DESIGN AND DEVELOP ALGAE BIOREACTOR FOR HYDROGEN PRODUCTION OF 1MW

By

Adeel Syed Mahmud

Moazzam Iqbal

Ramesha Hassan

School of Chemical and Materials Engineering

National University of Science and Technology

June, 2021

DESIGN AND DEVELOP ALGAE BIOREACTOR FOR HYDROGEN PRODUCTION OF 1MW.

By

Leader – 00000215508 Adeel Syed Mahmud
Member 1 – 00000219569 Moazzam Iqbal
Member 2 – 00000237574 Ramesha Hassan

School of Chemical and Materials Engineering

National University of Science and Technology

June, 2021

CERTIFICATE

This is to certify that work in this thesis has been completed by Mr. Adeel Syed Mahmud, Mr. Moazzam Iqbal and Ms. Ramesha Hassan under the supervision of Dr. Rafay Jafri at the school of Chemical and Materials Engineering (SCME), National University of Science and Technology, H-12, Islamabad, Pakistan.

Advisor:

Dr. Rafay Jafry
Department of Chemical Engineering
School of Chemical and Materials Engineering
National University of Sciences and Technology

Co-Advisor:

Dr. Umair Sikandar
Department of Chemical Engineering
School of Chemical and Materials Engineering
National University of Sciences and Technology

Submitted Through:

HOD-------------------------------------

Department of Chemical Engineering
School of Chemical and Materials Engineering
National University of Sciences and Technology

Principal/Dean ------------------------

School of Chemical and Materials Engineering
National University of Sciences and Technology

ABSTRACT

Biomass is being considered as a source for alternative and renewable fuels, such as hydrogen, and this is the reason extensive research is being carried out recently. Microalgae are one of the sources with huge potential. With high growth rates and a cultivation process known to be environmentally friendly, they are an ideal candidate for efficient production of hydrogen gas which is a clean energy provider.

The objective of this thesis is to design a bioreactor capable of producing enough hydrogen gas to provide a net power output greater than 1 MW. For this purpose, a new and ever improving technology of steam gasification in a bubbling fluidized bed reactor is utilized where algal biomass will be converted to syngas with high hydrogen content. This gas can be easily burned without extensive separation methods in a syngas gas turbine for the desired output of power. Extraction of hydrogen from microalgae provides us with a clean and renewable source of energy which can be scaled up according to industrial requirements.

DEDICATION

We would like to dedicate this project to our parents and our teachers who have struggled day and night in order to build us who we are today and inspired us to take on new challenges.

ACKNOWLEDGEMENT

We would like to thank Almighty Allah for his countless blessings on us. We are highly thankful to Dr. Rafay Jafri and Dr. Umair sikandar for helping us out throughout the project. Our Parents deserve a special praise for their support and keeping our spirits high.

-Authors

TABLE OF CONTENTS

LIST OF FIGURES

Figure 1 Process Flow Diagram 9

LIST OF TABLES

NOMENCLATURE

c_p	Heat Capacity
ΔT_m	Log Mean Temperature Difference
m	Meter
mm	Millimeter
cm	Centimeter
m^2	Meter square
m^3	Meter cube
cm^2	Centimeter square
cm^3	Centimeter cube
kJ	Kilo Joule
kg	Kilogram
h	Hour
c_v	Heat Capacity of Vapor
c_p	Heat Capacity of Liquid
c_s	heat capacity of Solid
mmol	Millimole
gmol	gram mole
L	Length
D	Diameter
H	Height
ð	Thickness
ε	Void Ratio

α	Selectivity
μ	Viscosity
ƛ	Latent Enthalpy of steam
θ	Stage Cut
y_1	Mole Fraction
y_2	Mole Fraction
T_{sb}	Temperature of Solids Leaving
T_{sa}	Temperature of Air Entering
T_{sv}	Temperature of Vaporization
N_t	Number of transfer Units
q_T	Energy
°C	Centigrade
°K	Kelvin
°F	Fahrenheit
P	Inlet Pressure
I.D	Internal Diameter
ρ	Density
MJ/kg	Mega Joule per kilogram
U	Overall Heat Transfer Coefficient
P_t	Tube Pitch
D_s	Shell Diameter
IB	Baffle Spacing
GS	Mean Velocity
F_t	Friction Factor

CO	Carbon Monoxide
CO_2	Carbon Dioxide
H_2	Hydrogen
N_2	NitrogenCH_4 Methane
Re	Reynolds Number
ΔP	Pressure Difference
PCE	Purchase Cost of Equipment
H_i	Internal Pipe Heat Transfer Coefficient
O.D	Outer diameter
A_s	Cross Sectional Area
Pr	Prandtl Number
Nu	Nusselt Number
d_e	Equivalent diameter
N_p	Number of Tube Passes
u_t	Tube Side Velocity
j_h	Heat Transfer Factor
s	Second
d_o	Tube Outside Diameter
BWG	Birmingham Wire Gauge
Q	Volumetric Flow Rate
F_{AC}	Flow rate of feed
h_{Ao}	Enthalpy of feed at entrance
h_A	Enthalpy of feed at exit
h_{So}	Enthalpy of air at entrance

h_S	Enthalpy of air at exit
F_S	Flow rate of air
m_{in}	Moisture in feed inlet
m_{out}	Moisture in feed outlet
$\Delta\dot{m}$	Change in moisture content
λ	Average moisture removal
A	Area of belt
b	width of belt
H	Height of conveyor duct
L	Length of conveyor
V	Velocity of belt
ρ	Density
X_{So}	Initial moisture fraction in feed
X_S	Final moisture fraction in feed
X_C	Ultimate analysis mass fraction of carbon
X_H	Ultimate analysis mass fraction of hydrogen
X_O	Ultimate analysis mass fraction of oxygen
X_N	Ultimate analysis mass fraction of nitrogen
$\emptyset_s$	Sphericity
U_{mf}	Minimum fluidization velocity
U_f	Fluidization velocity/superficial velocity
V_g	Volumetric flow rate
H_{fb}	Height of fluidized bed

TDH	Transport disengagement height
U_t	Terminal velocity
θ	Residence time
r	Rate of reaction
$\mathrm{D_c}$	Diameter of cyclone
d	Diameter of particle

CHAPTER 1.

INTRODUCTION

1.1 Background

The world is coping with challenges of high energy demands along with rapidly increasing fuel prices due to rising world population and industrialization which need to be dealt and for this, researchers are trying to come with economic and sustainable methods for producing energy. Fossil fuels lead to environmental pollution and thus, more attention is towards renewable energy sources as they are economic and environmentally friendly.

Microalgae contain high amounts of different carbohydrates, such as glycogen, starch, agar, and cellulose, etc. The major components of microalgae are lipids, proteins, and carbohydrates stored in the microalgae cell. Moreover, high lipid productivity is the main reason why microalgae can be used as an alternative feedstock. Operating conditions such as temperature and pressure should be monitored carefully.

1.2 Fuel potential of Microalgae:

Microalgae have a high growth rate that can convert solar energy into biomass.

Algae are being used to produce various kinds of biofuels, serving as biodiesel, bioethanol, **biogas,** and bio-hydrogen. Hydrogen is a clean fuel, environment friendly, renewable and a great replacement of fossil fuels and with high energy density than all other fuels.

1.3 History:

For the past few years, research about the potential of algal biofuels, to produce syngas and hydrogen via thermochemical routes was not given much importance. Now, with an improvement in our technologies, research has been focused on developing new methods for hydrogen production. One such method that has been gaining a lot of popularity is gasification. Using gasification for this purpose proves to be an effective method.

1.4 Sources of Algae:

Algae may be grown through the following ways.

- **Open ponds** – algae are placed in a pond with open air. Though it is simple, algae must compete with other living organisms.
- **Closed-loop** – just like open ponds, however they do not have access to open air and use CO_2.
- **Photo-bioreactors** – highly complicated and costly but provide greater amount of product and are closed.

Harvesting: After cultivation, the algae need to be harvested before any treatment. The most common method is using gravity settlement, or centrifuge followed by dryers such as conveyor dryers.

1.5 Biohydrogen:

Bio-hydrogen is obtained through steam reforming of methane, gasification of coal and biomass, and metabolic pathway of different microorganisms. This is a cleaner way to get energy from renewable sources. These micro-organisms are either photosynthetic or fermentative. Bio-hydrogen is formed by steam reforming of methane (biogas) produced by anaerobic digestion of organic waste. In the latter process, natural gas and steam react to produce hydrogen and carbon dioxide.

Bio-hydrogen can be made by bio photolysis, photo-fermentation, dark fermentation, combination of these processes or by water–gas shift reaction. Research is focused on increasing yields of biohydrogen from various strains of bacteria. Apart from these, microbial fuel cells can also be used.

1.6 Hydrogen from sustainable resources

According to research hydrogen can be made in the following ways:

- Biological Water Splitting
- Photo-electrochemical Water Splitting
- Conversion of Biomass and Wastes
- Solar Thermal Water Splitting
- Renewable Electrolysis

1.7 Uses of biomass:

Biomass acts as an alternative source of fuel. It can replace coal, oil, and gas as a clean and renewable alternative. Biomass can also be converted to hydrogen and other fuels which can then be used for electricity production.

1.8 Environmental aspects:

Biomass releases a lot fewer greenhouse gases than conventional fuels. Biomass contains no sulfur content, so the hazard of acid rain is eliminated and there will be no pollution from the sulfur oxides.

The atmospheric carbon dioxide is recycled when biomass is burned because the carbon dioxide emitted is equal to the amount absorbed during the biomass formation process. This aspect of biomass does not allow global warming to occur. There is no need to develop more landfills as the biomass goes through a recycle process.

1.9 Advantages of Algae:

- Algae is more efficient compared to any biofuel crops.
- It gives a high production rate.
- They are regenerative.
- It gives usable hydrocarbons for various products.

- Algae can grow in virtually all waters.
- Microalgae helps in reducing emissions as they need carbon dioxide to grow.

1.10 Biochemical composition of different microalgae

Sample	Biochemical composition (wt%)			HHV (MJ/kg)
	Lipids	Carbohydrates	Proteins	
Chlorella	15.78	16.10	46.80	22.18
Nannochloropsis	55.36	12.39	12.02	29.44
Pavlova	13.88	28.00	46.04	22.60
Scenedesmus	35.66	50.40	7.15	25.45

Table 1 Biochemical composition of different microalgae strains

1.11 Ultimate and proximate analysis

Microalgae name: Green Algae, Chlorella - Ultimate Analysis

Elements	Percentages
C	62.51
H	9.24
N	1.76
O	19.85

Table 2 Ultimate analysis of green algae

Proximate Analysis

Property	Percentages
Moisture Content	6.71
Volatile Matter	78.94
Fixed Carbon	7.95
Ash	6.65

Table 3 Proximate analysis of green algae

1.12 Availability

Microalgae cultivation is not a well-known concept in Pakistan, but the ease of cultivation and the environment needed are such that they can be cultivated in a variety of climates considering the large variety of species of microalgae present. In the future as further studies are being done and research takes place specifically in European countries, cultivation of microalgae will become a norm as they are a third-generation biofuel. The ease of cultivation and the lower costs associated with it can easily help mass produce microalgae for large scale processes such as gasification.

1.13 Representative Formula for Microalgae used

$$\mathbf{CH_aO_bN_c}$$

a = H/C   b = O/C c = N/C   as stated in the ultimate analysis.

So, the representative formula for Microalgae becomes:

$$\mathbf{CH_{0.148}O_{0.318}N_{0.028}}$$

1.14 Problem Statement

Microalgae is seen as a potential replacement of first- and second-generation biofuels and can also substitute non-renewable resources. We have designed a bioreactor that is capable of thermochemically converting microalgal biomass to a mixture of gases with a high hydrogen content that will be combusted in a gas turbine to generate a net power output of 1.5 MW.

CHAPTER 2.

PROCESS DESCRIPTION

2.1 Basic Chemistry

2.1.1 Gasification

A thermochemical process that involves the heating and converting of waste organic material i.e carbon containing compounds at a high temperature between 700–1100°C into syngas is termed as gasification. The process is facilitated by a constrained amount of oxygen, air, or steam to generate industrial gas mixtures called syngas.[5]

The gasification mechanism begins with the thermochemical degradation of biomass to Hydrogen, Carbon dioxide, Carbon monoxide, and light hydrocarbons gases with some unwanted products including water and tar. Low moisture content is suitable for gasification. That's the reason why algae whose moisture content is between 3.99 to 10% by weight after drying.[6]

2.1.2 Basic chemistry

The series of reactions that take place in the gasifier includes:

Partial oxidation:

The type of chemical reaction in which a sub-stoichiometric fuel-air mixture is set to partial combustion is termed partial oxidation. 6

$$C + 1/2\ O2 \leftrightarrow CO \qquad \Delta H = -268 MJ/kg\ mol$$

Complete oxidation:

In the complete oxidation, there is no production of carbon monoxide but only carbon dioxide is produced as follows:

$$C + O2 \leftrightarrow CO2 \qquad \Delta H = -393.5 MJ/kg\ mol$$

Water gas reaction:

Water gas reaction also occurs during gasification in which Carbon reacts with water gas to form Carbon monoxide and Hydrogen.

$C + H_2O \leftrightarrow CO + H_2$ $\Delta H = 118 MJ/kg\ mol$

Boudouard Reaction:

The redox reaction of a mixture of carbon monoxide and carbon dioxide is called a boudouard reaction at a given temperature at equilibrium. It is the disproportionation of carbon monoxide into carbon dioxide and graphite or vice versa.

The enthalpy of formation for carbon dioxide is higher than that of carbon monoxide while the formation entropy is much lower. The standard free energy of the formation of CO_2 from its component elements is almost constant, it is temperature independent, while the free energy of formation of CO decreases with the increase of temperature. The overall reaction is endothermic.

$C + CO_2 \leftrightarrow 2CO$ $\Delta H = 170 .7 MJ/kg\ mol$

Water gas shift reaction:

The reaction in which carbon monoxide and water vapor react together to form hydrogen gas and carbon dioxide is a water gas shift reaction. The equilibrium constant has an indirect relation with the temperature that increased conversion of carbon monoxide is observed at lower temperature.

$CO + H_2O \leftrightarrow CO_2 + H_2$ $\Delta H = -42 MJ/kg\ mol$

Methane formation:

Some of the CO will react with Hydrogen to form methane.

$CO + 3H_2 \leftrightarrow CH_4 + H_2O$ $\Delta H = -88 MJ/kgmol$

2.2 Process Flow Diagram:

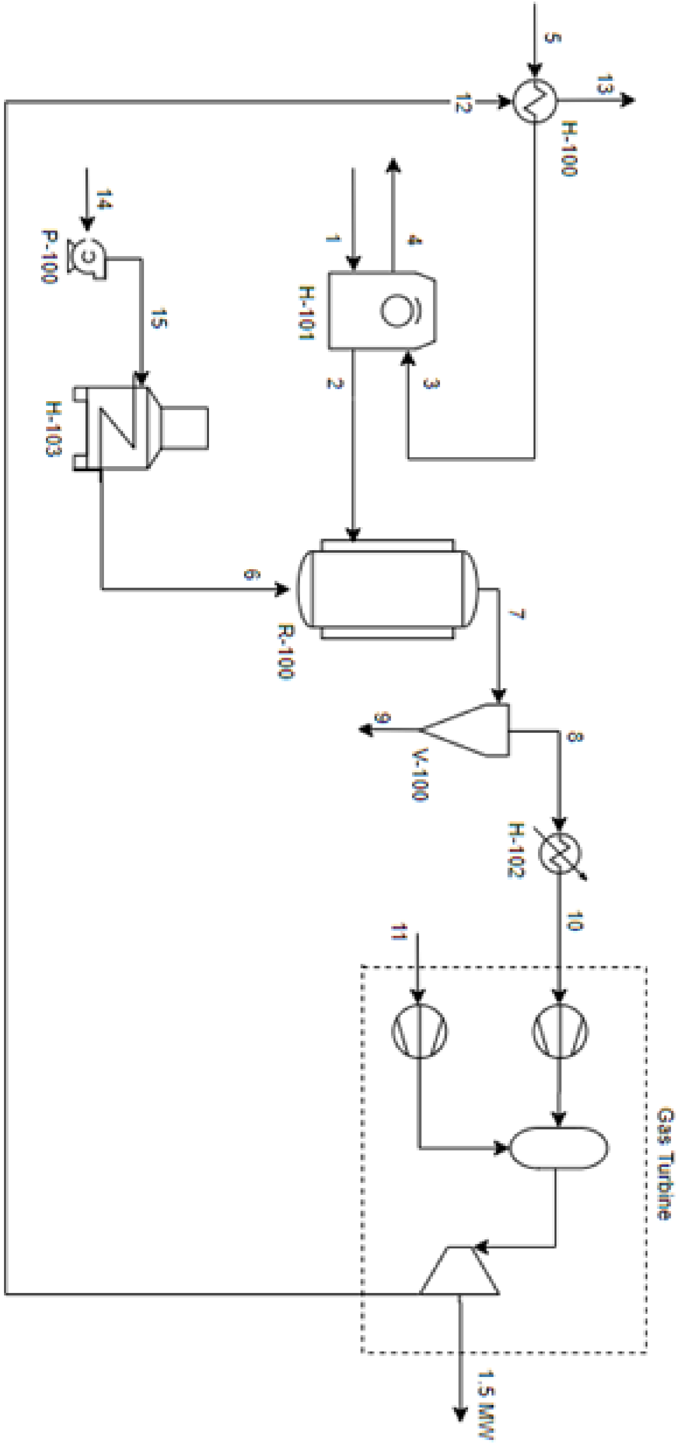


Figure 1 Process Flow Diagram

2.3 Process Description:

2.3.1 Feed Preparation:

We will take the algae after several steps, like cultivation, harvesting etc. Our algae will contain 30% moisture content which needs to be removed and reduced to 10% in order to do effective gasification. So for this purpose we will use a Heat exchanger and a dryer to first dry the algae.

2.3.2 Shell and Tube Heat Exchanger:

Air at 25 degree C will enter a shell and tube heat exchanger and leave at 120-degree C. The other fluid that will be used to heat it up will be the flue gases which will be the product of a gas turbine.

The air will enter and leave from the shell side of the heat exchanger and the flue gases will enter and leave from the tube side. This is because the flow rate of the flue gases was a lot more than the air that is why we used this arrangement. The reason to choose shell and tube heat exchanger over other heat exchangers is the same, our flow rates are very high so shell and tube heat exchanger -was the best possible choice.

2.3.3 Conveyor belt Dryer:

Now to remove the moisture content from the algae, we will use the conveyor belt dryer. We can use other drying equipment but they are costly and ineffective in case of microalgae. For instance, a spray dryer will take a lot of energy to dry the particles. Similarly, freeze drying will use a lot of energy to create vacuum and then dry. Rotary dryer is mostly used for paste like materials so it will also not be a very good choice for this purpose.

Also, a conveyor belt dryer uses low air velocity and temperature so it will not evaporate the volatile components from algae. Also, by using low velocity, the chances for the entrainment of solid particles are very less so we don't need to add any cyclones after the drying process to separate the solids from the air.

Hot Air from the heat exchanger will enter a conveyor belt dryer to dry the algae and reduce the 30% initial moisture content to 10% and will make our feed ready for the bubbling fluidized bed reactor.

2.3.4 Gasifier:

2.3.4.1 Bubbling fluidized bed reactor

Bubbling fluidized bed reactors are utilized in our project. It will help in the fluidization as well as the bubbling process. The process of fluidization refers to the process in which solid particles are molded to behave like fluid through suspension. Bubbling fluidization is achieved by increasing the velocity of the steam, 3 to 7 times of the minimum fluidization velocity. We have used bubbling fluidized bed reactor in our process.

2.3.4.2 Moving Bed

Moving-bed gasifiers are identified as a bed, where the fuel runs slowly downward under gravity as it is gasified. Usually, the arrangement is set concurrently. In the gasification zone, the hot gases are utilized for preheating and pyrolyzing the fuel that is flowing downward. Using the moving bed gasifier, the utilization of oxygen is relatively low but one of its shortcomings is the substantial generation of methane in syngas.

2.3.4.3 Entrained gasifiers

These gasifiers are characterized by separated fuel particles. These particles are carried by a gust in the concurrent flow of reactants and products. The residence time for these gasifiers is few seconds. The high temperature of up to 1600°C guarantees the very low methane content and high fuel conversion.

2.3.4.4 Hybrid and Novel Gasifiers

Other than the fundamental types of gasifiers, there are some gasifiers that are either the hybrid combination of the fundamental gasifiers or novel gasifiers like molten metal baths. One example of a hybrid gasifier is a transport reactor-based gasifier. It is the combination of both, the entrained flow gasifier and the fluidized bed.

2.3.4.5 Steam Gasification Process in bubbling fluidized bed:

Dried algae from the conveyor belt dryer will enter the bubbling fluidized bed reaction to react with steam coming at 800-degree C.

Here steam gasification will occur. Moreover, the process of pyrolysis and gasification is simultaneously done in the bubbling fluidized bed reactor. The first step is pyrolysis in which the algae molecules break down to give off syngas and char, this step is very fast so we ignore this step. As a result of pyrolysis, the resultant products i.e. char undergo steam gasification and result in syngas. The syngas comprises carbon dioxide, carbon monoxide, methane, water vapors, hydrogen and some other hydrocarbons.

In the gasification, several reactions occur that are already explained under the heading of chemistry of gasification. The hydrogen concentration is much higher when steam is used as a gasifying agent as compared to air gasifying agent. The fundamental reaction is water gas reaction that helps vitally in the production of hydrogen. Another disadvantage of using air as the gasifying medium is the high amount of nitrogen which results in a lower energy value gas and as some of the feed is burned, the net amount of fuel left for gasification is decreased.

2.3.5 Cyclone

After the products are collected from the reactor, they are shifted to the cyclone separator. This is specifically done to separate the solid particles from the gases, like char and other solid particles. The intermediate solid residue that is formed during pyrolysis and gasification process is char. Char is separated using the cyclone separator.

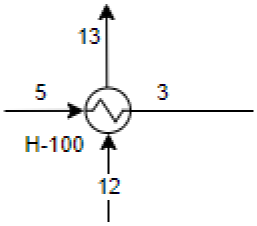
2.3.6 Gas turbine

The clean syngas will then be cooled down and sent to a gas turbine where three processes will occur, compression, combustion and expansion. The syngas and air will be compressed first to 16 pressure ratio and then combusted to generate flue gases, these high pressure and temperature flue gases will be expanded in the gas turbine to generate 1.5MW of power by spinning the blades of the turbine. The expanded low Pressure and Temperature flue gases will then go to the shell and tube heat exchanger to heat the air as discussed earlier. Thermal efficiency of the gas turbine is around 36.8%.

CHAPTER 3.

MATERIAL BALANCE

3.1 Heat Exchanger:



Inlet:

Shell side:

Air in= 10560Kg/h

Air out= 10560Kg/h

Tube Side:

Gas in= 13392kg/h

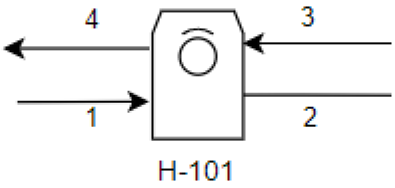
Gas out= 13392kg/h

Overall Balance= In=Out

13392kg/h+10560kg/h=13392kg/h+10560kg/

23952=23952

3.2 Dryer:



Inlet:

Wet Material= 1500kg/h

Air= 10560kg/h

Outlet:

Dried Material= 1100kg/h

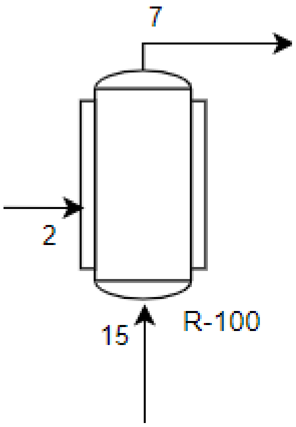
Air= 10960kg/h

Overall Balance=In=Out

1500+10560=1100+10960

12060=12060

3.3 Gasifier:



Inlet:

Dried Material= 1100kg/h

Steam= 863kg/h

Outlet:

Steam=171kg/h

CO=1209.6kg/h

CO2=312.4kg/h

H_2=164.2kg/h

CH_4=8kg/h

N_2=16.24kg/h

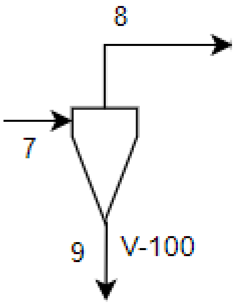
C + Ash= 81.56kg/h

Overall Balance=In=Out

1100+863=171+1209.6+312.4+164.2+8+16.24+81.56

1963=1963

3.4 Cyclone:



Inlet:

Producer gas=1963kg/h

Outlet:

Top Product= 1881.44kg/h

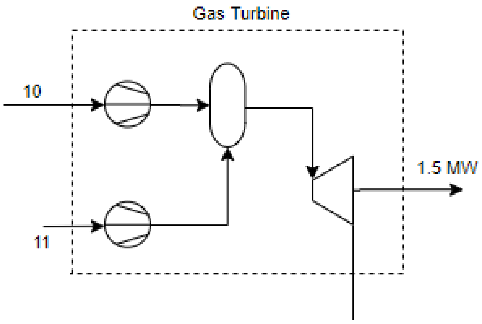
Bottom Product= 81.56kg/h

Overall Balance=In=Out

1963=1881.44+81.56

1963=1963

3.5 Gas Turbine:



Inlet:

Syngas= 1881.44kg/h

Air=11600kg/h

Outlet:

Exhaust Gases= 13498kg/h

Overall Balance=In=Out

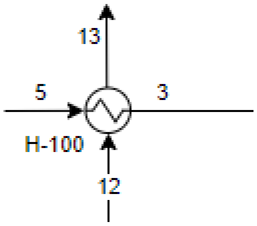
1881.44+11600=13498

13498=13498

CHAPTER 4.

ENERGY BALANCE

4.1 Heat Exchanger:



Shell side:

Air in Energy= -2995kJ/h

Air out Energy= 1.6 x 10^6 kJ/h

Tube Side:

Gas in Energy= -2.3 x 10^7kJ/h

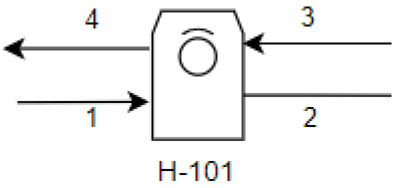
Gas out Energy= -2.498 x 10^7 kJ/h

Overall Balance=In=Out

-2995 + (-2.3 x 10^7) = 1.6 x 10^6 + (-2.498 x 10^7)

-2.3 x 10^7 = -2.33 x 10^7

4.2 Dryer:



Inlet:

Wet Material Energy= -7.6 x 10^6 kJ/h

Inlet Air Energy= 1.0 x 10^6 kJ/h

Outlet:

Dried Material Energy= -1.2 x 10^6 kJ/h

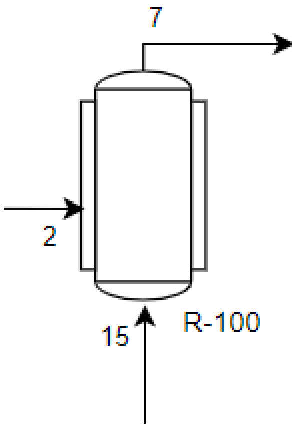
Outlet Air Energy= -5.25 x 10^6 kJ/h

Overall Balance=In=Out

-7.6 x 10^6 + 1.0 x 10^6 = -1.2 x 10^6 + (-5.25 x 10^6)

-6.6 x 10^6 = -6.5 x 10^6

4.3 Gasifier:



Inlet:

Dried Algae with 10% moisture energy= -1.2 x 10^6 kJ/h

Steam Energy= -1.02 x 10^7 kJ/h

Extra Energy Required= Q= 3.7 x 10^6 kJ/h

Outlet:

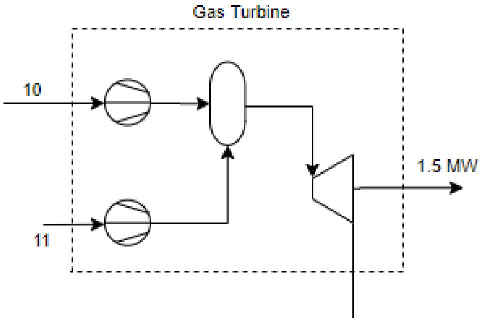
Producer Gas Energy= -6.2 x 10^6 kJ/h

Overall Balance=In=Out

-1.2 x 10^6 + (-1.02 x 10^7) + 3.7 x 10^6 = -6.2 x 10^6

-6.3 x 10^6 = -6.2 x 10^6

4.4 Gas Turbine:



Inlet:

Syngas + air energy before combustion= -7.6 x 10^6 kJ/h

Outlet:

Exhaust Gases energy after expansion= -2.31 x 10^7 kJ/h

Power Generated= 1.5MW= 1.55 x 10^7 kJ/h

Overall Balance=In=Out

-7.6 x 10^6 = -2.31 x 10^7 + 1.55 x 10^7

-7.6 x 10^6 = -7.6 x 10^6

4.4.1 Selection of gas turbine

KAWASAKI M1A-13A GT 1500 kW Gas Turbine CHP Plant

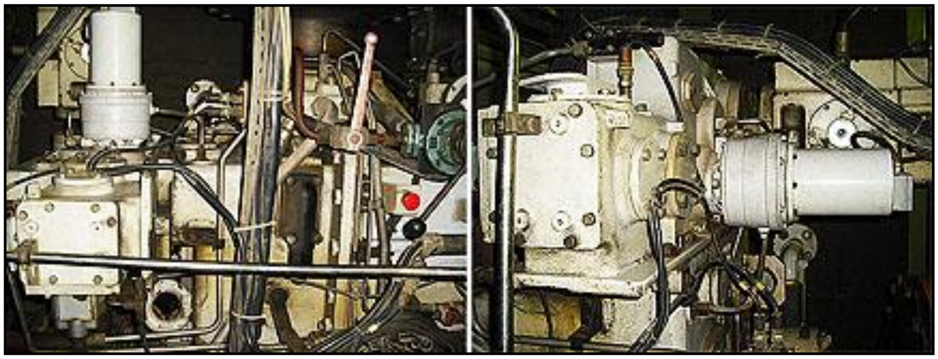
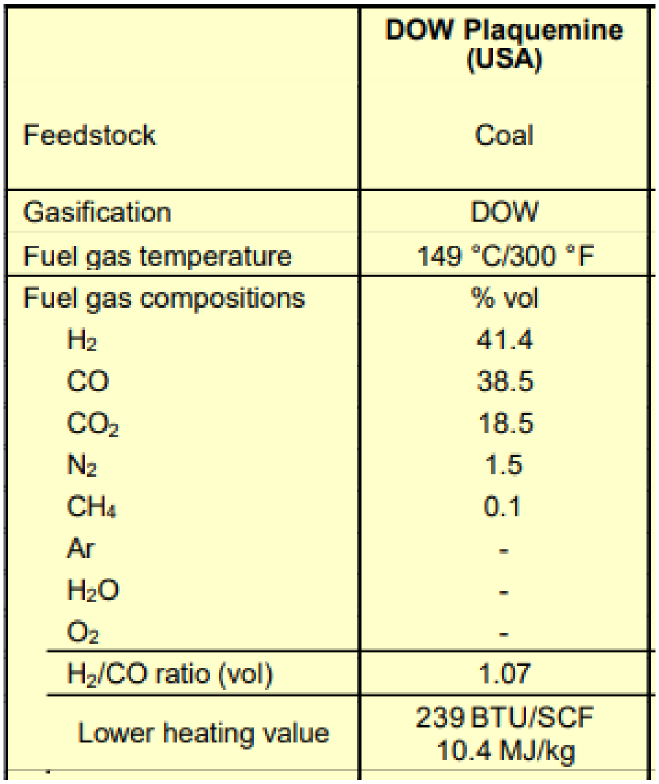


Figure 2 Selected Gas Turbine

Gas Turbine Specifications	
Power	1.5 MW
Thermal Efficiency	36.8%
Turbine Speed	22000 RPM
Fuel Consumption	0.523 kg/s
Air Consumption	3.21 kg/s
LHV of fuel	17.44 MJ/kg
Cooling water requirements	12 t/h
Pressure Ratio	16
Outlet Temperature	600°C
Engine Type	Designed for low quality

Table 4 Selected Gas Turbine Specification

Typical specifications:



	DOW Plaquemine (USA)
Feedstock	Coal
Gasification	DOW
Fuel gas temperature	149 °C/300 °F
Fuel gas compositions	% vol
H_2	41.4
CO	38.5
CO_2	18.5
N_2	1.5
CH_4	0.1
Ar	-
H_2O	-
O_2	-
H_2/CO ratio (vol)	1.07
Lower heating value	239 BTU/SCF 10.4 MJ/kg

Table 5 Typical Gas Turbine feed compositions

Process Stream Specifications

Feedstock	Algal Biomass
Gasification	BFB
Fuel gas composition	% vol
Hydrogen	57.0
Carbon Monoxide	30.0
Carbon Dioxide	4.9
Nitrogen	0.4
Methane	0.35
Water (Steam)	6.66
Hydrogen to Carbon Monoxide Ratio (vol)	1.9
LHV	16.55 MJ/kg

Table 6 Selected Gas Turbine feed compositions

CHAPTER 5.

DESIGNING

5.1 Heat exchanger

Flue gas (600 ∘ C)

Air:

In = 25 ∘ C

Out = 135 ∘ C

Shell

ID = 2.55m

Baffle space = 5 inch

Passes = 1

Tube

OD = 0.12m

Number and length = 15 and 3m

Passes = 1

Hot Fluid		Cold Fluid	Differences
600	Higher Temperature	135	465
450	Lower Temperature	25	425

150		110	40

Table 7 In and Out Temperatures

LMTD = 352.5 $\circ$ C

$S = \frac{110}{600-25} = 0.191$ $R = \frac{150}{110} = 1.36$

$F_T = 0.8$

$\triangle t = LMTD \ x \ F_T$ = 352.5 x 0.8

$\triangle t = 282$

<u>Shell</u>

Flow area, a_s= $\frac{\text{IDxC}'\text{B}}{144xP_T}$ = $3.405 \text{x} 10^{-3}$ m^2

Mass velocity = $\frac{10{,}960}{3.405 \text{x} 10^{-3}}$ = $3.2x10^6$ kg/hr.m^2

Re = $\frac{\text{DGs}}{\mu}$ = 25,9047.61

J_H= 98

C = 1.287

K = 0.024

$(\frac{\text{C}\mu}{k})^{0.5}$ = 4.10997

<u>Tube</u>

$$a_t = \frac{\text{No. of tubes} \times \text{flow area/tube}}{\text{No. of passes}}$$

a_t = $\frac{N_{ta'_t}}{144n}$

Flow area a'_t = $3.81924 \text{x} 10^{-3}$ m^2

Mass velocity = $\frac{10{,}560}{3.81924 \text{x} 10^{-3}}$ = $2.76x10^6$ kg/hr.m^2

$a_t = \frac{N_{t}a'_{t}}{144n}$ = 3.978x10^{-4} m^2

Re = 8420

J_H = 31

c = 1.012 KJ/kg K

μ = 0.0015

Heat Transfer surface

Total surface area = 61.5 m^2

$U_D = \frac{\text{Q}}{\text{A}\triangle\text{T}}$

Dirt Factor

$R_d = \frac{U_C - U_D}{U_C \cdot U_D}$ = 0.00318

Pressure Drop

Shell side:

For Re = 25,9047.61, f = 0.0018

No. of crosses = N+1 12L/B

$$\Delta P_s = \frac{fG_s^2 D_s (N+1)}{5.22 \times 10^{10} D_e s \phi_s} \text{ psi}$$

△P = 24.13 kPa

Tube side:

For Re = 8420, f = 0.000294

$$\Delta P_t = \frac{fG_t^2 Ln}{5.22 \times 10^{10} D s \phi_t}$$

△P = 63.43 kPa

5.2 Dryer

$$F_{AC}(h_{AO} - h_A) = F_S(h_S - h_{SO})$$

F_{AC} = 0.416 kg/s

h_{Ao} = 70.43 kJ/kg

h_A = 118.19 kJ/kg

h_{So} = 49.31 kJ/kg

h_S = 177 kJ/kg

F_s = 0.424 kg/s

m_{in} = 0.14 kg/s

m_{out} = 0.03 kg/s

$\Delta\dot{m}$ = 0.11 kg/s

λo = 0.00764 kg/(s*m^2)

X_{So} = 0.33

λ = 0.00252 kg/(s*m^2)

$$A = \frac{a \cdot \Delta m}{\lambda}$$

a = 0.25 m^2

$\Delta\dot{m}$ = 0.11 kg/s

λ = 0.00252 kg/(s*m^2)

A = 10.91 m^2

$$F_S = V_{Belt} \cdot b_{Belt} \cdot H_{Bed} \cdot \rho_{Product}$$

b (width of belt) = 0.75 m (assumed)

H (height) = 1 m (assumed)

ρ = 570 kg/m^3

V (belt) = 7.13$\times 10^{-4}$ m/s

Parameter	Value
F_S	0.424 kg/s
X_{So}	0.33
X_S	0.10
ρ	570 kg/m^3
h_{Ao}	70.43 kJ/kg
h_A	118.19 kJ/kg
h_{So}	49.31 kJ/kg
h_S	177 kJ/kg
P	101.3 kPa
Q	21.240 kW
$\Delta\dot{m}$	0.11 kg/s
λ (avg moisture removal)	0.00252 kg/(s*m^2)
λo	0.00764 kg/(s*m^2)
a	0.25m2
A	10.91 m^2
b (width of belt)	0.75 m
H (height)	1 m
L (length)	14.55 m
V (belt)	7.13$\times \mathbf{10^{-4}}$ m/s

Table 8 Dryer Specifications summary

5.3 Gasifier

The first step in designing is the elemental balances for output composition which will help us in determining the exact size of the gasifier. This is done by assuming equilibrium is achieved for all reactions which is an ideal case but will give us some idea to work with. Seven equations are formed with seven unknowns with a steam to biomass ratio set at 0.863. solving these equations gives us the total flowrate of feed as well as that of the outlet gases and its composition, useful for determining the LHV of the fuel gas.

Carbon
$$1.866FX_c = V_{CO} + V_{CO_2} + V_{CH_4}$$

Hydrogen
$$1.24FS + (11.21FX_H + 1.24FW) = V_{H_2} + V_{H_2O} + 2V_{CH_4}$$

Oxygen
$$F\left(\frac{S}{36} + \frac{X_O}{32} + \frac{W}{36} + \frac{AO_a}{32}\right) = \frac{(0.5V_{CO} + V_{CO_2} + 0.5V_{H_2O})}{22.4}$$

Nitrogen
$$0.8FX_N + 0.8FAN_a = V_{N_2}$$

Overall Balance
$$V_{CO} + V_{CO_2} + V_{H_2} + V_{CH_4} + V_{H_2O} + V_{N_2} = 1$$

Boudard Reaction
$$K_{pb} = \frac{P_{CO}^2}{P_{CO_2}}$$

Water gas Reaction
$$K_{pw} = \frac{P_{H_2}P_{CO}}{P_{H_2O}} = V_{H_2}V_{CO}\frac{P}{V_{H_2O}}$$

The values used in the equations are shown in the table below.

S = 0.863	$\mathbf{V_{CO}}$ **= 0.25**
A = 0	$\mathbf{V_{CO_2}}$ = 0.0345
$\mathbf{X_C}$ **= 0.6251**	$\mathbf{V_{H_2}}$ = 0.594
$\mathbf{X_H}$ **= 0.0924**	$\mathbf{V_{H_2O}}$ = 0.058
$\mathbf{X_N}$ **= 0.0176**	$\mathbf{V_{CH_4}}$ = 0.053
$\mathbf{X_O}$ **= 0.1985**	$\mathbf{V_{N_2}}$ = 0.004
W = 0.1671	F = 0.31kg/Nm3
$\mathbf{K_{pb}}$ **= 1.9**	
$\mathbf{K_{pw}}$ **= 2.617**	

Table 9 Unknowns and Knowns for elemental balance calculations

Next, using equations of fluidized beds, we find the velocity of the entering fluidizing and gasifying medium which is steam and this helps us to find the area of the base of the reactor and consequently, the diameter of the reactor as shown below.

$\emptyset_s$ = 0.8

ε_{mf} = 0.45

ρ_s = 2650 kg/ m^3

ρ_g = 0.205 kg/ m^3

μ_{steam} = 4.04×10^{-5} kg/ms

d_p = 350 μm

$$U_{mf} = \frac{d_p^2\ (\rho_s - \rho_g)\ g}{150\ \mu} \times \frac{\varepsilon_{mf}^3 \times \emptyset_s^2}{1-\varepsilon_{mf}}$$

U_{mf} = 0.131 m/s

U_f = 4 × U_{mf}

U_f = 0.523 m/s

Area of base = Volumetric flow rate (V_g) ÷ Superficial Velocity (U_f)

V_g = Mass Flow rate ÷ Density

= 863/3600 ÷ 0.205

= 1.17 m^3/s

Area = 1.17 ÷ 0.523

= 2.24 m^2

Diameter = 1.69 m

$H_{fb} = \frac{D-2.51D^{0.2}}{0.13D^{0.47}}$

= 1.12 m

TDH = $0.85U^{1.2}(7.33 - 1.2logU)$

= 3 m

Height of freeboard = TDH

$u^{*} = U_f \left[\frac{\rho_g^2}{\mu\,(\rho_s-\rho_g)\,g}\right]^{\frac{1}{3}}$

u^{*}= 0.179

$d^{*} = d_p \left[\frac{\rho_g\,(\rho_s-\rho_g)\,g}{\mu^2}\right]^{\frac{1}{3}}$

d^{*} = 5.34

$U_t = u^{*} \left[\frac{\mu\,(\rho_s-\rho_g)\,g}{{\rho_s}^2}\right]^{\frac{1}{3}}$

U_t= 1.13 m/s

The residence time approach assumes that the water gas reaction is the main gasification reaction,

$$C + H_2O \rightarrow CO + H_2$$

Where char is consumed by the steam gasification reaction for nth order kinetics.

$$\frac{1}{m}\frac{dC}{dt} = k[H_2O]^n$$

Where n=1

m = 0.28 kg (mass of biomass)

k = 0.003 (literature)

n_{H_2O} = 863000 ÷ 18

= 47933.3

1 mol water = 89.7 l

47933.3 mol = 4299617.01 l

C_{H_2O} = (47933.3 × 1000) ÷ 4299617.01

= 11.15 mol/m^3

$\frac{dC}{dt}$ = r = 0.0093 kg/s

$$\theta = \frac{C_o \times X}{r}$$

C_o = 0.174 kg (initial mass of carbon)

X = 0.9746

r = 0.0093

θ = 18.2s

$$V_{bed} = \frac{F(C) \times \theta}{(1-\varepsilon)\rho_s x_{char}}$$

$\varepsilon = 0.4$

ρ_s = 570 kg/m3

x_{char} = 0.05

V_{bed} = 2.5 m3

Area = 2.237 m2

H_{bed} = 1.12 m

Total bed height = 1.12 + 3

= 4.12 m

Total Volume = 9.22 m3

$$\Delta P = \left(1 - \varepsilon_{mf}\right)\left(\rho_s - \rho_g\right) \times g \times H_b$$

ΔP = 17.45 kPa

Design Summary	
Carbon Conversion	97.46%
S/B ratio	0.863
Operating Temperature	800°C
$\mathbf{H_2/CO}$	1.9
Cold Gas efficiency	76.4%
Fluidization Velocity	0.523 m/s
Diameter	1.69 m

Base Area	2.24 m^2
Rate	0.0093 kg/s
Residence Time	18.2 s
Bed Height	1.12 m
Freeboard Height	3 m
Total Volume	9.22 m^3
Pressure Drop	17.45 kPa
Bed Material	Silica sand

Table 10 Gasifier Design Summary

5.4 Cyclone

The basic method to design a cyclone separator for gas-solid systems is to use either of the two available standard designs of cyclone as in Stairmand's method. The flowrate of the exiting gas in our system is quite high so we have chosen a high throughput cyclone separator which has the dimensions shown in figures 9 and 10 in the appendix.

Particle size range: 100 – 200 μm

Flow rate of gas = 12677.42 m^3/h

= 3.52 m^3/s

Flow rate of gas per cyclone = 0.88 m^3/s

Area of inlet duct at 15 m/s = 0.88 / 15

= 0.059 m^2

D_c = 0.46 m

Now, using this diameter and values for the standard cyclone and our process, we calculate a scaling factor which will be used to determine the grade efficiency of the selected sized particles.

$$d_2 = d_1 \left[\left(\frac{D_{c_2}}{D_{c_1}} \right)^3 \times \frac{Q_1}{Q_2} \times \frac{\Delta\rho_1}{\Delta\rho_2} \times \frac{\mu_2}{\mu_1} \right]^{1/2}$$

$\mathbf{D_{c1}}$ = 0.203 m

$\mathbf{Q_1}$ = 669 $\mathbf{m^3}$/h

$\Delta\boldsymbol{\rho_1}$ = 2000

$\boldsymbol{\mu_1}$ = 0.018 mN s/$\mathbf{m^2}$

$\boldsymbol{\mu_2}$ = 0.033 mN s/$\mathbf{m^2}$

$\mathbf{D_{c2}}$ = 0.46 m

$\mathbf{Q_2}$ = 3169.4 $\mathbf{m^3}$/h

$\Delta\boldsymbol{\rho_2}$ = 570

Scaling factor = 2.3

$\mathbf{d_2}$ = 100

$\mathbf{d_1}$ = 43.5

Using figure 7 from the appendix, the grade efficiency of particles sized 43.5 micrometers is 100% which can easily be translated for 100 micrometer sized particles as in our system.

Grade Efficiency = 100%

Pressure drop:

$$\Delta P = \frac{\rho_f}{203}\left\{u_1^2\left[1 + 2\phi^2\left(\frac{2r_t}{r_e} - 1\right)\right] + 2u_2^2\right\}$$

Using values

Pressure drop = 3 mbar, within suitable range

CHAPTER 6.

ECONOMICS

Inflation index = 1.41

Cost in 2021 = Cost in 2004 x Inflation (1.41)

Cost of Heat exchanger = 24,800 x 1.41

= 35,000 $

Cost of dryer = 80,000 x 1.41

= 113,000 $

Cost of Gasifier = 15000 x $9.\mathbf{22}^{0.4}$ x 2 x 1.41

= 100,000 $

Cost of cyclones = 2304 x 4 x 1.41

= 13,000 $

Cost of Gas Turbine = 500,000 $

PCE = 0.76 million $

PPC = 3.15 x 0.76

= 2.3 million $

Fixed Capital = 1.4 x 2.3

= 3.26 million $

Working capital = 10% of Fixed Capital

= 0.33 million $

Total investment = 3.59 million $

Cost of raw material algae = 0.03 dollars/kg

= 0.24 million $

Variable costs

$\boldsymbol{f_1}$ = 0.24 million $

$\boldsymbol{f_2}$ = 16100

$\boldsymbol{f_3}$ = 0.394 million $

$\boldsymbol{f_4}$ = --

Fixed Costs

$\boldsymbol{f_5}$ = 0.161 million $

$\boldsymbol{f_6}$ = 30,000

$\boldsymbol{f_7}$ = --

$\boldsymbol{f_8}$ = 15,000

$\boldsymbol{f_9}$ = 9000

$\boldsymbol{f_{10}}$ = 0.2 million $

$\boldsymbol{f_{11}}$ = 32,200

$\boldsymbol{f_{12}}$ = --

Annual Operating Costs = 1.1 million $

Revenue

Production = 1.5 MW

= 1500 units

Cost of one unit = 0.147 $

Annual revenue = 0.147 x 1500 x 24 x 365

Annual revenue = 1.93 million $

$$\text{Payout Period} = \frac{\textbf{Total Investment}}{\textbf{Annual Revenue} - \textbf{Annual Operating Costs}}$$

$$= \frac{\mathbf{3.59}}{\mathbf{1.93 - 1.10}}$$

= 4.3 years

	Cost (USD $)
Fixed Capital	3.26 million
Total Investment	3.59 million
Variable Costs	0.65 million
Annual Operating Costs	1.10 million
Annual Revenue	1.93 million
Annual Profit	0.83 million
Payout Period	4.3 years

Table 11 Economic Analysis Summary

CHAPTER 7.

SIMULATION

Fluid package used was based on the gases that were being produced in the gasifier. Hence, due to the presence of acidic gases, the choice of Peng-Robinson equation of states was ideal.

Here we have represented the gasifier in a conversion reactor using mass balance values worked on earlier. The results coming from the conversion reactor were similar to our manual calculations. The gas turbine was also simulated using two compressors, a conversion reactor for combustion of syngas, and an expander. The net power production using this setup was near 1.5 MW which supported our calculations.

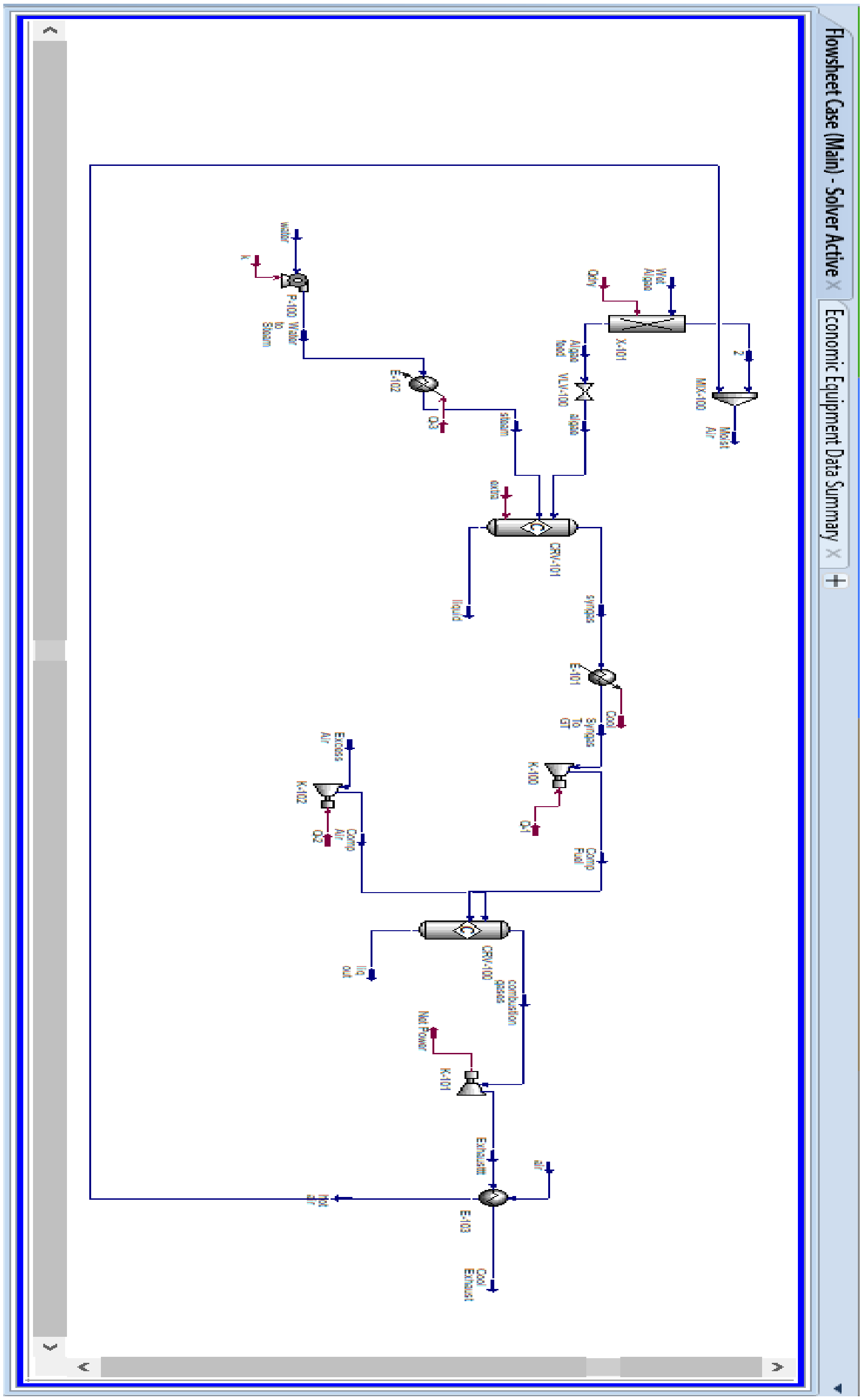


Figure 3 Simulation of Gasification and gas turbine

CHAPTER 8.

HAZOP STUDY

8.1 Introduction

HAZOP Analysis:

Hazards are a part of chemical processes. We cannot completely eliminate the threats. However, we can take actions to minimize these risks. The Hazop study carries out a detail risk assessment of what hazards might occur when a process is fully functional.

Hazard and operability study analysis (HAZOP) is a qualitative technique, done to determine the potential hazard of equipment. HAZOP is done on both on and off-site equipment. The main aim of the HAZOP is to identify the hazardous situations and to determine the efficient working of the safeguards to prevent unwanted scenarios or to control those scenarios.

HAZOP is important to operate the plant within safe limits. HAZOP is necessary before building a chemical plant. In other words, it's a legal requirement to quantify all the hazards and risk present on the plant.

Before conducting the HAZOP, our group reviewed all the critical parameters and piping and instrumentation diagrams including equipment.

The parameters included in the study are:

- Flow
- Temperature
- Pressure
- And the Guide words selected are:
- None
- More
- Less

- Other
- As well as
- Reverse

These guide words were applied to each equipment and the deviation of each parameter from their intention, possible causes and consequences were studied.

8.2 Dryer

Guide Word	Deviation	Possible Causes	Consequences	Action
None	No Flow	Valve malfunction, Line Fracture	No change in moisture content	Inspection of transfer lines Installation of bypass valve
Less of	Less Flow	Leakage, Line damage, Valve malfunctioning	Lower efficiency	Regular patrolling and inspection of transfer lines
More of	More Temperature	Equipment Failure	Unfavorable conditions in dryer	Installation of temperature sensor
Other	Equipment Failure	Pump damage	Product loss, line failure	Installation of bypass line for pump
More of	More flow	Valve malfunction	Filter clogging	Installation of flow control on outlet of reactor, Installation of bypass line
	More Pressure	Vent valve choked	Unfavorable reaction conditions, Lower yield	Install a Relieve valve (PSV)
Other	Equipment Failure	Power malfunction	Poor moisture control	Install a stabilizer

8.3 GASIFIER

Guide Word	Deviation	Possible Causes	Consequences	Action
More of	Higher pH More temperature	1. Valve fails causing high feed flow. 2. External fire. 3. Valve malfunctions causing increase in temperature of fluid entering Too much increase in temperature in reactor	Feed to Reactor enter at high temperature than normal.	Use high PH alarm. Use temperature alarming system in reactor.
	Higher pH	Valve malfunctioning	Feed to Reactor enter at high temperature than normal. Activity may affect if PH increases from the PH range. Reaction may be affected due to increase from the PH range.	Use high PH alarm.
Others	High solid content.	High amount of unreacted material arise from reactor	Damage to pump	Use strainer or filter before pump suction.
None	No flow	Failure of pump	Reactor runs empty. Damage to the pump. No feed to Reactor	Maintenance of valve and instrumentation.

8.4 HEAT EXCHANGER

Guide Word	Deviation	Possible Causes	Consequences	Action
As well as	Impurities	1. Presence of different containment in feed. Affect the performance of heat exchanger.	Affect the performance of heat exchanger.	Regular checking required.
Others	Maintenance	Accumulation in shell and tube side.	Resulting in corrosion.	Regular inspection should be done.
More of	More flow	Valve fails due to which feed flow rate increases.	Quantity of feed increases thereby increasing load on the tubes. Improper heat transfer	Regular checking and maintenance of valve
None	No flow	Failure of pump	No heat transfer. Same exit and entrance temperature.	Maintenance of valve and instrumentation. Proper inspection and maintenance of line should be implemented.

8.5 COOLER

Guide Word	Deviation	Possible Causes	Consequences	Action
None	No cooling water flow	Failure of inlet cooling water valve to open	Process fluid temperature is not lowered accordingly	Install Temperature indicator before and after the process fluid
More	More cooling water flow	Failure of inlet cooling water valve to close	Output of Process fluid temperature too low	Install Temperature indicator before and after the process fluid line
Less	Less cooling water flow	Pipe leakage	Process fluid temperature too low	Installation of flow meter
More of	More pressure on tube side	Failure of process fluid valve	Bursting of tube	Install high pressure alarm
	More Temperature	Cooler fails to operate Hot Weather or External fire.	Feed to Reactor enter at high temperature than normal. Temperature and pressure Increases, causing pump damage. Possible Pump failure.	Use temperature alarming system
Reverse	Reverse process fluid flow	Failure of process fluid inlet valve	Product off-set	Install check valve
Contamination	Process flow contamination	Contamination in cooling water	Output temperature too low	Proper maintenance and operator alert
Corrosion	Corrosion of pipe	Hardness of cooling water	Less cooling and crack of pipe	Proper maintenance

8.6 Safety Conclusions.

1. Once we have reviewed all the risks and hazards that might occur in the plant, we developed this HAZOP study. This enables us to operate according to a standard and to produce a high-quality product while ensuring the safety limits.
2. Having a complete knowledge about all the hazardous material and the operating conditions, HAZOP is necessary to ensure the safety of the workers.
3. It is important to operate the plant within safe limits and everyone should be careful in order to ensure a safe working environment.

CONCLUSION

The current research and technology based on the third-generation biofuels derived from algal biomass have been considered as the best alternative that avoids the disadvantages of first- and second-generation biofuels. Microalgae are looked upon as a promising feedstock for energy production and are thus called the third-generation fuel.

According to research microalgae are the untapped resource with more than 25,000 species of which only 15 are in use. Micro-algae are preferred due to their ability to grow in diverse environments. Micro-algae are considered an important biomass source; however, their consumption as an energy source is still quite low.

This project was designed to evaluate the process of biomass gasification. For this purpose, we have designed an overall process for hydrogen production from algal biomass using steam gasification as the main step. Theoretical hydrogen yield for the feedstock used in the study was determined. This project turned out to be feasible as the payback period is within a reasonable time frame. As the requirements for the project were suited for a typical pilot plant, we believe that after thorough research, this system could be scaled up to industrial level providing cheap and clean electricity using renewable resources and the benefits will eventually be seen at every household.

APPENDIX

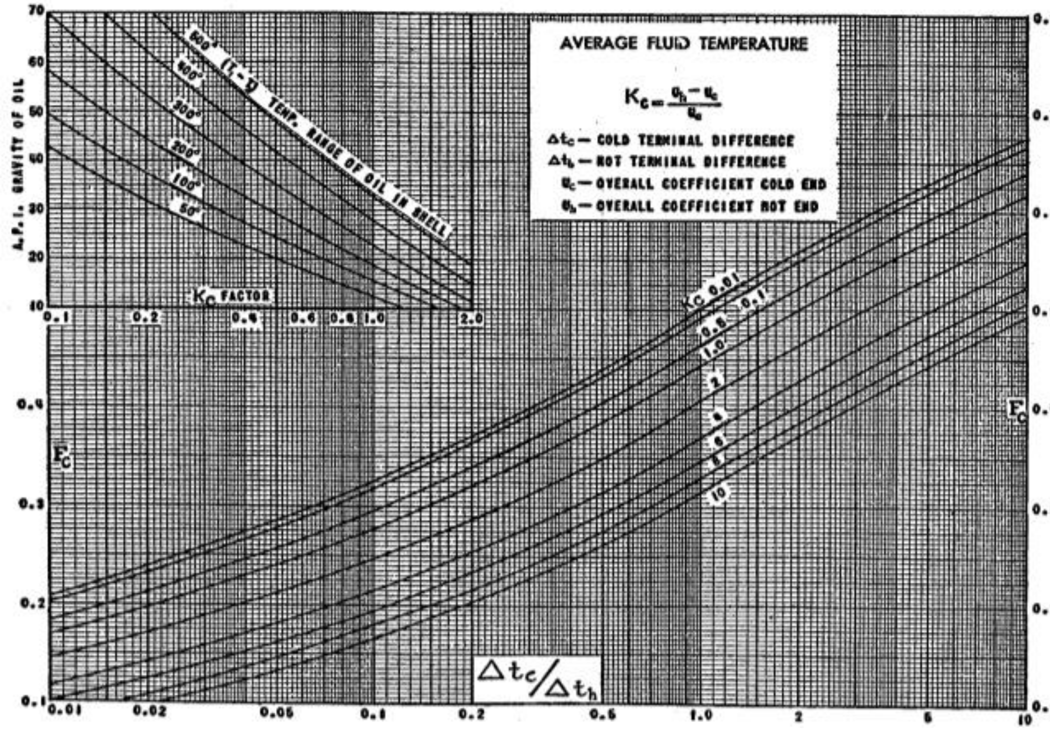


Figure 4 The calorific temperature F_C

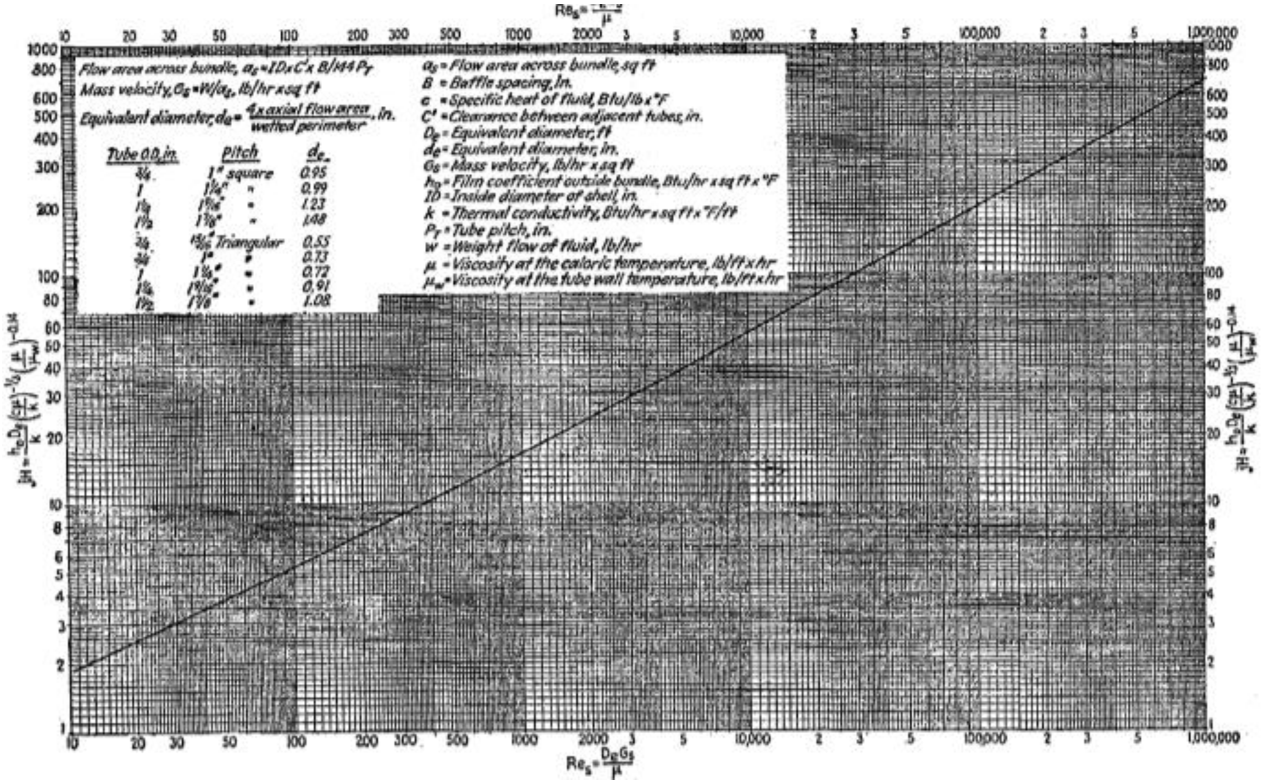
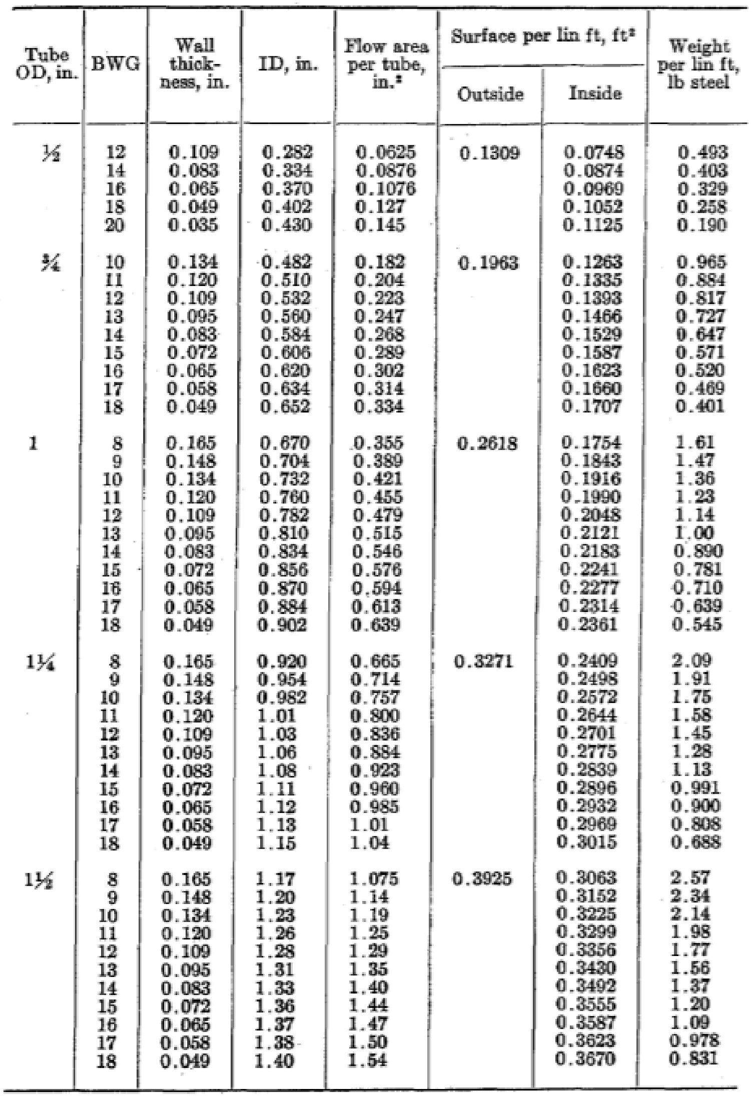


Figure 5 Shell side heat transfer curve for bundle with 25% cut segmental baffles



Tube OD, in.	BWG	Wall thick-ness, in.	ID, in.	Flow area per tube, in.^{2}	Surface per lin ft, ft^2		Weight per lin ft, lb steel
					Outside	Inside	
½	12	0.109	0.282	0.0625	0.1309	0.0748	0.493
	14	0.083	0.334	0.0876		0.0874	0.403
	16	0.065	0.370	0.1076		0.0969	0.329
	18	0.049	0.402	0.127		0.1052	0.258
	20	0.035	0.430	0.145		0.1125	0.190
¾	10	0.134	0.482	0.182	0.1963	0.1263	0.965
	11	0.120	0.510	0.204		0.1335	0.884
	12	0.109	0.532	0.223		0.1393	0.817
	13	0.095	0.560	0.247		0.1466	0.727
	14	0.083	0.584	0.268		0.1529	0.647
	15	0.072	0.606	0.289		0.1587	0.571
	16	0.065	0.620	0.302		0.1623	0.520
	17	0.058	0.634	0.314		0.1660	0.469
	18	0.049	0.652	0.334		0.1707	0.401
1	8	0.165	0.670	0.355	0.2618	0.1754	1.61
	9	0.148	0.704	0.389		0.1843	1.47
	10	0.134	0.732	0.421		0.1916	1.36
	11	0.120	0.760	0.455		0.1990	1.23
	12	0.109	0.782	0.479		0.2048	1.14
	13	0.095	0.810	0.515		0.2121	1.00
	14	0.083	0.834	0.546		0.2183	0.890
	15	0.072	0.856	0.576		0.2241	0.781
	16	0.065	0.870	0.594		0.2277	0.710
	17	0.058	0.884	0.613		0.2314	0.639
	18	0.049	0.902	0.639		0.2361	0.545
1¼	8	0.165	0.920	0.665	0.3271	0.2409	2.09
	9	0.148	0.954	0.714		0.2498	1.91
	10	0.134	0.982	0.757		0.2572	1.75
	11	0.120	1.01	0.800		0.2644	1.58
	12	0.109	1.03	0.836		0.2701	1.45
	13	0.095	1.06	0.884		0.2775	1.28
	14	0.083	1.08	0.923		0.2839	1.13
	15	0.072	1.11	0.960		0.2896	0.991
	16	0.065	1.12	0.985		0.2932	0.900
	17	0.058	1.13	1.01		0.2969	0.808
	18	0.049	1.15	1.04		0.3015	0.688
1½	8	0.165	1.17	1.075	0.3925	0.3063	2.57
	9	0.148	1.20	1.14		0.3152	2.34
	10	0.134	1.23	1.19		0.3225	2.14
	11	0.120	1.26	1.25		0.3299	1.98
	12	0.109	1.28	1.29		0.3356	1.77
	13	0.095	1.31	1.35		0.3430	1.56
	14	0.083	1.33	1.40		0.3492	1.37
	15	0.072	1.36	1.44		0.3555	1.20
	16	0.065	1.37	1.47		0.3587	1.09
	17	0.058	1.38	1.50		0.3623	0.978
	18	0.049	1.40	1.54		0.3670	0.831

Figure 6 Heat Exchanger Tube data

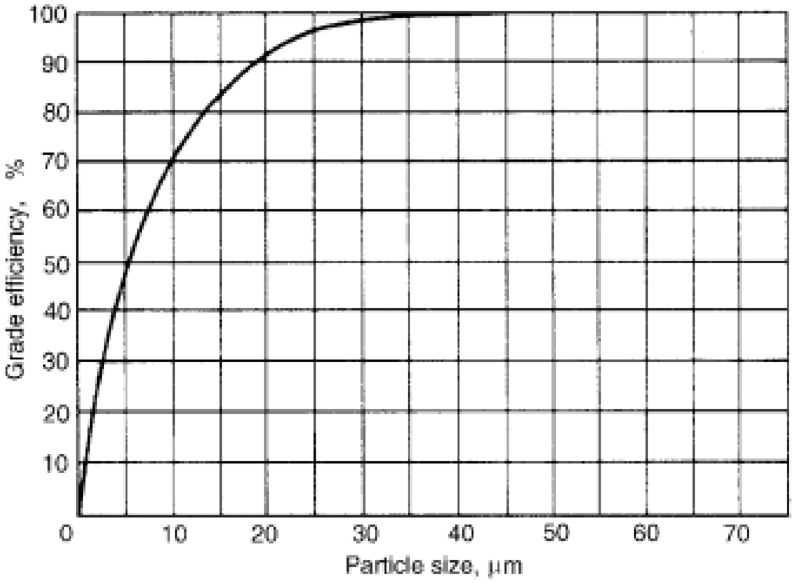


Figure 7 Performance Curves, standard conditions High gas rate cyclone

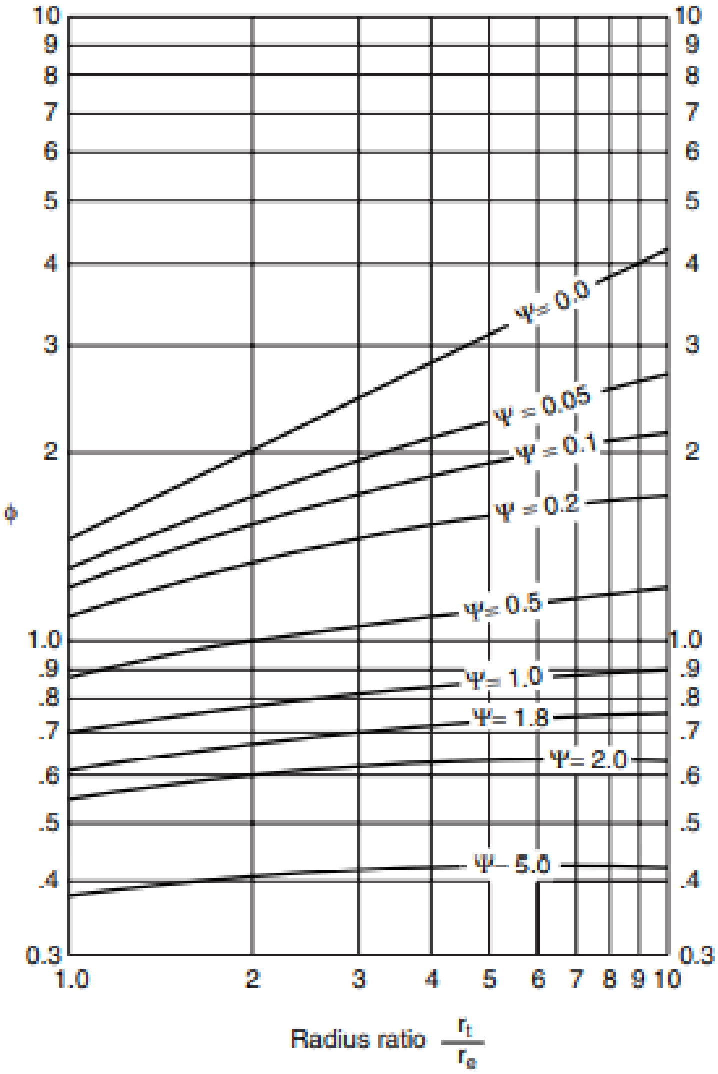


Figure 8 Cyclone pressure drop factor

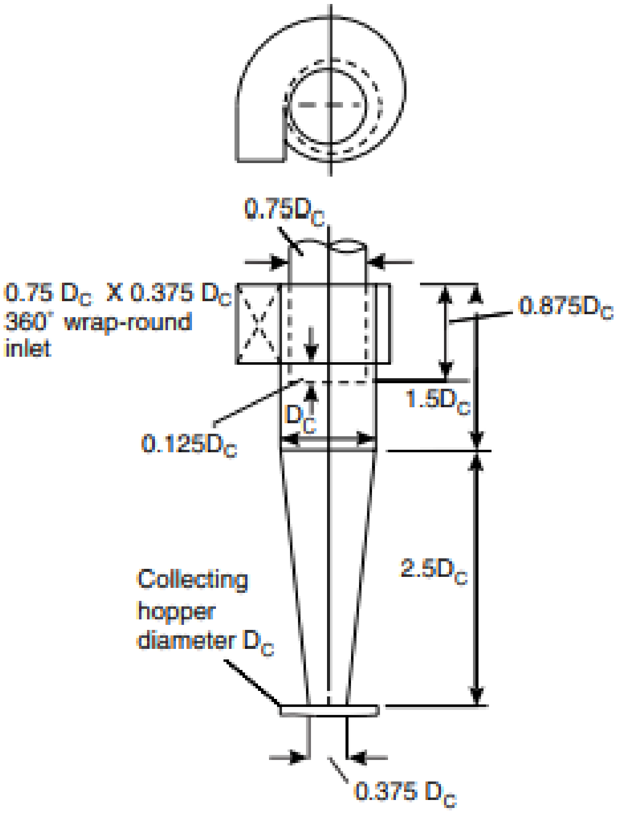


Figure 9 High gas flow rate cyclone dimensions

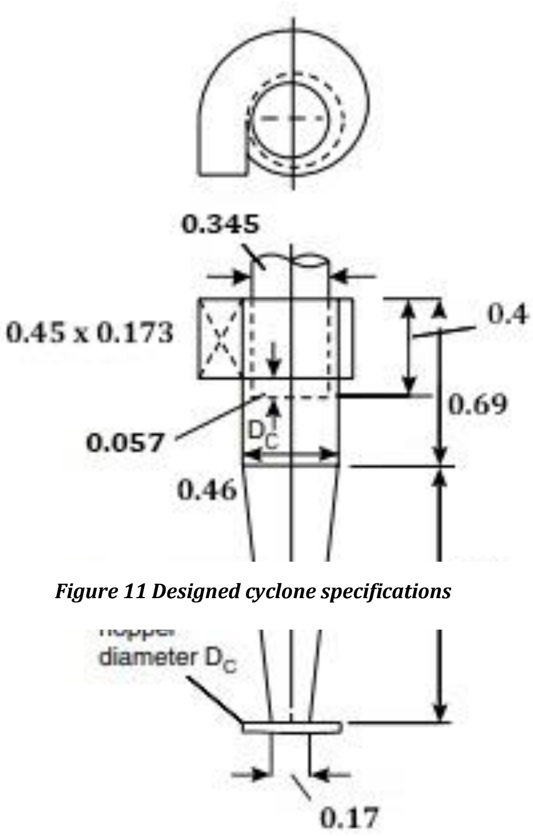


Figure 11 Designed cyclone specifications

Figure 10 Proposed high gas flow rate cyclone dimensions

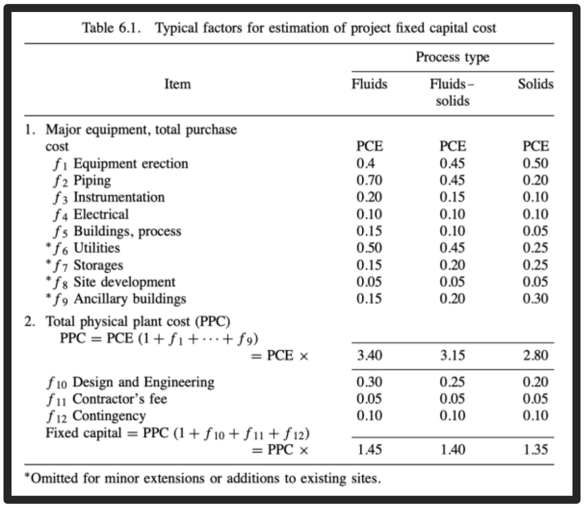


Table 6.1. Typical factors for estimation of project fixed capital cost

Item	Process type		
	Fluids	Fluids–solids	Solids
1. Major equipment, total purchase cost	PCE	PCE	PCE
f_1 Equipment erection	0.4	0.45	0.50
f_2 Piping	0.70	0.45	0.20
f_3 Instrumentation	0.20	0.15	0.10
f_4 Electrical	0.10	0.10	0.10
f_5 Buildings, process	0.15	0.10	0.05
*f_6 Utilities	0.50	0.45	0.25
*f_7 Storages	0.15	0.20	0.25
*f_8 Site development	0.05	0.05	0.05
*f_9 Ancillary buildings	0.15	0.20	0.30
2. Total physical plant cost (PPC)			
$PPC = PCE\,(1 + f_1 + \cdots + f_9)$			
$= PCE \times$	3.40	3.15	2.80
f_{10} Design and Engineering	0.30	0.25	0.20
f_{11} Contractor's fee	0.05	0.05	0.05
f_{12} Contingency	0.10	0.10	0.10
Fixed capital $= PPC\,(1 + f_{10} + f_{11} + f_{12})$			
$= PPC \times$	1.45	1.40	1.35

*Omitted for minor extensions or additions to existing sites.

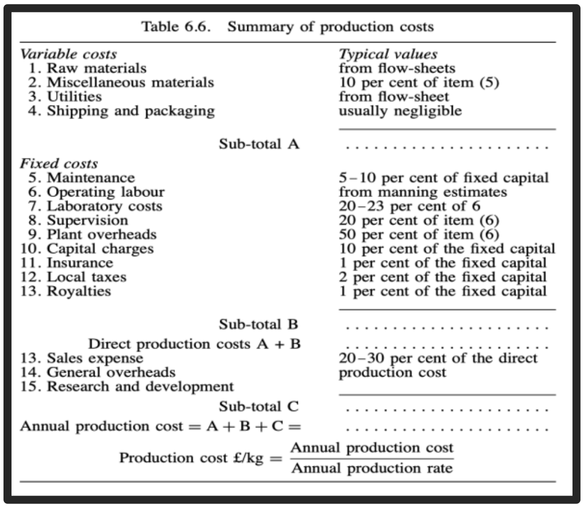
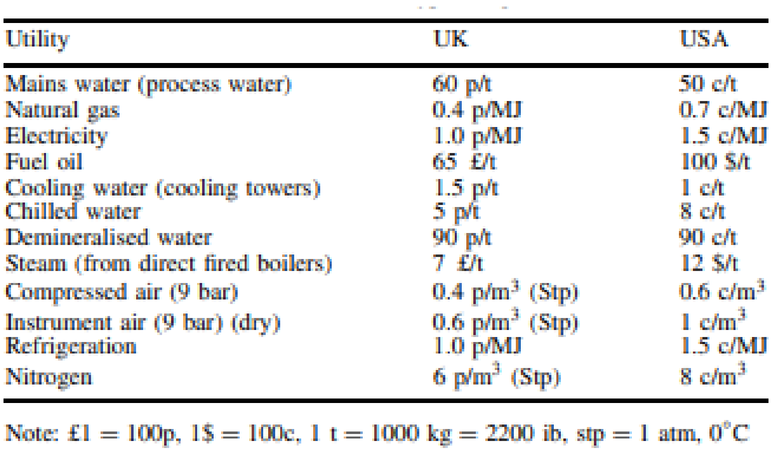


Table 6.6. Summary of production costs

Variable costs	*Typical values*
1. Raw materials	from flow-sheets
2. Miscellaneous materials	10 per cent of item (5)
3. Utilities	from flow-sheet
4. Shipping and packaging	usually negligible
Sub-total A	....
Fixed costs	
5. Maintenance	5–10 per cent of fixed capital
6. Operating labour	from manning estimates
7. Laboratory costs	20–23 per cent of 6
8. Supervision	20 per cent of item (6)
9. Plant overheads	50 per cent of item (6)
10. Capital charges	10 per cent of the fixed capital
11. Insurance	1 per cent of the fixed capital
12. Local taxes	2 per cent of the fixed capital
13. Royalties	1 per cent of the fixed capital
Sub-total B	....
Direct production costs A + B	....
13. Sales expense	20–30 per cent of the direct production cost
14. General overheads	
15. Research and development	
Sub-total C	....
Annual production cost = A + B + C =	....

$$\text{Production cost £/kg} = \frac{\text{Annual production cost}}{\text{Annual production rate}}$$

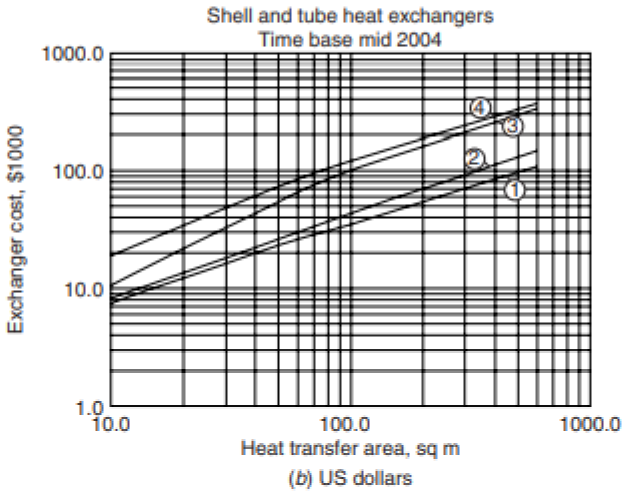
Figure 12 Factorial methods for economic analysis



Utility	UK	USA
Mains water (process water)	60 p/t	50 c/t
Natural gas	0.4 p/MJ	0.7 c/MJ
Electricity	1.0 p/MJ	1.5 c/MJ
Fuel oil	65 £/t	100 $/t
Cooling water (cooling towers)	1.5 p/t	1 c/t
Chilled water	5 p/t	8 c/t
Demineralised water	90 p/t	90 c/t
Steam (from direct fired boilers)	7 £/t	12 $/t
Compressed air (9 bar)	0.4 p/m^3 (Stp)	0.6 c/m^3
Instrument air (9 bar) (dry)	0.6 p/m^3 (Stp)	1 c/m^3
Refrigeration	1.0 p/MJ	1.5 c/MJ
Nitrogen	6 p/m^3 (Stp)	8 c/m^3

Note: £1 = 100p, 1$ = 100c, 1 t = 1000 kg = 2200 ib, stp = 1 atm, 0°C

Figure 13 Cost of utilities., typical figures mid-2004



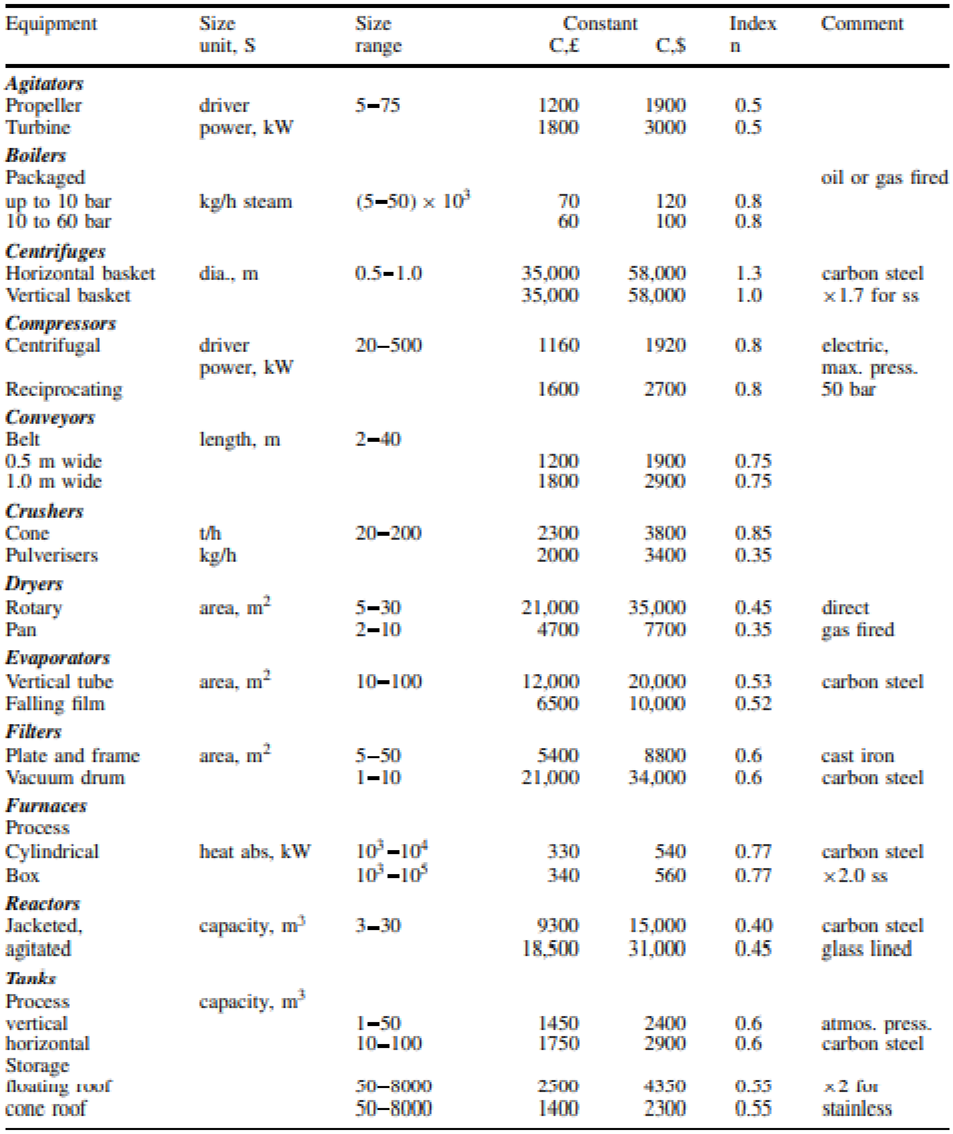
(*b*) US dollars

Materials	
Shell	**Tubes**
① Carbon steel	Carbon steel
② C.S.	Brass
③ C.S.	Stainless steel
④ S.S.	S.S.

Pressure factors	
1–10 bar	× 1.0
10–20	× 1.1
20–30	× 1.25
30–50	× 1.3
50–70	× 1.5

Type factors	
Floating head	× 1.0
Fixed tube sheet	× 0.8
U tube	× 0.85
Kettle	× 1.3

Figure 14 Shell and tube heat exchanger cost time base mid-2004



Equipment	Size unit, S	Size range	Constant		Index n	Comment
			C,£	C,$		
Agitators						
Propeller	driver power, kW	5–75	1200	1900	0.5	
Turbine			1800	3000	0.5	
Boilers						
Packaged						oil or gas fired
up to 10 bar	kg/h steam	$(5–50) \times 10^3$	70	120	0.8	
10 to 60 bar			60	100	0.8	
Centrifuges						
Horizontal basket	dia., m	0.5–1.0	35,000	58,000	1.3	carbon steel
Vertical basket			35,000	58,000	1.0	×1.7 for ss
Compressors						
Centrifugal	driver power, kW	20–500	1160	1920	0.8	electric, max. press. 50 bar
Reciprocating			1600	2700	0.8	
Conveyors						
Belt	length, m	2–40				
0.5 m wide			1200	1900	0.75	
1.0 m wide			1800	2900	0.75	
Crushers						
Cone	t/h	20–200	2300	3800	0.85	
Pulverisers	kg/h		2000	3400	0.35	
Dryers						
Rotary	area, m^2	5–30	21,000	35,000	0.45	direct
Pan		2–10	4700	7700	0.35	gas fired
Evaporators						
Vertical tube	area, m^2	10–100	12,000	20,000	0.53	carbon steel
Falling film			6500	10,000	0.52	
Filters						
Plate and frame	area, m^2	5–50	5400	8800	0.6	cast iron
Vacuum drum		1–10	21,000	34,000	0.6	carbon steel
Furnaces						
Process						
Cylindrical	heat abs, kW	$10^3–10^4$	330	540	0.77	carbon steel
Box		$10^3–10^5$	340	560	0.77	×2.0 ss
Reactors						
Jacketed,	capacity, m^3	3–30	9300	15,000	0.40	carbon steel
agitated			18,500	31,000	0.45	glass lined
Tanks						
Process	capacity, m^3					
vertical		1–50	1450	2400	0.6	atmos. press.
horizontal		10–100	1750	2900	0.6	carbon steel
Storage						
floating roof		50–8000	2500	4350	0.55	×2 for
cone roof		50–8000	1400	2300	0.55	stainless

Figure 15 Purchase cost of miscellaneous equipment, Cost basis mid-2004

REFERENCES

[1] Basu, P. (2010). Biomass gasification and pyrolysis: practical design and theory. Academic press.

[2] Li, J., Tian, Y., Zong, P., Qiao, Y., & Qin, S. (2020). Thermal cracking behavior, products distribution and char/steam gasification kinetics of seawater Spirulina by TG-FTIR and Py-GC/MS. *Renewable Energy*, *145*, 1761-1771.

[3] Jangsawang, W., Gupta, A. K., Kitagawa, K., & Lee, S. C. (2007). High temperature steam and air gasification of non-woody biomass wastes. *Asian Journal on Energy and Environment*, *8* (03), 601-609.

[4] Kim, Y. S., Lee, J. J., Kim, T. S., Sohn, J. L., & Joo, Y. J. (2010). Performance analysis of a syngas-fed gas turbine considering the operating limitations of its components. *Applied Energy*, *87* (5), 1602-1611.

[5] Popovic, M. (2019). Thermodynamic properties of microorganisms: determination and analysis of enthalpy, entropy, and Gibbs free energy of biomass, cells and colonies of 32 microorganism species. *Heliyon*, *5*(6), e01950.

[6] Shakya, R., Adhikari, S., Mahadevan, R., Shanmugam, S. R., Nam, H., & Dempster, T. A. (2017). Influence of biochemical composition during hydrothermal liquefaction of algae on product yields and fuel properties. *Bioresource technology*, *243*, 1112-1120.

[7] Kang, Z., Kim, B. H., Ramanan, R., Choi, J. E., Yang, J. W., Oh, H. M., & Kim, H. S. (2015). A cost analysis of microalgal biomass and biodiesel production in open raceways treating municipal wastewater and under optimum light wavelength. *Journal of microbiology and biotechnology*, *25*(1), 109-118.

[8] Duman, G., Uddin, M. A., & Yanik, J. (2014). Hydrogen production from algal biomass via steam gasification. *Bioresource technology*, *166*, 24-30.

[9] Kim, Y. S., Lee, J. J., Kim, T. S., Sohn, J. L., & Joo, Y. J. (2010). Performance analysis of a syngas-fed gas turbine considering the operating limitations of its components. *Applied Energy*, *87*(5), 1602-1611.

[10] Warnecke, R. (2000). Gasification of biomass: comparison of fixed bed and fluidized bed gasifier. *Biomass and bioenergy*, *18*(6), 489-497.

[11] Vincent, G. M. (2006). Key Combustion Issues Associated with Syngas and High-Hydrogen Fuels.

[12] Todd, D. M. (2000). Demonstrated applicability of hydrogen fuel for gas turbines.

[13] Alam, F., Mobin, S., & Chowdhury, H. (2015). Third generation biofuel from algae. *Procedia Engineering*, *105*, 763-768.

[14] Fatoni, R., Gajjar, S., Gupta, S., Handa, S., & Elkamel, A. (2014, January). Modeling biomass gasification in a fluidized bed reactor. In *Proceedings of the 2014 international conference on industrial engineering and operations management. Bali, Indonesia* (pp. 7-9).

[15] Latif, A. (1999). *A study of the design of fluidized bed reactors for biomass gasification* (Doctoral dissertation, University of London).