the furnace bed and heats up the surrounding air and generates flue gases from the fuel that is used. Not a lot of data is known about the input parameters of overbed burner but the boundary condition is that the flow rate must be defined for oxygen at 6% volume basis.

Table 1.8: Overview of Input parameters

Type of operation	**Heating up with burners**		**Heating up with burners and biomass**		**Normal operation**	
Input Parameters	**Air flow rate(in kg/h)**	**Temperature of flue gases(in ° C)**	**Air flow rate(in kg/h)**	**Temperature of flue gases(in ° C)**	**Air flow rate(in kg/h)**	**Temperature of flue gases(in ° C)**
Over Bed Burner	—	—	—	—	—	—

- Propane

In addition to the riser burner, the overbed burner is also present in the combustor which is above the furnace bed. This burner heats up the surrounding fluegases inside the combustor too. But the criteria for the burner is that the amount of oxygen that is remaining should be at 6% volume.Besides that , it is known that the amount of heat generated is 3.5 MW. Therefore, using lower heating value of propane , it can be calculated that the mass flow rate of fuel to be 0.0755 kg/sec.

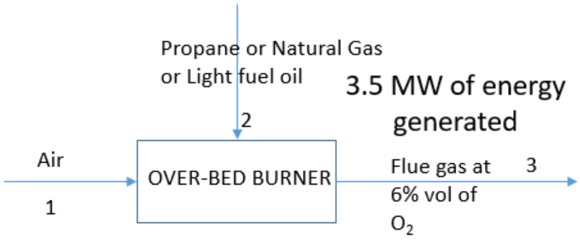


Figure 1.10: Over Bed Burner

The unknown quantity in this case ,is the amount of air that is needed to generate 3.5 MW.By iterative calculation, with a guess value of amount of air, the iterations are continued till the volume of oxygen reaches 6% at the exit of overbed burner.The chemical reaction is similar to the one in 1.8.After convergence,the flow rate of air is found to be 6082.42 kg/hr. The final values are tabulated below.

Component	**kg/sec**	**kmol/sec**	**kg/hr**	**kmol/hr**	**mol % or vol %**	**Nm^3/hr**
CO_2	0.22	0.00514	815.53	18.53	8.30	408.28
H_2O	0.12	0.0068	444.83	24.71	11.06	544.38
N_2	1.29	0.462	4665.21	166.61	74.62	3670.20
O_2	0.11	0.003	428.67	13.39	6	295.09

Table 1.9: Flue gases from overbed burner-for propane

- Methane:

Similarly,the flue gas calculations for methane is made and the results are tabulated below. Methane is used as the fuel in both the burners in this case and the convergence is done when the target value for the oxygen reaches 6%. The amount of air that is required for combustion of methane to produce 3.5MW of heat is found to be 6238.42 kg/hr when the oxygen at the products side has 6% volume.

Component	kg/sec	kmol/sec	kg/hr	kmol/hr	mol % or vol %	Nm^3/hr
CO_2	0.19	0.004	693	15.75	6.78	346.94
H_2O	0.15	0.008	567	31.5	13.57	693.88
N_2	1.32	0.047	4784.87	170.88	73.63	3764.34
O_2	0.12	0.003	445.55	13.92	5.99	306.709

Table 1.10: Flue gases from overbed burner-for methane

-Light Fuel Oil:
In this scenario, the fuel oil is used for the combustion of air. The lower heating value of LFO is valued at 42 MJ/kg. Since, for the riser burner the amount of flow rate of fuel is known, the unknown quantity that is the amount is air required is calculated similar to the previous cases.The total flow rate of flue gases from the burner is 4794.831 kg/hr. After the solution is converged at 6% vol of oxygen the amount of air is 5973.30 kg/hr.The results for the products composition is tabulated below:

Component	kg/sec	kmol/sec	kg/hr	kmol/hr	mol % or vol %	Nm^3/hr
CO_2	0.25	0.005	906.79	20.60	9.46	453.97
H_2O	0.102	0.006	366.59	20.36	9.35	448.63
N_2	1.27	0.045	4581.61	163.62	75.17	3604.43
O_2	0.11	0.003	417.91	13.05	6	287.68
SO_2	8.08*10$^-$5	1.26*10$^-$6	0.29	0.004	0.002	0.100

Table 1.11: Flue gases from overbed burner-for Light fuel oil

1.6 Char Calculations:

Char, which is formed during the fast pyrolysis of biomass in the reactor. The char from the reactor is pumped into the char combustor for combustion at high temperatures using the burners.Based on the process conditions in the reactor of the Hengelo biomass plant the char has the following composition under heating up conditions with biomass and normal operation of the plant respectively. There is not much char production during heating up process because the biomass is not yet used in the process.Consequently,the amount of flue-gases also vary.

Composition of Light fuel Oil	weight %	weight	Distribution ratio
C content	78.4	6.533	1
H content	5	5	0.7653
O content	16.4	1.025	0.1568
N content	0.16	0.0114	0.0017
Amount of ash	30 kg/h		

Table 1.12: Char composition during Heating up with biomass

The chemical formula of char can then be considered as $C_1H_{0.7653}O_{0.1568}$.
- During Heating up with biomass:
During the heating up with biomass,the char flow rate is found to be 900 kg/hr.The char reacts with oxygen inside the char combustor and the flue gases are formed.The following chemical reaction occurs in the combustor:

$$C_1H_{0.7653}O_{0.1568} + 1.1096O_2 \rightarrow 1CO_2 + 0.3736H_2O \tag{1.6}$$

Since the flow rate of char is a known quantity and the chemical formula of char is also available ,then as a result the composition of carbondioxide and water can also be calculated in kmol/hr. The values are tabulated below.

Component	kg/sec	kmol/sec	kg/hr	kmol/hr	mol % or vol %
CO_2	0.720	0.016	2592.38	58.917	0.723
H_2O	0.112	0.0062	405.81	22.545	0.276

Table 1.13: Char products during Heating up with biomass

The total amount of oxygen that is consumed is then given by 2098.196 kg/hr,which is used for the complete reaction of char in equation 1.6.

- During Normal Operation:

During the normal operation of the plant,the char flow rate is found to be 1100 kg/hr. Similar to the reaction in equation 1.6 occurs and the composition of the products are tabulated. During this operation, the burners are switched off and only the char is used to continue the combustion process.

Component	kg/sec	kmol/sec	kg/hr	kmol/hr	mol % or vol %
CO_2	0.883	0.020	3180.59	72.286	0.728
H_2O	0.135	0.0075	486.07	27.004	0.271

Table 1.14: Char products during normal operation

1.7 Non-Condensable Gases:

The non-condensable gases are the gases that are generated as a by-product when converting the biomass into bio oil. The non-condensable gases are passed from the condenser to the char combustor where they undergo combustion using either the riser air or overbed burner air.To be more clear about non-condensable gases, the volatile vapours, from the pyrolysis reactor, pass through the char separator(s), are then fed to the condensing system, or volatile vapour recovery unit, and the condensable vapours are separated from non-condensable gases. The volatile vapours are rapidly quenched to avoid secondary reactions which convert condensable vapours to non-condensable or permanent gases.In a typical large scale pyrolysis units we have one or two solid-vapor cyclone separators,because of which gives more residence time for the particles to be in the hot zone and increases the production of non-condensables.Common condensers used in biomass pyrolysis system include spray columns,shell and tube heat exchangers.Spray columns and towers have a similar design but in shell and tube we have different models.[5]

In this assignment, it is important to study about the non-condensables in two cases which is during the heating up of the system with biomass and during the normal operation. During the former stages , the amount of gases produced is less in comparison to the normal operation because of the difference in the flow rates of biomass which is explained in the operational modes.It should be noted that there are no non-condensable gases in the start-up only scenario since the there is no-biomass added.In the empyro system,the following are the flow rates and gas composition, under these operations.

-Under Heat up with biomass:

For the non-condensables, it is important to know how much oxygen is used for combustion process which is in the combustor.In the end it is required to calculate how much flue gases are exiting from the combustor using mass balance which will be explained in the next sections.The input values under this scenario is as shown below:

Gas components	**wt%**	**kg/hr**	**kg/sec**	**kmol/sec**	**kmol/hr**
N_2content	14	42	0.0116	0.0004	1.5
CO	26.3	78.9	0.021	0.0007	2.81
CO_2	55	165	0.0458	0.001	3.75
CH_4	2.4	7.2	0.002	0.0001	0.45
$C_2 + (C_2H_6)$	2.3	6.9	0.0019	$6.38 * 10^-5$	0.23
Total amount of gases		300			

Table 1.15: Non-condensable gases composition from start-up with biomass

Further, it is required to notice the reactions that occur under this composition. Consequently, the oxygen in the combustor reacts with hydrocarbons and carbon-monoxide with the following chemical reactions:

$$CO + 0.5O_2 \rightarrow 1CO_2 \tag{1.7}$$

$$CH_4 + 2O_2 \rightarrow 1CO_2 + 2H_2O \tag{1.8}$$

$$C_2H_6 + 3.5O_2 \rightarrow 2CO_2 + 3H_2O \tag{1.9}$$

Since, the flow rate of each composition is known, using the above chemical reactions we can calculate the quantity of products generated. The oxygen consumed will be sum of all the amount of oxygen used for each reaction.These results are given below:

Gas components	**kg/hr**	**kg/sec**	**kmol/sec**	**kmol/hr**
CO_2	164.02	0.045	0.001	3.72
H_2O	28.62	0.007	0.0004	1.59
Total amount of oxygen consumed	99.64	0.027	0.0008	3.11

Table 1.16: Products composition in start-up scenario

-Under Normal Operation:

As stated earlier in the operational modes that in the normal operation the amount of biomass is increased gradually to 5000 kg/hr which means that the non-condensables also increase. The flow rate of non-condensables in this case is given at 876 kg/hr. The composition of these gases is given below:

Gas components	**wt%**	**kg/hr**	**kg/sec**	**kmol/sec**	**kmol/hr**
N_2content	4.4	38.54	0.01	0.0003	1.37
CO	29	254.04	0.07	0.0025	9.07
CO_2	60.8	532.60	0.14	0.003	12.10
CH_4	2.9	25.40	0.007	0.0004	1.58
$C_2 + (C_2H_6)$	2.9	25.40	0.007	0.0002	0.84
Total amount of gases		876			

Table 1.17: Non-condensable gases composition from Normal Operation

Similar to equations 1.7,1.8,1.9, we have the same reactions as a consequence of combustion but the amount of oxygen consumed is different than in the previous scenario due to change in values of each gas component.The results are tabulated below:

Gas components	kg/hr	kg/sec	kmol/sec	kmol/hr
CO_2	543.58	0.15	0.003	12.35
H_2O	102.88	0.02	0.001	5.71
Total amount of oxygen consumed	341.62	0.09	0.002	10.67

Table 1.18: Products composition under normal operation

1.8 Preparation of Specifications for Supplier:

In order to request for a cleaning system from the supplier we have to organize our flue gas specifications. But now, from earlier sections the compositions of flue gases from each entity that is connected to the char combustor is known.So, the next step would be to see the exit flue gas compositions which is done by simply the mass balance. It is to be noted that the mass balance should be done for all the scenarios from start-up to the normal operations because the local circumstances of each case are different.The amount of gases and their composition from each entity that is the burners, flue gases from char and the non-condensables are available. Additionally, there is also mass flow of fluidization air and secondary air into the char combustor. Fluidization air is the air passed through the bubbling fluidized bed of the char combustor. The secondary air is used to compensate for the gases that went unreacted. It can be considered as an additional air that can be provided for complete combustion of the gases. In summary, the mass balance can be done in three cases which is shown in the following paragraphs below:

-Under Start-Up conditions

As explained in the operational modes , the start-up of the system begins by switching on the burners until the temperature inside the combustor is reached to a minimum level where there is no addition of the biomass.In this case,the combustor has no input from char and non-condensables. This is shown in the simple block diagram below:

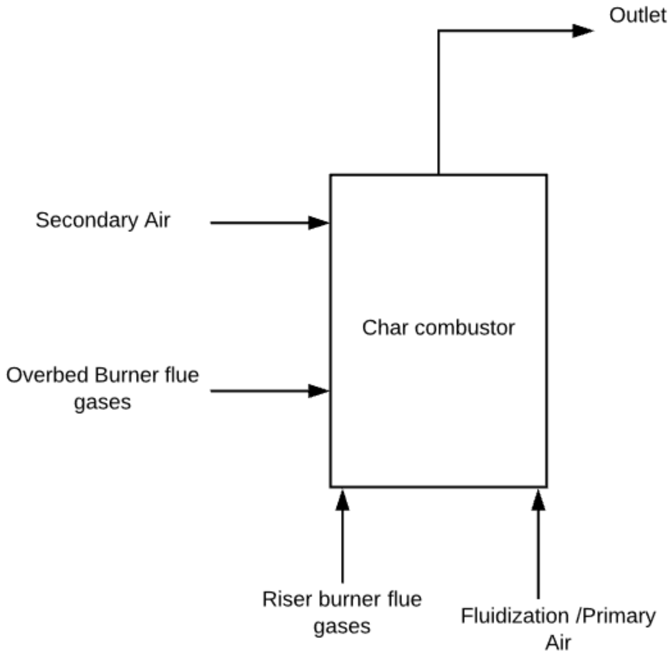


Figure 1.11: Start-up conditions

-Heating Up with Biomass:

In this case, the biomass is added into the system and temperature of the combustor rises to the peak level and the burners are switched off immediately.This condition remains with minor fluctuations inside the combustor until 2.5 days and then we have a burn-out condition. Therefore , we have flow of non-condensables from the condensor and the char flow into the system.

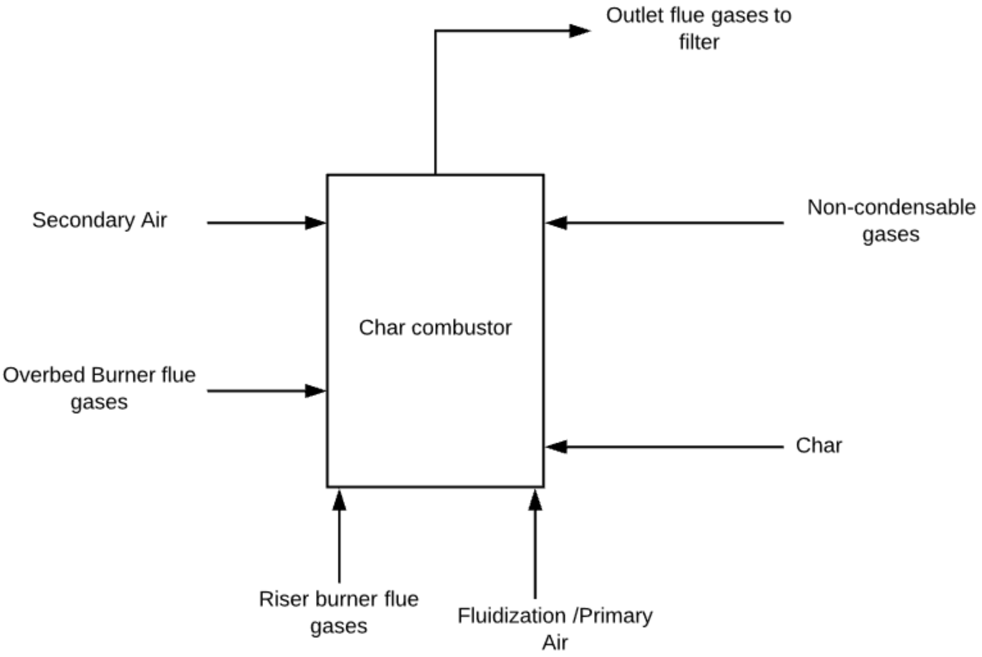


Figure 1.12: Heat up with Biomass and Burners

-Under Normal Operation:

It can be observed that the inlet flows into the combustor will be similar to second scenario with the exception that the flow rates of fluidization air, secondary or sand cooler air, the non-condensables are different.Figure **??** can be used again ,with just the change in values.

Therefore, after all these calculations the design specifications for each fuel type is summarised in the appendix.i.e for methane,propane and lightfuel oil.

1.9 Quotations from Supplier with Cost Analysis:

After the quantitative values for the flue-gases is known from the design ,the next step would be to ask for a quotation for a specific cleaning system from the supplier or manufacturer. The supplier might ask for a quantitative information about flue gases like the sulphur oxide emissions, NO_x emission data in which case the design specifications can be shown.Based upon the emission limits and regulations, the supplier suggests which cleaning system can be preferred in order to reach the emission limits in the clean gases.

In this assignment,since the light fuel oil is the critical case because it is the only case the sulphur emissions are present, we can focus the analysis on the cleaning system for light fuel oil. For CAPEX and OPEX costs , we have two different suppliers of BTG.Here named as supplier 1 and supplier 2.

From here on,the price data, electrical consumption and material required that will be represented is based on the information given by the supplier during the meetings and their preference and ofcourse the decision for the best available technology will be based on costs and suitability for passing the emission regulations.It should also be kept in mind that these regulations will be more constrained in the future. This reason will also be taken into account while selecting the best filter. The emission regulations for the scandinavian countries are as follows. [6]

Gas components	**Finland**	**Sweden**	**Netherlands**
Dust in mg/Nm^3	10	20	20
SO$_x$*in* mg/Nm^3	100	40	200
NO$_x$*in* mg/Nm^3	300	200	200

Table 1.19: Emission Limits at 6% Volume of O_2

Having known the emission limits , the next step would be to see the options available as explained in the literature review there can be possibility of technologies that can be combined,likewise:

Only Bag filter
Bag filter +Lime injection
Electrostatic precipitator
ESP plus cyclone
Cyclone plus Bag filter
Multi Cyclone

But based on the emission regulations,some of the combinations were not suitable since they cannot meet these limits. These combinations may not also work if there are SO_x emissions in the flue gases.Based on these reason some combinations can be ignored which were also stated by the suppliers of BTG. These combinations are:

Multi Cyclone
ESP plus cyclone

Also, it was explained previously that using a cyclone with a bag filter is not an advantage because it does not reduce the size of the filter and moreover , collection of fine particulates in the bags can occur over time which is also an issue since clogging takes place and reduces the efficiency of the filter and therefore increases the pressure drop in the filter which leads to increase in power consumption by the ID fan to maintain the set pressure. Hence, cyclone plus bag filter is also not a suitable option for the purpose. Therefore, the technologies that are suitable for further analyses are:

Only Bag filter
Bag filter +Lime injection
Electrostatic precipitator

Now, these technologies are compared by the data given by the suppliers and then calculate the costs incurred on the buyer which here is BTG by knowing initial investment costs and the operational costs of the system. A critical case will be explained here to get an overview of how the costs are made and this can be repeated similarly with few minor changes to other systems. The critical scenario is the Bag filter plus lime injection because it satisfies the reduction of sulphur emissions. Hence, it will be a good starting point.

1.9.1 Bag filter and Lime injection:

As the name suggests ,the buyer has to pay for both the bag filter and lime injection setup. This means as an initial investment ,the costs for the bag filter and lime dosing are important. The investment cost for the bag filter and lime are given by the supplier 1. But it also needs operational maintainance. Therefore , the OPEX costs. The operational costs for the bag filter are tabulated below based on certain assumptions for the empyro plant which are:

-Maintanence costs which includes replacement of bags
-Maintanence costs during unplanned stops which also incurs additional costs for stoppage of the plant.
- Number of Operational hours
-Pressure drop in the bag filter (given by the supplier)
-Electrical Consumption costs per kwh
-Additional Electrical costs due to pressure drop
-Costs for lime dosing

All of these costs are per annum basis.Since the flow rate varies based on the operation , so are the costs. Hence , for the three scenarios which is start up, heating up with burners and normal operation. Additionally, it should be observed that the flow rates mentioned in the appendix are for the NTP or Normal temperature and pressure conditions, but the temperature in the flue gas line is 240 $^{\circ}C$. Therefore, the flow rates will be much different and is reflected in the table of costs.The electrical costs is considered to be 0.1 euro/Kwh.

The flow rates at 240$^{\circ}C$ for lightfuel oil can be calculated as follows, for example with Bag filter plus lime injection during heating up with burners and biomass:
-Sum of flow rates in kmol/hr during this operational mode: 650.05 kmol/hr
- flow rate at 0° during heating up with burners and biomass is : 14319 Nm^3/hr

- Therefore,flow rate at 240 $^{\circ}C$:
= 650.05*8.314*513.15/103.1
= 26901 Am^3/hr

This is the actual flow rate in the flue gas line. Additionally, it should be observed that during the usage of lime only the additional electrical equipments for using the conveyor belts,motors etc are used, but not during without the lime. Hence the operational hours vary accordingly.The lime injection is used after 2.5 days for 8 hrs per week.It was also observed that it is required around 2.02 kg of lime per kg of sulphur emissions,according to supplier 1.There are also other additives with lime which are considered 3.1 kg/hr and 5 kg/hr for additive ii and remainder from additional injection respectively.

Number of operational hours of normal operation of the plant: 7500 hours per annum
Number of operational hours when using lime is : 416 hours per annum
Number of operational hours with burners and biomass is : 52 hrs per annum

It was also observed that electrical consumption mentioned in the quotations is for maximum operation. Therefore, equipments such as ID fan, other electrical devices are taken 80% of the rated value in KW, which is true for operational conditions.

The maintainance costs are of two types:
-During a planned stop of the plant the operational costs are 15000 euros.
-During an unplanned stop of the plant the operational costs are 45000 euros.
The additional costs during the unplanned stop is because of the costs that the company has to pay for the loss.In general, for the EMPYRO plant there are 4 start-up's in a year.

1.9.2 Table of Costs:

Costs for Bag filter plus Lime injection:

Costs	BAG FILTER PLUS LIME INJECTION		
	WITH LIME	**WITHOUT LIME INJECTION**	
OPERATIONAL MODE	**Start-up**	**Start-up with burners and biomass**	**Normal operation**
Flow rate of flue gases at 240 Celsius	25935	26901	28535
Number of operating hours per annum	416	52	7500
Pressure drop in mbar	12		
Additional ID fan power in KW required due to pressure drop	8.64	8.97	9.51
Electrical consumption of other equipments related to filter and dosing system or ESP in KW	6.21	0	0
Additional costs for Dosing per annum	1192	0	0
Electrical consumption costs for ID fan per annum	2762	345	49800
Electrical costs for other equipments	258.25	0	0
Additional Electrical costs in euros per annum due to pressure drop	359.63	46.63	7134
Maintainance costs per annum	«15000	15000	15000
Investment costs in euros based on supplier quotation	548140		
Total Investment (CAPEX)	548140		
OPEX costs	19314	15392	71934

Table 1.20: CAPEX and OPEX analysis for Bag filter plus lime injection

Note 1:
The maintainance costs mentioned in the table for bag filter does not include the costs for replacement of bags in the filter for every 3 years. This has to be considered additonally.

Note 2:
The maintainance costs mentioned in the table for bag filter during lime injection should be very less than when compared to other operational modes.Assuming the situation to be worst case scenario, the costs during the lime process was considered to be the same but in reality it is very less.

Costs for Bag filter:

Costs	**BAG FILTER**		
OPERATIONAL MODE	**Start-up**	**Start-up with burners and biomass**	**Normal operation**
Flow rate of flue gases at 240 Celsius	25935	26901	28535
Number of operating hours per annum	416	52	7500
Pressure drop in mbar	12		
Additional ID fan power in KW required due to pressure drop	8.64	8.97	9.51
Electrical consumption of other equipments related to filter and dosing system or ESP in KW	0		
Additional costs for Dosing per annum	0	0	0
Electrical consumption costs for ID fan per annum	2496	312	45000
Electrical costs for other equipments	0	0	0
Additional Electrical costs in euros per annum due to pressure drop	44.95	46.63	49.46
Maintainance costs per annum	15000	15000	15000
Investment costs in euros based on supplier quotation	470580		
Total Investment (CAPEX)	470580		
OPEX costs	17541	15359	60049

Table 1.21: CAPEX and OPEX analysis for Bag filter

Note:
The maintainance costs mentioned in the table for bag filter does not include the costs for replacement of bags in the filter for every 3 years. This has to be considered additonally.

Costs for ESP:

Costs	**ESP**		
OPERATIONAL MODE	**Start-up**	**Start-up with burners and biomass**	**Normal operation**
Flow rate of flue gases at 240 Celsius	25935	26901	28535
Number of operating hours per annum	416	52	7500
Pressure drop in mbar	3		
Additional ID fan power in KW required due to pressure drop	2.16	2.24	2.38
Electrical consumption of other equipments related to filter and dosing system or ESP in KW	12.32		
Additional costs for Dosing per annum	0	0	0
Electrical consumption costs for ID fan per annum	2496	312	45000
Electrical costs for other equipments	512	64	9240
Additional Electrical costs in euros per annum due to pressure drop	89.91	93.26	98.92
Maintainance costs per annum	0	0	0
Investment costs in euros based on supplier quotation	376000		
Total Investment (CAPEX)	376000		
OPEX costs	3098	470	54339

Table 1.22: OPEX and CAPEX analysis for ESP

Note:
There are no maintenance costs for ESP based on the supplier 2 information.

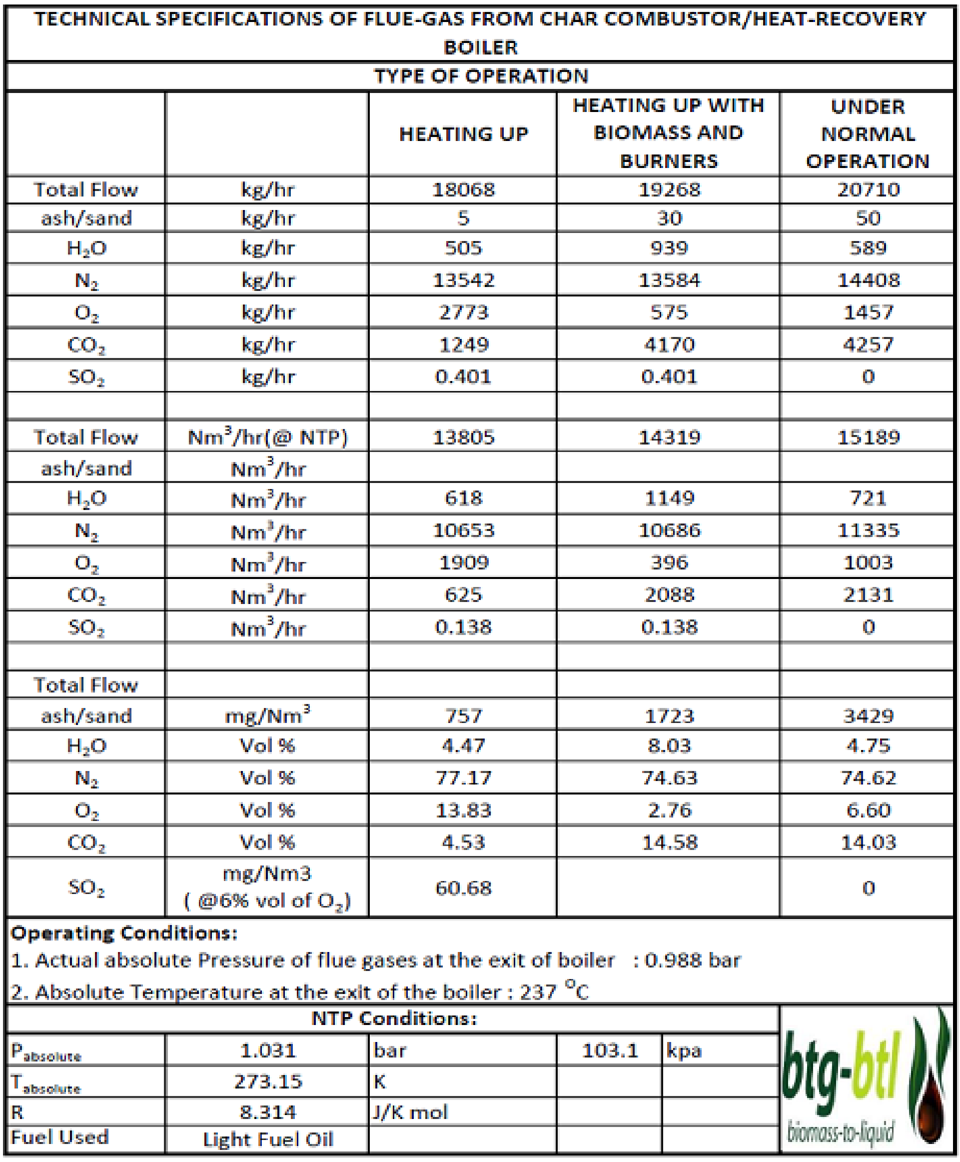
1.10 Conclusion:

-Observations based on Costs and availability:
The Best available technology based on the observations in the table of costs and emissions limit will be bag filter plus lime injection even though the cost for ESP is less,the critical case which is for the SO_x emissions cannot be solved for light fuel oil using ESP. The other advantage of choosing Bag filter plus lime injection is because it is very suitable for the EMPYRO plant as it already has a bag filter currently, so only lime injection equipment has to be added. Also, for the scandinavian countries, since the emission limits will be much constrained and will have very low values in future, bag filter plus lime injection can be very much flexible in meeting these limits.

2

Appendix

2.1 Design Specifications for LightFuel Oil



TECHNICAL SPECIFICATIONS OF FLUE-GAS FROM CHAR COMBUSTOR/HEAT-RECOVERY BOILER

TYPE OF OPERATION

		HEATING UP	HEATING UP WITH BIOMASS AND BURNERS	UNDER NORMAL OPERATION
Total Flow	kg/hr	18068	19268	20710
ash/sand	kg/hr	5	30	50
H_2O	kg/hr	505	939	589
N_2	kg/hr	13542	13584	14408
O_2	kg/hr	2773	575	1457
CO_2	kg/hr	1249	4170	4257
SO_2	kg/hr	0.401	0.401	0
Total Flow	Nm^3/hr(@ NTP)	13805	14319	15189
ash/sand	Nm^3/hr			
H_2O	Nm^3/hr	618	1149	721
N_2	Nm^3/hr	10653	10686	11335
O_2	Nm^3/hr	1909	396	1003
CO_2	Nm^3/hr	625	2088	2131
SO_2	Nm^3/hr	0.138	0.138	0
Total Flow				
ash/sand	mg/Nm^3	757	1723	3429
H_2O	Vol %	4.47	8.03	4.75
N_2	Vol %	77.17	74.63	74.62
O_2	Vol %	13.83	2.76	6.60
CO_2	Vol %	4.53	14.58	14.03
SO_2	mg/Nm3 (@6% vol of O_2)	60.68		0

Operating Conditions:

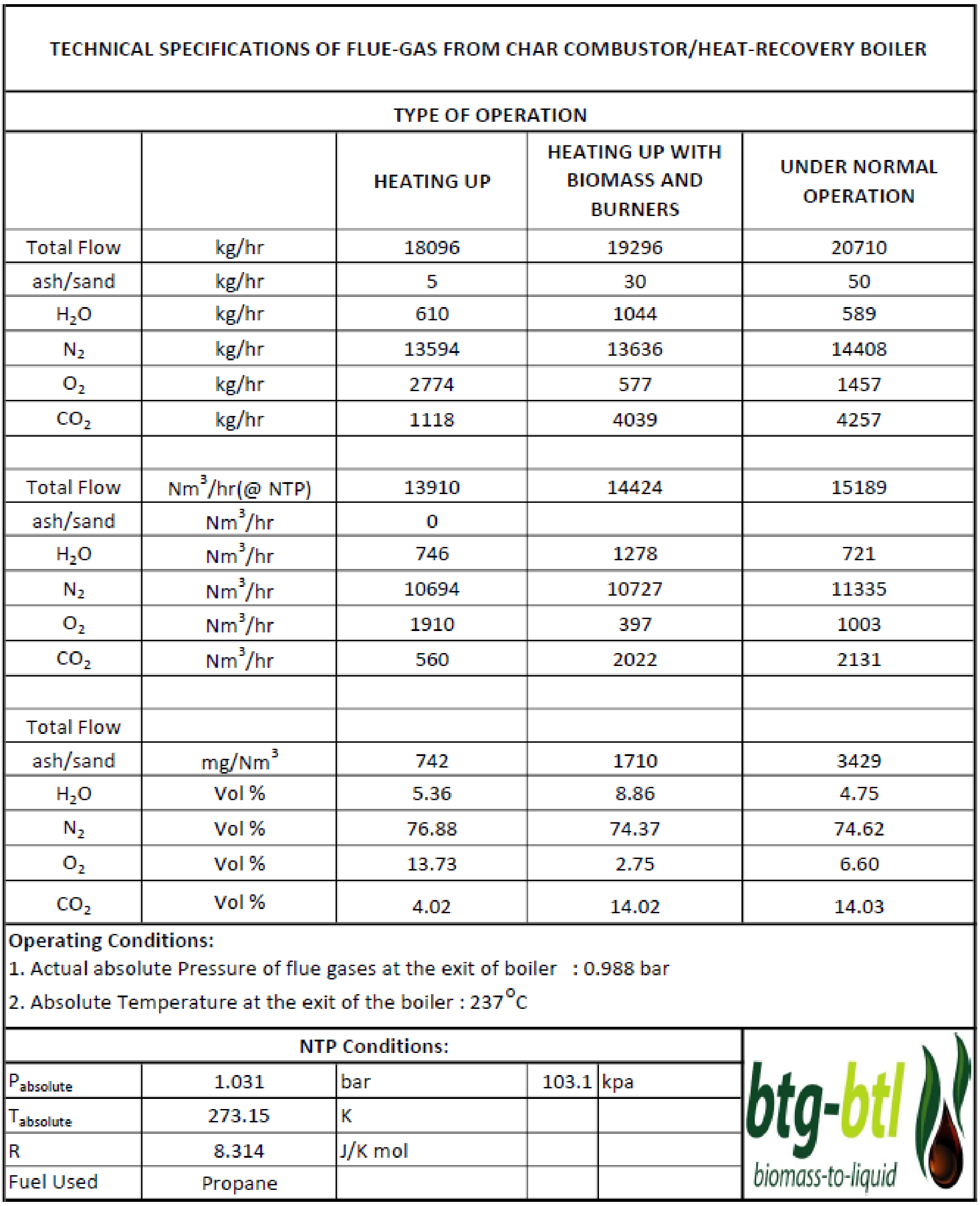
1. Actual absolute Pressure of flue gases at the exit of boiler : 0.988 bar
2. Absolute Temperature at the exit of the boiler : 237 °C

NTP Conditions:

$P_{absolute}$	1.031	bar	103.1	kpa
$T_{absolute}$	273.15	K		
R	8.314	J/K mol		
Fuel Used	Light Fuel Oil			

Figure 2.1: Specifications for LFO

2.2 Design Specifications for Propane



TECHNICAL SPECIFICATIONS OF FLUE-GAS FROM CHAR COMBUSTOR/HEAT-RECOVERY BOILER

		TYPE OF OPERATION		
		HEATING UP	HEATING UP WITH BIOMASS AND BURNERS	UNDER NORMAL OPERATION
Total Flow	kg/hr	18096	19296	20710
ash/sand	kg/hr	5	30	50
H_2O	kg/hr	610	1044	589
N_2	kg/hr	13594	13636	14408
O_2	kg/hr	2774	577	1457
CO_2	kg/hr	1118	4039	4257
Total Flow	Nm^3/hr(@ NTP)	13910	14424	15189
ash/sand	Nm^3/hr	0		
H_2O	Nm^3/hr	746	1278	721
N_2	Nm^3/hr	10694	10727	11335
O_2	Nm^3/hr	1910	397	1003
CO_2	Nm^3/hr	560	2022	2131
Total Flow				
ash/sand	mg/Nm^3	742	1710	3429
H_2O	Vol %	5.36	8.86	4.75
N_2	Vol %	76.88	74.37	74.62
O_2	Vol %	13.73	2.75	6.60
CO_2	Vol %	4.02	14.02	14.03

Operating Conditions:

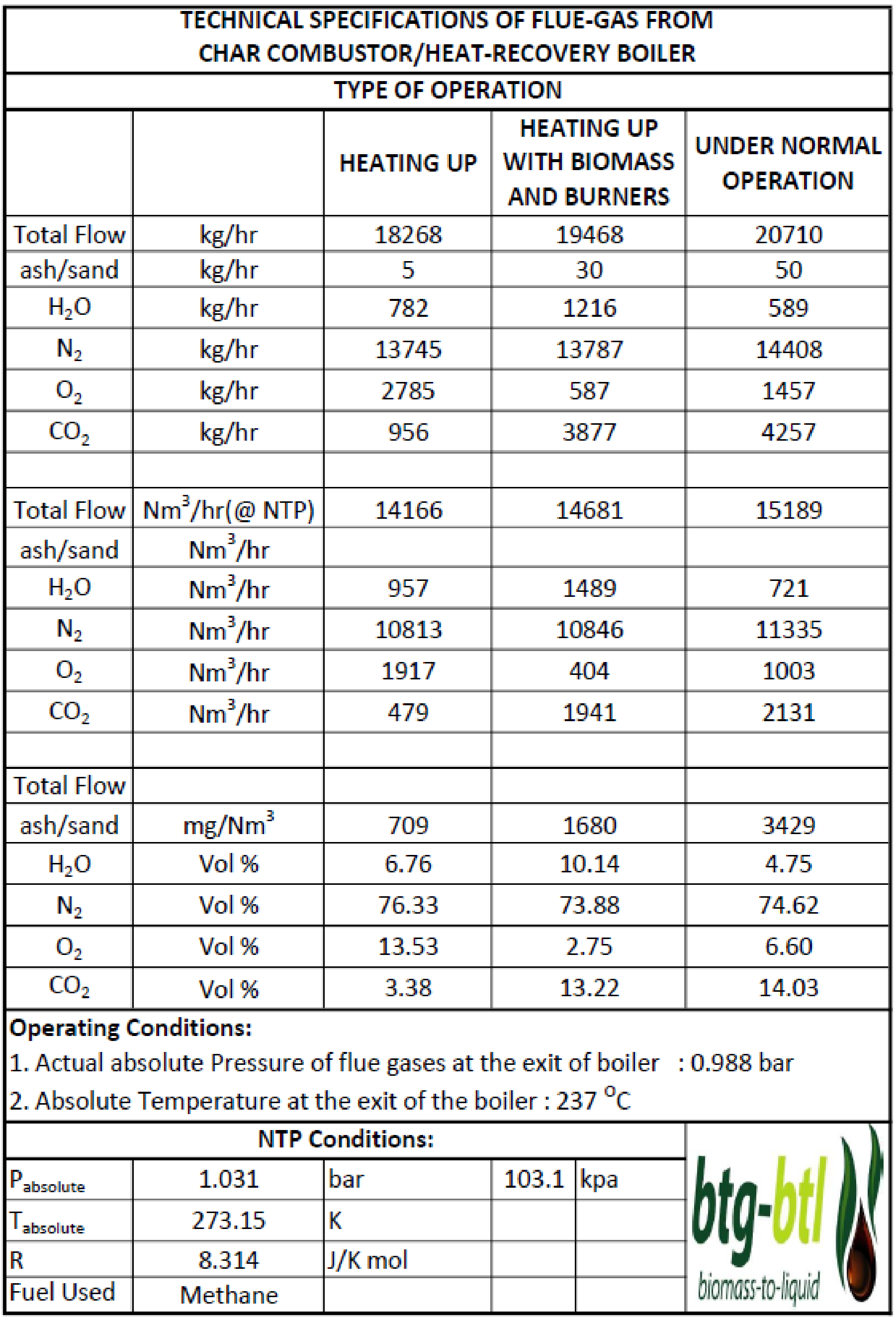
1. Actual absolute Pressure of flue gases at the exit of boiler : 0.988 bar
2. Absolute Temperature at the exit of the boiler : 237°C

NTP Conditions:				
$P_{absolute}$	1.031	bar	103.1	kpa
$T_{absolute}$	273.15	K		
R	8.314	J/K mol		
Fuel Used	Propane			

btg-btl
biomass-to-liquid

Figure 2.2: Specifications for Propane

2.3 Design Specifications for methane



TECHNICAL SPECIFICATIONS OF FLUE-GAS FROM
CHAR COMBUSTOR/HEAT-RECOVERY BOILER

TYPE OF OPERATION

		HEATING UP	HEATING UP WITH BIOMASS AND BURNERS	UNDER NORMAL OPERATION
Total Flow	kg/hr	18268	19468	20710
ash/sand	kg/hr	5	30	50
H_2O	kg/hr	782	1216	589
N_2	kg/hr	13745	13787	14408
O_2	kg/hr	2785	587	1457
CO_2	kg/hr	956	3877	4257
Total Flow	Nm^3/hr(@ NTP)	14166	14681	15189
ash/sand	Nm^3/hr			
H_2O	Nm^3/hr	957	1489	721
N_2	Nm^3/hr	10813	10846	11335
O_2	Nm^3/hr	1917	404	1003
CO_2	Nm^3/hr	479	1941	2131
Total Flow				
ash/sand	mg/Nm^3	709	1680	3429
H_2O	Vol %	6.76	10.14	4.75
N_2	Vol %	76.33	73.88	74.62
O_2	Vol %	13.53	2.75	6.60
CO_2	Vol %	3.38	13.22	14.03

Operating Conditions:

1. Actual absolute Pressure of flue gases at the exit of boiler : 0.988 bar
2. Absolute Temperature at the exit of the boiler : 237 °C

NTP Conditions:

$P_{absolute}$	1.031	bar	103.1	kpa
$T_{absolute}$	273.15	K		
R	8.314	J/K mol		
Fuel Used	Methane			

btg-btl
biomass-to-liquid

Figure 2.3: Specifications for Methane

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