TECHNICAL AND ECONOMICAL FEASIBILITY OF PRODUCTION OF ETHANOL FROM SUGAR CANE AND SUGAR CANE BAGASSE

Ç. Efe

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Summary

The primary aim of this study is to investigate and analyze the sugar-ethanol plants operating in Brazil to construct a raw model to gain better understanding and insight about the technical and economical aspects of the currently operating plants. And, the secondary aim is to combine the knowledge gathered from the currently existing plant with the future design idea of converting lignocellulose into ethanol together with sucrose.

The major products of the plants are ethanol and sugar. The proposed plant capacity is 5 million metric tones of sugar cane per year. The plant operation time is 174 days (4176 h) during the harvest season. The plant location is Sao Paulo state of Brazil.

The designed 2005 plant produces 99.88% anhydrous ethanol and white sugar with 165,956 and 379,401 ton yearly capacities respectively. The plant is self sufficient in terms of steam and electricity requirements. The results of economical analysis of the 2005 plant reveal that the required total capital investment is $102 million. The investment is determined using the location factor of 0.4 for Brazil. The net present value of the investment is $166 million using discount rate of 4%. The same analysis reveals that the maximum interest rate that investment can remain economical through out the investment period is 26.5%. The sensitivity analysis showed that the economical values are highly sensitive to sugar cane, ethanol and sugar prices.

The designed 2015 plant produces 99.89% anhydrous ethanol and white sugar with 396,872 and 370,405 ton yearly capacities respectively. For the future plant the microorganisms for fermentation are selected as recombinant *Sacharomyces cerevisiae* that can metabolize pentose and hexose sugars. the pretreatment method for the lignocellulose breakdown is selected as acid catalyzed steam explosion. The cellulose is hydrolyzed using cellulase enzyme. The plant is self sufficient in steam production, however, some portion of electricity demand is required to be purchased from the grid. The environmental impact of the plant is minimized by proposed waste treatment methods.

The investment requirement for 2015 plant is obtained as $133 million after including the location factor of 0.4 for Brazil. The net present value of 2015 plant is obtained as $370 million for discount rate of 4%. The pay put period for the 2015 plant is obtained as 1.4 years. The discounted cash flow analysis reveals that the plant investment can stand interest rates up to 38%. As in 2005 plant the cost of sugar cane and selling price of ethanol and sugar are the major disturbances for the economic parameters.

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List of Symbols

Symbol	Description	SI Units
A	Area	m^2
A_C	Sugar cane processing capacity	ton/h
A_{TF}	Total flow area	m^2
A_{TV}	Vessel cross sectional area	m^2
B	Brix (total dissolved solids) of juice	%
C	Cost of unit capacity (Economy)	$
C_e	Equipment purchase cost	$
C_E	Ethanol concentration	kg/l
C_p	Heat capacity	$kJ\ kg^{-1}{}^{o}C^{-1}$
C_s	Substrate concentration	kg/l
C_x	Biomass concentration	kg/l
D	Diameter	m
D_b	Tube bundle diameter	m
D_{Vmin}	Minimum vessel diameter	m
d_i	Tube inside diameter	m
d_o	Tube outer diameter	m
d_p	Particle diameter	m
f	Fiber fraction of bagasse	
F	Flow rate	m^3/s
F_M	Material factor (Economy)	
F_{VM}	Mass flow rate of vapor	kg/s
F_{VV}	Vapor flow rate of vapor	m^3/s
F_P	Pressure factor (Economy)	
g	Gravitational acceleration	$m\ s^{-2}$
H	Sugar moisture fraction	$W\ m^{-1}\ {}^{o}C^{-1}$
h_b	Boiling heat transfer coefficient	$Wm^{-2}{}^{o}C^{-1}$
h_{id}	Outside tube fouling coefficient	$Wm^{-2}{}^{o}C^{-1}$
h_{od}	Inside tube fouling coefficient	$Wm^{-2}{}^{o}C^{-1}$
h_o	Outside heat transfer coefficient	$Wm^{-2}{}^{o}C^{-1}$
H	Height	m
H_0	Humidity of incoming air	kg/kg
H_1	Humidity of out going air	kg/kg
I	Impurity fraction in bagasse	
K	Thermal conductivity	$Wm^{-1}{}^{o}C^{-1}$
k_e	The ratio of area of tubes to area of the tube bundle	
K_I	Inhibition constant	kg/l
K_s	Substrate affinity	kg/l
$L_{v,B}$	Latent heat of evaporation at boiling temperature	kJ/kg
$L_{v,T}$	Latent heat of evaporation at temperature T	kJ/kg
m_s	Maintenance coefficient for microorganism	$g\ g\ biomass^{-1}h^{-1}$
n	Sizing exponent (Economy)	
N_r	Number of tubes	
P	Pressure	bar
P_c	Critical pressure	bar
P_t	Tube pitch	m
q_E	Specific ethanol production rate	$g\ g\ biomass^{-1}h^{-1}$
q_s	Specific substrate uptake rate	$g\ g\ biomass^{-1}h^{-1}$
q_s^{max}	Maximum specific substrate uptake rate	$g\ g\ biomass^{-1}h^{-1}$
Q	Heat transfer/ heat flow rate	kJ/s

Symbol	Description	SI Units
Q'	Heat flux	$W\ m^{-2}$
r_{dis}	Discount rate	
r_i	Interest rate	
r_{tax}	Tax rate	
s	Sucrose fraction in bagasse	
S	The capacity of concerning equipment (Economy)	Unit of capacity
T	Temperature	ºC
T_c	Critical temperature	ºC
T_b	Normal boiling temperature	ºC
t_m	Mixing time	s
u_t	Water velocity inside the tubes	m/s
U	Overall heat transfer coefficient	$Wm^{-2o}C^{-1}$
u_g	Terminal settling velocity	m/s
V	Volume	m^3
VIS A/B	Coefficients for viscosity calculation	
v_{gs}	Superficial gas velocity	m/s
w	Water fraction in bagasse	
W	Weight	kg/h
y	Year	
Y_{SX}^{Max}	Maximum biomass yield on substrate	g/g
Y_{SX}	Biomass yield on substrate	g/g
Greek	**Description**	**SI Units**
β	Bagasse flow	kg/s
Γ	Condensate flow per unit length of tube	$kg\ s^{-1}\ m^{-1}$
κ	Langmuir constant	g/100g
ΔT_{LM}	Log mean temperature difference	ºC
ΔH	Enthalpy difference	kJ/s
$\Delta\rho$	Density difference	kg/m^3
μ	Specific growth rate	h^{-1}
μ_{max}	Maximum specific growth rate	h^{-1}
η	Viscosity of the fluid	$kg\ m^{-1}\cdot s^{-1}$
Θ	The adsorption capacity of monolayer	g/g solid
ρ	Density	kg/m^3
Σ	Sigma factor	m^{-2}
Subscripts		
L	Liquid	
E	Ethanol	
G	Glucose	
SS	Saturated steam	
X	Biomass	
Xy	Xylose	
V	Vapor	
W	Wall	

List of Abbreviations

Abbreviation	Description	Unit
CGCF	Cumulative gross cash flow	$
CNCF	Cumulative net cash flow	$
COD	Chemical oxygen demand	g O_2/g
DCFROR	Discounted cash flow rate of return	%
EPC	Equipment purchase cost	$
FC	Flow controller	
FCI	Fixed capital investment	$
GCF	Gross cash flow	$
IEC	Installed equipment cost	$
IF	Equipment installation factor	
IRef	Internet References (see references)	
LC	Level controller	
LMTD	Logarithmic mean temperature difference	ºC
NCF	Net cash flow	$
NCV	Net calorific value of bagasse	kJ/kg
NPV	Net present value	$
OHTC	Overall heat transfer coefficient	$Wm^{-2o}C^{-1}$
PC	Pressure controller	
pHC	pH controller	
POP	Pay out period	Year
RC	Ratio controller	
ROR	Rate of return	%
t.c.h.	Ton cane per hour	
TC	Temperature controller	

Definitions

The definitions are obtained from Sugar Engineers' web page (IRef-1)

Ash

Carbonated ash: The residue remaining after incineration at 650°C.

Conductivity ash: The conductivity ash of a product is the figure arrived at by correlating the specific conductance of the solution of that product with its sulphated ash.

Sulphated ash: The residue remaining after incineration at 650°C of a sample which was pre-treated with sulphuric acid.

Bagasse: The residue obtained after crushing cane in a mill is known as bagasse. Depending on the number of the mill it is referred to as first mill bagasse, second mill bagasse, etc. After a diffuser the residue is called diffuser bagasse. The final residue from a milling train or from the dewatering mills of a diffusion plant is called final bagasse or simply, bagasse.

Brix : total dissolved solids in the juice

Bulk Density: The mass of material per unit of total volume occupied. Some useful values are:

Product	Density[kg/m^3]
Shredded cane on the carrier	280
Cut whole stick cane loosely piled	200
Chopped cane from harvesters	350
Piled bagasse	200
Piled raw sugar	880

Extraction : The percentage ratio of sucrose in mixed juice to sucrose in cane. If based on pol it is referred to as Pol Extraction.

Fiber : The water insoluble matter of cane and bagasse from which the brix-free water has been removed by drying.

Imbibition : The process in which water or juice is put on bagasse to mix with and dilute the juice present in the latter. The water so used is termed imbibition water. General terms in use are: single imbibition, double imbibition, compound imbibition, depending on the manner in which the water and / or juice is added.

Insoluble Solids : Insoluble material in mixed juice or press water mud, determined gravimetrically by filtration according to a prescribed method.

Maceration : In the South African sugar industry the term is synonymous with imbibition. The latter is the preferred terminology.

Magma : A mixture of crystals and sugar liquor produced by mechanical means.

Massecuite : The mixture of crystals and mother liquor discharged from a vacuum pan. Massecuites are classified in order of descending purity as first, second, etc., or A, B, etc.

Molasses : The mother liquor separated from a massecuite by mechanical means. It is distinguished by the same prefixes as the massecuites from which it is separated.

Final molasses (B-Molasses): The mother liquor separated from the final massecuite by mechanical means.

Mud : The material removed from the bottom part of the clarifiers. The mud contains the settled insoluble solids.

Purity: The percentage ratio of sucrose (or pol) to the total soluble solids (or brix) in a sugar product.

Sucrose : The pure disaccharide α-D-glucopyranosyl-β-D-fructofuranoside, known commonly as sugar.

Sugar Cane : Botanically a tall grass of the genus *Saccharum* and agriculturally the crop produced from hybrids which are the progeny of a number of *Saccharum* species commonly referred to as cane. Specifically for the determination and distribution of sucrose in cane it is the raw material accepted at the mill for processing.

Syrup: Concentrated clear juice having a brix of between 60° and 70°.

Total Solids : The solids concentration of an aqueous solution determined either by drying or by analysing for the water content using the Karl Fischer method.

Vinasse : The bottom product of beer column which is rich in sugar cane organics and fermentation by products.

1. Introduction

The production of ethanol from crops by means of microorganisms is well known and has been applied by the human kind since the early times of history. However, the utilization of ethanol as a fuel goes back to late 19th and early 20th centuries. In the early phases of automotive industry, ethanol was one of the fuel options for internal combustion engines. But, the government regulations in favor of petroleum industry discouraged the ethanol producers and the fuel ethanol production was disregarded for several decades until the great depression in 1930's. With the increasing fossil fuel prices ethanol was again back stage until the fuel prices got back to low levels.

The real debut of ethanol was in early 1970's when Brazil started the ethanol program to produce ethanol as fuel additive to decrease the amount of petroleum they imported. Since then, Brazil is the major fuel ethanol producer on the world with current ethanol production capacity of 14 million m^3 per annum. The Brazilian ethanol industry is dependent on sugar cane as the raw material with cultivation capacity of 330 million tons per annum. The fuel for about 3 million vehicles running in Brazil is hydrated alcohol; anhydrous alcohol is mixed in a 24% proportion to supply the whole Brazilian fleet of 17 million vehicles. Alcohol is also used intensively in the beverages industry, as well as the chemical, pharmaceutical and cleaning products industries.

However, current worldwide interest in ethanol production is not only due to economical reasons. The exhaustion of the fossil fuel and the increasing greenhouse effect due to the high carbon dioxide emissions urged the nations to search for alternative fuels that can replace the fossil fuels and that can be compatible with petroleum in price. Ethanol with its bio-renewable nature and optimized production technology already proved itself as a replacement for fossil fuels in Brazil. Since the carbon dioxide generated during the combustion of ethanol originates from grains, it doesn't have a positive contribution to the greenhouse effect and it will be metabolized into organic carbon by next generation plants. Moreover, the high octane value of ethanol makes it suitable for lead free gasoline production. However, the cost of production affects the compatibility of ethanol against petroleum. In the earlier times of the ethanol industry in Brazil, the government subsidies played an important role in the survival of the plants

In recent years, there is an increasing interest in utilizing lignocellulosic biomass, which is the most abundant feedstock on the world for ethanol production. Almost 70% of cellulosic biomass can be hydrolyzed into pentose and hexose sugars, which can later on be fermented into ethanol.

In the following sections some information relating to sugar cane, sugar and ethanol is presented to gain some insight to the topic.

1.1. Sugar Cane

Sugar can is a gigantic grass, 2.5-7 m high, with a bamboo-like segmented stalk about 5 cm diameter takes 9-14 months to flower (ripen), with the soft fibers within the stalk containing the sugar. (IRef-11). Sugar cane originated on the island of New Guinea in the South Pacific and was widely used in ancient India. The Arabs brought the plant to medieval Spain and the first Spanish explorers carried it to the continent of America (IRef-10).

Currently, Brazil cultivates approximately 330 million ton sugarcane per annum (See Table 1.1). 25% of the produced sugar is exported. With the current export capacity Brazil is one of the major determining countries in sugar prices.

Table 1.1 The sugarcane production in Brazil (tons per annum) (IRef-9)

YEAR	SÃO PAULO	BRAZIL
2001/02	176,574,250	293,050,543
2002/03	192,486,643	320,650,076
2003/04	207,810,964	359,315,559

Sugarcane is itself an enormously efficient production unit; every ton has an energy potential that is equivalent to 1.2 barrels of petroleum. Brazil is the largest world producer, followed by India and Australia. On average, 55% of Brazilian sugarcane is turned into alcohol and 45% into sugar. Sugarcane is grown in Brazil's Central-South and North-Northeast regions, with two harvest periods. When planted for the first time, sugarcane takes between one year to one year and a half to be ready for harvesting and processing. This same plantation can be harvested up to five times, although significant investments must be made in each cycle to maintain productivity. The sugarcane raw material generates sugar, anhydrous alcohol (a gasoline additive) and hydrated alcohol for the internal and external markets, with different price and demand dynamics. (IRef-9).

1.2. Sugar

At least half of a person's daily energy requirement can be found in nature, in the form of sugars and starches. Due to its long relationship with sugarcane, Brazil became the world's largest sugar producer and exporter, with the lowest production costs, as a consequence of using leading technology and management (IRef-9).

Half of the Brazilian production is exported. White (refined) sugar, crystallized and demerara sugars are exported; Russia has been Brazil's largest customer for at least the last five years. São Paulo produces 60% of all Brazilian sugar and this State accounts for 70% of national exports. The Brazilian annual consumption of sugar is 52 kg per capita (the world average is 22 kg), using sugarcane planted on 2.35 million hectares of land. The annual sugar production of Brazil for last four years is presented in Table 1.2.

Table 1.2 The sugar production in Brazil (tons per anum)(IRef-9)

YEAR	SÃO PAULO	BRAZIL
2001/02	12,350,253	19,218,011
2002/03	14,347,908	22,567,260
2003/04	15,171,854	24,925,793

Various different sugar product are produced from sugar cane which can be named as (IRef-9),

Granulated refined sugar :due to its purity and absence of dyes, moisture or chunks, having well-defined crystals and a homogeneous particle-size distribution, finds considerable use in the pharmaceutical industry, in candy, in exceptional transparent syrups and in dry mixtures, where visual appearance, ease of flow and solubility are important.

Amorphous refined sugar**:** has low color values, good solubility, a fine particle-size distribution and excellent whiteness and is used for domestic consumption, in solid mixtures with instant solubility, cakes and candy, transparent and colorless syrups.

Castor sugar**:** Also known as confectioner's sugar, with very fine crystalline granules, it is produced at the plant and is not refined. It is used by the food industry in pastries, biscuits and drinks.

Inverted syrup **:** Containing 1/3 glucose, 1/3 fructose and 1/3 sucrose, it is an aqueous solution with a high degree of resistance to microbiological contamination and also to crystallization and moisture, that is used in fruit preserves in syrup, ice cream, sweets and toffees, liqueurs, jams, biscuits and carbonated drinks.

Simple syrup or liquid sugar**:** Transparent and clear, it is also an aqueous solution that is used when an absence of color is essential, in the case of clear drinks, sweets, candy and pharmaceutical products.

Organic sugar : This is a product with uniform particle-size distribution, produced with no chemical additives either in the agricultural or in the industrial phase, which can be found in clear and golden versions.

Some properties of crystal sugar is given in Table 1.3.

Table 1.3 Physical properties of crystal sugar

Chemical names	Sucrose, saccharose
Formula	$C_{12}H_{22}O_{11}$
Molar mass	342.3 g/mol
Specific gravity	1.587
Melting point	160 - 186 °C (under decomposition)
Crystal class	monoclinic spenoidal

1.3. Ethanol

Fuel alcohol (ethanol) is a clean and renewable product that contributes to the reduction of the greenhouse effect and lowers air pollution substantially, minimizing its impact on public health. The intensive use of alcohol in Brazil reduces the pollutant emissions, mainly those of carbon monoxide, sulfur oxides, and toxic organic compounds, such as benzene and lead compounds (IRef-9).

Brazil chose fuel alcohol as the alternative to decrease her dependency on external fuel sources and save foreign currency after the oil shock in 70's. A diversification program for the sugar industry was created with large public and private investments supported by the World Bank, allowing expansion of the sugarcane plantation area and construction of alcohol distilleries, either autonomous or attached to existing sugar plants. Today, Brazil produces ~12 million tons/year (about 62% in São Paulo), see Table 1.4 (IRef-9).

In the history of more than 25 years of large-scale use of alcohol, Brazil developed engine technology and product transportation and distribution logistics that are unique in the world. Nowadays, all Brazilian gasoline has a legal alcohol content requirement of 20% to 24%, with a variation of + or – 1%. The actual content is defined by the Inter-ministerial Council for Sugar and Alcohol (CIMA - Conselho Interministerial de Açúcar e Álcool), with the aim of balancing supply and demand.

Large-scale use of alcohol happened in two stages: initially, as a gasoline additive (anhydrous alcohol) at a level of 20%, later 22%. Starting in 1980, alcohol was used as a pure fuel for vehicle engines (hydrated alcohol) that had been adapted from gasoline models, and therefore didn't perform adequately.

Approximately 3 million vehicles are powered by hydrated alcohol, consuming 3.9 million ton/year. Anhydrous alcohol (4.3 million ton/year production) is used in a 22% proportion as an additive to gasoline. Over the last 22 years, savings amounted to 1.8 billion US dollars/year with the replacement by alcohol, equivalent to 200 thousand barrels of gasoline/day(IRef-9).

Table 1.4 The ethanol production in Brazil (tons per annum)(IRef-9)

YEAR	SÃO PAULO	BRAZIL
2001/02	5,629,143	9,101,931
2002/03	6,067,954	9,959,725
2003/04	6,965,571	11,684,068

1.4. Cogeneration of Electrical Power

Steam and heat are important in the process of obtaining sugar and alcohol. Steam, obtained from burning cane bagasse, drives the turbines that generate the electrical power, which makes the industrial units self-sufficient and able to sell the excess to concession holders.

In the State of São Paulo, the industry generates between 1,200 and 1,500 MW for its own consumption, 40 power plants produce surpluses of 158 megawatts and the energy derived from sugarcane already helps to illuminate several cities. The sugarcane agro-industry's potential for generating energy is in the region of 12,000 MW, the total potential installed in Brazil is 70,000 MW. In 2002, as a result of new projects, a further 300 MW should be added and, short-term, the industry could contribute an additional 4000 MW.

1.5. Properties of Ethanol

Ethanol or ethyl alcohol has been described as one of the most exotic synthetic oxygen-containing organic chemicals because of its unique combination of properties as a solvent, a germicide, a beverage, an antifreeze, a fuel, a depressant, and especially because of its versatility as a chemical intermediate for other organic chemicals (IRef-8).

Ethanol under ordinary condition is a volatile, flammable, clear, colorless liquid. Its odor is pleasant, familiar, and characteristic, as is its taste when it is suitably diluted with water. The physical and chemical properties of ethanol are primarily dependent upon the hydroxyl group. This group imparts polarity to the molecule and also gives rise to intermolecular hydrogen bonding. In the liquid state, hydrogen bonds are formed by the attraction of the hydroxyl hydrogen of one molecule and the hydroxyl oxygen of a second molecule. The effect of this bonding is to make liquid alcohol behave as though it were largely dimerized. This behavior is analogous to that of water, which however, is more strongly bonded and appears to exist in liquid clusters of more than two molecules (IRef-8). Some important physical properties of ethanol is given in Table 1.5.

Table 1.5 The important physical properties of ethanol (IRef-8)

Property	**Value**
Normal boiling point (°C)	78.32
Critical temperature, (°C)	243.1
Density (20 ºC, g/ml)	0.79
Heat of combustion at (25°C, J/g)	29,676.7
Autoignition temperature, (°C)	793.0
Flammable limits in air	
Lower, (vol%)	4.3
Upper, (vol%)	19.0

Fermentation processes from any material that contains sugar can derive ethanol. The many and varied raw materials used in the manufacture of ethanol via fermentation are conveniently classified under three types of agricultural raw materials: sugar, starches, and cellulose materials. Sugars (from sugar cane, sugar beets, molasses, and fruits) can be converted to ethanol directly. Starches (from grains, potatoes, root crops) must first be hydrolyzed to fermentable sugars by the action of enzymes from malt or molds. Cellulose from wood, agricultural residues, waste sulfite liquor from pulp and paper mills) must likewise be converted to sugars enzymatically or chemically generally by the action of mineral acids. Once simple sugars are formed, enzymes from yeast can readily ferment them to ethanol (IRef-8).

1.6. Scope of the Design

The scope of this project is to apply the current knowledge in sugar-ethanol industry to investigate the feasibility of sugar and ethanol production from sugar cane, to search for alternative usages for the side products of the sugar production and to improve the existing technology to get more sustainable and environmentally friendly processes.

The project is composed of two parts. In first part, sugar ethanol plant of current technology (year 2005) is investigated. In conventional plant (2005), 52% of the sucrose extracted from sugar cane is crystallized into sugar and the remaining is fermented into ethanol. Bagasse, the fiber remaining after the extraction, is burned to generate electricity. The vinasse ,which is the organics and salts rich stillage from distillation, the filter mud, which is rich in coagulated proteins, gums, organics, sulphate and phosphate salts, and the cogenerator ash, which is mainly composed of the mineral portion of the bagasse, is sent back to the fields to be used as fertilizer for the cane plantation.

The proposed 2015 plant is generated based on the current technology. However, in future case, the cellulose and hemicellulose fraction of lignocellulose is hydrolyzed into hexose and pentose sugars to be used as substrate for ethanol production. Also, some environmental concerns and sustainability of the plant is progressed. For example, vinasse, filter cake and the genetically modified yeast are combusted in cogeneration unit and streams rich in organic materials is treated in waste water treatment plant.

The proposed plant location is Sao Paulo state of Brazil. In Brazil, there are 324 existing sugar mills. The ten leading plants process 3.6 to 6.8 million tons of sugar cane per year. Using these plant capacities and the projected future world ethanol production capacities, the design capacity of the plant can be estimated. The future projection of world total ethanol production is given in Figure 1.1. The plant is proposed to have a capacity within the range of the top ten largest plants and the plant size is proposed to be sugar cane processing capacity of 5 million tons/year.

The ethanol processing plant yields yeast, carbon dioxide, vinasse, filter cake and bagasse as side products. Since CO_2 is bio-renewable and coming from plants, it is just released to atmosphere and it does not have positive contribution to green house effect. Using the existing technologies, the yeast output of the plant is minimized and the produced yeast is combusted in boiler house. Existing plants utilizes bagasse as fuel in electricity and steam generation to be used in the process as heating and electricity source.

In conventional plant, the main input streams to the plant are sugar cane and recycle water as main raw materials, limestone, and sulphite in the pretreatment of juice for production of sugar and sulphuric acid for the sterilization of the yeast recycle. In future case, in addition to the inputs listed for conventional plant, acidifying agents for pre-treatment and cellulase for cellulose hydrolysis are input streams to the process.

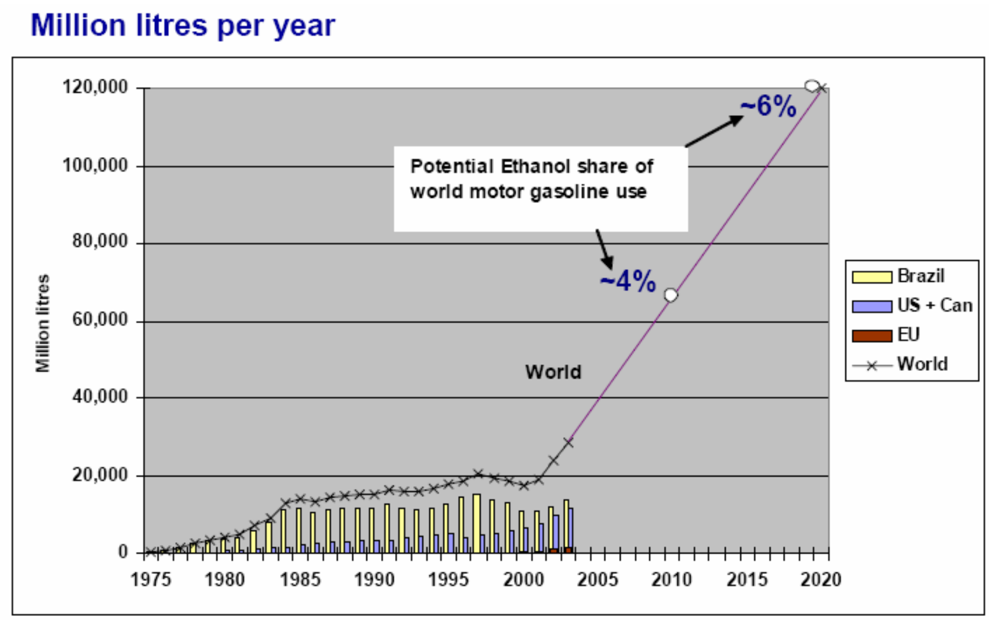


Figure 1.1 Ethanol production projection (Source International Energy Agency)

1.7. Additional Information on Report

For the clarity of the report, all values given in this report are in metric units. Unless, specified otherwise the compositions are mole percentages. The terminology relating to sugar production is described in descriptions section.

The 2005 plant design is carried by mimicking the current existing plant, the alternatives for 2015 plant can be found in Section 2. The revised basis of design is presented in Section 3. In this chapter, reader can find information relating to basic assumption, plant capacity and location, composition and flow rate information of incoming and out going streams, block schemes and pure component properties. In Section 4, the thermodynamic property methods and data relating to thermodynamics of the system is presented. The description of process and the flow scheme of the plants can be found in Section 5. In Section 6, the 2015 plant is analyzed to determine the process controls required. The mass and energy balances are presented in Section 7. The equipment designs are described in Sections 8 and 9. Section 10 is on safety and HAZOP analysis. The discussion on wastes generated in the plant is given in Section 11. The economical evaluations of the plants are presented in Sections 12 and 13. The discussions, conclusions and future recommendations on design can be found in Sections 14 and 15.

2. Process Options and Selection

In this chapter the process alternatives for 2015 plant are compared and the best one is selected. Since the 2005 plant is a reproduction of existing plants, it is not analyzed in this chapter. The process descriptions for the 2005 and 2015 plants can be found in Chapter 5.

2.1. Alternatives for Pretreatment of Bagasse

The degradation of the lignocellulosic materials requires two steps; the first step is the pretreatment for the solubilization of the hemicellulose and releasing the lignin and second step is the hydrolysis of cellulose.

The purpose of the pretreatment is to remove lignin and hemicellulose, reduce cellulose crystallinity, and increase the porosity and contact area of the materials to let the cellulase enzymes gain access to the cellulose molecules (Sun and Cheng, 2002). During the pretreatment of lignocellulosic material the hemicellulose fraction is also hydrolyzed. Hemicellulose monomers are a mixture of hexose sugars and substantial amounts of pentose sugars.

There are several different methods proposed by the researchers and engineering companies. Physical, physical-chemical, chemical and biological treatments have been proposed for the pretreatment of lignocellulosic materials. In following sections, different pretreatment alternatives will be discussed.

2.1.1. Mechanical Treatment

The crystallinity of cellulose can be reduced by chipping, grinding and milling. The size of the materials is usually 10-30 mm after chipping and 0.2-2 mm after milling or grinding. The power requirement is dependent on the final size of the particles. Conventional mechanical methods require 70% more energy than steam explosion to achieve the same size reduction (Holtzapple et al., 1989). Energy requirement for soft wood is reported to be in the range of 3.2-25 kWh/ton for 1.6-9.5 mm final particle sizes (Sun and Cheng, 2002). Mechanical treatment is beneficial for the size reduction, but, still further hydrolysis steps are required to hydrolyse hemicelluloses and remove lignin. Therefore, mechanical treatment can only be used as a part of pretreatment and considering that the bagasse is the out put of the mill tandems, mechanical treatment is not necessary for the current design.

2.1.2. Steam Explosion

Steam explosion refers to a treatment technology in which lignified biomass is exposed to high pressure steam followed by sudden (explosive) decompression. This is recognized as a hydrolytic pretreatment prior to biomass saccharification for ethanol production (Glasser and Wright, 1998). This pressure reduction triggers the explosive decomposition of the materials. Typical process conditions for steam explosion is 160-260°C (0.69-4.83 MPa) for several seconds or minutes before the material is exposed to atmospheric pressure. Optimal hemicellulose solubilization and hydrolysis can be achieved by high temperature and short residence times (270°C, 1 min) (Duff and Murray, 1996) or long residence times lower temperatures (190°C, 10 min) (Wright, 1998).

Steam explosion by itself suffers from low xylose yields. On the other hand, for acid catalyzed steam explosion, xylose yield of as high as 90% is reported (Wyman, 1996). Addition of H_2SO_4, SO_2 or CO_2 in steam explosion improves enzymatic hydrolysis, decrease the production of inhibitory compounds and lead to more complete removal of hemicellulose (Morjanoff and Gray, 1987). The optimal conditions of steam explosion pretreatment of sugarcane bagasse is reported as 220°C; 30 s residence time, water to solid ratio 2 and 1% H_2SO_4 (w/w dry bagasse). The sugar yield was 65 g/100 g treated bagasse. Released sugar comprised 38.9 g glucose, 0.6 g cellobiose, 22.1 g xylose and 3,5 g arabinose and 0.21 g furfural which is equivalent to 83% of the and hydroglucan and 84% of the anhydroxylan content of the untreated bagasse.

Steam Pretreatment at 200-210 °C with the addition of 1% SO_2 (w/w dry wheat straw) gave 95% glucose and 62% xylose yield based on the glycan and xylan available (Saha, 2003).

Steam explosion has some limitations such as, destruction of a portion of xylan fraction, incomplete disruption of lignin-carbohydrate matrix and generation of side products that may be inhibitory to the down stream fermentation processes (Sun and Cheng, 2002).

2.1.3. Ammonia Fiber Explosion (AFEX)

In this method, the lignocellulosic material is exposed to liquid ammonia at high temperature and pressure for a period of time and then the pressure is immediately reduced. In a typical AFEX process, the dosage of liquid ammonia is 1-2 kg ammonia/ kg dry biomass, temperature 90 °C, residence time is 30 min and moisture content is 15%. This method is reported to be successful for sugarcane bagasse. Over 90 % hydrolysis of cellulose and hemicellulose has been obtained after AFEX pretreatment of bagasse, however the hydrolysis of hemicelluloses are incomplete and the monomeric pentose sugar yield of the process is low. Over 99% recovery of ammonia is easily achieved in a proceeding recovery process (Sun Y. and J. Cheng, 2002).

AFEX does not significantly solubilize hemicellulose to its monomers as efficient as the acid pretreatment and acid catalyzed steam explosion (Mes-hartre *et al.*, 1988). On the other hand, the ammonia pretreatment does not require small particle sizes and does not produce inhibitors for the downstream biological processes, so intensive detoxification is not required. AFEX (Mes-hartre *et al.*, 1988).

2.1.4. CO_2 Explosion

The main advantage of acid catalyzed pretreatment methods is the hydrolysis of the hemicellulose with the presence of acid. Carbonic acid, which is formed when carbon dioxide is dissociated in water, can be used as an acid catalyst in pretreatment step. By using carbon dioxide, the major draw back of sulfuric acid that is neutralization after treatment can be eliminated and the reaction effluent can be neutralized by releasing the pressure in the reactor. On the other hand, the lowest pH that can be obtained by carbon dioxide is pH 3.1 at 50 bar (Hofland *et al*, 2000) which is not comparable with the pH that can be achieved with mineral acids. This level of acidity can already be achieved with the natural acidity of the hydrolyzate arising from the acetic acid released during the pretreatment (personal communication Wim de Laat, Nedalco).

CO_2 explosion has a similar structure with steam and ammonia explosion techniques. CO_2 forms carbonic acid and increases the hydrolysis rate. The yields are relatively low compared to steam and ammonia explosion treatment (75%). CO_2 explosion is more cost effective then ammonia explosion and does not cause formation of inhibitory compounds for the downstream biological processes (Sun Y. and J. Cheng, 2002).

2.1.5. Wet Oxidation

Wet oxidation is used for the oxidation of the organic material rich solutions. High pressure and temperatures are applied in the presence of oxygen to oxidize the organic compounds in feed stream to low molecular weight organic compounds carbon dioxide and water. In their study, Lissens et al (2004) showed that alkaline wet oxidation (T = 185 $^{\circ}$C, 12 bar O_2 pressure, 2 g l^{-1} Na_2CO_3 and 15 min) can be used as a pretreatment for efficient fermentation of high-lignin woody yard waste into bioethanol. The combined effect of high oxygen pressure (12 bars) and alkaline conditions (3.3 g Na_2CO_3 per 100 g dry matter waste) were the most decisive parameters for enhanced enzymatic carbohydrate degradation of wet oxidized high-lignin wood. The combination of alkali treatment and wet oxidation does not generate hydroxymethyl furfural and furfural.

The energy yield during the wet oxidation was 4,100 kJ/kg straw, when cellulose (11% w/w), hemicellulose (45% w/w), and lignin (74% w/w) are oxidized to CO_2, formic acid, acetic acid, and phenolic compounds (Thygesen *et al*.,2003).

Lissens *et. al* (2004) have measured low-molecular weight carboxylic acids (62 g formic acid/kg dry straw and 23 g acetic acid/kg dry straw) and phenolic compounds (40 g/kg straw) as being the main degradation products

2.1.6. Ozonolysis

The efficiency of ozonolysis treatment is dependent on the lignin percentage of the feedstock. The degradation is limited to lignin. Hemicellulose is slightly affected and cellulose is hardly affected. The advantage of this method is effective removal of lignin, less toxic material production and being carried out at room temperature and pressure. The large amount of ozone required increases the cost of the process. Beh-Ghedalia and Miron, (1981) reported the use of 20% (w/w dry mater) of ozone for the ozonolysis and 40% moisture substrate, whereas, Neely (1984) reported the use of 4-6% ozone and stated that for 33% moisture in lignocellulose the cellulose digestibility is 100% while for 50% moisture it is 80%. An enzymatic hydrolysis yield of 57% was obtained after ozonolysis treatment.

2.1.7. Acid Hydrolysis

Concentrated acids such as HCl or H_2SO_4 have been used to treat lignocellulosic materials. Dilute acid pretreatment processes has been successfully developed for lignocellulosic materials. There are two types of dilute acid pretreatment processes high temperature (greater than 160 °C), continuous flow processes for low solid loading (5-10% w/w reaction mixture) and low temperature (less than 160 °C), batch processes for high solid loadings (10-40%). 0.5-2% sulfuric acid is added to the feedstock and the mixture is heated to 140-160 °C for 5-20 min and most of the hemicellulose is hydrolyzed. The cost is higher compared to steam and ammonia explosion pretreatments, and, requires pH neutralization for downstream biological process steps (Sun and Cheng, 2002).

During acid hydrolysis the pentose sugars can degrade to furfural and the hexose sugars can degrade to hydroxylmthylfurfural. Acetic acid is also released during the hydrolysis of the hemicellulose. Large variety of lignin degradation products (aromatic, polyaromatic, phenolic and aldehylic) are released from lignin acid treatment (Mussatto *et al.,* 2004). The generation of these chemical can be inhibitory for the downstream biological process steps. Therefore, a detoxification step is required after acid hydrolysis.

Neureiter and his coworkers (2002), carried out dilute acid run at 170 °C with 25 mM sulfuric acid, 4% solid concentration and 10 minute retention time and obtained 93% xylan monomers recovery with 0.06, 0.44, 0.93 g/100g dry matter of hydroxymethy furfural, furfural and acetic acid respectively.

Chong and coworkers (2004), compared the use of nitric acid, hydrochloric acid and sulfuric acid in dilute acid treatment. They concluded that nitric acid and sulfuric acid has close yields where as hydrochloric acid gave lower yields. In this study the optimal operating condition was found as 122 °C with 6% (w/w dry matter), 10:1 liquor to solid ratio and 9.3 minute retention time

Peracetic acid usage was also reported for the pretreatment of sugarcane bagasse. Teixeira *et al.*(1999) reported that 21% peracetic acid concentration based on oven dried dry biomass gave the best results for the pretreatment of bagasse.

Robinson and his coworkers (2004) fractionated biomass carbohydrates very cleanly (99%) from unreacted insoluble lignin by a dilute acid hydrolysis (0.35–1.5% H_3PO_4), wherein the incipient aldoses are intercepted by catalytic hydrogenation (Ru/C) to produce a solution of C_5–C_6 polyols (xylitol, sorbitol) and anhydro polyols (sorbitan) in which no phenols could be detected (100 ppm).

2.1.8. Alkaline Hydrolysis

Almost complete enzymic degradation to glucose of the cellulose fraction is possible provided that first lignin and a considerable part of hemicellulose are removed. The material so obtained is readily hydrolyzed by a commercial mixture of cellulase and hemicellulase, giving a concentrated sugar solution composed for 80% of glucose and the remaining 20% of xylose, as the only products.

Six-gram samples of the dry material are suspended in 100 ml of 1% NaOH and kept under stirring for times ranging from 6 to 24 h. After this, hydrogen peroxide is added to the liquor. The residues were collected and washed with water, then with 10 mM acetic acid and finally washed thoroughly with water until neutral pH (Curreli et al., 2002).

Sun and co-workers (2004) sequentially extracted over 90% of the original hemicelluloses in the cell walls of bagasse with distilled water, 0.5 M NaOH, 0.5, 1.0, 1.5, 2.0 and 3.0% H_2O_2 at pH 11.5, and 2.0 M NaOH at 55 °C for 2 h. Meanwhile, the successive treatments also released 89.0% of the original lignin

Ammonia is also used for pretreatment to remove lignin (170°C 2.5-20% ammonia, 1h residence time) 65-85% delignification for switch grass. No hemicellulose hydrolysis is reported (Sun and Cheng, 2002).

2.1.9. Organosolv

In this method, an organic or aqueous organic solvent mixture with inorganic acid catalysis (HCl or H_2SO_4) is used to break the internal lignin and hemicellulose bonds. The organic solvents can be methanol, ethanol, acetone, ethylene glycol, triethylene glycol and tetrahydrolfurfuryl alcohol (Chum *et al.*, 1988). Organic acids such as oxalic, acetylsalicylic and salicylic acid can also be used as catalysis for organosolv processes (Sun and Cheng, 2002). At high temperatures (above 185°C), the addition of catalysis is unnecessary. Usually a high yield of xylose can be obtained by the addition of the acid. The solvents need to be recycled to reduce the cost. Delignification of the lignocellulose also occurs when organosolv treatment is used.

Recently, Copersugar (Brazil) developed a new organosolv process for the treatment of lignocelluloses. In this technique, bagasse is added to ethanol water mixture to a final ratio of 80/20 ethanol/water (including the water coming from bagasse) and 25 grams of H_2SO_4 is added per liter ethanol produced. The mixture is heated to 100-180°C at 25 bar pressure.

Chum (1988) and his coworkers reported 99.4% lignin removal, 100% xylan removal and 91% hexose recovery for 10:1 wood to liquor ratio, 0.5 M H_2SO_4, 70% methanol, 170 ºC and 1hr retention time.

Papatheofanous and coworkers (1995) fractionated wheat straw into cellulosic fibers, hemicellulose sugars and solid lignin oligomers through a two-stage, acid-catalyzed process. During the first stage, raw material was treated with dilute H_2SO_4 (0.25-1.25 mol/l) at reflux temperature (99.5±1.0°C) for 10-60 min. Subsequently, delignification was conducted with H_2SO_4 (1 mol/l) in aqueous ethanol (62.5-87.5% Ethanol v/v) at reflux temperature (81.0±2.0°C) for 90 min. Selective hydrolysis of about 50% of the straw hemicelluloses (w/w on original straw hemicelluloses), converted to water-soluble oligo- and monosaccharides (first stage), followed by delignification in 87.5% v/v Ethanol (second stage), led to optimal component fractionation efficiency with negligible cellulose loss (less than 2% w/w on original straw cellulose) and high lignin removal (more than 70% w/w on original straw lignin).

2.1.10. Biological Pretreatment

In biological pretreatment, microorganisms such as brown- white- and soft rod fungi are used to degrade lignin and hemicellulose in waste materials. Brown rod mainly attacks cellulose, while white and soft rod fungi attack both cellulose and lignin at low rates (35 percent hydrolysis in 15 days –4weeks). In order to prevent loss of cellulose, cellulase less mutants of the microorganisms are developed (Sun and Cheng, 2002). Under current conditions biological pretreatment is neither technologically nor economically feasible.

2.1.11. Selection of Pretreatment Method

Different process alternatives for the production of ethanol from bagasse involve the pretreatment, detoxification, delignification, cellulose hydrolysis and the fermentation of the pentose and hexose sugars in to ethanol. Since the cellulose is enzymaticaly hydrolyzed, and the cellulase requirement for the hydrolysis of certain amount of cellulose is fixed, the cost contribution of cellulase into production cost is approximately same for each alternative. Therefore, for the current section the major cost factor in rough economical evaluation for process selection is based on pretreatment, detoxification and delignification methods.

Based on the literature data, brief economical calculations are carried out by using the input and output streams of the system. The assumptions made during the calculations are as given in Section 2.1.11.1. Since this is just for comparing the different alternatives, the prices are not updated to current prices. In later economical evaluations, last updated prices will be used.

2.1.11.1. Assumptions

- Bagasse composition is taken as (%)

Glucan	Xylan	Galactan	Arabian	Lignin	Mannan	Extractives	Ash
41.20	21.50	0.9	1.90	25.20	0.60	4.70	4.0

- Ethanol yield on both pentose and hexoses is 91.4%.
- Water to bagasse ratio of incoming bagasse is 1:1
- Process water price is 0.153 Euro/ton
- Natural gas price is 0.0053 Euro/MJ
- Sulphuric acid price is 48.8 Euro/ton
- Sulphur dioxide price is 240 euro/ton

- Ammonium price is 138 Euro/ton
- Oxygen price is 40 Euro/ton
- Na_2CO_3 price is 445 Euro/ton
- NaOH price is 170 Euro/ton
- H_2O_2 price is 610 Euro/ton
- $Ca(OH)_2$ price is taken as 65 Euro/ton
- Solid disposal for gypsum 75 euro/ton
- 80% boiler efficiency
- Density of the Initial Bagasse mixture is 1 kg/l
- 1 kg/min bagasse inflow

The utility and raw material prices are taken from Chemical Market Report (2003) and Sinnot (1996). During the calculations, it is assumed that the processes other than the pretreatment method are same for all the alternatives. i.e. the hemicellulose hydrolysates and cellulose obtained from each pretreatment method follows the same downstream treatment to the end of fermentation. The cost calculations are based on the input and output streams of the different alternatives and does not include the cost for ethanol purification.

Since the cellulase requirement for products of each pretreatment method is assumed to be same, the nutrients required for cellulase production also aren't included in the economic comparisons.

To have some idea, the vessel volumes are also calculated based on the residence times. The material factors are taken 1.1 for corrosive media and 1.2 for high pressure media up to 5 MPa. Based on this information the cost calculations are as follows,

2.1.11.2. Steam Explosion

Acid Catalyzed

All the values are per kg of ethanol

	Requirement		**Contribution to Cost**	
Water	3.890	kg	0.0006	Euro
Sulphuric acid	0.039	kg	0.0019	Euro
$Ca(OH)_2$	0.029	kg	0.0019	Euro (neutralization)
Solid Disposal	0.059	kg	0.0044	Euro (gypsum)
Heat	11955	kJ	0.0633	Euro
Total			**0.0721**	**Euro**

Ethanol yield 0.257 kg ethanol/kg bagasse

SO_2 Catalyzed

	Requirement		**Contribution to Cost**	
Water	3.910	kg	0.0006	Euro
Sulphur dioxide	0.039	kg	0.0090	Euro
$Ca(OH)_2$	0.029	kg	0.0019	Euro (neutralization)
Solid Disposal	0.059	kg	0.0044	Euro (gypsum)
Heat	12011	kJ	0.0637	Euro
Total			**0.0796**	**Euro**

Ethanol yield 0.255 kg ethanol/ kg bagasse

Volume factor	3.01 liters/ min
Residence time	30 s
Vessel volume	1.5 liters
Steam recovery flash vessel	
Material factor	1.1

In steam explosion method, low pressure steam is obtained as by product and can be utilized as heat source in other process units. The heat required includes the energy required to heat the inflow to the process temperature.

2.1.11.3. Ammonia Fiber Explosion

All the values are per kg of ethanol

	Requirement		Contribution to Cost		
Ammonia	7.132	kg	0.010	Euro	(99% recovery)
Heat	17648	kJ	0.094	Euro	(Including the drying and recovery of ammonia)
Total			**0.104**	**Euro**	

Ethanol yield	0.28 kg ethanol/kg bagasse
Volume Factor	3.15 liter/ min
Residence time	30 min
Vessel volume	94.5 litres
Material Factor	1.1

Ammonia recovery is required, no need for detoxification

The recovery cost of ammonia is calculated as the amount of heat required for the phase change of ammonia and include in heat requirement. The AFEX treatment does not generate toxic materials. On the other hand, the AFEX products are not sugar monomers high portion is sugar oligomers.

2.1.11.4. Carbon Dioxide Explosion

The chemical cost is parallel to steam explosion with SO_2 and sulphuric acid. Since the CO_2 is by-product of fermentation the cost of carbon dioxide is zero. But, the compression of CO_2 to pressures up to 10 bar requires 243 kJ/ kg CO_2. That will add the cost significantly. As mentioned in previous sections, the most important obstacle of carbon dioxide explosion is the low acidity and consequently low conversions.

2.1.11.5. Wet Oxidation

All the values are per kg of ethanol

	Requirement		Contribution to Cost	
Water	66.33	kg	0.010	Euro
Na_2CO_3	0.140	kg	0.062	Euro
Oxygen	2.120	kg	0.085	Euro
Compression	708	kJ	0.008	Euro
		Total	**0.165**	**Euro**

Ethanol yield	0.236 kg ethanol/kg bagasse
Volume Factor	17.66 liter/ min
Residence time	15 min
Vessel volume	264.9 liters
Material Factor	1.2
Heat release	17372 kJ/kg ethanol
Detoxification	0.0974 kg acetic acid, 0.267 kg formic acid per kg ethanol. Detoxification is not included in the cost calculations

If air is used as the oxygen source the oxygen cost will be excluded from the cost. The compression energy requirement will increase.

2.1.11.6. Ozonolysis

Not enough information.

2.1.11.7. Acid Hydrolysis

All the values are per kg of ethanol

	Requirement		**Contribution to Cost**	
Water	10.47	kg	0.0016	Euro
H_2SO_4	0.215	kg	0.0105	Euro
Heat	13218	kJ	0.0700	Euro
$Ca(OH)_2$	0.162	kg	0.0111	Euro (neutralization)
Solid waste	0.298	kg	0.023	Euro ($CaSO_4$ disposal cost)
Total			**0.116**	**Euro**

Ethanol yield	0.287 kg ethanol/kg bagasse
Volume Factor	5.06 litre/ min
Residence time	10 min
Vessel volume	50.06 litres
Material Factor	1.1
Detoxification	0.002 kg HMF, 0.015 kg furfural and 0.033 kg acetic acid per kg ethanol. Detoxification is not included in the cost calculations.

2.1.11.8. Alkaline Hydrolysis

All the values are per kg of ethanol

	Requirement		**Contribution to Cost**	
Water	57.69	kg	0.009	Euro
H_2O_2	0.108	kg	0.038	Euro
NaOH	0.223	kg	0.066	Euro
Sulphuric acid	0.273	kg	0.0013	Euro (neutralization)
Heat	10173	kg	0.054	Euro
Total			**0.1683**	**Euro**

Ethanol yield	0.277 kg ethanol/kg bagasse
Volume Factor	18 liter/min
Residence time	120 min
Vessel volume	2160 litres
Material Factor	1.1

2.1.11.9. Organosolv

All the values are per kg of ethanol

	Requirement		**Contribution to Cost**	
Water	3.57	kg	0.001	Euro
H_2SO_4	0.025	kg	0.015	Euro
EtOH	28.50	kg	Recovery cost as heat	
Heat	30414	kJ	0.161	Euro
$Ca(OH)_2$	0.019	kg	0.0012	Euro (neutralization)
Gypsum	0.035	kg	0.0026	Euro (solid disposal)
Total			**0.182**	**Euro**

Ethanol yield	0.280 kg ethanol/kg bagasse
Volume Factor	11 liter/unit time (min)
Residence time	60 min
Vessel volume	660 liters
Material Factor	1.1

The recovery cost of ethanol is included in the heat required by calculating the energy required to evaporate given amount of ethanol.

2.1.12. Evaluation

Based on the preliminary economical evaluation given in previous sections, the acid catalyzed steam explosion techniques are the most economical ones with the less cost contribution to overall price. But, since economics is not the only factor for determination, a list of selection criteria is prepared as given in Table 2-1.

Based on these criteria, the steam explosion techniques came out to be the most feasible ones. On the other hand, in terms of sustainability the sulphuric acid and sulphur dioxide catalyzed pre-treatment methods have the disadvantage of neutralization and disposal requirements. Unlike sulphuric acid and sulphite catalyzed pretreatments, carbon dioxide catalysed steam explosion is easier to operate because carbon dioxide can be neutralized just by releasing the pressure.

But, the acidity that can be obtained by carbon dioxide can already be achieved by acetic acid released from the pre-treatment. Therefore, carbon dioxide steam explosion technique will not meet the demand of pre-treatment step

On the other hand, the operation pressures of steam explosion techniques cause operational problems when fibers involved. Although AFEX and alkaline hydrolysis techniques have relatively low pressures, the high amount of ammonia requirement of AFEX and high operational costs of alkaline hydrolysis makes them unfavorable as a pretreatment method. Therefore, under current technological conditions, the high pressures will be involved in the process.

The organosolv process, when compared to steam explosion techniques, is a favorable alternative. The high operational cost is related to the recovery of the ethanol. The costs can be reduced by heat integration. However, considering that the majority of the cost of steam explosion techniques are also due to the high energy requirement and can be recovered by heat integration, organosolv still can not have cost as low as steam explosions. Moreover, organosolv process also involves the use of acids and requires neutralization and disposal. The most important factor making organosolv unfavorable is the amount of ethanol circulated in the system. To produce 1 kg of ethanol 28 kg of ethanol is circulated in the system which means the process losses during the circulation will effect the overall process yields and cause a reduction in productivity.

Table 2.1 Selection of pretreatment method

Criteria	ω	**Mechanical**	**Steam Explosion**			**AFEX**	**Wet Oxidation**	**Ozonolysis**	**Acid Hydrolysis**	**Alkaline Hydrolysis**	**Organosolv**
			SO_2	**CO_2**	**H_2SO_4**						
Good optimizations	0.10	0.4	0.8	0.5	0.8	0.8	0.65	0.4	0.8	0.7	0.6
Flexibility , robustness	0.15	0.5	0.8	0.6	0.8	0.6	0.8	0.6	0.8	0.6	0.8
Sustainability and environment	0.15	0.8	0.5	0.8	0.5	0.5	0.6	0.5	0.5	0.5	0.5
Operability	0.10	0.4	0.7	0.7	0.7	0.8	0.8	0.5	0.9	0.7	0.7
Clients' satisfaction	0.12	0.4	0.7	0.9	0.6	0.5	0.7	0.5	0.6	0.6	0.7
Economics	0.18	0.4	0.9	0.8	0.85	0.7	0.3	0.5	0.65	0.5	0.5
Safety	0.10	1.0	0.7	0.8	0.7	0.5	0.6	0.5	0.7	0.7	0.7
Use of best available techniques	0.10	0.6	0.8	0.8	0.8	0.8	0.8	0.6	0.8	0.8	0.8
Total	1.00	0.555	**0.741**	**0.742**	**0.72**	0.641	0.633	0.515	0.704	0.617	0.649

Therefore, the current situation makes steam explosion techniques the more favorable option. Only problem with steam explosion techniques is the presence of acidifying agent, which is supposed to be neutralized. It is already known that steam explosion is self catalyzing method due to the presence of acetic acid. Therefore, by increasing the residence times and lowering the acidifying agent fraction in the pre-treatment media the mineral acid addition can be minimized.

2.2. Detoxification

When hemicellulose is degraded xylose, mannose, galactose, arabianose, glucose and acetic acid are liberated. At high temperature and pressure, xylose is further degraded to furfural and hexose sugars are degraded in to hydroxymethyl furfural. Also, hydroxymethyl furfural can be degraded into levulinic acid (Palmqvist and Hahn-Hagerdal, 2000b).

The inhibitors for fermentation can be listed as acetic acid and extractives from pretreatment i.e. terpenes, alcohols and tannins, sugar degradation products, i.e. furfural, hydroxymethyl furfural, levulinic acid, formic acid and humic substances, lignin degradation products, i.e. aromatic and polyaromatic compounds and fermentation products, i.e. ethanol, acetic acid, glycerol, lactic acid.

The need for detoxification is reported to decrease if high initial cell concentrations are used. The reason for this is the inhibitory action of the toxic materials is mainly on growth of the microorganism, not on the ethanol productivity.

Biological, physical, and chemical detoxification methods have been proposed to transform inhibitors into inactive compounds or to reduce their concentration.

The major inhibitory compound in pretreatment effluent is acetic acid. But the acetic acid resistances of the microorganisms are expected to rise in the close future and at pH of 4.5 to 5 the acetic acid concentrations of 2-3 g/l already do not have an inhibitory action on fermentation (Personal Communication Marco Kuyper and Musatto *et al*., 2004). Experimental studies show that lignin monomers in untreated hydrolysate is composed of mainly 4-hydroxy benzoic acid, vanillin and catechol and these compounds does not have an inhibitory effect up to 1 g/l concentration (Palmqvist and Hahn-hagerdal, 2000). The inhibition arising from furfural and hydroxymethyl furfural is reported to be 40-70% on growth for concentrations 1-3 g/l (Olsson and Hahn-Hagerdal, 1996).

In following sections, biological, physical and chemical detoxification techniques are given.

2.2.1. Biological Methods

Biological methods of treatment involve the use of specific enzymes or microorganisms that act on the toxic compounds present in the hydrolyzate and change their composition. Wood hydrolyzate detoxified with laccase and peroxidase enzymes of the white-rot fungus *Trametes versicolor* (Mussatto et al., 2004). These enzymes are effective on phenolic acids and phenolic monomers. The fact that the inhibition is significantly decreased by removing phenolic compounds by laccase, shows that the main origin of inhibition is from the phenolic compounds (Palmqvist and Hahn-Hagerdal, 2000a).

The filamentus soft-rod fungus *Trichoderma reesei* has been reported to degrade inhibitors in hemicellulose hydrolyzate obtained after steam pretreatment of willow. Acetic acid, furfural and benzoic acid derivatives are removed from the hydrolyzate by treatment with *T. reesei* (Palmqvist and Hahn-Hagerdal, 2000a).
Current enzyme technologies on detoxifying enzymes prevent biological detoxification be an economically and technologically feasible option for detoxification.

2.2.2. Physical and Chemical Methods

Several methods of detoxification, i.e., the removal of inhibitors from lignocellulosic hydrolyzate, have been proven to increase their fermentability. The addition of activated charcoal, extraction with organic solvents, ion exchange, ion exclusion, molecular sieves, overliming and steam stripping have been investigated (Olsson L. and B. Hahn-Hägerdal, 1996). The summary of detoxification methods are given in Table 2.2. Among the physical and chemical detoxification methods, overliming, evaporation, steam stripping and absorption are the most widely mentioned ones.

The activated charcoal absorption by itself removes the lignin degradation products by 95.4%. Combined with evaporation this treatment can reduce the toxic material concentrations to noninhibitory levels (Mussatto and Roberto, 2004). In NREL report (Alan *et al.*,2002), it is reported that 70% of furfural and HMF and 8% of acetic acid is removed by flash cooling of the pretreatment hydrolyzates. Later on, liming is applied to neutralize the media. The selection of detoxification method is described in Section 2.7.

Table 2.2 Detoxification procedures (Olsson L. and B. Hahn-Hägerdal, 1996).

Procedure	**Effect of Detoxification**
Steam stripping	Removal of volatiles (furfural, phenols) including acetic acid
Neutralization with CaO, NaOH, KOH; activated carbon; filtration	Reduction of acetic acid concentration
Neutralization to pH 6.5 or over titration to pH 10 with $Ca(OH)_2$,CaO, or KOH; removal of precipitate; H_2SO_4 to pH 6.5	Precipitation of acetate, heavy metals, furfurals, tannins, terpenes, phenolics
Ion exclusion chromatography (cation exchange resin)	Removal of aromatic monomers and dimmers
Ether extraction	Removal of furfural
Evaporation under vacuum	Removal of acetic acid
Ethylacetate extraction	Removal of lignin-degradation products
Molecular sieve	Partial removal of acetic acid, furfural, and soluble lignin
Mixed bed ion resins	Partial removal of acetic acid, furfural, and soluble lignin
Anion/cation exchange resins	Partial removal of acetic acid, removal of metal ions

2.3. Hydrolysis of Cellulose into C_6 Sugars

Three different approaches are applied for hydrolysis of cellulose in to glucose: dilute acid hydrolysis, concentrated acid hydrolysis and enzymatic hydrolysis of the cellulose. Dilute acid and concentrated acid treatments shows similarity with the pretreatment of the lignocellulose with same methods and are not favorable because of the mineral acid uses.

Cellulase enzyme is used for enzymatic hydrolysis. The combination of the enzymes endoglucanase, exoglucanase and β-glucosidase (cellobiase), collectively known as cellulase, are needed to break down cellulose into its constituent glucose monomers (Wheals et al., 1999).

The cellulase enzymes are commercially available, but, the high cost of enzymes increases the cost of ethanol production price. Recently Novozymes reported that the cost of enzymes for biomass-based fuel ethanol production has been reduced to $ 0.033-0.06 per kg in laboratory trials, a 30-fold reduction since 2001. Enzymes are no longer the main economic barrier in the commercialization of biomass technology. (Novozymes web page, 14 April 2005).

Since the latest technology in cellulase production is not published yet, as first step, the cellulase is assumed to be purchased from a supplier or it can be produced near the plant by license owning contractors. The process selection for the cellulose hydrolysis is carried out in Section 2.7.

2.4. Alternatives for the Fermentation of the C_5&C_6 Sugars to Ethanol

Several different processes are proposed for the fermentation of the bagasse hydrolysate to ethanol. Different process alternative are as follows,

2.4.1. Alternative 1

The hydrolysate from pretreatment is washed to remove the pentose sugars and by products of hemicellulose hydrolysis and lignin degradation. Depending on the type of pretreatment this stream can be detoxified before being sent to C_5 fermentation reactor. The solid containing cellulose and lignin is treated with cellulase enzyme to solubilize cellulose. The cellulase enzyme is produced in a separate fermentor and presented to the hydrolysis reactor. The hydrolyzed cellulose is separated from lignin and sent to hexose fermentation reactor. The fermentation medium of pentose and hexose fermentations can feed to the same distillation column from different feed point due to the differences in their ethanol concentrations. Since all the process steps are occurring independent from others, main advantage of this method is its flexibility. However, during the hydrolysis of cellulose high levels of glucose accumulating in the reactor inhibit the activity of the cellulase. Cellulase can also be purchased (See Figure 2.1).

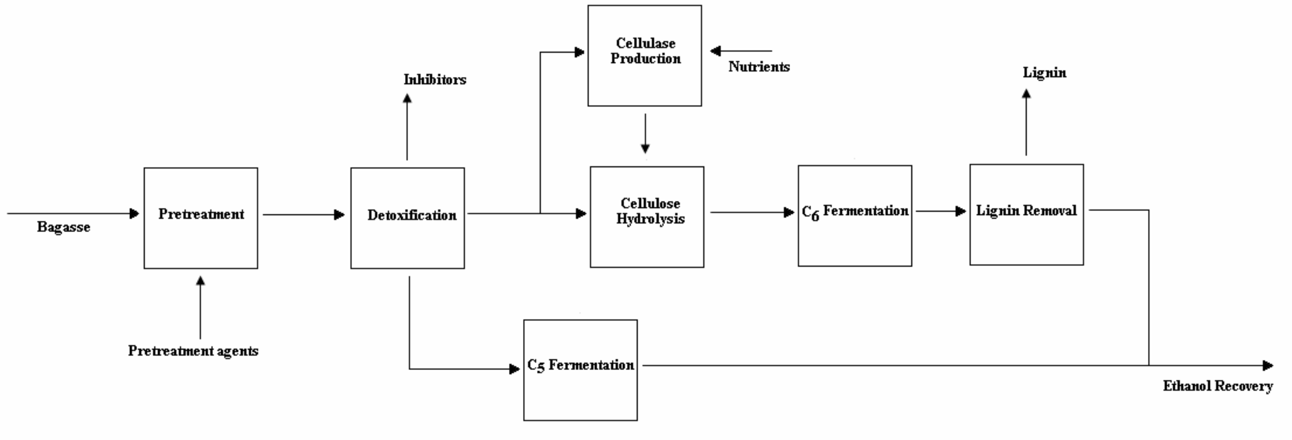


Figure 2.1 Lignocellulose to ethanol process alternative 1

2.4.2. Alternative 2

The treatment of the hydrolysate from pretreatment and C_5 fermentation processes are the same with Alternative 1, however, here the lignin and cellulose are directly fed to the fermentation reactor where the hydrolysis of cellulose and fermentation of ethanol occurs simultaneously. Since glucose is immediately converted to ethanol the inhibition of cellulose does not occur. Shorter fermentation times and reduced risk of contamination due to rapid conversion of glucose to ethanol is positive attributes for this alternative. On the other hand, the optimal conditions for ethanol fermentation and cellulose hydrolysis are at different temperatures and pH values. Therefore, the strains used for ethanol fermentation and cellulose hydrolysis should be carefully selected (See Figure 2.2).

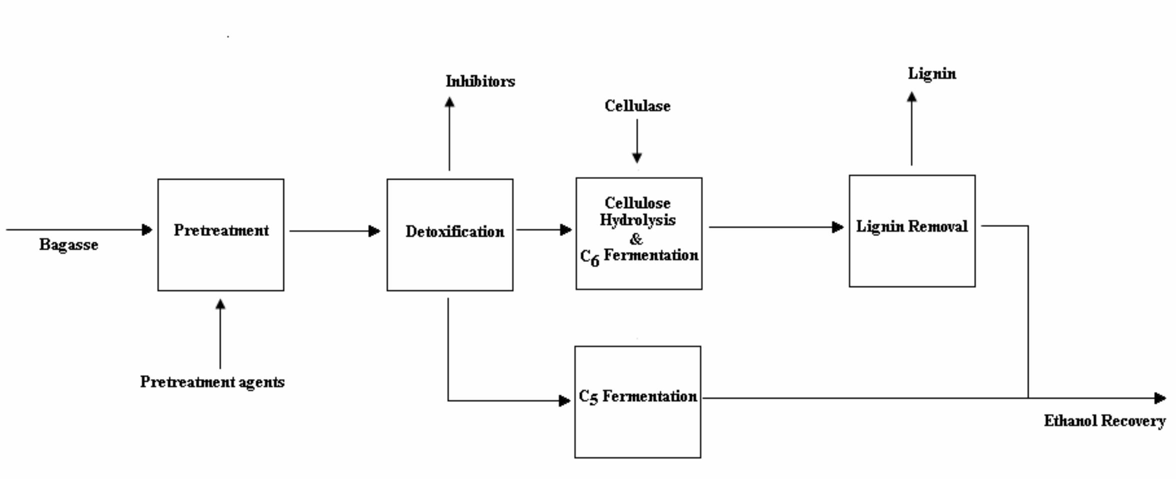


Figure 2.2 Lignocellulose to ethanol process alternative 2

2.4.3. Alternative 3

In this alternative the cellulose is hydrolyzed in a separate vessel and sent to a C_5&C_6 co-fermentor and lignin is removed after the hydrolysis step by means of solid liquid separation. Cellulose can be purchased or produced within the battery limits (See Figure 2.3).

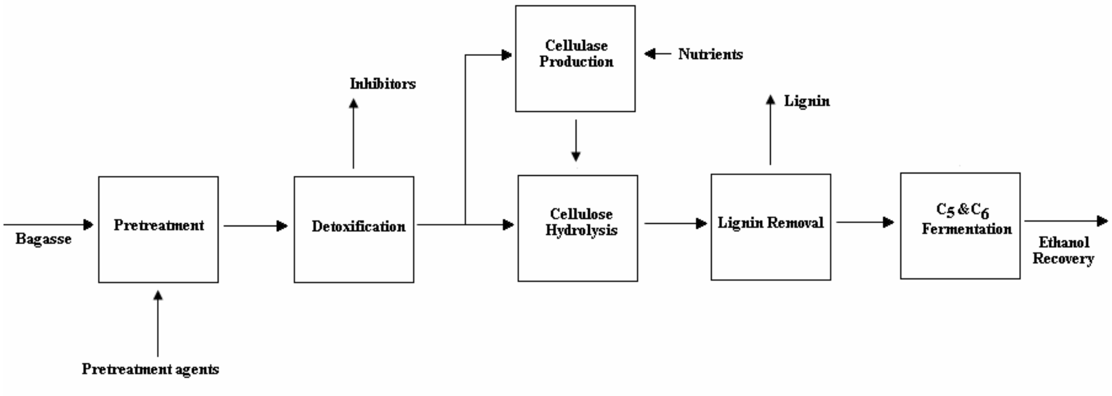


Figure 2.3 Lignocellulose to ethanol process alternative 3

2.4.4. Alternative 4

The process is same with Alternative 2, but, in this alternative the ethanol fermenting strains are genetically modified to secrete cellulase enzyme (See Figure 2.4).

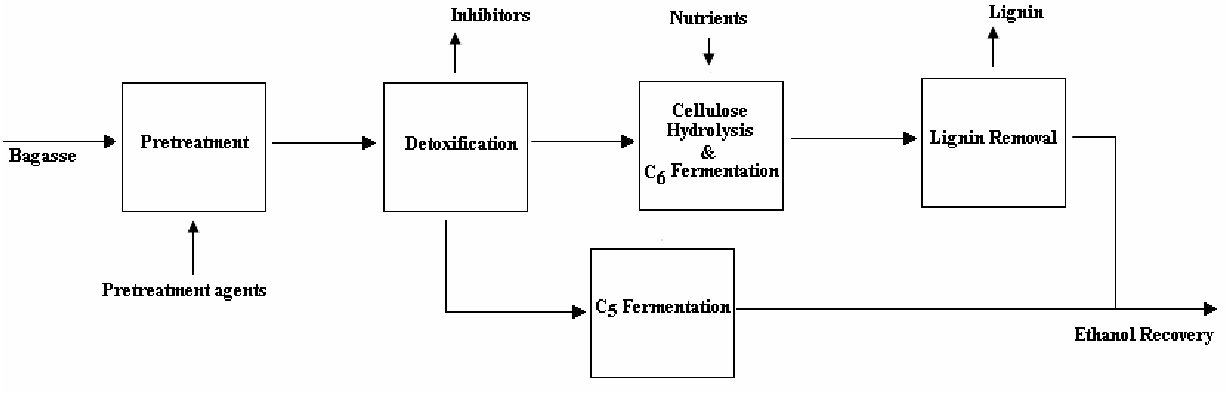


Figure 2.4 Lignocellulose to ethanol process alternative 4

2.4.5. Alternative 5

The hydrolyzate from pretreatment can be detoxified depending on the pretreatment and send to fermentor where saccharification, hexose and pentose fermentations to ethanol occur simultaneously. Saccharification and fermentation is carried out by separate microorganisms (See Figure 2.5).

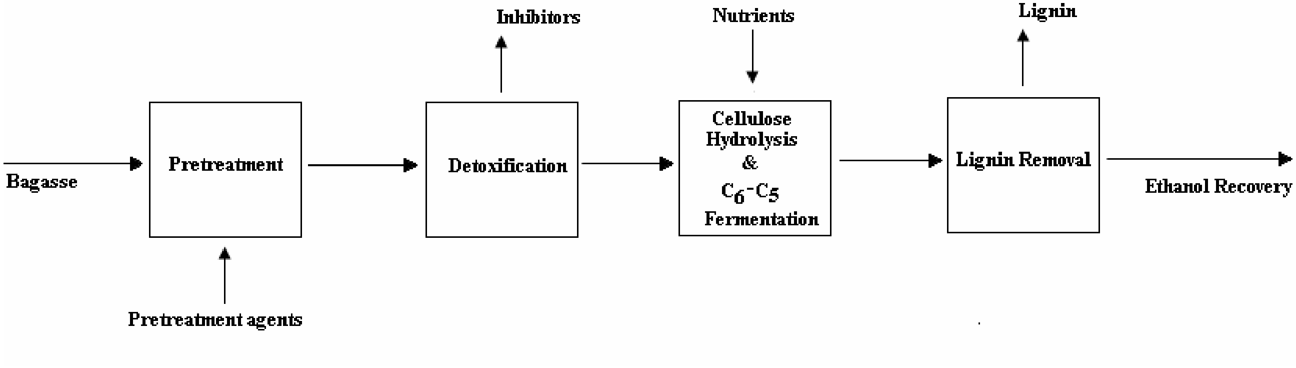


Figure 2.5 Lignocellulose to ethanol process alternative 5

2.4.6. Alternative 6

In this alternative the hydrolysis of cellulose and co-fermentation of C_5&C_6 sugars are carried out by one genetically modified organism. There is a group of microorganisms (*Clostridium, Cellulomonas, Trichoderma, Penicillium, Neurospora, Fusarium, Aspergilus,* etc.) showing a high cellulolytic and hemicellulolytic activity, which are also highly capable of ferment monosaccharides into ethanol (Szczodrak and Fiedurek, 1996). By genetically modifying these strains it could be possible to obtain strains that can hydrolyze cellulose and xylan and co-ferment C_5&C_6 sugars and that can withstand toxic compounds (See Figure 2.6).

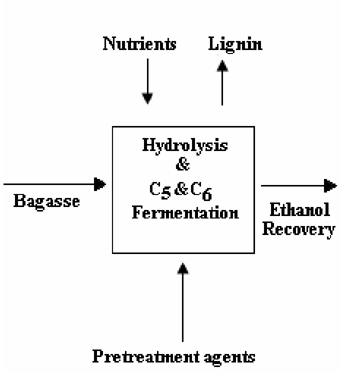


Figure 2.6 Lignocellulose to ethanol process alternative 6

2.4.7. Evaluation of the Alternatives

The characteristics of the microorganisms are the main determining factor to choose among the alternatives. Currently, existing microorganisms can utilize pentose and hexose sugars together. On the other hand, there aren't any microorganisms with high yields and productivities reported that can co-ferment hexose and pentose sugars and at the same time hydrolyze cellulose, xylan or both together. Therefore, alternatives 4, 5 and 6 are not feasible alternatives. The selection of the final alternative is carried out in Section 2.7.

2.5. Microorganism Alternatives for Pentose and Hexose Fermentation

In this section the recombinant and wild type microorganisms that can utilize hexose sugars, pentose sugars and both hexose and pentose sugars are presented to gain insight about the capabilities of different microorganisms.

2.5.1. Hexose Fermenting Organisms

The Microorganisms that can ferment hexose sugars efficiently are listed in Table 2.3.

Table 2.3 Performance of glucose fermenting organisms
(Olsson and Hanh- Hagerdal, 1996)

Strain	**Yield (g g^{-1})**	**Productivity (g $l^{-1}h^{-1}$)**
Saccharomyces cerevisiae	0.48	8
Zymomonas mobilis NRRL 04023	0.48	100

2.5.2. Pentose Fermenting Organisms

The Microorganisms that can ferment pentose are listed in Table 2.4.

Table 2.4 Performance of xylose fermenting organisms (Olsson and Hanh- Hagerdal, 1996)

Strain	**Yield (g g^{-1})**	**Productivity(g $l^{-1}h^{-1}$)**
Bacteria: Naturally occurring		
Clostridium Thermohydrosulfuricum 39 E	0.39	
Bacteria : Recombinant		
Erwinia Chrysanthemi B374 (*pcd*)	0.44	
Escherichia Coli B, pLO1297 (*pdc, adhB)*	0.49	0.70
Escherichia coli K011 (*pdc, adhB, frd*$^-$*)*	0.52	0.87
Klebsiella oxytoca M5A1 (*pdc, adhB)* [a]	0.46	0.96
Klebsiella planticola SDF20 *(pdc, pfl*$^{-1}$*)*	0.44	0.18
Zymomonas mobilis CP4 (pZB5) [b]	0.44	0.57
Yeasts : Naturally occurring		
Candida shehatae CBS 4705	0.48	0.19
Candida sp. CSIR-62 A/2	0.40	0.42
Pichia stipitis CBS 5776	0.45	0.34
Yeasts: recombinant		
Schizosaccharomyces pombe (xyi A)	0.42	0.19
Fungi		
Fusarium oxysporum VlT-D-80134 (3)	0.50	0.17
Paecilomyces sp. NFI **ATCC 20766 100**	0.40	0.24
Xylose fermentation with xylose isomerase and yeast		
Saccharomyces cerevisiae Bakers' yeast With Optisweet-p enzyme	0.50	1.00

Zymomonas mobilis has been shown to have ethanol productivity superior to yeast. In glucose medium this microorganism can achieve ethanol levels of 12% at yields up to 97% of theoretical yields. Also *Zymomonas* is the only genus that exclusively utilizes Entner-Doudoroff pathway anaerobically. Due to the low ATP yields in ED pathway biomass yields are also lower (Aristidou and Penttila, 2000). On the other hand, the low stability and the high optimal pH make this microorganism an unfavorable one for industrial application.

2.5.3. Pentose and Hexose Fermentation Together

The Microorganisms that can ferment pentose and hexose sugars together are listed in Table 2.5.

Table 2.5 Performance of lignocellulose hydrolyzate fermenting organisms (Olsson and Hanh-Hagerdal, 1996)

Strain	**Substrate**	**Yield (g g^{-1})**	**Productivity (g l^{-1}h^{-1})**
Bacteria			
Escherichia Coli B, pLOI 297	g, ga m, x	0.39	0.33
Escherichia coli K011	a, g, ga, m, x	0.43	0.67
Escherichia coli K011	a, g, x	0.51	0.42
Escherichia coli (Aristidou and Penttila, 2000)	g,x	0.5	1.80
Klebsiella oxytoca M5A1 (*pdc, adhB)* [a]	g,x	0.5	2.00
Yeasts			
Candida shehatae ATCC 22984	a, g, ga, m, x	0.37	0.36
Candida shehatae ATCC 22984 (R)	a, g, ga, m, x	0.45	1.27
Candica sp. *XF217*	a, g, x	0.4	0.44
Pichia stipitis CBS 5776 (R)	a, g, ga, m, x	0.44	0.78
Xylose fermentation with xylose isomerase and yeast			
Saccharomyces cerevisiae + *Lactobacillus brevis* XI	a, g, ga, m, x	0.40	0.11
Saccharomyces cerevisiae Bakers' yeast + Maxazyme	a, g, ga, m, x	0.41	0.37
Saccharomyces cerevisiae Bakers' yeast With Optisweet-p enzyme	a, g, ga, m, x		

2.5.4. Evaluation of Microorganisms

All the microorganisms given in Tables 2.3-2.5 are taken from a publication from 1996. These data are useful to have some insight about the microorganisms and their capabilities. As it seen, there are various recombinant and wild type microorganisms that can utilize pentose sugars and hexose sugars at high productivities. Furthermore, there are numerous recombinant microorganisms that can utilize both hexose and pentose sugars. The main disadvantage of these microorganisms is their low productivities which require high investment costs. Another disadvantage is their low stabilities which make the cell recycle difficult due to their low viabilities.

In more recent publications, higher yields, productivities and viabilities are reported for recombinant microorganisms. In his review article, Hahn-Hagerdal (2004) reported productivities of 1.13, 0.38, 0.43 g ethanol g biomass^{-1}h^{-1} respectively for *Zymomonas mobilis* 8β, *E-coli* KO11 and *P.stipitis*. When high cell densities are used the productivities would be higher. Therefore, for the following decade the capacities of the microrganisms can be expected to get closer to current hexose fermenting microorganisms.

Z.Mobilis is a promising bacterium for future ethanol production. However, there aren't any clear data on how viable this bacterium is. To have high cell concentrations in the fermentor and to spend less substrate for cell growth, high cell viability and stability are necessary microbial characteristics. If the stability of these microorganisms is high, they will be more favorable than yeast in close future. On the other hand, their high pH optimum is a difficult hurdle to overcome for industrial application.

Another important factor is to prevent the contamination of the recycle cells. Since *Saccharomyces* species are utilized in current technology, keeping the cells sterile is easy by applying high pH conditions where yeast survives, but, not the other microorganisms. Therefore, changing yeast with another microorganism will also arise the problem of maintaining the sterility of the microorganisms during the cell recycle.

The recent studies on recombinant xylose utilizing strains decreased the fermentation times to ferment mixed substrates. For mixed substrate concentrations of 100 g/l glucose and 20 g/l xylose times as low as 24 hour are reported for the complete fermentation of sugars (Kuyper *et al.*, 2005). Using high cell densities and further developments would yield fermentation times compatible with current hexose fermenting strains. Therefore, considering all this information selecting pentose and hexose co-fermenting recombinant yeast as the microorganism will not be a wrong choice for future application.

2.6. Waste Treatment and Cogeneration

The waste streams generated in the process are the filter mud from filters containing sugar cane organic ingredients, vinasse from distillation containing fermentation by products and sugar cane organic material, water from rectification, dehydration, evaporator, crystallizer and drying containing volatile organics and detoxification waster containing the toxic organics. Since the organic loadings in vinasse are high, the vinasse is proposed to be concentrated in an evaporator and then burned in cogeneration unit together with the filter cakes and the genetically modified yeast. The organic remaining organic rich streams are treated in waste treatment plant before being recycled to plant or sent back to the fields.

2.7.Selection of the Process for the Lignocellulose Hydrolysis Branch

The pretreatment method is selected as shown in Section 2.1. For the Detoxification method, based on the information given in Section 2.2, flash cooling and overliming will be applied. Since the pretreatment is carried out at high temperatures, no extra heat input is required for steam stripping and just by flashing the effluent of pretreatment the volatile toxic materials (acetic acid and sugar degradation products) can be stripped of the media. Then, the remaining toxic materials, which are phenolic lignin degradation products, can be absorbed by activated charcoal if the concentrations are above the threshold inhibition concentrations.

For the lignin removal, different alternatives are present. Lignin can be solubilized by alkaline and peroxide treatments (1) and removed in liquid stream. This option has the disadvantage of high toxic phenolic material formation, consequently, high detoxification load. When lignin is solubilized, extra processes should be included to recover lignin and send to cogeneration, which will increase the cost of operation. Also, in Section 2.1 the alkaline-peroxide treatment was proved to contribute significantly to production cost. Therefore, this option is not feasible. The second option (2) is to send lignin together with cellulose to a cellulose hydrolysis unit and recover lignin after fermentation in the bottom product of distillation column. However, the presence of solid particles in fermentor increases the viscosity of the media consequently increase the stress applied on microorganisms and power requirement for agitation and makes it impossible to recycle the cells for multiple uses. But, by applying this technique the cellulase hydrolysis and fermentation can be carried out in one vessel without a lignin removal step in between. As a third option (3), a similar alternative is chosen in which everything is the same with second option but, the lignin is removed from the effluent stream of cellulase hydrolysis just after the hydrolysis vessel and sent to co-generator. This alternative is not feasible if a simultaneous operation needs to be carried out for hydrolysis and fermentation units. On the other hand, it is more desirable than the second option in terms of pumping and agitation costs and cell recycling. As mentioned earlier, the optimal conditions for cellulose hydrolysis and fermentation are widely different from each other (45-50 ºC for hydrolysis and 30-35 ºC for fermentation). The studies on creating more thermotolerant microorganisms to ferment sugars to ethanol are still carried out by researchers and there isn't any industrially functional strains generated yet. Therefore, two separate vessels for hydrolysis and fermentation are necessary for optimal operation of both units. Consequently, removal after cellulose hydrolysis (third option) is the selected option for lignin removal.

Initially, cellulase for cellulose hydrolysis is assumed to be purchased from an external supplier. After sufficient information is obtained for recently developed high cellulase yielding microorganism, the cellulase production unit can be inserted into process to compare the feasibility of the two options.

The selection between having one co-fermentor for C_5&C_6 sugars together or fermenting them in two different fermentors is made after a preliminary calculation based on 5 different microorganisms, which can utilize hexose only, pentose only and hexose and pentose together. The microorganisms having high yield, high productivity or both are selected from the list presented in Tables 2.3-2.5 and given in Table 2.6 (this microorganisms are used just to see the effect of productivities and yields on investment cost they are not the selected microorganisms).

Table 2.6 Microorganisms for calculations of fermentation options

Strain		**Substrate**	**Yield ($g\ g^{-1}$)**	**Productivity ($g\ l^{-1}h^{-1}$)**
1	*Zymomonas mobilis* NRRL 04023	Hexose	0.48	100
2	*Candida shehatae* ATCC 22984	a, g, ga, m, x	0.47	1.27
3	*Escherichia coli* K011	A, g, x	0.51	0.42
4	*Escherichia coli* K011 (*pdc, adhB, frd$^-$)*	Pentose	0.52	0.87
5	*Klebsiella oxytoca* M5A1 (*pdc, adhB)* [a]	Pentose	0.46	0.96

For the calculations it is assumed that the bagasse is pretreated by steam explosion. Using the process parameters, the calculation results and assumptions in Section 2.1, the process flows are found as given in Figure 2.7 and 2.8 based on two cases. The first case (Figure 2.7) is the separate fermentors for hexose and pentose sugars and the second case (Figure 2.8) is co-fermentor for hexose and pentose sugars. The two cases are also separated in to two sub cases. In Case 1.1 the hexose fermenting organism is *Zymomonas mobilis* (1) and the pentose fermenting organism is *Escherichia coli* K011 (4) and in Case 1.2 the hexose fermenting organism is same but the pentose fermenting organism is *Klebsiella oxytoca* M5A1 (5). In Case 2.1 the co-fermenting microorganism is *Escherichia coli* K011 (3) and in Case 2.2 the co-fermenting organism is *Candida shehatae* ATCC 22984 (2). Based on the productivities and yields on substrates, the required vessel volumes and total ethanol yield on total initial fiber are calculated as given in Table 2.7.

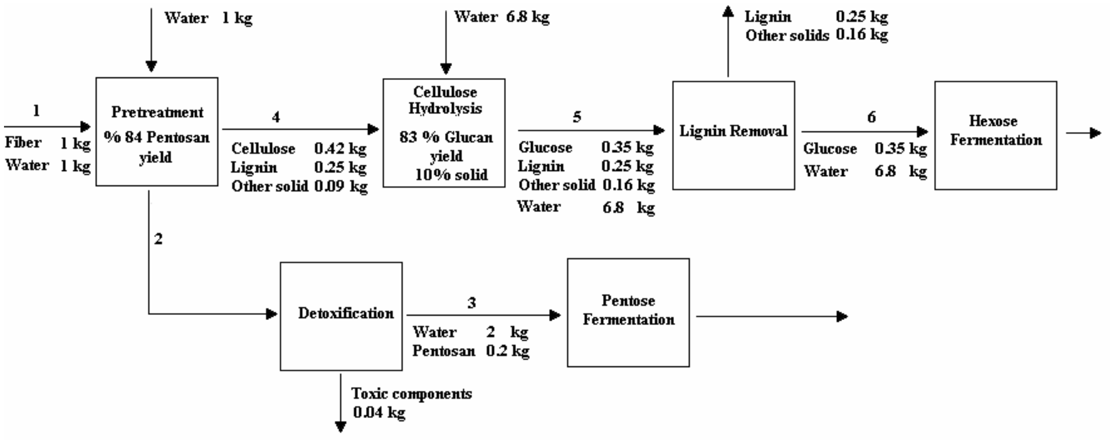


Figure 2.7 Flow sheet for Case 1 of fermentation options (other solids are nonhydrolyzed components and ash)

Table 2.7 Ethanol yields, reactor volumes and expected cost of fermentor for the cases

	Ethanol yield (kg ethanol/kg fiber)	Fermentor Volume (litre)	Reactor Cost*
Case 1.1	0.272	1.69^H+120^P	30 C_0
Case 1.2	0.258	1.69^H+96^P	26 C_0
Case 2.1	0.281	669	95 C_0
Case 2.2	0.248	195	40 C_0

H hexose fermentor volume and P is the pentose fermentor volume

* During cost calculations it is assumed that price of 1 l fermentor is C_0 and the fermentor prices are calculated using the equation Cost = $C_0\,(\text{Volume})^{0.7}$

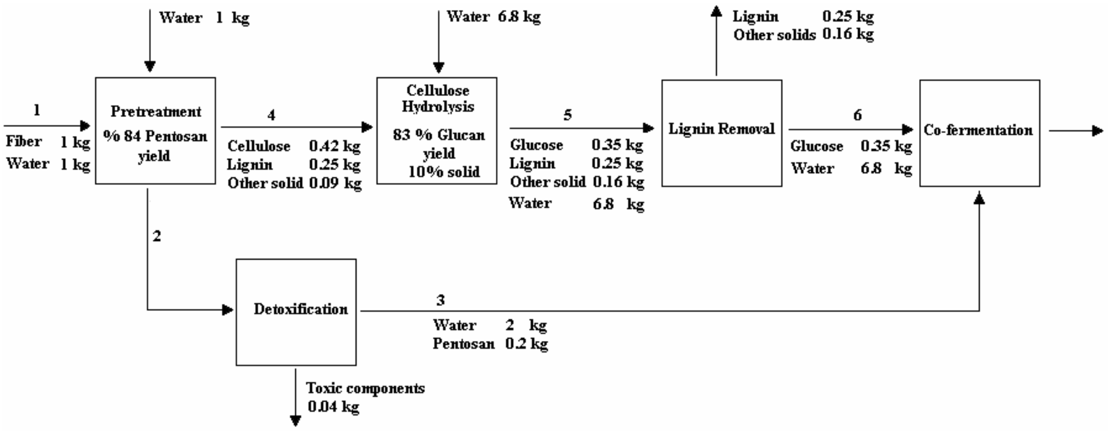


Figure 2.8 Flow sheet for Case 2 of fermentation options

Based on this information having two separate fermentors is more economically feasible than having a co-fermentor. And the productivity difference of microorganisms 4 and 5 does not have significant effect on vessel cost. Therefore, choosing the high ethanol yielding microorganism is more economical in terms of yearly operation income. On the other hand, considering that this literature values are for year 1996 and in more recent publications, (Aristidou and Penttila, 2000) productivities as high as 1.3 g ethanol g biomass^{-1}h^{-1} and yields close to theoretical maximum are reported for *E-coli*, it can be predicted for the close future to have even higher productivities with different organism. Therefore, considering their current close investment costs (40C_0 and 26 C_0), Case 2.2 is expected to be even more cost efficient then Case 1.2 in terms of both yearly cash flow and investment costs in following decade. As a result having a co-fermenting organism for the fermentation of pentose and hexose sugars together is the best option for the close future.

As a result the hexose and pentose fermentations will be carried out in one vessel by hexose and pentose co-fermenting microorganisms.

3. Basis of Design

3.1. Description of Design

This process is about the investigation of the feasibility of plant producing ethanol and sugar from sugar cane to gain better understanding of the technical and economical characteristics of the existing plant and to apply the knowledge gained from current technologies to future investments. In the early phases of ethanol industry, the government subsidies and support eased the investment on ethanol plant. However, currently the lack of subsidies is an important factor decreasing the competitiveness of the new investments with the already existing plants. Another factor in competitiveness is that existing plants have already depreciated and they can cut the prices more then a newly invested plant. Therefore, in the first step of this project the design of currently operating sugar ethanol plant will be carried out to determine the required technologies, investment and the feasibility of the plant with the current ethanol and sugar prices. Current plants are generating their own electricity and steam, by combusting the bagasse.

In the second step of the design, knowledge gained from the current technologies is applied to future design in which the sugar cane bagasse is hydrolyzed to pentose and hexose sugars to be fermented into ethanol. For the design of the future plant (2015 plant), the present alternatives are investigated and the best options for each step are determined and implemented into the existing sugar plant. Since the fibers in the 2015 design are converted into ethanol, the electricity and steam generation of the plant will be lower than in the current design which also affects the operation costs of the plant. The alternatives for 2015 plants can be found in Section 2. The description of the processes can be found in Section 5.

3.2. Block Schemes

The block schemes of the selected processes for 2005 and 2015 plants can be found in Appendix A.1.1& A.1.2

3.3. Pure Component Properties

The pure component properties for the components existing in the mass balances can be found in Appendix A.2.1.

3.4. Basic Assumptions

3.4.1. Plant Capacity

Since the harvesting season for the sugar cane lasts approximately 200 days the operation time for the plant is taken as 174 days based on the information given by Damen (2001). Calculating for 24 hours, total yearly operating hours is 4,176 hours.

3.4.1.1. Feedstock

Composition of the raw materials for conventional and future plan is given in Tables 3.1-3.8.

Table 3.1 Composition of sugar cane

Stream name:	Sugar Cane			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Water Soluble solids Fiber Dirt	73-76 10-16 11-16 0.5-1	68.0 17.0 14.0 1.0		The design values are from brazil. Values are in mass percentage
Total	100.0	100.0		
Total flow (kg/s)	332.6 kg/s		5,000,000 ton/year	

Table 3.2 Composition of recycle water stream

Stream name:	Recycle Water			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Water Organics	~100 <0.1%	100.0	(1)	(1) Organics are not included in mass balances Values are in mass percentage
Total	100.0	100.0		
Total flow (2005)	267.2 kg/s		4,015,500 ton/year	
Total flow (2015)	546.3 kg/s		8,213,000 ton/year	

Table 3.3 Composition of sulphuric acid stream

Stream name:	Sulphuric Acid			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Sulphuric acid Fe Cu Pb	98% 30 ppm 0.5 ppm 0.5 ppm	100.0		(IRef-13) Values are in mass percentage
Total	100.0	100.0		
Total flow (2005)	0.14 kg/s		2,134.8 ton/year	
Total flow (2015)	0.94 kg/s		14,101.5 ton/year	

Table 3.4 Composition of sulphur stream

Stream name:	Sulphur			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Sulphur Water Ash Acidity Organics Arsenicum Ferrum	99.5 (min) 0.05(max) 0.10(max) 0.005(max) 0.30(max) 0.01(max) 0.005(max)	100.0		(IRef-14) Values are in mass percentage
Total	100.0	100.0		
Total flow	0.108 kg/s		1,623.6 ton/year	

Table 3.5 Composition of cellulase

Stream name:	Cellulase			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Cellulase Water	100.00 0.00	0.23 99.77		The cellulase is mixed with recycle water Values are in mass percentage
Total	100.0	100.0		
Total flow	0.366 kg/s		5,502.3	

Table 3.6 Composition of flocculent

Stream name:	Flocculent			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Polyacrylamide Water	100.00 0.00	0.5 99.5		Values are in mass percentage
Total	100.0	100.0		
Total flow	0.001 kg/s		10.4 ton/year	

Table 3.7 Composition of limestone

Stream name:	Limestone			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
$CaCO_3$	96-98	98.0		(1) This values is taken for all the non $CaCO_3$ compounds Values are in mass percentage
$MgCO_3$	<2	2.0	(1)	
SiO_2	<2			
AlO_3	<0.4			
Fe_2O_3	<0.3			
Total	100.0	100.0		
Total flow (2005)	0.303 kg/s		4,555.2 ton/year	
Total flow (2015)	0.925 kg/s		13,906.1 ton/year	

Table 3.8 Composition of air

Stream name:	Air			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Nitrogen	75.7	79.0	(1)	(1) This stream is used to provide oxygen from air. (2) Components excluded in mass balance.
Oxygen	23.3	21.0		
Argon	0.93		(2)	
Carbon dioxide	0.03		(2)	
Total	100.0	100.0		
Total flow (2005)	4,074.2 kg/s		6,125,000 ton/year	
Total flow (2015)	2,369.4 kg/s		3,562,000 ton/year	

3.4.1.2. Products

The specifications of refined sugar crystals and dehydrated ethanol are given in Tables 3.9 and 3.10.

Table 3.9 Composition of sugar crystals

Stream name:	Sugar Crystals			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Sucrose	99.7	99.94		Values are in mass percentage
Water	0.05	0.02		
Ash	0.05	0.04		
SO_2	5ppm	trace		
Total	100.0	100.0		
Total flow (2005)	24.6 kg/s		370,401 ton/year	
Total flow (2015)	24.6 kg/s		370,405 ton/year	

Table 3.10 Composition of anhydrous ethanol

Stream name:	Sugar Crystals			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Ethanol	Min. 90.0%	99.86		Values are in mass percentage
Water	Max. 1.25%	0.12		
Volatiles	Max. 2.00%	0.02		
Acidity	Max. 30 ppm	trace		
Total chlorine	Max. 4 ppm	0.00		
Others	Max. 4 ppm	0.00		
Lead	Max. 2 ppm	0.00		
Phosphorus	Max. 0.2 ppm	0.00		
Sulfur	Max. 20 ppm	0.00		
Gum, heptane	Max. 5 ppm	0.00		
Total	100.0	100.0		
Total flow (2005)	11.0 kg/s		165,956 ton/year	
Total flow (2015)	26.4 kg/s		396,872 ton/year	

3.4.1.3. Waste

1. Conventional Plant: 2005 Case

The waste streams for 2005 plant are presented in Tables 3.11-3.20.

Table 3.11 Composition of Stream <143>

Stream name:	Exhaust gas <143>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Water		14.97		Values are in mass percentage
O_2		3.83		
N_2		63.78		
CO_2		17.30		
NO_2	0.32 mg/m^3	0.12		
SO_2	0.37 mg/m^3	0.002		
Total		100.00		
Total flow	486.5 kg/s		7,313,846 ton/year	

Table 3.12 Composition of Stream <146>

Stream name:	Exhaust gas <146>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Water		5.47		Values are in mass percentage
O_2		5.14		
N_2		58.16		
CO_2		31.14		
NO_2	0.32 mg/m^3	0.09		
SO_2	0.37 mg/m^3	0.001		
Total		100.00		
Total flow	1.587 kg/s		23,858.3 ton/year	

Table 3.13 Composition o Streams <144> &<145>

Stream name:	Ash <144>&<145>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Ash		100.0		Composition of Ash is as given in Table 5.5. Values are in mass percentage
Total		100.0		
Total	1.24 kg/s		18,641.7 ton/year	

Table 3.14 Composition of Stream <123>

Stream name:	Filter cake <123>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Bagasse		3.05		(1) Calcium salts are composed of $CaSO_3$, $CaSO_4$ and calcium phosphate precipitates (2) Sugar cane organics are the non sugars that are separated by clarification like gums and proteins. Values are in mass percentage
Calcium salts		2.32	(1)	
Dirt-Soil		21.75		
Sugar cane organics		14.73	(2)	
Polyacrylamide		0.004		
Sucrose		3.10		
Water		55.05		
Total		100.00		
Total flow	15.3 kg/s		230,014 ton/year	

Table 3.15 Composition of Streams <147>, <226>, <227> and <228>

Stream name:	Waste water <147>, <226>, <227> and <228>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Water		99.79		(1) This compounds are 0% in <147>, <226> and <228> Values are in mass percentage
Sugar cane volatile organics		0.15	(1)	
Sulpur dioxide		0.06	(1)	
Total		100.00		
Total flow	163.3 kg/s		2,454,987 ton/year	

Table 3.16 Composition of Stream <334>

Stream name:	Waste water <334>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Water		98.54		Values are in mass percentage
Ethanol		0.11		
Nitrogen		1.35		
Total		100.00		
Total flow	0.879 kg/s		13,214.5 ton/year	

Table 3.17 Composition of Stream <327>

Stream name:	Carbon dioxide <327>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Carbon dioxide		95.58		Values are in mass percentage
Water		4.40		
Ethanol		0.02		
Total		100.0		
Total flow	11.3 kg/s		170,090 ton/year	

Table 3.18 Composition of Stream <332>

Stream name:	Waste acid solution <332>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Water		99.9		Values are in mass percentage
Sulphuric acid		0.1		
Total		100.0		
Total flow	144.7 kg/s		2,175,362 ton/year	

Table 3.19 Composition of Stream <311>

Stream name:	Vinasse <311>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Water		80.25		Values are in mass percentage
Acetic acid		0.07		
Ethanol		Trace		
Glycerol		1.01		
Sugar cane organics		18.16		
Fusel oil components		0.41		
Succinic acid		0.10		
Total		100.00		
Total flow	17.0 kg/s		255,571.2 ton/year	

Table 3.20 Composition of Stream <328>

Stream name:	Yeast <328>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Yeast		100.00		
Total		100.00		
Total flow	3.52 kg/s		52,903.2 ton/year	

2. Future Plant:2015 Case

The waste streams the composition and the names of which are same with 2005 plant are <226>, <227>, <228> and <332>. The waste streams generated in 2015 are presented in Tables 3.21-3.28.

Table 3.21 Composition of Stream <139>

Stream name:	Exhaust Gas <139>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Water		4.83		Values are in mass percentage
O_2		6.93		
N_2		68.68		
CO_2		19.43		
NO_2	$0.32\ mg/m^3$	0.00		
SO_2	$0.37\ mg/m^3$	0.13		
Total		100.00		
Total flow	2.79 kg/s		41,913.7 ton/year	

Table 3.22 Composition of Stream <140>

Stream name:	Flash Vapor <140>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Water		100.00		Values are in mass percentage
Total		100.00		
Total flow	1.59 kg/s		23,963.6 ton/year	

Table 3.23 Composition of Streams <327>

Stream name:	Carbon dioxide <327>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Carbon dioxide		95.87		Values are in mass percentage
Water		4.12		
Ethanol		0.01		
Total		100.00		
Total flow	26.92 kg/s		404,764.6 ton/year	

Table 3.24 Composition of Stream <334>

Stream name:	Waste water <334>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Water		98.78		Values are in mass percentage
Ethanol		0.14		
Nitrogen		1.08		
Total		100.00		
Total flow	2.13 kg/s		32,081.7 ton/year	

Table 3.25 Composition of Stream <405>

Stream name:	Flash Vapor <405>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Acetic-acid		0.51		Values are in mass percentage
Furfural		1.51		
HMF		0.28		
Water		97.70		
Total		100.00		
Total flow	24.43 kg/s		367,300.9 ton/year	

Table 3.26 Composition of Stream <436>

Stream name:	Exhaust Gas <436>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Carb. dioxide Nitrogen NO_2 Oxygen Sulfur dioxide Water	 0.32 mg/m^3 0.37 mg/m^3	15.30 62.70 0.18 5.63 0.005 16.19		Values are in mass percentage
Total		100.00		
Total flow	279.85 kg/s		4,191,368 ton/year	

Table 3.27 Composition of Stream <433>

Stream name:	Ash <433>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Ash		100.0		Composition of Ash is as given in Table 5.5. Values are in mass percentage
Total		100.0		
Total flow	6.37 kg/s		95,703.9 ton/year	

Table 3.28 Composition of Stream <429>

Stream name:	Water Vapor <429>			
Components	Specification (%)		Notes	Additional Information (Also ref. Note numbers)
	Available	Design		
Acetic acid Ethyl alcohol Furfural Glycerin HMF Isoamylalcohol Lactic acid Succinic acid Water		0.31 trace 0.00 0.54 trace 0.05 0.05 0.06 98.99		Values are in mass percentage
Total		100.00		
Total flow	8.62 kg/s		129,634.7 ton/year	

3.4.2. Location

The proposed plant location is Sao Paulo state of Brazil.

3.4.3. Battery Limits

3.4.3.1. Units inside the plant

All the operation units are inside the battery limit. The feedstock is available from the fields and transportation cost is required for the feedstock. The product needs to be transported to the market. But, the transportation costs are assumed to be outside the battery limit. The units inside the battery limits are presented in Tables 3.29-3.30.

Table 3.29 Units inside the plant for 2005 case

Unit	Additional Information
Milling	For sugar extraction
Cogeneration	For Steam and electricity generation
Juice Clarification and Pretreatment	Impurity removal and sterilization
Sugar recovery and dewatering	To obtain crystalline sugar
Fermentation	Fermentation of sugars in to ethanol
Ethanol recovery and dewatering	To obtain anhydrous ethanol
Calciner	To provide the quick lime required

Table 3.30 Units inside the plant for 2015 case (in addition to the units in 2005 case)

Unit	Additional Information
Lignocellulose hydrolysis	To recover pentose and hexose sugars from lignocellulose
Detoxification	To remove the toxic materials liberated in hydrolysis
Hexose & pentose fermentation	To Ferment pentose sugars into ethanol

3.4.3.2. Units outside the plant

Cellulase production is outside the battery limits and the enzymes will be purchased. In 2005 case, the steam and electricity is produced with in the battery limits. Only utility requirement is the process water which can be reduced by efficient recycling. In 2015 case, since the amount of fibers to be combusted is lowered by almost 3/4, the energy generation by burning fibers does not meet the steam and electricity demands. Therefore, the steam generation is increased by reducing electricity generation and the electricity gap is filled by purchasing electricity form the grid. Also, in 2015 plant, the design of waste treatment is taken outside the battery limit, but, the waste treatment costs will be included in cost analysis. The process water requirement of 2015 can not be totally supplied by the recycle water. Therefore, process water will also be taken from outside the battery limit.

3.4.4. Definition Incoming and Outgoing Streams

3.4.4.1. Conventional Plant: 2005 case

The incoming and out going streams for 2005 plants are presented in Tables 3.32-33.

Table 3.31 Summary of incoming streams for 2005 case

Component	Stream	Price ($/ton)	Quantity (tons/year)	Total Cost ($/year)	Notes
Sugar cane	<101>	12.50	5,000,000.00	62,500,000.00	
Recycle water	<102>	0.00	1,400,000.00	0.00	
	<136>		2,017.51		
	<137>		233,856.67		
	<326>		207,057.77		
	<330>		2,173,332.38		
Limestone	<129>	6.78	4,556.68	30,894.29	
Sulphur	<106>	30.00	1,625.13	48,753.90	
Sulphuric acid	<330>	45.00	2,134.77	96,064.65	
Polyacrylamide	<136>	3,110.55	10.52	32,722.99	
Nitrogen	<323>		180.40		
Air	<105>		13,975.23		
	<127>		8,273.00		
	<141>		5,904,760.60		
	<223>		198,239.06		
Total			15,150,029.84	$62,708,435.83	

Table 3.32 Summary of outgoing streams for 2005 case

Component	Steam	Price ($/ton)	Quantity (tons/year)	Total Price ($/year)	Notes
Ethanol	<317>	400.00	165,955.91	66,382,364.00	
Sugar	<222>	176.00	370,400.84	65,190,547.84	
Exhaust gas	<143>		7,314,010.27		
	<146>		23,853.81		
Exhaust air	<225>		201,943.34		
Ash	<144>		18,620.62		
	<145>		21.05		
Filter cake	<123>		229,682.24		
Waste water	<147>		25,495.48		
	<226>		165,710.86		
	<227>		1,808,478.94		
	<228>		454,949.81		
	<334>		13,409.97		
Waste acid sln.	<332>		2,175,467.16		
Carbon dioxide	<327>		170,090.15		
Vinasse	<311>		255,571.20		
Fusel oil	<319>		1,703,261.78		
Yeast	<328>		52,897.22		
Total			15,149,820.65	$131,572,911.84	

3.4.4.2. Future Plant: 2015 Case

Table 3.33 Summary of incoming streams for 2015 case

Component	Stream	Price ($/ton)	Quantity (tons/year)	Total Cost ($/year)	Notes
Sugar Cane	<101>	12.5	5,000,000.00	62,500,000.00	
Water	<102>	0.00	1,400,000.00	0.00	
	<136>		2,017.51		
	<137>		233,856.67		
	<326>		456,795.94		
	<330>		5,232,444.48		
	<401>		87,988.65		
	<402>		700,561.25		
	<403>		303,490.80		
	<425>		510,473.40		
Limestone	<129>	6.78	13,907.58	94,293.39	
Sulphur	<106>	30.00	1,625.13	48,753.90	
Sulphuric acid	<330>	45.00	5,111.42	634,297.50	
	<402>		8,984.08		
Cellulase	<437>	3,419.68	5,499.23	18,805,627.55	
Polyacrylamide	<136>	3,110.55	10.52	32,722.99	
Nitrogen	<323>		345.77		
Natural gas	<128>		780.24		
Air	<105>	0.00	13,975.23	0.00	
	<127>		22,458.70		
	<223>		198,306.71		
	<434>		3,327,019.87		
Total			17,525,653.18	$82,115,695.33	

Table 3.34 Summary of outgoing streams for 2015 case

Component	Steam	Price ($/ton)	Quantity (tons/year)	Total Price ($/year)	Notes
Ethanol	<317>	400.00	396,872.01	158,748,804.00	
Sugar	<222>	176.00	370,405.35	65,191,341.60	
Exhaust Gas	<139>		41,909.17		
	<436>		4,192,111.84		
Exhaust Air	<225>		202,012.50		
Carbon dioxide	<327>		404,764.65		
Waste water	<140>		23,957.54		
	<226>		165,764.98		
	<227>		1,809,230.62		
	<228>		455,139.23		
	<334>		32,081.70		
	<405>		367,294.90		
	<429>		129,634.73		
Waste acid sln.	<332>		5,237,555.90		
Fusel oil	<335>		3,601,163.58		
Ash, gypsum, soil	<433>		95,700.89		
Total			17,525,599.59	$223,940,145.60	

3.5. Economic Margin

The preliminary economical evaluation of the processes is carried out based on the raw material costs as the main expanse. The revenues are calculated based on the sales revenue of sugar and ethanol. The raw material and product prices are as given in Tables 45-48.

The economical evaluation is carried out based on Equations 3.1-3.5.

Raw material cost : $$\sum_i Price\ of\ raw\ material\ i \times Yearly\ usage \qquad \text{Eqn 3.1}$$

Products Income : $$\sum_i Price\ of\ product\ i \times Yearly\ production \qquad \text{Eqn 3.2}$$

Net revenue : $$\sum Product\ income - \sum Raw\ material\ cost \qquad \text{Eqn 3.3}$$

Revenue after tax : $$Net\ revenue \times (1 - r_{tax}) \qquad \text{Eqn 3.4}$$

Net present value : $$\sum_n \frac{Net\ revenue}{(1+r_i)^n} \qquad \text{Eqn 3.5}$$

Where, r_{tax} is the tax rate, r_i is the interest rate and, n is the plant operation life which is 10 years. The corporate tax in Brazil is given as 34% (IRef-15) and the interest rate is taken as 18% based on the last eight years' interest rate profile of Brazil which is given in Figure 3.1.

Based on this information the yearly cash flows and net present values are calculated as given in Tables 3.36 & 3.37.

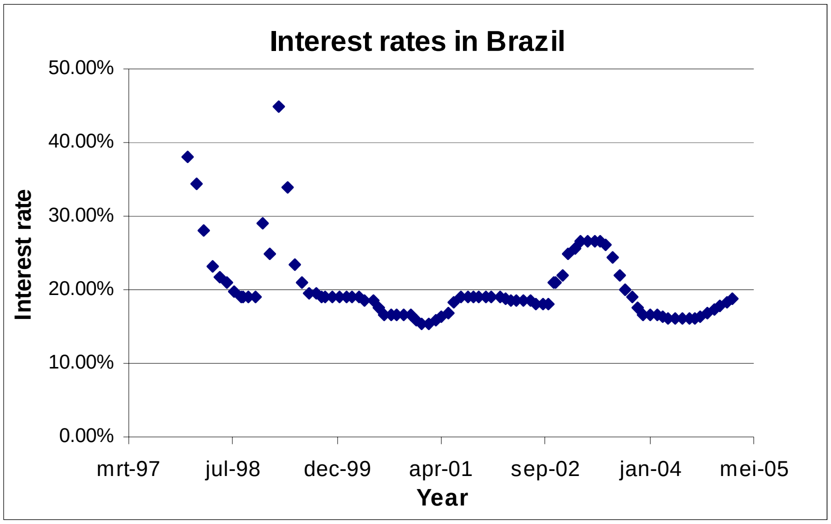


Figure 3.1 Interest rates profile in years for Brazil (source Central Bank of Brazil)

Table 3.35 Discounted cash flow analysis for 2005 plant

Year	Cost	Sales	Gross Income	Net Income	DCF
1	62.71	131.57	68.86	45.45	38.51
2	62.71	131.57	68.86	45.45	32.64
3	62.71	131.57	68.86	45.45	27.66
4	62.71	131.57	68.86	45.45	23.44
5	62.71	131.57	68.86	45.45	19.87
6	62.71	131.57	68.86	45.45	16.84
7	62.71	131.57	68.86	45.45	14.27
8	62.71	131.57	68.86	45.45	12.09
9	62.71	131.57	68.86	45.45	10.25
10	62.71	131.57	68.86	45.45	8.68
All values are in million $				**Total**	**204.25**

Table 3.36 Discounted cash flow analysis for 2015 plant

Year	Cost	Sales	Gross Income	Net Income	DCF
1	82.12	223.94	141.82	93.60	79.32
2	82.12	223.94	141.82	93.60	67.22
3	82.12	223.94	141.82	93.60	56.97
4	82.12	223.94	141.82	93.60	48.28
5	82.12	223.94	141.82	93.60	40.91
6	82.12	223.94	141.82	93.60	34.67
7	82.12	223.94	141.82	93.60	29.38
8	82.12	223.94	141.82	93.60	24.90
9	82.12	223.94	141.82	93.60	21.10
10	82.12	223.94	141.82	93.60	17.88
all values are in million $				**Total**	**420.65**

In conclusion for economical analysis, the investment on the plants should not exceed 204.3 and 420.7 million dollar for 2005 and 2015 plants, respectively, to have a profitable operation.

4. Thermodynamic Properties and Reaction Kinetics

The thermodynamic properties are obtained from the ASPEN Plus and Perry's Chemical Engineering Handbook (Green&Perry,1997). The sources other then these are mentioned in the references. The thermodynamic properties of the compounds existing in the plant are given in Appendix A.2.1.

4.1.Metabolic Reactions

The reactions relating to hexose metabolism are presented in Equations 8.8 to 8.16. The reactions relating to pentose metabolism are presented in Equations 9.7 to 9.15. And, the reactions relating to hydrolysis of the pentosans and hexosans are presented in Equations 9.1 to 9.5.

4.2.Kinetic Expressions

The kinetic expressions relating to hexose metabolism are presented in Section 8.6.2.3. And, the kinetic expressions relating to pentose metabolism are given in Table 9.18.

4.3.Property Methods

Based on ASPEN Plus 11.1 users manual the property method for streams containing alcohol is chosen as UNIQUAC property method for distillation, ethanol scrubber and fermentation gas composition calculations.

Acetic acid forms dimmers in vapor phase. ASPEN Plus user manual suggests the Hayden O'Connell equation of state for this behavior. Therefore, UNIQUAC-HOC is used as property method in designing the pretreatment collection flash tank.

4.4.Heat Capacities of the Compounds

The heat capacities of the compounds are presented in Table 4.1.

Table 4.1 The heat capacities of the compounds

Compound	Heat capacity ($kJ\ kg^{-1}{}^{o}K^{-1}$)	T Range (ºC)
Sulphur	$0.475 + 0.000840T$	0-368
CaO	$0.748 + 0.000360T$	0-900
$CaCO_3$	$0.824 + 0.000500T$	0-900
Ethanol	$2.576 + 0.00842T$	0-80
Nitrogen	$1.090 + 0.000150T$	23-4,723
Oxygen	$1.080 + 0.000034T$	23-4,723
Water (vapor)	$1.912 + 0.000035T$	23-2,223
Water (liquid)	$4.189 - 0.000600T$	
Carbon dioxide	$0.984 + 0.000260T$	23-923
Sulphur dioxide	$0.504 + 0.000350T$	23-923
NO_2	$0.157 + 0.000045T$	23-923
Molasses	$4.1868-0.0293B$*	
Sugar solutions	$4.1868-0.0251B$	
Bagasse (based on cellulose)	1.340	
Sugar crystals	1.185	
Sucrose	1,252	
Gypsum	1.084	
Dirt (based on sand)	0.800	

B is the brix of the juice

4.5.The Heat of Evaporation of Water

The heat of evaporation of water is calculated using the Watson equation which is given as

$$L_{v,T} = L_{v,B}\left[\frac{T_c - T}{T_c - T_b}\right]^{0.38} \qquad \text{Eqn 4.1}$$

$L_{v,T}$ is the latent heat of evaporation at temperature T (kj/kg), T_c is the critical temperature of water (647.3 K), T_b is the normal boiling temperature of water (373.2 K) and $L_{v,B}$ is the latent heat at normal boiling temperature (2,260 kJ/kg).

4.6.Normal Heats of Evaporation of the Compounds

The normal heats of evaporations of the compounds in vinasse are presented in Table A.9.7.

4.7.Heats of Combustion of the Compounds

The heats of combustions of the compounds are presented in Table A.9.9.

4.8.Combustion Reactions

The reactions relating the combustion of the compounds in vinasse are presented in Equations A.9.16.-A.9.26.

5. Process Structure and Description

In this section, the implementation of all information from the foregoing chapters to assess the various design criteria will be explained. The equipment and operations in the plant as well as the process flow scheme will be presented. The flow sheets for 2005 and 2015 plants can be found in Appendix 3.

5.1. Description of 2005 Plant

5.1.1. Milling

The sugar cane from fields is fed to the milling unit (M-101). The milling unit is mainly composed of cutting and shredding to cut the sugar cane in to smaller pieces and roller mills to disrupt the plant cells and to extract the cane juice by means of counter current water flow through the roller mills. Usually 5-6 roller mills are used in series that contain three rollers and another pressure roller. The output of the milling unit is sugar cane juice, which contains sucrose and the other soluble portion of the sugar cane, and bagasse, which is composed of unextracted sugar, fibers and water. In best milling practice, more than 95% of the sugar in the cane goes in to juice. Modern mills use 25-30% water for extraction of the sugar from sugar cane (Chen, 1985). The optimum imbibition is reached when the fiber to water weight ratio is 2.

Composition of the sugar cane varies from region to region. The possible intervals for the sugarcane composition are given in Table 5.1.

Table 5.1 Mass composition of sugar cane (Chen, 1985)

Millable Cane	**Cane (%)**
Water	73-76
Soluble solids	10-16
Fiber (dry)	11-16
Juice Constituents	Soluble Solids (%)
Sugars	75.0-92.0
Sucrose	70.0-88.0
Glucose	2.0-4.0
Fructose	2.0-4.0
Salts	3.0-4.5
Inorganic acids	1.5-4.5
organic acids	1.0-3.0
Organic Acids	1.5-5.5
Carboxylic acids	1.1-3.0
Amino acids	0.5-2.5
Other Organic Non sugars	
Protein	0.5-0.6
Starch	0.001-0.05
Gums	0.3-0.6
Waxes, fats, phosphatides	0.05-0.15
Others	3.0-5.0

For the calculations, the sugar cane composition is assumed as in Table 5.2. The basis for these assumptions is the values indicated in B-Basic report (2004).

Table 5.2 Assumed sugar cane mass composition

Bagasse	14%
Sugars	15%
Impurities	3%
Water	68%

The Impurity (non-sugars; soluble components other then sugar, bagasse and water content) composition is assumed as given in Table 5.3.The mineral constituent of the impurities are assumed as given in Table 5.4 based on the information taken from Chen (1985).

The most abundant organic and amino acids in sugar cane solids is aconitic acid ($C_3H_3(COOH)_3$) and aspartic acid ($C_4H_7NO_4$) , Based on this information, the molecular formula of the impurities is assumed as $CH_{1.1}ON_{0.04}S_{0.001}$ and ash content of the impurities is taken as 10.43%.

5.1.2. Cogeneration

Bagasse from mills is sent to bagasse handling unit (D-101) where the moisture content is reduced to a certain level. Then it is sent to cogeneration and boiler unit (F-103) where it is combusted as fuel for the generation of steam and electricity for plant use. The moisture content of bagasse coming out of the mill is w/w 45-55% and assumption of w/w 48% does not give significant errors (Hugot, 1972). The amount of soluble material in bagasse is given in the range w/w 2-4%, 3% is assumed for the calculations. The Efficiency of the boilers in cogeneration unit is inversely proportional with the moisture content of the bagasse. (i.e. for 52% moisture boiler efficiency is reported as 60.9% while for 32% the efficiency is 69.2%) (Hugot, 1972). Based on the same source the moisture content of the bagasse fed to the boilers is assumed as 12 w/w %.

Table 5.3 Assumed composition for non sugars (w/w % non sugars)

Salts	
Organic	13.40%
Inorganic	6.70%
Organic Acids	
Carboxylic acids	11.50%
Amino acids	5.75%
Proteins	3.16%
Starch	0.02%
Gums	2.59%
Waxes, fats, phosphates	0.57%
Soil and dirt	33.33%
Others*	22.98%

* This is composed of the vitamins, colorants, alcohols, other organic materials and minerals which are not stated in the list

Table 5.4 Mineral content of raw juice and clarified juice (w/w % Total non sugars)

Constituent	Raw Juice	Clarified Juice	% removal
Potassium (K_2O)	0.90	0.65	27.78
Sodium (Na_2O)	0.07	0.06	14.29
Sulfate (SO_3)	0.32	0.25	21.88
Chloride (Cl)	0.20	0.18	10.00
Calcium (CaO)	0.25	0.41	-39.00*
Magnesium (MgO)	0.26	0.26	0.00
Silica (SiO_2)	0.40	0.20	50.00
Phosphate (P_2O_5)	0.25	0.05	80.00
Iron (Fe_2O_3)	0.10	0.02	80.00
Sulfated ash	4.00	3.35	16.25
Conductivity ash	4.00	4.00	0.00
Total	10.75	9.52	11.44

* Increases because of the addition of CaO during clarification

The ultimate analysis of bagasse in weight basis is reported as 44-49% C, 6-7.1% H, 43.1-48% O, 0.4% N, 0.01% S and 1-2.7% ash based on different sources (Hugot, 1972). The author selected composition of (w/w) 46.8% C, 6.1% H, 44% O, 0.4% N, 0.01% S and 2.69% ash. Therefore, burning 1kg of bagasse yields, 1.716 kg CO_2, 0.549 kg H_2O, 0.013 kg NO_2, 0.0002 kg SO_2 , 0.0269 ash and requires 1.3051 kg O_2 in case of complete combustion. For complete combustion minimum amount of excess air is supposed to be supplied. On the other hand, the more air is supplied the more heat is lost with the exhaust gas from the boiler. Therefore, an optimal of 12% v/v of CO_2 in exhaust gas is recommended (Hugot, 1972). In a modern installation, it is possible to obtain flue gas temperatures below 200 ºC. Although it is possible to reach temperature as low as 130ºC, the economy of the system is not feasible. Most common furnace temperatures are reported as 1,100 ºC. The chemical composition of bagasse furnace ash is given in Table 5.5 and heat released from the bagasse during the combustion is calculated based on the data given in Table 5.6. Since water is evaporating during the combustion, its contribution to the heating value is negative. The heating value of bagasse can be calculated using the following equation

Table 5.5 Weight composition of bagasse furnace ash (Miles, 1996)

Silica (SiO_2)	44.76
Alumina (Al_2O_3)	16.99
Iron Oxide (Fe_2O_3)	13.58
Calcium Oxide (CaO)	4.29
Manganese Oxide (MnO)	4.00
Potash (K_2O)	3.99
Magnesium Oxide (MgO)	3.20
Phosphoric acid (P_2O_5)	2.61
Titanium Oxide (TiO_2)	2.53
Sulphate (SO_3)	2.00
Sodium Oxide (Na_2O)	0.73
Other undetermined	1.33

Table 5.6 Calorific values of bagasse constituents (Hugot, 1972)

Bagasse	19,260 kJ/kg
Sugars	16,559 kJ/kg
Impurities	17,166 kJ/kg
Water	-2,512 kJ/kg

$$N.C.V. = 19,260 \times f + 16,559 \times s + 17,166 \times i - 2,512 \times w \text{ (Hugot, 1972),} \qquad \text{Eqn. 5.1}$$

N.C.V. is the net calorific value of bagasse, f, s, i and w are respectively fiber, sugar, impurity and water mass fractions in bagasse. The calculated *N.C.V.* of bagasse for given weight composition (83% bagasse, 12% water, 5% sugar and non sugar) is 16,540.44 kJ/kg bagasse.

The air getting into the cogenerator is preheated by flue gasses and then the flue gasses are passed through the dewatering section (D-101) to remove the moisture content of the bagasse.

5.1.3. Juice Clarification

The dark green juice from the mills is acid and turbid with a pH of approximately 5.5. 35% of the juice is sent to ethanol production and the remaining part is feed to sugar production. These proportions are determined based on the assumption that 48% of the sucrose in the juice is converted to ethanol and the rest to sugar crystals with the contribution of the molasses from the crystallizers the total sucrose in the ethanol branch becomes 48%.

If the final desired sugar is export quality white sugar, a sulphitation unit operation is required to bleach the juice to remove the colorants. Approximately 400 ppm SO_2 (~0.0013 t/t cane) is required for the sulphitation. The sulphur dioxide is produced by burning elemental sulphur in sulphur burners (F-102) and absorbed in juice in sulphitation tower (C-101). Sulphitation can be done before liming or after liming. The important point is same final pH should be reached in the juice going to further clarification. If liming is done before sulphitation, settling is slower; volume of the mud is greater which requires greater capacity in filter press. Hot pre-liming decreases these disadvantages. On the other hand, a good coagulation requires that before the addition of lime, the sulphitation should be made at high acidity (pH 3.8-4.0) and low temperature (30-40 °C) or at low acidity (pH 5.1 5.3) and high temperature (70-75 °C) (Hugot,1972).

Following sulphitation, overliming takes place to neutralize the juice to pH 6-8. Milk of lime, about 0.5 kg CaO per ton of cane, neutralizes the natural acidity of the juice, forming insoluble lime salts. Heating the limed juice (95°C or above) coagulates albumin and some of the fats, waxes, and gums and the precipitate formed entraps the suspended solids as well as finer particles.

The lime milk is prepared in lime milk preparation vessels (V-101) by using the part of filtrate (Stream <126>) from the filters as solution media. The $Ca(OH)_2$ produced from $CaCO_3$. The lime stone is burned in calciner (F-101) to produce CaO which is later on blended with the juice from filters to prepare the lime milk. Calcite breaks down at 898 ºC and the dissociation enthalpy is 3,770-3,940 kJ/kg calcium oxide. The weight composition of limestone for sugar industry is 96-98% $CaCO_3$, <2% $MgCO_3$ and SiO_2, <0.4% AL_2O_3 and Fe_2O_3. The calciner will not be designed but the cost of investment and operation will be included in the economic calculations. Most factories have lime circulation systems, which consists mainly of two liming tanks with stirrers, in which the lime is mixed with water to desired density, usually 148 kg CaO per m^3 density, 1,116 kg/m^3 (Chen, 1985). The tanks should be stirred at a rate of 8-15 rpm. Some portion of bagasse is used as fuel in lime kiln.

After sulphitation, during liming and juice heating, polymer flocculant (Stream <136>) is added to the juice in clarification units. The role of flocculent is to improve the flocculation, increase settling rate, reduce mud volume and increase the purity of the juice. The most successful flocculating agent materials for raw sugar industry are the partially hydrolyzed polyacrylamides. The normal dosage of polymers in juice clarification is 2-4 ppm on weight of juice. The polymer is prepared as a stock solution of w/w 0.5-0.1% (Chen, 1985).

The juice is mixed with lime milk in liming tanks (V-102&103) for a 30 min residence time and then heated up to 103 ºC by steam in juice heaters (E-101&102). The hot juice is flashed upstream (V-104&105) to clarifiers to remove some of the water and air bubbles to ease the settling of the precipitate. The liquid effluents of the flash tanks are sent to clarifiers (S-101&102). The mud is separated by sedimentation in tray type clarifiers and then sent to rotary filter (S-103) unit to recover the juice. As previously mentioned some portion of the filtrate is used in lime milk preparation and the rest is sent to fermentation (Stream <125>). Water content of the mud sent to the filtration is between w/w 70-80%. The filter cake is sent back to fields as fertilizer (Chen,1985). w/w 40-50% of the impurities are removed with filter mud. The water content of the filter cake is w/w 50-60% and the sugar loss in filter cake is w/w 0.05-0.15% of the cane or w/w 0.5-1.2% of the sugar content of the cane. The amount of filter cake generated per tone of sugar cane is reported as 14-18 kg for filter presses and 27-64 kg for rotary vacuum filters.

The precipitate that is formed with the combined action of heat and liming is found to be amorphous calcium phosphate. Two stage precipitate first causes small particles to form, which grows and entraps the other non-sugars while it precipitates. X-ray diffraction studies show that the precipitates are composed of $CaSO_3$, $CaHPO_4$ and $CaHPO_4.2H_2O$ (Chen, 1985). Other than these inorganic salts, there exists a substantial amount of calcium organates which has molecular weight higher than 200. Iron and Alumina is completely removed a pH 8, but well defecated juices may contain 30 ppm Fe_2O_3 and 10 ppm of Al_2O_3. Potassium and sodium remains unchanged in clarification. Nitrogen removed by clarification is approximately one third of all nitrogenous compounds. 50% of available protein is precipitated by boiling.

5.1.4. Evaporation, Crystallization and Drying

The clarified juice (Stream <114>) is sent to multi effect evaporation section (E-201 to E-206) to increase the solid content of the juice. The evaporators are composed of 6 effects. The final effect pressure is set to 16.6 kPa. The sugar concentration is evaporation section is allowed to rise to w/w 68 %. Typical values for commercial plants are w/w 65-68% sugar. The condensates from the evaporators are sent to cooling tower to be cooled down and recycled as process water. The cooling tower design is out side the battery limit, but, the cost will be included in investment calculations. The last effect vapor (Stream <228>) is condensed in vacuum generators and taken outside the battery limit as condensed water.

In next step, the concentrated sugar solution is sent to vacuum boiling pans (E-207&E-208). The crystallizer is fed with the crude sugar from the second stage crystallizer to supply the seed for the crystallization of the sugar. The residence time in the pans are 3 and 4 hours respectively for E207&208. The concentrated juice (~80%) form the vacuum pans are send to crystallizers (V-201&V-202). The crystallization is carried out by cooling the oversaturated juice to 40ºC. To maintain continuous operation the pans E-207 and E-208 are supplied with 3 and 4 crystallizers respectively. The crystals are separated from the molasses by means of centrifuges (S-201&202). The molasses from the first pan (Streams <204>) is feed to second pans and the molasses from the second pan (Stream <220>) is feed to fermentors. The solid content of the molasses from the centrifuge varies between w/w 60-70%. The moisture content of the crystal sugar from the centrifuge is less than w/w 1% (Chou, 2000). The sugar recovery in each crystallizer is 60% and total sugar recovery is 84%.

The juice from the evaporators (Stream <201>) is at 56ºC and is heated to 75 ºC in first reheater (E-209). The effluents of the crystallizers (Streams <210> and <217>) are reheated to 55 ºC in E-

210&212 to decrease the viscosity and ease the centrifugation. The molasses from the first centrifuges (Stream <213>) are reheated to second pan operation temperature of 75ºC in E-211.

Crystalline sugar from centrifuge is send to driers (D-201). Rotary granulators are used for drying. Wet sugar with w/w 1% moisture enters a granulator counter current to air flow and exit at 50° with a moisture content of w/w 0.05%. Air enters and exits at 85 and 42°C respectively. Sugar also looses some of its moisture in following 24 hours after drying in silos by the help of conditioned air (Chou, 2000). The final sugar contains w/w 0.04% ash and w/w 0.02% moisture. The air is heated in air heaters E-213.

Table 5.7 Assumed weight composition of soluble materials in molasses (% of non sugars)

Carbohydrates (Gums, pentosans, starch, etc.)	10.52%
Ash (K_2O, CaO, MgO,Na_2O, SO_3,Cl,P_2O_5, SiO_2)	46.32%
Nitrogenous compounds	
Crude protein	12.63%
True Protein	3.16%
Amino acids	1.26%
Undefined nitrogenous	8.42%
Non nitrogenous organic acids (Aconitic, citric, malic, oxalic, glycolic, mesaconic, succinic, fumaric, tartaric)	16.84%
Waxes sterols and phosphates	0.84%

5.1.5. Fermentation

The clarified juice (Stream <120>) is mixed with filtered juice from filters (Stream <125>), and molasses from crystallization (Stream <220>). The temperature of the combined stream is 96 ºC and is to be cooled to fermentation temperature. The cooling of the broth is carried through in two stages: the hot broth is passed through an heat exchanger (regenerative, E-301) in countercurrent with the cold mixing broth, where the mixing broth is warmed and the broth for distillery is cooled. Final cooling, approximately 30 Cº, is carried out in exchangers E-302 using countercurrent water.

During the fermentation, sugar is converted to ethanol with 0.92 of the theoretical yield which corresponds to 0.46-0.47 g ethanol/g sugar. The remaining substrate is utilized by the microorganisms for maintenance, cell growth and production of by products such as succinic acid, glycerol, acetic acid, higher alcohols and acetaldehyde. Also, 0.44-0.48 g CO_2 / g sugar is produced during the ethanol fermentation. *Saccharomyces* species is used as the ethanol fermenting microorganism. The fermentation is carried out anaerobically to decrease cell growth and increase the ethanol productivity.

The fermentation is carried out in 6 vessels to maintain a continuous operation and to prevent requirement of extra storage tanks for broth and recycle biomass. The biomass is recycled to maintain high cell concentrations in the fermentors. The biomass is separated from the broth using continuous disc type centrifuges (S-301) and sent to biomass sterilization vessel (V-302), which are stirred vessels to blend the biomass with pH 2 acid solution. The residence time in the vessels are set to 15 minutes. After the sterilization vessel, the biomass is separated from the acid solutions using centrifuges S-302. The acid solution is taken outside the battery limit without treatment, but, the cost of waste treatment is included in the cash flows. The waste treatment methods are described in Section 11. The sulfuric acid requirement for this step is approximately 0.013 kg per kg ethanol produced. Since, yeast functions at low pH values they are not affected by the pH treatment. The biomass is recycled to fermentors. By recycling the cells, the long lag growth phase is eliminated and high cell concentrations are obtained in fermentors. Very short fermentation times (6-10 h) allows yeast to be recycled up to three times a day for as long as 200 days (Wheals *et al.*,1999). The fermentation parameters are given in Table 5.8.

Table 5.8 Typical fermentation parameters (Goldemberg, 1994)

Ethanol yield	91.4% of theoretical
Productivity	8 g/l.h
Yeast concentration	34 g/l
Yeast viability	90%*

* (Smart, 2000)

The carbon dioxide rich gas (Stream <303>) from the fermentors are passed through the ethanol scrubber column (C-301) to absorb the ethanol in the gas to lower the volatile organic emissions to atmosphere and to recover the lost ethanol. This column is not only feed with the gas from the fermentor, but, also with the carbon dioxide rich vapor distillate (Stream <S-312>) from the beer column. The absorption water is recycle water form the evaporators and boiling pans. The water and ethanol content of the vent gas is 1-2 w/w %. Carbon dioxide can be sold either to beverage and food industries or to dry ice production facilities. But, considering the large amount of carbon dioxide generated and the comparatively small market size available for carbon dioxide, there will not be enough demand for carbon dioxide. Therefore, carbon dioxide will be released to atmosphere. The water scrubber can remove 99.5% of ethanol in carbon dioxide stream. The environmental legislations set the upper limit of total volatile components to the atmosphere as 36.3 T/year (Alan *et al.*, NREL report, 2002). Therefore, the total ethanol released will be set accordingly.

5.1.6. Distillation, Rectification and Dehydration

The fermentation medium from the fermentor (Stream <306>) is mixed with the ethanol solution from the ethanol scrubbers (Stream <307>) and is sent to down stream processing units to obtain dehydrated ethanol. Ethanol separation form the wine is accomplished by conventional system using two sets of distillation column and one dehydration step which uses molecular sieves.

In beer column (C-302), ethanol is separated from the wine (initially at ~8%) as a vapor (~40%). The product is removed as a vapor stream from the 6th stage of the column. The carbon dioxide rich distillate (Stream <312>) is taken out the tower at temperature of 20ºC to reduce the amount of ethanol present in the vapor. The bottom product, stillage, has most of the salts (mainly from K) and suspension solids. But, most important of all is the significant amount of glycerol, acetic acid and fusel oils present in the bottom product. To recover the fusel oil and glycerol, another side draw (Stream <313>) is installed to column, one stage above the reboiler, to separate most of the fusel oil and glycerol from the vinasse (Stream <311>). Vinasse (the stillage) is entirely used as fertilizer; its ethanol content must not exceed 0.03%. Heating is accomplished by low pressure steam. In rectification column (C-303), the product from the beer column is concentrated to 92.5 %, and some impurities are removed. Bottom products (Stream <318>) are removed with the side draw of the beer column outside the battery limit to be further purified. The ethanol recovery in rectification column is 99.99%.

The final dehydration of ethanol is carried out in molecular sieve absorption columns (C-304). The distillate from the rectification column (Stream <314>) is taken out as liquid and cooled to adsorption column operation temperature of 60ºC in E-307. then feed to columns. The adsorption columns are working in tandem. The flow is passed through one of the columns for period of 2 h 10 minutes. After saturation, the flow is shifted to the other column. The columns are regenerated by using hot nitrogen at 200ºC. The nitrogen is circulated in the system. The wet nitrogen from regeneration (Stream <320>) is cooled to 30 ºC in E-308 and then sent to condenser (E-309) to condense the water to dry the nitrogen. For condensation refrigeration at 0ºC is used. E-308 is acting as an heat exchanger between the dry nitrogen (Stream <324>) and wet nitrogen. The dry nitrogen is heated up to 180 ºC before being send to adsorption column.

Molecular sieve dehydration technology utilizes micro-porous particles such as alumino-silicates, processing a very precise pore size. The pores make it possible to separate small molecules form large ones through selective adsorption. For example, ethanol dehydration is accomplished with molecular sieves which have a diameter of 3Å, which entraps water molecules which have a diameter of 2.5Å. Ethanol molecules which have a diameter of 4Å cannot enter. Therefore, flow around the material. Azeotropic distillation, extractive distillation and membrane separation techniques can also be applied for dehydration. However the operation costs are higher than molecular sieve dehydration.

5.2. Description of 2015 Plant

The 2015 case process is constructed based on current 2005 technology. As in 2005 case, 48 % of sucrose will be utilized for ethanol production and 52% will be crystallized into sugar crystals. On the other hand, in future case, the bagasse, lignocellulosic portion of sugar cane, will be hydrolyzed to pentose and hexose sugars and lignin. In addition, the organic waste generated in the process (i.e. vinasse, filter mud, microorganisms) will be also be combusted in cogeneration units instead of sending to fields untreated. The descriptions of the milling unit, the sugar section and downstream process of ethanol have the same structure with the 2005 plant with some variations in equipment sizes.

5.2.1. Lignocellulose Properties

The sugar cane bagasse is mainly composed of hemicellulose, cellulose and lignin. Cellulose is a linear crystalline polymer of β-D-glucose units. Hemicelluloses are heterogeneous linear and branched polymers of pentoses (D-xylose and L-arabinose), hexoses (D-galactose, D-glucose, D-mannose) and sugar acids (methyl and acetyl groups). Commonly occurring hemicelluloses are xylans, arabino-xylans, gluco-mannan, galacto-glucoman (Saha, 2003). Lignin is a complex, variable, hydrophobic, crosslinked, three dimensional aromatic polymers of *p*-hydroxyphenylpropanoid units connected by C-C and C-O-C links and they are extremely resistant to chemical and enzymatic degradations.

The process of converting the lignocellulose to ethanol requires delignification to liberate cellulose and hemicellulose from lignin, depolymerization of the carbohydrate polymers to produce free sugars and fermentation of hexose and pentose sugars to produce ethanol (lee, 1997). The content of lignocellulosic biomasses is different for different biomass sources. The reported different compositions for sugarcane bagasse are given in Tables 5.9-10. The assumed composition for sugar cane bagasse are given in Tables 5.9-5.10 as the design case values and the calculations will be carried out based on this assumptions.

Table 5.9 Composition of sugar cane bagasse

Reference	Cellulose (%)	Hemicellulose (%)	Lignin (%)	Others (%)
Brown, (2003)	35	25	20	20
Vacceri, (2004)	39	32	26	3
Chen, (1985)	38-43	27-32	10-23	0.6-12
Saha, (2003)	40	24	25	11
Design Case	40	30.8	25.2	4

Table 5.10 Composition of sugar cane bagasse

Reference	Glucan	Xylan	Galactan	Arabian	Lignin	Mannan	Extractives	Ash
Wyman, (1996)	40.20	21.10	0.50	1.90	25.20	0.30	4.40	4.0
Neureiter,(2002)	40.19	22.54	1.40	2.00	25.15	0.48		
Aquilar, (2002)	28.90	20.60	5.56		23.90			
Design Case	41.20	21.50	0.9	1.90	25.20	0.60	4.70	4.0

5.2.2. Pretreatment

The first step of the bagasse hydrolysis is pretreatment (R-401). This step is carried out to break the crystalline structure of lignocellulose, hydrolyze the easily hydrolyzing hemicellulose and increase the surface area for the cellulase enzyme to penetrate in to cellulose. Acid catalyzed steam explosion is used for the pretreatment. The conditions of steam explosion pretreatment of sugarcane bagasse is assumed as 190°C, 120 s residence time, water to solid ratio 2 and 1% H_2SO_4 (w/w dry bagasse). The recommended design for the pretreatment reactor is plug flow reactor. The fibers are first heated up to 100 ºC in steaming vessel and then mixed with hot acid solution (100ºC) and pumped in to the pretreatment vessel using high pressure feed pump. The high temperature effluent of the pretreatment reactor is flashed in down stream flash tank (V-401) to remove some of the volatile inhibitory compounds.

The liquid portion is removed from the fibers via screw presses (S-401) and passed through cooler (E-401) to cool the hydrolysate to overliming temperature of 50ºC. Before the detoxification, some of the hydrolyzate (Stream <409>) is bypassed to readjust the pH of the hydrolysate from 10 to 4.5. The fiber portion (Stream <421>) is transferred to cellulose hydrolysis vessels (V-404). The wash water for the screw presses are the portion of the conditioned hydrolysates (Stream <417>). The reason for this is to prevent the dilution of the hydrolysate and to prevent high reactor sizes due to the dilution.

5.2.3. Detoxification

The reason for the detoxification step is to remove the pretreatment by-products that can be inhibitory to fermentation. Since the fermentation is carried out at high cell densities the inhibition on growth will not be an obstacle. Moreover, most of furfural and HMF are removed in flash tank downstream of the pretreatment. The major inhibitory compound in pretreatment effluent is acetic acid. But, the acetic acid resistances of the microorganisms are expected to rise in close future and at pH of 4.5 to 5 the acetic acid concentrations of 2-3 g/l already does not have inhibitory action on fermentation (Personal Communication Marco Kuyper, Musatto *et al*., 2004). By diluting the hydrolysate with molasses clarified juice and filtrate from the filters the concentration of all the inhibitory compounds will be halved. The final concentration of lignin degradation products going to fermentors after overliming is obtained as 1.4 g/l (Super Pro mass balances). The experimental studies show that lignin monomers in untreated hydrolysate is composed of mainly 4-hydroxy benzoic acid, vanillin and catechol and this compounds does not have inhibitory effect up to 1 g/l concentration (Palmqvist and Hahn-hagerdal, 2000). Considering the high cell concentrations, 1.4 g/l will not have inhibitory effect on the fermentation, thus, the detoxification is not considered for the current case and the only detoxification will occur during the overliming. Several authors reported the enhancing effect of overliming on the fermentation performance by partial precipitation of phenolic compounds (Mussatto and Roberto, 2004).

Therefore, the detoxification is carried out by overliming to pH 10 in overliming tank V-402. This vessel is stirred tanks with volume that enables 30 min residence time. The lime milk is prepared in lime milk preparation tanks as described in Section 5.1.3. The limed hydrolysate is transferred to conditioning vessel (V-403) to readjust the pH to 4.5 to enhance the precipitate formation. As described before the pH is adjusted using the bypass hydrolysate stream (Stream <409>). The juice from V-403 is send to clarifiers (S-402) to remove the precipitates. The clarifiers have the same characteristics with juice clarifiers (S-101&102).

The mud from the clarifiers are send to rotary filter (S-403), there the cake is dehydrated to 55% water. The filter cake (Stream <416>) is sent to cogeneration due to the presence of the organic material inside the cake. The filtrate (Stream <415>) is remixed to hydrolyzate (Stream <413>). As previously mentioned some portion of the hydrolysate is utilized as wash water for the screw presses (S-401). And the remaining portion is sent to heater E-402 to heat up the hydrolysate to 54.4 ºC. The optimum temperature of hydrolysis vessels is 65 ºC. The reason of heating up to 54.4 ºC instead of 65 ºC is the high temperature of the fibers from the screw are able to heat up the hydrolysate to 65ºC.

5.2.4. Cellulose Hydrolysis

The detoxified liquid fraction after pretreatment contains the pentose sugars and this stream is used to dilute the fibers to 20.1% solids. The solid fraction from pretreatment contains the cellulase and the lignin fraction. This solid fraction is exposed to cellulase enzyme to break the cellulose in the hexose sugars. The cellulose hydrolysis vessels (V-404) are composed of 2 parallel vessel sets of 4 vessels. The reason for 4 vessels is to mimic plug flow behavior. The conversion of cellulose to glucose is 90 % total in four vessels. Due to the high residence time of 36 h, the flow is split to two and two sets of vessels are used to prevent large vessel volumes.

After cellulose hydrolysis, lignin residue (Stream <424>) is separated from the mixed sugar solution (Stream <423>) using screw presses (S-404) and sent to cogenerator to be combusted to generate energy and the liquid fraction which contains sugars is sent to fermentation (V-301) to produce ethanol. Similar with 2005 plant, 7 fermentors are utilized for fermentation. After the fermentation the downstream processes has the same structure with the conventional ethanol plant.

5.2.5. Waste Treatment and Cogeneration

The waste streams generated in the process are the filter mud (Streams <123> and <415>) from rotary filters (S-103 and S-403) containing sugar cane organic ingredients, vinasse (Stream <311>) from distillation containing fermentation by products and sugar cane organic material, yeast from fermentation and water from dehydration, evaporator, crystallizer and drying containing volatile organics and detoxification water containing the toxic organics. The waste water (condensates and vapors from evaporators, boiling pans, detoxification and dehydration) are treated by water treatment outside the battery limit, but the costs are included in cash flows.

The vinasse is concentrated in evaporator (E-403) and sent do cogeneration (F-401) as fuel. The lignin residue (Stream<424>), yeast (Stream<328>) and the filter mud (Streams <123> and<415>) are dehydrated in dewatering unit (D-401) using the flue gasses (stream <436>) from the cogenerator before being sent to furnace (F-401). Since the steam generation and electricity generation are not sufficient to meet the demand of the plant, the electricity generation is lowered and the steam generation is increased to meet the demand. The excess electricity demand is supplied by purchasing from the grid.

The solids from the furnace (ash, gypsum and dirt (Stream <434>)) are sent to land fill. The acid solution from the yeast sterilization unit (Stream<332>) is neutralized using lime and precipitated to be treated as land fill. The unit is outside the battery limit. But, the cost is included in the cash flows.

5.3. Process Flow Schemes

The process flow schemes of 2005 and 2015 plant can be found in Appendices A.3.1 and A.3.2. respectively.

5.4. Process Stream Summary

For process stream summaries for 2005 and 2015 plants are presented in Appendices A.4.1 and A.4.2 respectively.

5.5. Utilities

The utilities used in the plant are saturated steam, cooling water, chilled water, refrigeration and electricity. Saturated steam is needed to supply heat to the system. Cooling water is used for cooling. Power is needed for the pumps and electricity inside the plant. For the utility summaries see Appendices A.5.1 and A.5.5.

5.6. Process Yields

The incoming and out going streams for 2005 and 2015 plants are as presented in Figures 5.1 and 5.2.

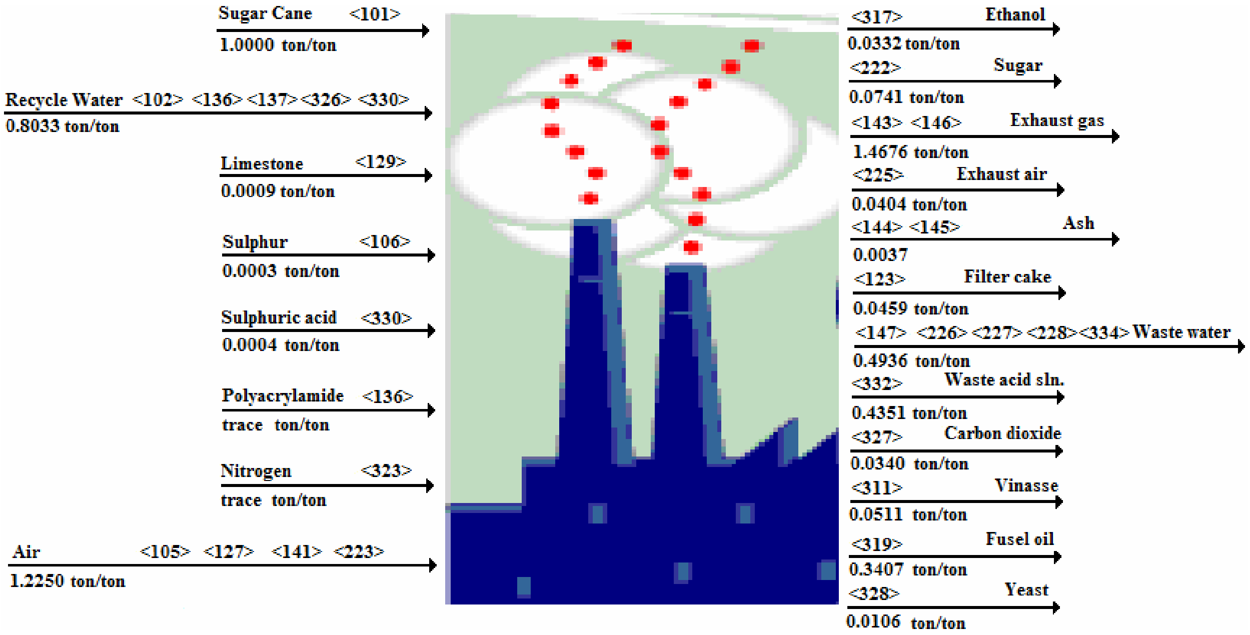


Figure 5.1 Process yields for 2005 plant (flows are given per ton of sugar cane)

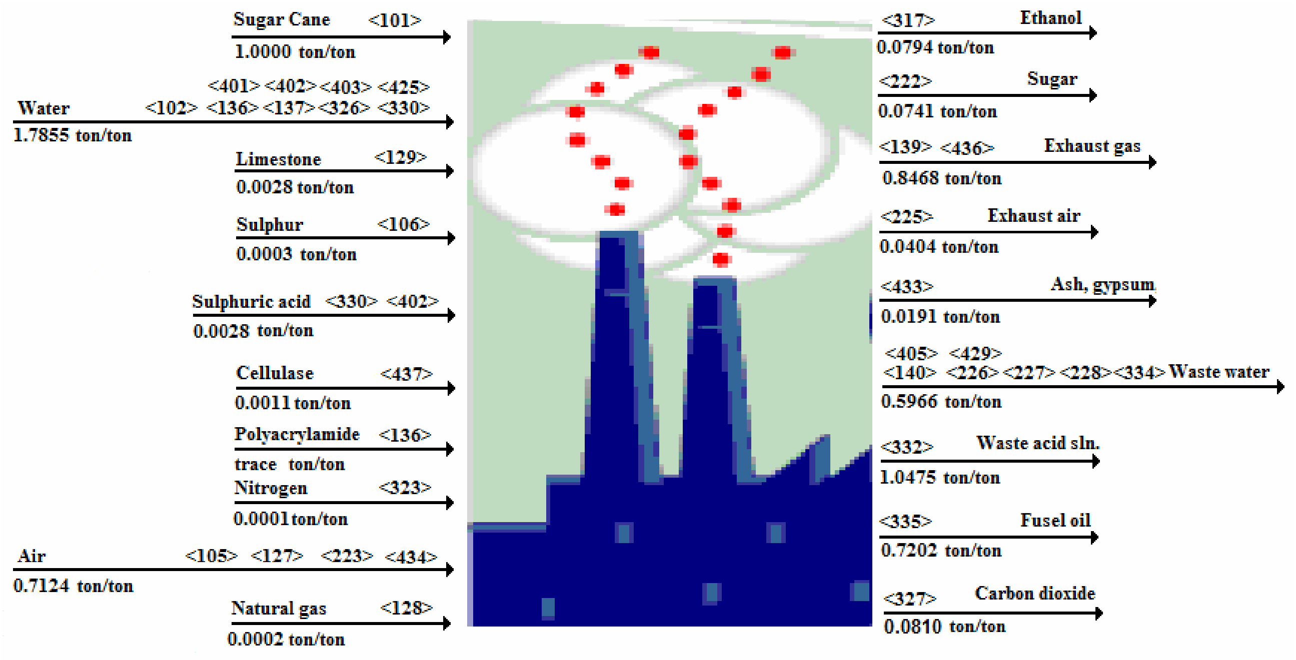


Figure 5.2 Process yields for 2015 plant (flows are given per ton of sugar cane)

6. Process Control

This chapter explains why certain combinations of control loops are chosen and positioned from a process point of view and how these choices influence the design. Since the 2015 plant covers the 2005 plants equipments only 2015 plant's control is described. As described previously, the complete process is divided into four sections: mill and clarification, juice evaporation and sugar crystallization, ethanol production and purification and hydrolysis sections. The control flow sheet of the whole process is shown in Appendix 6.

6.1. Flow Controllers (FC)

- Raw materials: Sugar cane (Stream<101>), limestone (Stream<129>), sulphur (Stream<106>), polyacrylamide (Stream<136>), cellulase (Stream <437>) and sulphuric acid (Streams <330> and <402>) should be weighted to maintain the process specifications before being send to system.
- Air Streams (Streams <105>,<127>, <223>,<434>), to prevent heat losses by excess air or incomplete combustion with less air flow, the air streams flowing into the furnaces or driers are to be controlled with flow controller to maintain the required air. Stream 223 is to be controlled to provide complete drying of the sugar crystals.
- Wash Water for Mills, filters and screw presses (Streams <102>, <127>, <425>), the washing water flow for mills is to be controlled to prevent any dilution with high flow rates or sugar losses with low flow rates
- Steam flow rates (Streams <401> and <403>)
- Distillate, bottoms and side products of the distillation columns (Streams <310>, <311>, <312>, <313>, <314>, <318>).

6.2. Level Controllers (LC)

All the vessels with continuous inflow require a level controller to prevent any flooding in case of any problem with pumping in/out. The level controllers are operated based on the liquid height in the vessel and the manipulated variable is the output flow from the vessel. The controller type is to be "fail-on" to prevent the overflow of the vessels in case of controller failure. The equipments that requires level controller are listed below.

- Liming Vessels (V-102 &103)
- Evaporators and boiling pans (E-201, E-202, E-203, E-204, E-205, E-206, E-207, E-208 and E-403)
- Overliming and conditioning Vessels (V-402&403)
- Cellulose hydrolysis bioreactors (BR-401)
- Crystallizers (V-201&202)

6.3. Ratio Controllers (RC)

The ratio controllers are necessary when the stream split into two or more streams. The streams that require ratio controller are

- Main Juice (Stream <103>), splits into <104> and <115>.
- Lime milk (Stream <132>), splits into <133>, <134> and <142>.
- Filtrate (stream <124>), splits into <125> and <126>.

- It is essential for the proper operation of the distillation columns to have constant reflux and boil up rates. The reflux and boilup ratios of the distillation columns are also adjusted with ratio and flow controller.
- The yeast bleed (Stream <328>).
- Conditioned hydrolysate (Stream <417>), splits into <418> and <419>.
- Stream <408> splits in to <409> and <410>.

6.4. Temperature Controllers (TC)

The equipments involving heat transfer are to be supplied with temperature controller to maintain constant operation conditions. The equipments and streams that require temperature controller are listed below.

- Evaporators and boiling pans (E-201, E-202, E-203, E-204, E-205, E-206, E-207, E-208 and E-403)
- Pretreatment reactor (R-401): The temperature is manipulated by steam input of the pretreatment. The controller should be fail-close control valve to prevent steam inflow in case of controller failure.
- All Heat exchangers
- Reboilers and condensers (E-303,E-304,E-305 and E-306)
- Fermentors (BR-301)
- Lime milk preparation tanks (V-101)
- Molecular sieve column (C-304).

6.5. Pressure Controllers (PC)

The equipment that requires a certain operation pressure other than atmospheric requires pressure controller. The equipments that require pressure controller are listed below.

- Evaporators and Boiling pans (E-201, E-202, E-203, E-204, E-205, E-206, E-207, E-208 and E-403), the pressure of the multi effect evaporators and the boilers should me monitored to provide the necessary vacuum for evaporation.
- Pretreatment reactor(R-401), the pressure in the pretreatment reactor should be monitored to keep the reactor pressure constant at set point. Reduction in pressure would cause evaporation of the water and affect the process yields and increase in pressure would cause leakages or explosions. The pressure should be controlled a vent valve. And, the controller should be fail to open valve to make sure that the vent valve is full open in case of any controller failure.

6.6. pH Controller (pHC)

The pH of the cellulase hydrolysis and fermentation bio reactors should be monitored to maintain the pH in optimal value. The pH can be manipulated with small portion of the pretreatment reactor effluent hydrolyzate.

7. Mass and Heat Balances

The process simulation is performed with the program SuperPro Designer v5.1, MS Excel and ASPEN Plus 11.1. The data for the mass balance were taken from the simulation files of these programs. With the program Microsoft Excel the heat balances were calculated based on the mass balances and the heat capacities of the components and the temperature of the stream.

The mass and heat balances are presented in Appendix 7.1&2. The mass and energy flow rates of the streams are given in kg/s and kW. The mass balance has been accurately solved, but, the heat balance has 0.3 to 1% error. This difference is introduced due to the round up in temperatures and using different tools for the design. The integration procedure of the MS Excel, SuperPro designer and ASPEN Plus introduced noticeable errors, i.e. during the simulation in ASPEN Plus 11.1 the sugar cane impurities are not defined in the system, this is why the temperatures given deviates from the real life case. Also, analysis shows that for large flow rates of juice, variation in second decimal digits also introduces noticeable errors in overall energy balances. Considering that all the temperatures are reported with one decimal digits, the errors generated can also be accounted for the round ups.

The reference temperature for the enthalpy calculations is taken as 30ºC.

8. Process and Equipment Design: 2005 Plant

Tools used in this process are Super Pro Designer simulation program v5.1, ASPEN Plus 11.1, MATLAB, Sugar engineers online calculation tools (IRef-1) and MS Excel spread sheet calculation. Most of the equipment design is carried out in MS Excel and ASPEN Plus and the parameters obtained from these calculations are implemented into Super Pro Designer to simulate the process. The selection and design of the 2005 plants is described in following sections.

The design of the heat exchanging equipment is carried out as described in Appendix A.8.13, unless specified in other method. The design of the pumps is carried out in Super Pro designer based on the head requirements.

All the units used in this chapter are metric units. The definitions of the technical terms relating to sugar production can be found in Definitions section

8.1. Mill (M-101)

The design of the milling equipments is outside the scope of the design. In this section, the dimension of the equipments that are required to process 5,000,000 ton of sugar cane per year will be given.

8.1.1. Cane Carriers

Cane carrier is a moving conveying belt that carries the cane into mills. It is composed of a horizontal portion and an inclined portion to elevate the feedstock to the mill feed location height. The length of the inclined portion of the carrier is 13.7-17 meter, and that of the horizontal portion is calculated using the equation (Hugot, 1972),

$$L_H = 4.877\sqrt[3]{A_C} \qquad \text{Eqn. 8.1}$$

L_H is the horizontal length in meters and A_C is the cane processing capacity in metric tones per hour. As it is presented in Section 8.2.1.6, the number of sugar mill tandems is 5, each having capacity of 240 ton cane per hour (t.c.h.). Using this capacity the horizontal length requirement for each carrier is calculated as 30.3 m.

The width of the carrier is adjusted to be the same as the mill width. The power requirement is calculated based on the following equation (Hugot, 1972),

$$P_i = \frac{Z_t + A}{10} \qquad \text{Eqn 8.2}$$

P_i is the installed power in horse power; Z_t is the total carrier length. The power requirement for each carrier is calculated as 20.2 kW.

8.1.2. Cane Knives

Cane knives assist the extraction in the mills by breaking up the cane in to smaller pieces. The important parameters relating to cane knives are the separation distance and number of blades, clearance, number of knife sets, speed of rotation and power requirement.

The pitch is the interval separating to adjacent blades and usually it is 5.08 cm (2 in). Number of blades is calculated using the following equation (Hugot, 1972),

$$N = \frac{L}{p} - 1 \qquad \text{Eqn 8.3}$$

N is the number of knives, L is the width of the carriers and p is the pitch of the knives.

The clearance of the knife set is the separation between the tip of the blades and the plane passing below the knives. Hugot (1972), report the clearance for the first knife set as 30.5 cm, 0.95 cm for second set and 0.64 cm for the third set. Currently, so few factories uses 3 sets of knives. The common number of knife sets is 2.

Speed of rotation of knife sets are reported as 400-750 rpm. The speed most frequently used is 500 rpm (Hugot, 1972). The recommended installed power requirement for engines is 2.6 kW/t.c.h. for each of the blade sets (Chen, 1986). Therefore, the total power requirement is calculated as 6,250 kW.

8.1.3. Magnetic separators

This equipment is installed to the end of the shedders to separate any metallic material before they are transferred to mills. An electromagnet is installed across the width of the chute feeding to crusher. The design of this equipment is out side scope of the design. But the cost will be included in economic calculations.

8.1.4. Shedders

Shedders tear the cane chips from the cane knives into shreds without extracting any juice. There are different types of shedders.

The Maxwell shedders consist of a steal cylinder 64 cm in diameter and length equal to that of mill rollers. It is furnished with number of longitudinal slots (generally 10) which are inserted with teeth 10-13 cm in length. Thus, the total diameter is about 75 cm over the tip of the teeth. There is approximately 0.32 to 0.48 cm separation between the teeth.

The Maxwell shedder is installed to the delivery side of the crusher. The speed of rotation is 450-600 rpm and, the mean power absorbed is about 0.4 kW / t.c.h. But, it is better to install a motor that can develop 0.45-0.52 kW / t.c.h. A set of teeth has a working life of 2,000-2,600 h.

Another type of shedder is Searby hammer mill shedder. This equipment can replace two rollers crusher. They are made in one diameter (1.07 m) and their lengths are between 0.92 and 2.14 m. They rotate at a speed of around 1,200 rpm. The power requirement is 1.7-2.3 kW/t.c.h. They have capacities up to 450 t.c.h.

8.1.5. Crushers

Hugot (1972) recommended using a hammer shedder to replace the 2 roller crusher and to install a mill crusher (three roller crusher) after the hammer shedder. A mill crusher is installed as a initial mill to the mill tandem. The speed of the mill crushers is 4.5 rpm. The pressure is the same with the heavies loaded mill in the mill tandem. The power requirement is determined based on an ordinary mill. However, it is increased by 20-25% due to the extra work required to break up the harder more resistant material.

8.1.6. Milling Machinery

The classical combination of three rollers arranged in triangular form is the standard milling unit for the industry. From 3 to 7 sets of three roller mills, described as 9 to 21 roller mills, are in use in industry. The most common ones are 15 to 18 roller mills. The rollers are 0.61-1.07 m in diameter and 1.22-2.14 m in length. The capacity of the mills can be calculated using the following equation,

$$A_C = 0.8 \times \frac{c \times n \times (1 - 0.06 n \times D) \times L \times D^2 \sqrt{N}}{f} \qquad \text{Eqn 8.4}$$

A is the mill capacity (t.c.h.), *c* is a coefficient which is equal to 1.25 for two sets of knives followed by a Searby shedder, *n* is the speed (rpm), *f* is the fiber fraction, *L* is the length (m), *D* is the diameter (m) and *N* is the number of rollers. Using diameter of 1.07 m, length of 2.14 m, fiber content of 0.14, 5 rpm speed and 18 rollers (including the mill crusher), the capacity of the tandem is calculated as 250 t.c.h. therefore to supply the capacity of 1,197 t.c.h. capacity 5 tandems are required. The mill parameters used here are taken from Hugot (1972).

Since the width of the mills is set as 2.14 m, the lengths of carriers and cane knives and shedders are also set to same value.

The pressure in the mill is given as 98-112 kg/ cm^2(Chen, 1985). The power consumption of whole mill tandem is given as an average value of 18 kW per ton fiber per hour per mill tandem composing of 6 mills (Hugot, 1972, IRef (Internet Reference) 1). Therefore, the power consumption of mill is obtained as 6,234 kW.

8.1.7. Intermediate Carriers

The mill tandems require fiber carriers in between the mills to transfer the fibers from one mill to other. The width of the carriers is the same with that of mills. Power requirements are expressed as average of 5% of the power requirement of the mill for slat type carriers. The design of these equipments is outside the scope of the project, but the capital investment and power requirements will be included during the calculations.

8.1.8. Bagasse Carrier

Bagasse carriers also have the characteristics similar to intermediate carriers. The power requirement is 122 W per meter total length of carrier. The length is dependent on the positioning in the plant. Therefore, an estimate of 30 meter is used for the length of the belt; therefore, the total length is 60 (upper and lower) and the corresponding power consumption is 12.2 kW.

8.2. Juice Clarification

8.2.1. Lime Kiln (F-101)

In a lime kiln, the lime stone is decomposed into carbon dioxide and CaO at high temperatures. At the temperature of decomposition, (898 ºC) theoretical heat requirement is 3,770-3,940 kJ/kg CaO (van der Poel *et al.*,1998). This heat can be obtained by burning bagasse in kilns. Capacity of the lime kilns are given as 15.4 to 51.4 kg $m^{-3}h^{-1}$, higher margin is selected as the design specification. For the current limestone processing capacity of 1,091 kg/h, the required operating volume for the kiln is calculated as 24 m^3 (including 10% safety).

The details of the calculations are given in Appendix A.8.1. And, the amount of bagasse required to supply the necessary heat is found as 222 kg bagasse/h (12% moisture). The exhaust gas will be cooled down to 320 ºC by heating up the feed air. The area required for heat exchanging is calculated as 75.5 m^2.

Table 8.1 Design specifications for F-101

Parameter	Unit	F-101
Number in series		1
Streams in		<127><128><129>
Streams out		<130><131><145>
Capacity	kg/h	1,091
Volume	m^3	24
Bagasse requirement	kg/h	222
Heat duty	kW	1,015
Feed air heater area	m^2	75.5
Operating pressure	atm	1
Operating temperature	ºC	898

8.2.2. Sulphitation

As it is previously described the sulphitation is applied when the desired final product is white sugar. With the presence of lime the sulphur dioxide precipitates as $CaSO_3$. Sulphur dioxide can be supplied in liquefied form or in gaseous form by burning sulphur. The Latter is economically more feasible and applied by most of the plants worldwide. Therefore, the sulphur dioxide will be obtained in sulphur burners.

8.2.2.1. Sulphur Furnaces (F-102)

The sulphur dioxide for the white sugar production is a gas generated by burning sulphur. 1 kg of sulphur requires 1 kg of oxygen and the exothermic reaction releases 9,281 kJ/kg of sulphur. Combustion takes place at 363ºC (Hugot, 1972). The theoretical minimum air requirement for complete combustion is 4.3 kg/kg Sulphur. However complete combustion is not possible with theoretical minimum. 100% excess air is recommended and the sulphur dioxide fraction in off gas is on average 14%. The sulphur consumption is reported as 0.5 kg per ton of cane.

In the presence of water the acidity of the media will be high and will cause corrosion. Therefore, it is necessary to use dry air for combustion. For this purpose the air is passed through lime to absorb moisture.

The capacities of the sulphur furnaces are given as 24.4 kg sulphur m^{-2}h^{-1}. Based on this information, the total furnace area required to burn 1,625 t sulphur /year (389 kg/h) is obtained as 16 m^2.

To check whether any heat integration is required for the incoming and outgoing streams the same methodology is applied with the lime kiln. Calculations are shown in Appendix A.8.2. The results indicate that the heat generated by the reaction is enough to heat up the inflowing sulphur and fresh air to combustion temperature. Therefore, air and sulphur can be directly sent to furnace. The exhaust air leaves furnace at 741ºC. The exhaust air does not need to be cooled because it is going to get in contact with juice in sulphur dioxide absorption column and it heats up the juice.

Table 8.2 Design specifications for F-102

Parameter	Unit	F-102
Number in series		1
Streams in		<104><105><106>
Streams out		<107><135>
Capacity	kg/h	389
Heat duty	kW	1,003
Furnace area	m^2	16
Operating pressure	atm	1
Operating temperature	ºC	741

8.2.2.2. Sulphitation Tower (C-101)

The sulphitation tower is operating to absorb the SO_2 generated in sulphur burners in juice from the mills. In Copersucar Co. Brazil, cold sulphitation is applied and the sulphitation is carried out before liming, that is, the sulphur dioxide is added to pH 3.8, then lime is added to adjust the pH to 7 and then the juice is heated to boiling and pumped to a clarifier.

Sulphitation apparatus consists of a rectangular tank divided into two unequal connected compartments. The juice to be supplicated is fed into the smaller compartment it is picked up by a pump which takes it to SO_2 aspirator. The sulphitation takes place by contact and mixing in the vertical descending column and returning back to the juice in the column. Oxygen, nitrogen and sulphur dioxide mixture is pumped in to the column from the bottom of the small compartment as bubbles and the exhaust gas is vented to atmosphere. The SO_2 content of the juice can be manipulated by varying the speed of the circulatory pump. In general the capacity of the pump is adjusted to 1.5 the volume of the juice to be treated.

The Equipment is simulated in Aspen Plus as an absorption column to find the required vessel dimensions. The calculations can be found in Appendix A.8.3. The packing is selected based on the system requirements. Raschig rings packings are much more chipper when compared to pall ring and saddles. But, the efficiency is lower. Since sulphur dioxide is highly soluble in water, the effect of efficiency will not be as high. For corrosive non-alkaline fluid the best packing material is ceramic material. Therefore, 50 mm ceramic raschig rings will be used as packing. The design specifications for the column are given in Table 8.3.

Table 8.3 Design specifications for C-101

Parameter	Unit	C-101
Number in series		1
Streams in		<104><105><106>
Streams out		<107><135>
Packing type		Ceramic rasching ring
Packing size	m	0.05
Height equivalent of stages	m	0.75
Number of stages		3
Column diameter	m	2.8
Column height	m	2.3
Stage efficiency		0.48
Packing requirement	ton	8.8
Operating pressure	atm	1

8.2.3. Lime Milk Preparation Vessels (V-101)

The lime milk preparation tanks are operating to dissolve CaO in some part of filtrates from the rotary filters (S-103) to prepare lime milk. Two tanks are operated for lime milk preparation, one of them is emptied while the other one is filled. They are provided with stirrers rotating at 8-10 rpm. Capacity of each tank should correspond to 1.5 to 2 hours of operating time (Hugot, 1972). As mentioned earlier, the concentration of CaO is 148 g CaO/L with density of 1,116 g/L. The CaO requirement is 0.0005 t/t cane, 0.599 t CaO/h. The required liquid volumetric flow rate is 4.185 m^3/h. The required water for the lime milk is supplied by portion of the juice from filtrate. The dimensions of the vessels are presented in Table 8.4. The shaft power requirement for the vessel is calculated using the rule of thumbs given by Sinnot (1999). For medium severity mixing conditions the power requirement is given in the range 1-1.5 kW/m^3. Since the speed of mixing in current design is low the lowest margin is taken as the design specification.

Table 8.4 Design specifications for V-101

Parameter	Unit	V-101
Number in parallel		2
Streams in		<126><121>
Streams out		<132>
Volume	m^3	6.9
Working volume	m^3	6.3
Height/diameter ratio		3
Diameter		1.43
Impeller diameter	m	0.48
Impeller speed	rpm	10
Power requirement	kW	6.3
Cooling area	m^2	6.2
Cooling water	ton/h	8.16
Operating pressure	atm	1
Operating temperature	ºC	90

Since the filtrate from the filter and CaO have 93ºC and 898ºC temperature respectively, the contents of the vessel are to be cooled to 90 ºC. The details of the calculations can be found in Appendix A.8.1.4

8.2.4. Liming Vessels (V-102 & V-103)

Liming vessels are operating to supply the necessary residence time for the liming to show effect before clarification. The reaction vessels are sized based on residence time of 30 minutes. The vessels are supplied with stirrers. The stirring speed is assumed as 4 revolutions per second due to the presence of solid precipitate. The results are given in Table 8.5.

Table 8.5 Design specifications for V-102&V-103

Parameter	Unit	V-102	V-103
Number in parallel		1	1
Streams in		<107><134>	<115><133>
Streams out		<108>	<116>
Working volume	m^3	364	196
Volume	m^3	400	216
H/D ratio		3	3
Diameter	m	5.7	4.7
Impeller speed	rpm	240	240
Power	kW	364	196
Operating pressure	atm	1	1
Operating temperature	ºC	44	45.2

8.2.5. Juice Heating Equipment

The juice heater consists of assembly of tubes. The juice circulates through the tubes and the heating agent outside them. The temperature of the juice coming from sulphitation is 45.2 ºC and the temperature of the juice in ethanol branch is 44ºC and these two streams are to be heated to 103 Cº in juice heaters. The specific heat of juice containing 17% solid is obtained as given in Equation 8.5

$$C_p = 3.7601 - 0.0006T \text{ kJ/kg C}^{\circ} \qquad \text{Eqn 8.5}$$

8.2.5.1. Sugar Branch Juice Heater (E-101)

Sugar branch juice is operating to heat up the juice sent to sugar production to temperature of 103 ºC using low pressure steam. Since the plant is producing excess amount of steam, low pressure steam will be used as heating agent in juice heaters. The latent heat of low pressure saturated steam at 275 kPa and 130.7 ºC is 2,171 kJ kg^{-1}. The flow rate of the juice is 216 kg/s. The heat duty of the heat exchanger is calculated using,

$$Q = 216\frac{kg}{s} \times \int_{45.2}^{103} (3.7852 - 0.0006T)dt = 47,257\,kW$$

The overall heat transfer coefficient for steam water systems is given as 1,500-4,000 W m^{-2} $^{o}C^{-1}$. For the design, the value is taken as 3,000 W m^{-2} $^{o}C^{-1}$. The calculations are carried out based on the equations given in Appendix A.8.13. The results are presented in Table 8.6

8.2.5.2. Ethanol Branch Juice Heater (E-102)

Ethanol branch juice is operating to heat up the juice sent to ethanol production to temperature of 103 ºC. Low pressure steam is used to heat up the juice going to the fermentation. The juice is first heated up to 103 ºC in juice heaters. The flow rate of juice is 116.2 kg/s. The required heat exchanger area to heat up the juice to 103 ºC is calculated using the same methodology for sugar branch juice heater. The results are presented in Table 8.6

Table 8.6 Design specifications for E-102&E-103

Parameter	Unit	E-101	E-102
Number in series		1	1
Streams in		<108>	<116>
Streams out		<109>	<117>
Heat duty	kW	47,257	25,648
Temperature	°C	45.2-103	44-103
OHTC	W m^{-2} $^{o}C^{-1}$	3,000	3,000
LMTD	°C	51.3	51.8
Heat transfer area	m^2	317	165
Heating agent	130.7 ºC	Steam	Steam
Heating agent requirement	t/h	78.4	42.5
Pressure	atm		1

8.2.6. Flash Tanks (V-104 & V-105)

Clarifiers are usually preceded by flash tanks which are cylindrical tanks just above ahead of the clarifier to remove the air bubbles from the suspended particles and to avoid the problems in settling due to the presence of the bubbles. The vessel pressure is atmospheric pressure. The sizing of the vessel is carried out by online flash vessel calculator provided by Sugar engineer's library (IRef-1). The dimension of the vessels, the water vapor removed and the inlet and outlet temperatures are as given in Table 8.7.

8.2.7. Clarifiers (S-101 & S-102)

Clarifier is the name given to continuous settlers and it is used to separate out the solids suspended in the cane juice. These solids originate from sand adhering to the cane stalks as well as from material inherent in the cane stalk. The separation takes place by allowing the solid particles to settle out onto a tray. The solids are swept from the tray into a mud compartment, from which it is pumped to filters for de-sweetening and dewatering (IRef-1).

Table 8.7 Design specifications for V-104&V-105

Parameter	Unit	V-104	V-105
Number in parallel		1	1
Streams in		<109>	<117>
Streams out		<110><111>	<118><119>
Inlet juice flow	ton/h	778.8	418.9
Vessel volume	m^3	23.9	13.2
Flash pressure	atm	1	1
Outlet temperature	ºC	100	100
Water vaporized	ton/h	3.7	2.0
Internal diameter	m	3.9	2.9
Height	m	2.0	2.0

Clarifiers are generally separated into several compartments to increase the settling area. The juice enters from the top tangentially in to a compartment half the diameter of the main clarifier, this is termed the flocculation chamber. The mud accumulates in the lower compartment. There is a central hallow shaft that circulates the plates at a speed of 12 rev/hour.

The major parameter for the clarifiers is the settling area. The speed of settling for an average juice is given as 7.5 cm/min. The largest clarifier capacity for four compartment DORR RAPIDORR type clarifiers is given as 158 ton cane per hour with 11 m diameter, 94.6 m^2 compartment area , 378.3 m^2 total area and total volume of 591 m^3. For multi feed type DORR clarifiers the maximum capacity of 185 ton cane per hour is reported for 5 compartments with diameter of 12.2 m, compartment area of 115 m^2, total area of 576.5 m^2 and total volume of 712.22 m^3(Hugot, 1972). As it is seen, RAPIDOR type clarifiers settling efficiency per m^2 is higher than multi feed type. The specifications of RAPIDORR will be considered for clarifier. Therefore, the number of clarifier equipment is eight each having area of 378 m^2. Three of these clarifiers will be operating at ethanol branch (S-102) and five of them will be operating at sugar branch (S-101). The design specifications of the clarifiers are given in Table 8.8.

Table 8.8 Design specifications for S-101&S-102

Parameter	Unit	S-101	S-102
Number in parallel		5	3
Streams in		<112>	<118>
Streams out		<113><114>	<120><121>
Settling area	m^2	378	378
Diameter	m	11	11
Volume	m^3	577	577
Power	kW	12	7.2

8.2.8. Rotary Vacuum Filter (S-103)

The defecated mud from clarifiers which has 25% solid concentration is filtered to recover the sugar solution in rotary filters. The viscosity of the juices which contain gums and waxes decreases with increasing temperature. Therefore, it would be advantageous to carry out the filtration at high temperatures preferably above 80ºC. This is why the condensates from the evaporators are used as the wash water.

The most common equipment for sulphitation processes is rotary vacuum filters. In rotary vacuum filter, the filter is composed of hollow drum rotating about a horizontal axis and partly submerged in the liquid to be filtered. Most drum filters operate at a rotation speed in the range of 0.1 to10 rpm (Perry and Green, 1997).

The achievable filter flux is 60-73 kg $m^{-2}h^{-1}$. The filtration area requirements are given as 0.2-2 m^2 per ton of cane the average value of 0.6 m^2 per ton of cane is given by Hugot (1972). For the given capacity 1,197 ton cane per hour the total area requirement is 718.2 m^2. The sizes of industrial rotary vacuum filter range from 0.37 to 186 m^2 (Perry and Green, 1997). Taking the maximum margin 180 m^2 the number of required equipment is obtained as 4.

The wash water requirement is given as 100-150% of the filter cake. The temperature of wash water is recommended to be 75-80ºC. In the calculations, the wash water to filter cake ratio is taken as 1:1. The condensates of evaporators are blended after heating up the juice in juice heaters. 56 t/h of the condensate of evaporators is used for washing. The total power requirement for filter installation is given to be approximately 0.8 kW/m^2 filtration area which corresponds to 578 kW power for 720 m^2. The design specifications for the rotary vacuum filters are given in Table 8.9.

Table 8.9 Design specifications for S-103

Parameter	Unit	S-103
Number in parallel		4
Streams in		<122><123>
Streams out		<124>
Type		Rotary vacuum filter
Filter area	m^2	180
Power requirement	kW	578

8.3. Multi Effect Evaporators (E-201, 202, 203, 204, 205, 206)

Multi effect evaporators are operated to concentrate the clarified juice from w/w 14-18% to w/w 60-70% solid content. Before the juice comes to evaporators its temperature is around boiling point (~100ºC).

More commonly used evaporators are; (1) Forced Circulation used for high tube scaling or fouling (high liquor concentration) liquors; (2) Rising Film used for low cost effects and where tube scaling/fouling from the liquor is low; (3) Falling Film used for all effects because of higher heat transfer even at low temperature difference between shell side and tube side streams and where scaling/fouling of tubes from liquor is low. The short tube vertical evaporators are most widely used commercial evaporators in cane sugar industry (Perry and Green, 1997).

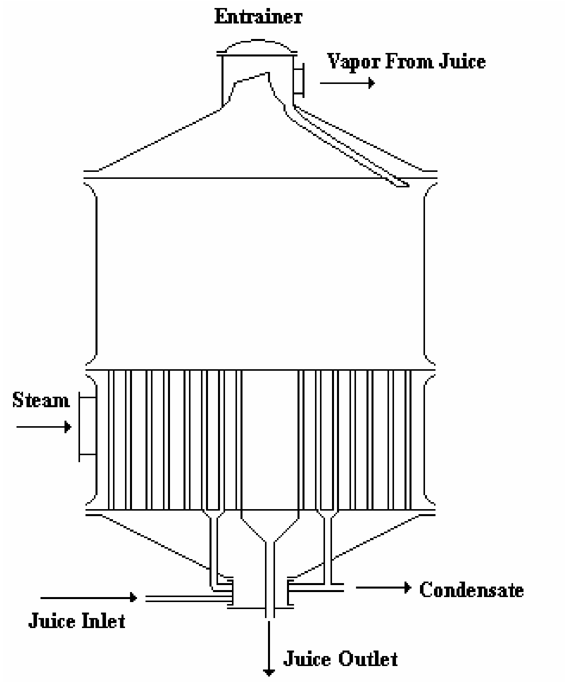
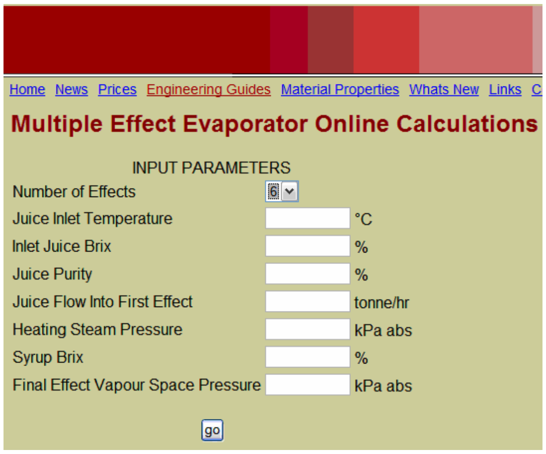


Figure 8.1 Evaporator scheme

The critical temperature limit to prevent the cane sugar caramelizing is 127 ºC. Therefore, the maximum steam temperature should be 125-130ºC. The lower temperature limit for the evaporators is the boiling point of the juice at the specified vacuum conditions. In sugar factories vacuum conditions to boil the juice at 42-52 ºC is already obtained (Hugot, 1972).

A standard multiple effect evaporator consists of a vertical cylinder, built on the tubular calandria across which the heat exchange takes place (Figure 8.1).The empty cylindrical section is used to prevent any entrainment generated during the juice boiling. The height of this space is recommended to be 2-2.5 times the height of the heat transfer tubes. The minimum diameter of the vessel is to be determined by considering the velocity of steam. The suitable steam velocities are reported to be 3.7-4.6 m/s the velocity should not exceed these values but can be lower. The length of the heat transfer tubes are 2-9 meters long. For short tube vertical evaporators the most common lengths are to the lower margin. For the design 2 m is used for the tubes. And, their diameters are reported to be within the range of 2.5-7.5 cm with wall thickness of 0.15-0.25 cm. The interval for short tube vertical evaporators is given as 5.1-7.5 cm.

Based on the information given above and the mass flows in the process, the multi effect evaporator design calculations were carried out using the online calculator (See Figure 8.2) provided by Sugar engineering web site (IRef-1). To operate the calculator, the user is to input the stream properties for the juice to be treated. The calculations in the web page are based on the evaporation design methodology presented in Hugot (1972) and Chen (1985). The necessary input values for the calculator is given in Table 8.10. The steam pressure corresponds to saturated pressure at 130ºC. And, the final effect vacuum is for boiling point of 56.6ºC.



Multiple Effect Evaporator Online Calculations

INPUT PARAMETERS

Number of Effects	6	
Juice Inlet Temperature		°C
Inlet Juice Brix		%
Juice Purity		%
Juice Flow Into First Effect		tonne/hr
Heating Steam Pressure		kPa abs
Syrup Brix		%
Final Effect Vapour Space Pressure		kPa abs

Figure 8.2 The online evaporator design input screen

Table 8.10 The input parameters for online evaporator design

Parameter	Value
Number of effects	6
Inlet juice brix (%)	16.2
Inlet juice purity (%)	93
Juice flow in to first effect (ton/h)	710.1
Heating steam pressure (kPa)	275
Syrup brix (%)	68
Final effect space pressure (kPa)	16.6

The results showing the steam consumption, vapor and juice flow from each effect, overall heat transfer coefficients and evaporation rates for each effect calculated by the online design tool is given in Tables 8.11-13,

Table 8.11 Vapor flow rates and properties of each effect evaporator

	Mass Flow (ton/h)	Pressure (kPa)	Temperature (°C)
Inlet steam	105.94	275.00	130.75
Vapor 1	72.81	231.94	125.13
Vapor 2	79.52	188.88	118.58
Vapor 3	86.31	145.82	110.68
Vapor 4	93.21	102.76	100.54
Vapor 5	100.36	59.70	85.98
Vapor 6	108.71	16.64	56.32

Table 8.12 Juice flow rates and properties of each effect evaporator

	Mass Flow (ton/h)	Brix (%)	Temperature (°C)
Inlet Juice	710.10	16.2	100.00
Syrup 1	637.29	18.05	125.49
Syrup 2	557.77	20.62	119.01
Syrup 3	471.45	24.4	111.22
Syrup 4	378.24	30.41	101.28
Syrup 5	277.88	41.4	87.16
Product	169.17	68	59.86

Table 8.13 Overall heat transfer coefficients and evaporation rates for each effect evaporator

Effect	Area (m^2)	*OHTC** ($kW\ m^{-2}K^{-1}$)	Evaporation Rate ($kg\ m^{-2}h^{-1}$)
1	3,758	3.233	19.38
2	2,697	2.683	29.48
3	3,123	2.120	27.64
4	3,673	1.549	25.38
5	4,458	0.979	22.51
6	5,982	0.409	18.17

*OHTC; overall heat transfer coefficient

The calculated areas require large equipment sizes. To obtain sizes close to industrial commercial values the juice stream will be split into two and two equipment sets will be used.
Using the information obtained from the online design tool, the minimum diameter requirement of each effect evaporator is calculated as given in Table 8.14. The velocity of vapor in each effect is set to 4.6 m/s.

Table 8.14 Minimum diameter calculations for each effect evaporator

Evaporator	Number	F_{VM} (kg/s)	ρ_{SS} (kg/m^3)	F_{VV} (m^3/s)	A_F (m^2)	D_{Vmin} (m)
E-201	2	10.11	1.52	6.64	1.44	1.36
E-202	2	11.04	1.07	10.37	2.25	1.69
E-203	2	11.99	0.83	14.41	3.13	2.00
E-204	2	12.95	0.60	21.66	4.71	2.45
E-205	2	13.94	0.35	39.45	8.58	3.30
E-206	2	15.10	0.09	169.16	36.77	6.84

F_{VM} is mass flow rate of vapor, ρ_{SS} is density of saturated steam (Smith *et al.,*1996), F_{VV} is the volumetric flow rate of vapor, A_F is vessel cross sectional area (F_{vv} steam velocity), D_{Vmin} minimum vessel diameter

The number of tubes and their dimensions are calculated as given in Table 8.15 The length of the tubes and their diameters are taken as 2 m and 5.1 cm based on the intervals previously stated. For the configuration of the tubes the pitch between the tubes should also be known. The pitch between tubes is determined by the following formula,

$$P_t = 0.952\frac{d_o}{\sqrt{k_e}} \qquad \text{Eqn 8.6}$$

d_o is the outer diameter if the tubes and k_e is the ratio of the total area of the tubes to ratio of the plane and has a value between 0.45-0.55. The value of 0.5 is assumed for these calculations.

As it is given in Table 8.15, the required vessel diameter to place the required number of tubes is larger than the vessel diameters given in Table 8.14 Therefore, the velocity of vapor will be lower than the critical limit with these vessel diameters. The larger diameters will be taken as the design values. The design specifications for the evaporators are given in Table 8.16.

Table 8.15 Number of tubes and required vessel diameter for each effect evaporator

	Area	Tube				N_T	A_{TF}	A_{TV}	D_V	P
	(m^2)	d_i (m)	d_o (m)	L (m)	A_{side} (m^2)		(m^2)	(m^2)	(m)	(m)
E-201	1,879.0	0.051	0.053	2.00	0.33	5,642	12.45	24.90	5.63	0.07
E-202	1,348.5	0.051	0.053	2.00	0.33	4,049	8.93	17.87	4.77	0.07
E-203	1,561.5	0.051	0.053	2.00	0.33	4,689	10.34	20.69	5.13	0.07
E-204	1,836.5	0.051	0.053	2.00	0.33	5,515	12.17	24.33	5.57	0.07
E-205	2,229.0	0.051	0.053	2.00	0.33	6,694	14.77	29.53	6.13	0.07
E-206	2,991.0	0.051	0.053	2.00	0.33	8,982	19.82	39.63	7.10	0.07

d_i is internal diameter, d_o outer diameter, L length of one tube, A_{side} is total side area of one tubes N_T is number of tubes, A_{TF} is the total flow area, A_{TV} is vessel cross sectional area calculated from $k_e x A_{TF}$, D_V vessel diameter and P is pitch.

Table 8.16 Design specifications for E-201-206

Parameter	Unit	E-201	E-202	E-203	E-204	E-205	E-206
Number in parallel		2	2	2	2	2	2
Streams in		<114>					
Streams out		<201><203><204><205><206><207><228>					
Heat duty	kW	63,888 (First Effect Steam)					
Temperature	°C	125.0	118.6	110.7	100.5	86.0	56.3
OHTC	$W\ m^{-2}\ {}^{o}C^{-1}$	3,233	2,683	2,120	1,549	979	409
LMTD	°C	5.7	6.4	7.9	10.2	14.5	29.7
Heat transfer area	m^2	1,879	1,348.5	1,561.5	1,836.5	2,229	2,991
Diameter	m	5.6	4.8	5.1	5.6	6.1	7.1
Height	m	6.0	6.0	6.0	6.0	6.0	6.0
Steam requirement	t/h	105.9					
Pressure	kPa	232.0	188.9	145.8	102.8	59.7	16.6

8.4. Crystallization

8.4.1. Sugar Boiling Pans (E-207 & E-208)

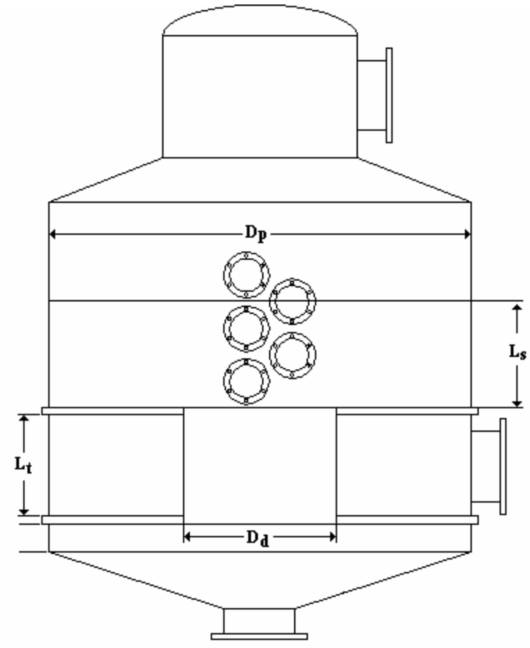


Figure 8.3 Boiling Pan Scheme

The concentrated sugar from evaporators are further concentrated in series of boilers. Vacuum pans are utilized to increase the sugar concentration to critical levels. The crystals formed in pans are separated by means of centrifuges. The mother liquor from the first pan centrifuges is named as A massecuites and still contains significant amount of sugars. Therefore, A massecuites are transferred in to second vacuum pan. The liquor after the sugar crystals are separated is named as B massecuites. The number of boilers can be increased, but, every time the quality of sugar solution decreases. Instead, the B molasses are sent to fermentation to be fermented into ethanol. The boiling times and crystal yields for first and second pans are given as 2.5-3, 3-4 h and 62.5, 60.3% respectively (Hugot, 1972). The recommended seeding amount is 1.6-16 g of sugar crystals per m^3 syrup. The sugar crystals should be free of moisture and finely grinded by ball mills. The sugars are added at the final phases of boiling and incubated for 20 more minutes to be sent to crystallizers.

The structure of the vacuum pans (Figure 8.3) are similar to the evaporators there is a tube section at the bottom of the vessel for heat transfer and there is a cylindrical section at top of the tubes where the circulation of the syrup occurs. The graining volume for batch pans is described as the minimum volume of concentrated juice to be introduced into vessel before turning on the steam. This can be interpreted as the amount of molasses that can totally cover the tubes. This value is usually 24-40 % of the working volume of the vessel. The juice circulation inside the pan can be natural circulation or forces circulation using stirrers. The power requirement for forced circulation is given as 0.75 kW/m^3 massecuites for three bladed propellers rotating at 56 rpm.

The evaporation rates are expressed in kg per square meter of heating surface per hour. The average evaporation rate values given for A massecuites and B massecuites are 32-72 and 11-46 kg m^{-2} h^{-1} respectively. The ratio of heating surface to working vessel volume is given to be 5-6.5 m^2/m^3 for steam at pressure of 48.26-103.42 kPa (Hugot,1972). The evaporation temperature is given in the interval 70-75ºC and is set to 75ºC which corresponds to 38.6 kPa pressure (Hugot, 1972). Low pressure steam will be used to heat the pans. The sugar crystal yield in first and second pan is assumed as 60% which yields 84% total sugar recovery in crystallization.

The boiling and crystallization of sugar can be carried out either batch vice or continuous manner. The continuous vacuum pans are operated since late 1940s. And currently there are number of sugar factories around Europe with continuous sugar boiling and crystallization facilities. With the development of the controlling systems, the continuous systems have more steady operation parameters and the operator requirements are lower. And also the down time for emptying and filling and the storage tank requirements are the added disadvantages of the batch systems on continuous systems. Therefore, for the current design continuous pans will be operated.

Unlike batch pans, the continuous pans are composed of several compartments. Different companies patented different designs for the continuous pans. Chen (1985), listed the preferred continuous systems as follows,

- Single stage forced crystallization evaporators, operating at low crystal contents, 10-30%, for organic and non-organic compounds, not sucrose.
- Combination of pre-evaporation of syrup at high temperature and high concentration, followed by seeding, cooling and gradually flashing to low temperatures with partial plug flow of the crystals through the system. De Vries systems is an example for this design.
- Using a series of vacuum pan unit systems, seed crystals are added to first compartment and concentrated juice flows through the compartments to the exit of the last cell.

The sugar boiling can also be conducted by converting the current design to continuous system. The Godchaux-Henderson sugar company system utilizes this system (Chen, 1985). To ensure continuous operation, seed slurry should be injected into the standard calandria pan which is equipped with mechanical circulation and the massecuites is removed by a magma pump. For the current design the last option will be considered.

In determining the volume of the pans the residence times are taken as 2.5 and 3 hours respectively. The parameters relating to juice boilers are given in Table 8.17. The pan design is carried out using the online tool (See Figure 8.4) provided by sugar engineering web page (IRef-1). The required parameters for calculation are given in Table 8.18. And the vessel dimensions are obtained as given in Table 8.19.

Table 8.17 The parameters used in the design of boiling pans

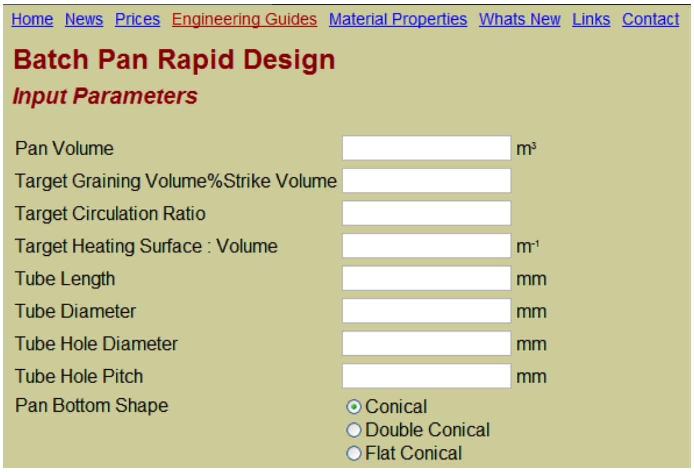
Parameter	Unit	E-207	E-208
Juice mass flow	ton/h	166.7	77.5
Density of syrup	kg/m^3	1,330.0	1,304.0
Volumetric flow	m^3/h	125.3	59.4
Residence time	h	2.5	3.0
Working volume	m^3	313.3	178.3
Total water evaporated	ton/h	25.2	14.4
Latent heat of vaporization at 75 ºC*	kJ/kg	2,321.5	2,321.5
Heat duty	kW	16,276.1	9,285.9
Evaporation rate	kg $m^{-2}h^{-1}$	70.0	45.0
Area required for evaporation	m^2	360.6	320.0
Latent heat of steam	kJ/kg	2,171.0	2,171.0
Steam requirement	ton/h	27.0	15.4
Crystal sugar yield	%	60.00	60.00

* Smith *et al*, (1996)

Table 8.18 The input parameters for online boiling pan design tool

Parameter	E-207	E-208
Graining volume percentage	30	30
Circulation ratio [a]	2.5	2.5
Heating surface to vessel volume ratio (m^{-1})	5.0	5.0
Tube length (m)[b]	1.0	1.0
Tube diameter (m)[c]	0.120	0.120
Tube hole diameter(m)	0.122	0.122
Tube hole pitch (m)[d]	0.134	0.134

a-The circulation ratio is given between 2-3. but since mechanical circulation is used the value is taken as 10
b-The length of the tubes are given in the interval 0.75-1.30 m
c-The recommended tube diameter for pans larger then 50 m^3 capacity
d-1.016+2.667*D* where *D* is the tube diameter in cm.



Batch Pan Rapid Design

Input Parameters

Pan Volume ☐ m³
Target Graining Volume%Strike Volume ☐
Target Circulation Ratio ☐
Target Heating Surface : Volume ☐ m⁻¹
Tube Length ☐ mm
Tube Diameter ☐ mm
Tube Hole Diameter ☐ mm
Tube Hole Pitch ☐ mm
Pan Bottom Shape ● Conical ○ Double Conical ○ Flat Conical

Figure 8.4 The online boiling pan design screen

Table 8.19 Design specifications for E-207&E-208

Parameter	Unit	E-207	E-208
Number in parallel		1	1
Streams in		<202>	<214>
Streams out		<208><209>	<215><216>
Pan inside diameter	m	9.6	7.4
Number of tubes		4,146	2,364
Down take diameter (D_T)*	m	2.4	1.81
Tube length	m	1.00	1.00
Heat duty	kW	16,276	5,324
LMTD	ºC	55.7	55.7
Heating surface	m^2	1,566.5	891.5
Strike level (L_S)*	m	2.9	2.9
Stirrer		Yes	Yes
Stirring speed	rpm	56	56
Stirring power	kW	234.8	133.5

*See Figure 8.3

8.4.2. Crystallizers (V-201 & V-202)

Crystallization is the process of mixing the molasses for a certain time after draining from the pans and before passing to centrifuges. The aim of the crystallization is to complete the formation of the sugar crystals and forcing further exhaustion of the mother liquor.

The residence times for crystallizers are given as 7 hours (extremes 3.72-15.68 h) for A massecuites and 14 hours (extremes 2.3-28.13) for the B massecuites which are the most frequently used values by different companies (IRef-1).

During crystallization the temperature of the molasses is dropped to 40ºC by using cooling water. The cooling can be carried out by cooling jacket, stationary cooling coils or by revolving cooling elements. In the latter one, the cooling water of proper temperature is circulated inside the hollow pipes through the stirring shaft. The stirring elements are acting as cooling surfaces (Chen, 1985). Typical design for crystallizers can be seen in Figure 8.5.

The speed of rotation of the stirrers is recommended as 0.5 rpm by Hugot (1972). The corresponding power requirement for the stirring is given as 223.8 W/m^3 of molasses.

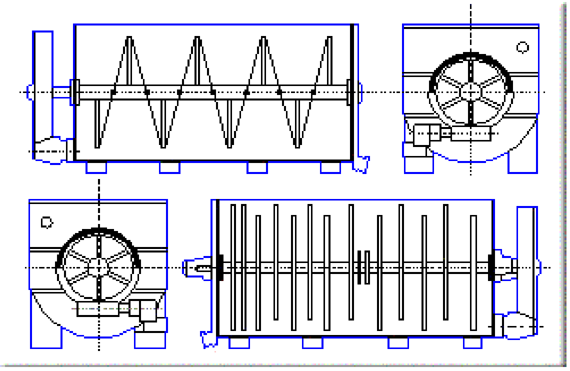


Figure 8.5 Horizontal crystallizer scheme

To maintain continuous operation, the most common application is to operate series of crystallizers in row (Chen, 1985). The number of crystallizers varies from one plant to other. For the current design the number is selected as 6 based on Mossman, Queensland sugar factory given in Chen (1985). The total residence time in each crystallizer is selected to be 1.2 and 2.4h to obtain 7.2 and 14.4 h total residence times respectively for first pan and second pan crystallizers. The cooling area to volume ratios are taken as 2 m^2/m^3 which is the average value given for the revolving cooling elements (Chen, 1985). To determine the cooling water requirement the amount of cooling is to be calculated. For that the specific heat of molasses is given as,

$$c = 4.1868\text{-}0.0293 \times B \text{ kJ/kg C}^{\text{o}} \qquad \text{Eqn 8.7}$$

The brix (B) of the massecuites is 80%. Therefore, the mean specific heat is 1.8463 kJ/kg Cº. Using this value and the total molasses flow the cooling required to lower the temperature of massecuites from 75 to 40ºC is obtained as 2,566.7 and 1,146.8 kW respectively for first and second pan crystallizers. For cooling, cooling water at 25 ºC is used. The exit temperature for the cooling water is set to 40 ºC. The specific heat of water at the temperature interval is 4.1774 kJ/kg Cº. Using this value the required cooling water amounts are calculated as 147.5 and 65.9 ton/h respectively for first pan and second pan crystallizers. The design specifications for the crystallizers are given in Table 8.20.

Table 8.20 Design specifications for V-201&V-202

Parameter	Unit	V-201	V-202
Number in series		6	6
Streams in		<208>	<215>
Streams out		<210>	<217>
Residence time	h	7.2	14.4
Mass flow	ton/h	141.5	63.1
Density of syrup	kg/ m^3	1,375.0	1,375.0
Working volume	m^3	123.5	110.1
Diameter of vessel	m	3.5	3.5
Height of vessel	m	12.9	11.5
Stirring speed	rpm	0.5	0.5
Power requirement	kW/vessel	27.6	24.7
Heating surface/volume	m^{-1}	2.0	2.0
Heating surface	m^2/vessel	247	220.2
Heat duty	kW	2,568	1,147
Cooling agent	Cooling water 25-40ºC		
Cooling agent flow	t/h	147.5	65.9

8.4.3. Reheaters (E-209, E-210, E-211&E-212)

8.4.3.1. First Pan Feed Heater (E-209)

The temperature of concentrated juice from the last effect evaporator is 56 ºC and is needed to be heated up to pan boiling temperature of 75 ºC before being sent to first boiling pan. Heat exchanger E-209 is utilized to heat up the juice. The condensate from the first effect evaporator is utilized as the heating agent. The inlet and outlet temperatures of condensate are 130.7- 90ºC respectively the average heat capacity of water at these temperature interval is 4.123 kJ kg^{-1} $^{o}C^{-1}$.

The brix of juice is 65%. Using Equation 8.7, the heat capacity of the juice is obtained as 2.2823 kJ kg^{-1} $^{o}C^{-1}$. The overall heat transfer coefficient for a reheater is about 79.5 W $m^{-2}$$^{o}C^{-1}$(Hugot,1972). The flow velocity of water in the finned tubes should be in the range 1.5 to 2 m/s, in order to achieve the stated *OHTC*. Care must be taken to ensure that the number of passes and length of each pass ensure the correct flow velocity without causing excessive pressure drops due to friction. (IRef-1). The mass flow rate of the juice in to E-209 is 166.7 t/h. The heat duty and design results are presented in Table 8.21.

8.4.3.2. Centrifuge Feed Heaters (E-210&E-212)

The massecuites from the crystallizers are at about 40°C. The reason of reheating the massecuites is to reduce the viscosity of the massecuites to make the separation of the crystal from the mother liquor in the centrifugal easier. It is considered that the viscosity of the massecuites falls 50% with

5ºC increase in temperature. Therefore, the massecuites will be reheated to 55 ºC in reheaters (E-210& E-212).

Massecuites specific heat capacity, in the temperature range, is about 1.8463 kJ kg^{-1} $ºC^{-1}$. The mass flow rate of massecuites to be heated is 143 ton/h from the first pan and 63 ton/h after the second pan. The total heat duties are 1,100 and 483.9 kW respectively for first (E-210) and second reheater (E-212). The heating agent is the condensing steam from the pans and first effect evaporator as mentioned in previous section. The heat duty and design results for E-210&212 are presented in Table 8.21.

8.4.3.3. First Pan Feed Heater (E-211)

After the centrifugation the sugar crystal free molasses from the first pan crystallizers is to be heated up the pan temperature of 75 ºC. Heat exchanger E-211 is utilized to heat up the juice from the first centrifuges (S-201). As described in previous section, the heating agent is some portion of the condensates from the first effect evaporator and boiling pans. The inlet temperature of the molasses is 55 ºC and the flow rate is 77.5 t/h. The heat capacity of the 64% solid molasses is obtained as 2.312 kJ/kg Cº using Equation 8.7. The heat duty and design results are presented in Table 8.21.

Table 8.21 Design specifications for E-209-212

Parameter	Unit	E-209	E-210	E-211	E-212
Number in series		1	1	1	1
Streams in		<201>	<210>	<213>	<217>
Streams out		<202>	<211>	<214>	<218>
Heat duty	kW	2,008	1,100	995.5	484
Temperature	°C	56-75	40-55	55-75	40-55
OHTC	W m^{-2} $ºC^{-1}$	80	80	80	80
LMTD	°C	43.7	62.7	44.3	61.7
Heat transfer area	m^2	574	204	281	98
Heating agent	Condensates from 1st effect evaporator 130.7-90ºC				
Heating agent requirement	t/h	42.7	23.4	21.2	10.3
Pressure	atm	1	1	1	1

8.4.4. Centrifuges (S-201&S-202)

The centrifuges have a perforated basket cone with angle of 34º and rotate at 2,000-2,600 rpm. The diameters are given in the range 0.61-0.81 m. the perforated screens has openings varying from 0.09mm for B massecuites to 0.06 mm for C massecuites. The maximum capacities are given in the range 7-14 ton/h for B massecuites and 7-9 ton/hour for the C massecuites. The corresponding power requirement is 20-30 kW (Hugot, 1972). The upper limits of these intervals are taken as the design parameters for the centrifuges the design parameters are listed in Table 8.22.

Table 8.22 Design specifications for S-201&S-202

Parameters	Units	S-201	S-202
Numbers in parallel		10	7
Streams in		<211>	<218>
Streams out		<212> <213>	<219> <220>
Bowl diameter	m	0.81	0.81
Max capacity	ton/h	14	9
Speed max	rpm	2,600	2,600
Motor max.	kW	30	30

8.5. Drying

The sugar crystals are dried in rotary drum driers using hot air. The entry temperature for the air in to drier is 85ºC and the exit temperature is 40ºC. The details for drum and air heater calculations can be found in Appendix A.8.5. The results are given in Table 8.23&24.

Table 8.23 Design specifications for D-201

Parameters	Units	D-201
Numbers in parallel		1
Streams in		<221><224>
Streams out		<222> <225>
Air requirement	t/h	47.5
Evaporation rate	kg $m^{-2}h^{-1}$	65
Evaporation area	m^2	14
Air flow velocity	m/s	4
Diameter	m	1.73
Power requirement	kW	13

Table 8.24 Design specifications for E-213

Parameter	Unit	E-213
Number in series		1
Streams in		<223>
Streams out		<224>
Heat duty	kW	593
Temperature	°C	30-80
OHTC	W m^{-2} $^{o}C^{-1}$	200
LMTD	°C	65
Heating agent		Saturated Steam (130.7 ºC)
Heat transfer area	m^2	41
Steam requirement	t/h	1
Pressure	atm	1

8.6. Fermentation

8.6.1. Juice Cooler (E-301&E-302)

The juice from clarifiers, filtrate from filters and molasses from crystallizers are at 100 ºC, 93ºC and 55 ºC respectively and their mixed temperature is 96 ºC, juice is to be cooled to fermentation temperature of 30ºC. For this, the hot juice is get in contact with the beer column feed to heat up the feed to 80ºC before distillation. During this operation the temperature of the juice is dropped to 45ºC. Later on, by utilizing another heat exchanger the temperature is dropped to fermentation temperature of 30ºC by using cooling water. The details of the Calculations can be found in Appendix A.8.1.2. The design results of the coolers are given in Table 8.25.

Table 8.25 Design specifications for E-301&E-302

Parameter	Unit	E-301	E-302
Number in series		1	1
Streams in		<301><308>	<301>
Streams out		<302><309>	<302>
Heat duty	kW	28,785	6,593
Temperature	°C	96-42.3	42.3-30
OHTC	W m^{-2} $^{o}C^{-1}$	1,400	1,100
LMTD	°C	12.8	12.6
Cooling agent		Stream <308>	Chilled water
Heat transfer area	m^2	1,606	476
Cooling agent	t/h		227.3
Pressure	atm	1	1

8.6.2. Fermentors (BR-301)

In Batch fermentors, sucrose in juice is hydrolyzed to glucose and glucose is fermented to microbial biomass, ethanol, carbon dioxide and fermentation by-products. The fermentation by-products are assumed as acetic acid, isoamyl alcohol, glycerol and succinic acid. There are several by-product of fermentation, but, for simplicity only the dominating components are considered. Fusel oil which is a side product of ethanol distilleries composes of various organics such as, methanol, ethyl acetate, n-propanol, iso-butanol, ethyl butyrate, 2-butanol, iso-amyl alcohol, n-butanol, n-propanol and isopropanol. Isoamylalcohol is reported as the dominant component for fusel oil (IRef-5). Therefore, for simplicity isoamylalcohol is taken as the model component for the fusel oil. The growth conditions in the vessel are 30ºC temperature and pH 4.5.

8.6.2.1. Reactions

The reaction for the growth of the microorganisms is obtained by using carbon, hydrogen, oxygen, nitrogen and degrees of reduction balances. The amount of the ATP required for the growth is found using the rule of thumb of "production of 0.61 carbon mole of biomass requires 1 mol of ATP".

$$0.175C_6H_{12}O_6 + 0.2NH_3 + 1.64ATP \rightarrow CH_{1.8}O_{0.5}N_{0.2} + 0.05CO_2 + 0.45H_2O + 65kJoule \qquad \text{Eqn 8.8}$$

The nitrogen source required for the growth is assumed to originate from juice impurities.

The ATP required for the growth is supplied by catabolic reaction given as

$$C_6H_{12}O_6 \rightarrow 2C_2H_6O + 2CO_2 + 2ATP \qquad \text{Eqn 8.9}$$

Equating the ATP's in both reactions and rearranging the reactions, the growth reaction can be written as

$$0.995C_6H_{12}O_6 + 0.2NH_3 \rightarrow CH_{1.8}O_{0.5}N_{0.2} + 1.64C_2H_6O + 1.69CO_2 + 0.45H_2O + 65kJoule \qquad \text{Eqn 8.10}$$

Using the carbon, hydrogen, oxygen and degree of reduction balances the reactions for by-product formation is obtained as

Ethanol : $C_6H_{12}O_6 \rightarrow 2C_2H_6O + 2CO_2$ Eqn 8.11

Glycerol : $C_6H_{12}O_6 + 4H^+ + 4e^- \rightarrow 2C_3H_8O_3$ Eqn 8.12

Succinic acid: : $C_6H_{12}O_6 + 2H_2O \rightarrow C_4H_6O_4 + 2CO_2 + 10H^+ + 10e^-$ Eqn 8.13

Acetic Acid : $C_6H_{12}O_6 + 2H_2O \rightarrow 2C_2H_4O_2 + 2CO_2 + 8H^+ + 8e^-$ Eqn 8.14

Isoamylalcohol : $C_6H_{12}O_6 \rightarrow 0.795C_5H_{12}O + 2.025CO_2 + 1.155H_2O + 0.15H^+ + 0.15e^-$ Eqn 8.15

The percentage distribution of glucose to products is assumed as given in Table 8.26 based on typical *Saccharomyces cerevisiae* metabolism taken from biochemical engineering hand book (Atkinson & Mavituna, 1991). Since the dominating components are taken for design, the minor components percentages are added to dominating ones to balance the mass flow.

The percentage of the substrate that is distributed between biomass and ethanol production is arranged in such a way that the total percentage of sugar utilized for ethanol production is 92%. And, the percentage of the substrate to biomass is determined in such a way that the bleed is 5% of the total yeast inside the fermentor which corresponds to yield value of Y_{SX} = 0.03 g biomass/gram glucose (for calculation see Appendix A.8.7.1). The glucose distributed between glycerol and acetic acid is arranged in such a way that the charge in the production reactions is balanced.

Table 8.26 Percentage substrate distribution among the reactions

Component	(%)
Acetic acid	0.82
Isoamylalcohol	1.00
Glycerol	2.18
Succinic acid	0.20
Ethanol	74.00
Biomass	21.80

And, for mass balances the sucrose conversion to glucose and fructose is taken as given in following reaction,

$$C_{12}H_{22}O_{11} + H_2O \rightarrow 2C_6H_{12}O_6 \qquad \text{Eqn 8.16}$$

The reactions are input to Super Pro designer simulator to obtain the mass balances and the results obtained from the Super pro designer are input to Aspen Plus 11.1 to simulate the thermodynamic behavior of the effluent streams i.e. how much ethanol, water and by products are evaporated with the vent carbon dioxide gas.

For ASPEN Plus simulation, UNIQUAC property method is used and simple flash drum is used to simulate the fermentor. Flash temperature is set to 30ºC and pressure is set to 1atm.

8.6.2.2. The Model

The details of the model can be found in Appendix A.8.7.2.

8.6.2.3. Parameter Estimation and Fermentation Time

μ_{max} and m_s are obtained as 0.0073 h^{-1} and 0.1799 g glucose/ (g biomass h) respectively. Then these values are input to MATLAB program to simulate the real results. In MATLAB simulation results are obtained as, $(C_x - C_{xo})/C_{xo} = 0.0542$, $C_E = 85.6\,kg/m^3$ and $C_S = 0\,kg/m^3$. The excess ethanol and biomass concentrations are due to neglecting the by product formation in the model. The resultant fermentation time is calculated as 7.5 h.

The details of the calculations can be found in Appendix A.8.7.3.

8.6.2.4. Fermentor Sizing

The fermentation time is obtained s 7.5 h in MATLAB and it will be assumed as 8 h for safety. Since the fermentors are operating batch vise, to maintain continuous overall operation the number of fermentors is needed to be determined first. The operation principle of four fermentors is arranged such that, one is in operation, one is being emptied, and the biomass from the emptied one is transferred to the one being filled. By doing so, the flow is continuously distributed between the fermentors, the need for intermediate storage tanks before the fermentor is eliminated, the intermediate biomass storage tank is also eliminated and the biomass sterilization can be carried out in continuous which also reduces the sterilization tank volume.

The optimal number of fermentors is obtained as 6 and the scheduling is given in Figure 8.6. In this design each fermentor is filled for 2 hours, then the fermentor ingredients are fermented for 8 hours and finally it takes 2 hours to empty the fermentors. The details of calculations to determine the number of fermentors is given in Appendix A.8.7.4.

Important parameters in fermentor sizing are the diameter to height ratio and the mixing time. Since the batch vessels do not have stirrers the mixing due to gas generation is the main mixing effect in the vessel. Therefore, the height to diameter ratios or the vessels are determined based on the final mixing obtained. Shorter the mixing time better the mixing. The details of the calculations can be found in Appendix A.8.7.4

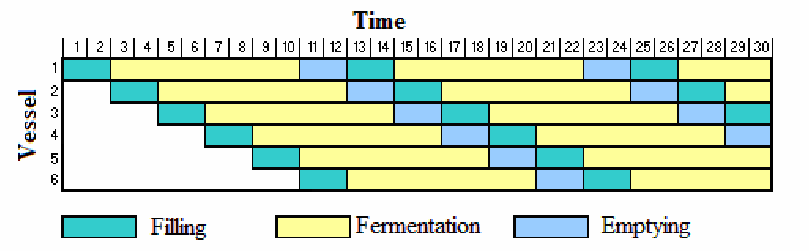


Figure 8.6 Scheduling for the fermentors

8.6.2.5. Fermentor Cooling

The details of the cooling design can be found in Appendix A.8.7.5. The final vessels design and dimensions are given in Table 8.27.

Table 8.27 Design specifications for BR-301

Parameter	Unit	BR-301
Number in row		6
Streams in		<302><333>
Streams out		<303><304>
Fermentation time	h	8
Working volume	m^3	1,311
Total volume	m^3	1442
Diameter of vessel	m	10
Height of vessel	m	18.4
Stirring		None
Power requirement	kW/vessel	None
Mixing time	s	106
Cooling surface	m^2/vessel	247
Heat duty	kW	1,360
Cooling agent		Chilled water at 10ºC
Cooling agent flow	t/h	78

8.6.3. Biomass Separation (S-301&S-302)

The fermentation broth from the fermentors is sent to centrifuges (S-301) to separate the biomass from broth to be recycled to fermentors. The centrifuges S-302 are utilized to remove the biomass from acid solution before being sent to fermentation.

Sedimentation type disc bowl centrifuges are widely used in industry for this purpose. The performance of the sedimentation centrifuges are described by sigma factor (Σ) which is equal to the cross sectional area of sedimentation tank giving the same clarification. The calculations relating to sigma value can be found in Appendix A.8.8. Based on the calculations, the disc type centrifuges proved to be more suitable for the operation. Perry's Chemical Engineering Handbook (Perry and Green, 1997) reports the maximum capacity of disk type centrifuges as 45 m^3/h with bowl diameter of 0.61 m and 5.6 kW motor power. The specifications of S-301 and S-302 are given in Table 8.28.

8.6.4. Yeast Acidification Tank (V-301)

The cell culture should be kept at low pH conditions, to prevent any microbial contamination in yeast culture during the recycling process. For this purpose, cells are separated from the fermentation broth by means of continuous centrifugation, transferred into stirred vessels and sulphuric acid solution is added in to vessel to lower the pH to 2. Approximately 13 g or sulphuric acid is required per kg of ethanol produced. The cell flow rate is 172 m^3/h. The ethanol flow out of the fermentor is 39.4 Mt/h therefore the required sulphuric acid is 512 kg/h. The required acid solution flow rate to have pH ~2 is 522 m^3/h. Including the volume of the cells the total volume flow will be 694 m^3/h. using the residence time of 15 minutes the required vessel volume is calculated as 190 m^3 including 10% safety. The design specifications for V-301 can be found in Table 8.29.

Table 8.28 Design specifications for S-301&S-302

Equipment ID	Units	S-301	S-302
Streams in		<304>	<331>
Streams out		<305> <306>	<332> <333>
Liquid flow	m^3/h	484	522
Speed max	Rpm	4 050	4 050
Max throughput	m^3/h	45	45
Motor max.	kW	5.6	5.6
Sigma factor		37,770	37,770
Number of units		11	12

Table 8.29 Design specifications for V-301

Parameter	Unit	V-301
Number in parallel		1
Streams in		<329>
Streams out		<331><330>
Residence time	h	0.25
Working volume	m^3	173
Total volume	m^3	190
Diameter of vessel	m	4.3
Height of vessel	m	13.1
Power requirement	kW	260

8.7. Ethanol Purification

8.7.1. Ethanol Scrubber (C-301)

The ethanol scrubber is utilized to absorb the evaporated ethanol from the carbon dioxide effluent of the fermentation and top product of beer column. The environmental regulations limit the emission of volatile organics to atmosphere with total of 36.4 ton/year. This is why the ethanol scrubber is designed accordingly. The design if the ethanol scrubber can be found in Appendix A.8.9. And, the design specifications are listed in Table 8.30.

Table 8.30 Design specifications for C-301

Parameter	Unit	C-301
Number in parallel		1
Streams in		<303><312><326>
Streams out		<307><327>
Number of stages		40
Height equivalent of stages	m	0.75
Efficiency		0.48
Packing type		Polypropylene pall ring
Packing size	m	0.05
Packing requirement	m^3	114
Diameter of column	m	2.2
Height of column	m	30
Water requirement	t/h	49.6
Section pressure drop	bar	0.15

8.7.2. Beer Column (C-302)

The ethanol solution from ethanol scrubber has weight percentage of 8.9% and the fermentation broth from the fermentor has 8.3%. When these two streams are mixed, the inlet concentration of ethanol becomes 8.4%.

Beer column is operated to concentrate the ethanol solution from fermentation and ethanol stripping column to 40-50%. The vapor side flow is withdrawn below the top stage since the top product of the column is mainly carbon dioxide, organic volatiles and water in gas phase. The top product of the column is recycled to ethanol stripping to recover the ethanol. The bottom side flow of the column contains most of the volatiles and fusel oil. The bottom product of the column (vinasse) contains the organic material from cane, and some portion of heavy fermentation by products like glycerol and salts. The ethanol lost in the bottom is in order of part per million. The most common application is to recover the fusel oil and sell, but, since fusel oil flow contains high portions of water the purification of the fusel oil is taken outside the battery limits and the fusel oil and other fermentation by products like glycerol and acetic acid is taken out as dilute solutions in water.

The description of design of beer column is carried out in ASPEN Plus 11.1 and the details of the design can be found in Appendix A.8.10, The design parameters of the beer column, the tray diameters and the split fractions of the products among the output streams are given in Tables 8.31-33.

Table 8.31 Tray information for C-302

Stage	Diameter (m)	Total Area (m^2)	Active Area (m^2)	Down comer Area (m^2)	Liquid flow (kg/hr)	Vapor flow (kg/hr)
2	2.35	4.35	3.48	0.43	47,920	62,162
3	2.24	3.93	3.15	0.39	40,257	63,460
4	2.66	5.56	3.82	0.87	564,120	40,963
5	2.66	5.57	3.83	0.87	562,931	52,081
6	3.85	11.62	9.30	1.16	554,864	50,892
7	3.80	11.36	9.09	1.14	547,664	143,825
8	3.79	11.27	9.02	1.13	542,172	136,625
9	3.78	11.23	8.98	1.12	538,495	131,134
10	3.78	11.21	8.96	1.12	536,259	127,456
11	3.78	11.20	8.96	1.12	534,983	125,221
12	3.78	11.19	8.95	1.12	534,278	123,944
13	3.77	11.19	8.95	1.12	533,897	123,240
14	3.77	11.19	8.95	1.12	533,690	122,858
15	3.77	11.19	8.95	1.12	533,578	122,652
16	3.77	11.19	8.95	1.12	533,514	122,539
17	3.77	11.19	8.95	1.12	533,475	122,476
18	3.77	11.19	8.95	1.12	533,445	122,436
19	3.77	11.19	8.95	1.12	183,411	122,406

Table 8.32 Split fractions of components among distillate, side draw and bottom

Component	<312>	<311>	<310>	<313>
Ethanol	0.104	Trace	0.895	Trace
Water	0.004	0.128	0.130	0.739
CO_2	0.999	Trace	0.001	Trace
Isoamylalcohol	0.002	0.103	0.149	0.747
Glycerol	Trace	0.344	Trace	0.656
Succinic acid	Trace	0.344	Trace	0.656
Acetic acid	0.003	0.104	0.162	0.731

Table 8.33 Design specifications for C-302

Parameter	Unit	C-302
Number in parallel		1
Streams in		<309>
Streams out		<310><311><313>
Number of Stages		20
Reflux ratio		3(mass)
Height of stages	m	1
Feed stage		4
Product stage		6
By product stage		19
Tray type		Nutter float valve
Tray efficiencies		0.48
Section pressure drop	bar	0.27
Condenser duty	kW	18,719
Distillate temperature	ºC	20
Reboiler duty	kW	76,886
Bottoms temperature	ºC	100

8.7.2.1. Beer Column Condenser (E-304)

The design of beer column condenser is described in Appendix A.8.10.1 and the design specifications are given in Table 8.34,

8.7.2.2. Beer Column Reboiler (E-303)

The design of beer column reboiler is described in Appendix A.8.10.2 and the design specifications are given in Table 8.34,

Table 8.34 Design specifications for E-303&E-304

Parameter	Unit	E-303	E-304
Number in series		1	1
Streams in		<311>	<312>
Streams out		<311>	<312>
Heat duty	kW	76,886	18,719
Temperature	°C	100	20
OHTC	$W\ m^{-2}\ {}^{o}C^{-1}$	1,281	1,229
LMTD	°C	30.7	19.9
Heat transfer area	m^2	1,953	765
Cooling/heating agent		Saturated steam at 130.7 ºC	Chilled water
Cooling/heating agent flow	t/h	127.5	149.3
Pressure	atm	1	1

8.7.3. Rectification Column (C-303)

Rectification column is operated to concentrate the 39.3% ethanol mixture coming from beer column to w/w 92.5% before being sent to molecular sieve adsorption. The design of rectification column was carried out in ASPEN plus. The details of the design of the rectification column can be found in Appendix A.8.11. and the design specifications are given in Tables 8.35&36.

8.7.3.1. Rectification Column Condenser (E-306)

The design of rectification column condenser is described in Appendix A.8.11.1 and the design specifications are given in Table 8.37.

8.7.3.2. Rectification Column Reboiler (E-305)

The design of rectification column reboiler is described in Appendix A.8.10.2 and the design specifications are given in Table 8.37.

8.7.4. Molecular Sieve Feed Cooler (E-307)

This cooler is operated to cool the condensate from the rectification column at 77.4 ºC to adsorption temperature of 60ºC. The design specifications of the heat exchanger are listed in Table 8.38. Using Equations A.8.11 &12, the mean heat capacity of the ethanol water mixture is obtained as 3.231 kJ kg^{-1}ºC^{-1} in the temperature interval 60-77.4ºC.

Table 8.35 Design specifications for C-303

Parameter	Unit	C-303
Number in parallel		1
Streams in		<310>
Streams out		<314><318>
Number of stages		32
Reflux ratio		4 (mass)
Height of stages	m	1
Feed stage		22
Tray type		Nutter float valve
Tray efficiencies		0.57
Section pressure drop	bar	0.27
Condenser duty	kW	56,679
Distillate temperature	ºC	77.4
Reboiler duty	kW	8,803
Bottoms temperature	ºC	100.1

Table 8.36 Tray information for C-303

Stage	Diameter (m)	Total area (m^2)	Active area (m^2)	Downcomer area (m^2)	Liquid flow (kg/hr)	Vapor flow (kg/hr)
2	4.39	15.10	12.08	1.51	170,926	214,530
3	4.36	14.96	11.97	1.50	169,746	213,832
4	4.34	14.82	11.85	1.48	168,496	212,652
5	4.32	14.66	11.73	1.47	167,161	211,402
6	4.30	14.50	11.60	1.45	165,716	210,067
7	4.27	14.33	11.47	1.43	164,132	208,622
8	4.24	14.15	11.32	1.41	162,370	207,038
9	4.21	13.94	11.15	1.39	160,379	205,276
10	4.18	13.71	10.97	1.37	158,090	203,285
11	4.14	13.46	10.76	1.35	155,405	200,996
12	4.09	13.16	10.53	1.32	152,185	198,312
13	4.04	12.81	10.25	1.28	148,220	195,091
14	3.98	12.44	9.95	1.24	143,184	191,126
15	3.92	12.06	9.65	1.21	136,551	186,090
16	3.83	11.55	9.24	1.15	127,444	179,457
17	3.71	10.81	8.65	1.08	114,448	170,350
18	3.53	9.78	7.82	0.98	96,598	157,354
19	3.36	8.86	7.08	0.89	81,035	139,505
20	3.29	8.52	6.82	0.85	75,296	123,941
21	3.28	8.45	6.76	0.85	74,026	118,202
22	1.30	1.32	1.05	0.13	73,493	16,075
23	1.29	1.31	1.05	0.13	72,982	15,399
24	1.29	1.31	1.05	0.13	72,648	14,888
25	1.29	1.31	1.05	0.13	72,465	14,554
26	1.29	1.31	1.05	0.13	72,316	14,371
27	1.29	1.31	1.04	0.13	72,229	14,222
28	1.29	1.31	1.04	0.13	72,181	14,135
29	1.29	1.31	1.04	0.13	72,155	14,087
30	1.29	1.30	1.04	0.13	72,138	14,061
31	1.29	1.30	1.04	0.13	72,127	14,044

Table 8.37 Design specifications for E-305&306

Parameter	Unit	E-305	E-306
Number in series		1	1
Streams in		<311>	<312>
Streams out		<311>	<312>
Heat duty	kW	8,803	56,679
Temperature	°C	100.1	77.4
OHTC	W m^{-2} $^oC^{-1}$	1,281	845
LMTD	°C	30.7	32.4
Heat transfer area	m^2	223.8	2,071
Cooling/heating agent		Saturated steam at 130.7 ºC	Cooling water (25-60ºC)
Cooling/heating agent flow	t/h	14.6	1,396
Pressure	atm	1	1

Table 8.38 Design specifications for E-307

Parameter	Unit	E-307
Number in series		1
Streams in		<314>
Streams out		<315>
Heat duty	kW	670
Temperature	°C	77.4-60ºC
OHTC	W m^{-2} $^oC^{-1}$	500
LMTD	°C	39.2
Heat transfer area	m^2	34.2
Cooling agent		Cooling water 25 to 40ºC
Water flow	t/h	38.6
Pressure	atm	1

8.7.5. Molecular Sieve Adsorption Column (C-304)

The rectified spirit from the rectifier is passed to one of the pair of molecular sieve beds until the bed is saturated. On a timed basis, the flow of rectified spirit is switched to the alternate bed of the pair. While one of the column is in operation the other is regenerated. Zeolite molecular sieves are adequate adsorbents for the removal of small amounts of water from organic solvents. In virtue of their small diameter (0.28 nm), the water molecules can easily penetrate the structural zeolite canals, while many organic molecules, such as ethanol (0.44 nm), are simultaneously excluded (Carmo and Gubilin, 1997).

Zeolite molecular sieves are crystalline, highly porous materials, which belong to the class of aluminosilicates. These crystals are characterized by a three-dimensional pore system, with pores of precisely defined diameter. The corresponding crystallographic structure is formed by tetrahedras of (AlO_4) and (SiO_4). These tetrahedras are the basic building blocks for various zeolite structures, such as zeolites A and X, the most common commercial adsorbents.

The life of molecular sieve may be around five to seven years. However, the operating cost is considerably less than azeotropic distillation. The details of the design calculations of molecular sieve adsorption column are described in Appendix A.8.12. The design specifications for adsorption column are listed in Table 8.40.

8.7.5.1. Cooling

Cooling is required to remove the heat of adsorption and to keep the adsorbent bed temperature at 60ºC. Details of the cooling system design can be found in Appendix A.8.12.1

Table 8.39 Physical and chemical analysis of the commercial 3 Å zeolites (Carmo and Gubilin, 1997)

Parameters	Value
Real density(g/ml)	2.765
Apparent density(g/ml)	1.21
Bulk density (g/ml) (Benson, 2003)	0.73
Porosity of particles	0.56
Porosity of bed	0.40
Equilibrium H_2O capacity (%) (Sigma Aldrich product catalog)	21
Regeneration Temperature (ºC)	175-260
Heat of adsorption (kJ/kg water) (Carmo and Gubilin, 1997)	2,365

8.7.5.2. Regeneration

The regeneration temperature for the zeolite adsorbents should be between 175-260 ºC. The regeneration gas can be either carbon dioxide or nitrogen. Since the process produces excess amount of carbon dioxide, the gas for regeneration can be selected as carbon dioxide. However, these streams contain water vapor and impurities like N_2O and SO_2. Also in presence of humidity carbon dioxide will form carbonic acid and increase the corrosion risk in the vessel which increases the investment cost. Therefore, the regeneration gas is selected as nitrogen. The column also needs to be heated to provide the heat required for evaporation of adsorbed water. The details of the regeneration calculations can be found in Appendix A.8.12.2.

Table 8.40 Design specifications for C-304

Parameter	Unit	C-304
Number in parallel		2
Streams in		<315>
Streams out		<316><320>
Packing type		3 Å zeolites
Packing size	m	0.005
Diameter of column	m	2.2
Height of column	m	7.8
Packing amount	ton	21.6
Operation temperature	ºC	60
Product purity	%	99.88
Regenerate		Nitrogen
Regeneration temperature	ºC	200
Cooling & heating area	m^2	181.5
Cooling water requirement	ton/h	60.5
Steam requirement	ton/h	3.8

8.7.6. Wet Nitrogen-Dry Nitrogen Heat Exchanger (E-308)

This heat exchanger is operated to heat up the dry nitrogen to 180ºC while cooling the wet nitrogen to 30ºC. This helps saving both cooling and heating agent usage. The design calculations for the heat exchanger can be found in Appendix A.8.12.2. The design specifications are presented in Table 8.41.

Table 8.41 Design specifications for E-308

Parameter	Unit	E-308
Number in series		1
Streams in		<320><324>
Streams out		<321><325>
Heat duty	kW	3,304
Temperature	°C	200-30 5-180
OHTC	$W\ m^{-2}\ {}^{o}C^{-1}$	300
LMTD	°C	23
Heat transfer area	m^2	467
Pressure	atm	1

8.7.7. Water condenser (E-309)

This condenser is operated to condense water to dry nitrogen. Refrigerant is used to cool the wet nitrogen to 5ºC. The design calculations for the heat exchanger can be found in Appendix A.8.12.2. The design specifications are presented in Table 8.42.

8.7.8. Product Cooler (E-310)

This cooler is operated to cool the product ethanol to from 60ºC to 30ºC. The average heat capacity of ethanol and cooling water at given interval is 2.955 and 4.17 kJ kg^{-1}ºC^{-1}. The design specifications are presented in Table 8.42.

Table 8.42 Design specifications for E-309&E-310

Parameter	Unit	E-309	E-310
Number in series		1	1
Streams in		<321>	<316>
Streams out		<322><334>	<317>
Heat duty	kW	1,107	1,057
Temperature	°C	5	60-30
OHTC	W m^{-2} ºC^{-1}	250	500
LMTD	°C	14	10.8
Heat transfer area	m^2	288	196
Cooling agent		Refrigeration	Cooling water
Cooling agent flow			60.8
Pressure	atm	1	1

8.8. Steam and Electricity production

Bagasse from mills is sent to bagasse handling unit where the moisture content is reduced to 12%. Then, it is sent to cogeneration unit where it is combusted as fuel for the generation of steam and electricity for plant use.

8.8.1. Bagasse Dewatering (D-101)

Bagasse is dewatered by passing the hot exhaust gas from the furnaces counter currently through the wet bagasse. The water content of the incoming wet bagasse which is to be evaporated to get 12% moisture is 140 ton/h which requires 93.7 MW of energy (average latent heat of water in drying temperature interval is calculated as 2,260 kj/kg using Wilson equation (Equation 4.1)).

As a first assumption, exit temperature of exhaust air is taken as 50ºC. At this temperature, saturation humidity of air is given as 0.086 kg water per kg dry air (Perry and Green, 1997). Therefore, the total air requirement to remove the specified moisture content is 1,616 ton/h which is the exhaust gas flow rate. This amount of air is to supply 93.7 MJ/s evaporation heat for evaporation. Thus, the minimum ingoing air streams temperature is calculated to be 213 ºC (mean heat capacity of the exhaust gas is 1.3 kJ/kgºC) which is far less than the inlet exhaust gas temperature (500 ºC). This means the exhaust gas from the furnaces has both the necessary moisture holding capacity and the heat capacity to evaporate the water. An energy balance is carried out to calculate the temperatures of the bagasse and air leaving the drying chamber. The results are given in Table 8.43. The temperature of the fibers leaving the drum is set to 100ºC. During the calculations the reference temperature is taken as 30 ºC. The heat capacities of the streams are calculated based on the individual heat capacities and the mass fractions of present compounds. The latent heat of evaporation of water is taken as 2,260 kJ/kg.

Using evaporation rate of 65 kg $m^{-2}h^{-1}$ the area requirement for the drum is obtained as 2,154 m^2. The design specifications for the drying drum are given in Table 8.44.

Table 8.43 Energy balances for drying chamber D-101

	In		Out	
	Fiber	**Flue gas**	**Fiber**	**Flues Gas**
Flow (kg/s)	95.04	447.62	56.15	486.51
Heat Capacity (kJ kg^{-1}°C^{-1})	2.705-0.0003T	1.130+0.00016T* 1.303+0.00011T*	1.678-0.00007T	1.193+0.00015T* 1.534+0.00005T*
Enthalpy (kJ/s)	3,534.1	328,535.8	6,576.1	325,493.7
Temperature (°C)	43.8	500	100	282.83

* The heat capacities are given for the cases of water as liquid and water as vapor

Table 8.44 Design specifications for D-101

Parameters	Units	D-201
Numbers in parallel		5
Streams in		<138><142>
Streams out		<139> <143>
Air requirement		Flue gas from furnace
Evaporation rate	kg m^{-2}h^{-1}	65
Evaporation area	m^2/drum	431
Diameter	m	4
Power requirement	kW/drum*	188

*Obtained by extrapolation using the values given in Appendix A.8.5.

8.8.2. Bagasse Burners

8.8.2.1. Amount of Steam Generated

The net calorific value of combusting bagasse is given in Section 5.1.2. In this section, using those values the quantity of the steam that can be generated will be calculated. The loss of heat in the furnace and boiler consists of the following factors (Hugot, 1972):

a) Latent heat of water formed by combusting hydrogen in the bagasse
b) Latent heat of water contained in the bagasse
c) Sensitive heat of flue gasses leaving the boiler
d) Heat loss due to radiation (neglected in the calculations)

The net calorific value of dry bagasse which is calculated as 16,540 kJ/kg bagasse using Eqn 5.1 has already taken into account the losses given in (a) and (b). The available heat is calculated based on the enthalpy balances around the furnace given as,

$$\Delta H_{Available} = \Delta H_{bagasse(12\%)} + \Delta H_{Fresh\ air} + \Delta H_{Bagasse\ burning} - \Delta H_{Flue\,gas} - \Delta H_{Ash} - \Delta H_{Fresh\ air(848-1100)}$$

The enthalpies are calculated using the reference temperature of 30°C.

Economizer

The economizer is the heat exchanger in which the flue gas is get in contact with cold fresh air to recover the lost heat in flue gas by heating up the fresh air. The air coming in to the furnace is preheated to 848 °C in economizer by getting in contact with flue gas. The exit temperature of the flue gas is set to 500 °C.

Table 8.45 Heat balances for the economizer

Component	Flow rate (kg/s)	Heat Capacity (kj/kg °C)	Initial Temp. (°C)	Final Temp. (°C)	Total Heat (MW)
Fresh Air	392.8	0.993+0.00013T	30	848	+337.3
Exhaust air	447.6	1.130+0.00016T	1,100	500	-337.3

The *LMTD* of the economizer is obtained as 350 ºC and the *OHTC* for flue gas-gas systems is taken as 0.1 kW $m^{-2o}C^{-1}$ based on the values given by Sinnot (1999). Using these values and the heat duty of 337.3 MW, the required area for heat exchanging is obtained as 9,643 m^2.

The incoming and out going enthalpies for the heat balances in furnace is given in Table 8.46.

Table 8.46 Heat loss calculations for furnace F-103

Component	Flow rate (kg/s)	Temperature (ºC)	Enthalpy (kJ/s)
$\Delta H_{Bagasse}$	56.2	100	6,576.1
$\Delta H_{Fresh\ air}$	392.8	848	337,330.7
$\Delta H_{Flue\ gas}$	447.6	1,100	589,216.2*
ΔH_{Ash}	1.2	1,100	1,778
Bagasse Burning	56.2	1,100	929,548

* The latent heat of water present in the flue gas is excluded since it is already counted in bagasse burning.

Thus, total available energy for steam production is obtained as,

$$Q_{Steam} = 929,548 + 6,576.1 + 337,330.7 - 589,216.2 - 1,778 = 682,460.3kW$$

The quality of the produced steam is given in Table 8.47.

Table 8.47 Properties of superheated steam at 86 bar and 525ºC

Parameter	Value
Temperature (ºC)	525
Enthalpy of saturated liquid at 300ºC and 86 bar (kJ/kg)	1,345.4
Enthalpy of superheated vapor at 525 ºC (kJ/kg)	3,453.0

Usually the incoming water has temperature of 90 ºC (Hugot, 1972) (the condensed steam from the heating equipment). Under this conditions, the heating required to heat up 1 kg of water at 90ºC to super heated steam at 86 bar and 525 ºC is calculated as,

First heat liquid water to 300 ºC

$$Q_{90-300^oC} = \int_{90}^{300} (4.189 - 0.0006T)dt = 855.12\, kJ / kg\ water$$

Then evaporate at 300 ºC and super heat to 525 ºC

$$Q_{300-525^oC} = (3,453 - 1,345.5)kJ / kg = 2,107.6\, kJ / kg\ water$$

The total energy requirement for 1 kg of water is obtained as 2,963 kJ/kg. For total available energy of 682.46 MW, the total amount of produced steam is calculated as 829.2 ton/h.

8.8.2.2. Furnace Dimensions

The combustion chamber includes the furnace plus the free space occupied by the gasses between leaving the furnace and reaching the boiler tubes. The volume of the combustion chamber should be determined based on the volume of the gases required for the combustion. The volume of the combustion chamber is proportional with the total heating surface of the boilers. Based on this information, the volume of the furnace is calculated by the following formula (Hugot,1972),

$$Volume\,of\ Furnace = \frac{\beta \times NCV_{Bagasse}}{1.062 \times 10^6 \dfrac{kJ}{m^3 h}} \qquad \text{Eqn 8.17}$$

β is the weight of the bagasse burned (kg/h) and NCV is the net calorific value of bagasse (kJ/kg). The resultant combustion chamber volume is calculated as 2,760 m^3.

The area requirement for heat exchanging and evaporation is calculated as 17,510 m^2 using evaporation rate of 49 kg $m^{-2}h^{-1}$ reported by Hugot (1972).

8.8.2.3. Chimney

The height of the chimney is recommended to be between 30-60 meters (Hugot, 1972). For the current plant 40 m is taken as the design value. For proper operation, the gas flow rate inside the chimney should be between 4.5-5 m/s. The total gas flow in the plant (gas from fermentors, sulphur and limestone burners and the furnace) is obtained as ~500 kg/s. which corresponds to 803 m^3/s. The diameter of the chimney is calculated as 14.3 m to obtain the specified gas flow rates.

8.8.2.4. Turbines

Multistage turbines are utilized for electricity production. The outlet steam pressure and temperature are 2.75 bar and 175ºC respectively. The outlet stem from the turbines are cooled to saturation pressure and used as process steam for heat exchangers, evaporators and distillation columns. Following equation is used to calculate the amount of the electricity produced.

$$P_E = \eta_t \eta_m \eta_a (H_i - H_o) m_s \qquad \text{Eqn 8.18}$$

Where, P_E is electricity produced (kW), η_t, η_m, η_a, are respectively turbine (0.66), mechanical (0.98) and electrical alternator (0.97) efficiencies, H_i and H_o are the inlet (3,453 kJ/kg) and outlet steam (2,815 kJ/kg) enthalpies. Inserting the variables in to equation yields electrical power production of

$$P_E = 0.66 \times 0.98 \times 0.97 \times (3{,}453 - 2{,}917) kJ / kg \times 230.33 kg / s = 77.4 MW$$

The efficiencies are taken from research of Damen (2001).

Table 8.48 Results for fuel combustion calculations

Parameter	Unit	Value
Heat supplied by fuel	MW	929.3
Excess air requirement	ton/h	1,414
Super heated steam temperature	ºC	525
Super heated steam pressure	bar	86
Exhaust gas temperature	ºC	500
Heat available for steam production	MW	682.5
Steam produced	ton/h	829.2
Electricity production	MW	77.4

Table 8.49 Design specifications for Cogeneration unit (F-103)

Parameter	Unit	
Furnace temperature	ºC	1,100
Furnace volume	m^3	2,760
Boiler heating area	m^2	17,510

9. Process and Equipment Design: 2015 Plant

Tools used in this process are Super Pro Designer simulation program v5.1, ASPEN Plus 11.1, MATLAB and MS Excel spread sheet calculation. Most of the equipment design is carried out in MS Excel and ASPEN Plus and the parameters obtained from these calculations are implemented into Super Pro Designer to simulate the process. The design of the 2015 plants is described in following sections.

All the units used in this section are metric units.

9.1. Mill, Clarification and Sugar Boiling

In 2015 plant, the designs of the equipments in mill, clarification and sugar boiling are going to be the same with some exceptions. The equipments, design of which is changed are mentioned in Section 9.8.

9.2. Pretreatment

The first step of the pretreatment of bagasse is pretreatment to break the crystalline structure of closely bound hemicellulose, cellulose and lignin. As it is described in Section 2.1, acid catalyzed steam explosion is more suitable for the pretreatment. The optimal conditions of pretreatment of sugarcane bagasse using steam explosion is reported as 200°C; 30 s residence time, water to solid ratio 2 and 1% H_2SO_4 (w/w dry bagasse) (Morjanoff and Gray, 1986). Optimal hemicellulose solubilization and hydrolysis can be achieved by high temperature and short residence times (270°C, 1 min) (Duff and Murray, 1996) or long residence times lower temperatures (190°C, 10 min) (Wright, 1998). To save from the operational costs of heating and to prevent large reactor size due to the short residence times, low temperature and long residence times will be used in the design.

There are some obstacles to be considered before starting the design. First one is the handling of fibers under high pressure and temperature conditions. The presence of fibers can cause clogging in the nozzles and time losses due to the maintenance requirements. The second obstacle is the short residence times for hydrolysis which requires continuous operation for lower investment costs and smaller equipment sizes. Another reason is batch operation requires several operators while continuous operation up to 600 ton pen day can be operated by one operator. Moreover, the digester shell corrosion can also be lowered in continuous operation since the reactor will be operating at relatively constant pressure and temperature conditions, thus, inhibiting corrosion (Britt, 1964). To overcome these obstacles, the solution is to be searched in pulp and paper industries. The alkaline and chemimechanical pulping operations is pulp and paper industries are utilizing continuous digesters that operate under high temperature and pressure conditions. The equipment and design utilized in these operations can be manipulated to fit into the current process. In the following sections, the implementation of the idea into current design is explained.

9.2.1. Continuous Digesters in Alkaline Pulping

Digesters are normally provided with a circulating system and heat exchanger to bring the contents to pulping temperature. However, direct injection of steam into a digester is also common. Forced circulation affords better contact between liquid and wood chips and also minimizes the load of evaporators by reducing the dilution of the liquor. In batch digesters, the chips are dumped into the reactor and the reactor is sealed to be heated up to reaction conditions. In continuous operation, there are some additional features like chip preheating and air removal, chip and liquor feeding, impregnation of chips with liquid, heating to cooking temperature, retention at cooking temperature and cooling and blowing the cooked chips to blow a tank (Britt, 1964).

In a continuous digester (Figure 9.1), the wood chips are fed from a small overhead surge bin through a chip meter and a rotary low pressure feeder into the steaming vessel. A slowly turning screw carries the chips through the horizontal steaming vessel to remove entrained air; the chips are discharged from a steaming vessel into chip chute connected to high pressure feeder. A successful method of feeding the chips continuously in to a pressure vessel had been one of the main stumbling blocks in development of continuous digester. In this feeder, the cooking liquor is used to flush the chips from the rotor pocket and to seal the feeder (Britt, 1964).

The pressure in the digester should be maintained above the vapor pressure of the liquor in given conditions to eliminate the flashing of the liquor. The first zone at the top is sized to allow the chips to impregnate in the cooking liquor for 45 minutes. The next zone is the cooking zone where the temperature is raised to cooking temperature. The third and the last zone is the reaction zone where the residence time is set according to reaction requirements (Britt, 1964). In the current design, the washing and cooling sections are not required. The reacted liquor is immediately discharged to atmospheric pressure to provide explosive decomposition of hemicellulose.

The major difference between the pulp and paper digester and hydrolysis digester is the difference in the residence times. But, the technology utilized for continuous digestion can be applied for the pretreatment reactors at short residence times.

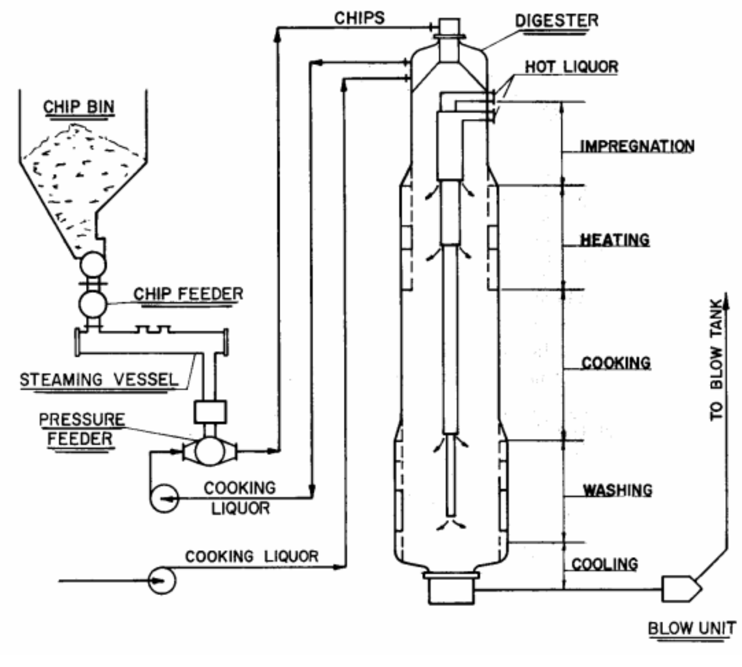


Figure 9.1 Digester scheme for pulp and paper industries (Britt, 1964)

As previously mentioned, the proposed optional treatment liquid to solid ratio for pretreatment is reported to be 2:1. However, if heating is supplied by direct steam injection, the pumping of the solids into the reactor will be an obstacle because the injected steam will contribute to the liquid portion when condensed and the liquor flow to be blended with fibers to ease pumping will be lower. To overcome this, the experimental data from a study on dilute acid hydrolysis will also be analyzed to implement into the current design.

Neureiter (2002) conducted experiments with bagasse under varying operation conditions. The different operating condition in study is listed in Table 9.1

In this study, the applied solid to liquid ratios are lower than the previously mentioned optima. Combining this information with the previously stated operating conditions (220ºC 30 sec 1% acid and 190ºC 2 minutes), it can be derived that the liquid percentage in the reactor can be increased further up to 20% solids in the reactor. Therefore, the heating in the reactor can be carried out by direct steam injection.

Table 9.1 Sugar yields of acid catalyzed pretreatment for different experimental conditions (Neureiter ,2002)

	pH	T	Time	Solid	Yields (% solids)							
		(ºC)	(min)	(%)	Glucose	Xylose	Galactose	Arabianose	Mannose	HMF	Furfural	Acetic
1	1.52	180	15	8	1.55	19.53	0.26	0.74	0.41	0.33	1.26	1.72
2	1.30	170	10	4	2.50	21.90	1.00	1.26	0.47	0.35	1.16	3.28
3	1.16	160	15	8	1.82	20.27	0.85	1.03	0.36	0.09	0.73	2.37
4	1.16	160	15	16	2.13	19.87	0.86	1.19	0.35	0.11	1.39	2.65
5	1.16	180	5	8	2.23	18.69	0.81	0.85	0.4	0.07	1.30	2.11

Thus, the resultant operating conditions in the reactor is set to 190ºC with 2 min residence time and 2:1 liquid to solid ratio (Aden *et al.*, 2002). The corresponding vapor pressure of water is 12.48 bar. The acid addition will be w/w 0.5% of the total liquid in the reactor after the steam injection. 0.5 % is obtained based on the experimental optimum of 1% on solids which corresponds to 0.5% on liquids for 2:1 liquid to solid ratio.

When the heating is supplied by direct steam injection, the heat transfer is completed in the order of seconds (Experimentally 2-7 s). By sending steam from various nozzles into the reactor, similar characteristic times can be obtained in large scale.

9.2.2. Design of Pretreatment Reactor

The pretreatment reactor will be designed based on the information given in Section 9.1. The chips from mill are blended with acid solution which is 1:1 ratio with the dry fibers.

The acid solution to be blended with fibers is heated up to 100ºC using low pressure steam by means of a heat exchanger or can be prepared using the condensates of the evaporators which are already at high temperatures. The fibers are also heated to 100 ºC by direct contact with steam. By doing so both, the fibers are heated to 100ºC and the liquid ratio is increased out side the reactor before pumping, which eases the pumping. Next, liquor is mixed with fibers using a similar set up as presented in Figure 9.1 and pumped in to vessel. Then, the vessel is quickly heated by pressurizing with superheated steam from the boilers. After residence time of 2 minutes, cooling is performed either by venting steam from the top of the vessel or by explosive discharge of the entire content of the vessel to a collection vessel (Morjanoff and Gray, 1986). The design would be based on the explosive discharge of vessel ingredients through a nozzle.

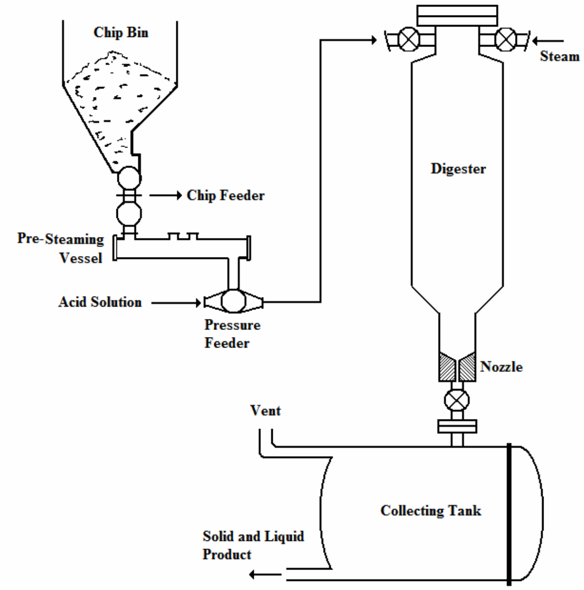


Figure 9.2 Digester scheme for steam explosion pretreatment for current design

Proposed digester-collection tank design by Puri and Mamers (1983) is implemented into the current design as given in Figure 9.2. This design is for small scale experimental design. In large scale, to increase the heat transfer the steam will be pumped to vessel from several inlets. Due to the risk of clogging and frequent maintenance requirement two sets of reactor can be used. When one is in operation, the other set will be maintained in case of clogging. The collecting tank is going to act as a flash tank where the pretreated slurry is flash cooled at atmospheric pressure. To have fast discharge of the reactor ingredients and to prevent problems arising from the clogging of the nozzles, several different nozzles should be built in each reactor.

The details of the calculations for required amount of steam and reactor sizing calculations can be found in Appendix A.9.1. The design specifications for the reactor are given in Table 9.2.

Table 9.2 Design specifications for R-401

Parameter	Unit	R-401
Number		2
Streams in		<403><404>
Streams out		<405><406>
Operation		Continuous
Reactor type		Plug flow
Volume	m^3	22
Steam requirement	ton/h	93.7
Solid content	%	30
Operating pressure	bar	14
Operating temperature	ºC	190

9.2.3. Material Balances

The 90% of the pentosans (xylan, arabinan) is hydrolyzed into pentose sugar monomers, 5% of hexosans are converted to hexose sugars (based on the conversions given in Table 9.1), 6% of pentosans is converted to furfural and 5% of the hexosans is converted to hydroxymetylfurfural, 1% of the lignin is solubilized to lignin degradation products and the acetyl groups are released as acetic acid. The composition of bagasse is as given in Table 9.3.

Table 9.3 Composition of sugar cane bagasse

Cellulose (%)	Hemicellulose (%)	Lignin (%)	Others (%)
40	30.8	25.2	4

Hexosan	Pentosan	Lignin	Extractives	Ash
42.70	23.40	25.20	4.70	4.0

Hemicellulose is composed of 76% pentosans, 9% hexosans and 15% extractives (assumption 50% is acetic acid and 50% is treated as other extractable i.e. impurity).

$$(Hexosan)_n + (n-1)\,H_2O \rightarrow n\,Glucose \qquad \text{Eqn 9.1}$$

$$(pentosan)_n + (n-1)\,H_2O \rightarrow n\,pentose \qquad \text{Eqn 9.2}$$

$$(Hexosan)_n + \rightarrow n\,HMF + 2n\,H_2O \qquad \text{Eqn 9.3}$$

$$(pentosan)_n + \rightarrow n\,furfural + 2n\,H_2O \qquad \text{Eqn 9.4}$$

$$(Lignin)_n + \rightarrow n\,Lignin\,Degradation \qquad \text{Eqn 9.5}$$

Hexosans and pentosans are assumed to be composed of 2,000 chains of pentose and hexose sugars. Therefore, 324 g hexosan yields 360 g hexose and 264 g pentosans yields 300 g pentose. These reactions and conversions values are input to Super Pro designer to generate the mass balances around the pretreatment.

9.2.4. Collecting Tank (V-401)

The collecting tank is operated for collecting the fiber residue and liquid hydrolysate after the explosive discharge. The details of vessel sizing calculations can be found in Appendix A.9.2. The design specifications for the reactor are given in Table 9.4. The heat released when fibers and sugars cooled down from 190 to 100 ºC causes an additional evaporation of 8.89 t/h.

Table 9.4 Design specifications for V-401

Parameter	Unit	V-401
Number in parallel		2
Streams in		<403><404>
Streams out		<405><406>
Volume	m^3	106
Height	m	8.0
Diameter	m	4.2
Operating pressure	bar	14
Operating temperature	ºC	190

9.2.5. Fiber Separation (S-401)

Due to the characteristics of the fibers the filtration of the fibers from the hydrolysate requires specific equipment. For that purpose, it is beneficial to concentrate on food and pulp and paper industries that process fibers. Screw presses are widely used for dehydration of different materials (See Table 9.5).

The horizontal continuous press is a dewatering machine with a conveying screw rotating inside a perforated cylinder. Incoming material is forced the length of the machine, liquid ("press liquor") is expelled through the cylinder wall. The remaining solids ("press cake") are discharged at the far end. (See Figure 9.3)

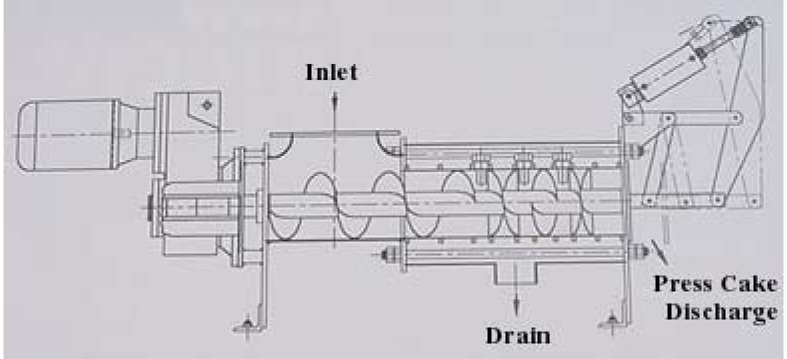


Figure 9.3 Drawing of screw press (IRef-3)

The press is available in sizes with screws ranging from 0.1 to 0.8 m in diameter. The screw has a graduated pitch, interrupted flight design. This screw interacts with stationary resistor teeth. These teeth assure that the material to be pressed does not co-rotate with the screw and is intermixed during passage through the press.

The graduated pitch screw gradually compresses the material as it passes through the main screen cylinder of the press. The press liquor is forced through the perforations in the screen.The press cake encounters resistance at the discharge of the cylinder in the form of a cone mounted on a ram. As the material reaches the discharge point, the cone exerts final pressing action to achieve maximum dewatering. The cone system allows easy adjustment of the resistance to cake discharge.

The press is fed from the top at the drive end through a rectangular flanged hopper. The press cake discharges at the far end, between the bed frame beams below the cone section. Press liquor drainage is through standard or sanitary pipe connections; these are attached to the pan under the screen section.

The capacity and the water removed during pressing can be varied by adjusting the cone pressure. Adjusting this pressure assures maximum dewatering over a wide range of feed rates, permitting good turndown from rated capacity (IRef-3).

The screens have various perforation sizes, commonly in the 0.03 to 0.25 cm diameter range. Selection is determined by the material being pressed and the desired pressing results. The industrial applications of screw press and the extent of dehydration are given in Table 9.5.

Table 9.5 Screw press applications (IRef-3)

Application	Moisture content (%)
Asbestos & water, slurry	57
Chopped bananas stem for paper	66
Coffee, bean chaff and spent grounds	55
Washed carpet fiber	46-52
Fiber glass, scrap wool with water	50-70
Grains and grasses	
Alfalfa, macerated, for liquid protein concentrate	70
Oat hulls, fiber additive for bread	50-65
Residential lawn clippings	51-62
Lemon and lime, washed peel or whole fruit	86
Manure, paunch and droppings, cow and pig	68-72
Mica flake	30
Pulp & paper	
Screen rejects	50
Fiber	30-35
Knots & shives, from paper pulp fiber production	45-64
Sawdust for pulping and burning	47
Clarifier fudge	35-50
Sludge, daf, paper mill, plastic recyclers	50
Soy bean flake	50-55
Chopped sugar cane leaves	60
Vegetables	76-82
Wood yard waste, bark, ground limps and dirt	52

Based on the information given in Tables 9.5, it is assumed that the fibers are dehydrated to 47% moisture taking the pulp and paper industries values as basis. The washing water requirement will be taken as 1:1 ratio with the filter cake. The equipment capacity is taken as 180 m^3/h, which is the maximum available (Table 9.6). The design specifications for the screw press is given in Table 9.7

Table 9.6 specifications of different capacity screw presses (IRef-3)

Model	Capacity t/hr	Capacity m^3/h	Screw rpm	Power kW	Length m	Width m	Height m	Weight kg
1	0.5 - 2.5	45	32	2.24	1.8	0.4	0.6	272
2	2-10	16	25	5.6	3.0	0.6	0.9	795
3	8-50	70	20	11.2	4.0	0.8	1.1	1,680
4	15 - 100	180	16	30	4.8	0.9	1.2	2,240

Table 9.7 Design specifications for S-401

Parameter	Unit	S-401
Number in parallel		2
Streams in		<406><418>
Streams out		<421><407>
Operation		Continuous
Cake flow	ton/h	204
Capacity	ton/h	102
Power	kW	60
Operation Temperature	ºC	84.4

9.2.6. Hydrolysate Cooler (E-401)

This cooler operates to cool down the hydrolysate from 84.4 ºC to overliming conditions of 50ºC. The average heat capacity of the hydrolysate is taken as 3.671 kJ kg^{-1}ºC^{-1} based on the average water (4.152 kJ kg^{-1}ºC^{-1}) and dissolved sugars (1.252 kJ kg^{-1}ºC^{-1}) heat capacities. The cooling agent is cooling water at 25ºC and the exit temperature of the cooling water is set to 40ºC (average heat capacity 4.17 kJ kg^{-1}ºC^{-1}). The specifications of the equipment are given in Table 9.8.

Table 9.8 Design specifications for E-401

Equipment ID	Unit	E-401
Number in series		1
Streams in		<407>
Streams out		<408>
Heat duty	kW	22,913
Operating temperature	°C	84.4-50
OHTC	$W\ m^{-2}\ {}^{o}C^{-1}$	1,400
LMTD	°C	33.8
Heat transfer area	m^2	384
Cooling agent	25-40ºC	Cooling water
Cooling agent requirement	t/h	1,045.3

9.3. Detoxification

9.3.1. Overliming (V-402) and Conditioning (V-403) Vessels

The hydrolysate from pretreatment is over limed to pH 10 at 50ºC to neutralize the acid and to partially remove the phenolic compounds (30%), furfural (20%), and HMF (20%) (Palmqvist and Hahn-hagerdal, 2000). Then, the pH is increased to 4.5 in a consecutive reactor to enhance the crystallization of and precipitation of the gypsum. The residence time in overliming and conditioning tanks are set to 30 minutes each. The pH of the hydrolysate is at high levels (~ pH=1.3), some portion of the hydrolysate will be bypassed to be mixed with the main stream downstream overliming to adjust the pH to 4.5. The neutralization reaction of sulphuric acid with calcium hydroxide is as given in following reaction,

$$Ca(OH)_2 + H_2SO_4 \rightarrow CaSO_4 + H_2O \qquad \text{Eqn 9.6}$$

Then inflowing stream to overliming vessel is neutralized and the pH is adjusted to 10 by addition of $Ca(OH)_2$. Total amount of required $Ca(OH)_2$ is obtained as 1,624 kg/h. The lime will be prepared in lime preparation tanks as w 16.7% $Ca(OH)_2$. Thus, the total lime milk flow into the overliming is obtained as 9.8 t/h.

The pH of effluent of the overliming vessel is 10. The amount of $Ca(OH)_2$ flow is 1.85 kg/h, the required amount of sulphuric acid to neutralize this amount of $Ca(OH)_2$ and to decrease the pH to 4.5 is calculated to be 3.22 kg/h. As described before, this sulphuric acid will be provided by by-pass stream separated from the hydrolysate before overliming. The amount of the hydrolysate to be bypassed is calculated as 796 kg/h.

The total volumetric flow into the overliming and conditioning tanks is 502 m^3. For residence time of 30 min the required vessel volume is obtained as 276 m^3 for each including the 10% safety factor. The power requirement for stirring is given as 98 W/m^3 (Aden *et al.*, 2002). The design specifications for the vessels can be found in Table 9.9.

Table 9.9 Design specification for V-402&V-403

Parameter	Unit	V-402&403
Number in parallel		1
Streams in		<141><409><410>
Streams out		<411><412>
Volume	m^3	276
Working Volume	m^3	251
Height	m	14
Diameter	m	5
Impeller diameter	m	1.5
Power	kW	24.6
Operating pressure	atm	1
Operating Temperature	ºC	50

9.3.2. Clarification (S-402)

These clarifiers are operated to remove the gypsum from the hydrolyzate. The slurry is assumed to have solid concentration of wt 30%. For this purpose two clarifiers will be used. The properties of the clarifiers are as described in Section 8.2.7. The design specifications are given in Table 9.10.

Table 9.10 Design specifications for S-402

Parameter	Unit	S-402
Number in parallel		2
Streams in		<412>
Streams out		<413><414>
Settling area	m^2	378
Diameter	m	11
Volume	m^3	577

9.3.3. Rotary Vacuum Filtration (S-403)

This equipment is used to concentrate the gypsum to 45% solids. Rotary vacuum filter is operated for the operation. For the determination of the filtration area, the maximum filtrate flux is taken as 73 kg $m^{-2}h^{-1}$ as described in Section 8.2.8. The total required filter area is obtained as 114 m^2 and the power requirement is 0.8 kW/m^2 which corresponds to 91.2 kW. The filter cake to water ratio is 1:1. The design specifications for the rotary filter is given in Table 9.11.

Table 9.11 Design specifications for S-403

Parameter	Unit	S-403
Number in parallel		1
Streams in		<414><428>
Streams out		<415><416>
Type		Rotary vacuum filter
Filter area	m^2	114
Power requirement	kW	91.2

9.4. Cellulose Hydrolysis

9.4.1. Hydrolysate Heater (E-402)

This heater operates to heat up the hydrolysate to 54.4ºC. The reason for not heating to 65 ºC is the temperature of the fibers (84.4 ºC) is able to heat up the stream to 65 ºC after mixing in the reactor. The average heat capacity of the hydrolysate is taken as 3.671 kJ $kg^{-1}{}^{o}C^{-1}$ based on the average water (4.152 kJ $kg^{-1}{}^{o}C^{-1}$) and dissolved sugars (1.252 kJ $kg^{-1}{}^{o}C^{-1}$) heat capacities. The heating agent for the heater is low pressure saturated stream at 275 kPa. The specifications of the equipment are given in Table 9.12.

Table 9.12 Design specifications for E-402

Equipment ID	Unit	E-402
Number in series		1
Streams in		<419>
Streams out		<420>
Heat duty	kW	1,473
Operating temperature	°C	49.9-54.4
OHTC	W m^{-2} $^{o}C^{-1}$	1,500
LMTD	°C	77.8
Heat transfer area	m^2	13
Heating agent		Saturated steam at 275 kPa
Heating agent requirement	t/h	2.45

9.4.2. Cellulose Hydrolysis Vessels (BR-401)

In this vessel, the lignin and cellulose rich fibers are blended with the conditioned hydrolysate from overliming reactor and 0.021 ton cellulase enzyme is added per ton of cellulose present in fiber. The activities of enzymes are as described in Section 2. The solid concentration in vessel is 20.1%. The vessels are supplied with stirring and they operate continuously. The residence time for 90% hydrolysis conversion is reported to be 36 h. To enhance the conversion and simulate plug flow 4 reactors will be operated in series. Actually the plug low behavior can be simulated by 3 vessels in series, but, since the volumetric flow rate is high, having less reactors in series will end up having more reactors in parallel. So, the trade of will make the final total size close to each other. Since the residence times are long, to prevent large reactor sizes the flow will be separated into two and two parallel vessel sets will be operated. The power consumption for stirring for this vessel is reported to be 20 W/m^3 (Aden *et al.*, 2002).

The hydrolysis of cellulose in to sugars is carried on based on reaction given in Equation 9.1 as described in Section 9.1.3.

To determine the sizes of the vessels, first the volumetric flow in to reactor should be determined. The hydrolysate flow rate in to reactor is 311 m^3/h (density is assumed to be 1,066 g/l). The liquid portion of the fibers has volume of 92.4 m^3/h. The density of the fibers is taken as 1,600 g/l which yields a fiber volume of 67.5 m^3/h. Some portion of the hydrolysate is used to prepare cellulase solution, but, since the cellulase solution is later going to be blended with the reactor ingredients there won't be a reduction or increase in the volume. Adding up the volumetric flows yields total flow of 471 m^3/h. Using 10% safety and residence time of 36 hours total vessel size is 18,648 m^3. There are 8 vessels and the required volume of one vessel is obtained as 2,331 m^3. The dimensioning of each vessel is given in Table 9.13.

Table 9.13 Design specification for BR-401

Parameter	Unit	BR-401
Number in parallel		2
Number in series in each set		4
Streams in		<420><421><438>
Streams out		<422>
Volume	m^3	2,331
Working volume	m^3	2,120
Height	m	20.6
Diameter	m	12
Impeller diameter	m	4
Number of impellers per vessel		5
Power	kW/vessel	42.4
Operating pressure	atm	1
Operating temperature	ºC	65

9.4.3. Fiber Residues Separation (S-404)

This operation is carried out using screw presses as described in Section 9.1.5. The design specifications are given in Table 9.14.

Table 9.14 Design specifications for S-404

Parameter	Unit	S-404
Number in parallel		1
Streams in		<422><427>
Streams out		<423><424>
Operation		Continuous
Cake flow	t/h	101.6
Capacity	t/h	101.6
Power	kW	90
Operation temperature	ºC	84.4

9.5. Fermentation

9.5.1. Juice Cooler (E-301)

The juice from clarifiers, filtrate from filters, molasses from crystallizers and hydrolysate are at 100 ºC, 93ºC, 55ºC and 57.1 ºC respectively and their mixed temperature is 79.5 ºC, juice is to be cooled to fermentation temperature of 30ºC. The heat capacity of the stream <301> is 3.426 kJ $kg^{-1}$$^{o}C^{-1}$. The design results of the coolers are given in Table 9.15

Table 9.15 Design specifications for E-301

Parameter	Unit	E-301
Number in series		1
Streams in		<301>
Streams out		<302>
Heat duty	kW	53,726
Temperature	°C	79.5-30
OHTC	W m^{-2} $^{o}C^{-1}$	1,400
LMTD	°C	16.7
Heat transfer area	m^2	2,144
Cooling agent		Cooling water 25-40ºC
Cooling agent flow	t/h	3,251
Pressure	atm	1

9.5.2. Fermentors (BR-301)

Different from the conventional plant, in 2015 plant, pentose sugars as well as hexose sugars are fermented into ethanol. Sucrose fermentation is carried out in similar manner with conventional plant. Genetically modified *Saccharomyces cerevisiae* that can ferment both pentose and hexose sugar is the microorganism utilized for the fermentation. The by-products are same with the conventional plant except xylitol which is one of the by products of the pentose metabolism. The microorganism is assumed to be further developed generations of the strains generated by Kuyper and his co-workers (2005). The pH of the fermentor media is 4.5 and the temperature is 30ºC.

9.5.2.1. Reactions

The reactions for hexose metabolism are same with that utilized in 2005 plant (Section 8.6.2.1.) The distribution of hexose sugars among the reactions are assumed to be the same with 2005 hexose fermenting organisms as given in Table 8.28.

The growth and catabolism reactions for pentose metabolism are taken from the study of Rivas and coworkers (2003) as given below.

Growth : $$0.219C_5H_{10}O_5 + 0.2NH_3 + 1.9ATP \rightarrow CH_{1.8}O_{0.5}N_{0.2} + 0.18H^+ + 0.095CO_2 + 0.405H_2O + 65kJ$$ Eqn 9.7

Catabolism : $$6C_5H_{10}O_5 \rightarrow 10C_2H_6O + 10CO_2 + 10ATP$$ Eqn 9.8

The ATP requirement for the growth is also taken from the same source. The heat of biomass growth reaction is assumed to be same with growth on hexose. Equating the ATP stoichiometic coefficients and adding two equations gives:

Growth : $$1.359C_5H_{10}O_5 + 0.2NH_3 \rightarrow CH_{1.8}O_{0.5}N_{0.2} + 1.9C_2H_6O + 1.995CO_2 + 0.405H_2O + 0.18H^+ + 65kJ$$ Eqn 9.9

The major by-products of xylose fermentation are glycerol, acetic acid, succinic acid, lactic acid and xylitol (Kuyper *et al.*, 2005). Using the component mass balances and degree of reduction balances, the reactions for by products are obtained as,

Ethanol : $6C_5H_{10}O_5 \rightarrow 10C_2H_6O + 10CO_2$ Eqn 9.10
Glycerol : $1.2C_5H_{10}O_5 + 4H^+ + 4e^- \rightarrow 2C_3H_8O_3$ Eqn 9.11
Succinic acid : $1.2C_5H_{10}O_5 + 2H_2O \rightarrow C_4H_6O_4 + 2CO_2 + 10H^+ + 10e^-$ Eqn 9.12
Acetic acid : $1.2C_5H_{10}O_5 + 2H_2O \rightarrow 2C_2H_4O_2 + 2CO_2 + 8H^+ + 8e^-$ Eqn 9.13
Lactic acid : $1.2C_5H_{10}O_5 \rightarrow 2C_3H_6O_3$ Eqn 9.14
Xylitol : $C_5H_{10}O_5 + 0.4375H_2O \rightarrow 0.875C_5H_{12}O_5 + 0.0625C_2H_6O + 0.5CO_2$ Eqn 9.15
(Rivas,2003)

The by-product yields on xylose for currently available xylose utilizing strains are reported as given in Table 9.16. (Kuyper *et al.*, 2005).

It is assumed that in industrial strains the stability, productivity and ethanol yields of the strains will be compatible with current hexose fermenting organisms. Therefore, biomass yields and ethanol yields of the pentose fermenting organisms are assumed to be the same with currently available hexose fermenting strains. That is the ethanol yield will be 92% of the theoretical maximum (0.47 g ethanol/g xylose) and the biomass yield is 0.03 g biomass/ g substrate. The rest of the by-products will be assumed accordingly.

Table 9.16 By-product yields on xylose

Component	(%)
Acetic acid	0.20
Xylitol	0.15
Glycerol	7.45
Succinic Acid	0.15
Ethanol	40.60
CO_2	41.60
Biomass	10.00
Lactic acid	0.72

The xylose distributed between the growth, succinic acid, glycerol and acetic acid producing reactions is arranged in such a way that the charges in the production reactions are balanced and the ethanol and biomass yields are satisfied. The results are given in Table 9.17.

The reactions are input to Super Pro Designer simulator to obtain the mass balances and the results obtained from the Super Pro Designer are input to Aspen Plus 11.1 to simulate the thermodynamic behaviour of the effluent streams i.e. how much ethanol, water and by products are evaporated with the vent carbon dioxide gas. For ASPEN Plus simulation UNIQUAC property method is used and simple flash drum is used to simulate the fermentor. Flash temperature is set to 30ºC and pressure is set to 1 atm.

Table 9.17 Distribution of pentose sugars among reactions

Reaction	(%)
Acetic acid	0.30
Xylitol	0.17
Glycerol	2.20
Succinic Acid	0.25
Ethanol	71.50
Biomass	24.86
Lactic acid	0.72

9.5.2.2. Model

The details of the description for model can be found in Appendix A.9.3.1.

9.5.2.3. Parameter Estimation

The details of the estimation of the parameters can be found in Appendix A.9.3.2. The results are presented in Table 9.18 and Figures 9.4&5.The total fermentation time required to consume all substrate is obtained as 9 hours.

Table 9.18 Parameters for the microorganism for currently available strains and future case values

Parameter	Current	Assumed 2015
μ_g^{max} (1/h)	0.225	0.0073
μ_{xy}^{max} (1/h)	0.11	0.0073
m_s (g substrate/gram biomass h)	0.276	0.1799
K_{sg} (g/l)	1.5	1.5
K_{sxy} (g/l)	8.5	8.5
K_I (g/l)	5.4	5.4

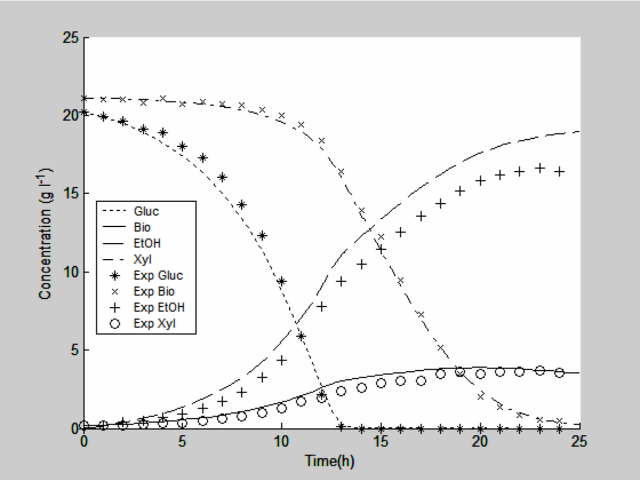


Figure 9.4 The simulation results for 20 g/l glucose and 20 g/l xylose. initial biomass is 0.17 g/l (Kuyper, *et al.*, 2005)

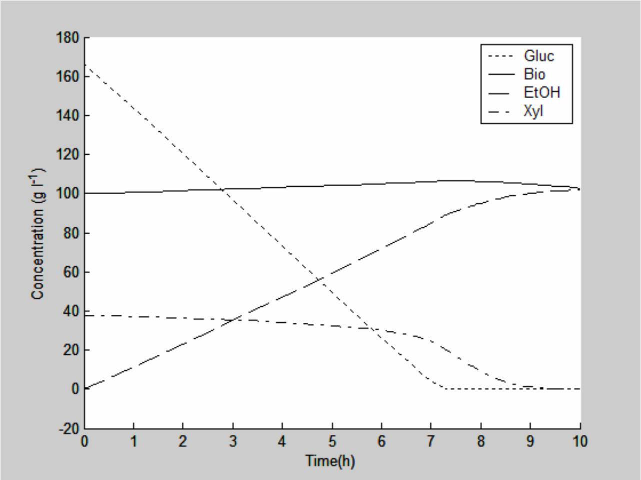


Figure 9.5 The simulation results for 166.7 g/l glucose and 37.8 g/l xylose initial biomass is 100 g/l (Design Case)

9.5.2.4. Fermentor Sizing

The details of fermentor sizing calculations can be found in Appendix A.9.3.3. The Scheduling is shown in Figure 9.6. The specifications can be found in Table 9.19.

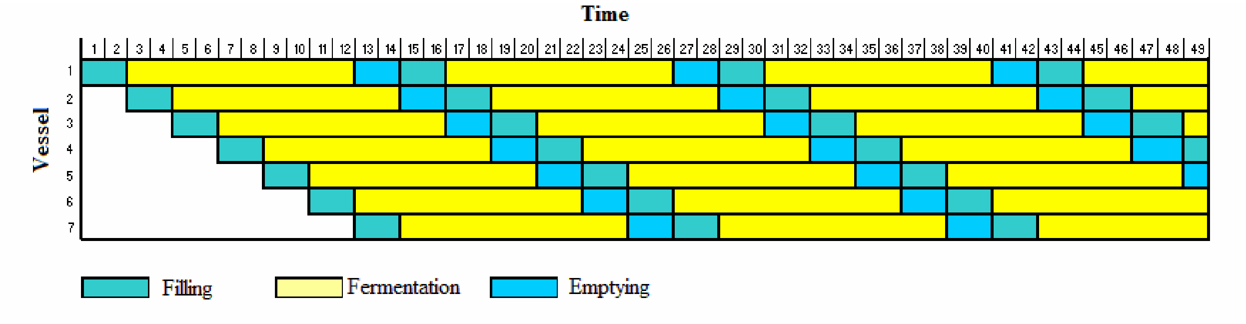


Figure 9.6 The scheduling for the fermentors

9.5.2.5. Cooling

The details of cooling jacket calculations can be found in Appendix A.9.3.4. The specifications are given in Table 9.19.

Table 9.19 Design specification for BR-301

Parameter	Unit	BR-301
Number in parallel		2
Number in series in each set		7
Streams in		<302><333>
Streams out		<303><304>
Volume	m^3	1,481
Working volume	m^3	1,345
Height	m	18.8
Diameter	m	10.0
Stirring		None
Cooling agent		Chilled water 10-25ºC
Cooling agent requirement	t/h	187
Cooling area	m^3/vessel	150.5
Operating pressure	atm	1
Operating temperature	ºC	65

9.5.3. Biomass Separation Centrifuges (S-301&S-303)

The fermentation broth from the fermentors is sent to centrifuges (S-301) to separate the biomass from broth to be recycled to fermentors. The centrifuges S-302 are utilized to remove the biomass from acid solution before being sent to fermentation.

Sedimentation type disc bowl centrifuges are widely used in industry for this purpose. The performance of the sedimentation centrifuges are described by sigma factor (Σ) which is equal to the cross sectional area of sedimentation tank giving the same clarification. The calculations relating to sigma value can be found in Appendix A.8.8. Based on the calculations, the disc type centrifuges proved to be more suitable for the operation. Perry's Chemical Engineering Handbook (Perry and Green, 1997) reports the maximum capacity of disk type centrifuges as 45 m^3/h with bowl diameter of 0.61 m and 5.6 kW motor power. The specifications of S-301 and S-302 can be found in Table 9.20.

Table 9.20 Design specifications for S-301&S-302

Equipment ID	Units	S-301	S-302
Streams in		<304>	<331>
Streams out		<305><306>	<332><333>
Liquid flow	m^3/h	974	1249
Speed max	rpm	4 050	4 050
Max throughput	m^3/h	45	45
Motor max.	kW	5.6	5.6
Sigma factor		37,770	37,770
Number of units		22	28

9.5.4. Yeast Acidification Tank (V-301)

The cell culture should be kept at low pH conditions to prevent any microbial contamination during the recycling process. For this purpose, cells are separated from the fermentation broth by means of continuous centrifugation, transferred into stirred vessels and sulphuric acid solution is added in to vessel to lower the pH to 2. Approximately 13 g or sulphuric acid is required per kg of ethanol produced. The amount of biomass to be recycled is 453 t/h. The ethanol flow out of the fermentor is 94.14 t/h. Therefore, the required sulphuric acid is 1,224 kg/h. And, the required acid solution flow rate is 1,249 m^3/h to obtain pH ~2. Including the volume of the cells, the total volume flow will be 1,650 m^3/h.For residence time of 15 minutes the required vessel volume is calculated as 312 m^3. The design specifications for V-301 can be found in Table 9.21.

Table 9.21 Design specifications for V-301

Parameter	Unit	V-301
Number in parallel		1
Streams in		<329><330>
Streams out		<331>
Residence time	h	0.25
Working volume	m^3	284
Total volume	m^3	312
Diameter of vessel	m	5
Height of vessel	m	15.9
Power requirement	kW	426

9.6. Ethanol Purification

9.6.1. Ethanol Scrubber (E-301)

The design of ethanol scrubber is carried out in similar manner as described in Appendix A.8.9. The results are presented in Table 9.22.

Table 9.22 Design specifications for C-301

Parameter	Unit	C-301
Number in parallel		1
Streams in		<303><312><326>
Streams out		<307><327>
Number of Stages		40
Height equivalent of stages	m	0.75
Efficiency		0.48
Packing type		Polypropylene pall ring
Packing size	m	0.05
Packing requirement	m^3	257
Diameter of column	m	3.3
Height of column	m	30
Water requirement	t/h	109.4
Section pressure drop	bar	0.15
Maximum stage liquid hold up	m^3	0.11

9.6.2. Beer Column Feed Heater (E-302)

This heat exchanger is utilized to heat the beer column feed from 30ºC to 80ºC. The heating agent is the fusel oil (<319> at 100 ºC) stream from the beer and rectification columns. The average heat capacity of the beer column feed at the temperature interval 30-80ºC is obtained as 4.033 kJ kg^{-1}ºC^{-1}. Since the fusel oil is water dominated the heat capacity will be taken as waters 4.147 kJ kg^{-1}ºC^{-1}. The exit temperature of hot stream is 41.5 ºC. The design Specifications for the exchanger can be found in Table 9.23.

Table 9.23 Design specifications for E-302

Parameter	Unit	E-302
Number in series		1
Streams in		<308><319>
Streams out		<309><335>
Heat duty	kW	58,133
Temperature	°C	30-80
OHTC	W m^{-2} ºC^{-1}	1,400
LMTD	°C	15.4
Heating agent		Stream <319>
Heat transfer area	m^2	2,696
Pressure	atm	1

9.6.3. Beer Column (C-302)

The design of beer column is carried out in similar manner with 2005 design as described in Section 8.7.2 and Appendix A.8.10. The results are presented in Tables 9.24-26.

Table 9.24 Design specifications for C-302

Parameter	Unit	C-302
Number in parallel		1
Streams in		<309>
Streams out		<310><311><313>
Number of stages		20
Reflux ratio		3 (mass)
Height of stages	m	1
Feed stage		4
Product stage		6
By product stage		19
Tray type		Nutter float valve
Tray efficiencies		0.48
Section pressure drop	bar	0.37
Condenser duty	kW	36,737
Distillate temperature	ºC	20
Reboiler duty	kW	165,553.2
Bottoms temperature	ºC	100

Table 9.25 Tray information for C-302

Stage	Diameter (m)	Total Area (m^2)	Active Area (m^2)	Down comer Area (m^2)	Liquid flow (kg/hr)	Vapor flow (kg/hr)
2	3.32	8.64	6.91	0.86	95,799	123,289
3	3.17	7.87	6.30	0.79	81,438	126,622
4	3.71	10.79	7.36	1.72	1,107,606	80,587
5	3.70	10.77	7.35	1.71	1,104,517	100,590
6	5.63	24.92	19.94	2.49	1,083,977	97,501
7	5.56	24.29	19.43	2.43	1,066,247	311,961
8	5.54	24.07	19.26	2.41	1,053,536	294,230
9	5.52	23.97	19.18	2.40	1,045,703	281,520
10	5.52	23.93	19.15	2.39	1,041,348	273,686
11	5.52	23.92	19.13	2.39	1,039,061	269,331
12	5.52	23.91	19.13	2.39	1,037,889	267,044
13	5.52	23.91	19.13	2.39	1,037,291	265,873
14	5.52	23.91	19.12	2.39	1,036,976	265,274
15	5.52	23.90	19.12	2.39	1,036,797	264,959
16	5.52	23.90	19.12	2.39	1,036,675	264,780
17	5.52	23.90	19.12	2.39	1,036,566	264,658
18	5.52	23.89	19.12	2.39	1,036,443	264,550
19	5.52	23.89	19.11	2.39	306,241	264,426

Table 9.26 Split fractions of components among distillate, side draw and bottom

Component	Distillate	Bottom	Product	By product
Ethanol	0.090	Trace	0.910	Trace
Water	0.004	0.056	0.151	0.789
CO_2	0.999	Trace	0.001	Trace
Isoamylalcohol	0.002	0.043	0.170	0.785
Glycerol	Trace	0.304	Trace	0.696
Succinic acid	Trace	0.304	Trace	0.696
Acetic acid	0.003	0.043	0.188	0.767
Lactic acid	Trace	0.290	0.001	0.708
Furfural	0.020	Trace	0.977	0.003
HMF	0.020	Trace	0.977	0.003

9.6.3.1. Beer Column Condenser (E-304)

The amount of the vapor to be condensed by the condenser is 92,467 kg/h. The condenser operates at atmospheric pressure. The vapor enters the condenser saturated at 76.3ºC and the condensation is complete at 20ºC. The condenser heat duty is 36,738 kW. Chilled water at 10 ºC will be used as the cooling agent the exit temperature of the chilled water is set to 40ºC.

The design of the beer column condenser is carried out in accordance with the Section 8.7.2.1. Since the operating conditions are close to each other the overall heat transfer coefficient will be taken as 1,229 W m^{-2} $^{o}C^{-1}$. The LMTD is also same with the value 19.9 ºC. The specifications of E-304 are given in Table 9.27.

Table 9.27 Design specifications for E-304

Parameter	Unit	E-304
Number in series		1
Streams in		<312>
Streams out		<312>
Heat duty	kW	36,738
Temperature	°C	20
OHTC	W m^{-2} $^{o}C^{-1}$	1,229
LMTD	°C	19.9
Heat transfer area	m^2	1,502
Cooling agent		Chilled Water
Cooling agent flow	t/h	1,056.2
Pressure	atm	1

9.6.3.2. Beer Column Reboiler (E-303)

The design of reboiler is carried out in similar manner with conventional plant beer column reboiler. The total amount to be reboiled is 264,226 kg/h. The boiling media is assumed to be pure water. The temperature of the input stream is 100.1ºC. Low pressure steam will be used as the heating agent. The reboiler heat duty is 165,553 kW. The temperature, pressure and the enthalpy of condensation of low pressure steam is 130.7 ºC, 275 kPa and 2,171 kJ/kg respectively. Since the operation conditions are same with 2005 plant beer column the overall heat transfer coefficient is selected as 1,284 W m^{-2} $^{o}C^{-1}$ as calculated in Section 8.7.2.2 and Appendix A.8.10.2. The design specifications are given in Table 9.28,

Table 9.28 Design specifications for E-303

Parameter	Unit	E-303
Number in series		1
Streams in		<311>
Streams out		<311>
Heat duty	kW	165,553
Temperature	°C	100
OHTC	W m^{-2} $^{o}C^{-1}$	1,281
LMTD	°C	30.7
Heat transfer area	m^2	4,168
Heating agent		Saturated steam at 130.7 ºC
Steam flow	t/h	274.5
Pressure	atm	1

9.6.4. Rectification Column (C-303)

The design of rectification column is carried out in similar manner with 2005 design as described in Section 8.7.3 and Appendix A.8.11. The results are presented in Tables 9.29-30.

Table 9.29 Design specifications for C-303

Parameter	Unit	C-303
Number in parallel		1
Streams in		<310>
Streams out		<314><318>
Number of stages		36
Reflux ratio		4(mass)
Height of stages	m	1
Feed stage		25
Tray type		Nutter Float valve
Tray efficiencies		0.57
Section pressure drop	bar	0.27
Condenser duty	kW	135,558
Distillate temperature	ºC	77.4
Reboiler duty	kW	25,728
Bottoms temperature	ºC	99.8

Table 9.30 Tray information for C-303

Stage	Diameter (m)	Total Area (m^2)	Active Area (m^2)	Down comer Area (m^2)	Liquid flow (kg/hr)	Vapor flow (kg/hr)
2	6.78	36.13	28.90	3.61	408,835	513,238
3	6.75	35.79	28.63	3.58	406,019	511,482
4	6.72	35.44	28.35	3.54	403,040	508,666
5	6.68	35.08	28.06	3.51	399,864	505,688
6	6.65	34.70	27.76	3.47	396,438	502,511
7	6.61	34.29	27.43	3.43	392,701	499,086
8	6.56	33.85	27.08	3.38	388,578	495,349
9	6.52	33.36	26.69	3.34	383,977	491,226
10	6.47	32.83	26.27	3.28	378,789	486,625
11	6.41	32.24	25.79	3.22	372,882	481,436
12	6.34	31.57	25.26	3.16	366,110	475,530
13	6.26	30.82	24.65	3.08	358,310	468,757
14	6.18	29.96	23.97	3.00	349,330	460,958
15	6.10	29.21	23.37	2.92	339,063	451,978
16	6.01	28.36	22.69	2.84	327,527	441,711
17	5.90	27.37	21.90	2.74	314,996	430,174
18	5.79	26.29	21.03	2.63	302,183	417,643
19	5.67	25.21	20.17	2.52	290,274	404,831
20	5.56	24.29	19.44	2.43	280,188	392,922
21	5.48	23.57	18.85	2.36	270,540	382,835
22	5.39	22.83	18.26	2.28	254,962	373,188
23	5.26	21.74	17.39	2.17	224,295	357,610
24	5.14	20.75	16.60	2.07	193,900	326,942
25	2.36	4.36	3.49	0.44	193,664	61,930
26	2.35	4.33	3.47	0.43	192,815	61,312
27	2.34	4.29	3.43	0.43	191,585	60,462
28	2.32	4.23	3.39	0.42	189,864	59,232
29	2.30	4.16	3.33	0.42	187,624	57,511
30	2.28	4.07	3.25	0.41	184,985	55,272
31	2.25	3.99	3.19	0.40	182,217	52,633
32	2.23	3.92	3.13	0.39	179,646	49,864
33	2.22	3.86	3.09	0.39	177,518	47,294
34	2.20	3.82	3.05	0.38	175,918	45,165
35	2.20	3.78	3.03	0.38	174,794	43,565

9.6.4.1. Rectifying Column Condenser (E-306)

The condenser operates at atmospheric pressure. The vapor enters the condenser saturated at 78.2 ºC and leaves at 77.5 ºC. The condenser heat duty is 135,558 kW. The inlet and outlet temperatures of cooling water are set to 25 and 60 ºC respectively. The average heat capacity of water between 25-60ºC is 4.164 kJ/kgºC. The design of the beer column condenser is carried out in accordance with the Section 8.7.3.1 & Appendix A.8.11.1. Since the operating conditions are close to each other the overall heat transfer coefficient will be taken as 845 W m^{-2} $^{o}C^{-1}$. The LMTD is also same with the value 32.4 ºC. The specifications of (E-306) are presented in Table 9.31.

Table 9.31 Design specifications for E-306

Parameter	Unit	E-306
Number in series		1
Streams in		<312>
Streams out		<312>
Heat duty	kW	135,558
Temperature	°C	77.4
OHTC	W m^{-2} $^{o}C^{-1}$	845
LMTD	°C	32.4
Heat transfer area	m^2	4,951,4
Cooling agent		Cooling water (25-60ºC)
Cooling agent flow	t/h	3,348.5
Pressure	atm	1

9.6.4.2. Rectification Column Reboiler (E-305)

The design of reboiler is carried out in similar manner with conventional plant beer column reboiler. The total amount to be reboiled is 42,084 kg/h. The boiling media is assumed to be pure water. The temperature of the input stream is 100ºC. Low pressure steam will be used as the heating agent. The reboiler heat duty is 25,728.6 kW. The temperature, pressure and the enthalpy of condensation of low pressure steam is 130.7 ºC, 275 kPa and 2,171 kj/kg respectively. Since the operation conditions are same with 2005 plant beer column the overall heat transfer coefficient is selected as 1,281 W m^{-2} $^{o}C^{-1}$ as calculated in Section 8.7.2.2 and Appendix A.8.10.2. The design specifications are given in Table 9.32,

Table 9.32 Design specifications for E-305

Parameter	Unit	E-305
Number in series		1
Streams in		<311>
Streams out		<311>
Heat duty	kW	25,729
Temperature	°C	100
OHTC	W m^{-2} $^{o}C^{-1}$	1,281
LMTD	°C	30.7
Heat transfer area	m^2	653.4
Heating agent		Saturated steam at 130.7 ºC
Heating agent flow	t/h	46.1
Pressure	atm	1

9.6.5. Molecular Sieve Feed Cooler (E-307)

This cooler is operated to cool the condensate from the rectification column at 77.4 ºC to adsorption temperature of 60ºC. The design specifications of the heat exchanger are listed in Table 9.33. Using Equations A.8.11 &12 the mean heat capacity of the ethanol water mixture is obtained as 3.231 kJ kg^{-1} $^{o}C^{-1}$ in the temperature interval 60-77.4ºC.

Table 9.33 Design specifications for E-307

Parameter	Unit	E-307
Number in series		1
Streams in		<314>
Streams out		<315>
Heat duty	kW	1,603
Temperature	°C	77.4-60
OHTC	W m^{-2} $^{o}C^{-1}$	500
LMTD	°C	39.2
Heat transfer area	m^2	81.8
Cooling agent		Cooling water 25 to 40ºC
Water flow	t/h	92.2
Pressure	atm	1

9.6.6. Molecular Sieve Adsorption Column (C-304)

The details of the design calculations of molecular sieve adsorption column are described in Appendix A.9.4. The design specifications for adsorption column is listed in Table 9.34.

9.6.6.1. Cooling

Cooling is required to remove the heat of adsorption and to keep the adsorbent bed temperature at 60ºC. Details of the cooling system design can be found in Appendix A.9.4.1

9.6.6.2. Regeneration

Due to the reasons states in Section 8.7.5.2 the regeneration temperature is set 200 ºC and the regenerate gas is nitrogen. The details for the regeneration can be found in Appendix A.9.4.2.

Table 9.34 Design specifications for C-304

Parameter	Unit	C-304
Number in parallel		2
Streams in		<315>
Streams out		<316><320>
Packing type		3 Å zeolites
Packing size	m	0.005
Diameter of column	m	3.2
Height of column	m	8.4
Packing amount	ton	49.3
Operation temperature	ºC	60
Product purity	%	99.89
Regenerant		Nitrogen
Regeneration temperature	ºC	200
Cooling & heating area	m^2	436
Cooling water requirement	ton/h	144.6
Steam requirement	ton/h	8.75

9.6.7. Wet Nitrogen-Dry Nitrogen Heat Exchanger (E-308)

This heat exchanger is operated to heat up the dry nitrogen to 180ºC while cooling the wet nitrogen to 30ºC. This helps saving both cooling and heating agent usage. The design calculations for the heat exchanger can be found in Appendix A.9.4.2. The design specifications are presented in Table 9.35.

9.6.8. Water condenser (E-309)

This condenser is operated to condense water to dry nitrogen. Refrigerant is used to cool the wet nitrogen to 5ºC. The design calculations for the heat exchanger can be found in Appendix A.9.4.2. The design specifications are presented in Table 9.36.

Table 9.35 Design specifications for E-308

Parameter	Unit	E-308
Number in series		1
Streams in		<320><324>
Streams out		<321><325>
Heat duty	kW	6,158
Temperature	°C	200-30 5-180
OHTC	W m^{-2} $^oC^{-1}$	300
LMTD	°C	23
Heat transfer area	m^2	893
Pressure	atm	1

Table 9.36 Design specifications for E-309

Parameter	Unit	E-309
Number in series		1
Streams in		<321>
Streams out		<322><334>
Heat duty	kW	1,977
Temperature	°C	5
OHTC	W m^{-2} $^oC^{-1}$	250
LMTD	°C	14
Heat transfer area	m^2	565
Cooling agent		Refrigeration
Pressure	atm	1

9.6.9. Product Cooler (E-310)

This cooler is operated to cool the product ethanol to from 60ºC to 30ºC. The average heat capacity of ethanol and cooling water at given interval is 2.955 and 4.17 kJ $kg^{-1o}C^{-1}$. The design specifications are presented in Table 9.37.

Table 9.37 Design specifications for E-310

Parameter	Unit	E-310
Number in series		1
Streams in		<316>
Streams out		<317>
Heat duty	kW	2,340
Temperature	°C	60-30
OHTC	W m^{-2} $^oC^{-1}$	500
LMTD	°C	10.8
Heat transfer area	m^2	433.3
Cooling agent		Cooling water
Cooling water flow	ton/h	134.6
Pressure	atm	1

9.7. Cogeneration

In 2015 design, in addition to burning the fibers to get energy, the cogeneration unit also has the function of burning the genetically modified yeasts, filter cakes and the concentrated vinasse to reduce the impacts on environment. The structure of the cogeneration unit is the same the microbial biomass (w/w 80% moisture) and lignin residues (w/w 48% moisture) are passed through drying chamber to reduce their moisture content to 12%. The vinasse is first concentrated in an evaporator to w/w 32% then sent to furnace together with dried yeast, filter cake and plant biomass residues. The analysis revealed that the amount of energy released in cogeneration is not sufficient to produce electricity demand of the plant. Therefore, the electricity gap is closed by purchasing from the grid.

9.7.1. Evaporator (E-403)

This evaporator is operated to concentrate the vinasse before being sent to boilers to be combusted. The details of the design of E-403 can be found in Appendix A.9.5.1 The specifications are given in Table 9.38

Table 9.38 Design specifications for E-403

Parameter	Unit	E-403
Number in series		1
Streams in		<311>
Streams out		<429><430>
Heat duty	kW	19,374.8
Temperature	°C	103
OHTC	$W\ m^{-2}\,^{o}C^{-1}$	979
LMTD	°C	27.7
Heat transfer area	m^2	714.4
Heating agent	Saturated steam at 275 kPa	
Heating agent flow	t/h	32.3
Pressure	atm	1

9.7.2. Fiber, Yeast and Filter Cake Drier (D-401)

This unit is operated to dry the wet yeast, filter cakes and fiber residues before being sent to cogeneration. The details of the calculation can be found in Appendix A.9.5.2. The design specifications are presented in Table 9.39.

Table 9.39 Design specifications for D-101

Parameters	Units	D-201
Numbers in parallel		3
Streams in		<431><435>
Streams out		<432> <436>
Air requirement		Flue gas from furnace
Evaporation rate	$kg\ m^{-2}h^{-1}$	65
Evaporation area	m^2/drum	485
Diameter	m	4
Power requirement	kW/drum*	188

*Obtained by extrapolation using the values given in Appendix A.8.5.

9.7.3. Heating Value of Feed and Steam Production

The details of the calculations for heating value of the fuels, steam production and the boiler efficiency calculations can be found in Appendix A.9.5.2. The results are given in Table 9.40.

Table 9.40 Results for fuel combustion calculations

Parameter	Unit	Value
Heat supplied by fuel	MW	453.1
Excess air requirement	ton/h	912.6
Exhaust gas temperature	^{o}C	350.0
Heat available for steam production	MW	354.37
Steam produced		
130ºC, 2.75 bar	ton/h	512.0
525ºC, 86 bar	ton/h	144.9

9.7.4. Economizer

The economizer is the heat exchanger to recover the heat from the exhaust gas to heat up the fresh air. The details of the design can be found in Appendix A.9.5.2.

9.7.5. Furnace Dimensions

The furnace volume is calculated as 1,536 m^3 using the same methodology with Section 8.8.8.2. The area requirement for heat exchanging and evaporation is calculated as 13,437 m^2 using evaporation rate of 49 kg $m^{-2}h^{-1}$ reported by Hugot (1972).

9.7.6. Chimney

The height of the chimney is recommended to be between 30-60 meters (Hugot, 1972). The height is assumed as 40 m. The velocity in the chimney should be between 4.5-5 m/s for proper operation. The total gas flow in the plant (gas from fermentors, sulphur and limestone burners and the furnace) is obtained as ~308.6 kg/s. which corresponds to 496 m^3/s. The diameter of the chimney is calculated as 11.2 m to obtain the specified gas flow rates.

9.7.7. Turbines

Multistage turbines are utilized for electricity production. The outlet steam pressure and temperature are 25 bar and 275ºC respectively. Following equation is used to calculate the amount of the electricity produced.

$$P_E = \eta_t \eta_m \eta_a (H_i - H_o) m_s$$

Where, P_E is electricity produced (kW), η_t, η_m, η_a, are respectively turbine (0.66), mechanical (0.98) and electrical alternator (0.97) efficiencies, H_i and H_o are the inlet (3,453 kJ/kg) and outlet steam (2,947.4 kJ/kg) enthalpies. Inserting the variables in to equation yields electrical power production of

$$P_E = 0.66 \times 0.98 \times 0.97 \times (3,453 - 2,947.4) kJ/kg \times 40.24 kg/s = 13.52 MW$$

The efficiencies are taken from research of Damen (2001).

Table 9.41 Results for electricity generation

Parameter	Unit	Value
Super heated steam temperature	ºC	525
Super heated steam pressure	bar	86
Steam for electricity	ton/h	144.9
Electricity production	MW	13.52

9.8. Equipments Manipulated in 2015 Design

Due to the manipulations in the design, design of some of the equipments in 2005 plant needs to be changed. The limestone processing capacity is changed and the sizing is to be renovated; lime milk capacities changed, thus, the lime milk preparation tanks are to be redesigned; the temperature of the condensates returning to boilers changed, thus, the reheaters before and after the boiling pans are to be redesigned. And, the juice heaters are no longer heated using steam instead they are heated using the exhaust gas from the boilers (F-401). The calculations are carried out in same manner with the 2005 design.

9.8.1. Lime Kiln (F-101)

The same methodology with Section 8.2.1 is used. For the current limestone processing capacity of 3330.3 kg/h, the required operating volume for the kiln is calculated as 71.3 m^3. In lime kiln, natural gas will be used as heating agent. The required amount of energy is 1,957.1 kW. Thus, 0.187 ton/h natural gas is required for the operation. The design specifications are presented in Table 9.42.

Table 9.42 Design specifications for F-101

Parameter	Unit	F-101
Number in series		1
Streams in		<127><128><129>
Streams out		<130><131><145>
Capacity	kg/h	3,330.3
Volume	m^3	71.3
Natural gas requirement	kg/h	187
Operating pressure	atm	1
Operating temperature	ºC	898

9.8.2. Lime Milk Preparation Vessels (V-101)

The volume flow in to vessel is 12.4 m^3/h. The specifications of the vessel are given in Table 9.43.

Table 9.43 Design specifications for V-101

Parameter	Unit	V-101
Number in parallel		2
Streams in		<121><126>
Streams out		<132>
Volume	m^3	20.5
Impeller speed	rpm	10
Power requirement	kW	20.5
Cooling area	m^2	18.7
Cooling water	ton/h	24.6
Operating pressure	atm	1
Operating temperature	ºC	90

9.8.3. Juice Heaters (E-102&E-103)

Since the plant is not producing enough steam to meet the demand, the juice heaters heated using the exhaust gas from the furnaces which is at 125 ºC after leaving the drying chamber. The major issue that needs attention is the highly corrosive nature of the exhaust gas due to the presence of the water vapor and carbon dioxide together. Therefore, the heat exchangers are to be made of stainless steel. The exhaust gas contains 160 ton/h of water vapor. It is assumed that the heating will be due to the condensation of the water vapor present in exhaust gas. The design specifications are presented in Table 9.44.

Table 9.44 Design specifications for E-102&E-103

Parameter	Unit	E-101	E-102
Number in series		1	1
Streams in		<108>	<116>
Streams out		<109>	<117>
Heat duty	kW	47,257	25,648
Temperature	°C	45.2-103	44-103
OHTC	$W\ m^{-2}\ ºC^{-1}$	2,000	2,000
LMTD	°C	44.9	45.3
Heat transfer area	m^2	527	283
Heating agent		125 ºC Furnace flue gasses	
Heating agent requirement*	t/h	78.4	42.5
Pressure	atm		1

* Steam present in the exhaust gas

9.8.4. Reheaters (E-209, E-210, E-211&E-212)

The heating agent for the reheaters is the condensates from the heating equipment (130.7-120 ºC). The specifications for the reheaters can be found in Table 9.45.

Table 9.45 Design specifications for E-209-212

Parameter	Unit	E-209	E-210	E-211	E-212
Number in series		1	1	1	1
Streams in		<201>	<210>	<213>	<217>
Streams out		<202>	<211>	<214>	<218>
Heat duty	kW	2,008	1,100	995.5	484
Temperature	°C	56-75	40-55	55-75	40-55
OHTC	$W\ m^{-2}\ {}^{o}C^{-1}$	80	80	80	80
LMTD	°C	59.7	77.8	60.2	77.8
Heat transfer area	m^2	421	178	207	78
Heating agent	Condensates from heating equipment 130.7-120ºC				
Heating agent requirement	t/h	175.7	96.3	87.1	42.4
Pressure	atm	1	1	1	1

10. Process Safety

Since the sugar ethanol plants are widely applied well known processes, in this section only the pretreatment section of the future plant will be analyzed in terms of safety and hazard.

10.1. Dow Fire and Explosion Index (FEI)

10.1.1. Introduction to Dow FEI and Selection of the Material Factor

The Dow Fire and Explosion Index is an evaluation method to study, which is the degree of hazard in case of possible fire or explosion occurs in a process. After the evaluation, correct actions can be taken in case the process has a considerable risk. If the degree of hazard is high, modifying the process or changing some parts completely is necessary. By applying the FEI, it is easier to identify which parts of the process have a higher risk (*Dow's Fire and Explosion Index Hazard Classification Guide, 7th edition, AIChE Technical Manual*).

The following table indicates the different degrees of hazards:

Table 10.1 Degree of hazard classification (Sinnot,1999)

FEI range	Degree of hazard
1-60	Light
61-96	Moderate
97-127	Intermediate
128-158	Heavy
>159	Severe

To apply the method two factors are considered: the material factor and the unit hazard factor. Material factor (MF) is the intrinsic rate of energy release from the burning, explosion, or other chemical reaction of the material. Unit hazard factor is the result of two products that take into account the general and special process hazards. The result of the product of these two factors will give the hazard value to the unit or process evaluated.

According to the Dow method, the component with the highest material factor that is present in significant quantities is the one used for calculating the Fire and Explosion Index.

In this process, only the ethanol and acetic acid are relevant for this evaluation; all the other components are present in a very low flow or their risk is irrelevant. The characteristics of these two components are shown in the following table

Table 10.2 Selection of the material factor (Sinnot, 1999)

Component	MF	Heat of combustion (MJ/kg)	N_h	N_f	N_r	Flash Point (°C)	Boiling point (°C)
Ethanol	16	26.75	0	3	0	12.8	78.0
Acetic acid	10	13.03	3	2	0	42.8	118.3

The selected component to be evaluated is ethanol. The reason for this selection is the high value of the material factor and its highest flow in the process.

10.1.2. Results and Conclusions from Dow FEI

In this section, the evaluation for the Pretreatment and hydrolysis and ethanol production and purification sections are considered and final results are summarized in Table 10.3.

- Unit: Pretreatment and hydrolysis and ethanol production and purification sections
- Material factor: 16

- ***General process hazards:*** The following items may increase the magnitude of a probable incident.
    A. *Exothermic chemical reactions:* Only exothermic reactions are the neutralization of the hydrolyzate solution and the hydrolysis of the lignocellulosic material the penalty factor is taken as 0.5

    B. *Endothermic processes:* Not applicable.

    C. *Material handling and transfer:*

    -Loading and unloading of Class I flammable material (flash point less then 38.7 ºC) penalty is 0.5

    -Use of centrifuges batch reactions or batch mixing in semi-open containers requires a 0.5 penalty. These process units are present in the system however the concentration of ethanol is well below flammability limits. Therefore, this factor is taken as 0.1.

    -Warehouse and yard storage covers general product storage other then tank farms and product hold tanks. Therefore, this factor is not applicable for the current set up.

    D. *Enclosed or indoor process units:* Not applicable

    E. *Access:* Adequate access will be provided, thus there is no penalty

    F. *Drainage and spill control:* Adequate drainage will be provided, thus no penalty is applied.

- ***Special process hazards:*** The following items may contribute to incidents that may increase the probability of a fire or explosion.

A. *Toxic material:* Sulphuric acid and acetic acid are substances that can cause irritation during inhalation and for eyes and skin. The phenolic lignin degradation products are toxic in case of skin contact. In case of ingestion, they can cause more severe injuries. Especially the concentration of sulphuric acid is high and at hazardous levels. Therefore, the factor for toxic materials is taken as 0.5.

B. *Sub-atmospheric pressure:* There aren't any units operating at sub-atmospheric pressures in these sections. Thus, the factor is taken as 0.

C. *Operation in or near flammable range:* Ethanol is stored in floating head storage tanks which reduces the risk, penalty is 0. In case of process upset and failures, the penalty is taken as 0.3. Necessary purging is assumed to be applied to the process equipment where necessary. For barge or tank car uploading the penalty is taken as 0.4 assuming that padding is used.

D. *Dust explosion:* Not applicable

E. *Pressure:* Ethanol is not subject to high pressure operations. Therefore, the penalty is taken as 0.

F. *Low temperature:* the whole process operates above 10°C, the value that corresponds to the transition temperature of the carbon steel. Thus, in this case there is no penalty applied.

G. *Quantity of flammable/unstable material:*

- Liquids or gasses in process: this category applies to flammable or combustible liquids (flash point less then 60ºC) in process units. The maximum flow of ethanol is ~92 ton/h. Considering the maximum amount of the ethanol and the heat of combustion the penalty is found in function of the potential energy release.

Maximum amount of Ethanol: 26.4 kg/s

Heat of combustion = 26.75 MJ/kg

Potential energy release = 26.37 kg × 26,750 kJ/kg × 1.055 Btu/kJ = 0.0007×10^9 Btu

This value is not detectable in the graph, thus, the penalty factor is 0

- Liquids or gasses in storage : This category applies to flammable or combustible liquids (flash point less then 60ºC) in storage. The capacity of the storage tanks are 3,985 tons each.

Maximum amount of Ethanol: 3,985,000 kg

Heat of combustion = 26.75 MJ/kg

Potential energy release = 26.37 kg × 26,750 kJ/kg × 1.055 Btu/kJ = 112.5×10^9 Btu

Thus the penalty factor is 1.1

H. *Corrosion and erosion*: Corrosion is possible due not only to the presence of some ions such as OH^- and SO_4^{2-} but also because of the corrosive character of the acetic acid. Hence the penalty considered is 0.20

I. *Leakage-joints and packing:* the process will be designed to minimize the possibly leakages. Thus a minimum factor is applied: 0.10

J. *Use of fired heaters:* The furnaces are assumed to be far enough from the storage tanks, therefore, the minimum penalty is taken as 0.1.

K. *Hot oil heat exchange system:* Heat exchange system only uses natural gas and steam as heating medium, hence this penalty is not applicable

L. *Rotating equipment:* Not applicable

The calculated value for the FEI is 124.32, as it can be checked in Table 10.3. According to the classification specified in Table 10.1 the process has a intermediate degree of hazard.

Using the unit hazard factor (7.77) and the material factor (16) the damage factor is obtained as 66% using the graph given in Dow Guide. The radius of exposure of any possible damage is read as 31.3 m. These results indicate that the current process units represent a 66% damage probability to surrounding circle of diameter 31.3 m. the most hazardous units are the storage tank fields therefore the location of the process equipments should be considered accordingly and the distances should be determined based on minimum of 31.2 m from the storage vessels to reduce the effect of explosions to the other process units.

Ethanol is Class I flammable liquid and requires special precautions for the handling and operation. The process equipment should be designed considering the flammability of ethanol.

Table 10.3 Dow and Fire Explosion Index evaluation

DOW FIRE AND EXPLOSION INDEX (FEI)	LOCATION Sao Paolo		DATE 28-09-2005
PLANT Ethanol production	PROCESS UNIT Pretreatment and hydrolysis and ethanol production and purification sections		
PROCESS: Production of ethanol from sugar cane and sugar cane bagasse			
MAIN MATERIALS IN THE PROCESS: Ethanol, carbon dioxide, acetic acid, sulphuric acid, glycerol, biomass, Ca(OH)$_2$ water,			
STATE OF OPERATION: DESIGN	Basic Material For MF: Ethyl alcohol		
MATERIAL FACTOR (MF)			**16**
1. GENERAL PROCESS HAZARD	Penalty range	Penalty used	
BASE FACTOR	1.00	1.00	
A. EXOTHERMIC CHEMICAL REACTIONS	0.30 to 1.25	0.50	
B. ENDOTHERMIC PROCESSES	0.20 to 0.40	-	
C. MATERIAL HANDLING AND TRANSFER	0.25 to 1.05	0.60	
D. ENCLOSED OR INDOOR PROCESS UNITS	0.25 to 0.90	–	
E. ACCESS	0.20 to 0.35	–	
F. DRAINAGE AND SPILL CONTROL	0.25 to 0.50	–	
GENERAL PROCESS HAZARDS FACTOR (F_1)		**2.10**	
2. SPECIAL PROCESS HAZARDS			
BASE FACTOR	1.00	1.00	
A. TOXIC MATERIALS	0.00 to 0.80	0.50	
B. SUB-ATMOSPHERIC PRESSURE (<500 mmHg)	0.50	–	
C. OPERATION IN OR NEAR FLAMMABLE RANGE			
1. Tanks farms and storage flammable liquids	0.50	–	
2. Process upset or purge failure	0.30	0.30	
3. Always in flammable range	0.80	0.40	
D. DUST EXPLOSION	0.25 to 2.00	–	
E. PRESSURE: OPERATING PRESSURE		–	
F. LOW TEMPERATURE	0.20 to 0.30	–	
G. QUANTITY OF FLAMMABLE / UNSTABLE MATERIAL:			
QUANTITY: 4 230 lbs Hc = 5 600 BTU/lb			
1. Liquids, gases and reactive materials in process		–	
2. Liquids and gases in storage		1.10	
3. Combustible solids in storage dust in process		–	
H. CORROSION AND EROSION	0.10 to 0.75	0.20	
I. LEAKAGE-JOINTS AND PACKING	0.10 to 1.50	0.10	
J. USE OF FIRED EQUIPMENTS		0.10	
K. HOT OIL HEAT EXCHANGE SYSTEM	0.15 to 1.15	–	
L. ROTATING EQUIPMENT	0.50	–	
SPECIAL HAZARD FACTOR (F_2)		**3.70**	
UNIT HAZARD FACTOR (F_1 x F_2 = F_3)		**7.77**	
FIRE AND EXPLOSION INDEX (F_3 x MF = F&EI)			**124.32**

10.2. Hazard and Operability Study (HAZOP)

The hazard and operability study is an evaluation method to indicate the potential hazards that may arise from deviations from the intended design conditions.

The following table indicates different expressions that are used to evaluate each part of the process and their meanings (Sinnot, 1999):

Table 10.4 Standard guide-words and their generic meanings

Guide word	Meaning
No or not	None of the design intent is achieved
More	Quantitative increase in a parameter
Less	Quantitative decrease in a parameter
As well as	All the design intentions are achieved and an additional activity takes place
Part of	Only some of the design intentions are achieved
Reverse	The logical opposite of the intention occurs
Other than	No part of the initial intention occurs. Something quite different takes place

The typical process parameters that may present some deviation are: flow, pressure, temperature, mixing, level, reaction, separation, operate and maintain. The HAZOP study is carried out for the main equipments of the process and the tables including the evaluation are listed as follows.

The major hazard factors in pretreatment section are the pretreatment reactor which operates at high pressures and temperatures and the acidic media circulating in the system. Therefore, the pretreatment reactor and the factors that might cause spill or overflow of acidic media are analyzed according to HAZOP criteria.

Vessel: Pretreatment reactor (R-401) and collection tank (V-401)
Intention: hydrolysis of the bagasse at 14 bar and 190°C

Table 10.5 HAZOP analysis for the pretreatment reactor and collection tank

Guide word	**Deviation**	**Cause**	**Consequences and actions**
Stream <403> transfers steam in the pretreatment reactor			
No or not	Pressure	Steam transfer failure, leakage	(1) There might be problem with the steam inflow. Install a flow controller. There isn't any hazard factor, but, process efficiency would drop in this case. In case of leakage, the high pressure and temperature steam will injure the operators. The operators should be supplied with necessary protection. (2) Install a sound alert of no flow which will inform the operator to check the reason.
No or not	Flow	Valve failure, or line fracture	
Less	Pressure	Low steam flow, valve failure or line fracture	See (1) and (2)
Less	Flow		
More	Flow	Increase in steam inflow	(3) If the amount of steam increases, the pressure and temperature inside the reactor will increase and cause explosion hazard, install flow controller to steam line and pressure& temperature controllers to reactor.
Reverse	Flow	Fall in line press, high pressure in reactor	(4) Install control valve to prevent reverse flow Also, see (1) and (2)

Table 10.5 (Continuing) HAZOP analysis for the pretreatment reactor and collection tank

Guide word	Deviation	Cause	Consequences and actions
Stream <404> supplies biomass			
No or not	Flow/pressure	Pump, valve failure Clogging Line leakage	(5) In case of line leakage, the acidic ingredients will spill and cause hazard for the operators. In case of low pressure, reverse flow will occur and the reactor ingredients will flow of the reactor, install one way valves to prevent reverse flow. Install flow controller and sound alert, that operator will be informed about the oscillations from the set points. Install proper drainage to let the spills flow to spill collection places. Supply the necessary emergency precautions in case of injuries. The line might be clogged due to the fiber, supply regular maintenance to the process equipment
Less	Flow/pressure		
More	Flow/pressure	Pump, valve failure	Install a flow controller and sound alert. The high flow rate will increase the pressure inside the reactor and can cause spills or explosions.
Pretreatment outlet nozzle			
No or not	Flow	-Nozzle clogged -No Input flow	(6) If the nozzles are clogged the pressure inside the reactor will increase and cause explosion hazard. To prevent this situation the outlet flow from the reactor should be monitored by flow controllers. - In case of no flow see (1) and (2)
Less	Flow	-Nozzle clogged -No Input flow	See (6)
More	Flow	The inlet flow has over pressure	Increase in outlet flow indicates pressure rise inside the reactor and might cause explosion hazard

The remaining equipments in pretreatment and hydrolysis section are vessels and the major hazard arising is the risk of spill or over flow. This is why a general HAZOP analysis will be carried out for the vessel equipments and the actions to be taken are discussed.

Vessel: Process equipment.
Intention: Mixing, neutralization, dehydrating.

Table 10.6 HAZOP analysis for the anonymous process equipments with spill and over flow risks.

Guide word	Deviation	Cause	Consequences and actions
In flow to equipment			
No or not	Pressure	Valve/pump failure, or line fracture	(1) In case of line leakage, the acidic ingredients will spill and cause hazard for the operators. In case of low pressure, reverse flow will occur and the equipment ingredients will flow of the equipment, install one way valves to prevent reverse flow. Install flow controller and sound alert, that operator will be informed about the oscillations from the set points. Install proper drainage to let the spills flow to spill collection places. Supply the necessary emergency precautions in case of injuries. The line/pump might be clogged due to the fiber, apply a regular maintenance to the equipments.
No or not	Flow		

Table 10.6 (continuing) HAZOP analysis for the anonymous process equipments with spill and overflow risks.

Guide word	**Deviation**	**Cause**	**Consequences and actions**
Less	Pressure	Low flow, valve/pump failure or line fracture	See (1)
Less	Flow		
More	Flow	Pump/Valve failure	(2) If the amount of flow increases, the level inside the vessel will increase and cause overflow hazard, install flow controller to line and level controller to the equipment if applicable.
Reverse	Flow		Install one way valve
Out flow from the equipment			
No or not	Flow/pressure	Pump, valve failure Clogging Line fracture	(3) In case of line leakage, the acidic ingredients will spill and cause hazard for the operators. In case of pump or valve failure the flow will be stopped and the liquid level inside the equipment will rise which will cause overflow. To prevent this level controller should be installed to the equipment and the equipments should be maintained frequently to prevent the failure of clogging. Sufficient drainage should be installed to have controlled flow of the spill. Necessary precautions should be taken in case of contact of operators with acidic media.
Less	Flow/pressure		
More	Flow/pressure	Pump, valve failure	Install a flow controller and sound alert. The high flow rate will decrease the liquid level inside the equipment and can cause insufficient residence times to complete the task.

Based on both FEI and HAZOP analysis, the importance of installation of a good control systems and hazard analysis is verified. In this way, it is more probable to get the design specifications and during the initial design it is also possible to improve the sections in which the process may present higher risks.

It is also of interest to develop a maintenance program for all the equipments, with special interest in those equipments that present a certain hazard during the operation or that may have an important influence in the process in case of failure.

Following these initial guidelines and taking into account the safety of the plant during an initial step, it is possible to minimize the accidents and situations under control during the operation of the plant.

11. Wastes

One of the objectives of this project is to reduce the impact to the environment to the minimum. As it is previously mentioned in current design the generated wastes (filter cake and vinasse) are sent back to fields as fertilizer without any treatment. But, considering that the filter cake contains calcium salts and the vinasse is acidic and has high COD (chemical oxygen demand) releasing these to environment does not comply with the sustainable operation. On the other hand, the exhaust gas from the cogeneration unit contains NO_x compounds and SO_2 the emission of this gas is should also be minimized to reduce the impact of air pollution. In the following sections the wastes generated in 2015 plant are mentioned and some solutions are suggested to deal with the problem.

11.1. Waste Water

The water output of the multi effect evaporators (E-201 to E-208) and vacuum pans (E-207 and 208) contains sulphur dioxide and volatile sugar cane organics in minor fractions. Since these streams are recycled to system and all the impurities present in these streams do not interfere with the normal operation, there is no need for treating them in waste water treatment. Therefore, the Streams <226>, <227> and <228> can directly be recycled to the process as process water. The Stream <140> which is the flash vapor from the flash drums V-104 and V-105 is water vapor and can be released to atmosphere.

The waste acid solution (Stream <332>) from the yeast acidification tank (V-101) is neutralized using lime and the generated gypsum is sent to land fill. The design of the equipment for this operation is outside the battery limits, but, the investment requirement and operational costs are included in economical evaluations (for details see Section 13.4.2.3 and Appendix A.11.5.2). The operation will be composed of 3 steps; first blending the acid solution with lime milk of required amount (~0.57 g CaO per g H_2SO_4) the mixture is blended for 30 min then send to settling. The clarified water is send back as process water and the gypsum is filtered and sent to land fill.

The remaining waste water streams are the condensed water (Stream <334>) from molecular sieve nitrogen condenser (E-309), the flash vapor (Stream <405>) from the pretreatment collection tank (V-401) and the organic rich vapor (Stream <429>) from the vinasse concentrating evaporator (E-403). The organic rich vapors are first super heated in boilers to be used as process steam. Then the condensates are treated in waste water treatment plant.

The proposed system for the waste water treatment is anaerobic digestion followed by aerobic digestion. Within anaerobic digestion, 90% of organic material is converted in to methane and carbon dioxide. Methane is produced at a maximum yield of 0.229 kg/ kg COD (at 25ºC) removed. One mole of carbon dioxide is produced per 3 mole of methane generated. Cell mass is produced at a yield of 30 g cell mass for each one kg of COD that is removed. All the sulphates entering to the anaerobic digestion are converted into hydrogen sulphide.

The effluent from the anaerobic digestion is sent to aerobic digestion where 90% of the remaining COD is removed. 60% of COD is converted into water, carbon dioxide nitrous oxide and sulphur dioxide and the remaining 30 percent is converted into cell mass. 50 % excess oxygen then COD is supplied into lagoon. Treated material is filtered to remove the sludge and the clarified water is recycled to the process. Part of the sludge is recycled to the anaerobic digestion to increase the cell load and the remaining is disposed in combustors (Alan *et al.*, NREL report, 2002).

The organic matter content in produced total waste water is calculated based on total COD using the following COD expressions.

Ethanol : $C_2H_6O + 3O_2 \rightarrow 2CO_2 + 3H_2O$ - COD = 2.0870 g O_2/g Eqn. 11.1

Acetic Acid : $C_2H_4O_2 + 2O_2 \rightarrow 2CO_2 + 2H_2O$ - COD = 1.0657 g O_2/g Eqn. 11.2

Glycerol : $C_3H_8O_3 + 3.5O_2 \rightarrow 3CO_2 + 4H_2O$ - COD = 1.2174 g O_2/g Eqn. 11.3

Furfural : $C_5H_4O_2 + 5O_2 \rightarrow 5CO_2 + 2H_2O$ - COD = 1.6653 g O_2/g Eqn. 11.4

HMF : $C_6H_6O_3 + 6O_2 \rightarrow 6CO_2 + 3H_2O$ - COD = 1.5238 g O_2/g Eqn. 11.5

Succinic acid : $C_4H_6O_4 + 3.5O_2 \rightarrow 4CO_2 + 3H_2O$ - COD = 0.9491 g O_2/g Eqn. 11.6

Isoamylalcohol : $C_5H_{12}O + 7.5O_2 \rightarrow 5CO_2 + 6H_2O$ - COD = 2.7273 g O_2/g Eqn. 11.7

Lactic acid : $C_3H_6O_3 + 3O_2 \rightarrow 3CO_2 + 3H_2O$ - COD = 1.0667 g O_2/g Eqn. 11.8

Xylitol : $C_5H_{12}O_5 + 5.5O_2 \rightarrow 5CO_2 + 6H_2O$ - COD = 1.1579 g O_2/g Eqn. 11.9

Hexose : $C_6H_{12}O_6 + 6O_2 \rightarrow 6CO_2 + 6H_2O$ - COD = 1.0667 g O_2/g Eqn. 11.10

Pentoses : $C_5H_{10}O_5 + 5O_2 \rightarrow 5CO_2 + 5H_2O$ - COD = 1.0667 g O_2/g Eqn. 11.11

Aromatics : $C_6H_6O + 7O_2 \rightarrow 6CO_2 + 3H_2O$ - COD = 2.383 g O_2/g Eqn. 11.12

Impurities : $CH_{1.1}ON_{0.04}S_{0.001} + 0.816O_2 \rightarrow CO_2 + 0.55H_2O + 0.04NO_2 + 0.001SO_2$ - COD = 0.752 g O_2/g Eqn. 11.13

The flows of the waste water streams together with their COD flows are given in Table 11.1.

Table 11.1 The flow rates of the waste water streams and their chemical oxygen demands

Component	COD (g O_2/g)	Stream <334> (kg/s)	Stream <405> (kg/s)	Stream <429> (kg/s)	Total (kg/s)	COD Flow (kg/s)
Acetic Acid	1.0657	0.000	0.125	0.027	0.152	0.1620
Ethyl Alcohol	2.0870	0.003	0.000	0.000	0.003	0.0063
Furfural	1.6653	0.000	0.370	0.000	0.370	0.6160
Glycearol	1.2174	0.000	0.000	0.047	0.047	0.0570
HMF	1.5238	0.000	0.068	0.000	0.068	0.1032
isoamylalcohol	2.7273	0.000	0.000	0.005	0.005	0.0125
Lactic Acid	1.0667	0.000	0.000	0.004	0.004	0.0047
Succinic Acid	0.9491	0.000	0.000	0.005	0.005	0.0047
Water	0.0000	2.108	23.869	8.535	34.512	0.0000
Total		24.432	8.623	2.111	35.166	0.9663

The total COD flow rate is 0.9663 kg/s with volumetric flow rate of 35.1 l/s. The proposed set up removes 99% of the COD. Therefore, after the treatment the effluent waste water has COD flow of 0.0096 kg/s which corresponds to 0.03 g COD/l. The waste water contamination limits for the Netherlands are given in Table 11.2 and they will be used as reference values. As it is presented in Table 11.2, the maximum allowable COD in waste water is 50 mg/l. When compared with the current design, it is seen that the resultant 30 mg/l COD in waste water for the current design is lower then the acceptable values.

Table 11.2 Effluent limits for wastewater treatment plants
(Kleerebezem R, TU Delft Kluyver Lab, personal communication)

Compound	Limit	Units
N-tot	10	mg N/L
N-NH_4	1	mg N/L
P-tot	1	mg P/L
COD tot	50	mg/L
Solids	20	mg/L

During the aerobic treatment, methane is generated at rate of 0.199 kg/s , and, can be combusted in cogeneration unit to generate energy for steam and electricity production. The net energy production by this operation will be 9.96 MW. The slurry bled from the aerobic digestion is also suggested to be combusted in cogeneration.

The waste water pollution of the plant will be reduced to minimal levels if the proposed systems are considered in the design. As previously mentioned, although the design of the waste water treatment systems are not inside the battery limits the costs are included in economic calculations (for details see Section 13.4.2.3 and Appendix A.11.5.3).

11.2. Waste Gas

The exhaust gas streams leaving the battery limits are Stream <139> from lime kiln and sulphur burners that contains carbon dioxide, sulphur dioxide, oxygen and nitrogen Stream <225> from sugar drier that contains water vapor and air, Stream <327> from the ethanol scrubber that contains carbon dioxide, water and trace amount of ethanol and Stream <436> which is the exhaust gas of the cogeneration and contains, carbon dioxide, nitrous oxides, sulphur dioxide and air.

Streams <225> and <327> do not need to be treated because the former is just moist air and the latter is already below the specifications in terms of organic material content (ethanol scrubber is designed accordingly).

The composition and flow of the remaining air streams are given in Table 11.3.

Table 11.3 the flow rates of waste gases.

	Stream <139> (kg/s)	Stream <439> (kg/s)	Total (kg/s)	Composition (Mass)	Concentration (mg/l)
CO_2	0.542	42.654	43.196	15.34%	1196.02
N_2	1.915	174.831	176.746	62.76%	4,893.76
NO_2	0.000	0.519	0.519	0.18%	14.37
O_2	0.193	15.695	15.888	5.64%	439.9
SO_2	0.004	0.014	0.018	0.01%	0.50
Water	0.135	45.136	45.271	16.07%	1,253.46
	2.788	278.850	281.637	100.00%	

The national air quality standards for Brazil are given in Table 11.4 (IRef-9).

Table 11.4 The maximum allowable limits for the pollutants in exhaust gas (IRef-9)

Pollutant	Sampling time	Primary μg/m³	Secondary μg /m³
Total particles in suspension	24 hours[1]	240	150
	MGA[2]	80	80
Inhaleable particles	24 hours[1]	150	150
	MAA[3]	50	50
Smoke	24 hours[1]	150	100
	MAA[3]	60	40
Sulfur dioxide	24 hours[1]	365	100
	MAA[3]	80	40
Nitrogen dioxide	1 hour[1]	320	190
	MAA[3]	100	100
Carbon monoxide	1 hour[1]	40.000 35ppm	40.000 35ppm
	8 hours[1]	10.000 9ppm	10.000 9ppm
Ozone	1 hour[1]	160	160
Notes	1 – Should not be exceeded more than once a year 2 - Annual geometric average 3 - Annual arithmetic average		

The nitrous oxide and sulphur dioxide flows in the current design are higher then the specified limits, thus, the pollutants in waste gas should be first absorbed using absorption set up then the effluent should be treated in waste treatment plant. The design of the treatment plant is outside the battery limits but the treatment costs are included in the cost analysis.

The suspending particles in the exhaust gas should be captured in bag houses by beg filters.

11.3. Waste Solids

The major waste solid streams are filter cake from S-103 (Stream <123>), yeast bleed from fermentors (Stream <328>) and filter cake from S-403 (Stream <416>). All these streams are sent to cogeneration to combust the organic materials present. The resultant solid waste from the cogeneration is Stream <433> which contains gypsum, other calcium salts, dirt and ash and will be sent to land fill.

11.4. Conclusion

If the proposed treatment methods are applied to the waste streams the effect of the plant to environment will be minimized and kept with in the specification. The design of the equipment can be considered in a separate study to have more clear view about the economical aspects.

12. Economy: 2005 Plant

12.1. Purchase Costs for the Process Equipments (EPC)

The details of the price determinations for the individual equipments can be found in Appendix A.10.1 to A.10.6. Since the equipment costs are based on previous years' costs, the costs are adjusted to 2005 costs using the inflation data from US. The yearly inflation rate for price adjustments is taken as 2.47% on average for the last 15 years. The details of the inflation calculations can be found in Appendix A.10.7. The purchase costs of the equipments (EPC) can be found in Tables A.10.11 to A.10.17 (Appendix A.10.9). The total equipment cost with old prices is obtained as $60,657,433, which corresponds to $74,327,355 in 2005 prices.

12.2. Installed Equipment Costs (IEC)

The details of the calculations for the installed equipment costs can be found in Appendix A.10.8. The equipment installation factors (EIF) and the installed equipment costs (EPC) can be found in Tables A.10.11 to A.10.17 (Appendix A.10.9). The total installed equipment cost is obtained as $ $114,707,807.

12.3. Fixed Capital Investment (FCI)

Fixed capital cost is estimated based on equipment purchase cost and installed equipment cost using the factors listed in Table 12.1. The procedure followed in selection of the cost factors can be found in Appendix A.10.10.

Table 12.1 Itemized fixed capital cost estimations

Major equipment total purchase cost (MEPC)		**Factor**	**Investment**
Equipment erection (*f_1*)	(IEC)	0.00	$114,707,807
Piping (*f_2*)	$(EPC\times)$ *	0.15	$11,149,103
Instrumentation (*f_3*)	$(EPC\times)$	0.15	$11,149,103
Electrical (*f_4*)	$(EPC\times)$	0.07	$5,202,915
Building (*f_5*)	$(EPC\times)$	0.25	$18,581,839
Service facilities (*f_6*)	$(EPC\times)$	0.20	$14,865,471
Yard improvement (*f_7*)	$(EPC\times)$	0.05	$3,716,368
Land (*f_8*)	$(EPC\times)$	0.05	$3,716,368
Physical Plant Cost (PPC)	$PPC = IEC + EPC \times (f_1 + f_2 + f_3 \ldots f_8)$		**$183,088,973.54**
Design And Engineering (*f_9*)	$(PPC\times)$	0.08	$14,647,118
Contractors Fee (*f_{10}*)	$(PPC\times)$	0.08	$14,647,118
Contingency (*f_{11}*)	$(PPC\times)$	0.05	$9,154,449
Fixed Capital Investment	$FCI = PPC \times (1 + f_9 + f_{10} + f_{11})$		$221,537,658

*EPC: Purchased equipment cost, IEC: installed equipment cost and FCI: Fixed capital investment

The working capital and start up costs, which are also included in the capital investment for the first year of operation, are calculated, based on the assumption that they are 15% of the fixed capital investment.

12.3.1. The Brazil Location Factor

The calculated investments are based on the US prices, and, are needed to be converted to investment in Brazil. Dr. ir. Peter Nosssin (DSM, Geleen, The Netherlands) kindly provided the location factor for investments in Brazil. They have calculated a location factor of 0.4 to correct for Brazil. This calculation is based on a SRI-method (that takes social, environmental and ethical criteria into account when investing) to estimate the location factor and the method is based on exchange rates and differences in purchasing power of US and Brazil. The resultant investment figures are presented in Table 12.2.

Table 12.2 Total capital investment

Cost Type	US investment	Brazil investment
Equipment purchase cost (2005)	$74,327,355	$29,730,942
Installed equipment cost	$114,707,807	$45,883,123
Fixed capital investment	$221,537,658	$88,615,063
Working Capital	$33,230,649	$13,292,259
Total Capital Investment (TCI)	$254,768,307	$101,907,323

12.4. Operational Costs

12.4.1. Fixed Operational Costs

The factors affecting the annual fixed operational cost are given in Table 12.3.

12.4.1.1. Labor cost

The operator requirement of the plant is calculated as 19 operators per shift. The number of shifts is 5, thus, the total operator requirement of the plant is obtained as 95 operators. The cost of one worker including the benefits is $2,796/year (6 months season). The details of the labor calculations can be found in Appendix A.10.10.1. The contribution of labor cost to fixed operational costs is given in Table 12.3.

Table 12.3 Annual fixed operational costs

Item	Contribution	Cost (per year)	Notes
Maintenance	7% of FCI	$5,126,491	(1)
Operating Labor	From manning estimates	$265,620	
Laboratory costs	15% of FCI	$38,878	(2)
Supervision	40% of labor cost	$103,676	(3)
Plant overheads	70% of labor cost	$181,432	(4)
Capital charges	0% of FCI	$0	(5)
Insurance	1% of FCI	$732,356	(6)
Local taxes	2% of FCI	$1,464,712	(7)
Royalties	0% of FCI	$0	(8)
Total		$7,188,848	

Notes:
(1) 5-9% of FCI for average processes with normal operating conditions (Peters and Timmerhaus, 1991)
(2) 10-20% of labor cost (Peters and Timmerhaus, 1991)
(3) 20 % of operating labor cost (Sinnot, 1999), taken as 40% because the labor cost is calculated for 6 months but plant needs whole year supervision.
(4) 50-70% of operating labor.
(5) depreciation is taken as a separate factor in cash flow analysis
(6) approximately 1 % of FCI (Peters and Timmerhaus, 1991)
(7) 1-2% of FCI (Peters and Timmerhaus, 1991)
(8) no patent costs
FCI: fixed capital investment

12.4.2. Variable Operational Costs

The variable operating cost items are given in Table 12.6 and the individual elements are described in following subsections.

12.4.2.1. Raw Material Costs

The yearly requirements of the raw materials and their costs are given in Table 12.4. Water is not counted in raw material costs because the evaporators (E-201-206) are generating enough condensates to meet the demand.

Table 12.4 Raw material costs

Raw material	Requirement (ton/year)	Unit Cost ($/ton)	Cost ($/year)	Notes
Sugar Cane	5,000,000	12.8	64,000,000	(1)
Limestone	4,557	6.78	30,896	(2)
Sulphur	1,625	30.00	48,750	(3)
Sulphuric acid	2,135	45.00	96,075	(3)
Polyacrylamide	11	3,110.6	34,216	(2)
Nitrogen	181	82.6	14,951	(4)
Total			64,244,888	
Notes	(1) Luis Cortes, UNICAMP, 2004, price is adjusted to 2005. (2) Aden *et al* (2002), corrected to 2005 (3) Chemical Market Reporter (IRef-18) (4) Sinnot (1999), corrected to 2005			

12.4.2.2. Utility Costs

The yearly requirements of the raw materials and their costs are given in Table 12.5. The plant is self sufficient in terms of electricity and steam. Therefore, they are not included in cost calculations.

Table 12.5 Utility costs

Utility	Requirement (unit/year)	Unit Cost ($/unit)	Cost ($/year)	Notes
Cooling water (ton)	19,333,460	0.01	193,335	(1)
Chilled water (ton)	3,519,533	0.08	281,563	(1)
Refrigeration (MJ)	16,642,195	0.005	83,211	(1)
Total			558,108	
Notes	(1) The prices are US prices			

12.4.2.3. Waste Treatment Costs

The waste treatment cost is taken as the limestone required to supply the lime to neutralize the waste acid solution. The lime stone requirement is 2,179 ton/ year which corresponds to $14,770/ year. The second factor in waste treatment is the land fill cost for the gypsum generated in the process which is 2,964 ton/year. The gypsum disposal cost is reported as $23/ton (Aden *et al*, 2002, price is corrected to 2005). The yearly gypsum disposal cost is obtained as $68,172. Therefore, the total waste treatment cost is obtained as $82,942.

Table 12.6 The variable costs

Item	Interval	Cost ($/year)
Raw materials	See Table 12.4	64,244,888
Operating material	15% of maintenance cost	768,974
Utilities	See Table 12.5	558,108
Waste treatment		82,942
Shipping& packaging	Negligible	0

12.4.3. General Operating Expenses

General operating expenses are composed of plant overheads, research and development costs, sales expense and reserves. And, their effect on operating costs is estimated as 20-30% of the direct production cost at the site. There aren't significant research and development costs for the current design. Also, the products are major goods that don't require intensive sales activity. Therefore, the contribution of this element to yearly cost is taken as 5% of the direct production costs (Fixed + variable). The yearly operational costs of the plant are given in Table 12.7.

Table 12.7 The annual costs

Item	Amount ($)
Fixed cost	7,188,848
Variable cost	65,634,912
General operating cost	3,641,188
Total operating cost	76,464,948

12.5. Revenue

The yearly income of the plant is listed in Table 12.8. The unused electricity is sold to grid. The electricity price is taken as 0.035 $/kWh.

Table 12.8The annual income

Product	Unit	Capacity (unit/year)	Price ($/Unit)	Income (€)	Notes
Ethanol	ton	165,956	$410	$68,022,008	(1)
Sugar	ton	370,401	$181	$66,798,087	(1)
Electricity	MJ		0.01	$10,642,578	(2)
			Total	$145,462,674	
Notes	(1)	Luiz Cortez, Presentation, UNICAMP, 2004, prices are corrected to 2005			
	(2)	IRef-19 price is corrected to 2005			

12.6. Cash Flow Analysis

The construction period of the plant is assumed to be 3 years. It is assumed that the first year expense is the engineering, construction and contingency costs (Table 12.1). In the second year, half of the remaining fixed capital investment is paid and the payment is finished in third year. The working capital and start-up costs are discounted from the first operating year (fourth year of investment) income. It is assumed that the plant is operated at 60 % capacity in the first operating year, at 80% in the second year and at 100% in the rest of the plant life. The plant life is assumed to be 10 year and depreciated 10% every operation year. The tax is taken as 34% for Brazil (IRef-15). The discounted cash flow is first calculated for an interest rate of 4% (1% higher then the current average US interest rate) and the net present value is obtained as $171 million. The details of the calculations are given in Appendix 10.12.

Then, the discounted cash flow analysis is carried out at different interest rates to find the discounted cash flow rate of return (DCFROR is the interest rate that makes the net present value(NPV) of the investment $0) for the investment. In Figure 12.1, the DCF at different interest rates are presented. It is shown that at an interest rate little less than 30% the end of project cash flow is 0. The interest rate which gives zero net present value of the investment is calculated as 26.46%. This value is an acceptable value since the usual DCFROR is reported as 20-30%. To calculate the DCFROR the Goal Seek add-in of MS Excel is used.

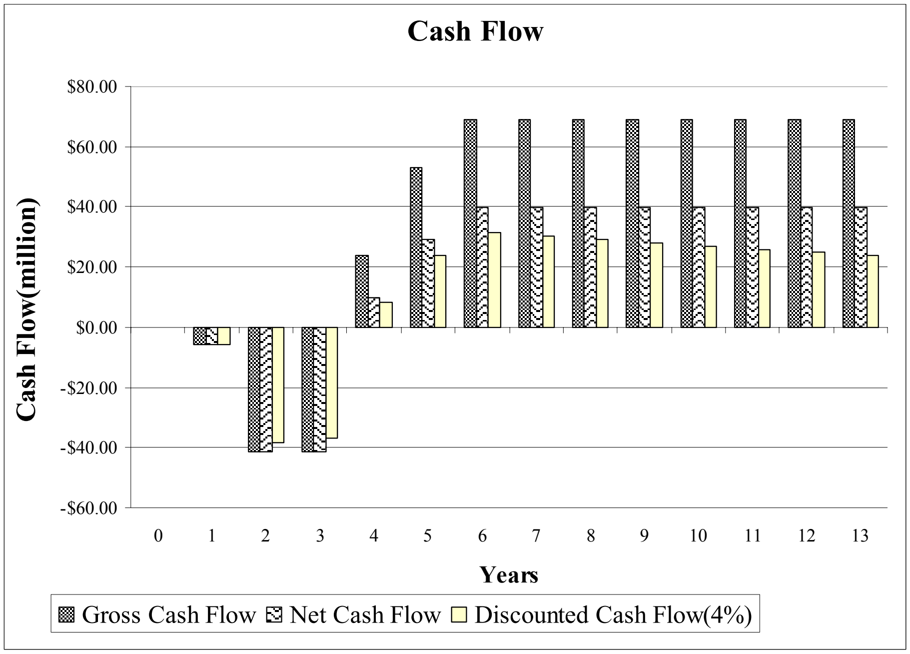


Figure 12.1Annual cash flow

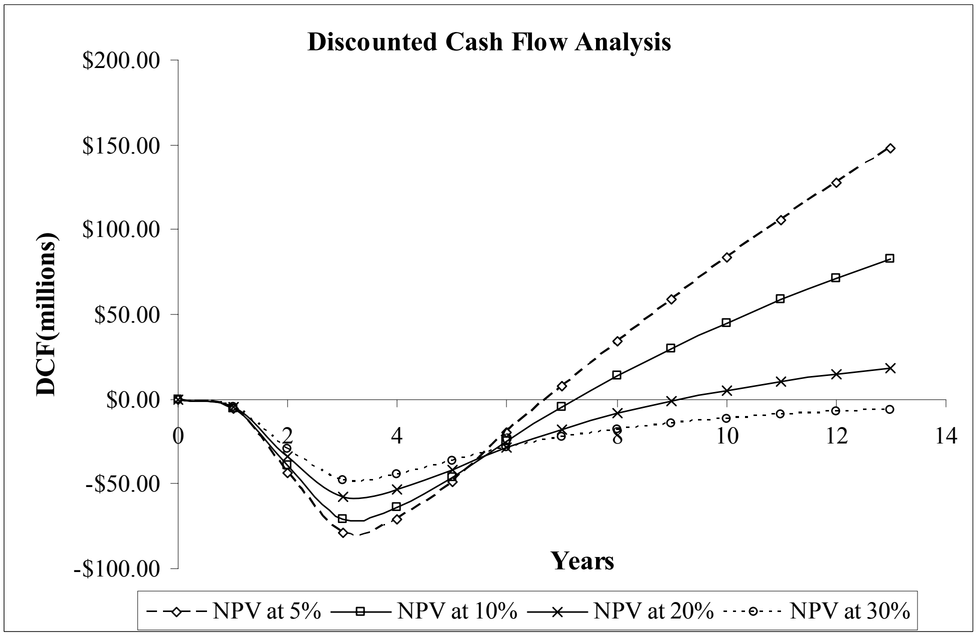


Figure 12.2 Discounted cash flow rate of return analysis

12.7. Payout Period (POP)

The payout time is calculated as 1.9 years using following equation

$$POP = \frac{Depreciable\ Fixed\ Capital\ Investment}{Avarage\ yearly\ profit + Depreciation} \qquad \text{Eqn 12.1}$$

Depreciable FCI is $88,615,064, yearly average profit is (net cash flow) $38,886,610 and the depreciation is 10% of fixed capital investment.

12.8. Rate of Return (ROR)

Rate of return is the ratio of annual profit to investment and calculated as follows,

$$ROR = \frac{CNCF}{Life\,of\;\; project \times Original\,investment} \times 100 \qquad \text{Eqn 12.2}$$

Using the equation the ROR is obtained as 19.7%.

12.9. Sensitivity Analysis

The actual cash flow achieved in any year can be affected by any change in raw material costs and other operating costs and will be very dependent on sales volume and price. Sensitivity analysis is the methodology to see the effects of uncertainties on the viability and feasibility of the project. For this purpose, some elements of the process are selected as the variables and manipulated in the range ±20% of the original value, to observe the effects on the profitability measures like NPV, POP, DCFROR and ROR. The manipulated variables to be analyzed is selected as,

- Location Factor (0.4-0.8)
- Labor Cost (±20%)
- Fixed Costs (±20%)
- Miscellaneous Costs (±20%)
- Sugar Cane Cost (±20%)
- Utility Cost (±20%)
- Ethanol Price (±20%)
- Sugar Price (±20%)
- Electricity Price (±20%)

It should be noted that during the analysis all the other variables kept constant at the values of initial analysis (Tables 12.1-12.8). During the sensitivity analysis, the interest rate for the discounted cash flow analysis is taken as 4% (1% higher than the current US inflation, 3.17% August 2005) (IRef-6).

12.9.1. Sensitivity on Location Factor

The sensitivity of the plant on location factor is given in Table 12.9 and Figures 12.3&5. The sensitivity analysis on location factor can be also interpreted as the sensitivity on the equipment purchase costs.

Table 12.9The sensitivity analysis on location factor

	NPV (4%)	POP (years)	DCFROR	ROR
0.4	$165,957,000	1.86	26.46%	19.65%
0.5	$124,240,000	2.35	18.88%	12.60%
0.6	$ 82,349,000	2.85	12.96%	7.89%
0.7	$ 38,483,000	3.36	7.86%	4.43%
0.8	$ (5,381.97)	3.88	3.49%	1.83%

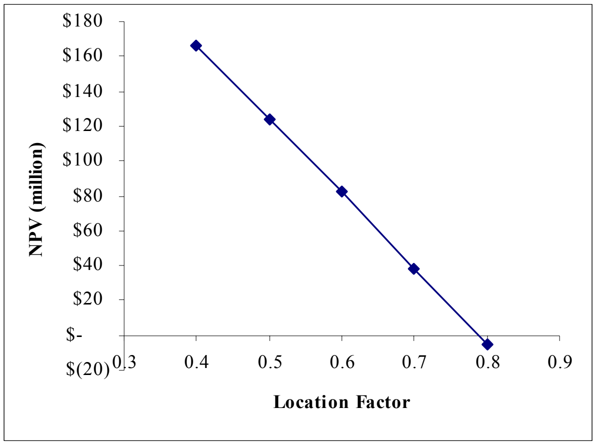


Figure 12.3 Effect of location factor on NPV

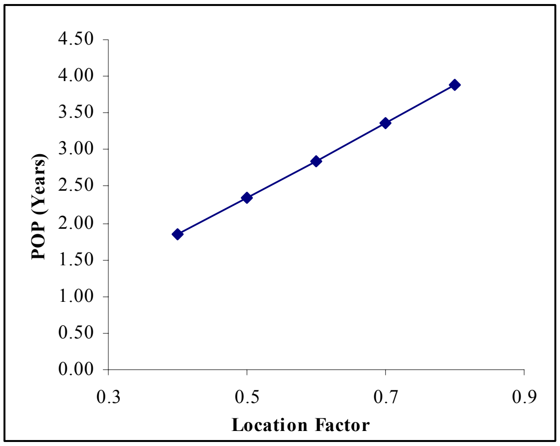


Figure 12.4 Effect of location factor on POP

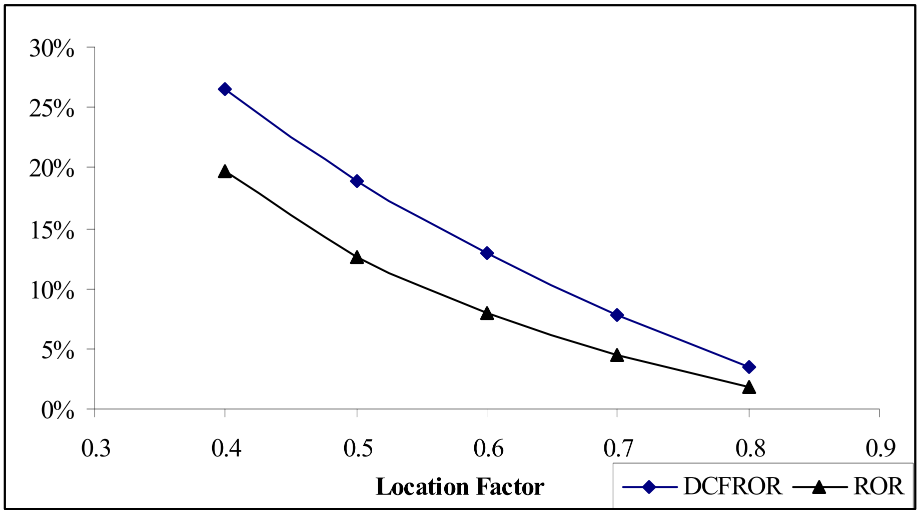


Figure 12.5 Effect of location factor DCFROR & ROR

12.9.2. Effect of Variables on NPV

The sensitivity of the NPV to changes in, labor cost, fixed costs, miscellaneous costs, sugar cane cost, utility cost, ethanol price, sugar price and electricity price are given in Table 12.10 and Figure 12.6.

Table 12.10 The sensitivity of NPV

	NPV (million)							
	Labor	Fixed Costs	Miscellaneous	Sugar Cane	Utilities	Ethanol	Sugar	Electricity
-20%	$166.5	$173.1	$169.4	$225.6	$166.4	$105.5	$106.5	$158.1
-10%	$166.2	$169.5	$167.6	$195.7	$166.2	$135.7	$136.2	$162.0
0%	$165.9	$165.9	$165.9	$165.9	$165.9	$165.9	$165.9	$165.9
10%	$165.6	$162.3	$164.2	$136.1	$165.6	$196.1	$195.6	$169.8
20%	$165.3	$158.7	$162.5	$106.2	$165.4	$226.3	$225.3	$173.7

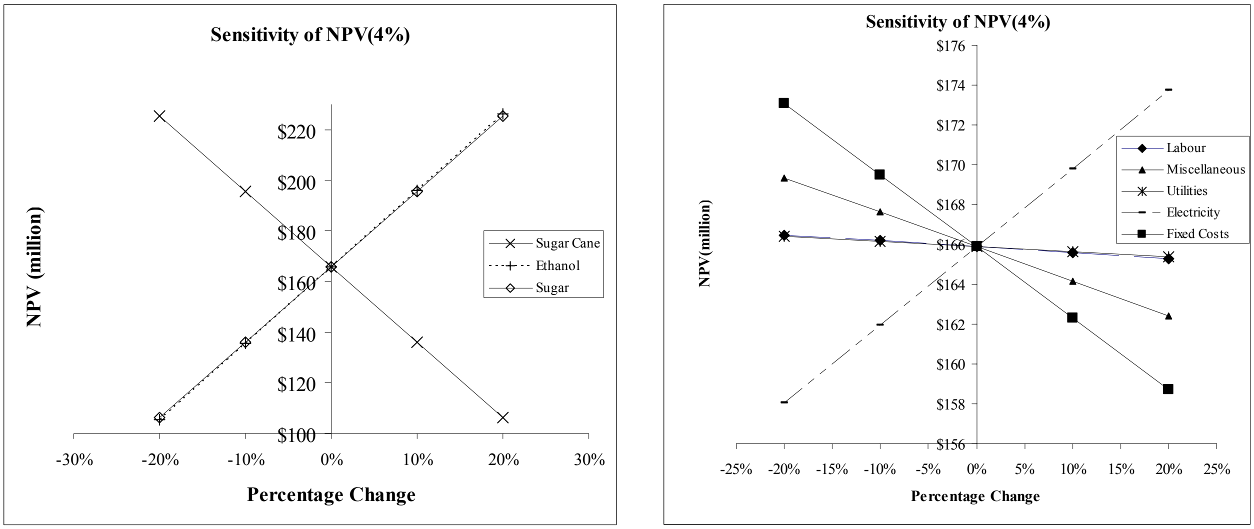


Figure 12.6 Sensitivity of NPV on specified variables

12.9.3. Effect of Variables on POP

The sensitivity of the POP to changes in, labor cost, fixed costs, miscellaneous costs, sugar cane cost, utility cost, ethanol price, sugar price and electricity price are given in Table 12.11 and Figure 12.7.

Table 12.11 The sensitivity of POP

	POP (years)							
	Labor	Fixed Costs	Miscellaneous	Sugar Cane	Utilities	Ethanol	Sugar	Electricity
-20%	1.85	1.82	1.84	1.57	1.85	2.29	2.28	1.90
-10%	1.85	1.84	1.85	1.70	1.85	2.05	2.05	1.88
0%	1.86	1.86	1.86	1.86	1.86	1.86	1.86	1.86
10%	1.86	1.88	1.87	2.05	1.86	1.70	1.70	1.83
20%	1.86	1.90	1.88	2.28	1.86	1.56	1.57	1.81

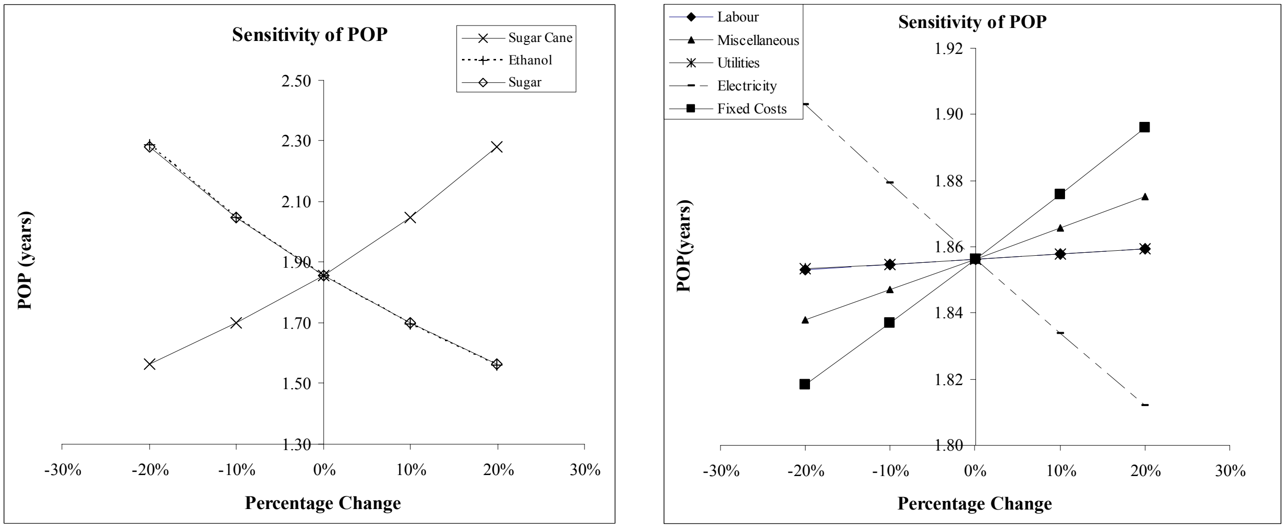


Figure 12.7 Sensitivity of POP on specified variables

12.9.4. Effect of Variables on DCFROR

The sensitivity of the DCFROR to changes in, labor cost, fixed costs, miscellaneous costs, sugar cane cost, utility cost, ethanol price, sugar price and electricity price are given in Table 12.12 and Figure 12.8.

Table 12.12 The sensitivity of DCFROR

	DCFROR							
	Labor	Fixed Costs	Miscellaneous	Sugar Cane	Utilities	Ethanol	Sugar	Electricity
-20%	26.52%	27.26%	26.84%	32.48%	26.51%	19.61%	19.72%	25.61%
-10%	26.49%	26.86%	26.65%	29.54%	26.48%	23.15%	23.20%	26.04%
0%	26.45%	26.45%	26.45%	26.45%	26.45%	26.45%	26.45%	26.45%
10%	26.42%	26.05%	26.26%	23.18%	26.43%	29.56%	29.52%	26.87%
20%	26.39%	25.64%	26.06%	19.68%	26.40%	32.52%	32.43%	27.28%

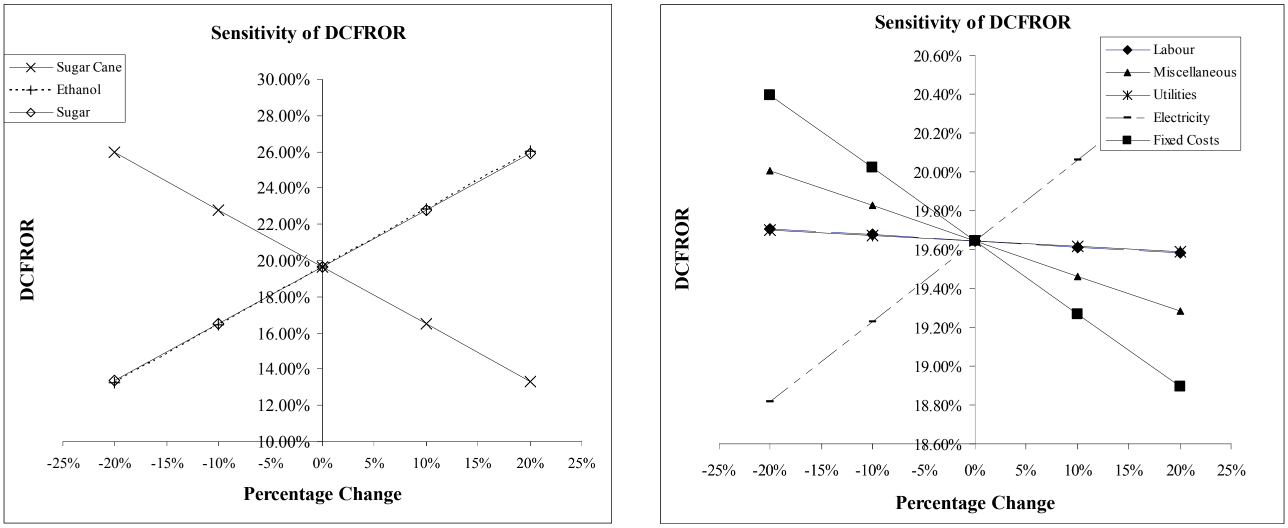


Figure 12.8 Sensitivity of DCFROR on specified variables

12.9.5. Effect of Variables on ROR

The sensitivity of the ROR to changes in, labor cost, fixed costs, miscellaneous costs, sugar cane cost, utility cost, ethanol price, sugar price and electricity price are given in Table 12.13 and Figure 12.9.

Table 12.13 The sensitivity of ROR

	ROR							
	Labor	Fixed Costs	Miscellaneous	Sugar Cane	Utilities	Ethanol	Sugar	Electricity
-20%	1.85	1.82	1.84	1.57	1.85	2.29	2.28	1.90
-10%	1.85	1.84	1.85	1.70	1.85	2.05	2.05	1.88
0%	1.86	1.86	1.86	1.86	1.86	1.86	1.86	1.86
10%	1.86	1.88	1.87	2.05	1.86	1.70	1.70	1.83
20%	1.86	1.90	1.88	2.28	1.86	1.56	1.57	1.81

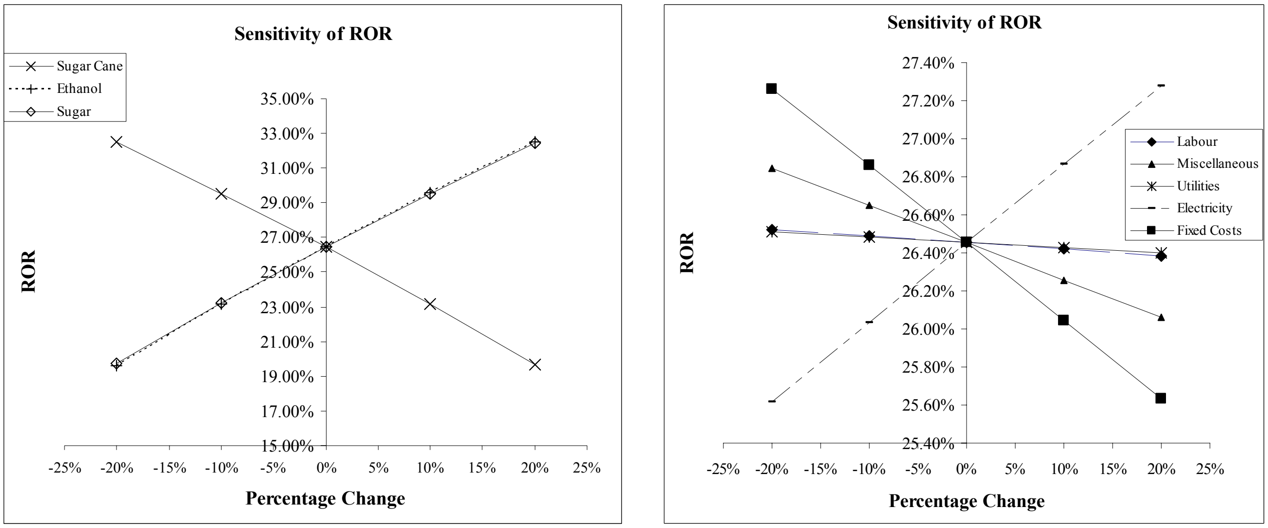


Figure 12.9 Sensitivity of DCFROR on specified variables

12.9.6. The Production Costs of Ethanol and Sugar

The production costs are determined based on gross income, net income (including tax and depreciation), net income + depreciation (cost excluding depreciation) and net income + tax (cost excluding tax but including depreciation) (See Table 12.14). The cost of ethanol is calculated by taking the 48% of the total cost and vice versa for the sugar cost.

Table 12.14 The production costs ($/ton)

	Gross	Net	Net+depr.	Net+tax
Ethanol	221.2	306.5	280.8	246.8
Sugar	107.4	148.8	136.3	119.8

13. Economy: 2015 Plant

13.1. Purchase Costs for the Process Equipments (EPC)

The details of the price determinations for the individual equipments can be found in Appendix A.11.1 to A.11.6. Since the equipment costs are based on previous years' costs, the costs are adjusted to 2005 costs using the inflation data from US. The yearly inflation rate for price adjustments is taken as 2.47% on average for the last 15 years. The details of the inflation calculations can be found in Appendix A.10.7. The purchase costs of the equipments (EPC) can be found in Tables A.11.4 to A.11.9 (Appendix A.11.7). The total equipment cost with old prices is obtained as $79,871,474, which corresponds to $94,502,977 in 2005 prices.

13.2. Installed Equipment Costs (IEC)

The details of the calculations for the installed equipment costs can be found in Appendix A.10.8. The equipment installation factors (EIF) and the installed equipment costs (EPC) can be found in Tables A.11.4 to A.11.9 (Appendix A.11.7). The total installed equipment cost is obtained as $ $134,791,741.

13.3. Fixed Capital Investment (FCI)

Fixed capital cost is estimated based on equipment purchase cost and installed equipment cost using the factors listed in Table 13.1. The procedure followed in selection of the cost factors can be found in Appendix A.10.10. The elements of fixed capital investment and total capital investment can be found in Table 13.2.

Table 13.1 Itemized fixed capital cost estimations

Major equipment total purchase cost (MEPC)		**Factor**	**Investment**
Equipment erection (*f_1*)	$(IEC)^*$	0.00	$134,791,741
Piping (*f_2*)	$(EPC\times)^*$	0.15	$14,175,447
Instrumentation (*f_3*)	$(EPC\times)$	0.15	$14,175,447
Electrical (*f_4*)	$(EPC\times)$	0.07	$6,615,208
Building (*f_5*)	$(EPC\times)$	0.25	$23,625,744
Service facilities (*f_6*)	$(EPC\times)$	0.20	$18,900,596
Yard improvement (*f_7*)	$(EPC\times)$	0.05	$4,725,149
Land (*f_8*)	$(EPC\times)$	0.05	$4,725,149
Physical Plant Cost (PPC)	$PPC = IEC + EPC \times (f_1 + f_2 + f_3 .... f_8)$		**$221,734,480**
Design And Engineering (*f_9*)	$(PPC\times)$	0.10	$22,173,448
Contractors Fee (*f_{10}*)	$(PPC\times)$	0.10	$22,173,448
Contingency (*f_{11}*)	$(PPC\times)$	0.15	$22,173,448
Fixed Capital Investment	$FCI = PPC \times (1 + f_9 + f_{10} + f_{11})$		$288,254,824

*EPC: Purchased equipment cost, IEC: installed equipment cost and FCI: Fixed capital investment

The working capital and start up costs, which are also included in the capital investment for the first year of operation, are calculated, based on the assumption that it is 15% of the fixed capital investment.

13.3.1. The Brazil Location Factor

The calculated investments are based on the US prices, and, are correct for the US, but, are not correct for Brazil, and, are to be converted to investment in Brazil. Dr. ir. Peter Nosssin (DSM, Geleen, The Netherlands) kindly provided the location factor for investments in Brazil. They have calculated a location factor of 0.4 to correct for Brazil. This calculation is based on a SRI-method (that takes social, environmental and ethical criteria into account when investing) to estimate the location factor and the method is based on exchange rates and differences in purchasing power of US and Brazil. The resultant investment figures are presented in Table 13.2.

Table 13.2 Total capital investment

Cost Type	US investment	Brazil investment
Equipment purchase cost (2005)	$94,502,977	$37,801,191
Installed equipment cost	$134,791,741	$53,916,696
Fixed capital investment	$288,254,824	$115,301,930
Working Capital	$43,238,224	$17,295,289
Total Capital Investment (TCI)	$331,493,048	$132,597,219

13.4. Operational Costs

13.4.1. Fixed Operational Costs

The factors affecting the annual fixed operational cost are given in Table 13.3.

13.4.1.1. Labor cost

The labor requirement of the plant is calculated as 23 per shift. The number of shifts is 5, thus, the total labor requirement of the plant is obtained as 115 operators. The cost of one worker including the benefits is $2,796/year (6 months season). The details of the labor calculations can be found in Appendix A.10.10.1. The contribution of labor cost to fixed operational costs is given in Table 13.3.

Table 13.3 Annual fixed operational costs

Item	Contribution	Cost (per year)	Notes
Maintenance	12% of FCI	$10,643,255	(1)
Operatinglabor	From manning estimates	$321,400	
Laboratory costs	15% of FCI	$48,210	(2)
Supervision	40% of labor cost	$128,560	(3)
Plant overheads	70% of labor cost	$224,980	(4)
Capital charges	0% of FCI	$0	(5)
Insurance	1% of FCI	$1,153,019	(6)
Local taxes	2% of FCI	$2,306,039	(7)
Royalties	0% of FCI	$0	(8)
Total		$14,825,463	

Notes:

(1) 5-9% of FCI for average processes with normal operating conditions (Peters and Timmerhaus, 1991), the maintenance costs of the current design is high due to the harsh operation conditions, i.e. presence of acids, high pressures and temperatures and fibers.
(2) 10-20% of labor cost (Peters and Timmerhaus, 1991)
(3) 20 % of operating labor cost (Sinnot, 1999), taken as 40% because the labor cost is calculated for 6 months but plant needs whole year supervision.
(4) 50-70% of operating labor.
(5) depreciation is taken as a separate factor in cash flow analysis
(6) approximately 1 % of FCI (Peters and Timmerhaus, 1991)
(7) 1-2% of FCI (Peters and Timmerhaus, 1991)
(8) no patent costs
FCI: fixed capital investment

13.4.2. Variable Operational Costs

The variable operating cost items are given in Table 13.6 and the individual elements are described in following subsections.

13.4.2.1. Raw Material Costs

The yearly requirements of the raw materials and their costs are given in Table 13.4. Water from the evaporators, boiling pans and waste treatment is not enough to meet the demand. There is a gap of 707,000 ton/ year and this is included in the raw material cost.

Table 13.4 Raw material costs

Raw material	Requirement (ton/year)	Unit Cost ($/ton)	Cost ($/year)	Notes
Sugar cane	5,000,000	12.80	64,000,000	(1)
Lime stone	13,908	6.78	94,293	(2)
Sulphur	1,625	30.00	48,754	(3)
Sulphuric acid	14,096	45.00	634,320	(3)
Cellulase	5,499	2,519	13,852,560	(5)
Polyacrylamide	11	3,111	32,723	(2)
Nitrogen	346	82.60	28,561	(4)
Water	707,000	0.20	141,400	(4)
Natural gas	780	263.80	205,827	(4)
Total			79,038,439	
Notes	(5) Luis Cortes, UNICAMP, 2004, price is adjusted to 2005.			
	(6) Aden *et al* (2002), corrected to 2005			
	(7) Chemical Market Reported (IRef-18)			
	(8) Sinnot (1999), US prices corrected to 2005			
	(9) $ 0.033-0.06/ kg ethanol, Novozymes web page, 14 April 2005, is taken as 0.06/ kg ethanol			

13.4.2.2. Utility Cost

The yearly requirements of the raw materials and their costs are given in Table 12.5. The plant is self sufficient in terms of steam, however, not in electricity and some portion of electricity is to be purchased from the grid.

Table 13.5 Utility costs

Utility	Requirement (unit/year)	Unit Cost ($/unit)	Cost ($/year)	Notes
Cooling Water (ton)	43,471,325	0.01	434,713	(1)
Chilled Water (ton)	5,191,60	0.08	415,328	(1)
Refrigeration (MJ)	29,721,427	0.005	148,607	(1)
Electricity (MJ)	126,847,804	0.01	1,268,478	
Total			2,267,127	
Notes	(1) The prices are US prices			

13.4.2.3. Waste Treatment Costs

The waste treatment cost is taken as the limestone required to supply the lime to neutralize the waste acid solution. The lime stone requirement is 5,216 ton/ year which corresponds to $35,362/year. The second factor in waste treatment is the land fill cost for the gypsum generated in the neutralization and the solids generated in furnaces (ash, gypsum, soil from F-401) which is 102,795 ton/year. The gypsum disposal cost is reported as $23/ton (Aden *et al*, 2002, price is corrected to 2005). The yearly gypsum disposal cost is obtained as $2,364,280. Therefore, the total waste treatment cost is obtained as $2,400,000/year. To account for the waste water treatment cost this value is increased by half and the total waste treatment cost is taken as $3,600,000.

Table 13.6 The variable costs

Item	Interval	Cost ($/year)
Raw Materials	See Table 13.4	79,038,439
Operating Material	15% of maintenance cost	1,596,488
Utilities	See Table 13.5	2,267,127
Waste treatment		3,600,000
Shipping& Packaging	Negligible	0
Total		86,502,054

13.4.3. General Operating Expenses

General operating expenses are composed of plant overheads, research and development costs, sales expense and reserves. And their effect on operating costs is estimated as 20-30% of the direct production cost at the site. There aren't significant research and development costs for the current design. Also the products are major goods that don't require intensive sales activity. Therefore, the contribution of this element to yearly cost is taken as 5% of the direct production costs (Fixed + variable). The yearly operational costs of the plant are given in Table 13.7.

Table 13.7 The annual costs

Item	Amount ($)
Fixed cost	14,825,463
Variable cost	86,502,054
General operating cost	5,066,376
Total operating cost	106,393,893

13.5. Revenue

The yearly income of the plant is listed in Table 13.8.

Table 13.8The annual income

Product	Unit	Capacity (unit/year)	Price ($/Unit)	Income (€)	Notes
Ethanol	ton	396,872	410	$162,717,524	(1)
Sugar	ton	370,405	180	$67,169,306	(1)
			Total	$229,886,830	
Notes	(3)	Luiz Cortez, Presentation, UNICAMP, 2004, prices are corrected to 2005			

13.6. Cash Flow Analysis

The construction period of the plant is assumed to be 3 years. It is assumed that the first year expense is the engineering, construction and contingency costs (Table 13.1). In the second year, half of the remaining fixed capital investment is paid and the payment is finished in third year. The working capital and start-up costs are discounted from the first operating year (fourth year of investment) income. It is assumed that the plant is operated at 60 % capacity in the first operating year, at 80% in the second year and at 100% in the rest of the plant life. The plant life is assumed to be 10 year and depreciated 10% every operation year. The tax is taken as 34% for Brazil (IRef-15). The discounted cash flow is first calculated for an interest rate of 4% (1% higher then the current average interest rate). The details of the calculations are given in Appendix 10.12.

Then, the discounted cash flow analysis is carried out at different interest rates to find the discounted cash flow rate of return (DCFROR) for the investment. In Figure 13.1, the DCF at different interest rates are presented. It is shown that at an interest rate little less than 40% the end of project cash flow is 0. The interest rate which yields zero net present value of the investment is calculated as

37.48%. This value is an acceptable value since the usual DCFROR is reported as 20-30%. To calculate the DCFROR the Goal seek add-in of MS Excel is used.

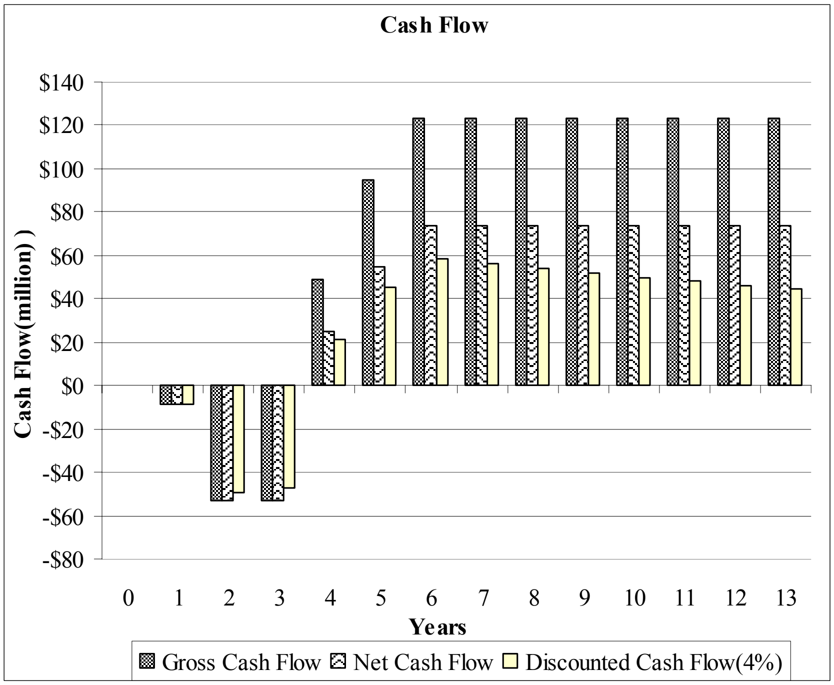


Figure 13.1Annual cash flow

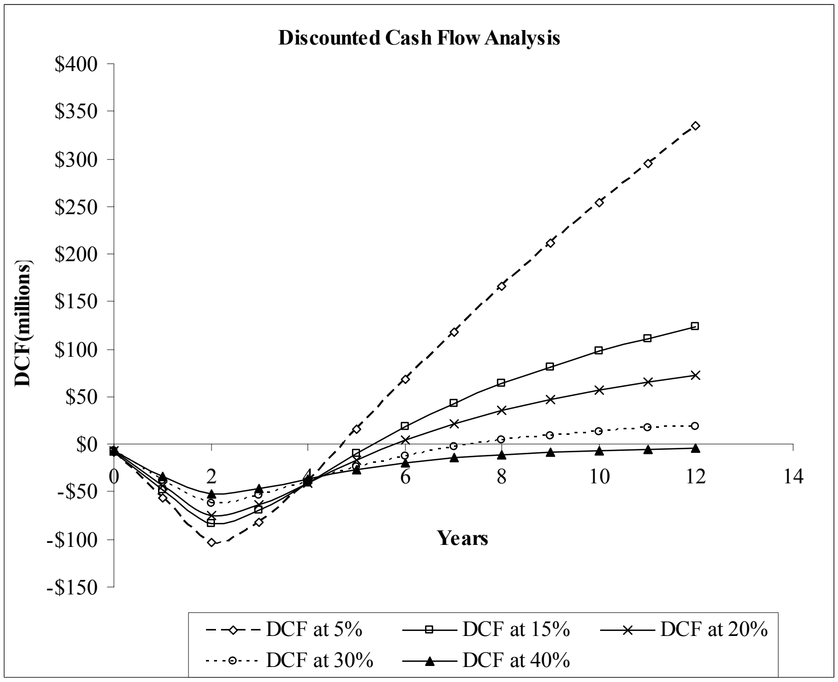


Figure 13.2 Discounted cash flow rate of return analysis

13.7. Payout Period (POP)

The payout time is calculated as 1.4 years using Equation 12.1. Depreciable FCI is $115,301,930, yearly average profit is (net cash flow) $73,895.410 and the depreciation is 10% of fixed capital investment.

13.8. Rate of Return (ROR)

Rate of return is the ratio of annual profit to investment and calculated as 32.2% using Equation 12.2.

13.9. Sensitivity Analysis

The actual cash flow achieved in any year can be affected by any change in raw material costs and other operating costs and will be very dependent on sales volume and price. Sensitivity analysis is the methodology to see the effects of uncertainties on the viability and feasibility of the project. For this purpose, some elements of the process are selected as the variables and manipulated in the range ±20% of the original value, to observe the effects on the profitability measures like NPV, POP, DCFROR and ROR. The manipulated variables to be analyzed is selected as,

- Location Factor (0.4-1)
- Labor Cost (±30%)
- Fixed Costs (±30%)
- Miscellaneous Costs (±30%)
- Sugar Cane Cost (±30%)
- Cellulase Cost (±30%)
- Utility Cost (±30%)
- Ethanol Price (±30%)
- Sugar Price (±30%)
- Electricity Price (±30%)
- Hemicellulose & cellulose conversion

It should be noted that during the analysis all the other variables kept constant at the values of initial analysis (Tables 13.1-13.8). During the sensitivity analysis the interest rate for the discounted cash flow analysis is taken as 4% (1% higher than the current US inflation, 3.17% August 2005) (IRef-6).

13.9.1. Sensitivity on Location Factor

The sensitivity of the plant on location factor is given in Table 13.9 and Figures 13.3&5. The sensitivity analysis on location factor can be also interpreted as the sensitivity on the equipment purchase costs.

Table 13.9The sensitivity analysis on location factor

	NPV (4%)	POP (years)	DCFROR	ROR
0.4	$370,116,960	1.35	37.48%	32.22%
0.5	$308,221,920	1.72	28.66%	22.17%
0.6	$246,326,870	2.11	21.90%	15.47%
0.7	$184,431,830	2.52	16.40%	10.68%
0.8	$120,398,570	2.94	11.58%	7.02%
0.9	$55,257,840	3.38	7.29%	4.14%
1.0	$(9,882.90)	3.84	3.44%	1.83%

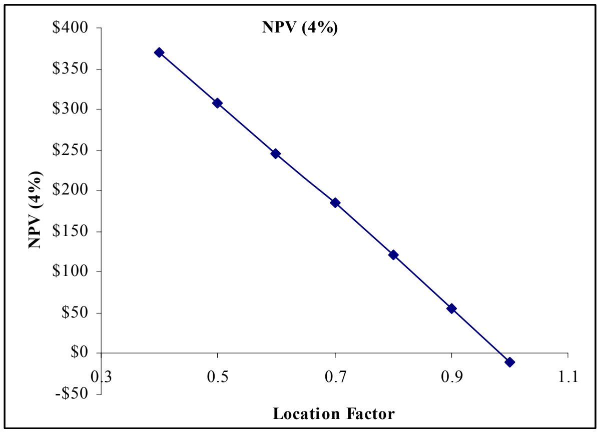


Figure 13.3 Effect of location factor on NPV

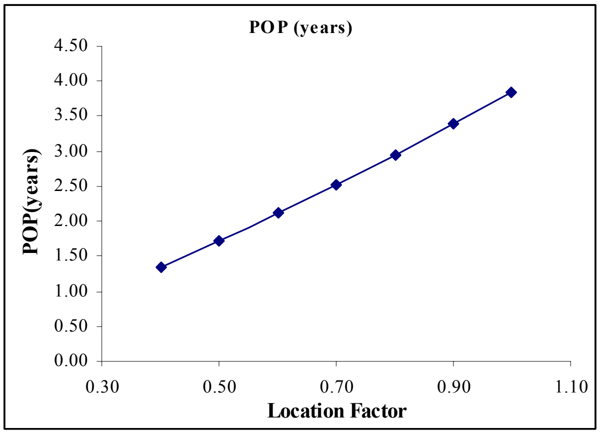


Figure 13.4 Effect of location factor on POP

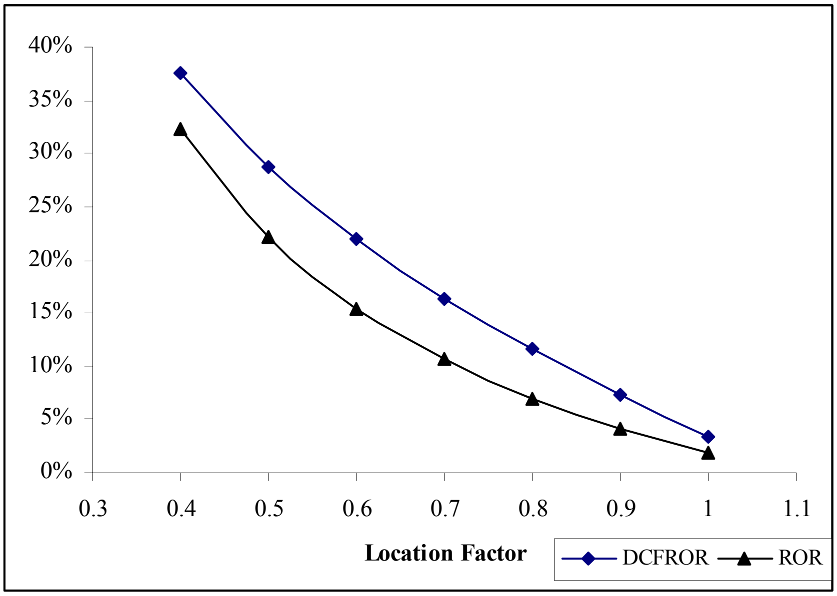


Figure 13.5 Effect of location factor DCFROR & ROR

13.9.2. Effect of Variables on NPV

The sensitivity of the NPV to changes in, labor cost, fixed costs, miscellaneous costs, sugar cane cost, utility cost, ethanol price, sugar price and electricity price are given in Table 13.10 and Figures 13.6&7.

Table 13.10 The sensitivity of NPV

NPV (million)									
	Labor	Fixed Costs	Miscellaneous	Sugar Cane	Cellulase	Utilities	Ethanol	Sugar	Electricity
-30%	$371.20	$392.34	$377.35	$459.64	$389.49	$373.29	$154.12	$280.95	$371.89
-20%	$370.84	$384.93	$374.94	$429.80	$383.04	$372.23	$226.12	$310.67	$371.30
-10%	$370.48	$377.53	$372.53	$399.96	$376.58	$371.17	$298.12	$340.40	$370.71
0%	$370.12	$370.12	$370.12	$370.12	$370.12	$370.12	$370.12	$370.12	$370.12
10%	$369.76	$362.71	$367.71	$340.28	$363.66	$369.06	$442.12	$399.84	$369.53
20%	$369.39	$355.30	$365.29	$310.43	$357.20	$368.00	$514.12	$429.56	$368.93
30%	$369.03	$347.89	$362.88	$280.59	$350.74	$366.95	$586.11	$459.28	$368.34

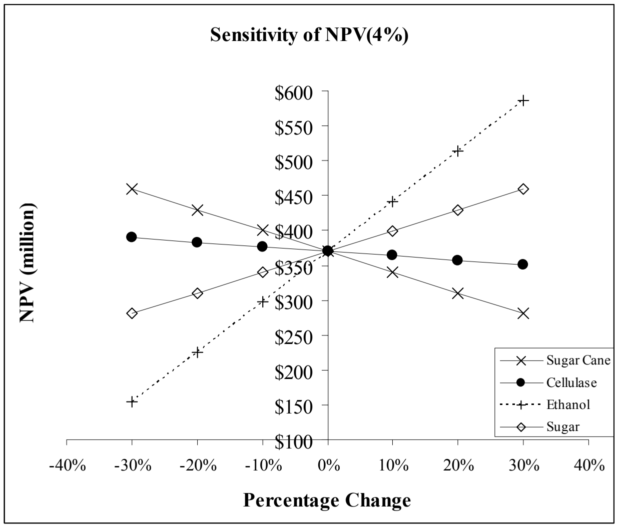


Figure 13.6 Sensitivity of NPV on specified variables

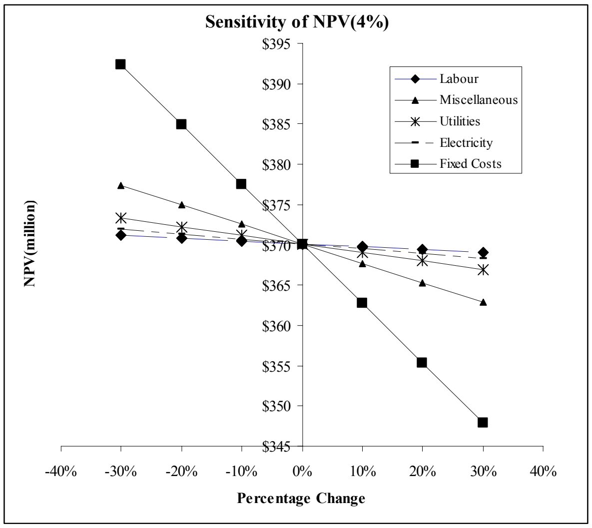


Figure 13.7 Sensitivity of NPV on specified variables

13.9.3. Effect of Variables on POP

The sensitivity of the POP to changes in, labor cost, fixed costs, miscellaneous costs, sugar cane cost, utility cost, ethanol price, sugar price and electricity price are given in Table 13.11 and Figures 13.8&9.

Table 13.11 The sensitivity of POP

	POP (Years)								
	Labor	Fixed Costs	Miscellaneous	Sugar Cane	Cellulase	Utilities	Ethanol	Sugar	Electricity
-30%	1.35	1.30	1.33	1.17	1.31	1.34	2.17	1.60	1.35
-20%	1.35	1.32	1.34	1.22	1.32	1.34	1.80	1.51	1.35
-10%	1.35	1.33	1.34	1.28	1.33	1.35	1.54	1.42	1.35
0%	1.35	1.35	1.35	1.35	1.35	1.35	1.35	1.35	1.35
10%	1.35	1.37	1.36	1.42	1.37	1.35	1.20	1.28	1.35
20%	1.35	1.38	1.36	1.51	1.38	1.35	1.08	1.22	1.35
30%	1.35	1.40	1.37	1.60	1.40	1.36	0.98	1.17	1.35

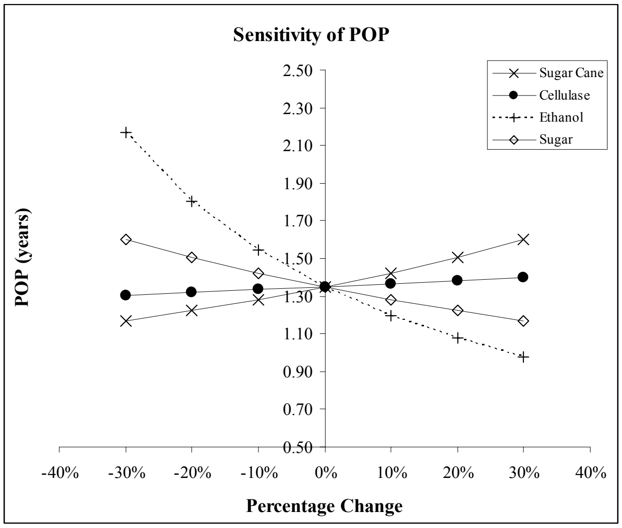


Figure 13.8 Sensitivity of POP on specified variables

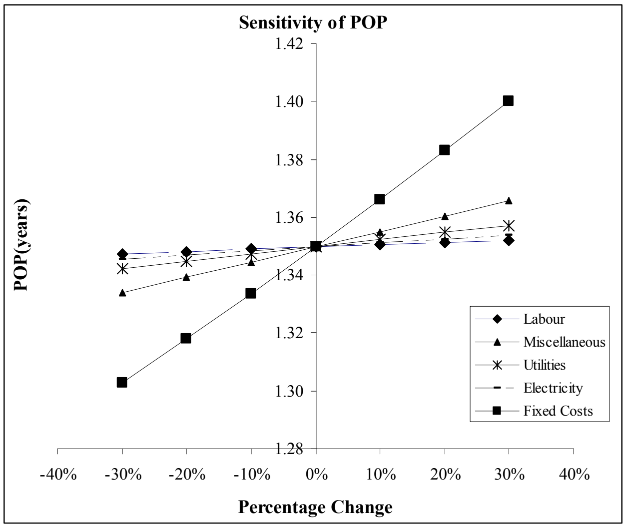


Figure 13.9 Sensitivity of POP on specified variables

13.9.4. Effect of Variables on DCFROR

The sensitivity of the DCFROR to changes in, labor cost, fixed costs, miscellaneous costs, sugar cane cost, utility cost, ethanol price, sugar price and electricity price are given in Table 13.12 and Figure 13.10&11.

Table 13.12 The sensitivity of DCFROR

	DCFROR								
	Labor	Fixed Costs	Miscellaneous	Sugar Cane	Cellulase	Utilities	Ethanol	Sugar	Electricity
-30%	37.56%	39.10%	38.01%	43.26%	38.77%	37.69%	20.96%	31.22%	37.60%
-20%	37.53%	38.56%	37.83%	41.38%	38.34%	37.62%	27.03%	33.38%	37.56%
-10%	37.51%	38.02%	37.66%	39.46%	37.91%	37.55%	32.48%	35.46%	37.52%
0%	37.48%	37.48%	37.48%	37.48%	37.48%	37.48%	37.48%	37.48%	37.48%
10%	37.45%	36.93%	37.30%	35.44%	37.04%	37.41%	42.14%	39.44%	37.44%
20%	37.43%	36.38%	37.12%	33.34%	36.60%	37.34%	46.52%	41.35%	37.40%
30%	37.40%	35.83%	36.94%	31.24%	36.16%	37.27%	50.67%	43.20%	37.36%

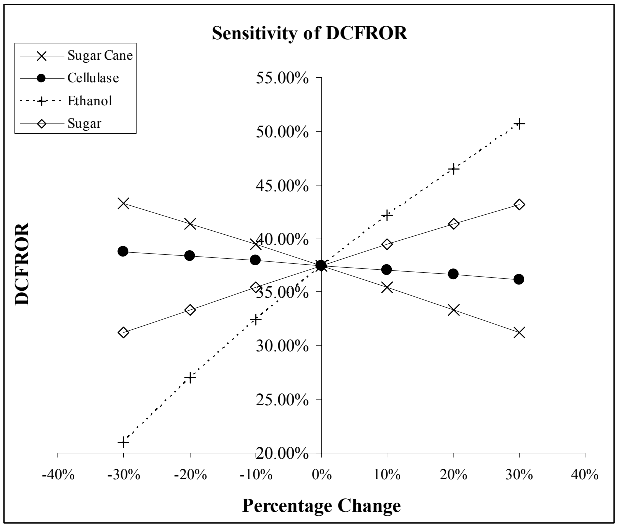


Figure 13.10 Sensitivity of DCFROR on specified variables

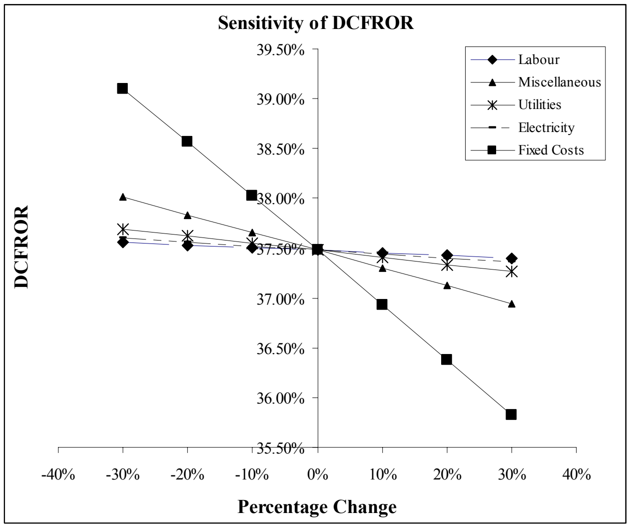


Figure 13.11 Sensitivity of DCFROR on specified variables

13.9.5. Effect of Variables on ROR

The sensitivity of the ROR to changes in, labor cost, fixed costs, miscellaneous costs, sugar cane cost, utility cost, ethanol price, sugar price and electricity price are given in Table 13.13 and Figure 13.12&13.

Table 13.13 The sensitivity of ROR

	ROR								
	Labor	Fixed Costs	Miscellaneous	Sugar Cane	Cellulase	Utilities	Ethanol	Sugar	Electricity
-30%	32.31%	34.01%	32.81%	39.50%	33.80%	32.48%	14.65%	24.97%	32.37%
-20%	32.28%	33.42%	32.61%	37.08%	33.27%	32.40%	20.51%	27.39%	32.32%
-10%	32.25%	32.82%	32.42%	34.65%	32.75%	32.31%	26.37%	29.81%	32.27%
0%	32.22%	32.22%	32.22%	32.22%	32.22%	32.22%	32.22%	32.22%	32.22%
10%	32.19%	31.63%	32.03%	29.80%	31.70%	32.14%	38.08%	34.64%	32.18%
20%	32.17%	31.03%	31.84%	27.37%	31.17%	32.05%	43.94%	37.06%	32.13%
30%	32.14%	30.44%	31.64%	24.95%	30.65%	31.97%	49.79%	39.48%	32.08%

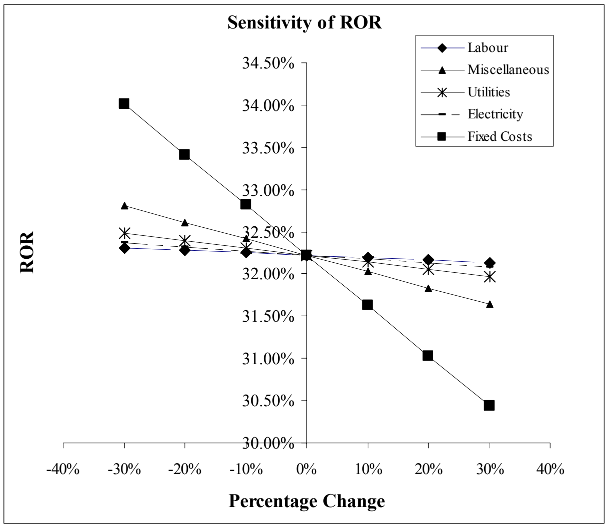


Figure 13.12 Sensitivity of ROR on specified variables

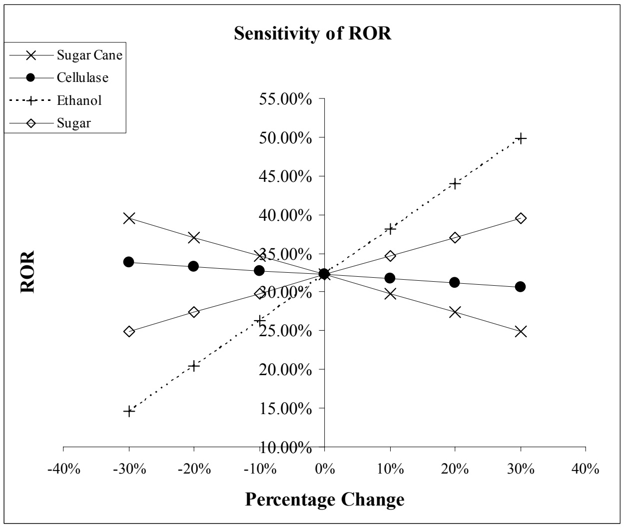


Figure 13.13 Sensitivity of ROR on specified variables

13.9.6. Sensitivity on Hemicellulose and Cellulose Conversions

The sensitivity of NPV, POP, ROR and DCFROR on hemicellulose and cellulose conversions are expressed in terms of ethanol conversions. Table 13.14 show the effect of the changes in conversion percentage of hemicellulose and cellulose on ethanol output of the system. Comparing the ethanol percentage changes given in Table 13.14 with the sensitivity of parameter on ethanol, the sensitivity on hemicellulose and cellulose can be obtained.

Table 13.14 Change in ethanol output with changing cellulose and hemicellulose conversions

		Cellulose Conversion									
		10%	**20%**	**30%**	**40%**	**50%**	**60%**	**70%**	**80%**	**90%**	**100%**
Hemicellulose Conversion	**10%**	51.95%	48.30%	44.65%	40.99%	37.34%	33.68%	30.03%	26.37%	22.72%	19.07%
	20%	49.11%	45.46%	41.80%	38.15%	34.50%	30.84%	27.19%	23.53%	19.88%	16.23%
	30%	46.27%	42.62%	38.96%	35.31%	31.66%	28.00%	24.35%	20.69%	17.04%	13.39%
	40%	43.43%	39.78%	36.12%	32.47%	28.82%	25.16%	21.51%	17.85%	14.20%	10.55%
	50%	40.59%	36.94%	33.28%	29.63%	25.98%	22.32%	18.67%	15.01%	11.36%	7.71%
	60%	37.75%	34.10%	30.44%	26.79%	23.14%	19.48%	15.83%	12.17%	8.52%	4.87%
	70%	34.91%	31.26%	27.60%	23.95%	20.30%	16.64%	12.99%	9.33%	5.68%	2.03%
	80%	32.07%	28.42%	24.76%	21.11%	17.46%	13.80%	10.15%	6.49%	2.84%	-0.81%
	90%	29.23%	25.58%	21.92%	18.27%	14.62%	10.96%	7.31%	3.65%	0.00%	-3.65%
	100%	26.39%	22.74%	19.08%	15.43%	11.78%	8.12%	4.47%	0.81%	-2.84%	-6.49%

The current conversion rates are taken as basis (90% hemicellulose and cellulose conversion) and the percentage deviations are calculated accordingly. For example, if the hemicellulose conversion rate is lowered to 50% and the cellulose conversion rate is lowered to 70%, the ethanol output will be lowered by 18.67%. To implement the result of this table to plant economics, Tables 13.10-13 are referred. Using the tables, a sensitivity factor on ethanol is obtained for each parameter. The sensitivity factor is defined as the ratio of change in economic parameter to the change in ethanol output.

Using the ethanol sensitivity results given in Table 13.10, it is seen that for 10% change in ethanol output the net present value changes 19.4% which yields a sensitivity factor of 1.94. Figure 13.6

show that the sensitivity of NPV on ethanol has a linear trend. Thus, it is assumed that the slop of the sensitivity line is 1.94. Taking the 18.67% change with reduction in lignocellulose conversion, the change in NPV is obtained as 36.22%. As a result, if the cellulose and hemicellulose conversions are reduced to 70% and 50% respectively, the NPV will be lowered by 36.22%. A similar approach can be applied to other parameters. The sensitivity factors of NPV, POP, ROR and DCFROR on ethanol are given in Table 13.15.

Table 13.15 The sensitivity factors on ethanol

	Sensitivity factor
NPV	1.940
POP	$2.5097E_{\%}^2-1.9124\ E_{\%}+1.3455^*$
ROR	1.816
DCFROR	1.334

$E_{\%}$ percentage change in ethanol

13.9.7. The Production Costs of Ethanol and Sugar

The production costs are determined based on gross income, net income (including tax and depreciation), net income + depreciation (cost excluding depreciation) and net income + tax (cost excluding tax but including depreciation) (See Table 13.16). The production cost of sugar production in 2015 and 2005 (See Table 12.14). The ethanol costs are calculated by reducing sugar costs from the total cost of 2015 plant.

Table 13.16 The production costs ($/ton)

	Gross	Net	Net+depr.	Net+tax
Ethanol	167.9	271.4	254.1	185.3
Sugar	107.4	148.8	136.3	119.8

14. Discussion and Conclusions

The primary aim of this study is to investigate and analyze the sugar-ethanol plants operating in Brazil to construct a raw model to gain better understanding and insight about the technical and economical aspects of the currently operating plants. And the secondary aim is to combine the knowledge gathered from the currently existing plant with the future design idea of converting lignocellulose into ethanol together with sucrose.

During the design the data was gathered from literature, internet and industrial clients taking part in the projects and a clear view of sugar – ethanol plant is obtained. However, there are some points that required special attention. In the following sections, the points of interests are going to be discussed.

The design calculations proved that both 2005 and 2015 plants are technically and economically feasible to invest on. And the plants are sufficient to produce the products of specified properties.

14.1. Discussion on 2005 Plant

14.1.1. Equipment Design

The milling section of the design is a straight forward application and is carried out same in all sugar ethanol plant. The variations occur in number of mills in one tandem. In the current design, the number is taken as 5+1 to meet the capacity of 5 million tons per year of sugar cane input. Since the product sugar is selected as white sugar the raw juice is to be bleached with SO_2 to help the removal of the colorant materials. This step can also be carried out via carbonation using carbon dioxide. However, sulphitation is more effective in removing the colorants that carbonation and more widely applied in Brazil then sulphitation. This is why sulphitation is selected as the bleaching.

The first point that needs attention is the overliming of the juice that goes to ethanol production (Stream <115>) Since the major reason of overliming and sulphitation is removal of the impurities for the refined sugar production, this step in ethanol branch can be skipped and a significant investment will be saved. However, since the currently available information reports the usage of overliming in ethanol branch the design is carried out accordingly. The design can be revised after discussion with the operators of sugar-ethanol plant.

Another point of interest is the multi effect evaporators. The number of effects is selected as 6 based on the plant visits of the clients. The number of effects should be determined based on the trade of between steam cost and investment requirement arising from the additional effect. The final design data shows that the plant is producing more steam than required by the plant. Since there is excess steam, the number of effects can be lowered by increasing the steam input to minimize the investment.

The most common operation of the vacuum pans is batch operation. But, after 80’s the continuous operation is introduced especially by the European sugar producers. Current design is also utilizes continuous operation of boiling pans. But, since the technology of the continuous operation is not well defined the detailed design of the equipments is carried out based on batch equipments. For the details see Section 8.4.1.

In Brazil, the fermentation step is carried out in batch, continuous and fed batch cultures. Since the juice is not concentrated the fed batch option is not feasible. In terms of investment, continuous operation is more profitable, however, the microorganism specific rates that is obtained from the literature is not able to simulate the fermentation behavior at high cell concentrations (i.e. the fermentation times and low cell yields). Thus, the parameters of the organisms are estimated based on the fermentation times and cell concentrations reported in literature and taken from industry. More reliable microbial parameters would lead to better modeling of the fermentation and the selection of the operation mode can be more reliable.

To prevent contamination during the recycle the cell culture is washed with low pH sulphuric acid solution. Current data states that the acid solution is used once and discarded. However, by reusing the acid solution several times, both the waste treatment loads and cost of operation can be lowered.

The distillation and rectification are strait forward and common in all plants. However final dehydration of the ethanol to anhydrous ethanol can be carried our using extractive distillation, azeotropic distillation and molecular sieve adsorption, due to the lover operation costs the molecular sieve adsorption is selected as the dehydration unit (for detailed analysis see Benson, 2003). However, the selection of the operation phase is selected as liquid due to the available equilibrium data for liquid phase adsorption. The vapor equilibrium data should be obtained to select the best operation phase.

The electricity generation in 2005 plant is obtained as 77.4 MW. This value is higher then the reported industrial value from Brazil. The deviations in the electricity output could be due to the lack of fit in furnace exhaust gas temperature assumption. If the release temperature of the furnace off gases is higher than the design specification, the electricity production will be lower.

The process water requirement of the plant is supplied with the condensates from the evaporation section and the plant is self sufficient in terms of water. Since the small amount of organic impurities present in condensates arises from the sugar cane, the evaporator vapor and condensate can be directly used as process water without further treatment. The water is cooled to process requirements via cooling tower. The design of the cooling tower is outside the battery limits, but, the investment is included in economic calculations.

In current design, the vinasse and filter cake is assumed to be sent back to fields without any treatment. But, the acid solution is neutralized outside the battery limits. The cost is included in the cash flow analysis. The fusel oils are leaving the battery limits as dilute solution of water. It is neither included as revenue nor as cost in the cash flow analysis.

The designed plant produces 99.88% anhydrous ethanol and white sugar with 165,956 and 370,401 ton yearly capacities respectively.

14.1.2. Economy

The results of economical analysis of the 2005 plant reveal that the required total capital investment is $102 million. The investment is determined using the location factor of 0.4 for Brazil. This factor is determined based on the social and economical comparison of Brazil with US and European countries. Based on this investment value, the net present value of the investment is $166 million using discount rate of 4%. The sensitivity analysis shows that the plant operates profitable for location factor of up to 0.7. The pay out periods does not exit 3.5 years up to location factor 0.7. This time period reported to be acceptable for chemical plants (Peters and Timmerhaus). The same analysis reveals that the maximum interest rate that investment can remain economical through out the investment period is 26.5% for location factor of 0.4. This value reduces down to 3.7% when the location factor is increased to 0.7. When compared to current US inflation rate (~3.17%) the plant still remains compatible with the market (See Figures 12.3 to 12.5 and Table 12.9 for the results).

The cash flow analysis reveals that the sugar cane cost dominates the yearly operational cost (87%). Therefore, the economic parameters are highly sensitive to sugar cane cost. The other cost factors does not have major effect in the economical parameters. The sensitivity analysis is carried out on net present value, pay out period, discounter cash flow rate of return and rate of return values. The results are presented in Tables 12.10 to 13 and Figures 12.6 to 9. Although the parameters are highly sensitive to sugar cane price, the net present value of the investment is $106 million even when the sugar cane price is 20% larger than current prices.

The parameters also have high sensitivity to revenues (ethanol and sugar). The sensitivity on electricity is lower, but, still higher than the sensitivities to minor operating cost like utilities, labor etc (See Tables 12.10 to 13 and Figures 12.6 to 9).

The production costs of sugar and ethanol are presented in Table 12.14 and reveals that the production cost of ethanol is higher then the current reported production costs ($190/ton). However, with current sugar cane cost the minimum production cost excluding everything, but, the sugar cane is approximately $140/ton.

The overall view of the 2005 plant proves to be economical and favorable for the investment even for investment cost of up to $170 million.

14.2. Discussion on 2015 Plant

14.2.1. Equipment Design

The milling, sugar production and ethanol production sections of the design are the same with 2015 plant. The major difference is in pretreatment and hydrolysis section and the fermentation sections.

The first point of interest is fermentation of pentose sugars. The recent studies (2004-2005) of Marco Kuyper and his coworkers in biotechnology department of TU Delft reveals that the recombinant *Sacharomyces cerevisiae* that can metabolize pentose and hexose sugars proves to be a favorable option for the ethanol production from lignocellulose hydrolyzates. Although the current production rates are lower when compared to wild type high productive hexose fermenting organisms, the future progress is promising to reach to productivities compatible for the existing wild type hexose fermenting microorganisms.

As described in Section 2, the acid catalyzed steam explosion proved to be more feasible both in terms of technical and economical aspects. The results of the design show that the ethanol production is increased to ~2.4 fold by the conversion of lignocellulose to ethanol. The only draw back is the presence of the acids and the fiber in the system. This is why the maintenance costs of the plant are taken higher in 2015 economical evaluation to account for the maintenance costs arising from the pretreatment section.

During the design of the pretreatment section, most of the time the pulp and paper industries are referred to be inspired to apply on the current design. The continuous operation of the pretreatment reactor is achieved by utilizing the design of continuous digesters in pulp and paper industries. The dewatering of the fibers is achieved using screw presses which are also applied in pulp and paper industries.

The cellulose hydrolysis unit is the one with more attention given in recent years. However, the final achievements of Novonzymes reduced the contribution of the cellulase enzyme to ethanol production to \$0.03 to \$0.06 per kg of ethanol produced. But, the contribution of cellulase enzyme to cost of production is even lowered when the design of the pretreatment is combined with the conventional plant (this study). Therefore, the most important out come of the design is the investment and operational cost reductions with combined operation.

The cogeneration unit in 2015 is utilized not only to generate steam and electricity, but, also to combust vinasse, filter cakes and recombinant yeast. The combustion of vinasse contributes to total energy generations with fraction of ~20%. This application reduced the waste output of the plant significantly. The steam generation is manipulated to recover energy for the production of electricity. The plant is producing 13.5 MW electricity while the demand is slightly higher than 20 MW. Therefore, the excess electricity need is supplied by purchasing from the grid. But, the steam production is supplying the steam demand of the plant.

The wastes generated in the plant are organic material free gypsum, dirt and ash which are send to landfill, organic material containing waste water and the NO_x rich exhaust gas from the furnaces. The waste waters are treated in waste water treatment plants outside the battery limits. The design is taken outside the battery limits but the costs are included in investment and operational costs. The exhaust gas should be refined using absorption to absorb the NO_x and SO_2. The waste treatment section requires special attention since it is not investigated in this project.

The process water requirement of the plant is higher than the water output from the evaporator condensates and waste water treatment. The major reason for this is the excess amount of water leaving the battery limits with the fusel oil stream. Thus, the excess process water is purchased from the grid.

The overall view of 2015 design proves to be technically feasible with the current situation. However, for the clarity pilot plant studies should be carried out to gain better insight of pretreatment reactor operation.

The designed 2015 plant produces 99.89% anhydrous ethanol and white sugar with 396,872 and 370,405 ton yearly capacities respectively.

14.2.2. Economy

The investment requirement for 2015 plant is obtained as \$133 million after including the location factor of 0.4 for Brazil. The net present value of 2015 plant is obtained as \$370 million which is more then double the present value for 2015 plant. The reason for this is when the lignocellulose hydrolysis is included in the design the investments did not change significantly while the revenues are almost doubled. The pay-out period for the 2015 plant is obtained as 1.4 which means the plant recovers the investment in less then 2 years. The analysis shows that the plant remains profitable as long as the location factor is increased up to 1. The discounted cash flow analysis reveals that the plant investment can stand interest rates up to 38% and 7.3% for location factor of 0.4 and 0.9 respectively (See Figures 13.3 to 13.5 and Table 13.9 for the sensitivity analysis on location factor).

As in 2005 plant, the cost of sugar cane and selling price of ethanol and sugar are the major disturbances for the economic parameters. However, even a negative deviation of 30% in sugar cane price from the current prices yields a net present value of $150 million which proves to be a profitable investment. In 2015 plant, the economic parameters are not as sensitive to sugar cane and sugar prices as the ethanol prices. But, still has major effects. The variations of ±30% in ethanol and sugar prices yielded ±60% in the net present value of the investments. The sensitivity on cellulase prices is not as high as the sugar cane and product prices. The variations of ±30% in cellulase price yielded ±6% in net present value of the investment. The rest of the cost factors (i.e. labor, fixed cost etc.) do not have major effect on the economic parameter. The variations in fixed cost influenced the parameters most. But, the sensitivity is still ±6% for variations of ±30%. (See Figures 13.6 to 13.13 and Tables 13.10 to 13.13 for the results)

The production cost of ethanol (~$167/ton) is significantly lower than the 2005 plant production costs. This shows that the combined plant design enhances the favorability of the investment on sugar ethanol plant (See table 13.11 for the production costs).

15. Recommendations

There are several points that should be considered for future studies following the current one. The points that needs special attention are listed below.

- Due to the time limitations, some of the options are not considered during the design. For future studies following options can be considered to enhance the technical and economical attributes of the plant
    - In 2015 plant, the ethanol branch clarification can be skipped and the sugar that is diverted into ethanol branch can be directly sent to pretreatment unit in bagasse with out being extracted. By doing so, the wash water requirement of the mills can be skipped and the juice that is sent to the sugar production can be more concentrated. This will reduce the operation and investment costs of sugar branch and eliminate the investment cost of ethanol branch juice clarification branch.
      In this option, the major draw back is the inhibition of the cellulase enzyme due to the presence of hexose sugars. But, using innovative designs, this drawback can be eliminated.
    - In another design, the ethanol production from sucrose can be operated during harvest season while the ethanol production from bagasse is carried out whole year.
    - Another alternative might be operating the sugar ethanol plant during the harvest season and store the bagasse. And operate the lignocellulose to ethanol section in the off season. This option can lower the investment costs since the equipment sizes will be smaller.
    - Since the electricity generation of the plant is not sufficient the conversion of lignocellulose to C_5&C_6 sugars can be lowered to have larger lignocellulose residues. When the conversions are lowered the acid utilization in pretreatment can also be skipped. And steam explosion with out acid catalysis can be used.

- The growth characteristics of hexose and hexose-pentose metabolizing organisms should be determined under high cell concentrations for better approximation of the fermentation unit.
- The effects of inhibition mechanism at high cell concentrations should be investigated to check the validity of detoxification assumptions.
- The molecular sieve adsorption column needs the equilibrium data of vapor phase water adsorption for better comparison of liquid phase and vapor phase adsorption.
- The boiler specifications should be specified properly to estimate the electricity and steam generations more accurate.
- Although a preliminary heat integration is carried out, a more detailed pinch analysis can lower the energy lost further and increase the electricity output of the 2015 plant.
- More expertise is required for the investment calculations because the prices of the equipments are not obtained from the suppliers, but, from the cost models from the literature.

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A.1. Process Block Schemes

A.1.1. 2005 Block Scheme

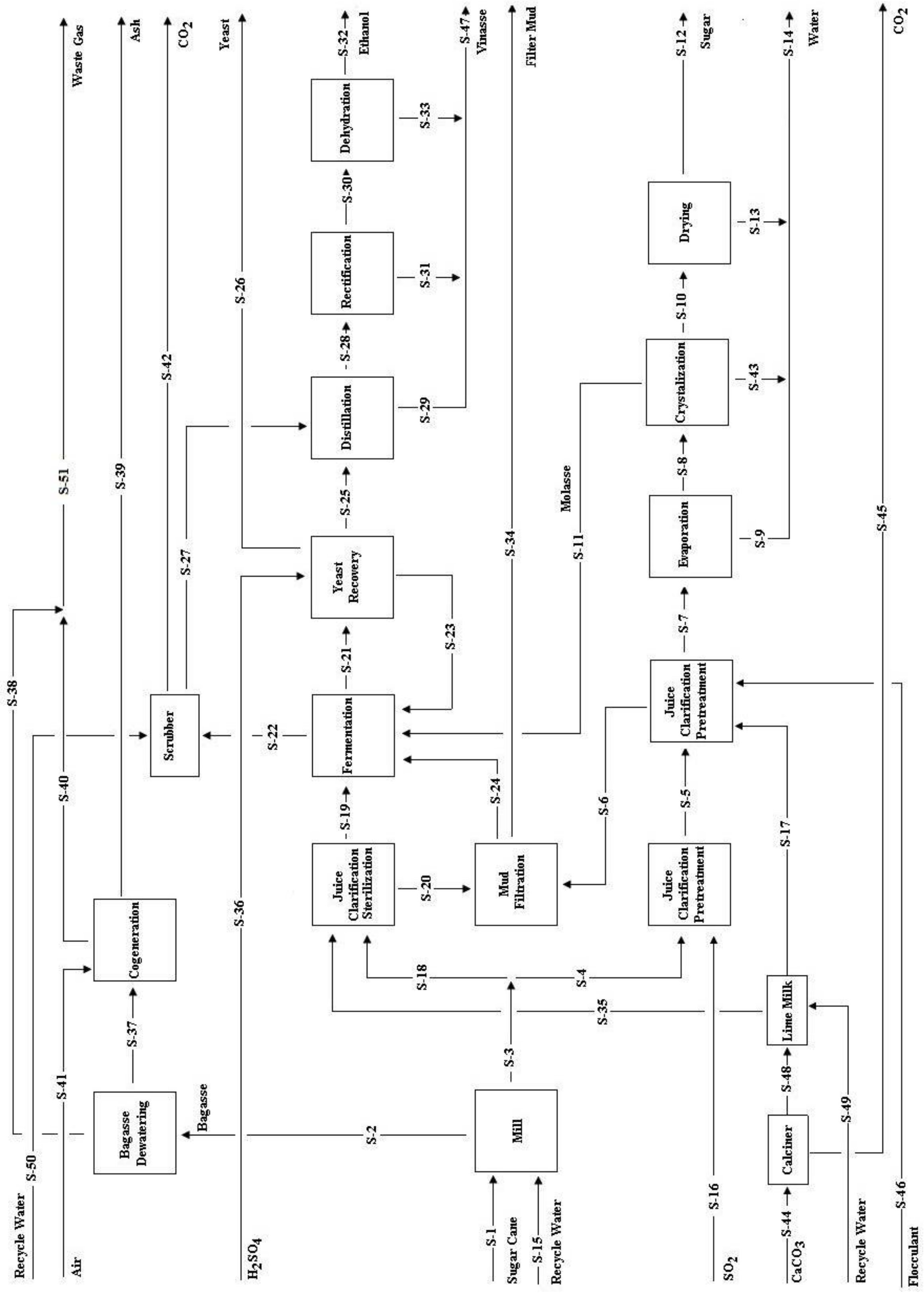


Figure A.1.1 2005 process block Scheme

A.1.2. 2015 Block Scheme

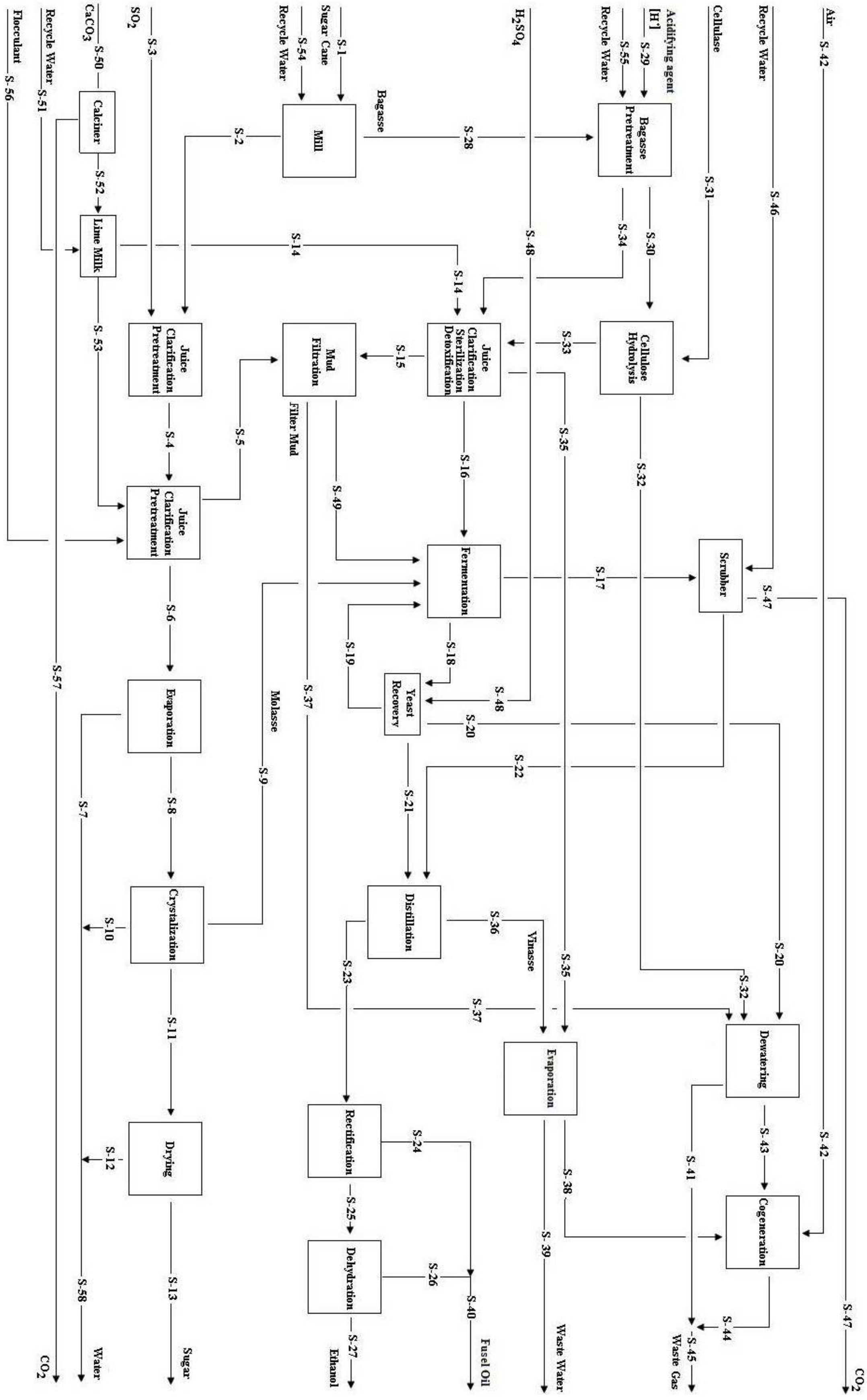


Figure A.1.2 2015 process block Scheme

A.2. Thermodynamic and Pure Compound Properties

A.2.1. Thermodynamic Properties

Table A.2.1 The thermodynamic properties of compounds existing in mass balances

Component Name	Technological Data						Notes
	Formula	State (1)	Gibbs Energy (1) kJ mol^{-1}	Enthalpy (1) kJ mol^{-1}	Entropy (1) J °C^{-1} mol^{-1}	Heat Capacity (1) J mol^{-1} K^{-1}	
Arabinose	$C_5H_{10}O_5$	S	n.a.	n.a.	n.a.	n.a.	
Acetic acid	$C_2H_4O_2$	L	-390.2	-484.8	159.9	124.3	(4)
Calcium hydroxide	$Ca(OH)_2$	S	-899.2	-986.6	83.4	87.5	(4)
Carbon dioxide	CO_2	G	-394.6	-393.4	213.8	37.1	(4)
Ethanol	C_2H_6O	L	-175.0	-277.9	160.8	111.5	(4)
Glucose	$C_6H_{12}O_6$	S	n.a.	n.a.	n.a.	n.a.	(3)
Polyacrylamide	$-[C_6H_8NO_3Na]-$	S	n.a.	n.a.	n.a.	n.a.	(3)
Furfural	$C_5H_4O_2$	L	n.a.	n.a.	n.a.	n.a.	(3)
Glycerol	$C_3H_5(OH)_3$	L	-488.5	-676.0	n.a.	218.3$^{(5)}$	
Calcium suphate	$CaSO_4$	S	-1,322.7	-1,435.1	106.8	99.7	(4)
Hydroxymethylfurfural	$C_6H_6O_3$	L	n.a.	n.a.	n.a.	n.a.	(3)
Calcium carbonade	$CaCO_3$	S	-1,129.6	-1,207.7	92.9	81.9	(4)
Calcium oxide	CaO	S	-604.4	-635.5	39.8	42.8	(4)
Isoamylalcohol		L	n.a.	n.a.	n.a.	n.a.	(3)
Lactic acid		L	-517.2	-687.0	n.a.	n.a.	
Mannose	$C_6H_{12}O_6$	S	n.a.	n.a.	n.a.	n.a.	(3)
Methane	CH_4	G	-50.7	-66.9	186.3	35.6	(4)
Microorganism	$CH_{1.8}O_{0.5}N_{0.2}$	S	-67.0	-91.0	n.a.	n.a.	(6)
Nitrogen	N_2	G	0.0	0.0	191.6	29.1	(4)
Nitrogen dioxide	NO_2	G	33.2	51.3	240.1	29.86	(4)
Oxygen	O_2	G	0.0	0.0	205.1	29.4	(4)
Sucrose	$C_{12}H_{22}O_{11}$	S	n.a.	n.a.	n.a.	n.a.	(3)
Succinic acid	$C_4H_6O_4$	S	-690.2	-909.0	n.a.	141.5$^{(5)}$	(6)
Sulphur	S	S	-	-	31.8	22.7	(4)
Sulphur dioxide	SO_2	G	-297.0	-300.4	248.3	39.9	(4)
Sulphuric acid	H_2SO_4	S	-	-	-	-	
Water	H_2O	L	-243.7	-286.0	76.0	84.4	(4)
Xylose	$C_5 H_{10}O_5$	S	n.a.	n.a.	n.a.	n.a.	(3)

Notes:

(1) Properties of formation at standard conditions 25ºC and 1bar
(2) Molecular formula as representative
(3) Data cannot be found
(4) Weast C. R., (1983)
(5) Perry and Green (1997)
(6) Advanced Course on Environmental Biotechnology, Kluyvers Laboratory, Technical University Delft.

A.2.2. Pure Compound Properties

Table A.2.2 The pure compound properties of compounds existing in mass balances

Component Name	Technological Data					Medical Data		Notes
	Formula	Mol. Weight g/mol	Boiling Point ºC	Melting Point ºC	Liquid Density kg/m^3	MAK Value mg/m^3	LD$_{50}$ Oral g/kg	
Arabinose	$C_5H_{10}O_5$	150.13	n.a	157.00	1.58	n.a.	n.a.	(1)
Acetic acid	$C_2H_4O_2$	60.05	118.20	16.60	1,050.00	n.a.	n.a.	
Calcium hydroxide	$Ca(OH)_2$	74.09	D	580.00	2,240.00^S	5.0	7.34	(2)
Carbon dioxide	CO_2	44.01	-78.50	-207.00	1,032.00	n.a.	90.00	(4)
Ethanol	C_2H_6O	46.06	78.00	-114.00	780.00	1,900.0	5.00	(5)
Galactose	$C_6H_{12}O_6$	180.08	n.a.	167.00	n.a.	n.a.	440.0	(6)
Glucose	$C_6H_{12}O_6$	180.08	n.a.	150.00	1.54	n.a.	25.80	(7)
Polyacrylamide	$-[C_6H_8NO_3Na]-$	165.00	-	-	1,190.00	n.a.	11.20	(8)
Furfural	$C_5H_4O_2$	96.08	161.70	-36.50	1.1567	20.0	65.00	(9)
Glycerol	$C_3H_5(OH)_3$	92.09	182.00	18.00	1.25	10.0	12.60	(10)
Calcium suphate	$CaSO_4$	136.14	D	1,450.00	1,043.00^S	10.0	n.a.	(11)
Hydroxymethylfurfural	$C_6H_6O_3$	126.00	n.a.	n.a.	n.a.	n.a.	n.a.	
Calcium carbonade	$CaCO_3$	100.09	n.a.	900.00^D	2.71	10.0	n.a.	(12)
Calcium oxide	CaO	56.10	2,850.00	2,570.00	3,200.00^S	5.0	n.a.	(13)
Isoamylalcohol	$C_5H_{12}O$	88.20	132.00	-117.00	800.00	370.0	0.10	(24)
Lactic acid	$C_3H_6O_3$	90.04	122.00	16.80	1,249.00^S	n.a.	3.73	
Mannose	$C_6H_{12}O_6$	180.08	n.a.	132.0	1.54	n.a.	n.a.	(14)
Methane	CH_4	16.04	-161.00	-182.5	422.62	n.a.	n.a.	(15)
Micro organism	$CH_{1.8}O_{0.5}N_{0.2}$	24.60	n.a.	n.a.	1,050.00	n.a.	n.a.	
Nitrogen	N_2	28.01	-195.80	-210.00	808.61	n.a.	n.a.	(16)
Nitrogen dioxide	NO_2	46.05	21.10	-11.20	1,443.00	H.T.	H.T.	(17)
Oxygen	O_2	32.00	-182.80	-218.64	1,141.00	n.a.	n.a.	(18)
Sucrose	$C_{12}H_{22}O_{11}$	342.30	D	190.00^D	1,586.20^S	n.a.	29.70	(19)
Succinic acid	$C_4H_6O_4$	118.10	235.00	188.00	1,570.00^S	n.a.	2.26	(20)
Sulphur	S	32.06	445.00	116.00	n.a.	n.a.	n.a.	(25)
Sulphur dioxide	SO_2	64.06	-10.20	-75.50	1,455.00	n.a.	n.a.	(21)
Sulphuric acid	H_2SO_4	98.08	337.00	10.00	1,800.00	n.a.	2.14	(22)
Water	H_2O	18.01	100.00	0.00	950.00	n.a.	n.a.	(23)
Xylose	$C_5H_{10}O_5$	150.06	n.a.	140.00	1.53	n.a.	n.a.	

Notes:
1) Density is specific gravity at 25ºC (IRef-19)
2) IRef-20
3) IRef-21
4) Density at at -20 °C and 19.7 bar, LD$_{50}$ is given for 5 minutes exposure (IRef-22)
5) IRef-23
6) IRef-24
7) Density is specific gravity at 25ºC (IRef-25)
8) Density is for 50% aqueous solution

The boiling point is at 3.33 kPa. At ambient pressure polymerize, density is solid density at ambient conditions (IRef-26)
9) Density is specific gravity at 25ºC (IRef-27)
10) Density is specific gravity at 25ºC (IRef-28)
11) IRef-29
12) Density is specific gravity at 25ºC (IRef-30)
13) Density is specific gravity at 25ºC (IRef-31)
14) Density is specific gravity at 25ºC (IRef-32)
15) Density at 1.013 bar at boiling point (IRef-22)
16) Density at 1.013 bar at boiling point (IRef-22)
17) Density at 1.013 bar at boiling point (IRef-22)
18) Density at 1.013 bar at boiling point (IRef-22)
19) IRef-33
20) Density at 25ºC (IRef-34)
21) Density at 1.013 bar at boiling point (IRef-22)
22) IRef-35
23) Density is specific gravity at 25ºC (IRef-36)
24) IRef-37
25) IRef-38

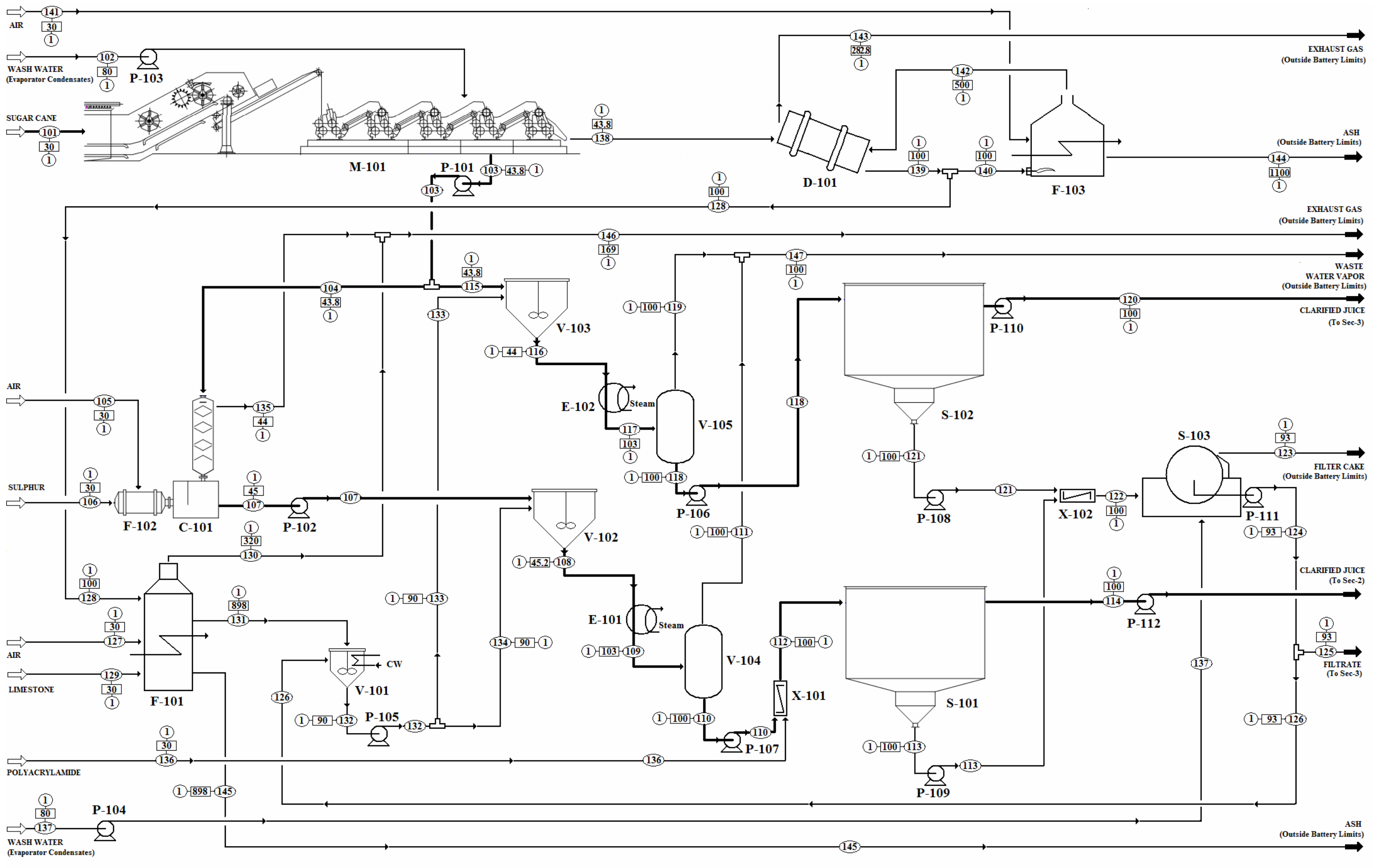
D: Decomposes – S: solid density- H.T.: highly toxic

A.3. Process Block Schemes

A.3.1. 2005 Process Block Schemes

2005 process block scheme is composed of three sections

- Mill & Clarification (Section 1)
- Juice evaporation & Crystallization (Section 2)
- Ethanol production & Purification (Section 3)



PROCESS EQUIPMENT SUMMARY

D-101 : Bagasse Dewatering Drum Drier
C-101 : Sulphitation Column
E-101 : Sugar Branch Juice Heater
E-102 : Ethanol Branch Juice Heater
F-101 : Lime Kiln
F-102 : Sulphur burner
F-103 : Cogeneration Boiler

M-101 : Roller Mills
S-101 : Clarifier&Settler
S-102 : Clarifier&Settler
S-103 : Rotary Vacuum Filter
P-101 : Circulation Pumps
P-102 : Sulphitated Juice Pumps
P-103 : Mill Wash Water Pumps

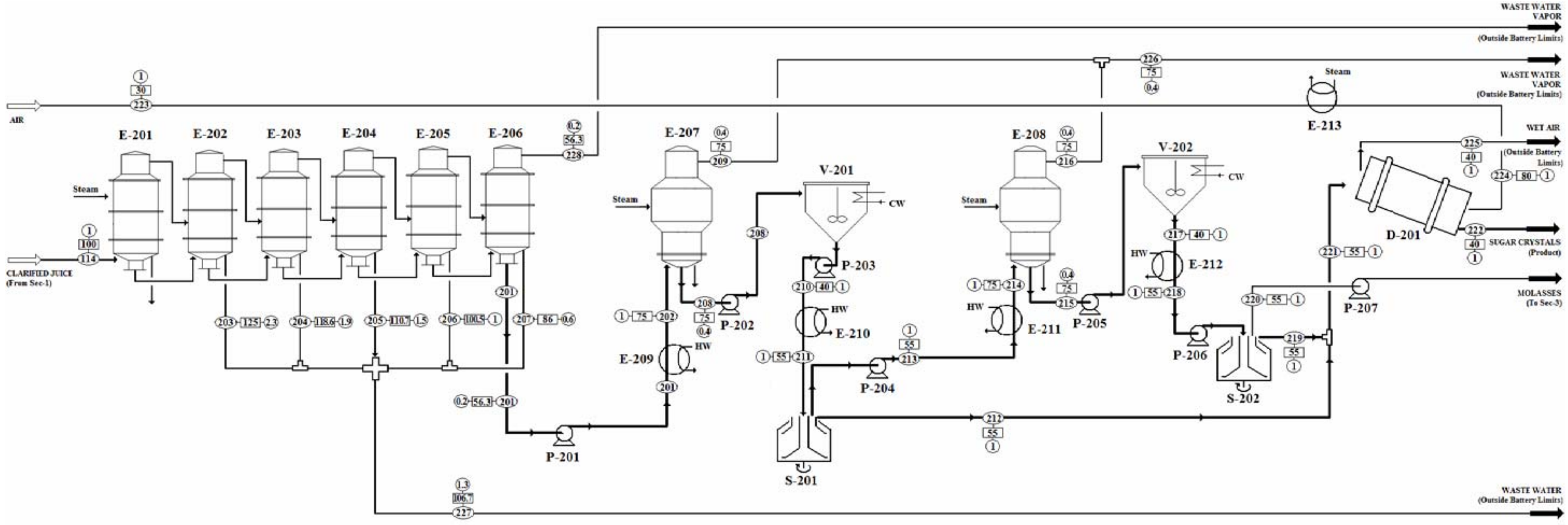
P-104 : Rotary Filter Wash Water Pump
P-105 : Lime Milk Pumps
P-106&107 : Limed Juice Pumps
P-108&109 : Clarifier Mud Pumps
P-110&112 : Clarified Juice Pumps
P-111 : Rotary Filter filtrate Pump
V-101 : Lime Milk Preparation Vessel

V-102&103 : Liming Tanks
V-104&105 : Flash Tanks
X-101&102 : Static Mixers

PROCESS FLOW SCHEME

PROJECT NAME	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse

DESIGNER	DATE	2005 SECTION 1 : MILL AND CLARIFICATION SECTION
Çağrı Efe	28-09-2005	◯ Stream Number ▭ Temperature (°C) ◯ Pressure (bar)

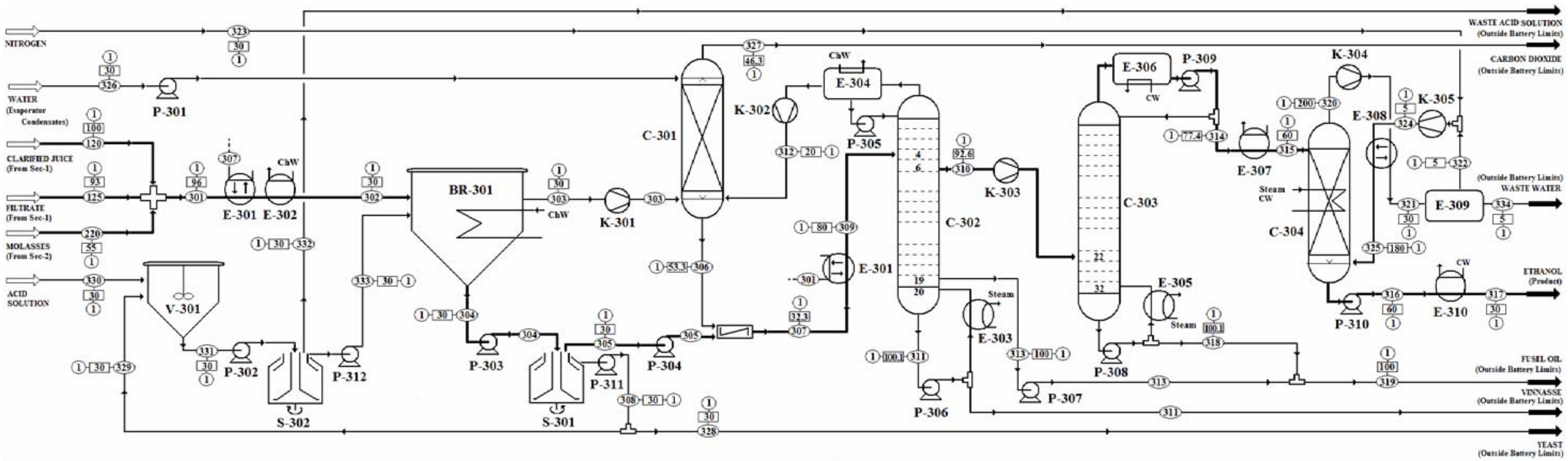


PROCESS EQUIPMENT SUMMARY

D-201	: Sugar Rotary Drier
E-201	: 1st Effect Evaporator
E-202	: 2nd Effect Evaporator
E-203	: 3rd Effect Evaporator
E-204	: 4th Effect Evaporator
E-205	: 5th Effect Evaporator
E-206	: 6th Effect Evaporator
E-207	: 1st Boiling Pan
E-208	: 2nd Boiling Pan
E-209	: Juice Heater
E-210	: Massecuites Reheater
E-211	: Molasse Reheater
E-212	: Massequites Reheater
E-213	: Drier Air Heater
P-201	: Juice Pumps
P-202	: Massecuite Pumps
P-203	: Massecuite Pumps
P-204	: Molasses Pumps
P-205	: Massecuite Pumps
P-206	: Massecuite Pumps
P-207	: Molasses Pumps
S-201	: Crystal Sugar Separation Centrifuges
S-202	: Crystal Sugar Separation Centrifuges
V-201	: 1st Boiling Pan Sugar Crystallizers
V-202	: 2nd Boiling Pan Sugar Crystallizers

PROCESS FLOW SCHEME

PROJECT NAME	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse	
DESIGNER	DATE	2005 SECTION 2 : JUICE EVAPORATION AND CRYSTALLIZATION ⬭ Stream Number ▭ Temperature (℃) ○ Pressure (bar)
Çağrı Efe	28-09-2005	



PROCESS EQUIPMENT SUMMARY

BR-301	: Fermentors	**E-303**	: Beer Column Reboiler	**E-310**	: Ethanol Cooler	**P-302**	: Biomass Pump	**P-309**	: Rectification Condenser Pump
C-301	: Ethanol Scrubber	**E-304**	: Beer Column Condenser	**K-301**	: CO2 Blower	**P-303**	: Fermentation Broth Pump	**P-310**	: Product Pump
C-302	: Beer Column	**E-305**	: Rectification Column Reboiler	**K-302**	: CO2 Blower	**P-304**	: Yeast Free Broth Pump	**P-311&312**	: Biomass Pump
C-303	: Rectification Column	**E-306**	: Rectification Column Condenser	**K-303**	: Product Vapor Blower	**P-305**	: Beer Condenser Pump	**S-301**	: Yeast Separation Centrifuges
C-304	: Molecular Sieve Column	**E-307**	: Molecular Sieve Feed Cooler	**K-304**	: Wet Nitrogen Blower	**P-306**	: Beer Reboiler Pump	**S-302**	: Yeast Separation Centrifuges
E-301	: Beer Column Feed Heater	**E-308**	: Wet Nitrogen/Dry Nitrogen Heat Ex.	**K-305**	: Dry Nitrogen Blower	**P-307**	: Fusel Oil Pump	**V-301**	: Yeast Sterilization Tank
E-302	: Juice Cooler	**E-309**	: Nitrogen Drying Condenser	**P-301**	: Water Pump	**P-308**	: Rectification Reboiler Pump		

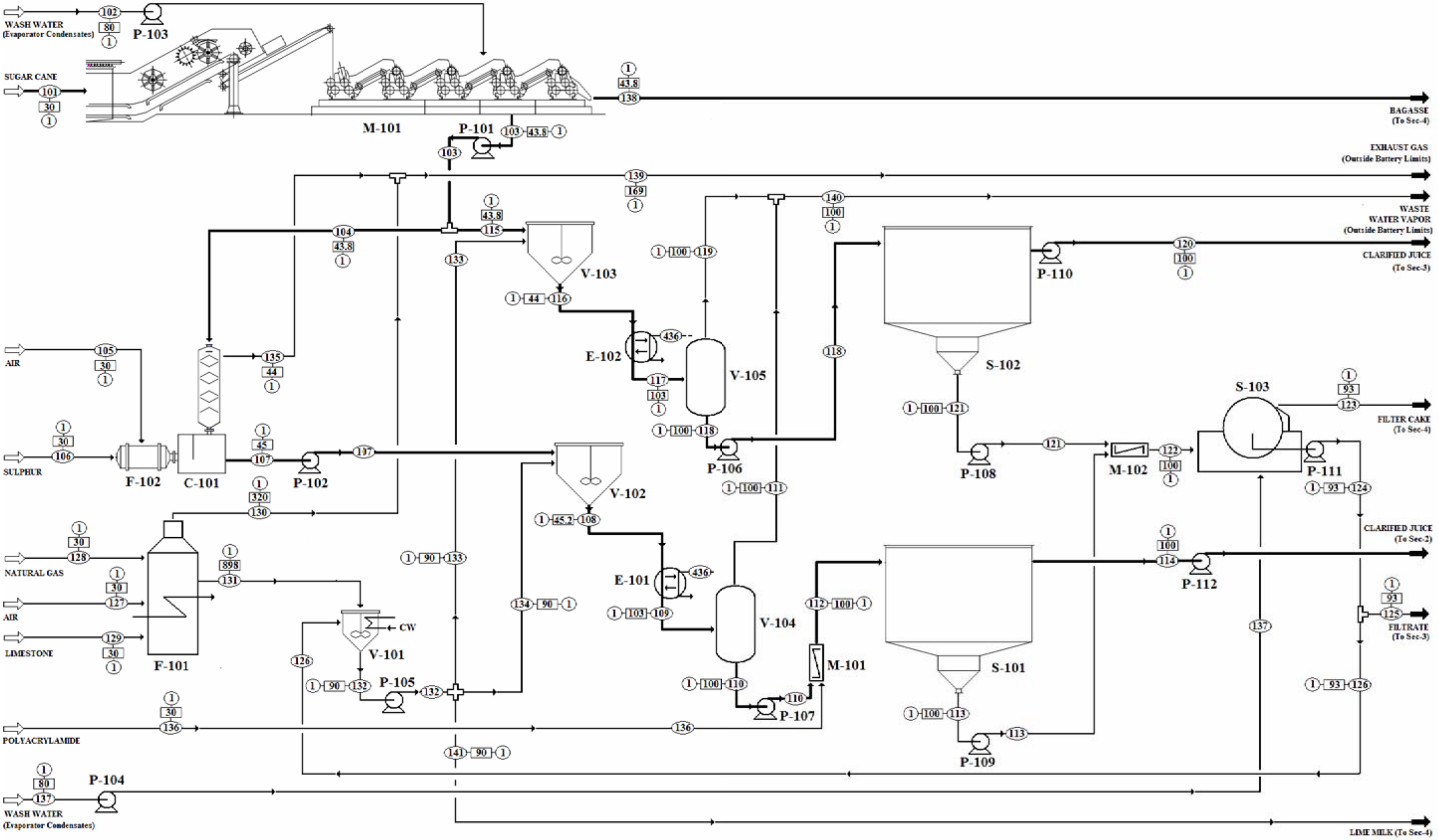
PROCESS FLOW SCHEME

PROJECT NAME	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse	
DESIGNER	DATE	2005 SECTION 3 : FERMENTATION AND ETHANOL PURIFICATION SECTION
Çağrı Efe	28-09-2005	⬭ Stream Number ▭ Temperature (°C) ○ Pressure (bar)

A.3.2. 2015 Process Block Schemes

2015 process block scheme is composed of four sections

- Mill & Clarification (Section 1)
- Juice evaporation & Crystallization (Section 2)
- Ethanol production & Purification (Section 3)
- Pretreatment & Hydrolysis (Section 4)



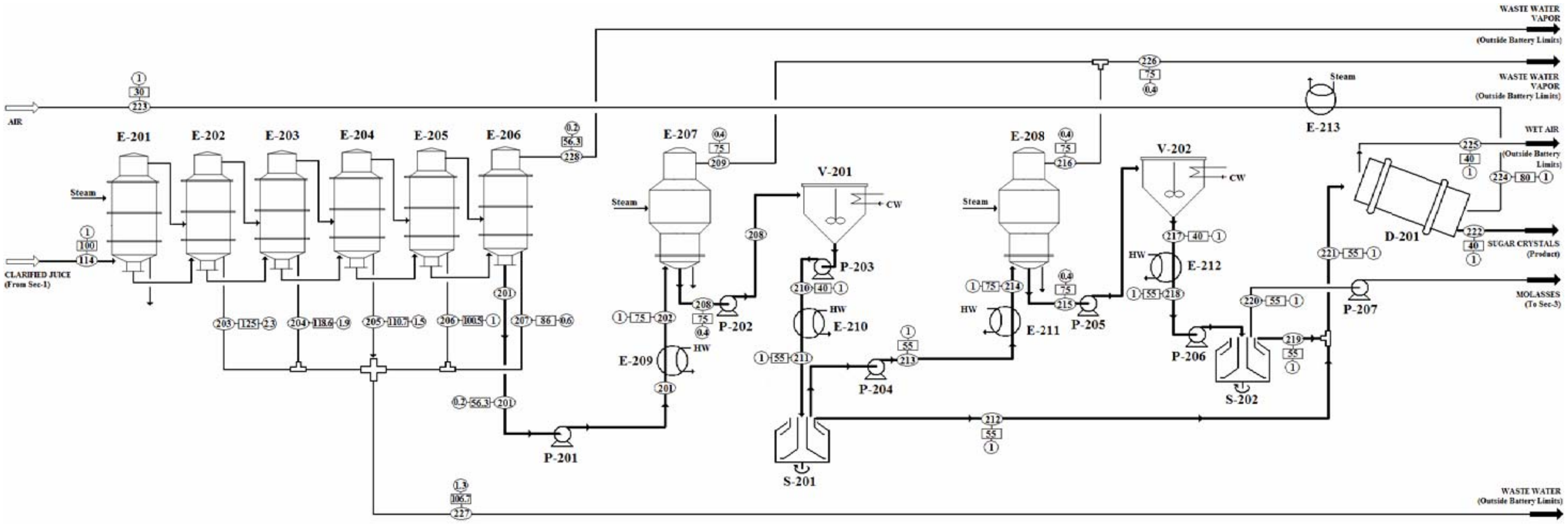
PROCESS EQUIPMENT SUMMARY

C-101	: Sulphitation Column
E-101	: Sugar Branch Juice Heater
E-102	: Ethanol Branch Juice Heater
F-101	: Lime Kiln
F-102	: Sulphur burner
M-101	: Roller Mills
S-101	: Clarifier&Settler
S-102	: Clarifier&Settler
S-103	: Rotary Vacuum Filter
P-101	: Circulation Pumps
P-102	: Sulphitated Juice Pumps
P-103	: Mill Wash Water Pumps
P-104	: Rotary Filter Wash Water Pump
P-105	: Lime Milk Pumps
P-106&107	: Limed Juice Pumps
P-108&109	: Clarifier Mud Pumps
P-110&112	: Clarified Juice Pumps
P-111	: Rotary Filter filtrate Pump
V-101	: Lime Milk Preparation Vessel
V-102&103	: Liming Tanks
V-104&105	: Flash Tanks
X-101&102	: Static Mixers

PROCESS FLOW SCHEME

PROJECT NAME	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse

DESIGNER	DATE	2015 SECTION 1 : MILL AND CLARIFICATION SECTION
Çağrı Efe	28-09-2005	⬭ Stream Number ▭ Temperature (°C) ○ Pressure (bar)

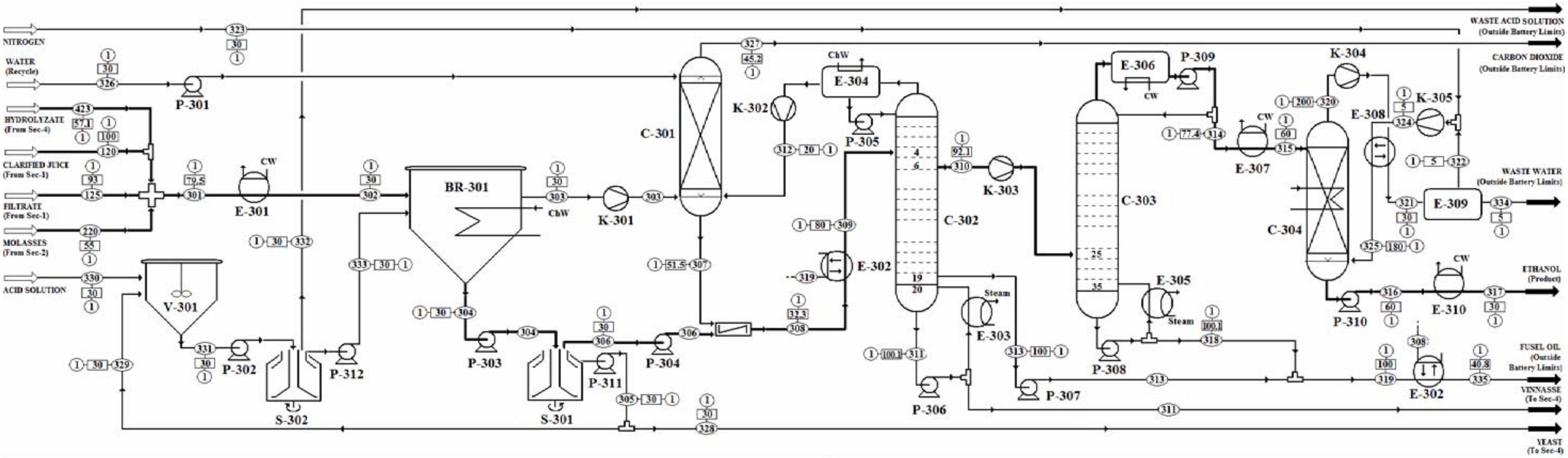


PROCESS EQUIPMENT SUMMARY

D-201	: Sugar Rotary Drier	**E-207**	: 1st Boiling Pan	**P-201**	: Juice Pumps	**S-201**	: Crystal Sugar Separation Centrifuges
E-201	: 1st Effect Evaporator	**E-208**	: 2nd Boiling Pan	**P-202**	: Massecuite Pumps	**S-202**	: Crystal Sugar Separation Centrifuges
E-202	: 2nd Effect Evaporator	**E-209**	: Juice Heater	**P-203**	: Massecuite Pumps	**V-201**	: 1st Boiling Pan Sugar Crystallizers
E-203	: 3rd Effect Evaporator	**E-210**	: Massecuites Reheater	**P-204**	: Molasses Pumps	**V-202**	: 2nd Boiling Pan Sugar Crystallizers
E-204	: 4th Effect Evaporator	**E-211**	: Molasse Reheater	**P-205**	: Massecuite Pumps		
E-205	: 5th Effect Evaporator	**E-212**	: Massequites Reheater	**P-206**	: Massecuite Pumps		
E-206	: 6th Effect Evaporator	**E-213**	: Drier Air Heater	**P-207**	: Molasses Pumps		

PROCESS FLOW SCHEME

PROJECT NAME	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse	
DESIGNER	DATE	2015 SECTION 2 : JUICE EVAPORATION AND CRYSTALLIZATION ⬭ Stream Number ▭ Temperature (°C) ○ Pressure (bar)
Çağrı Efe	28-09-2005	

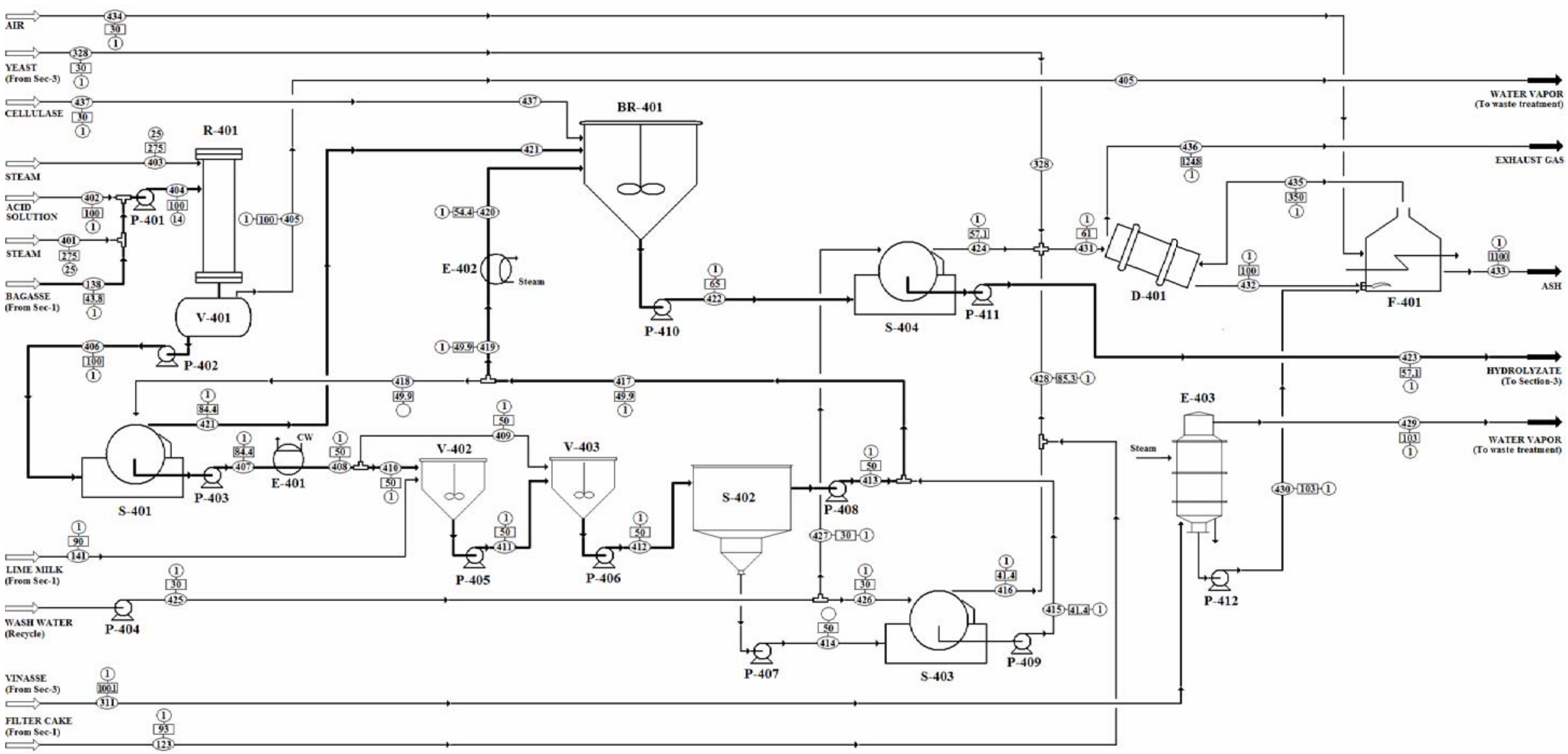


PROCESS EQUIPMENT SUMMARY

BR-301 : Fermentors
C-301 : Ethanol Scrubber
C-302 : Beer Column
C-303 : Rectification Column
C-304 : Molecular Sieve Column
E-301 : Juice Cooler
E-302 : Beer Column Feed Heater
E-303 : Beer Column Reboiler
E-304 : Beer Column Condenser
E-305 : Rectification Column Reboiler
E-306 : Rectification Column Condenser
E-307 : Molecular Sieve Feed Cooler
E-308 : Wet Nitrogen/Dry Nitrogen Heat Ex.
E-309 : Nitrogen Drying Condenser
E-310 : Ethanol Cooler
K-301 : CO2 Blower
K-302 : CO2 Blower
K-303 : Product Vapor Blower
K-304 : Wet Nitrogen Blower
K-305 : Dry Nitrogen Blower
P-301 : Water Pump
P-302 : Biomass Pump
P-303 : Fermentation Broth Pump
P-304 : Yeast Free Broth Pump
P-305 : Beer Condenser Pump
P-306 : Beer Reboiler Pump
P-307 : Fusel Oil Pump
P-308 : Rectification Reboiler Pump
P-309 : Rectification Condenser Pump
P-310 : Product Pump
P-311&312 : Biomass Pump
S-301 : Yeast Separation Centrifuges
S-302 : Yeast Separation Centrifuges
V-301 : Yeast Sterilization Tank

PROCESS FLOW SCHEME

PROJECT NAME	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse	
DESIGNER	DATE	2015 SECTION 3 : FERMENTATION AND ETHANOL PURIFICATION SECTION
Çağrı Efe	28-09-2005	⬭ Stream Number ▭ Temperature (℃) ○ Pressure (bar)



PROCESS EQUIPMENT SUMMARY

BR-401	: Cellulose Hydrolysis Vessels	**P-402**	: Fiber&Hydrolyzate Pumps	**P-409**	: Filtrate Pump	**S-403**	: Rotary Vacuum Filter
D-401	: Dewatering Rotary Drum	**P-403**	: Screw Press Permeate Pumps	**P-410**	: Hydrolyzate Pump	**S-404**	: Screw Press
E-401	: Hydrolyzate Cooler	**P-404**	: Water Pump	**P-411**	: Hydrolysate Pump	**V-401**	: Pretreatment Collection Tank
E-402	: Clarified Hydrolyzate Heater	**P-405**	: Hydrolyzate Pumps	**P-412**	: Concentrated Vinasse Pump	**V-402**	: Over Liming Vessel
E-403	: Vinasse Evaporator	**P-406**	: Hydrolyzate Pumps	**R-401**	: Pretreatment Reactor	**V-403**	: Conditioning Vessel
F-401	: Boiler Furnace	**P-407**	: Clarifier Mud Pump	**S-401**	: Screw Press		
P-401	: High Presure Fiber Pump	**P-408**	: Clarified Hydrolyzate Pump	**S-402**	: Clarifier&Settler		

PROCESS FLOW SCHEME

PROJECT NAME	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse	
DESIGNER	DATE	2015 SECTION 4 : BAGASSE PRETREATMENT AND HYDROLYSIS SECTION
Çağrı Efe	28-09-2005	⬭ Stream Number ▭ Temperature (℃) ○ Pressure (bar)

A.4. Process Steam Summaries

A.4.1. Process Stream Summary of 2005 Plant

STREAM Nr.		<101>		<101>		<102>		<103>		<104>		<105>		<106>		<107>	
Name :		Sugar Cane		Sugar Cane		Imbibition Water		Raw Juice		Raw Juice-Sugar		Air		Sulphur		Sulphitated Juice	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																
Ash	40.00																
Bagasse	257.84			46.562	0.181			0.303	0.001	0.303	0.001					0.303	0.001
Yeast	24.60																
$Ca(OH)_2$	74.09																
Ca-salts	136.14																
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01																
Dirt	40.00			3.326	0.083			2.162	0.054	2.162	0.054					2.162	0.054
Ethanol	46.06																
Glycerol	92.09																
Impurity	33.15			6.652	0.201			3.768	0.114	3.768	0.114					3.768	0.114
Isoamylalcohol	82.20																
N_2	28.01											0.734	0.026				
N_2O	46.05																
O_2	32.00											0.195	0.006				
Polyacrylamide	165.00																
Succinic Acid	118.10																
Sucrose	342.30			49.888	0.146			30.806	0.090	30.806	0.090					30.806	0.090
Sugar Cane	109.90	332.589	3.026														
SO_2	64.06															0.216	0.003
H_2SO_4	98.08																
Sulphur	32.06													0.108	0.003		
Water	18.02			226.160	12.551	93.125	5.168	177.899	9.872	177.899	9.872					177.881	9.871
Total		332.589	3.026	332.589	13.161	93.125	5.168	214.937	10.131	214.937	10.131	0.930	0.032	0.108	0.003	215.136	10.134
Enthalpy	kW	-		-		19,351.310		15,817.285		10,281.182		-		-		11,566.290	
Phase		S		S		L		L		L		G		S		L	
Press.	Bar	1		1		1		1		1		1		1		1	
Temp	°C	30.0		30.0		80.0		43.8		43.8		30.0		30.0		45.0	

STREAM Nr.		<108>		<109>		<110>		<111>		<112>		<113>		<114>		<115>	
Name :		Overlimed Juice		Overlimed Juice		Overlimed Juice		Flash Vapor		Overlimed Juice		Clarifier Mud		Clarified Juice-Sugar		Raw Juice-Ethanol	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																
Ash	40.00																
Bagasse	257.84	0.303	0.001	0.303	0.001	0.303	0.001			0.303	0.001	0.303	0.001			0.163	0.001
Yeast	24.60																
$Ca(OH)_2$	74.09																
Ca-salts	136.14	0.231	0.002	0.231	0.002	0.231	0.002			0.231	0.002	0.231	0.002				
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01																
Dirt	40.00	2.162	0.054	2.162	0.054	2.162	0.054			2.162	0.054	2.162	0.054			1.164	0.029
Ethanol	46.06																
Glycerol	92.09																
Impurity	33.15	3.753	0.113	3.753	0.113	3.753	0.113			3.753	0.113	1.479	0.045	2.274	0.069	2.029	0.061
Isoamylalcohol	82.20																
N_2	28.01																
N_2O	46.05																
O_2	32.00																
Polyacrylamide	165.00									0.001	0.000	0.001	0.000				
Succinic Acid	118.10																
Sucrose	342.30	30.856	0.090	30.856	0.090	30.856	0.090			30.856	0.090	1.543	0.005	29.313	0.086	16.588	0.048
Sugar Cane	109.90																
SO_2	64.06	0.107	0.002	0.107	0.002	0.091	0.001	0.016	0.000	0.091	0.001			0.091	0.001		
H_2SO_4	98.08																
Sulphur	32.06																
Water	18.02	178.578	9.910	178.578	9.910	177.488	9.850	1.089	0.060	177.622	9.857	12.434	0.690	165.189	9.167	95.792	5.316
Total		215.989	10.172	215.989	10.172	214.883	10.111	1.105	0.061	215.018	10.119	18.151	0.796	196.868	9.323	115.735	5.455
Enthalpy	kW	11,741.837		59,021.141		56,523.117		2,497.300		56,523.117		4,170.265		52,352.715		5,536.018	
Phase		L		L		L		G		L		Slurry		L		L	
Press.	Bar	1		1		1		1		1		1		1		1	
Temp	°C	45.2		103.0		100.0		100.0		100.0		100.0		100.0		43.8	

STREAM Nr.		<116>		<117>		<118>		<119>		<120>		<121>		<122>		<123>	
Name :		Overlimed Juice		Overlimed Juice		Overlimed Juice		Flash Vapor		larified Juice-Ethan		Clarifier Mud		Clarifier Mud		Filter Cake	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																
Ash	40.00																
Bagasse	257.84	0.163	0.001	0.163	0.001	0.163	0.001					0.163	0.001	0.466	0.002	0.466	0.002
Yeast	24.60																
$Ca(OH)_2$	74.09																
Ca-salts	136.14	0.124	0.001	0.124	0.001	0.124	0.001					0.124	0.001	0.355	0.003	0.355	0.003
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01																
Dirt	40.00	1.164	0.029	1.164	0.029	1.164	0.029					1.164	0.029	3.326	0.083	3.326	0.083
Ethanol	46.06																
Glycerol	92.09																
Impurity	33.15	1.962	0.059	1.962	0.059	1.962	0.059			1.189	0.036	0.773	0.023	2.252	0.068	2.252	0.068
Isoamylalcohol	82.20																
N_2	28.01																
N_2O	46.05																
O_2	32.00																
Polyacrylamide	165.00													0.001	0.000	0.001	0.000
Succinic Acid	118.10																
Sucrose	342.30	16.615	0.049	16.615	0.049	16.615	0.049			15.784	0.046	0.831	0.002	2.374	0.007	0.475	0.001
Sugar Cane	109.90																
SO_2	64.06																
H_2SO_4	98.08																
Sulphur	32.06																
Water	18.02	96.167	5.337	96.167	5.337	95.576	5.304	0.591	0.033	88.886	4.933	6.690	0.371	19.124	1.061	8.417	0.467
Total		116.195	5.475	116.195	5.475	115.604	5.442	0.591	0.033	105.859	5.015	9.745	0.428	27.896	1.224	15.290	0.624
Enthalpy	kW	5,630.545		31,261.371		29,925.711		1,335.660		27,308.876		2,616.739		6,787.004		2,999.711	
Phase		L		L		L		G		L		Slurry		Slurry		Slurry	
Press.	Bar	1		1		1		1		1		1		1		1	
Temp	°C	44.0		103.0		100.0		100.0		100.0		100.0		100.0		93.0	

STREAM Nr.		<124>		<125>		<126>		<127>		<128>		<129>		<130>		<131>	
Name :		Filtrate		Filtrate		Filtrate		Air		Dry Bagasse		Limestone		F-101 Exhaust Gas		Calcium Oxide	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																
Ash	40.00																
Bagasse	257.84									0.051	0.000						
Yeast	24.60																
$Ca(OH)_2$	74.09																
Ca-salts	136.14																
$CaCO_3$	100.09											0.297	0.003				
CaO	56.08															0.166	0.003
CO_2	44.01													0.233	0.005		
Dirt	40.00																
Ethanol	46.06																
Glycerol	92.09																
Impurity	33.15									0.001	0.000	0.006	0.000				
Isoamylalcohol	82.20																
N_2	28.01							0.435	0.016					0.435	0.016		
N_2O	46.05													0.001	0.000		
O_2	32.00							0.116	0.004					0.038	0.001		
Polyacrylamide	165.00																
Succinic Acid	118.10																
Sucrose	342.30	1.899	0.006	1.822	0.005	0.077	0.000			0.003	0.000						
Sugar Cane	109.90																
SO_2	64.06																
H_2SO_4	98.08																
Sulphur	32.06																
Water	18.02	26.263	1.457	25.194	1.398	1.069	0.059			0.007	0.000			0.041	0.002		
Total		28.162	1.463	27.016	1.403	1.146	0.060	0.550	0.019	0.062	0.001	0.303	0.003	0.747	0.024	0.166	0.003
Enthalpy	kW	7,019.661		6,733.959		285.702		-		7.721		-		246.192		132.085	
Phase		L		L		L		G		S		S		G		S	
Press.	Bar	1		1		1		1		1		1		1		1	
Temp	°C	93.0		93.0		93.0		30.0		100.0		30.0		320.0		898.0	

STREAM Nr.		<132>		<133>		<134>		<135>		<136>		<137>		<138>		<139>	
Name :		Lime Milk		Lime Milk		Lime Milk		F-102 Exhaust Gas		Polyacrylamide		S-103 Wash Water		Bagasse		Dry Bagasse	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																
Ash	40.00																
Bagasse	257.84													46.097	0.179	46.097	0.179
Yeast	24.60																
$Ca(OH)_2$	74.09	0.220	0.003	0.077	0.001	0.143	0.002										
Ca-salts	136.14																
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01																
Dirt	40.00																
Ethanol	46.06																
Glycerol	92.09																
Impurity	33.15													0.855	0.026	0.855	0.026
Isoamylalcohol	82.20																
N_2	28.01							0.734	0.026								
N_2O	46.05																
O_2	32.00							0.087	0.003								
Polyacrylamide	165.00									0.001	0.000						
Succinic Acid	118.10																
Sucrose	342.30	0.077	0.000	0.027	0.000	0.050	0.000							2.494	0.007	2.494	0.007
Sugar Cane	109.90																
SO_2	64.06																
H_2SO_4	98.08																
Sulphur	32.06																
Water	18.02	1.015	0.056	0.355	0.020	0.660	0.037	0.018	0.001	0.134	0.007	15.556	0.863	45.594	2.530	6.707	0.372
Total		1.312	0.060	0.459	0.021	0.853	0.039	0.839	0.030	0.135	0.007	15.556	0.863	95.040	2.742	56.153	0.584
Enthalpy	kW	270.082		94.257		175.547		34.442		-		3,232.454		3,534.100		6,576.105	
Phase		L		L		L		G		L		L		S		S	
Press.	Bar	1		1		1		1		1		1		1		1	
Temp	°C	90.0		90.0		90.0		44.0		30.0		80.0		43.8		100.0	

STREAM Nr.		<140>		<141>		<142>		<143>		<144>		<145>		<146>		<147>	
Name :		Dry Bagasse		Air		F-103 Exhaust Gas		Wet Exhaust Gas		Ash		Ash		Exhasut Gas		Flash Vapor	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																
Ash	40.00									1.239	0.031	0.001	0.000				
Bagasse	257.84	46.046	0.179														
Yeast	24.60																
$Ca(OH)_2$	74.09																
Ca-salts	136.14																
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01					84.181	1.913	84.181	1.913					0.233	0.005		
Dirt	40.00																
Ethanol	46.06																
Glycerol	92.09																
Impurity	33.15	0.855	0.026														
Isoamylalcohol	82.20																
N_2	28.01			310.289	11.078	310.289	11.078	310.289	11.078					1.169	0.042		
N_2O	46.05					0.599	0.013	0.599	0.013					0.001	0.000		
O_2	32.00			82.482	2.578	18.630	0.582	18.630	0.582					0.126	0.004		
Polyacrylamide	165.00																
Succinic Acid	118.10																
Sucrose	342.30	2.492	0.007														
Sugar Cane	109.90																
SO_2	64.06					0.009	0.000	0.009	0.000							0.016	0.000
H_2SO_4	98.08																
Sulphur	32.06																
Water	18.02	6.700	0.372			33.916	1.882	72.803	4.040					0.059	0.003	1.680	0.093
Total		56.092	0.583	392.771	13.655	447.624	15.468	486.511	17.626	1.239	0.031	0.001	0.000	1.587	0.054	1.696	0.093
Enthalpy	kW	6,568.384		-		328,535.800		325,493.700		1.778		0.001		280.634		3,832.960	
Phase		S		G		G		G		S		S		G		G	
Press.	Bar	1		1		1		1		1		1		1		1	
Temp	°C	100.0		30.0		500.0		282.8		1,100.0		898.0		169.0		100.0	

STREAM Nr.		<201>		<202>		<203>		<204>		<205>		<206>		<207>		<208>	
Name :		Conjentrated Juice		Conjentrated Juice		E-202 Condensate		E-203 Condensate		E-204 Condensate		E-205 Condensate		E-206 Condensate		A Massecuites	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																
Ash	40.00																
Bagasse	257.84																
Yeast	24.60																
$Ca(OH)_2$	74.09																
Ca-salts	136.14																
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01																
Dirt	40.00																
Ethanol	46.06																
Glycerol	92.09																
Impurity	33.15	2.047	0.062	2.047	0.062	0.031	0.001	0.033	0.001	0.036	0.001	0.039	0.001	0.042	0.001	2.047	0.062
Isoamylalcohol	82.20																
N_2	28.01																
N_2O	46.05																
O_2	32.00																
Polyacrylamide	165.00																
Succinic Acid	118.10																
Sucrose	342.30	29.313	0.086	29.313	0.086											29.313	0.086
Sugar Cane	109.90																
SO_2	64.06					0.012	0.000	0.013	0.000	0.015	0.000	0.016	0.000	0.017	0.000		
H_2SO_4	98.08																
Sulphur	32.06																
Water	18.02	14.950	0.830	14.950	0.830	20.222	1.122	22.085	1.226	23.978	1.331	25.886	1.437	27.869	1.547	7.938	0.441
Total		46.310	0.977	46.310	0.977	20.265	1.123	22.132	1.227	24.029	1.332	25.941	1.438	27.929	1.548	39.298	0.588
Enthalpy	kW	2,669.438		4,625.596		7,973.968		8,112.807		8,027.220		7,582.362		6,485.117		3,251.960	
Phase		L		L		L		L		L		L		L		L	
Press.	Bar	0.17		1		2.3		1.9		1.5		1		0.6		0.4	
Temp	°C	56.3		75.0		125.1		118.6		110.7		100.5		86.0		75.0	

STREAM Nr.		<209>		<210>		<211>		<212>		<213>		<214>		<215>		<216>	
Name :		E-207 Vapor		A Massacuites		A Massecuites		Wet Sugar Crystals		A Molasses		A Molasses		B Massecuites		E-208 Vapor	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																
Ash	40.00																
Bagasse	257.84																
Yeast	24.60																
$Ca(OH)_2$	74.09																
Ca-salts	136.14																
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01																
Dirt	40.00																
Ethanol	46.06																
Glycerol	92.09																
Impurity	33.15			2.047	0.062	2.047	0.062	0.007	0.000	2.040	0.062	2.040	0.062	2.040	0.062		
Isoamylalcohol	82.20																
N_2	28.01																
N_2O	46.05																
O_2	32.00																
Polyacrylamide	165.00																
Succinic Acid	118.10																
Sucrose	342.30			29.313	0.086	29.313	0.086	17.588	0.051	11.725	0.034	11.725	0.034	11.725	0.034		
Sugar Cane	109.90																
SO_2	64.06																
H_2SO_4	98.08																
Sulphur	32.06																
Water	18.02	7.011	0.389	7.938	0.441	7.938	0.441	0.179	0.010	7.759	0.431	7.759	0.431	3.748	0.208	4.011	0.223
Total		7.011	0.389	39.298	0.588	39.298	0.588	17.775	0.062	21.524	0.526	21.524	0.526	17.512	0.304	4.011	0.223
Enthalpy	kW	17,691.805		690.037		1,761.929		554.917		1,206.992		2,187.089		1,479.457		9,991.678	
Phase		G		L&S		L&S		S		L		L		L		G	
Press.	Bar	0.38		1		1		1		1		1		0.4		0.4	
Temp	°C	75.0		40.0		55.0		55.0		55.0		75.0		75.0		75.0	

STREAM Nr.		<217>		<218>		<219>		<220>		<221>		<222>		<223>		<224>	
Name :		A Massecuites		A Massecuites		Wet Sugar Crystals		B Molasses		Wet Sugar Crystals		Sugar Crystals		Air		Air	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																
Ash	40.00																
Bagasse	257.84																
Yeast	24.60																
$Ca(OH)_2$	74.09																
Ca-salts	136.14																
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01																
Dirt	40.00																
Ethanol	46.06																
Glycerol	92.09																
Impurity	33.15	2.040	0.062	2.040	0.062	0.003	0.000	2.037	0.061	0.010	0.000	0.010	0.000				
Isoamylalcohol	82.20																
N_2	28.01													10.417	0.372	10.417	0.372
N_2O	46.05																
O_2	32.00													2.769	0.087	2.769	0.087
Polyacrylamide	165.00																
Succinic Acid	118.10																
Sucrose	342.30	11.725	0.034	11.725	0.034	7.035	0.021	4.690	0.014	24.623	0.072	24.623	0.072				
Sugar Cane	109.90																
SO_2	64.06																
H_2SO_4	98.08																
Sulphur	32.06																
Water	18.02	3.748	0.208	3.748	0.208	0.072	0.004	3.676	0.204	0.251	0.014	0.005	0.000				
Total		17.512	0.304	17.512	0.304	7.110	0.025	10.402	0.279	24.885	0.086	24.638	0.073	13.186	0.458	13.186	0.458
Enthalpy	kW	334.532		806.909		227.787		593.130		797.190		782.702		-		715.566	
Phase		L&S		L&S		S		L		S		S		G		G	
Press.	Bar	1		1		1		1		1		1		1		1	
Temp	°C	40.0		55.0		55.0		55.0		55.0		40.0		30.0		85.0	

STREAM Nr.		<225>		<226>		<227>		<228>		<301>		<302>		<303>		<304>	
Name :		Humid Air		Boiler Vapor		Condensates		E-206 Vapor		Juice, Filtrate& Molasses				CO_2			
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05													trace	trace	0.128	0.002
Ash	40.00																
Bagasse	257.84																
Yeast	24.60															14.147	0.575
$Ca(OH)_2$	74.09																
Ca-salts	136.14																
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01													8.449	0.192	2.368	0.054
Dirt	40.00																
Ethanol	46.06													0.134	0.003	10.898	0.237
Glycerol	92.09															0.523	0.006
Impurity	33.15					0.182	0.005	0.046	0.001	3.226	0.097	3.226	0.097			3.087	0.093
Isoamylalcohol	82.20													trace	trace	0.091	0.001
N_2	28.01	10.417	0.372														
N_2O	46.05																
O_2	32.00	2.769	0.087														
Polyacrylamide	165.00																
Succinic Acid	118.10															0.052	0.000
Sucrose	342.30									22.296	0.065	22.296	0.065				
Sugar Cane	109.90																
SO_2	64.06					0.073	0.001	0.018	0.000								
H_2SO_4	98.08																
Sulphur	32.06																
Water	18.02	0.246	0.014	11.023	0.612	120.041	6.662	30.198	1.676	117.755	6.535	117.755	6.535	0.147	0.008	170.475	9.460
Total		13.433	0.472	11.023	0.612	120.296	6.668	30.262	1.677	143.277	6.697	143.277	6.697	8.730	0.203	201.769	10.428
Enthalpy	kW	308.613		27,683.483		38,181.474		74,978.637		34,635.965		-		-		-	
Phase		G		G		L		G		L		L		G		L	
Press.	Bar	1		0.38		1.3		0.17		1		1		1		1	
Temp	°C	40.0		75.0		106.0		56.3		96.0		30.0		30.0		30.0	

STREAM Nr.		<305>		<306>		<307>		<308>		<309>		<310>		<311>		<312>	
Name :		Ethanol Solution		Scrubber Ethanol		Ethanol Solution		Yeast		Beer Column Feed		Beer Colm. Product		Vinasse		Berr Colm. Distilate	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05	0.128	0.002	0.001	0.000	0.129	0.002			0.129	0.002	0.021	0.000	0.012	0.000	trace	trace
Ash	40.00																
Bagasse	257.84																
Yeast	24.60							14.147	0.575								
$Ca(OH)_2$	74.09																
Ca-salts	136.14																
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01	2.368	0.054	0.152	0.003	2.520	0.057			2.520	0.057	0.002	0.000			2.518	0.057
Dirt	40.00																
Ethanol	46.06	10.898	0.237	1.395	0.030	12.293	0.267			12.293	0.267	11.026	0.239	trace	trace	1.263	0.027
Glycerol	92.09	0.523	0.006			0.523	0.006			0.523	0.006			0.172	0.002		
Impurity	33.15	3.087	0.093			3.087	0.093			3.087	0.093			3.087	0.093		
Isoamylalcohol	82.20	0.091	0.001	trace	trace	0.091	0.001			0.091	0.001	0.014	0.000	0.070	0.001	trace	trace
N_2	28.01																
N_2O	46.05																
O_2	32.00																
Polyacrylamide	165.00																
Succinic Acid	118.10	0.052	0.000			0.052	0.000			0.052	0.000			0.017	0.000		
Sucrose	342.30																
Sugar Cane	109.90																
SO_2	64.06																
H_2SO_4	98.08																
Sulphur	32.06																
Water	18.02	113.881	6.320	13.871	0.770	127.752	7.089	56.594	3.141	127.752	7.089	16.992	0.943	13.642	0.757	0.450	0.025
Total		131.028	6.713	15.419	0.804	146.447	7.516	70.741	3.716	146.447	7.516	28.055	1.183	17.000	0.853	4.231	0.110
Enthalpy	kW	-		1,444.504		1,444.504		-		30,675.581		54,317.164		4,292.795		2,635.803	
Phase		L		L		L		Slurry		L		G		L		G	
Press.	Bar	1		1		1		1		1		1		1		1	
Temp	°C	30.0		53.3		32.2		30.0		80.0		92.6		100.0		20.0	

STREAM Nr.		<313>		<314>		<315>		<316>		<317>		<318>		<319>		<320>	
Name :		Fusel Oil		Rectification Prod.		Molec. Sieve Feed		Ethanol		Ethanol		Fusel Oil		Fusel Oil		Wet Nitrogen	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05	0.096	0.002	trace	trace	trace	trace	trace	trace	trace	trace	0.021	0.000	0.117	0.002		
Ash	40.00																
Bagasse	257.84																
Yeast	24.60																
$Ca(OH)_2$	74.09																
Ca-salts	136.14																
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01			0.002	0.000	0.002	0.000	0.002	0.000	0.002	0.000						
Dirt	40.00																
Ethanol	46.06	0.004	0.000	11.025	0.239	11.025	0.239	11.024	0.239	11.024	0.239	0.001	0.000	0.005	0.000	0.001	0.000
Glycerol	92.09	0.351	0.004									trace	trace	0.351	0.004		
Impurity	33.15																
Isoamylalcohol	82.20	0.007	0.000									0.014	0.000	0.021	0.000		
N_2	28.01															12.000	0.428
N_2O	46.05																
O_2	32.00																
Polyacrylamide	165.00																
Succinic Acid	118.10	0.035	0.000									trace	trace	0.035	0.000		
Sucrose	342.30																
Sugar Cane	109.90																
SO_2	64.06																
H_2SO_4	98.08																
Sulphur	32.06																
Water	18.02	96.668	5.364	0.892	0.050	0.892	0.050	0.013	0.001	0.013	0.001	16.100	0.893	112.768	6.258	0.939	0.052
Total		97.161	5.370	11.919	0.289	11.919	0.289	11.039	0.240	11.039	0.240	16.136	0.894	113.297	6.264	12.940	0.481
Enthalpy	kW	28,189.839		1,758.562		376.035		266.256		-		3,177.372		31,367.211		7,888.184	
Phase		L		L		L		L		L		L		L		G	
Press.	Bar	1		1		1		1		1		1		1		1	
Temp	°C	100.0		77.4		60.0		60.0		30.0		100.0		100.0		200.0	

STREAM Nr.		<321>		<322>		<323>		<324>		<325>		<326>		<327>		<328>	
Name :		Wet Nitrogen		Dry nitrogen		Nitrogen Make up		Dry nitrogen		Dry nitrogen		Scrubber Water		CO_2		Yeast Bleed	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																
Ash	40.00																
Bagasse	257.84																
Yeast	24.60															0.703	0.029
$Ca(OH)_2$	74.09																
Ca-salts	136.14																
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01													10.814	0.246		
Dirt	40.00																
Ethanol	46.06	0.001	0.000											0.002	0.000		
Glycerol	92.09																
Impurity	33.15																
Isoamylalcohol	82.20																
N_2	28.01	12.000	0.428	11.988	0.428	0.012	0.000	12.000	0.428	12.000	0.428						
N_2O	46.05																
O_2	32.00																
Polyacrylamide	165.00																
Succinic Acid	118.10																
Sucrose	342.30																
Sugar Cane	109.90																
SO_2	64.06																
H_2SO_4	98.08																
Sulphur	32.06																
Water	18.02	0.939	0.052	0.060	0.003			0.060	0.003	0.060	0.003	13.773	0.764	0.498	0.028	2.816	0.156
Total		12.940	0.481	12.048	0.431	0.012	0.000	12.060	0.432	12.060	0.432	13.773	0.764	11.314	0.273	3.519	0.185
Enthalpy	kW	-		-822.456		-		-823.273		6,265.314		-		1,313.753		-	
Phase		G		G		G		G		G		L		G		Slurry	
Press.	Bar	1		1		1		1		1		1		1		1	
Temp	°C	30.0		5.0		30.0		5.0		180.0		30.0		46.3		30.0	

STREAM Nr.		<329>		<330>		<331>		<332>		<333>		<334>					
Name :		Recyle Yeast		Acid Sln.		Acid & Yeast		Waste Acid Sln.		Recycle Yeast		Waste Water					
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																
Ash	40.00																
Bagasse	257.84																
Yeast	24.60	13.444	0.547			13.444	0.547			13.444	0.547						
$Ca(OH)_2$	74.09																
Ca-salts	136.14																
$CaCO_3$	100.09																
CaO	56.08																
CO_2	44.01																
Dirt	40.00																
Ethanol	46.06											0.001	0.000				
Glycerol	92.09																
Impurity	33.15																
Isoamylalcohol	82.20																
N_2	28.01											0.012	0.000				
N_2O	46.05																
O_2	32.00																
Polyacrylamide	165.00																
Succinic Acid	118.10																
Sucrose	342.30																
Sugar Cane	109.90																
SO_2	64.06																
H_2SO_4	98.08			0.142	0.001	0.142	0.001	0.142	0.001								
Sulphur	32.06																
Water	18.02	53.778	2.984	144.565	8.022	198.343	11.007	144.565	8.022	53.778	2.984	0.879	0.049				
Total		67.222	3.531	144.707	8.024	211.929	11.555	144.707	8.024	67.222	3.531	0.892	0.049				
Enthalpy	kW	-		-		-		-		6,265.314		-92.218					
Phase		Slurry		L		L		L		-		L					
Press.	Bar	1		1		1		1		1		1		1		1	
Temp	°C	30.0		30.0		30.0		30.0		30.0		5.0					

A.4.2. Process Stream Summary of 2015 Plant

STREAM Nr.		<101>		<101>		<102)		<103>		<104>		<105>		<106>		<107>		<108>		<109>	
Name :		Sugar Cane		Sugar Cane		Imbibition Water		Raw Juice		Raw Juice		Air		Sulphur		Sulphitated Juice		Overlimed Juice		Overlimed Juice	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																				
Ash	40.00																				
Bagasse	257.84			46.562	0.181			0.466	0.002	0.303	0.001					0.303	0.001	0.303	0.001	0.303	0.001
Yeast	24.60																				
$Ca(OH)_2$	74.09																				
Ca-salts	136.14																	0.231	0.002	0.231	0.002
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01																				
Cellulase	500.00																				
cellulose	324.00																				
Dirt	40.00			3.326	0.083			3.326	0.083	2.162	0.054					2.162	0.054	2.162	0.054	2.162	0.054
Ethanol	46.06																				
Furfural	96.08																				
Glycerol	92.09																				
Hemicellulose	264.00																				
Hexosan	324.00																				
Hexoses	180.09																				
HMF	126.00																				
Impurity	33.15			6.652	0.201			5.796	0.175	3.768	0.114					3.768	0.114	3.757	0.113	3.757	0.113
Isoamylalcohol	82.20																				
Lactic Acid	90.04																				
Lignin	174.40																				
Lignin Monomere	90.90																				
Methance	16.04																				
N_2	28.01											0.734	0.026								
N_2O	46.05																				
O_2	32.00											0.195	0.006								
Pentosan	264.00																				
Pentoses	150.06																				
Polyacrylamide	165.00																				
Succinic Acid	118.10																				
Sucrose	342.30			49.888	0.146			47.394	0.138	30.806	0.090					30.806	0.090	30.857	0.090	30.857	0.090
Sugar Cane	109.90	332.589	3.026																		
SO_2	64.06															0.213	0.003	0.104	0.002	0.104	0.002
H_2SO_4	98.08																				
Sulphur	32.06													0.108	0.003						
Water	18.02			226.160	12.551			273.691	15.188	177.899	9.872					177.881	9.871	178.583	9.910	178.583	9.910
Xylitol	152.00					93.125	0.613														
Total		332.589	3.026	332.589	13.161	93.125	0.613	330.673	15.586	214.937	10.131	0.930	0.032	0.108	0.003	215.132	10.134	215.995	10.172	215.995	10.172
Enthalpy	kW	-		-		19,351.310		15,817.285		10,281.182		-		-		11,506.290		11,743.193		58,792.743	
Phase		S		S		L		L		L		G		S		L		L		L	
Press.	Bar	1		1		1		1		1		1		1		1		1		1	
Temp	°C	30.0		30.0		80.0		43.8		43.8		30.0		30.0		45.0		45.2		103.0	

STREAM Nr.		<110>		<111>		<112>		<113>		<114>		<115>		<116>		<117>		<118>		<119>	
Name :		Overlimed Juice		Flash Vapor		Overlimed Juice		Clarifier Mud		Clarified Juice		Raw Juice		Overlimed Juice		Overlimed Juice		Overlimed Juice		Flash Vapor	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																				
Ash	40.00																				
Bagasse	257.84	0.303	0.001			0.303	0.001	0.303	0.001			0.163	0.001	0.163	0.001	0.163	0.001	0.163	0.001		
Yeast	24.60																				
$Ca(OH)_2$	74.09																				
Ca-salts	136.14	0.231	0.002			0.231	0.002	0.231	0.002					0.124	0.001	0.124	0.001	0.124	0.001		
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01																				
Cellulase	500.00																				
cellulose	324.00																				
Dirt	40.00	2.162	0.054			2.162	0.054	2.162	0.054			1.164	0.029	1.164	0.029	1.164	0.029	1.164	0.029		
Ethanol	46.06																				
Furfural	96.08																				
Glycerol	92.09																				
Hemicellulose	264.00																				
Hexosan	324.00																				
Hexoses	180.09																				
HMF	126.00																				
Impurity	33.15	3.757	0.113			3.757	0.113	1.480	0.045	2.276	0.069	2.029	0.061	1.964	0.059			1.964	0.059		
Isoamylalcohol	82.20															1.964	0.024				
Lactic Acid	90.04																				
Lignin	174.40																				
Lignin Monomere	90.90																				
Methance	16.04																				
N_2	28.01																				
N_2O	46.05																				
O_2	32.00																				
Pentosan	264.00																				
Pentoses	150.06																				
Polyacrylamide	165.00					0.001	0.000	0.001	0.000												
Succinic Acid	118.10																				
Sucrose	342.30	30.857	0.090			30.857	0.090	1.543	0.005	29.314	0.086	16.588	0.048	16.615	0.049			16.615	0.049		
Sugar Cane	109.90															16.615	0.151				
SO_2	64.06	0.104	0.002			0.104	0.002			0.104	0.002										
H_2SO_4	98.08																				
Sulphur	32.06																				
Water	18.02	177.547		1.036	0.057	177.681	9.860	12.438	0.690	165.244	9.170	95.792	5.316	96.170	5.337	96.170	5.337	95.612	5.306	0.558	0.031
Xylitol	152.00																				
Total		214.959	0.262	1.036	0.057	215.094	10.122	18.156	0.796	196.938	9.326	115.735	5.455	116.200	5.475	116.200	5.543	115.642	5.444	0.558	0.031
Enthalpy	kW	56,456.817		2,341.360		56,456.817						5,536.747		5,635.987		31,266.813		1,261.080		30,005.342	
Phase		L		G		L		Slurry		L		L		L		L		L		G	
Press.	Bar	1		1		1		1		1		1		1		1		1		1	
Temp	°C	100.0		100.0		100.0		100.0		100.0		43.8		44.0		103.0		100.0		100.0	

STREAM Nr.		<120>		<121>		<122>		<123>		<124>		<125>		<126>		<127>		<128>		<129>	
Name :		Clarified Juice		Clarifier Mud		Clarifier Mud		Filter Cake		Filtrate		Filtrate		Filtrate		Air		Natural Gas		Limestone	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																				
Ash	40.00																				
Bagasse	257.84			0.163	0.001	0.466	0.002	0.466	0.002												
Yeast	24.60																				
$Ca(OH)_2$	74.09																				
Ca-salts	136.14			0.124	0.001	0.355	0.003	0.355	0.003												
$CaCO_3$	100.09																			0.907	0.009
CaO	56.08																				
CO_2	44.01																				
Cellulase	500.00																				
cellulose	324.00																				
Dirt	40.00			1.164	0.029	3.326	0.083	3.326	0.083												
Ethanol	46.06																				
Furfural	96.08																				
Glycerol	92.09																				
Hemicellulose	264.00																				
Hexosan	324.00																				
Hexoses	180.09																				
HMF	126.00																				
Impurity	33.15			0.774	0.023	2.254	0.068	2.254	0.068											0.019	0.001
Isoamylalcohol	82.20	1.190	0.014																		
Lactic Acid	90.04																				
Lignin	174.40																				
Lignin Monomere	90.90																				
Methance	16.04																	0.052	0.003		
N_2	28.01															1.180	0.042				
N_2O	46.05																				
O_2	32.00															0.314	0.010				
Pentosan	264.00																				
Pentoses	150.06																				
Polyacrylamide	165.00					0.001	0.000	0.001	0.000												
Succinic Acid	118.10																				
Sucrose	342.30			0.831	0.002	2.374	0.007	0.475	0.001	1.899	0.006	1.661	0.005	0.238	0.001						
Sugar Cane	109.90	15.784	0.144																		
SO_2	64.06																				
H_2SO_4	98.08																				
Sulphur	32.06																				
Water	18.02	88.919	4.934	6.693	0.371	19.131	1.062	8.418	0.467	26.268	1.458	22.979	1.275	3.289	0.183						
Xylitol	152.00																				
Total		105.894	5.093	9.749	0.428	27.905	1.224	15.294	0.624	28.167	1.463	24.640	1.280	3.526	0.183	1.494	0.052	0.052	0.003	0.925	0.010
Enthalpy	kW	27,389.177		2,616.739		6,332.046		2,543.688		7,020.936		6,141.921		879.015		-		-		-	
Phase		L		Slurry		Slurry		S		L		L		L		L		G		S	
Press.	Bar	1		1		1		1		1		1		1		1		1		1	
Temp	°C	100.0		100.0		100.0		93.0		93.0		93.0		93.0		30.0		30.0		30.0	

STREAM Nr.		<130>		<131>		<132>		<133>		<134>		<135>		<136>		<137>		<138>		<139>	
Name :		F-101 Exhaust Gas		Calcium Oxide		Lime Milk		Lime Milk		Lime Milk		F-102 Exhaust Gas		Polyacrylamide		Filter Wash Water		Wet Bagasse		Exhaust Gas	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																				
Ash	40.00																				
Bagasse	257.84																	46.097	0.179		
Yeast	24.60																				
$Ca(OH)_2$	74.09					0.671	0.009	0.077	0.001	0.143	0.002										
Ca-salts	136.14																				
$CaCO_3$	100.09																				
CaO	56.08			0.508	0.009																
CO_2	44.01	0.542	0.012																	0.542	0.012
Cellulase	500.00																				
cellulose	324.00																				
Dirt	40.00																				
Ethanol	46.06																				
Furfural	96.08																				
Glycerol	92.09																				
Hemicellulose	264.00																				
Hexosan	324.00																				
Hexoses	180.09																				
HMF	126.00																				
Impurity	33.15			0.019	0.001	0.019	0.001	0.002	0.000	0.004	0.000							0.855	0.026		
Isoamylalcohol	82.20																				
Lactic Acid	90.04																				
Lignin	174.40																				
Lignin Monomere	90.90																				
Methance	16.04																				
N_2	28.01	1.180	0.042									0.734	0.026							1.915	0.068
N_2O	46.05																				
O_2	32.00	0.106	0.003									0.087	0.003							0.193	0.006
Pentosan	264.00																				
Pentoses	150.06																				
Polyacrylamide	165.00													0.001	0.000						
Succinic Acid	118.10																				
Sucrose	342.30					0.238	0.001	0.027	0.000	0.051	0.000							2.494	0.007		
Sugar Cane	109.90																				
SO_2	64.06											0.004	0.000							0.004	0.000
H_2SO_4	98.08																				
Sulphur	32.06																				
Water	18.02	0.117	0.006			3.126	0.173	0.358	0.020	0.665	0.037	0.018	0.001	0.134	0.007	15.556	0.863	45.594	2.530	0.135	0.007
Xylitol	152.00																				
Total		1.945	0.064	0.526	0.010	4.053	0.184	0.464	0.021	0.863	0.039	0.843	0.030	0.135	0.007	15.556	0.863	95.040	2.742	2.788	0.094
Enthalpy	kW	1,017.022		402.243		854.046		99.240		180.923		34.442		-		3,232.454		3,534.100		1,051.464	
Phase		G		S		Lime Milk		L		L		G		L		L		S		G	
Press.	Bar	1		1		1		1		1		1		1		1		1		1	
Temp	°C	320.0		898.0		90.0		90.0		90.0		44.0		30.0		80.0		43.8		169.0	

STREAM Nr.		<140>		<141>		<201>		<202>		<203>		<204>		<205>		<206>		<207>		<208>	
Name :		Flash Vapor		Lime milk		Concentrated Juice		Concentrated Juice		E-202 Condensate		E-203 Condensate		E-204 Condensate		E-205 Condensate		E-206 Condensate		A Massecuites	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																				
Ash	40.00																				
Bagasse	257.84																				
Yeast	24.60																				
$Ca(OH)_2$	74.09			0.451	0.006																
Ca-salts	136.14																				
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01																				
Cellulase	500.00																				
cellulose	324.00																				
Dirt	40.00																				
Ethanol	46.06																				
Furfural	96.08																				
Glycerol	92.09																				
Hemicellulose	264.00																				
Hexosan	324.00																				
Hexoses	180.09																				
HMF	126.00							2.049	0.016			0.034	0.000	0.036	0.000	0.039	0.000	0.042	0.000	2.049	0.016
Impurity	33.15			0.012	0.000	2.049	0.062			0.031	0.001										
Isoamylalcohol	82.20																				
Lactic Acid	90.04																				
Lignin	174.40																				
Lignin Monomere	90.90																				
Methance	16.04																				
N_2	28.01																				
N_2O	46.05																				
O_2	32.00																				
Pentosan	264.00																				
Pentoses	150.06																				
Polyacrylamide	165.00																				
Succinic Acid	118.10																				
Sucrose	342.30			0.160	0.000			29.314	0.086											29.314	0.086
Sugar Cane	109.90					29.314	0.267														
SO_2	64.06									0.014	0.000	0.015	0.000	0.017	0.000	0.018	0.000	0.019	0.000		
H_2SO_4	98.08																				
Sulphur	32.06																				
Water	18.02	1.594	0.088	2.102	0.117	14.955	0.830	14.955	0.830	20.229	1.123	22.093	1.226	23.986	1.331	25.895	1.437	27.879	1.547	7.941	0.441
Xylitol	152.00																				
Total		1.594	0.088	2.725	0.124	46.317	1.158	46.317	0.932	20.274	1.124	22.141	1.227	24.039	1.332	25.952	1.438	27.940	1.548	39.304	0.543
Enthalpy	kW	3,602.440		573.883		2,602.598		4,559.340		7,962.234		8,113.791		8,028.356		7,577.314		8,157.815		3,137.183	
Phase		G		L		L		L		L		L		L		L		L		L	
Press.	Bar	1		1		0.17		1		2.3		1.9		1.5		1		0.56		0.38	
Temp	°C	100.0		90.0		56.3		75.0		125.0		118.6		110.7		100.5		86.0		75.0	

STREAM Nr.		<209>		<210>		<211>		<212>		<213>		<214>		<215>		<216>		<217>		<218>	
Name :		E-207 Vapor		A Massecuite		A Massecuites		Wet Sugar Crystals		A Molasses		A Molasses		B Massecuites		E-208 Vapor		B Massecuite		B Massecuite	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																				
Ash	40.00																				
Bagasse	257.84																				
Yeast	24.60																				
$Ca(OH)_2$	74.09																				
Ca-salts	136.14																				
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01																				
Cellulase	500.00																				
cellulose	324.00																				
Dirt	40.00																				
Ethanol	46.06																				
Furfural	96.08																				
Glycerol	92.09																				
Hemicellulose	264.00																				
Hexosan	324.00																				
Hexoses	180.09																				
HMF	126.00			2.049	0.016	2.049	0.016	0.007	0.000	2.042	0.016	2.042	0.016	2.042	0.016			2.042	0.016	2.042	0.016
Impurity	33.15																				
Isoamylalcohol	82.20																				
Lactic Acid	90.04																				
Lignin	174.40																				
Lignin Monomere	90.90																				
Methance	16.04																				
N_2	28.01																				
N_2O	46.05																				
O_2	32.00																				
Pentosan	264.00																				
Pentoses	150.06																				
Polyacrylamide	165.00																				
Succinic Acid	118.10																				
Sucrose	342.30			29.314	0.086	29.314	0.086	17.588	0.051	11.726	0.034	11.726	0.034	11.726	0.034			11.726	0.034	11.726	0.034
Sugar Cane	109.90																				
SO_2	64.06																				
H_2SO_4	98.08																				
Sulphur	32.06																				
Water	18.02	7.014	0.389	7.941	0.441	7.941	0.441	0.180	0.010	7.761	0.431	7.761	0.431	3.749	0.208	4.013	0.223	3.749	0.208	3.749	0.208
Xylitol	152.00																				
Total		7.014	0.389	39.304	0.543	39.304	0.543	17.775	0.061	21.529	0.481	21.529	0.481	17.516	0.258	4.013	0.223	17.516	0.258	17.516	0.258
Enthalpy	kW	17,697.608		683.037		1,754.929		554.915		1,199.873		2,171.089		1,361.968		10,095.968		217.053		695.211	
Phase		G		S&L		S&L		S		L		L		L		G		S&L		S&L	
Press.	Bar	0.38		1		1		1		1		1		0.38		1		1		1	
Temp	°C	75.0		40.0		55.0		55.0		55.0		75.0		75.0		75.0		40.0		55.0	

STREAM Nr.		<219>		<220>		<221>		<222>		<223>		<224>		<225>		<226>		<227>		<228>	
Name :		Wet Sugar Crystals		B Molasses		Wet Sugar Crystals		Dry Sugar		Air		Air		Wet Air		Boiler Vapor		Condensates			
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																				
Ash	40.00																				
Bagasse	257.84																				
Yeast	24.60																				
$Ca(OH)_2$	74.09																				
Ca-salts	136.14																				
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01																				
Cellulase	500.00																				
cellulose	324.00																				
Dirt	40.00																				
Ethanol	46.06																				
Furfural	96.08																				
Glycerol	92.09																				
Hemicellulose	264.00																				
Hexosan	324.00																				
Hexoses	180.09																				
HMF	126.00	0.003		2.039	0.016	0.010	0.000	0.010	0.000									0.182	0.001	0.046	0.000
Impurity	33.15																				
Isoamylalcohol	82.20																				
Lactic Acid	90.04																				
Lignin	174.40																				
Lignin Monomere	90.90																				
Methance	16.04																				
N_2	28.01									10.421	0.372	10.421	0.372	10.421	0.372						
N_2O	46.05																				
O_2	32.00									2.770	0.087	2.770	0.087	2.770	0.087						
Pentosan	264.00																				
Pentoses	150.06																				
Polyacrylamide	165.00																				
Succinic Acid	118.10																				
Sucrose	342.30	7.035	0.021	4.690	0.014	24.624	0.072	24.624	0.072												
Sugar Cane	109.90																				
SO_2	64.06																	0.083	0.001	0.021	0.000
H_2SO_4	98.08																				
Sulphur	32.06																				
Water	18.02	0.072	0.004	3.677	0.204	0.251	0.014	0.005	0.000					0.247	0.014	11.026	0.612	120.081	6.664	30.208	1.676
Xylitol	152.00																				
Total		7.110	0.025	10.406	0.234	24.885	0.086	24.639	0.072	13.191	0.459	13.191	0.459	13.437	0.472	11.026	0.612	120.346	6.666	30.275	1.677
Enthalpy	kW	197.887		497.512		752.702		308.492		-		602.698		1,046.907		27,793.576		39,839.510		74,184.447	
Phase		S		L		S		S		G		G		G		G		L		G	
Press.	Bar	1		1		1		1		1		1		1		1		1.3		0.17	
Temp	°C	55.0		55.0		55.0		40.0		30.0		85.0		40.0		75.0		106.7		56.3	

STREAM Nr.		<301>		<302>		<303>		<304>		<305>		<306>		<307>		<308>		<309>		<310>	
Name :		Fermentor Feed		Fermentor Feed		CO_2		Fermentation Broth		Yeast		Ethanol Solution		Scrubber Ethanol Sln.		Beer Colm. Feed		Beer Colm. Feed		Berr Colm. Prod.	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05	0.833	0.014	0.833	0.014	0.002	0.000	1.102	0.018			1.102	0.018	0.005	0.000	1.107	0.018	1.107	0.018	0.211	0.004
Ash	40.00	1.602	0.040	1.602	0.040			1.602	0.040			1.602	0.040			1.602	0.040	1.602	0.040		
Bagasse	257.84																				
Yeast	24.60							28.650	1.165	28.650	1.165										
$Ca(OH)_2$	74.09																				
Ca-salts	136.14																				
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01					21.085	0.479	4.732	0.108			4.732	0.108	0.352	0.008	5.084	0.116	5.084	0.116	0.005	0.000
Cellulase	500.00	0.002		0.002	0.000			0.002	0.000			0.002	0.000			0.002	0.000	0.002	0.000		
cellulose	324.00	0.009		0.009	0.000			0.009	0.000			0.009	0.000			0.009	0.000	0.009	0.000		
Dirt	40.00																				
Ethanol	46.06					0.393	0.009	25.991	0.564			25.991	0.564	2.980	0.065	28.971	0.629	28.971	0.629	26.376	0.573
Furfural	96.08	0.102		0.102	0.001	trace	trace	0.102	0.001			0.102	0.001	0.002	0.000	0.104	0.001	0.104	0.001	0.102	0.001
Glycerol	92.09							1.252	0.014			1.252	0.014			1.252	0.014	1.252	0.014	trace	trace
Hemicellulose	264.00	0.007		0.007	0.000			0.007	0.000			0.007	0.000			0.007	0.000	0.007	0.000		
Hexosan	324.00																				
Hexoses	180.09	19.617		19.617	0.109																
HMF	126.00	0.019		0.019	0.000	trace	trace	0.019	0.000			0.019	0.000	trace	trace	0.019	0.000	0.019	0.000	0.019	0.000
Impurity	33.15	5.055	0.152	5.055	0.152			4.817	0.145			4.817	0.145			4.817	0.145	4.817	0.145		
Isoamylalcohol	82.20					trace	trace	0.178	0.002			0.178	0.002	trace	trace	0.178	0.002	0.178	0.002	0.031	0.000
Lactic Acid	90.04							0.074	0.001			0.074	0.001	trace	trace	0.074	0.001	0.074	0.001	trace	trace
Lignin	174.40	0.055		0.055	0.000			0.055	0.000			0.055	0.000			0.055	0.000	0.055	0.000		
Lignin Monomere	90.90	0.383		0.383	0.004			0.383	0.004			0.383	0.004			0.383	0.004	0.383	0.004		
Methance	16.04																				
N_2	28.01																				
N_2O	46.05																				
O_2	32.00																				
Pentosan	264.00																				
Pentoses	150.06	10.373		10.373	0.069																
Polyacrylamide	165.00																				
Succinic Acid	118.10							0.130	0.001			0.130	0.001			0.130	0.001	0.130	0.001	trace	trace
Sucrose	342.30	24.790	0.072	24.790	0.072																
Sugar Cane	109.90																				
SO_2	64.06																				
H_2SO_4	98.08	0.0002	0.0000	0.0002	0.0000			0.0002	0.0000			0.0002	0.0000			0.0002	0.0000	0.0002	0.0000		
Sulphur	32.06																				
Water	18.02	228.640	12.688	228.640	12.688	0.365	0.020	335.373	18.611	114.533	6.356	220.840	12.255	30.527	1.694	251.367	13.949	251.367	13.949	38.535	2.138
Xylitol	152.00							0.016	0.000			0.016	0.000			0.016	0.000	0.016	0.000		
Total		291.488	12.967	291.488	13.151	21.845	0.508	404.494	20.675	143.183	7.521	261.311	13.154	33.866	1.767	295.177	14.921	295.177	14.921	65.279	2.716
Enthalpy	kW	46,997.943		-		1,381.589		-		-		-		2,928.190		2,928.190		58,151.924		128,755.983	
Phase		L		L		G		L		Slurry		L		L		L		L		G	
Press.	Bar	1		1		1		1		1		1		1		1		1		1	
Temp	°C	76.8		30.0		30.0		30.0		30.0		30.0		51.5		32.3		80.0		92.1	

STREAM Nr.		<311>		<312>		<313>		<314>		<315>		<316>		<317>		<318>		<319>		<320>	
Name :		Vinasse		Beer Colm. Dist.		Fusel Oil		Rectification Dist.		Molec Sieve Feed		Ethanol		Ethanol		Fusel Oil		Fusel oil		Wet Nitrogen	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05	0.038	0.001	0.003	0.000	0.855	0.014	trace	trace	trace	trace	trace	trace	trace	trace	0.211	0.004	1.066	0.018		
Ash	40.00	1.602	0.040																		
Bagasse	257.84																				
Yeast	24.60																				
$Ca(OH)_2$	74.09																				
Ca-salts	136.14																				
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01			5.079	0.115			0.005	0.000	0.005	0.000	0.005	0.000	0.005	0.000						
Cellulase	500.00	0.002	0.000																		
cellulose	324.00	0.009	0.000																		
Dirt	40.00																				
Ethanol	46.06	trace	trace	2.590	0.056	0.004	0.000	26.373	0.573	26.373	0.573	26.370	0.573	26.370	0.573	0.003	0.000	0.007	0.000	0.003	0.000
Furfural	96.08	trace	trace	0.002	0.000	trace	trace	trace	trace	trace	trace	trace	trace	trace	trace	0.102	0.001	0.102	0.001		
Glycerol	92.09	0.370	0.004			0.882	0.010									trace	trace	0.882	0.010		
Hemicellulose	264.00	0.007	0.000																		
Hexosan	324.00																				
Hexoses	180.09																				
HMF	126.00	trace	trace	trace	trace	trace	trace	trace	trace	trace	trace	trace	trace	trace	trace	0.019	0.000	0.019	0.000		
Impurity	33.15	4.817	0.145																		
Isoamylalcohol	82.20	0.006	0.000	trace	trace	0.141	0.002									0.031	0.000	0.172	0.002		
Lactic Acid	90.04	0.021	0.000	trace	trace	0.054	0.001									trace	trace	0.054	0.001		
Lignin	174.40	0.055	0.000																		
Lignin Monomere	90.90	0.383	0.004																		
Methance	16.04																				
N_2	28.01																			23.000	0.821
N_2O	46.05																				
O_2	32.00																				
Pentosan	264.00																				
Pentoses	150.06																				
Polyacrylamide	165.00																				
Succinic Acid	118.10	0.038	0.000			0.092	0.001									trace	trace	0.092	0.001		
Sucrose	342.30																				
Sugar Cane	109.90																				
SO_2	64.06																				
H_2SO_4	98.08	0.000	0.000																		
Sulphur	32.06																				
Water	18.02	11.199	0.621	0.888	0.049	200.745	11.140	2.132	0.118	2.132	0.118	0.024	0.001	0.024	0.001	36.402	2.020	237.147	13.160	2.224	0.123
Xylitol	152.00	0.016	0.000																		
Total		18.563	0.817	8.562	0.221	202.773	11.167	28.510	0.691	28.510	0.691	26.399	0.574	26.399	0.574	36.768	2.025	239.541	13.192	25.227	0.945
Enthalpy	kW	4,548.984		4,571.397		58,843.386		4,215.140		2,604.238		2,340.767		-		10,654.800		69,145.909		5,888.845	
Phase		L		G		L		L		L		L		L		L		L		G	
Press.	Bar	1		1		1		1		1		1		1		1		1		1	
Temp	°C	100.1		20.0		100.0		77.4		60.0		60.0		30.0		100.0		100.0		200.0	

STREAM Nr.		<321>		<322>		<323>		<324>		<325>		<326>		<327>		<328>		<329>		<330>	
Name :		Wet Nitrogen		Dry Nitrogen		Makeup Nitrogen		Dry nitrogen		Dry nitrogen		Absorption Water		CO_2		Yeast Bleed		Recycle Yeast		Acid Solution	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																				
Ash	40.00																				
Bagasse	257.84																				
Yeast	24.60															1.680	0.068	26.970	1.096		
$Ca(OH)_2$	74.09																				
Ca-salts	136.14																				
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01													25.812	0.587						
Cellulase	500.00																				
cellulose	324.00																				
Dirt	40.00																				
Ethanol	46.06	0.003	0.000											0.002	0.000						
Furfural	96.08																				
Glycerol	92.09																				
Hemicellulose	264.00																				
Hexosan	324.00																				
Hexoses	180.09																				
HMF	126.00																				
Impurity	33.15																				
Isoamylalcohol	82.20																				
Lactic Acid	90.04																				
Lignin	174.40																				
Lignin Monomere	90.90																				
Methance	16.04																				
N_2	28.01	23.000	0.821	22.977	0.820	0.023	0.001	23.000	0.821	23.000	0.821										
N_2O	46.05																				
O_2	32.00																				
Pentosan	264.00																				
Pentoses	150.06																				
Polyacrylamide	165.00																				
Succinic Acid	118.10																				
Sucrose	342.30																				
Sugar Cane	109.90																				
SO_2	64.06																				
H_2SO_4	98.08																			0.340	0.003
Sulphur	32.06																				
Water	18.02	2.224		0.116	0.006			0.116	0.006	0.116	0.006	30.385	1.686	1.110	0.062	6.653	0.369	107.880	5.987	348.050	19.315
Xylitol	152.00																				
Total		25.227	0.821	23.093	0.827	0.023	0.001	23.116	0.828	23.116	0.828	30.385	1.686	26.924	0.648	8.333	0.437	134.850	7.083	348.390	19.318
Enthalpy	kW	-		-639.749		-		-639.749		3,886.630		-		3,024.700		-		-		-	
Phase		G		G		G		G		G		L		G		Slurry		Slurry		L	
Press.	Bar	1		1		1		1		1		1		1		1		1		1	
Temp	°C	30.0		5.0		30.0		5.0		180.0		30.0		45.2		30.0		30.0		30.0	

STREAM Nr.		<331>		<332>		<333>		<334>		<335>		<401>		<402>		<403>		<404>		<405>	
Name :		Yeast&Acid		Waste Acid Solution		Recycle Yeast		Waste Water		Fusel Oil		Steam		Sulphuric Acid		Steam		Pretreatment Feed		Flash Vapor	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05									1.066	0.018									0.125	0.002
Ash	40.00																				
Bagasse	257.84																	46.097	0.179		
Yeast	24.60	26.970	1.096			26.970	1.096														
$Ca(OH)_2$	74.09																				
Ca-salts	136.14																				
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01																				
Cellulase	500.00																				
cellulose	324.00																				
Dirt	40.00																				
Ethanol	46.06							0.003	0.000	0.007	0.000										
Furfural	96.08									0.102	0.001									0.370	0.004
Glycerol	92.09									0.882	0.010										
Hemicellulose	264.00																				
Hexosan	324.00																				
Hexoses	180.09																				
HMF	126.00									0.019	0.000									0.068	0.001
Impurity	33.15																	0.855	0.026		
Isoamylalcohol	82.20									0.172	0.002										
Lactic Acid	90.04									0.054	0.001										
Lignin	174.40																				
Lignin Monomere	90.90																				
Methance	16.04																				
N_2	28.01							0.023	0.001												
N_2O	46.05																				
O_2	32.00																				
Pentosan	264.00																				
Pentoses	150.06																				
Polyacrylamide	165.00																				
Succinic Acid	118.10									0.092	0.001										
Sucrose	342.30																	2.494	0.007		
Sugar Cane	109.90																				
SO_2	64.06																				
H_2SO_4	98.08	0.340	0.003	0.340	0.003									0.598	0.006			0.598	0.006		
Sulphur	32.06																				
Water	18.02	455.930		348.050	19.315	107.880	5.987	2.108	0.117	237.147	13.160	5.853	0.325	46.600	2.586	20.188	1.120	98.046	5.441	23.869	1.325
Xylitol	152.00																				
Total		483.240	1.100	348.390	19.318	134.850	7.083	2.134	0.118	239.541	13.192	5.853	0.325	47.197	2.592	20.188	1.120	148.091	5.659	24.432	1.331
Enthalpy	kW	-		-		-		-221.039		10,748.725		16,290.000		13,566.495		56,187.530		33,390.595		49,828.240	
Phase		L		L		Slurry		L		L		G		L		G		L		G	
Press.	Bar	1		1		1		1		1		25		1		1		1		1	
Temp	°C	30.0		30.0		30.0		5.0		40.8		275.0		100.0		275.0		100.0		100.0	

STREAM Nr.		<406>		<407>		<408>		<409>		<410>		<411>		<412>		<413>		<414>		<415>	
Name :		Pretreatment Effl.		Rqw Hydrolyzate		Rqw Hydrolyzate		Hydrolyzate Bypass		Rqw Hydrolyzate		Overlimed Hydroly.		Conditioned Hydroly.		Clarified Hydroly.		Gypsum		Filtrate	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05	0.833	0.014	1.045	0.017	1.045	0.017	0.002	0.000	1.044	0.017	1.044	0.017	1.045	0.017	1.033	0.017	0.013	0.000	0.013	0.000
Ash	40.00	1.844	0.046	2.311	0.058	2.311	0.058	0.004	0.000	2.308	0.058	2.308	0.058	2.311	0.058	2.284	0.057	0.028	0.001	0.024	0.001
Bagasse	257.84																				
Yeast	24.60																				
$Ca(OH)_2$	74.09											0.001	0.000								
Ca-salts	136.14											0.828	0.006	0.829	0.006			0.829	0.006		
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01																				
Cellulase	500.00																				
cellulose	324.00	17.517		0.088	0.000	0.088	0.000	0.000	0.000	0.088	0.000	0.088	0.000	0.088	0.000			0.088	0.000	0.000	0.000
Dirt	40.00																				
Ethanol	46.06																				
Furfural	96.08	0.133		0.156	0.002	0.156	0.002	0.000	0.000	0.156	0.002	0.156	0.002	0.156	0.002	0.125	0.001	0.031	0.000		
Glycerol	92.09																				
Hemicellulose	264.00	1.420		0.007	0.000	0.007	0.000			0.007	0.000	0.007	0.000	0.007	0.000			0.007	0.000		
Hexosan	324.00																				
Hexoses	180.09	2.187		2.743	0.015	2.743	0.015	0.004	0.000	2.739	0.015	2.739	0.015	2.743	0.015	2.710	0.015	0.033	0.000	0.033	0.000
HMF	126.00	0.024		0.029	0.000	0.029	0.000			0.029	0.000	0.029	0.000	0.029	0.000	0.023	0.000	0.006	0.000		
Impurity	33.15	1.814	0.055	2.279	0.069	2.279	0.069	0.003	0.000	2.276	0.069	2.289	0.069	2.292	0.069	2.264	0.068	0.028	0.001	0.028	0.001
Isoamylalcohol	82.20																				
Lactic Acid	90.04																				
Lignin	174.40	11.036		0.055	0.000	0.055	0.000	0.000	0.000	0.055	0.000	0.055	0.000	0.055	0.000			0.055	0.000	0.000	0.000
Lignin Monomere	90.90	0.581		0.658	0.007	0.658	0.007	0.001	0.000	0.657	0.007	0.657	0.007	0.658	0.007	0.461	0.005	0.197	0.002		
Methance	16.04																				
N_2	28.01																				
N_2O	46.05																				
O_2	32.00																				
Pentosan	264.00																				
Pentoses	150.06	10.373		13.011	0.087	13.011	0.087	0.020	0.000	12.992	0.087	12.992	0.087	13.011	0.087	12.854	0.086	0.157	0.001	0.157	0.001
Polyacrylamide	165.00																				
Succinic Acid	118.10																				
Sucrose	342.30	2.494	0.007	3.186	0.009	3.186	0.009	0.005	0.000	3.181	0.009	3.341	0.010	3.346	0.010	3.305	0.010	0.041	0.000	0.041	0.000
Sugar Cane	109.90																				
SO_2	64.06																				
H_2SO_4	98.08	0.598	0.006	0.598	0.006	0.598	0.006	0.001	0.000	0.597	0.006			0.000	0.000	0.000	0.000				
Sulphur	32.06																				
Water	18.02	92.992		117.660	6.529	117.660	6.529	0.177	0.010	117.483	6.520	119.804	6.648	119.981	6.658	118.529	6.578	1.452	0.081	1.985	0.110
Xylitol	152.00																				
Total		143.847	0.128	143.826	6.800	143.826	6.800	0.216	0.010	143.611	6.790	146.336	6.920	146.552	6.930	143.587	6.837	2.965	0.093	2.282	0.113
Enthalpy	kW	36,964.363		32,127.397		13,967.487		20.977		13,946.510		14,520.393		14,541.370		14,348.586		192.784		118.461	
Phase		L		L		L		L		L		L		L		L		L		L	
Press.	Bar	1		1		1		1		1		1		1		1		1		1	
Temp	°C	100.0		84.4		50.0		50.0		50.0		50.0		50.0		50.0		50.0		41.4	

STREAM Nr.		<416>		<417>		<418>		<419>		<420>		<421>		<422>		<423>		<424>		<425>	
Name :		Gypsum		Hydrolysate		Wash Hydrolyzate		Hydrolysate		Hydrolysate		Fibre		BR-401 Effluent		C_5&C_6 Sugars		Fibre Residue		Wash Water	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05			1.045	0.017	0.406	0.007	0.639	0.011	0.639	0.011	0.195	0.003	0.833	0.014	0.833	0.014				
Ash	40.00	0.004	0.000	2.308	0.058	0.897	0.022	1.411	0.035	1.411	0.035	0.430	0.011	1.840	0.046	1.602	0.040	0.238	0.006		
Bagasse	257.84																				
Yeast	24.60																				
$Ca(OH)_2$	74.09																				
Ca-salts	136.14	0.829	0.006																		
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01																				
Cellulase	500.00													0.366	0.001	0.002	0.000	0.364	0.001		
cellulose	324.00	0.087		0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	17.429	0.054	1.743	0.005	0.009	0.000	1.734	0.005		
Dirt	40.00																				
Ethanol	46.06																				
Furfural	96.08	0.031		0.125	0.001	0.049	0.001	0.076	0.001	0.076	0.001	0.026	0.000	0.102	0.001	0.102	0.001				
Glycerol	92.09																				
Hemicellulose	264.00	0.007										1.413	0.005	1.413	0.005	0.007	0.000	1.406	0.005		
Hexosan	324.00																				
Hexoses	180.09			2.743	0.015	1.067	0.006	1.677	0.009	1.677	0.009	0.510	0.003	19.617	0.109	19.617	0.109				
HMF	126.00	0.006		0.023	0.000	0.009	0.000	0.014	0.000	0.014	0.000	0.005	0.000	0.019	0.000	0.019	0.000				
Impurity	33.15			2.292	0.069	0.891	0.027	1.401	0.042	1.401	0.042	0.425	0.013	1.826	0.055	1.826	0.055				
Isoamylalcohol	82.20																				
Lactic Acid	90.04																				
Lignin	174.40	0.055		0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	10.981	0.063	10.981	0.063	0.055	0.000	10.926	0.063		
Lignin Monomere	90.90	0.197		0.461	0.005	0.179	0.002	0.282	0.003	0.282	0.003	0.102	0.001	0.383	0.004	0.383	0.004				
Methance	16.04																				
N_2	28.01																				
N_2O	46.05																				
O_2	32.00																				
Pentosan	264.00																				
Pentoses	150.06			13.011	0.087	5.059	0.034	7.953	0.053	7.953	0.053	2.421	0.016	10.373	0.069	10.373	0.069				
Polyacrylamide	165.00																				
Succinic Acid	118.10																				
Sucrose	342.30			3.346	0.010	1.301	0.004	2.045	0.006	2.045	0.006	0.609	0.002	2.654	0.008	2.654	0.008				
Sugar Cane	109.90																				
SO_2	64.06																				
H_2SO_4	98.08			0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000			0.000	0.000	0.000	0.000				
Sulphur	32.06																				
Water	18.02	1.478		120.515	6.688	46.856	2.600	73.659	4.088	73.659	4.088	22.189	1.231	94.104	5.222	113.066	6.274	12.983	0.720	33.956	1.884
Xylitol	152.00																				
Total		2.694	0.006	145.869	6.950	56.714	2.702	89.155	4.248	89.155	4.248	56.734	1.402	146.255	5.603	150.549	6.575	27.651	0.800	33.956	1.884
Enthalpy	kW	73.983		14,467.047		4,143.122		10,323.925		11,711.461		8,979.928		20,691.389		18,697.566		1,993.823		-	
Phase		Slurry		L		L		L		L		S		L		L		S		L	
Press.	Bar	1		1		1		1		1		1		1		1		57.1		1	
Temp	°C	41.4		49.9		49.9		49.9		54.4		84.4		65.0		57.1				30.0	

STREAM Nr.		<426>		<427>		<428>		<429>		<430>		<431>		<432>		<433>		<434>		<435>	
Name :		Wash Water		Wash Water		Filter Cake		E-403 Vapor		Vinasse Syrup		Solid Residues		Dry Fuel		Dirt, Gypsum, Ash		Air		Dry Exhaust Gas	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05							0.027	0.000	0.011	0.000										
Ash	40.00					0.004	0.000			1.602	0.040	0.242	0.006	0.242	0.006	1.856	0.046				
Bagasse	257.84					0.466	0.002					0.466	0.002	0.466	0.002						
Yeast	24.60											1.680	0.068	1.680	0.068						
$Ca(OH)_2$	74.09																				
Ca-salts	136.14					1.184	0.009					1.184	0.009	1.184	0.009	1.184	0.009				
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01																			42.654	0.969
Cellulase	500.00									0.002	0.000	0.364	0.001	0.364	0.001						
cellulose	324.00					0.087	0.000			0.009	0.000	1.821	0.006	1.821	0.006						
Dirt	40.00					3.326	0.083					3.326	0.083	3.326	0.083	3.326	0.083				
Ethanol	46.06																				
Furfural	96.08					0.031	0.000					0.031	0.000	0.031	0.000						
Glycerol	92.09							0.047	0.001	0.323	0.004										
Hemicellulose	264.00					0.007	0.000			0.007	0.000	1.413	0.005	1.413	0.005						
Hexosan	324.00																				
Hexoses	180.09																				
HMF	126.00					0.006	0.000					0.006	0.000	0.006	0.000						
Impurity	33.15					2.254	0.068			4.816	0.145	2.254	0.068	2.254	0.068						
Isoamylalcohol	82.20							0.005	0.000	0.002	0.000										
Lactic Acid	90.04							0.004	0.000	0.017	0.000										
Lignin	174.40					0.055	0.000			0.055	0.000	10.981	0.063	10.981	0.063						
Lignin Monomere	90.90					0.197	0.002			0.383	0.004	0.197	0.002	0.197	0.002						
Methance	16.04																				
N_2	28.01																	174.831	6.242	174.831	6.242
N_2O	46.05																			0.745	0.016
O_2	32.00																	46.474	1.452	15.466	0.483
Pentosan	264.00																				
Pentoses	150.06																				
Polyacrylamide	165.00					0.001	0.000					0.001	0.000	0.001	0.000						
Succinic Acid	118.10							0.005	0.000	0.034	0.000										
Sucrose	342.30					0.475	0.001					0.475	0.001	0.475	0.001						
Sugar Cane	109.90																				
SO_2	64.06																			0.015	0.000
H_2SO_4	98.08									0.000	0.000					0.000	0.000				
Sulphur	32.06																				
Water	18.02	2.011		31.944	1.773	9.896	0.549	8.535	0.474	2.663	0.148	29.532	1.639	3.257	0.181					18.864	1.047
Xylitol	152.00									0.016	0.000										
Total		2.011		31.944	1.773	17.987	0.715	8.623	0.475	9.938	0.342	53.972	1.953	27.697	0.495	6.366	0.138	221.306	7.694	252.575	8.757
Enthalpy	kW	-		-		2,617.671		17,629.809		2,135.463		4,744.500		3,056.300		6,406.900		-		139,488.800	
Phase		L		L		Slurry		G		L		Slurry		S		S		G		G	
Press.	Bar	1		1		1		1		1		1		1		1		1		1	
Temp	°C	30.0		30.0		85.3		103.0		103.0		61.0		100.0		1,100.0		30.0		350.0	

STREAM Nr.		<436>		<437>																	
Name :		Wet Exhaust Air		Cellulase																	
COMP	MW	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s	kg/s	kmol/s
Acetic Acid	60.05																				
Ash	40.00																				
Bagasse	257.84																				
Yeast	24.60																				
$Ca(OH)_2$	74.09																				
Ca-salts	136.14																				
$CaCO_3$	100.09																				
CaO	56.08																				
CO_2	44.01	42.654	0.969																		
Cellulase	500.00			0.366	0.001																
cellulose	324.00																				
Dirt	40.00																				
Ethanol	46.06																				
Furfural	96.08																				
Glycerol	92.09																				
Hemicellulose	264.00																				
Hexosan	324.00																				
Hexoses	180.09																				
HMF	126.00																				
Impurity	33.15																				
Isoamylalcohol	82.20																				
Lactic Acid	90.04																				
Lignin	174.40																				
Lignin Monomere	90.90																				
Methance	16.04																				
N_2	28.01	174.831	6.242																		
N_2O	46.05	0.745	0.016																		
O_2	32.00	15.466	0.483																		
Pentosan	264.00																				
Pentoses	150.06																				
Polyacrylamide	165.00																				
Succinic Acid	118.10																				
Sucrose	342.30																				
Sugar Cane	109.90																				
SO_2	64.06	0.015	0.000																		
H_2SO_4	98.08																				
Sulphur	32.06																				
Water	18.02	45.139																			
Xylitol	152.00																				
Total		278.850	7.711	0.366	0.001																
Enthalpy	kW	141,177.000		-																	
Phase		G		S																	
Press.	Bar	1		1		1		1		1		1		1		1		1		1	
Temp	°C	124.8		30.0																	

A.5. Utilities Summary

A.5.1. 2005 Plant

The utility requirements of the equipments in 2005 plant are given in Tables A.5.1-5.8.

Table A.5.1Mill equipments (2005)

Equipment	Name	Electricity	Heating			Cooling		
			MPS	LPS	HW	CW	ChW	RFG
		kW	ton/h			ton/h		kW
Cane carriers	M-101	101.0						
Cane knives	M-101	6,250.0						
Magnetic separators	M-101							
Crushers	M-101	1,256.8						
Mills	M-101	5,027.4						
Intermediate carriers	M-101	40.4						
Bagasse carriers	M-101	61.0						
Feed pumps	P-103	19.0						
Circulatory pumps	P-101	1.3						
Sum		12,757.0	0.0	0.0	0.0	0.0	0.0	0.0
Notes		MPS: Medium pressure steam at 25 bar 275 ºC LPS : Low pressure steam at 2.75 bar and 130.6 ºC HW : Hot water at 130.6 ºC CW : Cooling Water at 25ºC ChW : Chilled Water at 10ºC RFG: refrigeration 0ºC						

Table A.5.2 Clarification equipments (2005)

Equipment	Name	Electricity	Heating			Cooling		
			MPS	LPS	HW	CW	ChW	RFG
		kW	ton/h			ton/h		kW
Sulpur burner	F-102							
Lime kiln	F-101							
Sulphitation tower	C-101							
Lime milk prep. tanks	V-101	6.3				8.16		
Liming tank	V-102	364.0						
Liming tank	V-103	196.0						
Juice heater	E-101			78.4*				
Juice heater	E-102			42.5*				
Flash tank	V-104							
Flash tank	V-105							
Clarifiers	S-101	19.2						
Rotary filter	S-103	578.0						
	P-102	29.0						
	P-104	0.5						
	P-105	0.17						
	P-106	17.4						
	P-107	9.4						
	P-108	0.5						
	P-109	1.1						
	P-110	26.3						
	P-111	0.8						
	P-112	26.6						
Sum		1,289.47	0.0	0.0	0.0	8.16	0.00	0.00
Notes		See Notes of Table A.5.1 *this steam is supplied by the exhaust gas this is why is not counted in total steam consumption						

Table A.5.3 Evaporation equipments (2005)

Equipment	Name	Electricity	Heating			Cooling		
			MPS	**LPS**	**HW**	**CW**	**ChW**	**RFG**
		kW	**ton/h**			**ton/h**		**kW**
1st Effect evaporator	E-201			105.94				
2nd Effect evaporator	E-202							
3rd Effect evaporator	E-203							
4th Effect evaporator	E-204							
5th Effect evaporator	E-205							
6th Effect evaporator	E-206					2,361.3		
Pumps	P-201	10.0						
Sum		10.0	0.0	105.94	0.0	2,361.3	0.0	0.0
Notes		See Notes of Table A.5.1						

Table A.5.4 Sugar boiling, crystallization and drying equipments (2005)

Equipment	Name	Electricity	Heating			Cooling		
			MPS	**LPS**	**HW**	**CW**	**ChW**	**RFG**
		kW	**ton/h**			**ton/h**		**kW**
First boiling pan	E-207	234.8		27.0		312.4		
Second boiling pan	E-208	133.5		15.4		178.5		
First crystallizer	V-201	165.6				147.5		
Second crystallizer	V-202	148.2				65.9		
First centrifuge	S-201	300.0						
Second centrifuge	S-202	210.0						
Reheater	E-209				42.7			
Reheater	E-210				23.4			
Reheater	E-211				21.2			
Reheater	E-212				10.3			
Air heater	E-213			1.0				
Rotary drier	D-201	13.0						
Sugar carrier		12.2						
Pumps	P-202	8.2						
Pumps	P-203	3.3						
Pumps	P-204	1.23						
Pumps	P-205	3.7						
Pumps	P-206	0.55						
Pumps	P-207	2.7						
Sum		1,236.98	0.0	43.4	97.6	704.3	0.0	0.0
Notes		See Notes of Table A.5.1						

Table A.5.5 Fermentation equipments (2005)

Equipment	Name	Electricity	Heating			Cooling		
			MPS	**LPS**	**HW**	**CW**	**ChW**	**RFG**
		kW	**ton/h**			**ton/h**		**kW**
Juice cooler	E-301							
Juice cooler	E-302						227.3	
Fermentors	V-301						78.0	
first centrifuge	S-301	61.6						
second centrifuge	S-302	67.6						
Yeast sterilization tank	V-302	260.0						
Scrubber column	C-301							
Sum		389.2	0.0	0.0	0.0	0.0	305.3	0.0
Notes		See Notes of Table A.5.1						

Table A.5.6 Ethanol dehydration equipments (2005)

Equipment	**Name**	**Electricity**	**Heating**			**Cooling**		
			MPS	**LPS**	**HW**	**CW**	**ChW**	**RFG**
		kW	**ton/h**			**ton/h**		**kW**
Beer column	C-302							
Rectification column	C-303							
Molecular sieve column	C-304		3.8			60.5		
Beer condenser	E-304						149.3	
Rectification condenser	E-306							
Molecular sieve condenser	E-309							1,107.0
Beer reboiler	E-303			127.5				
Rectification reboiler	E-305			14.6		1,396.0		
Molecular sieve feed cooler	E-307					38.6		
Nitrogen exchanger	E-308							
Product cooler	E-310					60.8		
Pumps	P-301	5.9						
Pumps	P-302	5.9						
Pumps	P-303	5.5						
Pumps	P-304	30.5						
Pumps	P-305	4.3						
Pumps	P-306	1.5						
Pumps	P-307	27.9						
Pumps	P-308	5.7						
Pumps	P-309	34.9						
Pumps	P-310	4.3						
Pumps	P-311	11.7						
Pumps	P-312	16.2						
Pumps	K-301	325.0						
Pumps	K-302	158.0						
Pumps	K-303	288.4						
Pumps	K-304	150.5						
Pumps	K-305	425.5						
Sum		1,501.7	3.8	142.1	0.0	1,555.9	149.3	1,107.0
Notes		See Notes of Table A.5.1						

Table A.5.7 Cogeneration equipments (2005)

Equipment	**Name**	**Electricity**	**Heating**			**Cooling**		
			MPS	**LPS**	**HW**	**CW**	**ChW**	**RFG**
		kW	**ton/h**			**ton/h**		**kW**
Rotary driers	D-101	940.00						
Boiler	F-103	16.56						
Turbine generator	F-103							
Cooling tower		280.00						
Sum		1,236.56						
Notes		See Notes of Table A.5.1						

Table A.5.8 2005 total utility requirement

Equipment	**Name**	**Electricity**	**Heating**			**Cooling**		
			MPS	**LPS**	**HW**	**CW**	**ChW**	**RFG**
		kW	**ton/h**			**ton/h**		**kW**
2005 Total		18,406.71	3.80	412.34	97.60	4,629.66	454.6	1,107.00
Notes		See Notes of Table A.5.1						

A.5.2. 2015 Plant

The utility requirement of the equipments in 2005 plant is given in Tables A.5.9-5.15. The equipments that are same with 2005 plant are not shown and can be found in Appendix A.5.1.

Table A.5.9 Clarification equipments (2015)

Equipment	Name	Electricity	Heating			Cooling		
			MPS	LPS	HW	CW	ChW	RFG
		kW	ton/h			ton/h		kW
Juice heater	E-101			78.4*				
Juice heater	E-102			42.5*				
Lime milk prep. Tanks	V-101	20.50				24.60		
Sum		20.50	0.00	0.00	0.00	24.60	0.00	0.00
Notes		See Notes of Table A.5.1 *this steam is supplied by the exhaust gas this is why is not counted in total steam consumption						

Table A.5.10 Sugar boiling, crystallization and drying equipments (2015)

Equipment	Name	Electricity	Heating			Cooling		
			MPS	LPS	HW	CW	ChW	RFG
		kW	ton/h			ton/h		kW
Reheater	E-209				175.7			
Reheater	E-210				96.3			
Reheater	E-211				87.1			
Reheater	E-212				42.4			
Sum		0.00	0.00	0.00	401.5	0.00	0.00	0.00
Notes		See Notes of Table A.5.1						

Table A.5.11 Fermentation equipments (2015)

Equipment	Name	Electricity	Heating			Cooling		
			MPS	LPS	HW	CW	ChW	RFG
		kW	ton/h			ton/h		kW
Juice cooler	E-301					3,251.00		
Product cooler	E-302							
Fermentors	V-301						187.00	
First centrifuge	S-301	123.20						
Second centrifuge	S-302	156.80						
Yeast sterilization tank	V-302	426.00						
Scrubber column	C-301							
Sum		706.00	0.00	0.00	0.00	3,251.00	187.00	0.00
Notes		See Notes of Table A.5.1						

Table A.5.12 Ethanol dehydration equipments (2015)

Equipment	Name	Electricity	Heating			Cooling		
			MPS	LPS	HW	CW	ChW	RFG
		kW	ton/h			ton/h		kW
Molecular sieve column	C-304		8.75			144.6		
Beer condenser	E-304						1056.2	
Rectification condenser	E-306					3348.5		
Molecular sieve condenser	E-309							1977
Beer reboiler	E-303			274.5				
Rectification reboiler	E-305			46.1				
Molecular sieve feed cooler	E-307					92.2		
Nitrogen exchanger	E-308							

Table A.5.12 (Continuation) Ethanol dehydration equipments (2015)

Equipment	Name	Electricity	Heating			Cooling		
			MPS	LPS	HW	CW	ChW	RFG
		kW	ton/h			ton/h		kW
Product cooler	E-310	13.09						
Pumps	P-301	13.44						
Pumps	P-302	11.06						
Pumps	P-303	64.94						
Pumps	P-304	6.03						
Pumps	P-305	2.59						
Pumps	P-306	63.48						
Pumps	P-307	23.90						
Pumps	P-308	72.15						
Pumps	P-309	10.34						
Pumps	P-310	22.22						
Pumps	P-311	32.47						
Pumps	P-312	632.64						
Pumps	K-301	266.11						
Pumps	K-302	1121.6						
Pumps	K-303	303.43						
Pumps	K-304	852.27						
Pumps	K-305	13.09						
Sum		3,511.76	8.75	320.60	0.00	3,585.30	1,056.20	1,977.00

Table A.5.13 Pretreatment and hydrolysis equipments (2015)

Equipment	Name	Electricity	Heating			Cooling		
			MPS	LPS	HW	CW	ChW	RFG
		kW	ton/h			ton/h		kW
Pretreatment Reactor	R-401		93.70					
Collection flash tank	V-401							
Screw Press	S-401	60.00						
Screw Press	S-401							
Hydrolysate Cooler	E-401					1,045.30		
Hydrolysate heater	E-402			2.45				
Overliming Vessel	V-402	24.69						
Conditioning Vessel	V-403	24.69						
Cellulose Hydrolysis vessel	BR-401	329.29						
Clarifiers	S-402	4.89						
Rotary Vacuum Filter	S-403	91.29						
Fibre Conveyors		12.29						
Fibre Conveyors		12.29						
Pumps	P-401	482.56						
Pumps	P-402	7.76						
Pumps	P-403	31.67						
Pumps	P-404	0.92						
Pumps	P-405	31.67						
Pumps	P-406	21.11						
Pumps	P-407	0.13						
Pumps	P-408	43.73						
Pumps	P-409	0.69						
Pumps	P-410-A	41.92						
Pumps	P-410-B	3.99						
Pumps	P-411	44.23						
Pumps	P-412	0.39						
Sum		1,269.48	93.70	2.45	0.00	1,045.30	0.00	0.00
Notes		See Notes of Table A.5.1						

Table A.5.14 Cogeneration equipments (2015)

Equipment	Name	Electricity	Heating			Cooling		
			MPS	LPS	HW	CW	ChW	RFG
		kW	ton/h			ton/h		kW
Rotary driers	D-101	564.00						
Boiler	F-103	13.93						
Evaporator	F-103			32.3				
Cooling tower		599.00						
Sum		1,176.93	0.00	32.3	0.00	0.00	0.00	0.00
Notes		See Notes of Table A.5.1						

Table A.5.15 2015 total utility requirement

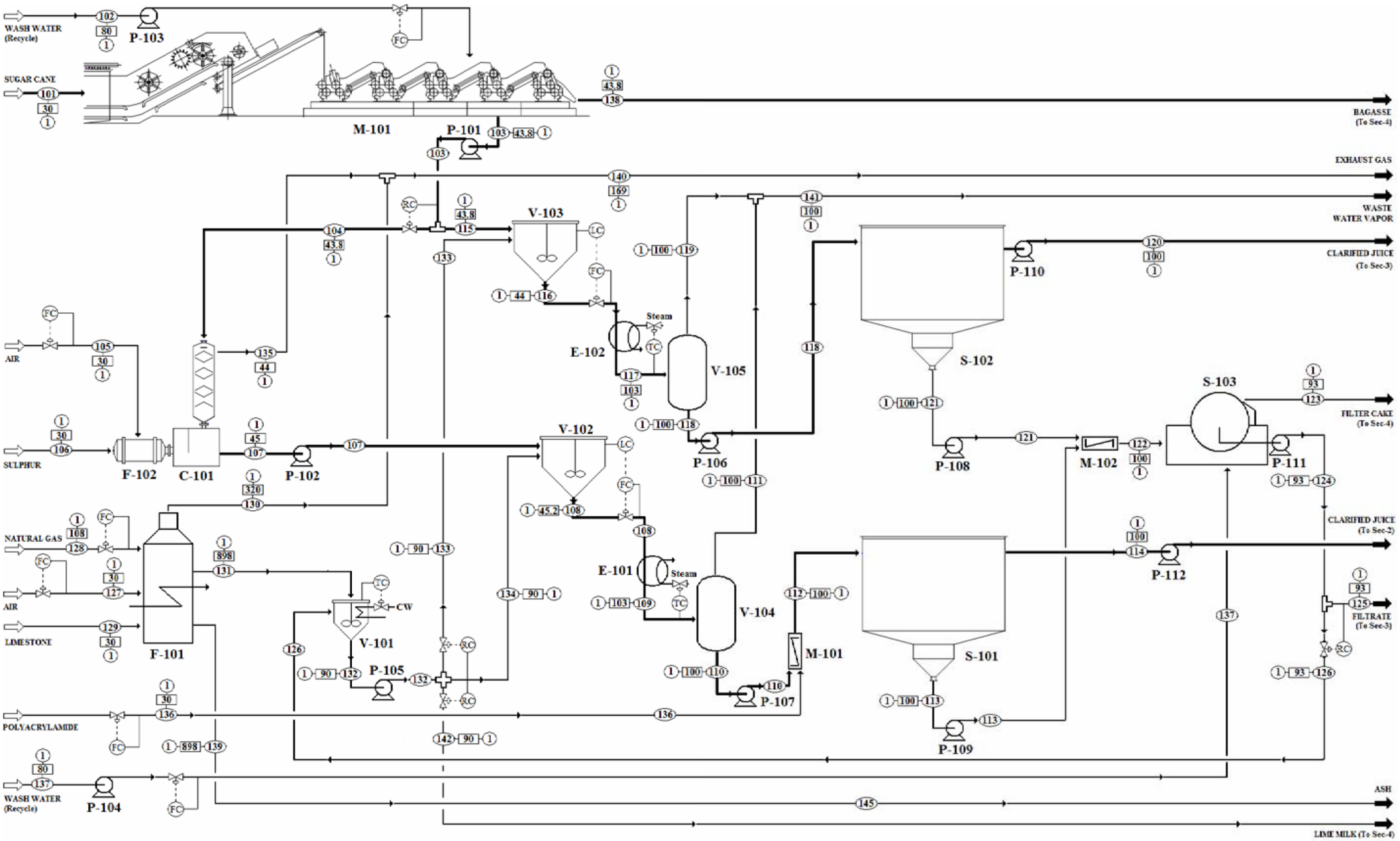
Equipment	Name	Electricity	Heating			Cooling		
			MPS	LPS	HW	CW	ChW	RFG
		kW	ton/h			ton/h		kW
2015 Total		21,957.62	102.45	504.81	401.50	10,971.80	1,243.20	1,977.00
Notes		See Notes of Table A.5.1						

A.6. Process Control Schemes

2015 process control block scheme is composed of four sections

- Mill & Clarification (Section 1)
- Juice evaporation & Crystallization (Section 2)
- Ethanol production & Purification (Section 3)
- Pretreatment & Hydrolysis (Section 4)

FC : Flow Controller
RC : Ratio Controller
PC : Pressure Controller
TC : Temperature Controller
LC : Level Controller

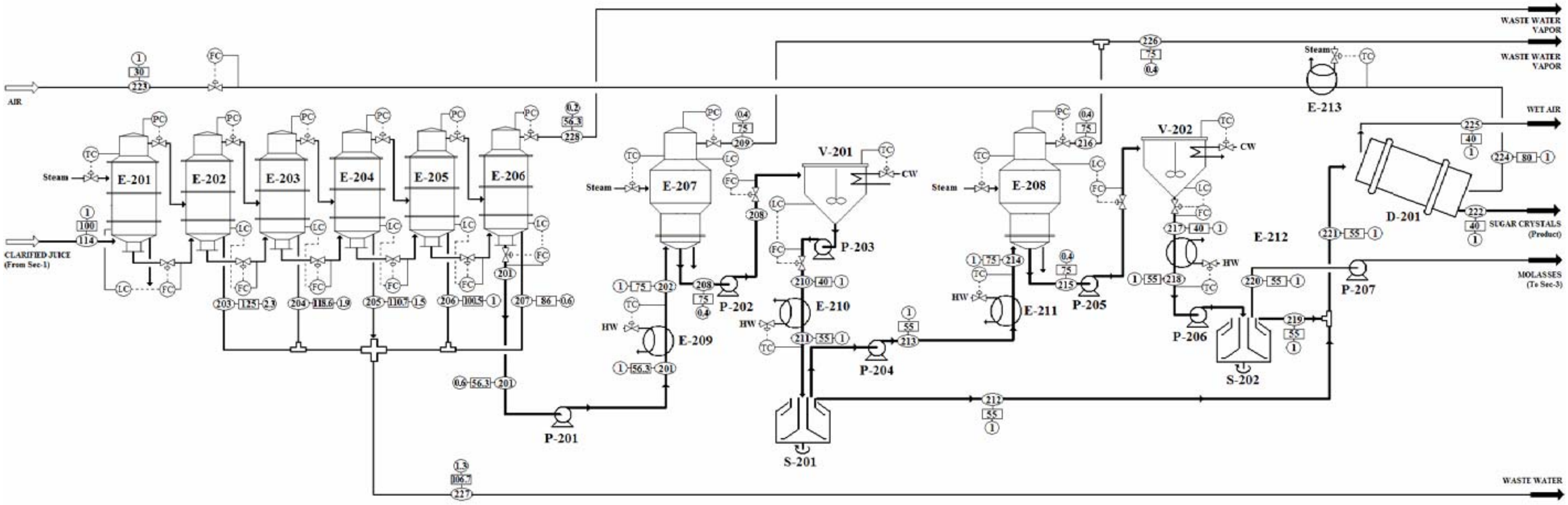


PROCESS EQUIPMENT SUMMARY

C-101	: Sulphitation Column
E-101	: Sugar Branch Juice Heater
E-102	: Ethanol Branch Juice Heater
F-101	: Lime Kiln
F-102	: Sulphur burner
M-101	: Roller Mills
S-101	: Clarifier&Settler
S-102	: Clarifier&Settler
S-103	: Rotary Vacuum Filter
P-101	: Circulation Pumps
P-102	: Sulphitated Juice Pumps
P-103	: Mill Wash Water Pumps
P-104	: Rotary Filter Wash Water Pump
P-105	: Lime Milk Pumps
P-106&107	: Limed Juice Pumps
P-108&109	: Clarifier Mud Pumps
P-110&112	: Clarified Juice Pumps
P-111	: Rotary Filter filtrate Pump
V-101	: Lime Milk Preparation Vessel
V-102&103	: Liming Tanks
V-104&105	: Flash Tanks
X-101&102	: Static Mixers
FC	: Flow Controller
LC	: Level Controller
RC	: Ratio Controller
TC	: Temperature Controller
PC	: Pressure Controller

PROCESS CONTROL SCHEME

PROJECT NAME	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse	
DESIGNER	DATE	2015 SECTION 1 : MILL AND CLARIFICATION SECTION
Çağrı Efe	28-09-2005	⬭ Stream Number ▭ Temperature (°C) ○ Pressure (bar)

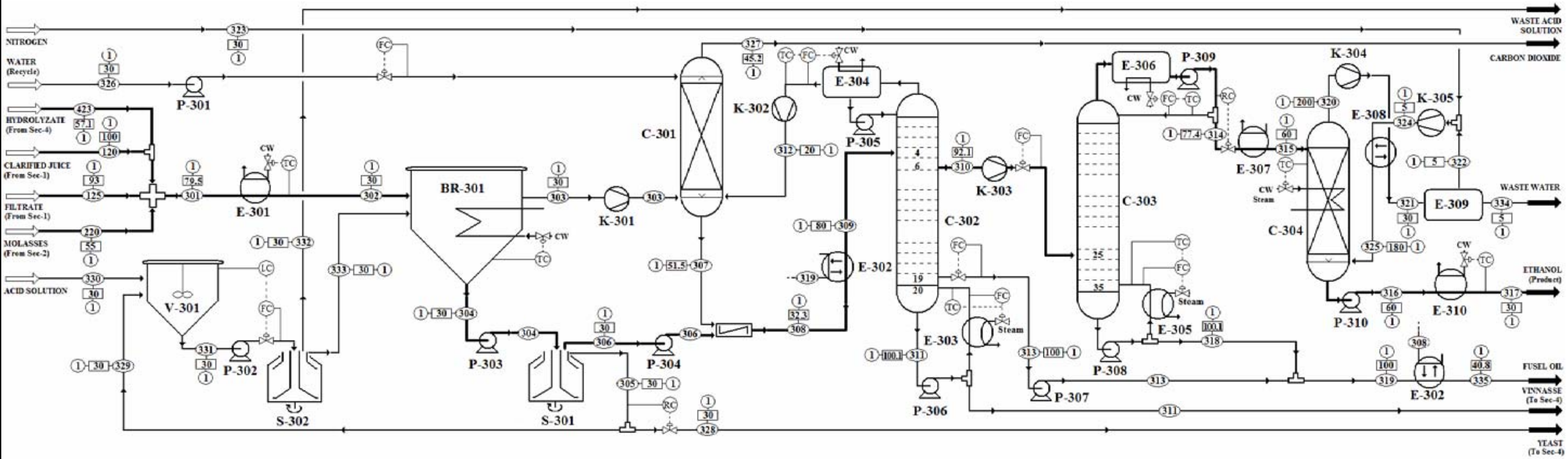


PROCESS EQUIPMENT SUMMARY

D-201	: Sugar Rotary Drier	**E-207**	: 1st Boiling Pan	**P-201**	: Juice Pumps	**S-201**	: Crystal Sugar Separation Centrifuges
E-201	: 1st Effect Evaporator	**E-208**	: 2nd Boiling Pan	**P-202**	: Massecuite Pumps	**S-202**	: Crystal Sugar Separation Centrifuges
E-202	: 2nd Effect Evaporator	**E-209**	: Juice Heater	**P-203**	: Massecuite Pumps	**V-201**	: 1st Boiling Pan Sugar Crystallizers
E-203	: 3rd Effect Evaporator	**E-210**	: Massecuites Reheater	**P-204**	: Molasses Pumps	**V-202**	: 2nd Boiling Pan Sugar Crystallizers
E-204	: 4th Effect Evaporator	**E-211**	: Molasse Reheater	**P-205**	: Massecuite Pumps	**FC**	: Flow Controller
E-205	: 5th Effect Evaporator	**E-212**	: Massequites Reheater	**P-206**	: Massecuite Pumps	**LC**	: Level Controller **TC** : Temperature Controller
E-206	: 6th Effect Evaporator	**E-213**	: Drier Air Heater	**P-207**	: Molasses Pumps	**RC**	: Ratio Controller **PC** : Pressure Controller

PROCESS CONTROL SCHEME

PROJECT NAME	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse	
DESIGNER	DATE	2015 SECTION 2 : JUICE EVAPORATION AND CRYSTALLIZATION
Çağrı Efe	28-09-2005	⬭ Stream Number ▭ Temperature (°C) ○ Pressure (bar)

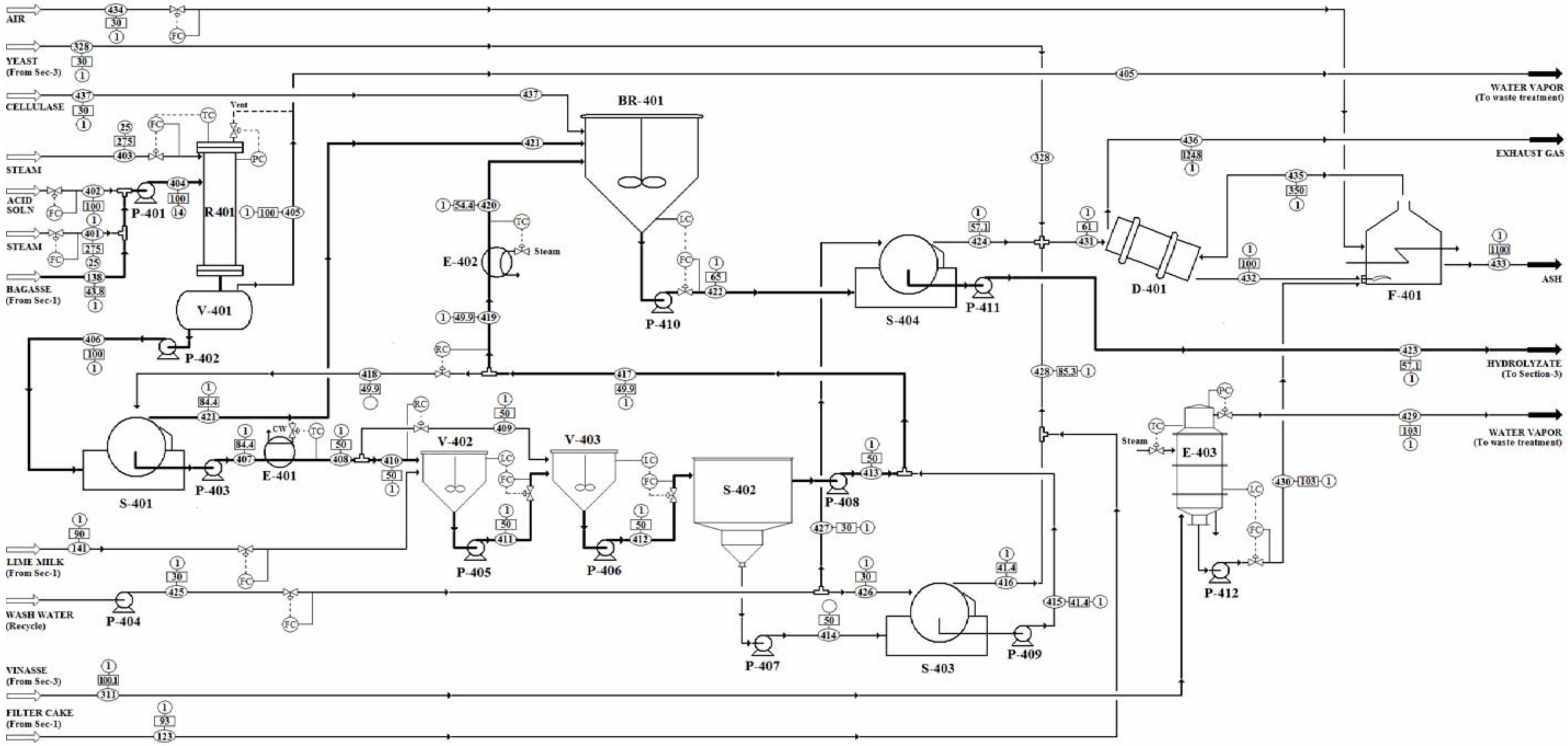


PROCESS EQUIPMENT SUMMARY

BR-301 : Fermentors
C-301 : Ethanol Scrubber
C-302 : Beer Column
C-303 : Rectification Column
C-304 : Molecular Sieve Column
E-301 : Juice Cooler
E-302 : Beer Column Feed Heater
E-303 : Beer Column Reboiler
E-304 : Beer Column Condenser
E-305 : Rectification Column Reboiler
E-306 : Rectification Column Condenser
E-307 : Molecular Sieve Feed Cooler
E-308 : Wet Nitrogen/Dry Nitrogen Heat Ex.
E-309 : Nitrogen Drying Condenser
E-310 : Ethanol Cooler
K-301 : CO2 Blower
K-302 : CO2 Blower
K-303 : Product Vapor Blower
K-304 : Wet Nitrogen Blower
K-305 : Dry Nitrogen Blower
P-301 : Water Pump
P-302 : Biomass Pump
P-303 : Fermentation Broth Pump
P-304 : Yeast Free Broth Pump
P-305 : Beer Condenser Pump
P-306 : Beer Reboiler Pump
P-307 : Fusel Oil Pump
P-308 : Rectification Reboiler Pump
P-309 : Rectification Condenser Pump
P-310 : Product Pump
P-311&312 : Biomass Pump
S-301 : Yeast Separation Centrifuges
S-302 : Yeast Separation Centrifuges
V-301 : Yeast Sterilization Tank

PROCESS CONTROL SCHEME

PROJECT NAME	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse	
DESIGNER	DATE	2015 SECTION 3 : FERMENTATION AND ETHANOL PURIFICATION SECTION
Çağrı Efe	28-09-2005	⬭ Stream Number ▭ Temperature (℃) ○ Pressure (bar)



PROCESS EQUIPMENT SUMMARY

BR-401	Cellulose Hydrolysis Vessels
D-401	Dewatering Rotary Drum
E-401	Hydrolyzate Cooler
E-402	Clarified Hydrolyzate Heater
E-403	Vinasse Evaporator
F-401	Boiler Furnace
P-401	High Presure Fiber Pump
P-402	Fiber&Hydrolyzate Pumps
P-403	Screw Press Permeate Pumps
P-404	Water Pump
P-405	Hydrolyzate Pumps
P-406	Hydrolyzate Pumps
P-407	Clarifier Mud Pump
P-408	Clarified Hydrolyzate Pump
P-409	Filtrate Pump
P-410	Hydrolyzate Pump
P-411	Hydrolysate Pump
P-412	Concentrated Vinasse Pump
R-401	Pretreatment Reactor
S-401	Screw Press
S-402	Clarifier&Settler
S-403	Rotary Vacuum Filter
S-404	Screw Press
V-401	Pretreatment Collection Tank
V-402	Over Liming Vessel
V-403	Conditioning Vessel

PROCESS CONTROL SCHEME

PROJECT NAME	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse	
DESIGNER	DATE	2015 SECTION 4 : BAGASSE PRETREATMENT AND HYDROLYSIS SECTION
Çağrı Efe	28-09-2005	⬭ Stream Number ▭ Temperature (°C) ○ Pressure (bar)

A.7. Mass and Heat Balances

A.7.1. 2005 Design Mass and Heat Balances

HEAT & MASS BALANCE FOR STREAMS TOTAL 2005 Section 1- Mill and Clarification										
IN					EQUIPM. IDENTIF.	OUT				
Section		EQUIPMENT				EQUIPMENT			Section	
Mass kg/s	Heat kW	Mass kg/s	Heat kW	Stream Nr.		Stream Nr.	Mass kg/s	Heat kW	Mass kg/s	Heat kW
332.589		332.589		<101>	**M101**	<103>	330.673	15,817.285		
93.125	19,351.310	93.125	19,351.310	<102>		<138>	95.040	3,534.100		
425.714	19,351.310	425.714	19,351.310		Total		425.713	19,351.385		
		214.937	10,281.182	<104>	**F-102&C-101**	<107>	215.136	11,566.290		
0.930		0.930		<105>		<135>	0.839	34.442	0.839	34.442
0.108		0.108		<106>						
	1,002.300		1,002.300		**Combustion**					
1.038	1,002.300	215.975	11,283.482		Total		215.975	11,600.732	0.839	34.442
0.550		0.550		<127>	**F-101**	<130>	0.747	246.192	0.747	246.192
		0.062	7.721	<128>		<131>	0.166	132.085		
0.303		0.303		<129>		<145>	0.001	0.001	0.001	0.001
	1,025.480		1,025.480		**Combustion**					
					Reaction			654.200		
0.853	1,025.480	0.915	1,033.201		Total		0.914	1,032.478	0.748	246.193
		1.146	285.702	<126>	**V-101**	<132>	1.312	270.082		
		0.166	132.085	<131>						
	-47.288		-47.288		**Cooling**			100.489		100.489
	-47.288	1.312	370.499		Total		1.312	370.571		100.489
		215.136	11,566.290	<107>	**V-102**	<108>	215.989	11,741.837		
		0.853	175.547	<134>						
		215.989	11,741.837		Total		215.989	11,741.837		
		115.735	5,536.018	<115>	**V-103**	<116>	116.195	5,630.545		
		0.459	94.527	<133>						
		116.194	5,630.545		Total		116.195	5,630.545		
		215.989	11,741.837	<108>	**E-101**	<109>	215.989	59,021.141		
	56,298.310		56,298.310		**Steam**			9,018.270		9,018.270
	56,298.310	215.989	68,040.147		Total		215.989	68,039.411		9,018.270
		116.195	5,630.545	<116>	**E-102**	<117>	116.195	31,261.371		
	30,519.691		30,519.691		**Steam**			4,888.865		4,888.865
	30,519.691	116.195	36,150.236		Total		116.195	36,150.236		4,888.865
		215.989	59,021.141	<109>	**V-104**	<110>	214.883	56,523.117		
						<111>	1.105	2,497.300	1.105	2,497.300
		215.989	59,021.141		Total		215.988	59,020.417	1.105	2,497.300
		116.195	31,261.371	<117>	**V-105**	<118>	115.604	29,925.711		
						<119>	0.591	1,335.660	0.591	1,335.660
		116.195	31,261.371		Total		116.195	31,261.371	0.591	1,335.660
		214.883	56,523.117	<110>	**S-101**	<113>	18.151	4,170.265		
0.135		0.135		<136>		<114>	196.868	52,352.715	196.868	52,352.715
0.135		215.018	56,523.117		Total		215.018	56,522.980	196.868	52,352.715
		115.604	29,925.711	<118>	**S-102**	<120>	105.859	27,308.876	105.859	27,308.876
						<121>	9.745	2,616.739		
		115.604	29,925.711		Total		115.604	29,925.615	105.859	27,308.876
		18.151	4,170.265	<113>	**S-103**	<123>	15.290	2,999.711	15.290	2,999.711
		9.745	2,616.739	<121>		<125>	27.016	6,733.959	27.016	6,733.959
15.556	3,232.454	15.556	3,232.454	<137>		<126>	1.146	285.702		
15.556	3,232.454	43.451	10,019.458		Total		43.452	10,019.372	42.305	9,733.670
		95.040	3,534.100	<138>	**D-101**	<139>	56.153	6,576.105		
		447.624	328,535.800	<142>		<143>	486.511	325,493.700	486.511	325,493.700
		542.664	332,069.900		Total		542.664	332,069.805	486.511	325,493.700
		56.091	6,568.384	<140>	**F-103**	<142>	447.624	328,535.800		
392.771		392.771		<141>		<144>	1.239	1.778	1.239	1.778
	1,004,198.300		1,004,198.300		**Combustion**					
					Steam			682,460.300		682,460.300
392.771	1,004,198.300	448.862	1,010,766.684		Total		448.863	1,010,997.878	1.239	682,462.078
836.067	1,115,580.557				**Total**				836.065	1,115,472.258
OUT-IN :									0.000	-108.298

HEAT & MASS BALANCE FOR STREAMS TOTAL

2005 Section 2- Juice Evaporation and Crystallization

IN					EQUIPM. IDENTIF.	OUT				
Section		EQUIPMENT					EQUIPMENT		Section	
Mass kg/s	Heat kW	Mass kg/s	Heat kW	Stream Nr.		Stream Nr.	Mass kg/s	Heat kW	Mass kg/s	Heat kW
196.868	52,352.715	196.868	52,352.715	<114>	**E-201, E-202**	<201>	46.310	2,669.438		
					E-203,E-204	<203>	20.265	7,973.968	20.265	7,973.968
					E-205,E-206	<204>	22.132	8,112.807	22.132	8,112.807
						<205>	24.029	8,027.220	24.029	8,027.220
						<206>	25.941	7,582.362	25.941	7,582.362
						<207>	27.929	6,485.117	27.929	6,485.117
						<228>	30.262	74,978.637	30.262	74,978.637
	76,075.646		76,075.646		**Steam**			12,187.941		12,187.941
196.868	128,428.361	196.868	128,428.361		Total		196.867	128,017.490	150.558	125,348.052
		46.310	2,669.438	<201>	**E-209**	<202>	46.310	4,625.596		
	4,911.686		4,911.686		**Hot Water**		0.000	2,955.552		2,955.552
	4,911.686	46.310	7,581.124		Total		46.310	7,581.148		2,955.552
		46.310	4,625.596	<202>	**E-207**	<208>	39.298	3,251.960		
						<209>	7.011	17,691.805	7.011	17,691.805
	19,388.250		19,388.250		**Steam**			3,105.750		3,105.750
	19,388.250	46.310	24,013.846		Total		46.310	24,049.515	7.011	20,797.555
		39.298	3,251.960	<208>	**V-201**	<210>	39.298	690.037		
	-854.783		-854.783		**Cooling**			1,707.222		1,707.222
	-854.783	39.298	2,397.177		Total	0.000	39.298	2,397.259		1,707.222
		39.298	690.037	<210>	**E-210**	<211>	39.298	1,761.929		
	2,691.650		2,691.650		**Hot Water**			1,619.670		1,619.670
	2,691.650	39.298	3,381.687		Total		39.298	3,381.599		1,619.670
		39.298	1,761.929	<211>	**S-201**	<212>	17.775	554.915		
						<213>	21.524	1,206.992		
		39.298	1,761.929		Total		39.298	1,761.907		
		21.524	1,206.992	<213>	**E-211**	<214>	21.524	2,187.089		
	2,438.589		2,438.589		**Hot Water**			1,467.393		1,467.393
	2,438.589	21.524	3,645.581		Total		21.524	3,654.482		1,467.393
		21.524	2,187.089	<214>	**E-208**	<215>	17.512	1,479.457		
						<216>	4.011	9,991.678	4.011	9,991.678
	11,055.263		11,055.263		**Steam**			1,768.207		1,768.207
	11,055.263	21.524	13,242.352		Total		21.523	13,239.342	4.011	11,759.885
		17.512	1,479.457	<215>	**V-202**	<217>	17.512	334.532		
	-381.909		-381.909		**Cooling**			762.994		762.994
	-381.909	17.512	1,097.548		Total		17.512	1,097.526		762.994
		17.512	334.532	<217>	**E-212**	**218>**	17.512	806.909		
	1,184.786		1,184.786		**Hot Water**			712.932		712.932
	1,184.786	17.512	1,519.318		Total		17.512	1,519.841		712.932
		17.512	820.909	<218>	**S-202**	<219>	7.110	227.787		
						<220>	10.402	593.130	10.402	593.130
		17.512	820.909		Total		17.512	820.917	10.402	593.130
13.186		13.186		<223>	**E-213**	<224>	13.186	715.566		
	840.158		840.158		**Steam**		0.000	134.583		134.583
13.186	840.158	13.186	840.158		Total		13.186	850.149		134.583
		24.885	797.190	<221>	**D-201**	<222>	24.638	308.613	24.638	308.613
		13.186	715.566	<224>		<225>	13.433	1,203.150	13.433	1,203.150
		38.071	1,512.756		Total		38.071	1,511.763	38.071	1,511.763
210.054	169,702.051				**Total**				210.054	169,370.731
OUT-IN :									0.000	-331.321

HEAT & MASS BALANCE FOR STREAMS TOTAL
2005 Section 3- Ethanol Production and Purification

IN					EQUIPM. IDENTIF.	OUT				
Section		EQUIPMENT				EQUIPMENT			Section	
Mass kg/s	Heat kW	Mass kg/s	Heat kW	Stream Nr.		Stream Nr.	Mass kg/s	Heat kW	Mass kg/s	Heat kW
105.859	27,308.876	105.859	27,308.876	<120>	**E-301**	<302>	143.277			
27.016	6,733.959	27.016	6,733.959	<125>	**E-302**	<309>	146.447	30,675.581		
10.402	593.130	10.402	593.130	<220>						
		146.447		<307>						
	-5,274.623		-5,274.623		**Cooling**			-1,317.235		-1,317.235
143.277	29,361.342	289.724	29,361.342		Total		289.724	29,358.346		-1,317.235
		143.277		<302>	**BR-301**	<303>	8.730			
144.707		144.707		<330>	**V-301**	<305>	131.028			
					S-301	<328>	3.519		3.519	
					S-302	<332>	144.707		144.707	
					Reaction			-1,358.012		
	-1,810.033		-1,810.033		**Cooling**			-452.021		-452.021
144.707	-1,810.033	287.984	-1,810.033		Total		287.984	-1,810.033	148.226	-452.021
		8.730		<303>	**C-301**	<306>	15.419	1,444.504		
		4.231	2,635.803	<312>		<327>	11.314	1,313.753	11.314	1,313.753
13.773		13.773		<326>						
13.773		26.734	2,635.803		Total		26.733	2,758.257	11.314	1,313.753
		146.447	30,675.581	<309>	**C-302**	<310>	28.055	54,317.164		
					E-303	<311>	17.000	4,292.795	17.000	4,292.795
					E-304	<312>	4.231	2,635.803		
						<313>	97.161	28,189.839	97.161	28,189.839
	-3,464.589		-3,464.589		**Cooling**			1,728.562		1,728.562
	76,889.583		76,889.583		**Steam**			14,666.042		14,666.042
	73,424.994	146.447	104,100.575		Total		146.447	105,830.205	114.161	48,877.238
		28.055	54,317.164	<310>	**C-303**	<314>	11.919	1,758.314		
					E-305	<318>	16.136	3,177.372	16.136	3,177.372
					E-306					
	-8,090.014		-8,090.014		**Cooling**			48,417.933		48,417.933
	10,484.017		10,484.017		**Steam**			1,679.406		1,679.406
	2,394.003	28.055	56,711.167		Total		28.055	55,033.026	16.136	53,274.711
		11.919	1,758.314	<314>	**C-304**	<317>	11.039		11.039	
0.012		0.012		<323>	**E-307**	<334>	0.892	-92.218	0.892	-92.218
					E-308					
					E-309					
					E-310					
	-223.692		-223.692		**Cooling**			446.902		446.902
	-350.606		-350.606		**Cooling**			1,749.248		1,749.248
			-1,107.000		**Refrigeration**					
	3,154.937		3,154.937		**Heating**			1,059.659		1,059.659
	2,110.000		2,110.000		**Adsorbtion**					
					Evaporation			2,095.000		2,095.000
0.012	4,690.639	11.931	5,341.953		Total		11.931	5,258.591	11.931	5,258.591
301.769	108,060.945				**Total**				301.768	106,955.037
OUT-IN :									0.000	-1,105.909

A.7.2. 2015 Design Mass and Heat Balances

HEAT & MASS BALANCE FOR STREAMS TOTAL
2015 Section 1- Mill and Clarification

IN						OUT				
Section		EQUIPMENT			EQUIPM. IDENTIF.	EQUIPMENT			Section	
Mass kg/s	Heat kW	Mass kg/s	Heat kW	Stream Nr.		Stream Nr.	Mass kg/s	Heat kW	Mass kg/s	Heat kW
332.589		332.589		<101>	**M101**	<103>	330.673	15,817.285		
93.125	19,351.310	93.125	19,351.310	<102>		<138>	95.040	3,534.100	95.040	3,534.100
425.714	19,351.310	425.714	19,351.310		Total		425.713	19,351.385	95.040	3,534.100
		214.937	10,281.182	<104>	**F-102&C-101**	<107>	215.132	11,566.290		
0.930		0.930		<105>		<135>	0.843	34.442	0.843	34.442
0.108		0.108		<106>						
	1,002.300		1,002.300		**Combustion**					
1.038	1,002.300	215.975	11,283.482		Total		215.975	11,600.732	0.843	34.442
1.494		1.494		<127>	**F-101**	<130>	1.945	1,017.022	1.945	1,017.022
0.052		0.052		<128>		<131>	0.526	402.243		
0.925		0.925		<129>						
	2,073.465		2,073.465		**Combustion**					
					Reaction			654.200		
2.471	2,073.465	2.471	2,073.465		Total		2.471	2,073.465	1.945	1,017.022
		3.526	879.015	<126>	**V-101**	<133>	0.464	99.240		
		0.526	402.243	<131>		<134>	0.863	180.923		
						<141>	2.725	573.883	2.725	573.883
	-142.560		-142.560		**Cooling**			284.813		284.813
	-142.560	4.052	1,138.698		Total		4.052	1,138.859	2.725	858.696
		215.132	11,566.290	<107>	**V-102**	<108>	215.995	11,743.193		
		0.863	180.923	<134>						
		215.995	11,747.213		Total		215.995	11,743.193		
		115.735	5,536.747	<115>	**V-103**	<116>	116.200	5,635.987		
		0.464	99.240	<133>						
		116.199	5,635.987		Total		116.200	5,635.987		
		215.995	11,743.193	<108>	**E-101**	<109>	215.995	58,792.743		
	56,125.180		56,125.180		**Steam**			9,071.892		9,071.892
	56,125.180	215.995	67,868.373		Total		215.995	67,864.635		9,071.892
		116.200	5,635.987	<116>	**E-102**	<117>	116.200	31,266.813		
	30,519.691		30,519.691		**Steam**			4,888.865		4,888.865
	30,519.691	116.200	36,155.678		Total		116.200	36,155.678		4,888.865
		215.995	58,792.743	<109>	**V-104**	<110>	214.959	56,456.817		
						<111>	1.036	2,341.360	1.036	2,341.360
		215.995	58,792.743		Total		215.995	58,798.177	1.036	2,341.360
		116.200	31,266.813	<117>	**V-105**	<118>	115.604	30,005.342		
						<119>	0.558	1,261.080	0.558	1,261.080
		116.200	31,266.813		Total		116.162	31,266.422	0.558	1,261.080
		214.959	56,456.817	<110>	**S-101**	<113>	18.156	3,715.307		
0.135		0.135		<136>		<114>	196.938	52,740.715	196.938	52,740.715
0.135		215.094	56,456.817		Total		215.094	56,456.022	196.938	52,740.715
		115.604	30,005.342	<118>	**S-102**	<120>	105.894	27,389.177	105.894	27,389.177
						<121>	9.749	2,616.739		
		115.604	30,005.342		Total		115.643	30,005.916	105.894	27,389.177
		18.156	3,715.307	<113>	**S-103**	<123>	15.294	2,543.688	15.294	2,543.688
		9.749	2,616.739	<121>		<125>	24.640	6,141.921	24.640	6,141.921
15.556	3,232.454	15.556	3,232.454	<137>		<126>	3.526	879.015		
15.556	3,232.454	43.461	9,564.500		Total		43.460	9,564.624	39.934	8,685.609
444.914	112,161.840				**Total**				444.913	111,822.958
OUT-IN :									0.000	-338.881

HEAT & MASS BALANCE FOR STREAMS TOTAL

2015 Section 2- Juice Evaporation and Crystallization

IN						OUT				
Section		EQUIPMENT			EQUIPM. IDENTIF.	EQUIPMENT			Section	
Mass kg/s	Heat kW	Mass kg/s	Heat kW	Stream Nr.		Stream Nr.	Mass kg/s	Heat kW	Mass kg/s	Heat kW
196.938	52,740.715	196.938	52,740.715	<114>	**E-201, E-202**	<201>	46.317	2,602.598		
					E-203,E-204	<203>	20.274	7,962.234	20.274	7,962.234
					E-205,E-206	<204>	22.141	8,113.791	22.141	8,113.791
						<205>	24.039	8,028.356	24.039	8,028.356
						<206>	25.952	7,577.314	25.952	7,577.314
						<207>	27.940	8,157.815	27.940	8,157.815
						<228>	30.275	74,184.447	30.275	74,184.447
	76,075.646		76,075.646		**Steam**			12,187.941		12,187.941
196.938	128,816.361	196.938	128,816.361		Total		196.938	128,814.496	150.621	126,211.898
		46.317	2,602.598	<201>	**E-209**	<202>	46.317	4,559.340		
	4,911.686		4,911.686		**Hot Water**			2,955.552		2,955.552
	4,911.686	46.317	7,514.284		Total		46.317	7,514.892		2,955.552
		46.317	4,559.340	<202>	**E-207**	<208>	39.304	3,144.183		
						<209>	7.014	17,697.608	7.014	17,697.608
	19,388.250		19,388.250		**Steam**			3,105.750		3,105.750
	19,388.250	46.317	23,947.590		Total		46.318	23,947.541	7.014	20,803.358
		39.304	3,144.183	<208>	**V-201**	<210>	39.304	583.037		
	-854.783		-854.783		**Cooling**			1,707.222		1,707.222
	-854.783	39.304	2,289.400		Total	0.000	39.304	2,290.259		1,707.222
		39.304	583.037	<210>	**E-210**	<211>	39.304	1,654.929		
	2,691.650		2,691.650		**Hot Water**			1,619.670		1,619.670
	2,691.650	39.304	3,274.687		Total		39.304	3,274.599		1,619.670
		39.304	1,654.929	<211>	**S-201**	<212>	17.775	454.915		
						<213>	21.529	1,199.873		
		39.304	1,654.929		Total		39.304	1,654.788		
		21.529	1,199.873	<213>	**E-211**	<214>	21.529	2,171.089		
	2,438.589		2,438.589		**Hot Water**			1,467.393		1,467.393
	2,438.589	21.529	3,638.462		Total		21.529	3,638.482		1,467.393
		21.529	2,171.089	<214>	**E-208**	<215>	17.516	1,361.968		
						<216>	4.013	10,095.968	4.011	10,095.968
	11,055.263		11,055.263		**Steam**			1,768.207		1,768.207
	11,055.263	21.529	13,226.352		Total		21.529	13,226.143	4.011	11,864.175
		17.516	1,361.968	<215>	**V-202**	<217>	17.516	217.053		
	-381.909		-381.909		**Cooling**			762.994		762.994
	-381.909	17.516	980.059		Total		17.516	980.047		762.994
		17.516	217.053	<217>	**E-212**	218>	17.516	695.211		
	1,191.843		1,191.843		**Hot Water**			712.932		712.932
	1,191.843	17.516	1,408.896		Total		17.516	1,408.143		712.932
		17.516	695.211	<218>	**S-202**	<219>	7.110	197.787		
						<220>	10.406	497.512	10.406	497.512
		17.516	695.211		Total		17.516	695.299	10.406	497.512
13.191		13.191		<223>	**E-213**	<224>	13.191	602.698		
	718.768		718.768		**Steam**			115.713		134.583
13.191	718.768	13.191	718.768		Total		13.191	718.411		134.583
		24.885	652.702	<221>	**D-201**	<222>	24.639	208.492	24.639	208.492
		13.191	602.698	<224>		<225>	13.437	1,046.907	13.437	1,046.907
		38.076	1,255.399		Total		38.076	1,255.399	38.076	1,255.399
210.129	169,975.718				**Total**				210.128	169,992.687
OUT-IN :									0.000	16.969

HEAT & MASS BALANCE FOR STREAMS TOTAL
2015 Section 3- Ethanol Production and Purification

IN					EQUIPM. IDENTIF.	OUT				
Section		EQUIPMENT				EQUIPMENT			Section	
Mass kg/s	Heat kW	Mass kg/s	Heat kW	Stream Nr.		Stream Nr.	Mass kg/s	Heat kW	Mass kg/s	Heat kW
105.894	27,389.177	105.894	27,389.177	<120>	**E-301**	<302>	291.488			
24.640	6,141.921	24.640	6,141.921	<125>						
10.406	497.512	10.406	497.512	<220>						
150.549	17,154.460	150.549	18,697.566	<423>						
	-15,583.130		-15,583.130		**Cooling**			37,643.870		37,643.870
291.489	35,599.940	291.489	37,143.046		Total		291.488	37,643.870		37,643.870
		291.488		<302>	**BR-301**	<303>	21.845	1,381.589		
348.390		348.390		<330>	**V-301**	<306>	261.311			
					S-301	<328>	8.333		8.333	
					S-302	<332>	348.390		348.390	
	4,638.300		4,638.300		**Reaction**					
	-4,339.439		-4,339.439		**Cooling**			-1,083.691		-1,083.691
348.390	298.861	639.878	298.861		Total		639.879	297.898	356.723	-1,083.691
		21.845	1,381.589	<303>	**C-301**	<307>	33.866	2,928.190		
		8.562	4,571.397	<312>		<327>	26.924	3,024.700	26.924	3,024.700
30.385		30.385		<326>						
30.385		60.792	5,952.986		Total		60.790	5,952.890	26.924	3,024.700
		261.311	0.000	<306>	**C-302**	<310>	65.279	128,755.983		
		33.866	2,928.190	<307>	**E-302**	<311>	18.563	4,548.984	18.563	4,548.984
	69,145.909		69,145.909	<319>	**E-303**	<312>	8.562	4,571.397		
					E-304	<313>	202.773	58,843.386	202.773	58,843.386
						<335>		10,748.725		10,748.725
	-24,509.710		-24,509.710		**Cooling**			12,228.449		12,228.449
	197,301.945		197,301.945		**Steam**			31,763.195		31,763.195
	241,938.144	295.177	244,866.334		Total		295.177	251,460.119	221.336	118,132.739
		65.279	128,755.983	<310>	**C-303**	<314>	28.510	4,215.140		
					E-305	<318>	36.768	10,654.800	36.768	10,654.800
					E-306					
	-19,405.020		-19,405.020		**Cooling**			116,137.142		116,137.142
	33,135.226		33,135.226		**Steam**			5,534.365		5,534.365
	13,730.206	65.279	142,486.189		Total		65.278	136,541.447	36.768	132,326.307
		28.510	4,215.140	<314>	**C-304**	<317>	26.399		26.399	
0.023		0.023		<323>	**E-307**	<334>	2.134	-221.039	2.134	-92.218
					E-308					
					E-309					
					E-310					
	-534.312		-534.312		**Cooling**			1,067.471		1,067.471
	-780.026		-780.026		**Cooling**			1,558.369		1,558.369
	-837.977		-837.977		**Cooling**			4,180.848		4,180.848
	-1,977.000		-1,977.000		**Refrigeration**					
	6,289.000		6,289.000		**Heating**			1,012.488		1,012.488
	5,040.000		5,040.000		**Adsorbtion**					
					Evaporation			4,119.000		
0.023	7,199.685	28.533	11,414.825		Total		28.533	11,717.137	28.533	7,726.958
670.287	298,766.836				**Total**				670.284	297,770.883
OUT-IN :										-995.953

HEAT & MASS BALANCE FOR STREAMS TOTAL
2015 Section 4- Pretreatment and Hydrolysis

IN						OUT				
Section		EQUIPMENT			EQUIPM. IDENTIF.	EQUIPMENT			Section	
Mass kg/s	Heat kW	Mass kg/s	Heat kW	Stream Nr.		Stream Nr.	Mass kg/s	Heat kW	Mass kg/s	Heat kW
95.040	3,534.100	95.040	3,534.100	<138>	**R-401**	<405>	24.432	52,614.240	24.432	52,614.240
5.853	16,290.000	5.853	16,290.000	<401>	**V-401**	<406>	143.847	36,964.364		
47.197	13,566.495	47.197	13,566.495	<402>						
20.188	56,187.530	20.188	56,187.530	<403>						
168.278	89,578.125	168.278	89,578.125		Total		168.278	89,578.604	24.432	52,614.240
		143.847	36,964.364	<406>	**S-401**	<407>	143.826	32,127.397		
		56.714	4,143.122	<418>		<421>	56.734	8,979.928		
0.000	0.000	200.561	41,107.486		Total		200.560	41,107.325	0.000	0.000
		143.826	32,127.397	<407>	**E-401**	<408>	143.826	13,967.487		
	-6,057.659		-6,057.659		**Cooling**			12,102.251		12,102.251
0.000	-6,057.659	143.826	26,069.738		Total		143.826	26,069.738	0.000	12,102.251
		143.826	13,967.487	<408>	**V-402**	<413>	143.587	14,348.586		
2.725	573.883	2.725	573.883	<141>	**V-403**	<414>	2.965	192.784		
					S-402					
2.725	573.883	146.551	14,541.370		Total		146.552	14,541.370	0.000	0.000
		2.965	192.784	<414>	**S-403**	<415>	2.282	118.461		
2.011		2.011		<426>		<416>	2.694	73.983		
2.011	0.000	4.976	192.784		Total		4.976	192.444	0.000	0.000
		89.155	10,323.975	<419>	**E-402**	<420>	89.155	11,711.461		
	1,670.983		1,670.983		**Steam**			283.497		283.497
	1,670.983	89.155	11,994.958		Total		89.155	11,994.958	0.000	283.497
		89.155	11,711.461	<420>	**BR-401**	<423>	150.549	18,697.566	150.549	18,697.566
		56.734	8,979.928	<421>	**S-404**	<424>	27.651	1,993.823		
31.944		31.944		<427>						
0.366		0.366		<437>						
32.310	0.000	178.199	20,691.389		Total		178.200	20,691.389	150.549	18,697.566
18.563	4,548.984	18.563	4,548.984	<311>	**E-402**	<429>	8.623	21,892.809	8.623	21,892.809
						<430>	9.938	2,135.463		
	23,216.222		23,216.222		**Steam**			3,737.527		3,737.527
18.563	27,765.206	18.563	27,765.206		Total		18.561	27,765.799	8.623	25,630.336
15.294	2,543.688	15.294	2,543.688	<123>	**D-401**	<432>	27.697	3,056.300		
		2.694	73.983	<416>		<436>	278.850	141,177.000	278.850	141,177.000
		27.651	1,993.823	<424>						
8.333		8.333		<328>						
		252.575	139,488.800	<435>						
23.627	2,543.688	306.547	144,100.295		Total		306.547	144,233.300	278.850	141,177.000
		9.938	2,135.463	<430>	**F-401**	<433>	6.366	6,406.900	6.366	6,406.900
		27.697	3,056.300	<432>		<435>	252.575	139,488.800		
221.306		221.306		<434>						
	495,118.500		495,118.500		**Combustion**					
					Steam			354,365.900		354,365.900
221.306	495,118.500	258.941	500,310.263		Total		258.941	500,261.600	6.366	360,772.800
468.820	611,192.726				**Total**				468.819	611,277.689
OUT-IN :										84.964

A.8. Process and Equipment Design: 2005 Plant

Tools used in this process are Super Pro Designer simulation program v5.1, ASPEN Plus 11.1, and MS Excel spread sheet calculation. Most of the equipment design is carried out in MS Excel and ASPEN Plus and the parameters obtained from these calculations are implemented into Super Pro Designer to simulate the process.

A.8.1. Lime Kiln (F-101)

The factors that are to be considered during the calculation of bagasse requirement for the furnace is

- Heat required for the reaction
- Heat required to heat up the inflowing air to furnace temperature
- Heat required to heat up the $CaCO_3$ to furnace temperature
- Heat required to heat up the bagasse to furnace temperature
- Heat leaving furnace with exhaust gas
- Heat leaving furnace with CaO and ash

The flow rates of each component and their heat capacities are given in Table A.8.1

Table A.8.1 The calculations for bagasse requirement for F-101

Component	Flow rate (kg/s)	Heat Capacity (kj/kg ºC)	Initial Temp. (ºC)	Final Temp. (ºC)	Total Heat (kJ/s)
$CaCO_3$	0.303	0.824+0.00050T	30	888	273.9
$CaCO_3$	0.303	0.824+0.00050T	888	898	3.9
Fresh Air	8.53β	0.993+0.00013T	30	888	$7,704.2\beta$
Fresh Air	8.53β	0.993+0.00013T	888	898	94.6β
Bagasse	β	1.34	108	898	$1,058.6\beta$
CaO+ash	$0.170+0.0269\beta$	0.747+0.00036T	898		$133.6+21.1\beta$
Exhaust air	$0.133+9.5031\beta$	1.037+0.00017T	898		$127.4+9,105.4\beta$
Reaction	0.170	3855 kJ/kg	898	898	654.2
Bagasse Burning	β	16540kj/kg	898	898	$+16,540\beta$

*heat capacities of the fresh air and exhaust gas are obtained by using the individual heat capacities of the gas stream component and their mass compositions.

The flow rates are calculated using the exhaust gas yields and theoretical air requirements of bagasse given in Section 5.1.2 β is the required amount of bagasse. The fresh air can be heated up using the exhaust gas, but, using 10ºC minimum temperature difference the fresh air can be heated up to 888ºC by using the exhaust gas of 898ºC. Therefore, the heat duty to heat up calcium carbonate and bagasse from 30ºC to 898ºC , heat required to heat up fresh air from 888ºC to 898ºC and the heat required for the reaction are supplied by burning bagasse. Writing the energy balance,

$$Bagasse(108 \rightarrow 898^{\circ}C) + CaCO_3(30 \rightarrow 898^{\circ}C) + Fresh\,air(888 \rightarrow 898^{\circ}C) + \Delta H_{reaction} = \Delta H_{Bag.Comb.}$$

$$1,058.6\beta + 273.88 + 3.85 + 94.61\beta + 663.06 = 16,540\beta$$

The amount of bagasse required to supply the necessary heat is found as 222 kg 12% moisture bagasse/h. Using this value, the heat requirement to heat up the fresh feed air is obtained as 470 kJ/s. This is much less then the heat supplied by exhaust gas (682.7 kJ/s) when it is cooled down to 40ºC Therefore, the exhaust gas will be cooled down to 320 ºC by heating up the feed air. The area required for heat exchanging is calculated using overall heat transfer coefficient of 75 W m^{-2} ºC^{-1} which is the average value of 50-100 W m^{-2} ºC^{-1} reported by Sinnot (1999) for air cooling of gasses.

To determine the total heating area requirement following equations are used,

$$Q = AU\Delta T_{LM} \quad \text{Eqn A.8.1}$$

A is heat exchanger area (m^2), *U* is the overall heat transfer coefficient (W m^{-2} ºC^{-1}), *Q* is power required or generated (kW), ΔT_{LM} is the log mean temperature difference for counter current flow (ºC) which is calculated based on equation,

$$\Delta T_{LM} = \frac{(T_{hot,in} - T_{cold,out}) - (T_{hot,out} - T_{cold,in})}{\ln\left[\frac{(T_{hot,in} - T_{cold,out})}{(T_{hot,out} - T_{cold,in})}\right]} \quad \text{Eqn A.8.2}$$

Using the inlet and outlet temperatures of fresh air and exhaust gas, the log mean temperature difference is calculated as 83ºC. Using the total transferred heat of 470 kW, the resultant area requirement is found as 75.5 m^2.

A.8.2. Sulphur Furnace (F-102)

To check whether any heat integration is required for the incoming and outgoing streams the same methodology is applied with the lime kiln.

Table A.8.2 Heat balance calculations for sulphur burner F-102

Component	Flow rate (kg/s)	Heat Capacity (kj/kg ºC)	Initial Temp. (ºC)	Final Temp. (ºC)	Total Heat (kJ/s)
Sulphur	0.108	0.475+0.00084T	30	363	23.0
Fresh Air	0.929	0.993+0.00013T	30	353	305.4
Fresh Air	0.929	0.993+0.00013T	353	363	9.7
Exhaust air	1.037	0.883+0.00147T	363	741	664.2
Reaction	0.108	9,280.97 kJ/kg	363	363	1,002.3

As the results indicate, the heat generated by the reaction is enough to heat up the inflowing sulphur and fresh air to combustion temperature. Therefore, air and sulphur can be directly sent to furnace. After heating, there will be an excess heat of 664.23 kJ/s. This heat will leave the furnace with exhaust air at 741ºC. The exhaust air does not need to be cooled because it is going to get in contact with juice in sulphur dioxide absorption column and it heats up the juice.

A.8.3. Sulphitation Tower (C-101)

The Equipment is simulated in Aspen Plus as an absorption column to find the required vessel dimensions. Radfrac distillation column model is used without condenser and reboiler. The number of stages is determined by manipulating the number of stages to get the required SO$_2$ absorption. The packing is selected based on the system requirements. Raschig rings packings are much more chipper when compared to pall ring and saddles. But, the efficiency is lower. Since sulphur dioxide is highly soluble in water, the effect of efficiency will not be as high. For corrosive non-alkaline fluid the best packing material is ceramic material. Therefore, 50 mm ceramic raschig rings will be used as packing. The properties of raschig rings are given in Table A.8.3

Table A.8.3 Properties of column packings

Property	Value
Size (mm)	50
Bulk Density (kg/m^3)	651
Surface area (m^2/ m^3)	95
Packing Factor (m^{-1})	210

The height equivalent of each stage is taken as 0.75m. For the simulation PRWS property method is applied as recommended by Aspen Plus user manual and 99.999% of SO_2 is absorbed in 3 stages with 0.48 Murphree efficiencies. Since the solubility of SO_2 in water is high, operation at atmospheric pressure is sufficient. The ASPEN simulation results gives the column diameter as 2.76 m and the required column height is 2.25 m. Based on this information, the required amount of packing is obtained as 8.8 ton.

A.8.4. Lime Milk Preparation Vessels (V-101)

A.8.4.1. Cooling

Since the filtrate from the filter and CaO have 93ºC and 898ºC temperature respectively, the contents of the vessel are to be cooled to 90 ºC prevent boiling of the lime milk. The heat capacity of CaO and juice are given in following equations

$$C_{p,CaO}=0.748+0.0004T \text{ kJ kg}^{-1}{}^{o}C^{-1} \quad \text{Eqn A.8.3}$$

$$C_{p,filtrate}=4.1868-0.0251B-0.0006T \text{ kJ kg}^{-1}{}^{o}C^{-1} \quad \text{Eqn A.8.4}$$

The CaO and filtrate flow in the vessel are 598 and 4,126 kg/h respectively. Using the flow rates, inlet temperatures and heat capacities, the heat duty of the system is obtained as 142 kW. Cooling water at 25 ºC is used as cooling agent and the exit temperature of the cooling water is set to 40 ºC. The mean heat capacity of cooling water at this temperature interval is 4.1774 kJ kg^{-1}ºC^{-1} which yields cooling water requirement of 8.16 t/h.

The vessel will be cooled with jacket. The over all heat transfer coefficients for jacketed vessels are given in the interval 200-500 W m^{-2}ºC^{-1} for water-dilute aqueous solutions system. 400 W m^{-2}ºC^{-1} which is close to upper interval is selected as design value. The LMTD is calculated as 57.2 ºC and the required jacket area is calculated as 6.2 m^2.

A.8.5. Rotary Drum Dryer (D-101)

The sugar crystals are dried by counter current hot air in inclined horizontal rotary driers. The amount of the air required to dry sugar is calculated using the equation,

$$W_A = \frac{W_S h}{H_1 - H_0} \quad \text{Eqn.A.8.5}$$

W_A is required air (kg/h), W_S is the weight of sugar to be dried (kg/h), h is the moisture fraction in sugar and H_0 and $H_1$are the weight of water vapour contained in saturated air at inlet and outlet temperatures respectively (kg/kg air). The entry temperature for the air in to drier is 85ºC and the exit temperature is 40ºC. The corresponding saturated air moisture content is given as 55.8 g/kg air at 40ºC. The moisture content at 85ºC is taken as the saturated air moisture content at 30 ºC (20.24 g/kg air) because the air is heated from 30ºC to 85ºC without any moisture addition. The total water content in the sugar crystals is 887 kg/h and 98% of moisture will be removed. Therefore, the required air amount is calculated to be 23 t/h. It should be noted that this is the minimum amount of air required.

The air heater consists of tubes with or without fins which are supplied with steam. The heating surface requirement for the heat exchanger is calculated using the equation,

$$A = \frac{Q}{U \Delta T_{LM}} \qquad \text{Eqn A.8.6}$$

A is the heating surface (m^2), *Q* is the heat transfer rate (*W*), *U* is the overall heat transfer coefficient with value 200 W m^{-2} ºC^{-1} (the given interval is 30-300 W m^{-2} ºC^{-1} for steam gas systems) and *LMTD* is log mean temperature difference (ºC) . The temperature of the air leaving the heaters is in the order of 70-95ºC. Value of 85ºC is used for the design.

Q is calculated based the heat required to evaporate given amount of water in the dryer. It can be calculated using the following equations,

$$Q = 0.98 W_S h L_{v,T} \qquad \text{Eqn A.8.7}$$

$L_{v,T}$ is the latent heat of evaporation at temperature *T* in kj/kg and, for water, it is calculated using the equation the coefficient 0.98 is coming from the percentage of water evaporation. Watson Equation is used to calculate the latent heat of evaporation at specified temperature.

$$L_{v,T} = L_{v,B} \left[\frac{T_c - T}{T_c - T_b} \right]^{0.38} \qquad \text{Eqn A.8.8}$$

T_c is the critical temperature of water (647.3 K), T_b is the normal boiling temperature of water (373.2 K) and $L_{v,B}$ is the latent heat at normal boiling temperature (2,260 kJ/kg). Using an average drier temperature of 50ºC the latent heat of water is calculated as 2,408.8 kJ/kg. And, the total heat required to evaporate the moisture content of sugar is calculated as 593 kW. The amount of air required to supply this much heat is calculated as 47.5 ton/h using,

$$W_A = \frac{Q_A}{c_A (T_1 - T_0)} \qquad \text{Eqn A.8.9}$$

c_A is the average heat capacity of air in given interval 1 kj/kg ºC.

Low pressure saturated steam at 130.7 ºC is used as heating agent. Using the value of 593 kW and *LMTD* of 73 ºC, the required heating area is calculated as, 40.6 m^2. The required amount for steam is calculated as 0.98 t/h using the enthalpy of 2,171 kJ/kg steam.

Since the amount of air required to supply the heat of evaporation (47.5 t/h) is higher then the amount calculated to remove the moisture (23 t/h), the higher result flow rate will be taken as basis.

The diameter of the dryer drum is related to the quantity of air required for drying. The length of the dryer drum is related to the time required to affect the transfer of heat from the drying air stream to the material being processed and the time required to affect the transfer of the mass of water evaporated from the material to the drying air stream.

The evaporation rates of indirect heated rotary drum driers are given as an average value of 65 kg $m^{-2}h^{-1}$(IRef-2). Using this value the required evaporation area for moisture removal is calculated as 14 m^2.

The air flow velocity in the driers is to be in the order of 2-4.9 m/s (Hugot, 1972). For air flow rate of 47.5 t/h (9.43 m^3/s), to obtain air flow velocity of 4 m/s, the required cross sectional area is calculated as 2.4 m^2. Thus, the diameter of the drum should be 1.73 m.

The power requirements for drum driers evaporating 500 and 1,200 kg water/h are given as 7.5 and 15 kW respectively (IRef-3). For current capacity of 850 kg/h the value of 13 kW is taken as the power consumption.

A.8.6. Fermentation

A.8.6.1. Juice Coolers (E-301&E-302)

The juice from clarifiers, filtrate from filters and molasses from crystallizers are at 100 ºC, 93ºC and 55 ºC respectively and their mixed temperature is 96 ºC, juice is to be cooled to fermentation temperature of 30ºC. For this, the hot juice is contacted with the beer column feed to heat up the feed to 80ºC before distillation. During this operation the temperature of the juice is dropped to 45ºC. Later on, by utilizing another heat exchanger the temperature is dropped to fermentation temperature of 30ºC by using chilled water.

<u>*First Cooler (E-301)*</u>
The dissolved solid concentration of the juice entering the coolers is 17.8%. Using Equation A.8.10, the heat capacity of the juice is obtained as 3.74 kJ kg^{-1}ºC^{-1}.

$$C_{p,Juice} = 4.1868 - 0.0251B \text{ kJ kg}^{-1}{}^{\circ}\text{C}^{-1} \qquad \text{Eqn A.8.10}$$

The heat capacity curve of ethanol is obtained from ASPEN Plus as given in Figure A.8.1. For simplicity the heat capacity of ethanol is assumed as linear and the obtained equation is as following,

$$C_{p(ethanol)} = 2.576 + 0.00842T \text{ kJ kg}^{-1}{}^{\circ}\text{C}^{-1} \qquad \text{Eqn .A.8.11}$$

$$C_{p(water)} = 4.189 - 0.0006T \text{ kJ kg}^{-1}{}^{\circ}\text{C}^{-1} \qquad \text{Eqn .A.8.12}$$

Therefore, the specific heat capacity of 9% ethanol solution is obtained as

$$C_{p(broth)} = 4.044 - 0.0002T \text{ kJ kg}^{-1}{}^{\circ}\text{C}^{-1}$$

The flow rate of fermentation broth and ethanol solution from stripping column is 538.7 t/h and the temperature of the mixture is 32.3 ºC (from ASPEN Plus). 28,785 kW power increases the final temperature of the mixture to 80ºC.

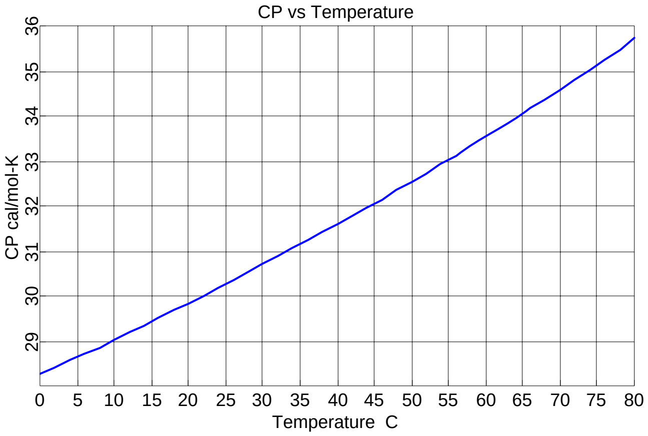


Figure A.8.1 Ethanol heat capacity curve

The mass flow rate of hot juice into the coolers is 516 t/h. Using the heat capacity and the heat duty of the system, the exit temperature of the hot juice is obtained as 42.3 ºC.

The overall heat transfer coefficient for heat exchangers for similar systems is given between 800-1,500 W $m^{-2o}C^{-1}$. A value of 1,400 W $m^{-2o}C^{-1}$ close to the upper bound is selected as the design value. The *LMTD* is calculated as 12.8ºC. Using this parameters and the heat duty the required heat transfer area is calculated as 1,606 m^2.

Second Cooler (E-302)

In the second cooler the sugar solution is cooled from 42.3ºC to fermentation temperature of 30ºC. Chilled water at 10ºC is used as cooling agent. The exit temperature of chilled water is set to 35ºC. Using the heat capacity information given the extracted heat is obtained as 6,593 kW. The amount of chilled water required for cooling is calculated as 227.3 t/h. The overall heat transfer coefficients of brine-water cooling systems are given as 600-1.200 W $m^{-2o}C^{-1}$. Value of 1,100 W $m^{-2o}C^{-1}$ is assumed. The *LMTD* is calculated as 12.6 ºC. Based on this information the heat transfer area requirement is calculated as 476 m^2.

A.8.7. Fermentors (BR-301)

A.8.7.1 Calculation of Biomass Yield on Substrate

The percentage of the substrate that is distributed between biomass and ethanol production is arranged in such a way that the total percentage of sugar utilized for ethanol production is 92% of total. And, the percentage of the substrate to biomass is determined in such a way that the bleed is 5% of the total yeast inside the fermentor which corresponds to yield value of Y_{SX} = 0.03 g biomass/gram glucose. This yield value is found using: Sucrose inflow concentration of 166.04 kg/m^3 which yields 174.78 kg/m^3 glucose. Assuming fermentor volume of V the amount of glucose in the fermentor is 174.78V kg. Initial biomass concentration is 100 kg/m^3 and the initial biomass amount is 100V kg which is equal to 95% of the outgoing biomass. Therefore, the excess biomass produced in one batch is obtained as 5.244V kg. Dividing this value with the initial glucose amount yields 0.03 g biomass/g glucose.

A.8.7.2. Model

For simplicity in the model the by-products are excluded and it is assumed that all the glucose is converted into ethanol and biomass. As a result, the reactions for the model are,

$$0.995C_6H_{12}O_6 + 0.2NH_3 \rightarrow CH_{1.8}O_{0.5}N_{0.2} + 1.64C_2H_6O + 1.69CO_2 + 0.45H_2O \qquad \text{Eqn A.8.13}$$

$$C_6H_{12}O_6 \rightarrow 2C_2H_6O + 2CO_2 \qquad \text{Eqn A.8.14}$$

The specific rates of the micro organisms are required to estimate the microbial activity of the micro organisms. Specific glucose uptake (q_s) rate is calculated using,

$$q_s = q_s^{\max} \frac{C_s}{K_s + C_s} \qquad \text{Eqn A.8.15}$$

Specific ethanol productivity (in g ethanol/(g biomass h)) is determined based on growth and ethanol production reactions as,

$$q_E = 3.0667\mu + 0.51m_s \qquad \text{Eqn A.8.16}$$

m_s is the maintenance coefficient and also the specific rate of catabolism reaction in g g biomass^{-1} h^{-1} and for anaerobic growth of micro organism on glucose it is given as 0.04 g (g biomass h)$^{-1}$. Since m_s represents the specific rate of glucose, it is multiplied by the theoretical yield of ethanol in catabolism reaction.

Specific growth rate of the micro organism is determined by using Herbert-Pirt equation which is defined as,

$$q_s = \frac{\mu}{Y_{SX}^{\max}} + m_s \qquad \text{Eqn A.8.17}$$

When this equation is rearranged and the equation for substrate uptake and ethanol productivity is inserted into equation following specific growth rate expression is obtained,

$$\mu = Y_{SX}^{\max} q_s^{\max} \frac{C_s}{K_s + C_s} - Y_{SX}^{\max} m_s \qquad \text{Eqn A.8.18}$$

Where, μ is the specific growth rate in h^{-1}, $q_s^{\max}$ is the maximum specific substrate uptake rate of the microorganism in g (g biomass h)$^{-1}$, K_s is the substrate affinity of the microorganism in g/l, and $Y_{SX}^{\max}$ is the maximum biomass yield on substrate. In Biochemical Engineering Handbook (Atkinson and Mavituna,1991), the K_s for anaerobic growth of yeast is given as 0.025 g/l. Using the growth reaction, $Y_{SX}^{\max}$ is obtained as 0.1374 g biomass/g substrate. $q_s^{\max}$ can be estimated by using,

$$q_s^{\max} = \frac{\mu^{\max}}{Y_{SX}^{\max}} + m_s \qquad \text{Eqn A.8.19}$$

The mass balances to be solved for batch fermentation are,

For glucose uptake , $\frac{dC_s}{dt} = -q_s C_x$ Eqn A.8.20

For growth, $\frac{dC_x}{dt} = \mu C_x$ Eqn A.8.21

For ethanol, $\frac{dC_E}{dt} = q_E C_x$ Eqn A.8.22

The initial conditions are $t = 0$ $C_S = C_{S0}$, $C_X = C_{X0}$, $C_E = 0$, concentrations are obtained from the super pro designer mass balances.

A.8.7.3 Parameter Estimation and Fermentation Time

The parameters μ_{max} and m_s are estimated using the yield values on substrate. The known design parameters used in calculations are given in Table A.8.4.

Table A.8.4 List of know parameters for fermentation modelling

Parameter	Value
Inflowing sucrose concentration	166.04 kg/m^3
Inflowing biomass concentration	100.00 kg/m^3
Biomass yield on substrate	0.03 g/g
Fermentation time	4-12 h
Selected Fermentation time	7 h (Goldemberg,1994)

Since the system is batch system, as first estimation the microorganism is assumed to grow with μ_{max}. Using this information the mass balances are integrated and the following equations are obtained

$$C_x = C_{xo} e^{\mu_{\max} t} \quad \text{Eqn A.8.23}$$

$$C_E = C_{xo}(a_1 \mu_{\max} + 0.51 m_s)\frac{(e^{\mu_{\max} t} - 1)}{\mu_{\max}} \quad \text{Eqn A.8.24}$$

$$C_{S0} = C_{xo}\left(\frac{\mu_{\max}}{Y_{SX}^{\max}} + m_s\right)\frac{(e^{\mu_{\max} t} - 1)}{\mu_{\max}} \quad \text{Eqn A.8.25}$$

From the yield values we have,

$$\frac{C_x - C_{xo}}{C_{xo}} = 0.0498, \; C_E = 81.9\, kg/m^3$$

and,

$$C_{S0} = \frac{166 g\, sucrose/l}{342\, g\, sucrose/mol\, sucrose} \times \frac{2\, mol\, glucose}{1\, mol\, sucrose} \times \frac{180\, g\, glucose}{1\, mol\, glu\cos e} = 174.74\, kg/m^3 .$$

Using this values biomass and ethanol equations are solved in MS Excel solver and μ_{max} and m_s are obtained as 0.0073 h^{-1} and 0.1799 g glucose/ (g biomass h) respectively. Then, these values are input to substrate equation to see it satisfies the result. There was a 7.4 kg/m^3 difference between the real value and simulation which corresponds to 4.3% of the total substrate. The reason for the difference is the by products formation which is not included in the model.

Then, these values are input to MATLAB program to simulate the real results. Using MATLAB, simulation results are obtained as, $(C_x - C_{xo})/C_{xo} = 0.0542$, $C_E = 85.6\, kg/m^3$ and $C_S = 0\, kg/m^3$.

The excess ethanol and biomass concentrations are due to neglecting the by product formation in the model. The resultant fermentation time is calculated as 7.5 h.

A.8.7.4. Fermentor Sizing

Determination of number of fermentors:

In first trial, volume of each fermentor is adjusted to maintain 2 h of volumetric juice flow. The scheduling is as shown in Figure A.8.2. Under these conditions the total cycle time of one fermentor is 12 h, 2 h filling, 8 h fermentation and 2 h emptying. Therefore, total 6 fermentors are required for the operation. The emptying time can be chosen shorter, but, shorter the emptying time, larger the number of centrifuges to remove the biomass and larger the flow rate which increases the pump capacity and power requirement.

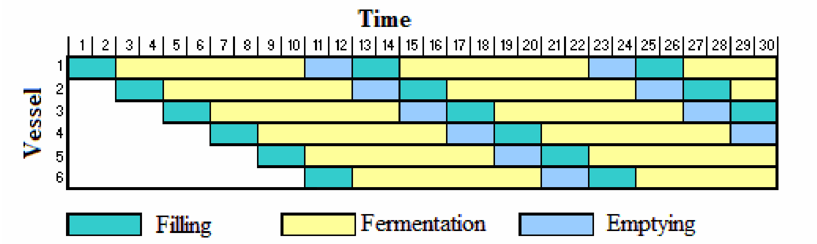


Figure A.8.2 Scheduling for 6 fermentors

Using the mass balances from Super pro designer, total volumetric flow per hour is obtained as 484.2 m^3 (density is 1,065 kg/m^3). Since there are 6 vessels, the volume of each fermentor is adjusted to maintain 2 hours of volumetric flow which corresponds to 968 m^3 of working volume.

Volume of 968 m^3 excludes volume of the initial biomass in the reactor. Using the concentration of 400 kg wet biomass/m^3 (100 kg dry biomass/m^3) and the assumed biomass density of 1.131 ton/m^3, the volume contribution of biomass to 1 m^3 broth is calculated as

$$1m^3 \times 0.400\frac{ton\ wet\ biomass}{m^3}\Big/1.131\frac{ton\ wet\ biomass}{m^3\ wet\ biomass} = 0.356m^3$$

Therefore, the final volume of each fermentor is obtained as 1442 m^3 including 10 % safety.

Using the same methodology, the required sizes of fermentors for different number of vessels and their costs are calculated and presented in Table A.8.5 for comparison.

Table A.8.5 Equipment costs for different number of fermentors

Number	Cycle Time (h)	Volume (m^3)	Total Volume (m^3)	Cost (10^6\$)	Total Cost (10^6\$)
10	10	721	7,210	0.64	6.4
7	11.2	1,154	8,076	0.89	6.3
6	12	1,442	8,653	1.05	6.3
4	16	2890	11,537	1.72	6.9
3	24	5,770	17,300	2.80	8.4

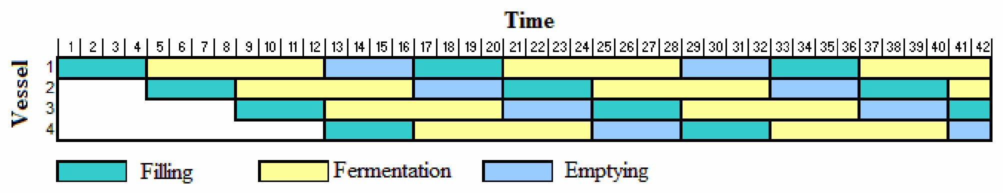


Figure A.8.3 Scheduling for 4 fermentors

As it is seen in cost analysis as the number of vessels is decreased from 10 to 3 first the investment cost is decreased till 6 vessels then it starts increasing. Therefore, the number of fermentors is taken as 6.

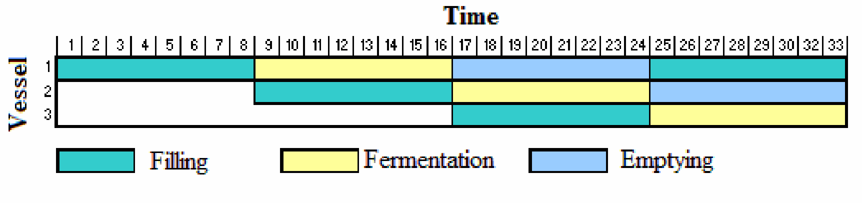


Figure A.8.4 Scheduling for 3 fermentors

Mixing Time Calculations

Important parameters in fermentor sizing are the diameter to height ratio and the mixing time. Since the batch vessels do not have stirrers the mixing due to gas generation is the main mixing effect in the vessel. Therefore, the height to diameter ratios or the vessels are determined based on the final mixing obtained. Shorter the mixing time better the mixing.

The mixing time (t_m) is calculated using,

$$t_m = 11.\frac{H}{D}(g.v_{gs}D^{-2})^{-0.33} \qquad \text{Eqn A.8.26}$$

v_{gs} is the superficial gas velocity, g is the gravitational acceleration (9.82 m s^{-2}) D is diameter (m) and H is height of the reactor (m). Superficial gas velocity is calculated using

$$v_{gs} = \frac{Volumetric\ gas\ flow rate}{cros\sec tional\ Area\ of\ vessel} \qquad \text{Eqn A.8.27}$$

The volumetric gas flow rate is obtained as 5.03 m^3/s using ASPEN simulation. For H/D ratio of 2 the superficial gas velocity is calculated as 0.06 m/s and the mixing time is obtained as 106 s with a gas hold up of 0.09. Considering the 8 hr of batch time, the mixing is sufficient.

A.8.7.5. Fermentor Cooling

As it is previously mentioned, one mol of biomass growth produces 65 kJ heat. It is required to install a cooling system to keep the fermentor temperature constant at 30ºC. the rate of biomass production is 2,530 kg/h which corresponds to 1.86 MW energy. Some of this heat is removed with the water and ethanol evaporation with the off carbon dioxide gas. The ASPEN simulation yield that 486 kg/h ethanol and 544 kg/h water is removed with the off gas. The latent heat of evaporation of ethanol and water can be found using Watson equation (Equation 4.1),

The critical temperature and latent heat at boiling point of water and ethanol are 647.3 and 516.2 K and 2,260 and 843 kJ/ kg respectively. Using these values and Watson equation, the latent heat of evaporation of water and ethanol at 30ºC are obtained as 2,464 and 929 kJ/kg respectively. Using these latent heats and ethanol and water flow rates with off gas, the heat removed due to evaporation is calculated as

$$Q_{evap} = 544\frac{kg}{h}\times 2,467\frac{kJ}{kg\ water} + 486\frac{kg}{h}\times 929\frac{kJ}{kg\ ethanol} = 1,793\frac{MJ}{h} = 0.5MW$$

As a result, the heat to be removed by cooling is obtained as,

$$Q_{Excess} = Q_{Growth} - Q_{Evaporation} = 1.86MW - 0.5MW = 1.36MW$$

The cooling agent is chilled water at 10ºC. The outlet temperature of cooling water is set to 25ºC. The heat capacity of water is given with Equation A.8.12. Using the inlet and outlet temperatures and the heat capacity the cooling water requirement is calculated as 78 t/h.

Coils or jackets can be utilized for cooling, since the fermentor media is highly viscous due to the presence of high biomass concentrations, the cooling from the walls might not be sufficient. Therefore, coils are recommended as cooling system. The overall heat transfer coefficient for coils for water-aqueous solutions systems is given in the interval 400-700 W $m^{-2o}C^{-1}$. Due to the presence of biomass, 500 W $m^{-2o}C^{-1}$ value close too lower margin is selected as the design value. The *LMTD* for the system is calculated as 11 ºC. Using these values and the heat duty the area required for cooling is calculated as 247 m^2.

A.8.8. Centrifuges (S-301)

The performance of the sedimentation centrifuges are described by sigma factor (Σ) which is equal to the cross sectional area of sedimentation tank giving the same clarification.

According to the equation that defines the throughput for a centrifuge the sigma factor can be calculated:

$$F = 2 \cdot u_g \cdot \Sigma \qquad \text{Eqn A.8.28}$$

Where:

$$u_g = \frac{\Delta\rho \cdot d_p^2 \cdot g}{18 \cdot \eta} \qquad \text{Eqn A.8.29}$$

F is Liquid flow through the centrifuge (0.14 m^3/s), u_g the terminal velocity of the particles settling under gravity through the liquid to be calculated (m/s), $\Delta\rho$ is the density difference between solid and liquid 136 kg/m^3 (1131 kg/m^3 cells density and 995 kg/m^3 water density), d_p is the diameter of the solid particle ($5x10^{-6}$ m), η is the viscosity of the liquid (10^{-3} kg/m·s) and g is the gravity constant (9.81 m/s^2).

Using this information the terminal velocity and the corresponding sigma value are calculated as 1.85 $x10^{-6}$ m/s and 37.77 $x10^4$ m^2 respectively.

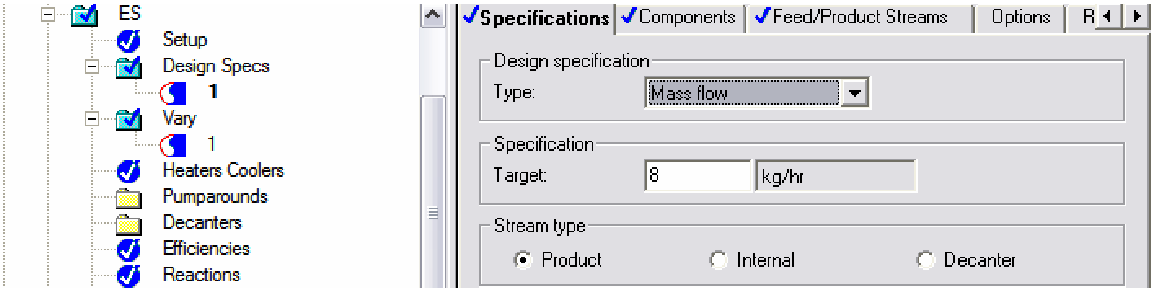
To select the best centrifuge type following parameter is required,

$$\frac{F}{\Sigma} = 3.9 \times 10^{-6} \qquad \text{Eqn A.8.30}$$

According to Sinnot (1999), for this value and the corresponding volumetric flow the disc type centrifuges are most suitable options for the current system.

A.8.9. Ethanol Scrubber (C-301)

When designing the scrubber, ASPEN plus RadFrac column is used and the property method is, UNIQUAC which is suggested by ASPEN Plus user manual for alcohol separation systems. The condenser and reboiler are specified as none. The gas effluent is feed from the bottom and the absorption water is feed from the top stage. During the calculation the built in optimization tool of ASPEN is used to manipulate the flow rate of feed absorption water to get the specified ethanol amount of 8 kg/h in gas effluent (36.4 ton/year maximum allowable organic material). To do this, in "Design specifications" selection in menu (see Figure A.8.5) mass flow of ethanol in gas stream is set to 8 kg/h and in "Vary" selection the manipulated variable is set as water feed with specified upper and lower bounds.



Design specification
Type: Mass flow

Specification
Target: 8 kg/hr

Stream type
● Product ○ Internal ○ Decanter

Figure A.8.5 The design specification screen for ASPEN scrubber design

The reason why water is set as the manipulated variable is the excess amount of water released in sugar branch. The number of stages is varied to get the less number of stages which yields the necessary design specifications.

The recommended packing sizes for columns with diameter larger than 0.9 m is 50-75 mm. Therefore, 51 mm polypropylene plastic packing is utilized. Although they are not suitable for distillation, plastic packing is reported to be suitable for organic solvent absorption at moderate temperatures (Sinnot, 1999). Also, for the same packing size (50 mm) the cost of plastic packing is the lowest ($400 /m^3) when compared to saddles ($960 /m^3) and stainless steal pall rings ($1,360 /m^3). Therefore, pall ring polypropylene packing is used for the adsorption. The bulk density of 51 mm packing is given as 68 kg/m^3 with 102 m^2/m^3 specific surface area and 82 m^{-1} packing factor (Sinnot,1999).

The typical values for height of transfer units for 50 mm packing are given within the range 0.6-1 m. For the current design acetone water absorption system is taken as basis with value 0.75 m (Sinnot, 1999).

The more water is added to ethanol the more water is to be evaporated during the distillation. Therefore, the trade of between the cost of having one more stage and the cost of evaporating extra water input due to having one less stage is to be known. To observe the effect, the number of stages is reduced from 12 to 10. The corresponding water requirement increased from 58.9 ton/h to 60.5 ton/h.

As it is previously mentioned the cost of packing is $400 /m^3 (1998 prices). Using the column dimensions packing requirement for one stage is calculated as 2.64 m^3 which corresponds to $1,056. The casing of one column is calculated based on vertical storage tank cost as following

$$Cost = 2,400\$ \times (2.64)^{0.6} = \$4,297 \ \text{(1998 values)}$$

Since we need just the side surfaces 50% of the cost is taken as the shell cost. The cost coefficients are taken from Sinnot (1999). And, total contribution of one stage to cost is calculated as \$3,204/stage. For two stages this value corresponds to \$6,410.

The energy cost due to removal of extra water can be calculated as following; The difference in water requirement is 1.56 t/h. Cost of steam is \$12 /ton since energy produced with in the battery limits, 30% of this value is taken as cost which is equal to \$3.6 /ton. And, it is assumed that 1 ton steam evaporates 1 ton water. Finally, cost of vaporizing extra water is obtained as 5.6 \$/h which corresponds to \$23,450 /year which is much more then investment cost.

The same methodology is applied by varying the number of stages and the results are given in Figure A.8.6 and Table A.8.6. In Table A.8.6, the investment cost of vessel is calculated as previously described. But, it is multiplied by two to include the auxiliary material requirement.

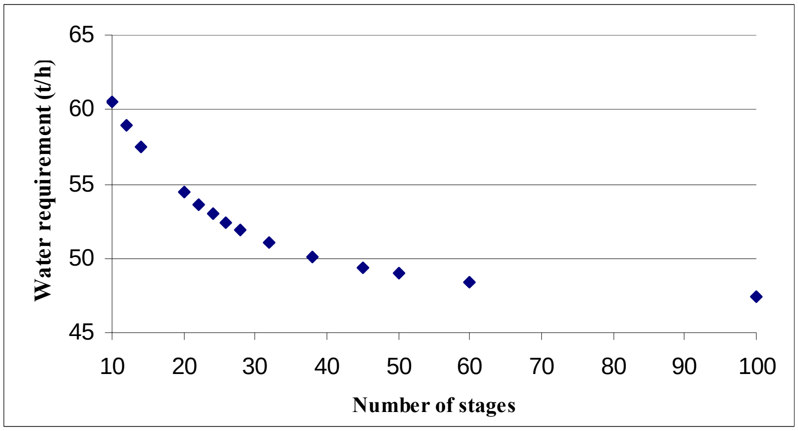


Figure A.8.6 The water requirement as function of number of stages

Table A.8.6 The optimization of number of stages for ethanol scrubber

# of stages	Water Flow (t/h)	Vessel Volume (m^3)	Packing Cost (\$)	Vessel Cost (\$)	Investment Cost (\$)	$\Delta_{\text{Investment}}$ (Stage i-10) (\$)	Evaporation Cost (\$/year)	$\Delta_{\text{Evaporation}}$ (Stage i-10) (\$/year)
10	60.48	26.47	10,590	34,272	44,861	0	909,232	0
12	58.92	31.77	12,708	38,233	50,941	6,080	885,780	-23,452
14	57.50	37.06	14,826	41,938	56,764	11,903	864,432	-44,800
20	54.43	52.95	21,179	51,946	73,125	28,264	818,279	-90,953
22	53.65	58.24	23,297	55,003	78,301	33,439	806,553	-102,679
24	52.94	63.54	25,415	57,951	83,366	38,505	795,879	-113,353
26	52.41	68.83	27,533	60,802	88,335	43,474	787,911	-121,321
28	51.90	74.13	29,651	63,567	93,218	48,356	780,244	-128,988
32	51.08	84.72	33,887	68,869	102,756	57,895	767,916	-141,316
38	50.13	100.60	40,241	76,349	116,590	71,729	753,634	-155,598
45	49.35	119.13	47,654	84,501	132,155	87,293	741,908	-167,324
50	48.99	132.37	52,948	90,015	142,964	98,103	736,496	-172,736
60	48.41	158.85	63,538	100,421	163,959	119,098	727,777	-181,456
100	47.37	264.74	105,897	136,438	242,335	197,473	712,142	-197,090

- Results are calculated using feed water temperature of 25ºC. In design the water temperature is 30ºC

As it is seen in Table A.8.6, the investment cost is always lower than the operation cost for any number of stages. Therefore, based on Figure A.8.6, the number of stages is selected as 40 where the curve starts to have a linear trend.

Using all the given information above, the dimensioning of the column is carried out in ASPEN plus and the results are given in Table A.8.7.

Table A.8.7 The design parameters for ethanol scrubber

Column diameter:	2.2	m
Column Height	30.00	m
Water Requirement	50.7	t/h
Maximum fractional capacity:	0.62	
Maximum capacity factor:	0.09	m/sec
Section pressure drop:	0.15	bar
Average pressure drop / Height:	0.066	bar/m
Maximum stage liquid holdup:	0.11	m^3
Surface area:	1.02	cm^3/ml

A.8.10. Beer Column (C-302)

The design of beer column is carried out in ASPEN plus using UNIQUAC property method which is suggested by ASPEN Plus user manual for alcohol separation systems. For initial simulation, the number of stages and the reflux ratio is taken from NREL report (Aden *et al*, 2002) as 32 and 3 respectively. The plate type is selected as nutter float valve with 0.48 Murphree tray efficiency and the tray spacing is selected as 1 m. The feed stage is 4th stage from top and the vapor side flow is taken from 8th stage from top. The feed temperature is set to 80ºC. The flow rate of the product side flow (101 t/h) is determined accordingly to set the ethanol concentration to 40% and the flow rate of the distillate is determined by built-in optimization tool of ASPEN plus to set the condenser temperature 20ºC. The reason why condenser temperature is set to 20ºC is the limitation of the cooling agents (Chilled water at 10ºC). The by product side flow is set to 350 t/h. The ethanol loss with top vapor product is ~11%. It is possible to decrease the amount of the ethanol in top product, but, this operation requires operation below 0ºC which requires refrigeration and increases the operation cost. Instead, the condenser temperature is set top 20ºC and the lost ethanol is recovered by adsorption column. The condenser is set to partial vapor-liquid and the vapor fraction in the condenser is set to 1. To iterate the condenser temperature, the temperature of the stage 1 is set to 20ºC from "Design specifications" as shown in Figure A.8.5 and the distillate rate is iterated between 10,000-30,000 kg/h by inputting these boundaries in "Vary" selection. And, the distillate rate that gives 20ºC condenser temperature is obtained as 15,651 kg/h.

Table A.8.8 The optimal number of trays for C-302

# of stages	Feed Stage	Product Stage	Ethanol Lost (%)	% Ethanol in Product
10	2	4	1.34	38.8
12	3	6	1.42	38.7
14	4	6	0.43	39.1
16	4	6	0.13	39.3
18	4	6	0.04	39.3
20	4	6	0.03	39.3

After the calculation, the ethanol mass fraction in product stream is obtained as 39.3% and the ethanol loss in bottom product is 0.003%. However, 0.003% is a low value for product loss. Thus, the number of stages of the column is manipulated manually to find the minimum number of stages satisfying current design specifications. In varying the total number of stages, the feed point location should also be changed to get proper results. The feed point locations are determined such that a smooth concentration profile is obtained in the column. False and correct feed point locations are shown in Figure A.8.7. The ethanol loss for different number of stages is given in Table A.8.8.

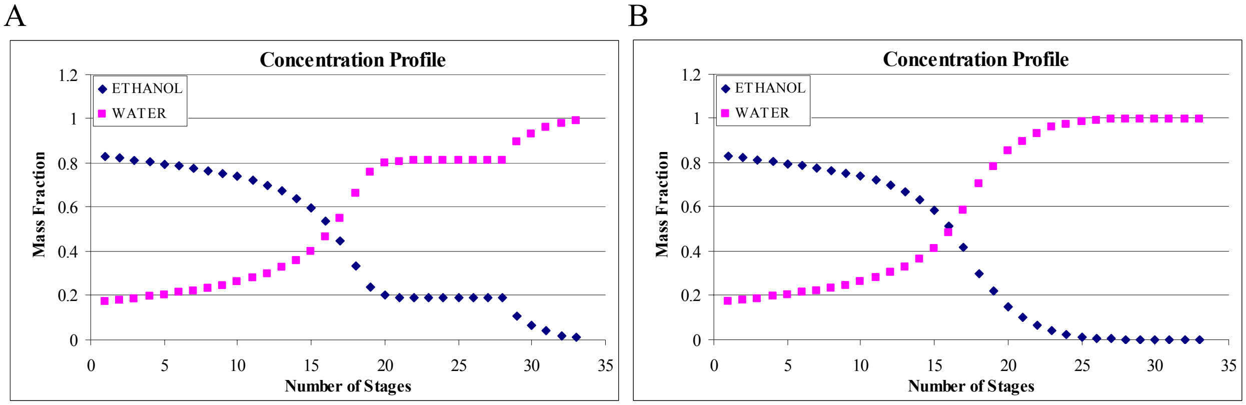


Figure A.8.7. False (A) and correct (B) concentration profiles

Based on these results, the number of stages is selected as 20 with feed and products stages 4 and 6 respectively. The flow rates of distillate, side draws and bottoms are obtained as 15,538 kg/h, 101,000 kg/h, 350,000 kg/h and 61,024 respectively. The heat duties of condenser and reboiler are obtained as 18,719 and 76,886 kW respectively, with condenser and reboiler temperatures of 20ºC and 100ºC respectively. The product vapor temperature is 92.6ºC.

The dimensioning of the column is carried out by inputting the necessary data in to "Tray sizing" selection of ASPEN plus RadFrac. The reason why Nutter Float valve type trays selected was that type of tray gave the smallest column diameter. The pressure drop in the column is calculated as 0.27 bars.

A.8.10.1 Beer Column Condenser (E-304)

The amount of the vapour to be condensed by the condenser is 46,629 kg/h. the condenser operates at atmospheric pressure. The vapour enters the condenser saturated at 76.3ºC and the condensation is complete at 20ºC. The condenser heat duty is 18,719 kW. Chilled water will be used as the cooling agent. The average heat capacity of chilled water between 10 - 40ºC is 4.1793 kJ/kgºC. The amount of required chilled water is calculated as 149.3 kg/s using the following equation.

$$18,719 kJ/s = m_{Ch} \times 4.1793 \times (40-10)$$

As an initial estimate overall heat transfer coefficient of 1,300 $Wm^{-2o}C^{-1}$ is used (recommended interval is 1,000-1,500 W m^{-2} $^{o}C^{-1}$).

The Log mean temperature difference is calculated as 20.3 ºC using Equation A.8.43. The correction factor for log mean temperature difference is obtained as 0.98 for one shell pass four tube passes. Therefore, the corrected *LMTD* is 19.9ºC. Using the mean temperature difference, heat duty and the initial estimate for overall heat transfer coefficient initial area is calculated as,

$$Area = \frac{18,719 \times 10^3 W}{1300 W/m^{2\,o}C \times 27.6\,^oC} = 724\,m^2$$

Using tube outer diameter of 2.5 cm and tube length of 6 m the surface area of one tube is calculated as 0.393 m^2. The required number of tubes is found as 1,536 by dividing total area by single tube surface area.

Using square pitch the pitch can be found as $P_t = 1.25 \times 0.025 = 0.031 m$

The tube bundle diameter is calculated as 1.45 m using the following equation

$$D_b = d_0 \left[\frac{number\ of\ tubes}{K_1} \right]^{1/n_1} \quad \text{Eqn A.8.31}$$

d_o is outer diameter of tubes in m, K_1 and n_1 are constants determined by system. For four passes and square pitch, K_1 and n_1 have values of 0.158 and 2.263 respectively. Number of tubes in centre row is calculated as

$$N_r = \frac{D_b}{P_t} = 46$$

Shell Side Coefficient

To calculate the shell side heat transfer coefficient the wall temperature of tubes should be calculated. For that condensing temperature is assumed as 3,500 W m^{-2} $^{o}C^{-1}$.

Shell side mean temperature is $\frac{76+20}{2} = 48\,^{o}C$

Tube side mean temperature is $\frac{40+10}{2} = 25\,^{o}C$

The wall temperature is calculated as 39.5 ºC calculated by assuming that condensation heat flux is equal to total heat flux using the following equations.

$$U_{cond}(T_{Shell} - T_{Wall}) = U_{Overall}(T_{Shell} - T_{Tube}) \quad \text{Eqn A.8.32}$$

$$3500(48 - T_{Wall}) = 1300(48 - 25)$$

The mean temperature of the condensate is calculated as $\frac{48+39.5}{2} = 43.7\,^{o}C$

The liquid condensate at 43.7ºC is assumed to be composed of half ethanol half water. The physical properties at specified temperature are (Geankoplis,1993)&(Sinnot, 1999) given in Table A.8.9. Viscosities are calculated using the correlation

$$Log[\eta] = VISA \times \left(\frac{1}{T} - \frac{1}{VIS\ B} \right) \quad \text{Eqn A.8.33}$$

Table A.8.9 The physical properties of ethanol and water at condenser (E-304) conditions

Property	Ethanol	Water	Mixture
VIS A	686.64	300.88	-
VIS B	658.25	283.16	-
Viscosity kg $m^{-1}s^{-1}$	0.0008	0.0006	0.0007
Density kg/m3	772	991	881.5
Thermal Conductivity (kl) W/m ºC	0.17	0.63	0.4

Vapor density is obtained from ASPEN as 1.6 kg/m^3. The mean condensing overall heat transfer coefficient is calculated using

$$h_o = 0.95 k_L \left[\frac{\rho_L(\rho_L - \rho_v) g}{\eta_L \Gamma_h} \right]^{1/3} \times N_r^{-1/6} \qquad \text{Eqn A.8.34}$$

h_o is shell side condensation coefficient (W m^{-2} ^{o}C^{-1}), g is the gravitational acceleration (m s^{-2}), k_L is liquid thermal conductivity (W m^{-1} ^{o}C^{-1}), ρ_L is liquid density (kg m^{-3}), μ_L is Liquid viscosity (Nsm^{-2}), ρ_v is vapor density (kg m^{-3}), N_r is the average number of tubes in a centre tube row, Γ is the condensate flow per unit length of tubes which is defined as,

$$\Gamma = \frac{Total\ Condensate\ Flow\,(kg/s)}{Length\ of\ one\ tube \times Number\ of\ tubes} \qquad \text{Eqn A.8.35}$$

Γ is calculated as 1.4 x 10^{-3} kg/s m. The corresponding condensation overall heat transfer coefficient is calculated as 4,032 W m^{-2} ^{o}C^{-1} which is close to assumed value of 3,500 W m^{-2} ^{o}C^{-1} so the T_{wall} does not need to be corrected.

Tube-side Coefficient

The tube side coefficient is calculated using the correlation given for water,

$$h_i = \frac{4200 \times (1.35 + 0.02T) u_t^{0.8}}{d_i^{0.2}} \qquad \text{Eqn A.8.36}$$

h_i is tube side heat transfer coefficient (W m^{-2} ^{o}C^{-1}), u_t is the water velocity (m s^{-1}), T is water temperature (ºC), d_i is tube internal diameter (mm)

The internal diameter of the tubes is 0.02 m. Using this diameter, the flow cross sectional area is calculated as 3.14x10^{-4} m^2. The number of tubes is 1,536. Since the number of passes is 4 the total flow are a is calculated from,

$$3.14 \times 10^{-4} \times \frac{1,536}{4} = 0.12 m^2$$

The tube side mean temperature is already calculated as 25 ºC. The density of water at that temperature is 999.95 kg/m^3. The mass flow was obtained as 149.3 kg/s. Using this values, the flow velocity is calculated as 2.6 m/s. Putting all this values in equation yields the tube side heat transfer coefficient of 9,298 W m^{-2} ^{o}C^{-1}.

Overall Heat Transfer Coefficient

The overall heat transfer coefficient is calculated using,

$$\frac{1}{U} = \frac{1}{h_o} + \frac{1}{h_{od}} + \frac{d_o \times \ln \frac{d_o}{d_i}}{2 \times k_w} + \frac{d_o}{d_i}\frac{1}{h_{id}} + \frac{d_o}{d_i}\frac{1}{h_i} \qquad \text{Eqn A.8.37}$$

h_{od} and h_{id} are the outer and inside fouling factors, (assumed as 6,000 W m^{-2} $^oC^{-1}$), k_w is thermal conductivity of tube wall material (for the current case 50 W m^{-1} $^oC^{-1}$).

Inserting all this values into equation and calculating the equation yields an overall heat transfer coefficient of 1,229 W m^{-2} $^oC^{-1}$, which is close enough to initial estimate. The resultant heat transfer area is 765 m^2.

A.8.10.2. Beer Column Reboiler (E-303)

The design of reboiler is carried out in similar manner with condenser. The total amount to be reboiled is 122,439 kg/h. The boiling media is assumed to be pure water. The temperature of the input stream is 100ºC. Low pressure steam will be used as the heating agent. The reboiler heat duty is 76,886 kW. The temperature, pressure and the enthalpy of condensation of low pressure steam is 130.7 ºC, 275 kPa and 2,171 kj/kg respectively. The reboiler is a kettle type reboiler and the design is carried out as described by Sinnot (1999).

The required amount of steam is found as 35.4 kg/s by dividing the heat duty by enthalpy of steam. The interval for steam water vaporizers' overall heat transfer coefficient is given as 1000-1500 W m^{-2} ºC^{-1}. The initial estimate is taken as 1,300 W m^{-2} ºC^{-1}. Since one side of the tubes is condensing while the other side is evaporating the mean temperature difference is calculated as the difference of hot side and cold side which yields 30.74ºC. The initial area requirement is found as 1,948 m^2 by using the following equation,

$$76.886 \times 10^6 W = A\, m^2 \times 1{,}300\, Wm^{-2}\ {}^oC \times 30.74^o C$$

The selected tubes have 0.02 m inside and 0.025 m inside diameters with nominal length of 8 meters. The surface area of one tubes is calculated as 0.63 m^2 and the required total number of tubes is obtained as 3,100. The tube bundle diameter is calculated in same manner with condenser and obtained as 1.97 m.

Boiling Coefficient

The boiling coefficient is calculated using the following equation

$$h_b = 0.104 P_c^{0.69} Q'^{0.7} \left[1.8 \left(\frac{P}{P_c} \right)^{0.17} + 4 \left(\frac{P}{P_c} \right)^{1.2} + 10 \left(\frac{P}{P_c} \right)^{10} \right] \qquad \text{Eqn .A.8.38}$$

h_b is boiling heat transfer coefficient (W m^{-2} $^oC^{-1}$), P_c is the critical pressure of the liquid to be evaporated (bar), P is the operating pressure (bar), Q' is the heat flux (W m^{-2}).

Since it is assumed that the bottom product is water, the critical pressure of water (220.5 bar) is used for the calculations. The heat flux can be obtained based on the estimated area using the heat duty as

$$q = \frac{76{,}886\, kW}{1922\, m^2} = 40\, kW\, m^{-2}$$

Inserting the values in the equation yields boiling coefficient of 5,209 W m^{-2} $^oC^{-1}$.

Overall Heat Transfer Coefficient

The overall heat transfer coefficient is calculated using Equation A.8.35. But, the descriptions of terms are different for boiling. Here, the outer heat transfer coefficient (h_o) is the boiling coefficient and the inner (h_i) is the condensation heat transfer coefficient of steam which is suggested as 8,000 W m^{-2} ^{o}C^{-1} (Sinnot, 1999). Inner and outer fouling coefficients are taken as 6,000 W m^{-2} ^{o}C^{-1}. The conductivity of tubing material is taken as given in condenser design. The resultant overall heat transfer coefficient is 1,281 W m^{-2} ^{o}C^{-1} which is in agreement with the initial estimate. Using the resultant heat transfer coefficient the required heat transfer area is obtained as 1,953 m^2.

A.8.11.Rectification Column (C-303)

Rectification column is operated to concentrate the 39.3% ethanol mixture coming from beer column to w/w 92.5% before being sent to molecular sieve adsorption. The design of rectification column was carried out in ASPEN plus. To find the initial estimates for required number of stages and reflux ration DSTWU column is used. The initial guess for the reflux ratio is set to 4. The pressure in the column is set to atmospheric pressure. Light key (ethanol) recovery in top stage is set to 99.9% and heavy key recovery in distillate is set to 5%. The results of the simulation are given in Table A.8.10.

The results of DSTWU simulation is input to RadFrac column as the initial estimates and the number of stages are manipulated to find the minimum number of stages that yields 0.05% ethanol loss in bottom product. In RadFrac column, the specified number of stages, reflux ratio and feed stage positions are input, the condenser is set to partial vapor liquid and the distillate vapor fraction is set to 1. As the initial guess, the distillate flow rate is specified as 42,900 kg/h. The RadFract distillation column is utilized in simulation of the distillation column. To obtain the product ethanol composition of 92.5%, the built-in optimization tool of ASPEN plus is utilized. In the “design specifications” tab, the composition of ethanol in distillate is set to 92.5% and in the “Vary” tab of the simulation window the distillate flow rate is set as manipulated variable. The lower and upper bounds are set to 42,000 and 43,900 kg/h respectively. In Table A.8.11, different number of stages and the corresponding ethanol losses are given

Table A.8.10 Inputs for DSTWU column in ASPEN simulation

Minimum reflux ratio:	3.2	
Actual reflux ratio:	4.0	
Minimum number of stages:	17.63	
Number of actual stages:	32.8	
Feed stage:	29.1	
Number of actual stages above feed:	28.1	
Reboiler heating required:	23,632.0	kW
Condenser cooling required:	44,970.0	kW
Distillate temperature:	78.23	C^o
Bottom temperature:	99.95	C^o
Distillate to feed fraction:	0.243	

Table A.8.11 Ethanol loss as a function of number of stages

Number of Stages	Ethanol Loss (%)
34	0.004
32	0.01
30	0.04
28	0.13

As a result, the number of stages is set to 32 with feed point location of stage 22 and reflux ratio of 4. The flow rates of distillate and bottoms are obtained as 42,915 and 58,085 kg/h respectively. The heat duties of condenser and reboiler are obtained as 56,679 and 8,803 kW respectively, with condenser and reboiler temperatures of 77.5 ºC and 100ºC respectively. The vapor fraction of the distillate is 0 that is distillate is collected as liquid.

The dimensioning of the column is carried out by inputting the necessary data into tray sizing selection of ASPEN plus RadFrac. The spacing between the trays is set to 1 m.

A.8.11.1. Rectification Column Condenser (E-306)

The amount of the vapor to be condensed by the condenser is 171,624 kg/h. the condenser operates at atmospheric pressure. The vapor enters the condenser saturated at 78.2 ºC and leaves at 77.5 ºC. The condenser heat duty is 56,679 kW. Cooling water will be used as the cooling agent. The inlet and outlet temperatures of cooling water are set to 25 and 60 ºC respectively. The average heat capacity of chilled water between 25-60ºC is 4.1836 kJ/kgºC. The required amount of water is calculated as 387.8 kg/s. The same methodology is applied with Appendix A.8.10.1.

The initial estimate for overall heat transfer coefficient is taken as 800 W m^{-2} $^{o}C^{-1}$. The Log mean temperature difference is calculated as 32.4 ºC. The correction factor for log mean temperature difference is obtained as 1 for one shell pass four tube passes. Therefore, the corrected *LMTD* is same. Using these values, initial area is calculated as 2,188 m^2.

Using tube outer diameter of 2.5 cm and tube length of 8 m, the surface area of one tube is calculated as 0.628 m^2. The required number of tubes is found as 3,483 by dividing total area by single tube surface area. Using square pitch, the pitch is found as 0.031 m. The tube bundle diameter is calculated as 2.1 m. Number of tubes in centre row is calculated as 67.

Shell Side Coefficient
Condensing temperature is assumed as 1,700 W m^{-2} $^{o}C^{-1}$.

Shell side mean temperature is 78 ºC
Tube side mean temperature is 42.5 ºC
Wall temperature is calculated as 61.2 ºC
Mean temperature of the condensate is 69.5 ºC

The liquid condensate at 69.5 ºC is assumed to be composed of 92.5% ethanol and 7.5% water. The physical properties at specified temperature are given in Table A.8.12.

Table A.8.12 The physical properties of ethanol and water at condenser conditions (E-306)

Property	Ethanol	Water	Mixture
Viscosity kg $m^{-1}s^{-1}$	0.0005	0.0004	0.0005
Density kg/m3	759	984	776
Thermal Conductivity (kl) W/m ºC	0.16	0.63	0.4

Vapor density is obtained from ASPEN as 1.43 kg/m^3. Γ is calculated as 1.7 x 10^{-3} kg/s m. The corresponding condensation overall heat transfer coefficient is calculated as 1,765 W m^{-2} $^{o}C^{-1}$ which is close to assumed value of 1,700 W m^{-2} $^{o}C^{-1}$ so the T_{wall} does not need to be corrected.

Tube-side coefficient

The internal diameter of the tubes is 0.02 m. Using this diameter, the flow cross sectional area is calculated as 3.14x10^{-4} m^2. The total flow area is 0.27 m^2. The tube side mean temperature is 42.5 ºC. The density of water at that temperature is 993.7 kg/m^3. The mass flow was obtained as 387.1 kg/s. Using this values, the flow velocity is calculated as 1.43 m/s. The tube side heat transfer coefficient is calculated as 6,734 W m^{-2} ºC^{-1}.

Overall Heat Transfer Coefficient

And finally, the overall heat transfer coefficient is calculated as 845 W m^{-2} ºC^{-1} which is close enough to initial estimate. The resultant heat transfer area is 2,071 m^2.

A.8.11.2. Rectification Column Reboiler (E-305)

The design of reboiler is carried out in similar manner with beer column reboiler. The total amount to be reboiled is 14,033 kg/h. The boiling media is assumed to be pure water. The temperature of the input stream is 100ºC. Low pressure steam will be used as the heating agent. The reboiler heat duty is 8,803 kW. The temperature, pressure and the enthalpy of condensation of low pressure steam is 130.7 ºC, 275 kPa and 2,171 kj/kg respectively. Since the operation conditions of the beer column reboiler and rectification column reboiler are similar, the overall heat transfer coefficient is selected as 1,281 W m^{-2} ºC^{-1} as calculated for beer column reboiler. The resultant heat transfer area is obtained as 210 m^2. The required amount of steam is found as 4.05 kg/s by dividing the heat duty by enthalpy of steam.

A.8.12.Molecular Sieve Adsorption Column (C-304)

Due to the dependence of adsorption dynamics on liquid flow rates, the dimensioning of the vessel requires the mathematical model of the system. For this purpose, the MATLAB program which is kindly provided by Joukje Houwing is used. The details of the model can be found in research article of Joukje *et al.* (2003). Equilibrium data for water-adsorbent system is taken from published article of Carmo and Gubulin (1997). Nonlinear isotherm is represented by Langmuir isotherm which is given as,

$$q_{H_2O} = \frac{\Theta \kappa C_{H_2O}}{1+\kappa C_{H_2O}} \qquad \text{Eqn .A.8.39}$$

q_{H_2O} is the solid phase concentration (g water/g solid), Θ is the capacity of the monolayer (g water/g solid), κ is the Langmuir constant, and C_{H_2O} is the water percentage of the mixture The parameters are as given in Table A.8.13.

Table A.8.13 The equilibrium data for molecular sieves

Temperature (ºC)	Θ (g_{water}/g_{ads})	κ (1/% water)	Diffusivities (x10^{-9} m^2/s)
25	0.249	0.317	1.67
40	0.238	0.153	2.50
50	0.220	0.090	3.88
60	0.200	0.052	5.70

The supplier of molecular sieves reports the equilibrium capacity of the adsorbents 0.21. To check weather the literature data yields the commercial capacities, the parameters are inserted in to equation and the equilibrium capacity is calculated as,

$$q^*_{H_2O} = \frac{0.2 \times 0.052 \times 7.5}{1 + 0.052 \times 7.5} = 0.056 \frac{kg\ water}{kg\ adsorbent}$$

Which is far less then commercial values. As a second trial, the data for 25ºC is inserted,

$$q^*_{H_2O} = \frac{0.249 \times 0.317 \times 7.5}{1 + 0.317 \times 7.5} = 0.175 \frac{kg\ water}{kg\ adsorbent}$$

This is still less then commercial value, but, closer. Therefore, the final kinetic parameter that will be used in design is taken as in Table A.8.14.

Table A.8.14 The equilibrium data for molecular sieves

Temperature (ºC)	Θ (g_{water}/g_{ads})	κ (1/% water)	Diffusivities (x10^{-9} m^2/s)
60	0.249	0.317	5.70

Since the concentration units used in MATLAB program is kg/m^3 the parameters are needed to be converted to corresponding units. Using the density of water-ethanol solution as 725 kg/m^3 (From ASPEN simulation) Θ and κ is obtained as,

$$\Theta' = 0.298 \frac{kg\ water}{kg\ adsorbent} \times 2{,}765 \frac{kg\ adsorbent}{m^3 adsorbent} = 688.5 \frac{kg\ water}{m^3 adsorbent}$$

$$\kappa' = 0.317 \frac{1}{\%\,water} \frac{\%\,water}{0.01\,water} \times \frac{1}{725} \frac{m^3\ solution}{kg\ solution} = 4.4 \times 10^{-2} \frac{m^3\ solution}{kg\ water}$$

The concentration of water in 7.5% water ethanol solution is obtained as 54.3 kg/m^3 from ASPEN simulation. The total volumetric flow rate is 0.0165 m^3/s. The time required for regeneration of the beds is taken as 2 h, but, the regeneration temperature is higher then the operation temperature Therefore, some cooling time should also be included in the operation time. This is why the total operation time is set to 2 h 10 min (7,800 s) for each absorbent bed. The properties and parameters required by the program is listed in Table A.8.15.

Table A.8.15 The input parameters for molecular sieve adsorption column model

Parameter	Value	Unit
L (m)	3-9 (Manipulated)	m
D (m)	1-4 (Manipulated)	m
Bed Porosity	0.4	
Zeolite adsorbent diameters	0.002	m
Liquid flow rate	0.0165	m^3.s^{-1}
Molecule diameter of water	2.8x10^{-10}	m
Diffusivity of water	5.7x10^{-9}	m^2.s^{-1}
Density of solution	725	kg.m^{-3}
Viscosity of solution	0.011	Pa.s
Feed water concentration	54.25	kg.m^{-3}

Diameter and the height of the bed are manipulated to get the necessary break even point. The break even point is determined as 5.4 kg/m^3 which is 10% of the inlet concentration. The simulation results are presented in Figures A.8.8-A.8.9. Setting the break even point to 5.4 yields an average concentration of 0.9 kg/m^3 which yields an outlet composition of 99.88% ethanol (see Figure A.8.9-A) the concentration that gives A_1=A_2 is taken as the average concentration) the resultant dimensions of the adsorption column is obtained as 2.2 m diameter and 7.8 m length. Using this information the total amount of required adsorbent material is calculated as 21.6 ton. The ratio of unused bed to used bed is obtained as 35% (Figure A.8.9-B $A_4/[A_3+A_4]$).

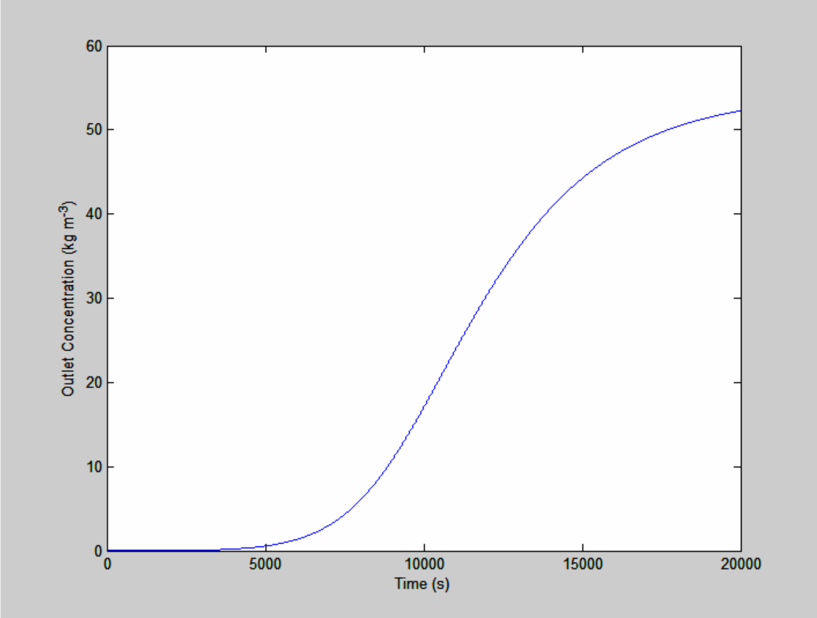


Figure A.8.8. Outlet concentration profile of molecular sieve adsorption column

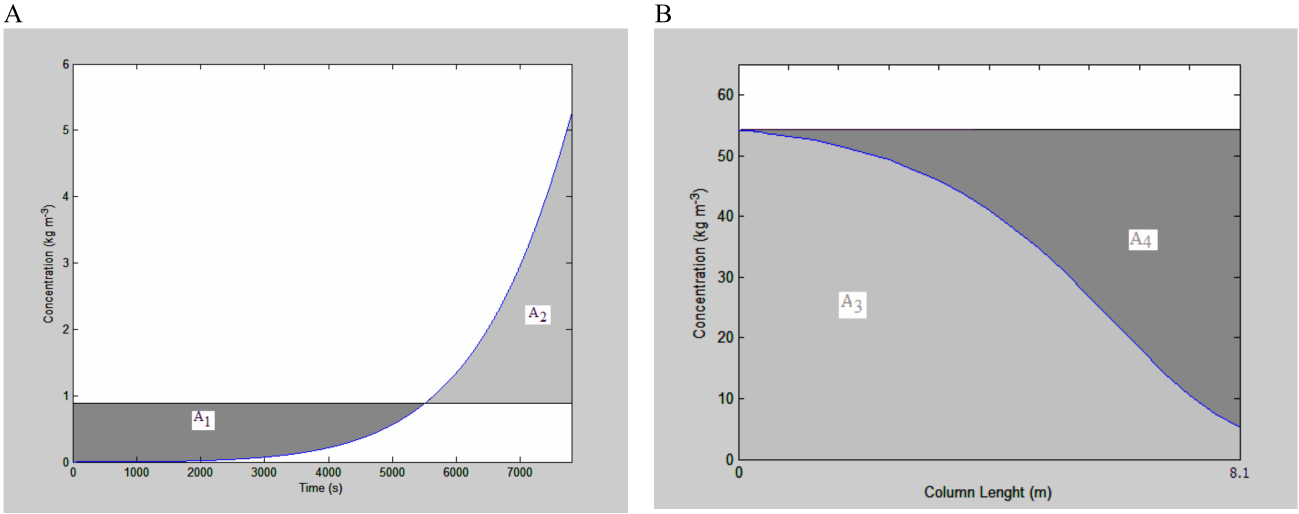


Figure A.8.9. Breakeven point (A) and concentration profile in column (B)

A.8.12.1. Cooling

Cooling is required to remove the heat of adsorption and to keep the adsorbent bed temperature at 60ºC. The heat of adsorption is 2,365 kJ/kg water. For adsorption rate of 3,212 kg water/h the heat duty of the cooling system is obtained as 2,110 kW. Cooling water at 25ºC is used as cooling agent and the exit temperature of the cooling water is set to 55ºC. The corresponding cooling water requirement is obtained as 60.5 t/h.

The overall heat transfer coefficient for water-organic solvents is given in the interval 250-750 W m^{-2}ºC^{-1} (Sinnot, 1999). A value of 750 W m^{-2}ºC^{-1} is taken for the calculations. The LMTD is calculated as 15.5ºC. Using the LMTD and overall heat transfer coefficient, the area requirement is obtained as 181.5 m^2.

A.8.12.2. Regeneration

The regeneration temperature for the zeolite adsorbents should be between 175-260 ºC. The regeneration gas can be either carbon dioxide or nitrogen. Since the process produces excess amount of carbon dioxide, the gas for regeneration can be selected as carbon dioxide. However, these streams contain water vapor and impurities like N_2O and SO_2. Also in presence of humidity carbon dioxide will form carbonic acid and increase the corrosion risk in the vessel which increases the investment cost. Therefore, the regeneration gas is selected as nitrogen. The velocity of the gas is set to 3 m/s (Sowerby and B.D. Crittenden, 1988) which corresponds to total gas flow rate of 11 m^3/s (12 kg/s = 43.2 t/h). The make up nitrogen is assumed to be 0.1% of the total requirement (180 t/year).

The spent nitrogen can be dehydrated to be reused in the process. Taking the nitrogen flow as 42.3 t/h and the water to be desorbed as 3.2 t/h, the composition of the moist nitrogen is obtained as 7.5% water. The dehydration is carried out by condensing the water by refrigerant reagents at low temperatures. ASPEN simulation shows that the water content of the nitrogen stream can be lowered from 7.5% to 0.5 % by condensing at 5ºC.

Desorption heating

Desorption temperature is set to 200 ºC. Due to the evaporation of water the column is to be heated up by steam to provide the energy requirement for evaporation and to keep the bed temperature constant at 200ºC. The nitrogen gas is assumed to enter the column at 180ºC. The heat of evaporation of water at 200ºC is 1,933 kj/kg, using this value the heat required to evaporate 3.2 t/h of water is obtained as 1,724 kW. Using nitrogen heat capacity given in Equation A.8.40, heating required to heat up nitrogen from 180 to 200ºC is obtained as 367 kW. The heating agent is superheated steam from the boilers at 275ºC and 2500 kPa. The enthalpy of condensation for superheated steam is 1,985 kJ/kg which yields steam requirement of 3.8 t/h. The mean temperature difference is 45ºC and the overall heat transfer coefficient is taken as 300 W $m^{-2\circ}C^{-1}$ (Sinnot, 1999). Using total power requirement of 2,091 kW, the resultant heat exchanging area requirement is calculated as 155 m^2. This area requirement is lower then the cooling area requirement; therefore, the cooling system will be switched to heating system during the regeneration.

Nitrogen heater

To recover the heat and to save refrigeration energy the dry nitrogen from condenser with temperature of 5ºC is heated up to 180ºC using the moist nitrogen from the regeneration as hot utility. Meanwhile, the moist nitrogen is cooled down from 200ºC to 30ºC. The heat capacity of nitrogen is given as (Perry and Green, 1997),

$$C_{p,N_2} = 1.512 + 0.0002T \frac{kJ}{kg^\circ C} \qquad \text{Eqn A.8.40}$$

The nitrogen heat capacity is taken as an average value of 1.53 kJ $kg^{-1\circ}C^{-1}$. The heat duty to heat up the nitrogen is calculated as 3,304 kW. The *LMTD* is 23 ºC and the overall heat transfer coefficient for gas-gas system is taken as 300 W $m^{-2\circ}C^{-1}$. The resultant heat exchanging area requirement is calculated as 467 m^2.

The moist nitrogen is further condensed in condenser at 5ºC. ASPEN simulation shows that the refrigeration requirement of condenser is 1,007 kW. The *LMTP* for condenser is 14 ºC and the overall heat transfer coefficient is assumed based on air gas-brine systems. The corresponding heat transfer coefficient is taken as 250 W m^{-2} ºC$^{-1.}$ The resultant heat exchanging area is obtained as 288 m^2.

A.8.13. Design of Heat Exchanging Equipments

The heat duty of the system is determined by using following equations,

$$Q = mC_p(T_{out} - T_{in})$$

Eqn A.8.41

Q is power required or generated (kW), *m* is the mass flow of the media to be heated/cooled (kg/s), C_p is the heat capacity of the media (kJ kg^{-1}ºC), T_{out} and T_{in} are the exit and inlet temperatures of the media (ºC).

To determine the total heating area requirement following equations are used,

$$Q = AU\Delta T_{LM}$$

Eqn A.8.42

A is heat exchanger area (m^2), *U* is the overall heat transfer coefficient (W m^{-2} ºC^{-1}), ΔT_{LM} is the log mean temperature difference for counter current flow (ºC) which is calculated based on equation,

$$\Delta T_{LM} = \frac{(T_{hot,in} - T_{cold,out}) - (T_{hot,out} - T_{cold,in})}{\ln\left[\frac{(T_{hot,in} - T_{cold,out})}{(T_{hot,out} - T_{cold,in})}\right]}$$

Eqn A.8.43

A.9. Process and Equipment Design: 2015 Plant

All the units used in this section are metric units.

A.9.1. Pretreatment Reactor (R-401)

The flow of bagasse (50% water) is 5.7 ton/min. With the contribution of acid solution the flow rate will be 8.55 ton/min. To calculate the required reactor size the amount of steam to be injected should be calculated. For this, a heat balance calculation is carried out for all streams. The inflow rates of dry fibers, water and acid solution is 2.85 ton/min each. The specific heat capacity of bagasse is not known, it will be assumed as the heat capacity of cellulose which is given as 1.34 kJ/kg ºC (Perry and Green, 1997). The acid solution will be assumed as pure water and the specific heat of water is,

$$C_{p(water)} = 4.189 - 0.0006T \text{ kJ/ kg ºC} \qquad \text{Eqn A.9.1}$$

The steam will be extracted from the turbines and properties of superheated steam from boiler are given in Table A.9.1.

Table A.9.1 Properties of superheated steam at 25 bar and 275ºC

Parameter	Value
Temperature (ºC)	275
Enthalpy of Saturated Liquid at 224ºC and 25 bar (kJ/kg)	961.96
Enthalpy of Superheated Vapour (kJ/kg)	2,947.4

First the steam requirement for the fiber steaming vessel is calculated;

Heat required to heat up fibers to 100ºC is

$$Q_{f(43.8\to 100^\circ C)} = 2,850\frac{kg}{\min}\times 1.32\frac{kJ}{kg^\circ C}\times(100-43.8) = 211.4\frac{MJ}{\min}$$

Heat required to heat up water in the fiber to 100ºC

$$Q_{w(43.8\to 100^\circ C)} = 2,850\frac{kg}{\min}\times\left[4.189\times(100-43.8)-0.0003\times(100^2-43.8^2)\right] = 664.04\frac{MJ}{\min}$$

This heat will be supplied by superheated steam at 275 ºC and 25 bar. Assume that the total required steam is S ton/min

Then, heat supplied by condensing at 224 ºC will be

$$Q_C = S\times 1,000\frac{kg}{\min}\times(2947.4-961.96)\frac{kJ}{kg} = 1,985.4\times S\frac{MJ}{\min}$$

And, the heat supplied by water cooling from 224ºC to 100 ºC is

$$Q_{Cool} = S\times 1,000\frac{kg}{\min}\times\left[4.189\times(224-100)-0.0003\times(224^2-100^2)\right] = 507.4\times S\frac{MJ}{\min}$$

Therefore, the total heat supplied by S ton steam is 2,492.8 S MJ/min

Based on this information, the steam requirement to heat up the fibers to 100 ºC is calculated as 0.35 ton/min. This yields fiber flow of 2.85 ton/min and liquid flow of 6.05 ton/min into the pressure feeder where the acid solution is blended with fibers. Similar calculations are carried out to find the amount of steam injected into the reactor

Heat required to heat up fibers to 190ºC is

$$Q_{f(100\rightarrow 190^\circ C)} = 2{,}850\frac{kg}{\min}\times 1.32\frac{kJ}{kg^\circ C}\times(190-100) = 338.6\frac{MJ}{\min}$$

Heat required to heat up water in the fiber to 190ºC is

$$Q_{w(100\rightarrow 190^\circ C)} = 3{,}200\frac{kg}{\min}\times\left[4.189\times(190-100)-0.0003\times(190^2-100^2)\right] = 1{,}181.4\frac{MJ}{\min}$$

And heat required to heat up the acid solution to 190ºC is

$$Q_{a(100\rightarrow 190^\circ C)} = 2{,}850\frac{kg}{\min}\times\left[4.189\times(190-100)-0.0003\times(190^2-100^2)\right] = 1{,}052.2\frac{MJ}{\min}$$

This heat will be supplied by steam, condensing at 224 ºC ,

$$Q_{Condense} = S\times 1{,}000\frac{kg}{\min}\times(2{,}947.4-961.96)\frac{kJ}{kg} = 1{,}985.4\times S\frac{MJ}{\min}$$

And, the heat supplied by cooling from 224ºC to 190 ºC is

$$Q_{Cool} = S\times 1{,}000\frac{kg}{\min}\times\left[4.189\times(224-190)-0.0003\times(224^2-190^2)\right] = 138.2\times S\frac{MJ}{\min}$$

Therefore the total heat supplied by S ton steam is 2,123.6S MJ/min. Using this information, the amount of steam to be injected in to reactor is obtained as, 1.21 ton/min.

Using this information, total liquid flow in to vessel is calculated as 7.34 ton/min. This yields solid concentration of 28% which is an acceptable value (30% (Aden *et al.*, 2002), 33% (Morjanoff and Gray, 1986)).

The density of the fibers will be assumed as 1,600 kg/m^3. The liquid density of water in reactor conditions is calculated using

$$Density_{water} = \frac{18\times C_1}{C_2^{\left[1+\left(1-\frac{T}{C_3}\right)^{C_4}\right]}}\frac{kg}{m^3} \qquad \text{Eqn A.9.2}$$

Where, C_1= 4.391, C_2= 0.2487 C_3=647.13 and C_2=0.2534 (Perry and Green, 1997), Temperature is in Kelvin. The density of water at 190 ºC is calculated as 874 kg/m^3. The volume contribution of fibers and water to flow is

$$V_{fiber} = \frac{2{,}850\frac{kg}{\min}}{1{,}600\frac{kg}{\mathrm{m}^3}} = 1.78\frac{m^3}{\min} \text{ and } V_{water} = \frac{7{,}260\frac{kg}{\min}}{874\frac{kg}{\mathrm{m}^3}} = 8.30\frac{m^3}{\min}.$$

Therefore, the total volume flow into reactor is 10.08 m^3/min and the residence time in the reactor is 2 min. The required vessel volume is calculated as 22 m^3 with 10% safety. The reactor should have a plug flow behavior to prevent loss of conversion due to mixing.

The vapor pressure of the reactor ingredients at 190 ºC is 12.55 bar. Therefore, the pressure in the reactor is set to 14 bars to keep the reactor contents in liquid phase.

A.9.2. Collecting Tank (V-401)

The Collecting tank is designed as a flash tank and the flash calculations are carried out in ASPEN plus. The Hayden O'Connell equation of state is used to simulate the flash tank to account for the fact that acetic acid forms dimmers in vapor phase, (recommended by ASPEN user manual). Therefore UNIQUAC-HOC is used as property method in designing the Collection flash tank. Since HMF properties are not present in ASPEN plus, during the simulation HMF is assumed to have similar properties to furfural. The conditions for the flash tank are set to 100ºC and 1 atm pressure. In simulation, for simplicity, only liquid components are considered, i.e. acetic acid, furfural, HMF and water. Percent vaporization of acetic acid, furfural, HMF and water is 13.05, 73.5, 73.5 and 20.43% respectively. The vapor flow rate is 36.5 m^3/s and liquid flow rate is 0.106 m^3/s.

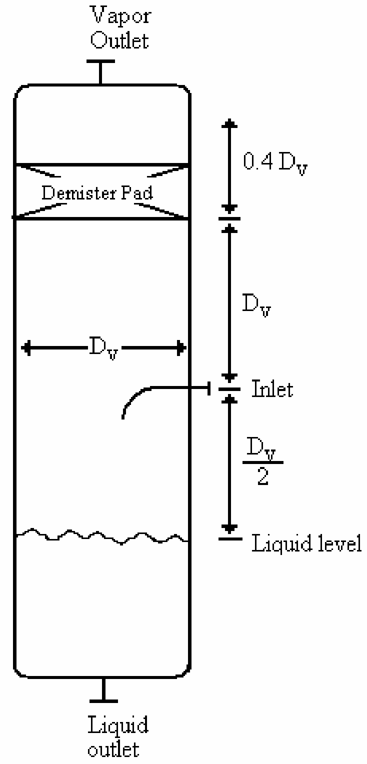


Figure A.9.1 Vertical flash tank

For the Flash Separations a vertical gas liquid separator is used. The dimensions are calculated using the following equations.

$$D_v = \sqrt{\frac{4V_v}{\pi u_s}} \qquad \text{Eqn A.9.3}$$

$$u_t = 0.07\left[\frac{\rho_l - \rho_v}{\rho_v}\right]^{0.5} \qquad \text{Eqn A.9.4}$$

D_v is the minimum vessel diameter (m), V_v is gas or vapour volumetric flow rate (m^3/s), u_s is equal to settling velocity (u_t)

The liquid density is 917 kg/m^3 and vapour density is 0.61 kg/ m^3. The calculated u_t is 2.71 m/s. Using this information, the dimensions are obtained as given in Table A.9.2.

TableA.9.2 Dimensions of V-401

D_v	4.14	m
Vapour section height	6.21	m
Vapour section volume	83.52	m^3
Vapour residence time	2.30	s
Liquid section volume	0.24	m^3
Liquid section height	0.018	m
Total Height	7.9	m

A.9.3. Fermentors (BR-301)

A.9.3.1. The Model

For simplicity in the model the by-products are excluded and it is assumed that all the glucose and pentose is converted into ethanol and biomass. As a result, the reactions for the model are, the reactions given in Equations 8.10,8.11 and 9.9,9.10

The specific rates of the micro organisms are required to mimic the microbial activity of the micro organisms. In diauxy growth, the substrate which yields the highest specific growth rate dominates and consumed first. In current case, hexose (glucose) dominates on pentose (xylose). To implement this into model, inhibition of glucose on xylose is added with a switching function and the specific uptake (q_i) rates are calculated using,

$$q_G = q_G^{\max} \frac{C_G}{K_G + C_G} \qquad \text{Eqn A.9.5}$$

$$q_{Xy} = q_{Xy}^{\max} \frac{C_{Xy}}{K_{Xy} + C_{Xy}} \frac{1}{1 + \frac{C_G}{K_I}} \qquad \text{Eqn A.9.6}$$

q_i is substrate uptake rate in g g biomass^{-1}h^{-1}, C_i is the concentration of substrate i in g/l, K_i is the substrate affinity for substrate i in g/l, q_i^{max} is the maximum uptake rate for substrate i in g g biomass $^{-1}$h^{-1} and K_I is the inhibition constant of glucose on xylose in g/l.

Specific ethanol productivity is determined based on growth and ethanol production reactions. The specific ethanol rates of hexose and pentose metabolisms are as given below

$$q_{EG} = 3.0667\mu_G + 0.51 m_s \qquad \text{Eqn A.9.7}$$

$$q_{EXy} = 3.5528\mu_{Xy} + 0.51 m_s \qquad \text{Eqn A.9.8}$$

q_{Ei} is the specific ethanol productivity on substrate i in g g biomass $^{-1}$h^{-1}, μ_i is the specific growth rate for growth on substrate i in h^{-1} and m_s is the maintenance coefficient in g g biomass^{-1} h^{-1}.

At the beginning of the fermentation, where glucose inhibits the xylose metabolism, the specific ethanol productivity on glucose is used by the model. When the inhibition is lowered and the xylose metabolism is active the specific ethanol productivity for xylose is activated. For simplicity, the maintenance coefficients for xylose and glucose metabolism are assumed to be the same.

Herbert-Pirt equation for mixed substrates is used to determine the specific growth rate of the microorganism, similar to ethanol specific rates the specific growth rate function is also switches growth on xylose after the glucose inhibition is over. The specific growth rates are calculated using,

$$\mu_G = Y_{GX}^{\max} q_G + Y_{XyX}^{\max} q_{Xy} - Y_{GX}^{\max} m_s \quad \text{Eqn A.9.9}$$

$$\mu_{Xy} = Y_{GX}^{\max} q_G + Y_{XyX}^{\max} q_{Xy} - Y_{XyX}^{\max} m_s \quad \text{Eqn A.9.10}$$

$Y_{iX}^{\max}$ and is the maximum biomass yield on substrate *i*.

Using the growth reactions, $Y_{GX}^{\max}$ and $Y_{XyX}^{\max}$ are obtained as 0.1374 and 0.1207 g biomass/g substrate respectively. q_i^{max} can be estimated using the maximum specific growth rates of the micro organism,

$$q_i^{\max} = \frac{\mu_i^{\max}}{Y_{iX}^{\max}} + m_s \quad \text{Eqn A.9.11}$$

The mass balances to be solved for the model are,

Glucose uptake : $\frac{dC_G}{dt} = -q_G C_x$ Eqn A.9.12

Xylose uptake : $\frac{dC_{Xy}}{dt} = -q_{Xy} C_x$ Eqn A.9.13

Growth : $\frac{dC_x}{dt} = \mu C_x$ Eqn A.9.14

Ethanol : $\frac{dC_E}{dt} = q_E C_x$ Eqn A.9.15

The initial conditions are t=0 $C_G = C_{G0}$, $C_{Xy} = C_{Xy0}$, $C_X = C_{X0}$, $C_E=0$.

A.9.3.2. Parameter Estimation

Since the micro organisms are assumed to perform same with the current hexose fermenting organisms, the parameters of hexose fermenting micro organisms are used in model. But, to test the reliability of the proposed model the experimental data provided by Marco Kuyper is used. In the experiments, the initial glucose, xylose and biomass concentrations are 20, 20 and 0.17 g/l respectively. The parameters are estimated by trial and error by changing the parameters manually based on the start up values obtained by applying linear regression to experimental data. The set of data which gives the best fit is presented in Table A.9.3 and the results are shown in Figure A.9.2. The primary reason for the lack of fit in biomass and ethanol data might be the exclusion of by-products from the model. And, the secondary reason might me the maximum 5 % variation in the experimental data reported by the author.

The results show that the model is proper to estimate the mixed substrate metabolism. In second step, the parameters of currently available hexose fermenting strains are input to model to find the fermentation time required. The total hexose (including hydrolyzed sucrose) and pentose concentrations entering the fermentor are obtained from Super Pro designer as, 166.7 and 37.8 g/l respectively. The initial biomass composition of 100 g/l and the initial substrate concentrations are input to model and the system is simulated using the parameters presented in Table 9.4. Only the substrate affinities and the inhibition constant are taken from the experimentally estimated parameters. The results are shown in Figure A.9.3. The total fermentation time required to consume all substrate is obtained as 9 hours.

Table A.9.3Parameters for the microorganism

Parameter	Value
μ_g^{max} (1/h)	0.225
μ_{xy}^{max} (1/h)	0.11
m_s (g substrate/gram biomass h)	0.276
K_{sg} (g/l)	1.5
K_{sxy} (g/l)	8.5
K_I (g/l)	5.4

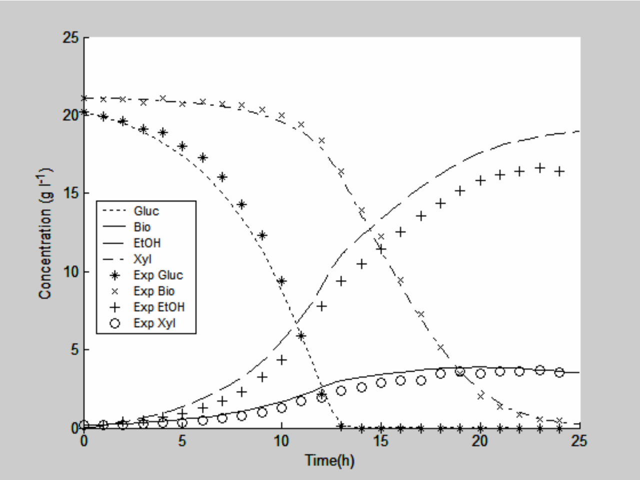


Figure A.9.2 The simulation results for 20 g/l glucose and 20 g/l Xylose. initial biomass is 0.17 g/l

Table A.9.4 Parameters used for fermentation time requirement

Parameter	Value
μ_g^{max} (1/h)	0.0073
μ_{xy}^{max} (1/h)	0.0073
m_s (g substrate/gram biomass h)	0.1799
K_{sg} (g/l)	1.5
K_{sxy} (g/l)	8.5
K_I (g/l)	5.4

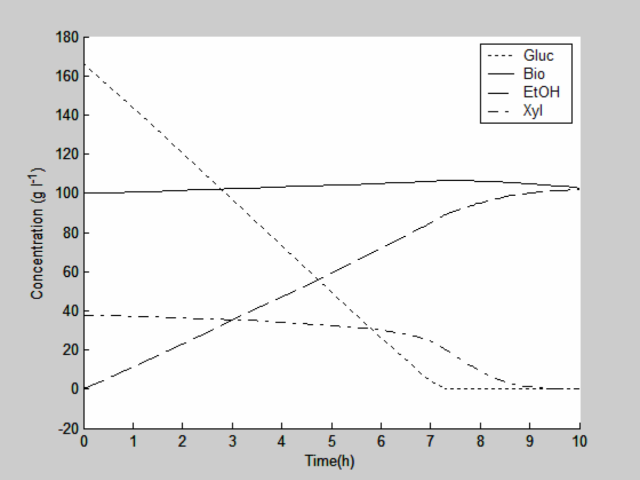


Figure A.9.3 The simulation results for 166.7 g/l glucose and 37.8 g/l xylose initial biomass is 100 g/l with hexose fermenting micro organism parameters.

A.9.3.3. Fermentor Sizing

The fermentors are sized using the same methodology described in Appendix A.8.7.4. Using the mass balances from Super Pro designer total volumetric flow per hour is obtained as 993 m^3/h. The volumetric flow is large, therefore, the flow will be divided in two streams and two sets of fermentors will operate in parallel. The fermentation time will be set to 10 h for safety margin. The same methodology with Section A.8.7.4 is applied to determine the number of vessels in series and the results are given in Table A.9.5.

Table A.9.5 Costs for different number of fermentors

Number	Cycle Time (h)	Volume (m^3)	Total Volume (m^3)	Cost (10^6\$)	Total Cost (10^6\$)
10	12.5	1,851	18,514	1.25	12.5
7	14	2,962	20,736	1.75	12.25
6	15	3,702	22,217	2.05	12.3
4	20	7,406	29,623	3.36	13.4
3	30	13,465	44,435	5.5	16.5

The results show that for 7 fermentors in series, the investment requirements are minimum. The total cycle time is 14 h, which corresponds to 2 h flow in each reactor. The scheduling is as given in Figure 9.6.

The scheduling is done in such a way that as soon as one reactor is filled, the second one is started to be filled and the biomass from the third is also being transferred into the one being filled. To calculate the total volume of one reactor, the contribution of presence of biomass should also be included. In Appendix A.8.7.4, the contribution of biomass to 1 m^3 volume is calculated as 0.356 m^3. Since the flow rate is high the fermentor sizes for 7 fermentors have volume of 2,962 m^3 each. To have smaller fermentor volumes, the flow is separated into two streams and two fermentor sets are operated in parallel. The resultant fermentor volume is obtained as 1,481 m^3 including the 10% safety.

The equipment sizing, mixing time and gas hold up calculations are carried out in the same manner with Appendix A.8.7.4. The volumetric gas flow rate is obtained as 12.6 m^3/s from ASPEN. For H/D ratio of 2, the superficial gas velocity is calculated as 0.04 m/s and the mixing time is obtained as 147 s with a gas hold up of 0.066. Considering the 10 hr of batch time, the mixing is sufficient. The gas hold up is in accordance with the assumed 10% safety. The final vessels dimensions are given in Table 9.19.

A.9.3.4. Cooling

The growing microorganisms generate 65 kJ of energy per mole of biomass. The total amount of biomass produced is 6,047 kg/h. In each set of fermentors, 5 fermentors are in operation at a time (see scheduling Figure 9.6). Therefore, the biomass growth per vessel is 604.7 kg/h. The heat generated due to biomass formation is 443.83 kW/vessel. Part of this energy is removed with the evaporating air and ethanol in off-gas due to heat of evaporation. From Aspen simulation, the rate of ethanol and water evaporation is obtained as 139.5 and 131.1 kg/h/vessel respectively. The heat of evaporation of water and ethanol are obtained as 2,260 and 838 kj/kg. Using these values, the cooling is calculated as 118.1 kW/vessel. Therefore, the amount of heat to be removed is 325.74 kJ/s/vessel. The temperature in the fermentors is to be kept at 30 ºC, due to this reason, the cooling agent is set as chilled water at 10ºC. Taking the heat capacity of the chilled water as 4.1793 kJ/kg ºC and the exit temperature as 25 ºC, the required amount of chilled water is obtained as 5.2 kg/s. LMTD is calculated as 10.82 ºC. The over all heat transfer coefficient for jacketed vessels is given between 200-500 W $m^{-2} {}^{o}C^{-1}$ for water dilute aqueous solutions. Since we have biomass in the fermentor and the mixing is not highly turbulent the lowest margin is taken as 200 W $m^{-2} {}^{o}C^{-1}$ the required cooling area is calculated as 150.5 m^2 for each vessel.

A.9.4. Molecular Sieve Adsorption Column (C-304)

Design of the molecular sieve adsorption unit is carried out the same way as 2005 plant. The operation conditions, parameters and the equilibrium expressions are as given in Section 8.7.4 & Appendix A.8.12.

The concentration and the volumetric flow rates are 54.3 kg/m^3 and 0.039 m^3/s respectively. The operation time is set to 2 h 10 min (7,800 s) for each absorbent bed. The properties and parameters required by the MATLAB program is listed in Table A.9.6.

Table A.9.6 Parameters input to MATLAB program

Parameter	Value	Unit
L (m)	3-9 (Manipulated)	m
D (m)	1-4 (Manipulated)	m
Bed Porosity	0.4	
Zeolite adsorbent diameters	0.002	m
Liquid flow rate	0.0165	m^3.s^{-1}
Molecule diameter of water	2.8x10^{-10}	m
Diffusivity of water	5.7x10^{-9}	m^2.s^{-1}
Density of solution	725	kg.m^{-3}
Viscosity of solution	0.011	Pa.s
Water concentration	54.25	kg.m^{-3}

Diameter and the height of the bed are manipulated to get the necessary break even point. First the diameter is set to 4 m to get the same flow velocity with the design in Section A.8.12. However, the retention time (5,800 s) was lower then the design specification (7,800 s). Therefore, diameter is reduced and length is increased until the desired retention time is obtained. The break even point is determined as 5.4 kg/m^3 which is 10% of the inlet concentration. The simulation results are presented in Figures A.9.4&5. Setting the break even point to 5.4 yields an average concentration of 0.8 kg/m^3 which yields and outlet composition of 99.89% ethanol (see Figure A.9.5-A, the concentration that gives $A_1=A_2$ is taken as the average concentration) the resultant dimensions of the adsorption column is obtained as 3.2 m diameter and 8.4 m length. Using this information the total amount of required adsorbent material is calculated as 49.3 ton. The ratio of unused bed to used bed is obtained as 30% (Figure A.9.5-B $A_4/[A_3+A_4]$).

A.9.4.1. Cooling

Cooling is required to remove the heat of adsorption and to keep the adsorbent bed temperature at 60ºC. The heat of adsorption is 2,365 kJ/kg water. For adsorption rate of 7,672 kg water /h, the heat duty of the cooling system is obtained as 5,040 kW. Cooling water at 25ºC is used as cooling agent and the exit temperature of the cooling water is set to 55ºC. The corresponding cooling water requirement is obtained as 144.6 t/h.

The overall heat transfer coefficient for water-organic solvents is given in the interval 250-750 W m^{-2o}C^{-1} (Sinnot, 1999). A value of 750 W m^{-2o}C^{-1} is taken for the calculations. The LMTD is calculated as 15.4ºC. Using the LMTD and overall heat transfer coefficient, the area requirement is obtained as 436 m^2.

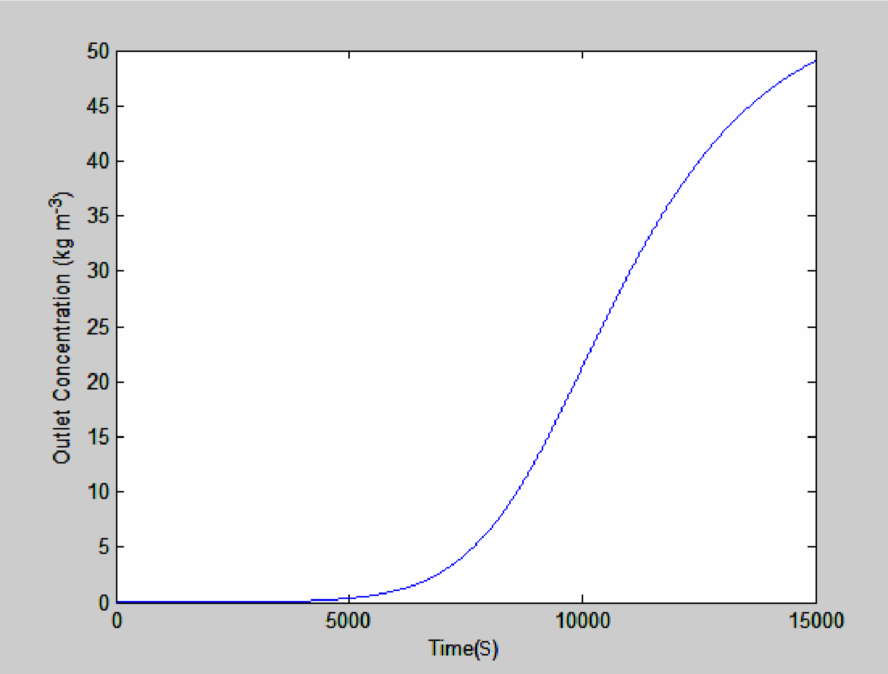


Figure A.9.4 Outlet concentration profile of molecular sieve adsorption column

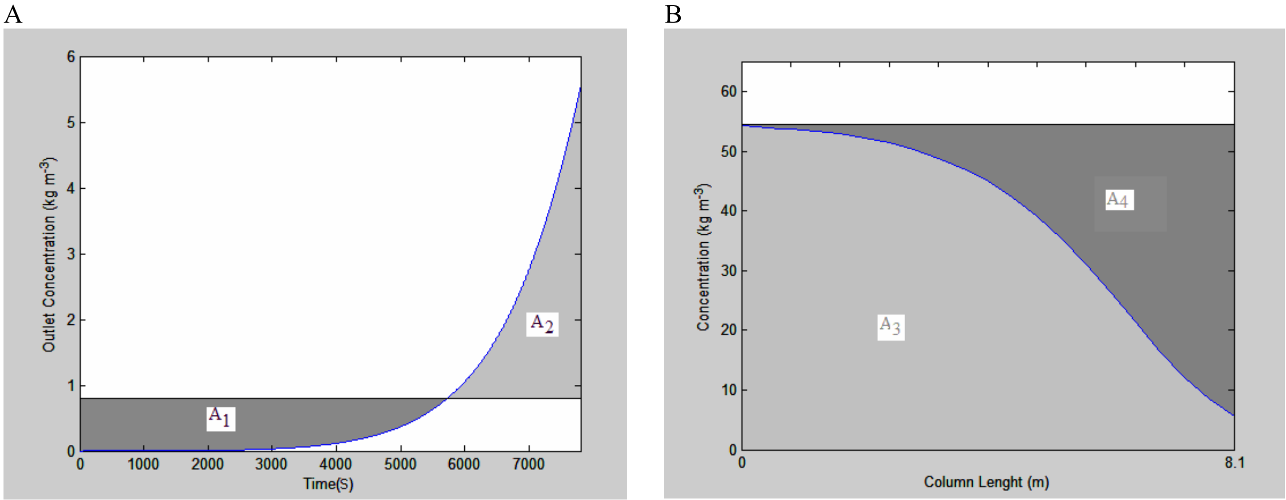


Figure A.9.5. Breakeven point (A) and concentration profile in column (B)

A.9.4.2. Regeneration

Due to the reasons stated in Section A.8.12.2, the regeneration temperature is set to 200 ºC. The regeneration gas is nitrogen. The total gas flow rate is of 21 m^3/s (23 kg/s = 82.8 t/h). The make up nitrogen is assumed to be 0.1% of the total requirement (346 t/year).

The spent nitrogen can be dehydrated to be reused in the process. Taking the nitrogen flow as 82.5 t/h and the water to be desorbed as 7.7 t/h, the composition of the moist nitrogen is obtained as 9.3% water. The dehydration is carried out by condensing the water by refrigerant reagents at low temperatures. ASPEN simulation shows that the water content of the nitrogen stream can be lowered from 9.3% to 0.5 % by condensing at 5ºC.

Desorption heating

Desorption temperature is set to 200 ºC. Nitrogen gas is assumed to enter the column at 180ºC. The heat of evaporation of water at 200ºC is 1,933 kj/kg. Using this value, the heat required to evaporate 7.7 t/h of water is obtained as 4,119 kW. Using nitrogen heat capacity given in Equation A.8.40, heating required to heat up nitrogen from 180 to 200ºC is obtained as 703 kW. The heating agent is superheated steam from the boilers at 275ºC and 25 MPa. The enthalpy of condensation for superheated steam is 1,985 kJ/kg which yields steam requirement of 8.75 t/h. The mean temperature difference is 45ºC and the overall heat transfer coefficient is taken as 300 W m^{-2}ºC^{-1} (Sinnot, 1999). Using total power requirement of 2,091 kW, the resultant heat exchanging area requirement is calculated as 357 m^2. This area requirement is lower then the cooling area requirement; therefore, the cooling system will be switched to heating system during the regeneration.

Nitrogen heater

To recover the heat and to save refrigeration energy the dry nitrogen from condenser with temperature of 5ºC is heated up to 180ºC using the moist nitrogen from the regeneration as hot utility. Meanwhile, the moist nitrogen is cooled down from 200ºC to 30ºC.

The heat duty to heat up the nitrogen is calculated as 6,158 kW. The LMTD is 23 ºC and the overall heat transfer coefficient for gas-gas system is taken as 300 W m^{-2}ºC^{-1}. The resultant heat exchanging area requirement is calculated as 893 m^2.

The moist nitrogen is further condensed in condenser at 5ºC. ASPEN simulation shows that the refrigeration requirement of condenser is 1,977 kW. The LMTP for condenser is 14 ºC and the overall heat transfer coefficient is assumed based on air gas-brine systems. The corresponding heat transfer coefficient is taken as 250 W m^{-2} º$C^{-1.}$ The resultant heat exchanging area is obtained as 565 m^2.

A.9.5. Cogeneration

A.9.5.1. Evaporator (E-403)

The vinasse is first concentrated to 65% water in reboiler of the distillation column then, further concentrated to 32% moisture in the evaporators. The evaporation percentages of the volatile components in vinasse are given in Table A.9.7.

Table A.9.7 The percent evaporation of the compounds and their heat of evaporations

Component	Evaporation (%)	ΔH_E (kJ/kg)
Acetic Acid	71.2	395
Ethyl Alcohol	78.4	842
Furfural	72.2	462
Glycerol	12.7	644
HMF	72.2	462
isoamylalcohol	75.0	516
Lactic Acid	20.9	840
Succinic Acid	12.6	799
Water	76.2	2,264

The heat duty of the evaporator is calculated using the normal boiling point heat of evaporations (from ASPEN database) of the individual components and found as 19,374.8 kW. The evaporator operates at atmospheric pressure and the heating agent is low pressure saturated steam at 130ºC. The enthalpy of steam at this temperature is 2,171 kJ/kg. Therefore, the required amount of steam is 32.3 t/h. To calculate the heating surface requirement for the evaporator, the overall heat transfer coefficients of 5th effect juice evaporator (E-205) is taken as basis since the water content of vinasse is similar to 5th effect juice evaporator (E-205). The heat transfer coefficient is 979 W m^{-2} ºC^{-1}. The boiling temperature of the evaporator contents is taken as 103 ºC. Thus, the mean temperature difference is 27.7 ºC. Using this values the area required for the evaporator is calculated as 714.4 m^2.

A.9.5.2. Fibre, Yeast and Filter Cake Drier (D-401)

The exhaust air from the furnace (F-401) is used to dry the fibre residues, yeast and filter cakes (Solid fuel). The temperature of the exhaust is set to 350 ºC and the temperature of the dry solid fuel leaving the drum is set to 100ºC. Based on this assumptions, and the heat capacity information the enthalpies of each stream entering and leaving the drum is calculated and the exit temperature of the exhaust air is obtained as 124.8 ºC (See Table A.9.8). During the enthalpy balances, the reference temperature is set to 30 ºC and the enthalpy of evaporation of water is taken as 2,260 kJ/kg.

TableA.9.8 Energy balances for drying chamber D-101

	In		**Out**	
	Solid fuel	**Flue gas**	**Solid Fuel**	**Flues Gas**
Flow (kg/s)	57.97	252.57	27.69	486.51
Heat Capacity (kJ kg^{-1}ºC^{-1})	2.851-0.0003T	1.133+0.00015T* 1.303+0.00011T*	1.581-0.00007T	1.204+0.00014T* 1.573+0.00004T*
Enthalpy (kJ/s)	4,744.5	139,488.8	3,056.3	141,177.0
Temperature (ºC)	61	350	100	124.8

* The heat capacities are given for the cases of water as liquid (0-100ºC) and water as vapour (>100ºC)

The total amount of evaporated water is 94.6 ton/h. Using the evaporation rate of 65 kg m^{-2}h^{-1}, the area requirement for the drum driers is obtained as 1,455 m^2.

A.9.5.3. Net Heating Value of Feed Fuel and Steam Production

The heat generated by burning lignin, hemicellulose, cellulose, cellulase and lignin degradation products are assumed to have same heat of combustion with bagasse which is given as 19,260 kJ/kg (Hugot, 1972). Microbial biomass is assumed to have heat of combustion of sucrose given by Hugot (1972). The heat of combustion of the organic compounds obtained from Perry's Chemical Engineering Handbook (Perry and Green, 1997) and Handbook of Chemistry and Physics (Weast, 1984) and listed in Table A.9.9.

The combustion reactions are as given in following reactions. The lignin, hemicellulose and cellulose are assumed to be composed of 48.3% C, 6.3% H and 45.4% O.

Ethanol : $C_2H_6O+3O_2 \rightarrow 2CO_2+3H_2O$ Eqn A.9.16
Acetic Acid : $C_2H_4O_2+2O_2 \rightarrow 2CO_2+2H_2O$ Eqn A.9.17
Glycerol : $C_3H_8O_3+3.5O_2 \rightarrow 3CO_2+4H_2O$ Eqn A.9.18
Furfural : $C_5H_4O_2+5O_2 \rightarrow 5CO_2+2H_2O$ Eqn A.9.19
HMF : $C_6H_6O_3+6O_2 \rightarrow 6CO_2+3H_2O$ Eqn A.9.20
Succinic acid : $C_4H_6O_4+3.5O_2 \rightarrow 4CO_2+3H_2O$ Eqn A.9.21

Isoamylalcohol : $C_5H_{12}O + 7.5O_2 \rightarrow 5CO_2 + 6H_2O$ Eqn A.9.22

Lactic acid : $C_3H_6O_3 + 3O_2 \rightarrow 3CO_2 + 3H_2O$ Eqn A.9.23

Xylitol : $C_5H_{12}O_5 + 5.5O_2 \rightarrow 5CO_2 + 6H_2O$ Eqn A.9.24

Impurities : $CH_{1.1}ON_{0.04}S_{0.001} + 0.786O_2 \rightarrow CO_2 + 0.55H_2O + 0.02N_2O + 0.001SO_2$ Eqn A.9.25

Yeast : $CH_{1.8}O_{0.5}N_{0.2} + 1.25O_2 \rightarrow CO_2 + 0.9H_2O + 0.1N_2O$ Eqn A.9.26

Table A.9.9 Heats of combustion for the compounds present in fuel

Component	ΔH_C (kJ/kg)
Fibres	19,260
Acetic Acid	17,709
Ethyl Alcohol	26,847
Furfural	50,924
Glycerol	18,067
HMF	50,924*
Isoamylalcohol	36,108
Lactic Acid	15,203
Succinic Acid	12,644
Xylitol	15,603
Impurities	17,166**
Microbial Biomass	16,559

* Assumed based on furfural
**Hugot(1972)

Based on the reactions, how much combustion products are produced and how much oxygen is required when 1 kg of organic material is burnt is calculated. The results are presented in Table A.9.10.

There are two flows in to furnace, one is concentrated syrup of vinasse and the other is the dehydrated filter cakes from rotary filters (S-103 and S-403), biomass and fibres mixture. The assumption is the moisture content of the solids is reduced to 12% as in 2005 plant. The content of these two flows and the heat generated by combustion are given in Table A.9.11.

Table A.9.10 Combustion products of the compounds present in fuel

Compound	O_2 (kg/kg)	CO_2 (kg/kg)	H_2O (kg/kg)	N_2O (kg/kg)	SO_2 (kg/kg)
Ethanol	2.087	1.913	1.174	-	-
Acetic Acid	1.067	1.467	0.600	-	-
Glycerol	1.218	1.435	0.783	-	-
Furfural	1.667	2.292	0.375	-	-
HMF	1.524	2.095	0.429	-	-
Succinic acid	0.949	1.492	0.457	-	-
Isoamylalcohol	2.727	2.500	1.227	-	-
Lactic acid	1.067	1.467	0.600	-	-
Xylitol	1.158	1.447	0.711	-	-
Impurities	0.847	1.482	0.333	0.030	0.002
Fibre	1.338	1.771	0.567	-	-
Bagasse	1.3051	1.716	0.549	0.013	0.0002
Sucrose	1.223	1.544	0.579	-	-
Biomass	1.626	1.788	0.659	0.179	-

Table A.9.11 The available heat in fuel after drying

Compound	ΔH_C-ΔH_E (kJ/kg)	Syrup (kg/h)	Syrup Heat (kW)	Fibres (kg/h)	Fibre Heat (kW)
Acetic Acid	17,314	39.2	188.7	0.0	0.0
Ash	0	5,765.9	0.0	870.0	0.0
Bagasse	19,260	0.0	0.0	1,676.3	8,967.9
Yeast	16,559	0.0	0.0	6,047.3	27,815.9
Ca-salts	0	0.0	0.0	4,261.2	0.0
Cellulase	19,260	6.6	35.2	1,310.4	7,010.7
Cellulose	19,260	31.4	167.8	6,557.0	35,080.1
Dirt	0	0.0	0.0	11,973.2	0.0
Ethyl Alcohol	26,005	0.0	0.1	0.0	0.0
Furfural	50,462	0.0	0.0	112.4	1,575.1
Glycerin	17,423	1,162.6	5,626.5	0.0	0.0
Hemicellilose	19,260	25.4	136.0	5,085.8	27,208.9
HMF	50,462	0.0	0.0	20.6	288.2
Impurity	17,166	17,338.5	82,675.6	8,105.7	38,650.7
isoamylalcohol	35,592	5.5	54.4	0.0	0.0
Lactic Acid	14,363	59.4	237.1	0.0	0.0
Lignin	19,260	197.7	1,057.4	39,530.4	211,487.5
Lignin monomers	19,260	1,380.1	7,383.7	710.5	3,801.1
Polyacrylamide	19,260	0.0	0.0	2.4	12.9
Succinic Acid	12,644	120.7	424.1	0.0	0.0
Sulfuric Acid	0	0.6	0.0	0.0	0.0
Sucrose	16,559	0.0	0.0	1,709.0	7,860.8
Water	-2,512	9,586.7	-6,689.4	11,726.8	-8,182.7
Xylitol	15,603	56.3	243.9	0.0	0.0

The calculations yields total gross heat generation of 453.1 MW heating and the contribution of vinasse syrup to this energy is 91.5 MW. The power generation is calculated including the heat of evaporation of the water as negative contribution. The heats of evaporations of the volatile components (Table A.9.8) are reduced from their heats of combustion (Table A.9.9). To calculate the available heat for steam production, the heat losses for heating up the inflowing air and the fuel contents is to be calculated.

The Required Amount of Air

The amount of exhaust gas and inflowing air is to be known to calculate the heat loss with the exhaust gas. The ratio of excess air to theoretical air requirement is taken as 1.51 in accordance with the 2005 design. The theoretical minimum air requirement for the complete combustion of the feed is calculated as 527.6 ton/h. Based on this, the excess air requirement is obtained as 796.7 t/h which yields exhaust gas flow of 912.6 ton/h. The fresh air is heated up using the exhaust air from the furnace to 1031 ºC and the rest of the heating occurs in furnace (See Table A.9.13).

The available heat is calculated based on the enthalpy balances around the furnace given as,

$$\Delta H_{Available} = \Delta H_{fiber\&filter\ cake(12\%)} + \Delta H_{vinasse(32\%)} + \Delta H_{Fresh\ air} + \Delta H_{Fuel\,burning} - \Delta H_{Flue\,gas} - \Delta H_{Ash\&solids}$$

The enthalpies are calculated using the reference temperature of 30ºC. The incoming and out going enthalpies for the heat balances in furnace is given in Table A.9.12.

Thus, total available energy for steam production is obtained as,

$$Q_{Steam} = 453{,}118.5 + 3{,}056.3 + 235{,}628.0 + 1{,}462.7 - 332{,}492.8 - 6{,}406.9 = 354{,}365.9kW$$

Economizer

The economizer is the heat exchanger in which the flue gas is get in contact with cold fresh air to recover the lost heat in flue gas by heating up the fresh air. The air coming in to the furnace is preheated to 1,031 ºC in economizer by getting in contact with flue gas. The exit temperature of the flue gas is set to 350 ºC.

The *LMTD* of the economizer is obtained as 163 ºC and the *OHTC* for flue gas systems is taken as 0.1 kW m^{-2}ºC^{-1} based on the values given by Sinnot (1999). Using these values and the heat duty of 235.6 MW, the required area for heat exchanging is obtained as 14,455 m^2.

Table A.9.12 Heat loss calculations for furnace F-103

Component	Flow rate (kg/s)	Heat Capacity (kJ kg^{-1}ºC^{-1})	Temperature (ºC)	Enthalpy (kJ/s)
$\Delta H_{Fibre\ \&\ filter\ cake}$	27.69	1.581-0.00007*T*	100	3,056.3
$\Delta H_{Vinasse}$	9.94	2.027-0.00020*T*	103	1,462.7
$\Delta H_{Fresh\ air}$	221.5	0.993+0.00013T	1031	235,628.0
$\Delta H_{Flue\ gas}$	447.6	1.133+0.00015T** 1.303+0.00011*T***	1,100	332,492.8*
ΔH_{Ash}	6.37	0.94	1,100	6,406.9
Fuel combustion			1,100	453,118.5

* The latent heat of water present in the flue gas is excluded since it is already counted in fuel burning.

** The heat capacities for vapour (T>100ºC) and liquid water (T<100ºC).

Table A.9.13 Heat balances for the economizer

Component	Flow rate (kg/s)	Heat Capacity (kj/kg ºC)	Initial Temp. (ºC)	Final Temp. (ºC)	Total Heat (MW)
Fresh air	221.5	0.993+0.00013T	30	1031	235.6
Exhaust air	252.6	1.133+0.00015T	1100	350	-235.6

Steam Production

In 2015 design, the amount of produced steam is set according to the demand of the plant. In the plant two steam qualities is required, the major need is low pressure steam at 275 kPa, the properties of which is given in Table A.9.14.

Table A.9.14 Properties of low pressure steam at 275 kPa and 130.7ºC

Parameter	Value
Temperature (ºC)	130.4
Pressure (bar)	2.75
Enthalpy of saturated liquid (kJ/kg)	549
Enthalpy of saturated vapour (kJ/kg)	2720

The incoming water to boilers is assumed to have temperature of 120 ºC (condensates from the heating equipment). Under this conditions the heating required to heat up 1 kg of water at 120ºC to saturated steam at 275 kPa is obtained as,

First heat liquid water to 130 ºC, $Q_{90-300^oC} = \int_{120}^{130} (4.189 - 0.0006T)dt = 41.1 kJ/kg\ water$

Then, evaporate at 130.7 ºC, $Q_{evaporation} = (2,720 - 549) kJ/kg = 2,171 kJ/kg\ water$

The total energy requirement for 1 kg of water is obtained as 2,212.1 kJ/kg.

The quality of medium pressure steam for pretreatment reactor is given in Table A.9.15.

Table A.9.15 Properties of medium pressure steam at 25 bar and 275ºC

Parameter	Value
Temperature (ºC)	275
Enthalpy of saturated liquid at 224ºC and 25 bar (kJ/kg)	961.96
Enthalpy of superheated vapour (kJ/kg)	2,947.4

The heat required to produce 1 kg of steam at 25 bar and 275ºC is calculated in similar manner and obtained as 2,399.6 kJ/kg.

Also the water vapours from the pretreatment flash tank (V-401), vinasse evaporator (E-403) and clarification flash vessels (V-102&103) are super heated and compressed to low pressure steam at 130.7 ºC and 275 kPa. The total amount of water vapour at 100ºC produced in the plant is 124 t/h. the energy required for this operation is obtained as 2.03 MW using.

$$Q_{100-130.7^oC} = \int_{100}^{130.7} (1.912 + 0.00004T)dt = 58.84 kJ/kg\ water\ vapor$$

The medium pressure (25 bar) steam demand of the plant is 102.5 t/h (for pretreatment and molecular sieve regeneration). The energy requirement to produce this amount is 68.32 MW.

For total available energy of 354.37 MW, 70.35 MW is used and the remaining 284.02 MW is utilized to produce low pressure steam. Using the energy requirement of 2,212.1 kJ/kg , the amount of produced steam is calculated as 462.21 t/h.

Therefore, the total steam generation is 688.71 ton/h. However, the steam demand of the plant is 614 ton/h, 105 ton/h at 25 bar and the rest is at 2.75 bar.

The steam production can be lowered and some of the steam can be produced at higher pressure and temperature to be used in electricity generation. For this, the proposed set up is as given below,

- Steams from process equipments will be heated up the 130.6 ºC at 275 bar (total 124 ton low pressure steam/h). This operation requires 2.03 MW energy.
- To meet the remaining low pressure steam requirement, 388 ton/h more steam is to be generated. This requires 238.11 MW energy.

- And the remaining energy (114.23 MW) is used to produce of high pressure and temperature steam for electricity production. The exit steam quality from the turbine generator will be 25 bar and 275 ºC which is the steam quality for pretreatment and molecular sieve regeneration

Based on this information the quality of steam produced for electricity generation is taken as given in Table A.9.16.

Table A.9.16 Properties of superheated steam at 86 bar and 525ºC

Parameter	Value
Temperature (ºC)	525
Enthalpy of Saturated Liquid at 300ºC and 86 bar (kJ/kg)	1,345.4
Enthalpy of Superheated Vapor at 525 ºC (kJ/kg)	3,453.0

The energy required to produce 1 kg of steam at 86 bar and 525 ºC is obtained as ,

First heat liquid water to 300 ºC, $Q_{120-300^{o}C} = \int_{120}^{300} (4.189 - 0.0006T)dt = 731.34\, kJ/kg\ water$

Then evaporate at 300 ºC and super heat to 525 ºC,
$Q_{300-525^{o}C} = (3,453 - 1,345.5) kJ/kg = 2,107.6\, kJ/kg\ water$

The total energy requirement for 1 kg of water is obtained as 2,838.94 kJ/kg. For total available energy of 114.23 MW, the total amount of produced steam is calculated as 144.85 t/h.

A.10. Economical Evaluation: 2005 Plant

The sources used during the equipment purchase cost estimations are Sinnott R.K. (1999), Peters and Timmerhaus (1991), Super Pro Designer 5.1 cost analysis tool and Matche online database (IRef-4). Majority of the prices are calculates using the cost model given in Equation A.10.1. (Sinnot, 1999).

$$C_e = F_m \times F_p \times C \times S^n \qquad \text{Eqn A.10.1}$$

C_e is purchased equipment cost ($), F_m is the material factor, F_p is the pressure factor, S is the characteristic capacity parameter for that equipment and n is index for that type of equipment

The prices referring to Peters and Timmerhaus (1991) are from year 1990, the prices taken from Matches (IRef-4) are from year 2003, the prices taken from super pro designer are adjusted to 2005, and that taken from Sinnot, (1999) are from 1998. During the investment cost calculations the prices will be converted to current values using the US inflation rate.

The prices of the pumps are determined based on centrifugal pumps by using Super Pro Designer database. The pumps have one spare; therefore, the costs are doubled during the calculations.

A.10.1. Mill (M-101)

Milling Equipment

There are 5 mill tandems each has 1 crusher followed by 5 roller mills. The capacity of the crushers is the highest with 250 ton/h raw sugar cane. But, after the crusher the juice is extracted and only wet bagasse is processed in roller mills this is why the capacity of the roller mills is 71.4 t/h (48% moisture).

The price of the milling equipment is determined based on the roll crushers. Peters and Timmerhaus (1991), reports the price of the roll crushers with capacity of 250 ton/h including the motor and drive as $280,000. Thus, the price of the crushers is taken as this value. The price of the roller crushers with capacity of 72 ton/h is reported as $75,000 (including the motor and the drive) by the same source.

Cane Carriers

The prices of cane and bagasse carriers are determined based on belt conveyors. Their lengths are taken as 30 m each. The capacity of belt conveyors with 0.75 m width is given between 240-641 ton/h for speeds varying between 45 m/s to 160 m/s. For capacity of 250 ton cane/h the width of 0.75 m is sufficient. The bagasse flow is 70 ton/h, therefore, the belt width of 0.4 m is sufficient since the capacity for 0.4 m wide belt filters are given in the interval of 65 to 215 ton/h for speeds varying between 45 m/s to 160 m/s. The prices for the conveyor belts are calculated using the correlation formula and the parameters given by Sinnot, (1999) The parameters are given for 0.5 m and 1 m wide conveyors. To obtain 0.75 m, the average of *C* is taken (Table A.10.1).

There are also intermediate carriers between each mill equipment which corresponds to 5 conveyors for each tandem (25 total). The lengths of each conveyor are taken as 3 meters and the width is same with the bagasse carriers.

Table A.10.1 The prices of the conveyors

	Width(m)	S	Unit	F_m	F_p	C ($)	n	Price($)
Cane carriers	0.75	30	m	1	1	2,400	0.75	30,995
Bagasse carriers	0.4	30	m	1	1	1,520	0.75	19,630
Intermediate carriers	0.4	3	m	1	1	1,520	0.75	3,465

Cane Knives

There are 2 knife sets for each tandem and the prices for each knife are taken as one third of the roller crusher with the capacity of 250 ton/h. There reasoning behind that is each knife sets has one rotating shaft while mills has three although the processing capacity is the same. Thus, the price for a knife set is taken as $90,000.

Magnetic Separators

There is one magnetic separator per mill tandem. The price is taken as $10,400 based on the value reported by Aden *et al,*(2002).

A.10.2.Juice Clarification

A.10.2.1. Lime Kiln (F-101)

The cost of the lime kiln is determined based on the process furnaces. The heat duty of the kiln is 1,015 kW. The coefficients for process furnaces to be used in Eqn. A.10.1 are as given in Table A.10.2 and the cost is calculated as $74,350 (Sinnot, 1999).

Air Blower

The capacity of the blower is 28.2 m^3/min. The cost of centrifugal fan of this capacity is reported as $410 (tube-axial fan) (Peters and Timmerhaus, 1991).

A.10.2.2. Sulphur Burner (F-102)

Burner

The cost of the sulphur burner is determined based on process furnaces. The heat duty is 1,003 kW. The coefficients for the calculations are taken from Sinnot, (1999) (See Table A.10.2). Since SO_2 is acidic the material factor is taken as 2 for stainless steal. The cost of the equipment is obtained as $147,344.

Air Blower

The capacity of the blower is 47.67 m^3/min. The cost of centrifugal blower of this capacity is reported as $460 (tube-axial blower) (Peters and Timmerhaus, 1991).

Heavy duty SO_2 Blower

The capacity of this blower is 53.22 m^3/min. The cost of centrifugal heavy duty blower of this capacity is reported as $18,500 (Peters and Timmerhaus, 1991).

Table A.10.2 The prices of the furnaces

	S	Unit	F_m	F_p	C($)	n	Price($)
Lime Kiln	1,015	kW	1	1	360	0.77	74,350
Sulphur Burner	1,003	kW	1	2	360	0.77	147,344

A.10.2.3. Sulphitation Tower (C-101)

The price of the sulphitation tower is determined based on the column shell cost and the packing cost.

Column

The dimensions of column are obtained as 2.3 m height and 2.8 m diameter. Since the column is operating at atmospheric pressures and the column is not high, the thickness of the shell is taken as 0.75 cm. Peters and Timmerhaus (1991) reports the costs of the column towers (excluding interiors) in terms of their weights. To determine the weight of the column, the surface area and the density is required. The density of steel is 7.8 (Perry and Green, 1997). The surface area is calculated using side area which is calculated as described in Eqn A.10.2 and the area of one head is calculated using Equation A.10.3. The area of the heads is calculated by assuming that heads have the area of one third of the hemisphere which has same diameter with the column.

$$A_{Side} = \pi D \times L \qquad \text{Eqn A.10.2}$$

$$A_{Head} = \frac{\pi D^2}{6} \qquad \text{Eqn A.10.3}$$

Using these equations, the total surface area is calculated as 28.44 m^2. Using shell thickness of 0.75 cm the volume and weight of the required steel is calculated as 0.21 m^3 and 1.66 ton respectively. But, the total area will be taken 10% more to take the contribution of the supports and skirt into account. Thus, total column weight is obtained as 1.83 ton. The price of the column with this weight is obtained as $42,000 for 304SS column including supports and the skirt.

Packing

The cost of the ceramic rasching ring packings with 50 mm size is reported as $334.6/$m^3$ (peters &Timmerhaus,1991). The total packing requirement is 14.2 m^3. Thus, the cost of packing is obtained as $4,738.7.

A.10.2.4. Lime Milk Preparation Vessels (V-101)

Vessels

The capacity of the vessels is 6.9 m^3. The cost of mixing tanks for this capacity is given as $21,000 including the drive for the agitator. (Peters and Timmerhaus, 1991).

Cooling

To account for the cooling the cost of cooling coils is assumed to be the half the cost of double pipe heat exchanger with the same area. The cost of the stainless steal double pipe heat exchanger with 6.2 m^2 heat exchanging area is $2,050, therefore, the cooling system is assumed to cost $1,025 for each vessel (Peters and Timmerhaus, 1991).

A.10.2.5. Liming Vessels (V-102&V-103)

Vessels

The capacities of the vessels are 400 and 216 m^3 respectively for V-102 and V-103. The cost of mixing tanks for this capacity is given as $210,000 and $150,000 including the drive for the agitator. (Peters and Timmerhaus, 1991).

A.10.2.6. Juice Heaters (E-101&E-102)

The juice heaters have areas of 317 and 165 m^2 respectively for E-101 and E-102. The costs are estimated based on shell and tube heat exchangers. The cost of is determined based on stainless steal tubes (SS-304) and carbon steel shell as $171,300 and $100,900 respectively (Mathces, IRef-4).

A.10.2.7. Flash Tanks (V-104&V-105)

The costs of the flash vessels are determined based on vertical pressure vessels. The costs are given based on diameter and height. The heights are 2 meter and the diameters are 3.9 and 2.9 m respectively. The corresponding costs for carbon steal tanks operating at atmospheric pressures are obtained as $10,500 and $9,000 respectively (Sinnot, 1999).

A.10.2.8. Clarifiers (V-104&V-105)

Aden *et al.* (2002) determined the cost of clarifier that processes 97 ton/h of inflow as $124,991. In current design each equipment processes 154.8 ton/h. Therefore, the cost of the equipments will be determined based on their capacities. Same source reports the scaling factor for the cost determination as n=0.51. The cost of one clarifying equipment is calculated as $158,630 using the following equation, (Base cost is the cost of the equipment with known capacity)

$$Cost = Cost_{base}\left(\frac{Capacity}{Capacity_{base}}\right)^{n} \qquad \text{Eqn A.10.4}$$

A.10.2.9. Rotary Vacuum Filter (S-103)

The costs for the rotary vacuum filters are determined using the formula given in Equation A.10.4. Perry and Green, (1997) reports the base capacity, base cost and the exponent as 9.3 m^2 and $63,300 and 0.48 respectively. Using these values and filter area of one filter as 180 m^2, the cost of one filter equipment is obtained as $262,460 including the motor.

A.10.3.Evaporation and Crystallization

A.10.3.1. Evaporators and Boiling Pans

(E-201, E-202, E-203,E-204, E-205, E-206, E207, E-208)

The costs of the evaporators and boiling pans are determined based on vertical tube evaporators. The coefficients for Equation A.10.1 are as given in Table A.10.3 (Sinnot, 1999). The material used is cast iron thus the material factor is taken as 1.

Table A.10.3 The prices of the evaporators and boiling pans

	Area	Unit	F_m	F_p	C ($)	n	Price($)
E-201	1,879.0	m^2	1	1	11,500	0.53	624,999
E-202	1,348.5	m^2	1	1	11,500	0.53	524,226
E-203	1,561.5	m^2	1	1	11,500	0.53	566,598
E-204	1,836.5	m^2	1	1	11,500	0.53	617,466
E-205	2,229.0	m^2	1	1	11,500	0.53	684,221
E-206	2,991.0	m^2	1	1	11,500	0.53	799,614
E-207	1,566.5	m^2	1	1	11,500	0.53	785,998
E-208	891.5	m^2	1	1	11,500	0.53	440,981

A.10.3.2. Crystallizers (V-201&V-202)

The cost of crystallizers are determined based on carbon steel agitated tanks, the cooler cost will be determined separately. There are 6 crystallizers in each set. The volumes of the vessels are 123.5 and 110.1 m^3 respectively for V-201 and V-202. The corresponding costs of the equipments are obtained as $73,000 and $70,000 per crystallizer respectively for V-201 and V-202 including the drive for the mixing (Peters and Timmerhaus, 1991).

Cooling

The cooling equipment cost is calculated based on the assumption that the cost will be parallel with the cost of pipes having the same heat exchanging area. The heat exchanging areas are 247 and 220 m^2 respectively for V-201 and V-202. The outer diameter of the tubes is taken as 5 cm. The required pipe lengths to supply the necessary areas are obtained as 1,572.5 and 1,400.6 m respectively for V-201 and V-202. Perry and Green, (1997) reports the cost of piping of specified diameter based on the Equation A.10.4. The base capacity is 152 mm, the base cost is \$30.6/m the exponent is 1.33. Thus, cost of 1 m of piping with 5 cm diameter is obtained as \$6.9. The piping costs are \$10,850 and \$9,664 per crystallizer. The cost of cooling installation is assumed to be twice the cost of the piping. Thus, the total cost of cooling installation is obtained as \$21,700 and \$19,328.

A.10.3.3. Centrifuges (S-201&S-202)

Sinnot, (1999) reports the prices of the centrifuges based on the diameter. The coefficients for Equation A.10.1 are given as *C*= \$58,000 and exponent is 1.3. The capacity (*S*) is the bowl diameter. The diameters of the centrifuges are 0.81 m each. The corresponding cost is \$44,102 per centrifuge.

A.10.3.4. Molasses Reheaters (E-209, E-210, E-211&E-212)

The Molasses Reheaters have areas of 574, 204, 271 and 98 m^2 respectively for E-209, E-210, E-211 and E-212. The costs are estimated based on shell and tube heat exchangers. The cost of is determined based on carbon steal tubes steel shell as \$88,000, \$43,500, \$51,000 and \$29,000 respectively (Sinnot,1999).

A.10.3.5. Rotary Drum Drier (D-201)

Perry and Green (1997) reports the costs of the rotary driers based on the Equation A.10.4. But, the prices are excluding the motor drive. The same source also gives the costs of the engines of given power. The capacity used in the calculations for drum and the motor are evaporation area (m^2) and engine power (kW). The base capacities and the exponents for drum and engine are given in Table A.10.4.

Table A.10.4 The costs of rotary drier.

	Capacity	Unit	Base Capacity		Base Cost ($)	n	Price($)
Drum	14	m^2	9.3	1	73,000	0.45	87,753
Engine	13	kW	7.5	1	12,300	0.56	16,737

A.10.3.6. Rotary Drum Air Heater (E-213)

The area of the air heater is 41 m^2. The air heater has structure similar to shell and tube exchanger and the costs of air heater is calculated based on shell and tube heat exchangers. The cost of carbon steel shell and tube heat exchanger with fixed U tube configuration with the specified area is reported as \$18,100 (Matches, IRef-4).

A.10.3.7. Sugar Carriers

The flow rate of sugar is 88.7 ton/h. Peters and Timmerhaus (1991) reports that for this capacity a belt conveyor width of 0.45 m is sufficient. The length is assumed to be 30 m. The corresponding cost is obtained as \$32,000.

A.10.3.8. Sugar Storage Silos

The flow rate of sugar is 88.7 ton/h. The residence time of sugar in storage is assumed as one week the density of bulk sugar crystals is 800 kg/m^3. Therefore the required storage volume is 18,603 m^3.

The costing of the silos are carried out based process buildings. The height of the sugars should not exceed 3 meter. Thus, the total area of the storage is 6,201 m^2. Peters and Timmerhaus (1991), reports the construction cost ware house as \$303/m^2. The total cost is obtained as \$1,879,000

A.10.4.Fermentation

A.10.4.1. Juice Coolers (E-301&E-302)

The areas of the coolers are 1,606 and 476 m^2 respectively for E-301 and E-302. The costs are estimated based on floating head shell and tube heat exchangers. The cost of is determined based on carbon steal tubes steel shell as \$274,000 and \$96,000 respectively (Matches, IRef-4). The cost of E-301 is determined based on two heat exchangers with area 803 m^2 since the area is higher that the upper margin of the cost estimation tool.

A.10.4.2. Fermentors (V-301)

The volumes of the vessels are 1,442 m^3 each. The costing is carried out based on ordinary process tanks. Brown (2003), reports the costs of vertical process vessels based on Equation A.10.4. The base capacity for a fermentor is given as 177 m^3 and the base cost is given as \$176,000 for stainless steel fermentor, the sizing exponent n=0.71 using this values the cost of one vessel is obtained as \$780,400

<u>Cooling</u>
The cooling equipment cost is calculated in using the same methodology with the crystallizers cooling. The cooling area is 247 m^2. The outer diameter of the tubes is taken as 5 cm. The required pipe lengths to supply the necessary areas are obtained as 1,572 m. Using the pipe cost of \$6.9/m obtained in Appendix A.10.3.2, the piping cost is obtained as \$10,846. The cost of cooling installation is assumed to be twice the cost of the piping. For Stainless steel material, the costs are to be multiplied by 2. Thus, the total cost of cooling installation is obtained as \$43,384.

A.10.4.3. Centrifuges (S-301&S-302)

The diameters of the centrifuges are 0.61 m. Using the same costing methodology with Appendix A.10.3.5 the cost of each centrifuges is obtained as \$30,504. Since the centrifuges operating in corrosive media the material is selected as stainless steel. The material factor for stainless steel is 2. But, the cost of centrifuges includes everything including the motors and the casings using an material factor of 2 over estimates the equipment cost. The cost of a motor with 5.6 kW (centrifuge power requirement) is obtained as \$12,443 using the same methodology in Appendix A.10.3.7. Subtracting this amount from the centrifuge cost yields \$18,061. The assumption is 75% of the equipment is stainless steel. Using these information the material factor is assumed as 1.44 based on the cost of equipment (\$30,504)

A.10.4.4. Yeast Sterilization Tank (V-302)

The volume of the sterilization tank is 190 m^3. The costing calculations are carried out based on agitated vessel cost reported by Peters and Timmerhaus (1991). And, the cost is obtained as \$128,000 for stainless steal mixing vessel including the motor and agitator.

A.10.4.5. Columns (C-301, C-302, C-303, C-304)

The costs of the columns are determined based on vertical pressure vessels (Sinnot, 1999). The prices are reported based on the height and diameter of the vessels.

C-301

Column: Column diameter is 2.2 and the column height is 30m the corresponding carbon steal column cost is obtained as $105,000. The material factor is 2 for stainless steal material. The cost for supports and skirt is assumed to be 10% of the total cost. Thus, the cost of column is $231,000.

Packing: The volume of the packing is 114 m^3. The cost of polypropylene pall ring packing is $400/$m^3$. The corresponding packing cost is obtained as $45,600.

C-302

Column: The column has 2 different diameter sections (See Table 8.33). Upper section has diameter of 2.7 m and height of 4 m. And, the lower section has diameter of 3.8 m and height of 15 m. Therefore, the cost of the column is assumed as the total cost of two different columns with specified diameters and heights. The cost of carbon steel column with 3.8 m diameter and 14 m height is obtained as $102,000. And, the cost of carbon steel column with 2.7 m diameter and 4 m height is obtained as $12,000. The total cost of the column is obtained as $114,000 (Sinnot,1999). The factor for stainless steel is 2 and the factor for the supports and skirts is 1.1. Thus, the total cost of column is obtained as $250,800.

Trays: The costing is carried out based on valve plates based on their diameters. The cost of carbon steel valve tray with 2.7 m diameter is obtained as $880 and that of 3.8 m is obtained as $3,500 per plate. The material factor for stainless steal is 1.7. The costs of the stainless steel valve plates with diameter 2.7 and 3.8 m is obtained as $1,496 and $5,950 respectively (Sinnot, 1999). The cost of plates with larger diameter than the upper margin of the cost model is obtained by extrapolation.

C-303

The column has 5 different diameter sections (See Table 8.38). The sections their diameter number of trays in each section and height of each section is given in Table A.10.5. The costing of the column is carried out in similar manner with beer column (C-302). There are 5 sections and the cost of each section is determined based on a complete column with 2 heads. Then the cost is multiplied by the percentage of the side surfaces to the total surface. Each section is priced based on the side surface are by determining the proportion of heads to side surface cost using their areas. For example, for the top and the bottom sections the costing is carried out based on one head and side surface cost and the other sections are priced based on only the side surface because both heads are to be open. The percentage of head to total surface is calculated as follows. It is assumed that heads are flat surface,

$$\%head\ Cost = \frac{D}{D+4H}$$

This is for one head, for two heads the percentage is to be multiplied by two. For the top and bottom sections only one head is subtracted from the total but for the intermediate sections two heads are reduced.

The prices given in Table A.10.5 are based on carbon steel material. For stainless steel material tray costs are to be multiplied by 1.7 and the column cost is to be multiplied by 2 (Sinnot, 1999). The total cost is multiplied by 1.1 to account for the supports and skirt cost. The total cost of stainless steel valve trays is obtained as $166,260 and the cost of stainless steel column is obtained as $258,706.

Table A.10.5 The column cost calculations for C-303 (Prices are given for carbon steel)

Section	Trays	Number of trays	Diameter (m)	Height (m)	Tray Cost ($/tray)	Column Cost ($)	Head %	Section Cost ($)
1	2-8	7	4.4	7	5,700	47,000	13.6	40,618
2	9-13	5	4.2	5	4,500	32,000	34.7	20.892
3	14-17	4	4.0	4	4,000	28,000	40.0	16,800
4	18-21	4	3.5	4	3,150	22,000	35.9	14,103
5	21-31	10	1.3	10	680	26,000	3.2	25,181
Total	2-31	30	-	-	97,800	155,000		117,594

C-304

*Column***:** Column diameter is 2.2 and the column height is 7.8 m. The corresponding carbon steal column cost is obtained as $31,000. The cost for supports and skirt is assumed to be 10% of the total cost. Thus, the cost of one column is $34,100.

Packing: The packing weight is 31,400 kg. Benson, (2003), reports the cost of molecular sieve zeolites as $10.3. The packing cost for one column is $324,374.

Cooling/Heating: The cooling/heating equipment cost is calculated in using the same methodology with the crystallizers cooling (Appendix A.10.3.2). The cooling area is 181.5 m^2. The outer diameter of the tubes is taken as 5 cm. The required pipe lengths to supply the necessary areas are obtained as 1,156 m. Using the pipe cost of $6.9/m obtained in Appendix A.10.3.4, the piping cost is obtained as $7,972. The cost of cooling installation is assumed to be twice the cost of the piping. Thus, the total cost of cooling installation is obtained as $15,945.

A.10.4.6. Condensers (E-304, E-306, E-309)

The costs of the condensers are determined based on floating head shell and tube heat exchangers. Equation A.10.4 is used to estimate the costs of the equipments with areas larger that the cost model. Perry and Green (1997) reports the exponent for Equation A.10.4 for heat exchanger cost calculation as 0.59.

Beer Column Condenser (E-304)

The heat exchanging area for the condenser is 765 m^2. The cost of shell and tube heat exchanger with SS shell and tubes with given area is obtained as $325,000 (Sinnot, 1999)

Rectification Column Condenser (E-306)

The heat exchanging area for the condenser is 2,071 m^2. Since the area is larger than the upper margin of the costing model given by Sinnot (1999), the condenser will be priced based on Equation A.10.4. The material is carbon steel. The cost of shell and tube heat exchanger with CS shell and tubes with 1,000 m^2 area is obtained as $125,000 (Sinnot, 1999). Therefore, the cost of the rectification column condenser is obtained as $192,070.

Molecular Sieve Column Condenser (E-309)

The heat exchanging area for the condenser is 288 m^2. The cost of shell and tube heat exchanger with CS shell and tubes with given area is obtained as $56,000 (Sinnot, 1999).

A.10.4.7. Reboilers (E-303 & E-305)

The costs of the reboilers are determined based on floating head shell and tube heat exchangers. Equation A.10.4 is used to estimate the costs of the equipments with areas larger that the cost model.

Perry and Green (1997) report the exponent for Equation A.10.4 for heat exchanger cost calculation as 0.59.

Beer Column Reboiler (E-303)

The heat exchanging area for the reboiler is 1,953 m^2. Since the area is larger than the upper margin of the costing model given by Sinnot (1999), the condenser will be priced based on Equation A.10.4. The material is stainless steel. The cost of shell and tube heat exchanger with SS shell and tubes with 1,000 m^2 area is obtained as \$435,000 (Sinnot, 1999). Therefore, the cost of the rectification column condenser is obtained as \$645,000.

Rectification Column Reboiler (E-306)

The heat exchanging area for the reboiler is 224 m^2. The material is carbon steel. The cost of shell and tube heat exchanger with CS shell and tubes with 224 m^2 area is obtained as \$47,000(Sinnot, 1999).

A.10.4.8. Heat Exchangers (E-307&E-308)

The cost of E-307 is determined based on floating head shell and tube heat exchanger with CS material. The heat exchanging area of E-307 is 34.2 m^2 and the corresponding cost is \$15,000.

The cost of E-308 is determined based on U tube shell and tube heat exchanger with CS material. The heat exchanging area of E-308 is 467 m^2 and the corresponding cost for CS floating head heat exchangers is \$72,000. the correction factor for U tube exchangers is 0.85. Consequently, the cost of E-308 is obtained as \$57,600.

A.10.4.9. Product Cooler (E-310)

The heat exchanging area for the condenser is 196 m^2. The cost of shell and tube heat exchanger with CS shell and tubes with given area is obtained as \$43,000 (Sinnot, 1999)

A.10.4.10. Ethanol Storage Tanks

The ethanol storage tanks are sized based on one week residence time. The ethanol volumetric flow rate is 55m^3/h which corresponds to 9,240 m^3/week. The cost of storage tanks are expressed using Equation A.10.1. The total number of storage Tanks are taken as 2 with 4,620 m^3 volume. The coefficients for floating roof tanks are given as C=\$2,900 and n=0.55. The corresponding cost for each tank is obtained as \$300,574.

A.10.5. Cogeneration

A.10.5.1. Rotary Driers (D-101)

The cost of the rotary driers is calculated using the same methodology with Appendix A.10.3.7. and the results are given in Table A.10.6.

Table A.10.6 The costs of rotary drier.

	Capacity	Unit	Base Capacity		Base Cost ($)	n	Price($)
Drum	431	m^2	9.3	1	73,000	0.45	536,587
Engine	188	kW	7.5	1	12,300	0.56	74,713

A.10.5.2. Steam Boilers (F-103)

The cost of boiler is determined based on the correlations given by Peters and Timmerhaus (1991). The maximum boiler capacity in the model is 454 ton/h. Therefore, the costing of the boiler is carried out based Equation A.10.4. Sinnot (1999) reports the sizing exponent for the boilers as 0.8. The cost of the boiler producing 454 ton steam/h (80 bars) is obtained as \$5,100,000. The price includes complete boiler, feed water deaerator, boiler feed pumps, chemical injection system, stack and shop assembly labor. Inserting the base capacity and cost in Equation A.10.4 yields the cost of boiler producing 829.2 ton/h steam as \$8,257,582.

A.10.5.3. Economizer (F-103)

The cost of the economizer is determined based on shell and tube heat exchangers. The area of the exchanger is 9,643 m^2. The model given by Sinnot (1999) predicts the cost of shell and tube heat exchangers with maximum size of 1,000 m^2. The cost of SS shell and tube heat exchanger with 1,000 m^2 area is obtained as \$310,000. Perry and Green (1997) reports the exponent for Equation A.10.4 for heat exchanger cost calculation as 0.59. using the values for 1,000 m^2 heat exchanger the cost of economizer is obtained as \$1,180,448.

A.10.5.4. Turbine Generators

The cost of turbine generator is obtained using the costs and capacities given in NREL report (Adan *et al.*, 2002). The electricity generation capacity of their plant is 30.3 MW and the sizing exponent for the turbines is given as 0.71. The cost of the turbine used in their study is reported as \$7,422,186. using the base capacity and the cost the cost for turbine generating 77.5 MW electricity is obtained as \$14,458,118 (2002 prices).

A.10.5.5. Chimney

The height of the chimney is 40 m and the diameter is 14.3m. To estimate the cost the wall thickness is taken as 0.4 meter on average which yields an external diameter of 15.1 m. The chimney is assumed to be a hollow cylinder and the volume of the wall of the chimney is obtained as the difference of the volume of the cylinder with 15.1 m diameter with that of 14.3 m diameter. The side wall volume of the chimney is obtained as 739 m^3 using,

$$Volume\,of\;Chimney\,Wall = \frac{\pi D_{ext}^2 H}{4} - \frac{\pi D_{int}^2 H}{4}$$

The costing is determined based on reinforced concrete wall costs. The costs of the reinforced concrete walls with 0.3 meter thickness is given as \$87.6/ m^2 which gives the cost as \$292.1/m^3 including the labor for the construction (Peters and Timmerhaus, 1991). Using this value the cost of chimney is obtained as \$215,833.

A.10.5.6. Cooling Tower

The cost of cooling tower is calculated based on the capacity of 6,000 m^3 water/h capacity (See utilities description). Peters and Timmerhaus (1991) reports the cost of cooling tower with this capacity as \$430,000.

A.10.6. Pumps

The capacities power requirements and the costs of pumps used in the plant are given in Table A.10.7. Each pump has a spare.

Table A.10.7 The power requirements and costs of the pumps

Pump	Number	Capacity (m^3/s)	Head (m)	Power (kW)	Cost ($/unit)
P-101	10	0.018	3	0.27	2,940
P-102	2	0.202	10	29.00	59,980
P-103	40	0.062	3	0.95	13,327
P-104	2	0.016	3	0.50	3,715
P-105	2	0.001	10	0.17	2,596
P-106	2	0.202	6	17.40	18,091
P-107	2	0.109	6	9.40	14,110
P-108	2	0.009	4	0.50	21,347
P-109	2	0.019	4	1.10	21,458
P-110	2	0.102	18	26.30	3,868
P-111	2	0.028	2	0.80	4,692
P-112	2	0.190	9.8	26.60	5,405
P-201	4	0.017	15	5.00	10,355
P-202	2	0.029	20	8.20	26,817
P-203	2	0.022	9.8	3.30	11,347
P-204	2	0.029	3	1.23	16,705
P-205	2	0.013	20	3.70	18,123
P-206	2	0.013	3	0.55	8,016
P-207	2	0.100	18	2.70	15,869
P-301	2	0.014	30	5.90	30,139
P-302	2	0.205	2	5.90	30,039
P-303	2	0.192	2	5.50	29,204
P-304	2	0.125	17	30.50	61,249
P-305	2	0.015	20	4.30	26,263
P-306	2	0.051	2	1.50	16,507
P-307	2	0.097	20	27.90	59,096
P-308	2	0.020	20	5.70	29,721
P-309	2	0.081	32	34.90	64,624
P-310	2	0.015	20	4.30	26,263
P-311	2	0.062	13	11.70	41,732
P-312	2	0.059	19	16.20	47,552
K-301	2	6.510	4	325.00	7,108
K-302	2	3.157	4	158.00	5,517
K-303	2	35.516	1	288.40	17,599
K-304	2	11.000	1	150.50	10,861
K-305	2	11.000	3	425.50	10,861

A.10.7.Adjustment of Prices to 2005

The historical United States inflation data is obtained from web page Inflationdata.com (IRef-6) and given in Table A.10.8. The customer price index profile for Brazil for the last 10 years is taken from Brazil Central Bank database (IRef-7) and is given in Table A.10.9.

Using the data the average of the last 10 years' inflation is obtained as 2.47% and 9.45% for US and Brazil respectively and these values will be used as yearly inflation. The adjustment of the prices from base year to 2005 is carried out using the following equation,

$$Cost_{2005} = Cost_y \times (1 + r_i)^{2005-y} \quad \text{Eqn A.10.5}$$

y is the base year and r_i is the inflation rate.

Table A.10.8 Historical US inflation data (IRef-6)

Period	Average
2004	2.68%
2003	2.27%
2002	1.59%
2001	2.83%
2000	3.38%
1999	2.19%
1998	1.55%
1997	2.34%
1996	2.93%
1995	2.81%
1994	2.61%

Table A.10.9 Historical Brazil inflation data (IRef-7)

Period	Average
2005	5.5%
2004	9.28%
2003	8.73%
2002	18.92%
2001	9.33%
2000	7.33%
1999	14.17%
1998	0.61%
1997	5.04%
1996	8.37%
1995	16.72%

A.10.8. Equipment Installation Factors

Installation of equipment includes the materials and the labor required for the installation. Labor, material, support structures, electrical wiring and control panels, piping and valves for water, gas and steam utilities are included in the installation cost of the equipment.

The installed cost of each of the equipments is determined by multiplying the purchased cost of equipment by the installation factor.

$$IEC = EPC \times IF \qquad \text{Eqn A.10.6}$$

IEC is installed equipment cost, *EPC* is equipment purchase cost and *IF* is installation factor. The installation factors are taken from NREL report (Aden *et al,* 2002). The installation factors for different equipments are given in Table A.10.10.

In the mentioned study (Aden, 2002), the installed equipment cost includes piping, instrumentation and electrical equipment materials.

Table A.10.10 Installation factors for the equipments

Equipment	Multiplier	Equipment	Multiplier
Agitators – Carbon Steel (CS)	1.3-1.4	Pumps – Lobe	1.2-1.4
Agitators – Stainless Steel (SS)	1.2-1.3	Pumps – Centrifugal, Carbon Steel	2.8
Boilers	1.3	Pumps – Centrifugal, Stainless Steel	2
Compressors (motor driven)	1.3	Pressure Vessels – Carbon Steel	2.8
Cooling Towers	1.2	Pressure Vessels – Stainless Steel	1.7
Distillation columns – Carbon Steel	3	Tanks – Field Erected, Carbon Steel	1.4
Distillation columns – Stainless Steel	2.1	Tanks – Field Erected, Stainless Steel	1.2
Filters	1.4	Solids Handling Equipment	1.2-1.4
Heat Exchangers (S&T)	2.1	Turbo Generator	1.5

A.10.9. Equipment Purchase Costs and Installed Equipment Costs (IEC)

Table A.10.11 Equipment purchase and installed equipment costs for mill unit (2005)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
Cane carriers	M-101		$154,976	1990	$223,467	1.30	$290,508
Cane knives	M-101	SS	$900,000	1990	$1,297,758	1.30	$1,687,085
Magnetic separators	M-101		$52,000	2002	$55,949	1.30	$72,734
Crushers	M-101	SS	$1,400,000	1990	$2,018,734	1.30	$2,624,354
Mills	M-101	SS	$1,875,000	1990	$2,703,661	1.30	$3,514,760
Intermediate carriers	M-101		$3,465	1998	$4,110	1.30	$5,343
Bagasse carriers	M-101		$98,151	1998	$116,432	1.30	$151,362
Feed pumps	P-103	CS	$29,404	2005	$29,404	2.80	$82,331
Circulatory pumps	P-101	SS	$533,080	2005	$533,080	2.00	$1,066,160
Milling section sum			$5,046,076		$6,982,596		$9,494,636

Table A.10.12 Equipment purchase and installed equipment costs for clarification unit (2005)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
Sulpur burner	F-102	SS	$147,345	1998	$174,788	1.40	$244,703
Lime kiln	F-101	SS	$74,350	1998	$88,198	1.40	$123,477
Sulphur dioxide blower	F-102		$18,500	1990	$26,676	1.30	$34,679
Sulphur burner air blower	F-102		$460	1990	$663	1.30	$862
Lime kiln blower	F-101		$410	1990	$591	1.30	$769
Sulphitation tower	C-101	SS	$42,000	1990	$60,562	2.10	$127,180
Sulphitation tower packing	C-101	Cer.	$4,378	1998	$5,193	2.10	$10,906
Lime milk preparation tanks and agitation	V-101	SS	$42,000	1990	$60,562	1.40	$84,787
V-101 cooling	V-101	SS	$2,050	1990	$2,956	2.10	$6,208
Sugar branch liming tank	V-102	SS	$210,000	1990	$302,810	1.40	$423,934
Ethanol branch liming tank	V-103	SS	$150,000	1990	$216,293	1.40	$302,810
Sugar branch juice heater	E-101	SS/CS	$171,000	2003	$179,552	2.10	$377,059
Ethanol branch juice heater	E-102	SS/CS	$100,900	2003	$105,946	2.10	$222,487
Ethanol branch flash tank	V-104	CS	$10,500	1998	$12,456	1.40	$17,438
Sugar branch flash tank	V-105	CS	$9,000	1998	$10,676	1.40	$14,947
Clarifiers	S-101/2	CS	$1,269,040	2002	$1,365,418	1.50	$2,048,127
Rotary vacuum filter	S-103	CS	$1,049,836	1996	$1,307,652	1.40	$1,830,712
Pumps	P-102	SS	$119,960	2005	$119,960	2.00	$239,920
Pumps	P-104	CS	$7,430	2005	$7,430	2.80	$20,803
Pumps	P-105	CS	$5,191	2005	$5,191	2.80	$14,535
Pumps	P-106	CS	$36,182	2005	$36,182	2.80	$101,310
Pumps	P-107	CS	$28,220	2005	$28,220	2.80	$79,016
Pumps	P-108	CS	$42,694	2005	$42,694	2.80	$119,542
Pumps	P-109	CS	$42,916	2005	$42,916	2.80	$120,164
Pumps	P-110	CS	$7,737	2005	$7,737	2.80	$21,663
Pumps	P-111	CS	$9,383	2005	$9,383	2.80	$26,273
Pumps	P-112	CS	$10,809	2005	$10,809	2.80	$30,266
Clarification unit sum			$3,612,290		$4,231,513		$6,644,575

Table A.10.13 Equipment purchase and installed equipment costs for evaporation unit (2005)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
1st Evaporator	E-201	CS	$1,249,997	1998	$1,482,812	2.1	$3,113,905
2nd Evaporator	E-202	CS	$1,048,453	1998	$1,243,730	2.1	$2,611,832
3rd Evaporator	E-203	CS	$1,133,196	1998	$1,344,257	2.1	$2,822,939
4th Evaporator	E-204	CS	$1,234,932	1998	$1,464,941	2.1	$3,076,376
5th Evaporator	E-205	CS	$1,368,441	1998	$1,623,317	2.1	$3,408,966
6th Evaporator	E-206	CS	$1,599,229	1998	$1,897,089	2.1	$3,983,887
Pumps	P-201	CS	$41,419.28	2005	$41,419	2.8	$115,974
Evaporation unit sum			$10,676,972		$7,675,667		$19,133,880

Table A.10.14 Equipment purchase and installed equipment costs for crystallization and drying unit (2005)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
1st boiling pan	E-207	CS	$567,559	1998	$673,268	2.1	$1,413,863
2nd boiling pan	E-208	CS	$420,981	1998	$499,389	2.1	$1,048,718
1st crystallizer	V-201	CS	$438,000	1990	$631,575	1.4	$884,205
2nd crystallizer	V-202	CS	$420,000	1990	$605,620	1.4	$847,868
1st crystallizer cooling	V-201	CS	$130,200	1990	$187,742	2.1	$394,259
2nd crystallizer cooling	V-202	CS	$115,970	1990	$167,223	2.1	$351,168
1st centrifuge	S-201	CS	$441,020	1998	$523,161	1.4	$732,426
2nd centrifuge	S-202	CS	$308,714	1998	$366,213	1.4	$512,698
Reheater	E-209	CS	$88,000	2003	$104,390	2.1	$219,219
Reheater	E-210	CS	$43,500	2003	$51,602	2.1	$108,364
Reheater	E-211	CS	$51,000	2003	$60,499	2.1	$127,048
Reheater	E-212	CS	$29,000	2003	$34,401	2.1	$72,243
Air heater	E-213	CS	$18,100	2003	$21,471	2.1	$45,089
Rotary drier	D-201	CS	$87,753	1996	$109,303	1.4	$153,025
Rotary drier engine	D-201	CS	$16,737	1996	$20,847	1.4	$29,186
Sugar carrier			$64,000	1990	$92,285	1.4	$129,199
Sugar silo			$1,879,000	1990	$2,709,429	1	$2,709,429
Pumps	P-202	CS	$26,817	2005	$26,817	2.8	$75,087
Pumps	P-203	CS	$11,347	2005	$11,347	2.8	$31,772
Pumps	P-204	CS	$16,705	2005	$16,705	2.8	$46,773
Pumps	P-205	CS	$18,123	2005	$18,123	2.8	$50,743
Pumps	P-206	CS	$8,016	2005	$8,016	2.8	$22,444
Pumps	P-207	CS	$15,869	2005	$15,869	2.8	$44,434
Crystallization and drying unit sum			$5,216,410		$6,955,298		$10,049,262

TableA.10.15 Equipment purchase and installed equipment costs for fermentation unit (2005)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
Juice cooler	E-301	CS	$274,000	2003	$287,703	2.10	$604,176
Juice cooler	E-302	CS	$96,000	2003	$100,801	2.10	$211,682
Fermentors	V-301	SS	$4,682,400	1998	$5,554,509	1.2	$6,665,411
Fermentor cooling		SS	$260,304	1990	$375,346	1.20	$450,415
1st centrifuge	S-301	SS	$483,183	1998	$573,177	1.40	$802,448
2nd centrifuge	S-302	SS	$527,109	1998	$625,285	1.40	$875,398
Yeast sterilization tank	V-302	SS	$128,000	1990	$184,570	1.40	$258,398
Scrubber column	C-301	SS	$231,000	1998	$274,024	2.10	$575,451
Scrubber column packing	C-301	Plastic	$45,600	1998	$54,093	2.10	$113,596
Fermentation unit sum			$6,727,596		$8,029,508		$10,556,975

Table A.10.16 Equipment purchase and installed equipment costs for ethanol purification unit (2005)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
Beer column	C-302	SS	$250,800	1998	$297,512	2.1	$624,775
Beer column trays	C-302	SS	$95,234	1998	$112.972	2.1	$237,240
Rectification column	C-303	SS	$258,706	1998	$306,891	2.1	$644,470
Rectification column trays	C-303	SS	$166,260	1998	$197,226	2.1	$414,175
Molecular sieve column	C-304	CS	$68,200	1998	$80,902	2.1	$169,895
Molecular sieve packing	C-304	Zeolite	$648,748	2002	$698,017	2.1	$1,465,837
Molecular sieve cooling	C-304	CS	$31,891	1990	$45,985	2.1	$96,569
Beer column condenser	E-304	SS	$325,000	1998	$385,532	2.1	$809,617
Rectification col. condenser	E-306	CS	$192,070	1998	$227,844	2.1	$478,471
Nitrogen dehumidifier	E-309	CS	$56,000	1998	$66,430	2.1	$139,503
Beer column reboiler	E-303	SS	$645,000	1998	$765,133	2.1	$1,606,779
Rectification col. reboiler	E-305	CS	$47,000	1998	$55,754	2.1	$117,083
Molecular sieve feed cooler	E-307	CS	$15,000	1998	$17,794	2.1	$37,367
Nitrogen heater	E-308	CS	$57,600	1998	$68,328	2.1	$143,489
Ethanol storage tanks		CS	$601,148	1998	$713,113	1.4	$998,358
Pumps	P-301	CS	$22,303	2005	$22,303	2.8	$62,448

Table A.10.16 (Continuing)Equipment purchase and installed equipment costs for ethanol purification unit (2005)

Pumps	P-302	SS	$60,078	2005	$60,078	2	$120,156
Pumps	P-303	SS	$58,408	2005	$58,408	2	$116,816
Pumps	P-304	SS	$122,498	2005	$122,498	2	$244,996
Pumps	P-305	SS	$52,526	2005	$52,526	2	$105,052
Pumps	P-306	SS	$33,014	2005	$33,014	2	$66,028
Pumps	P-307	SS	$118,192	2005	$118,192	2	$236,384
Pumps	P-308	SS	$59,442	2005	$59,442	2	$118,884
Pumps	P-309	CS	$47,822	2005	$47,822	2.8	$133,901
Pumps	P-310	CS	$19,435	2005	$19,435	2.8	$54,417
Pumps	P-311	SS	$83,464	2005	$83,464	2	$166,928
Pumps	P-312	SS	$95,104	2005	$95,104	2	$190,208
Pumps	K-301	SS	$34,118	2005	$34,118	1.3	$44,354
Pumps	K-302	SS	$26,482	2005	$26,482	1.3	$34,426
Pumps	K-303	SS	$84,475	2005	$84,475	1.3	$109,818
Pumps	K-304	CS	$21,722	2005	$21,722	1.3	$28,239
Pumps	K-305	CS	$21,722	2005	$21,722	1.3	$28,239
Ethanol purification unit sum			$4,462,461		$5,051,246		$9,895,932

Table A.10.17 Equipment purchase and installed equipment costs for cogeneration and aux. process equipments (2005)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
Bagasse drier	D-101	CS	$2,682,935	1996	$3,341,802	1.4	$4,678,523
Bagasse drier engines	D-101		$373,565	1996	$465,304	1.4	$651,426
Steam boiler	F-103	CS	$8,257,582	1990	$11,907,043	1.3	$15,479,156
Air heater (Economizer)	F-103	SS	$1,498,900	1998	$1,778,074	2.1	$3,733,955
Turbine generator	F-103		$14,458,118	2002	$15,556,145	1.5	$23,334,217
Chimney			$215,833	1990	$311,221	1	$311,221
Cooling tower			$430,000	1990	$620,040	1.2	$744,048
Cogeneration and auxiliary equipment sum			$27,916,933		$33,979,629		$48,932,546

A.10.10. Fixed Capital Investment

The factors effecting the fixed capital investment are divided in three subheadings: direct, indirect and working capital costs which are described as follows (Peters and Timmerhaus, 1991),

<u>Direct Costs:</u>

1. <u>Purchase cost of equipments</u>: Cost of all equipment listed in complete flow sheet.
2. <u>Equipment installation</u>: Installation of all equipment listed on complete flow sheet, structural supports and installation.
    - This element is already is included in installed equipment costs. Thus, this factor is taken as 0.
3. <u>Instrumentation and control</u>: Purchase, installation calibration and computer tie.
    - Sinnot (1999) reports the cost of this element for solid fluid processing plants as 15% of the equipment purchase cost (EPC).
    - Peters and Timmerhaus (1991) report the cost of this element as 6-30% of the EPC.
    - Since the design is not highly automated, for the current calculations a value of 15% of the purchased equipment cost is taken as the bases.
4. <u>Piping</u>: Process piping, hangers, fittings and valves and insulation.
    - Sinnot (1999) reports the contribution of this element as 45% of the EPC.
    - Peters and Timmerhaus (1991) reports the cost of this element as 31% of the EPC.
    - This element is already included in installed equipment cost. However, to account for the service facilities a factor of 15% of EPC is taken as the contribution to investment.
5. <u>Electrical equipment and materials</u>: Electrical equipment, switches motors, conduit, wire, fittings, grounding, instrument and control wiring and panels.

- Sinnot (1999) reports the contribution of this element as 10% of the EPC.
- Peters and Timmerhaus (1991) reports the cost of this element as 10-15% of the EPC.
-This element is already included in installed equipment cost. However, to account for the service facilities a factor of 7% of EPC is taken as the contribution to investment.

6. Buildings: Process buildings, administration and office, medical, cafeteria, garage, product ware house, parts warehouse, guards safety, fire station, personal building, research and control laboratory, maintenance shops and building services.
    - Sinnot (1999) reports the contribution of this element as 30% of the EPC.
    - Peters and Timmerhaus (1991) reports the contribution of this element as 22% of the EPC for a new plant at an existing site.
    - For the calculations the contribution of this element to cost is taken as 25% of the EPC.
7. Yard improvements: site development, grading, roads, walk ways, rail roads, fences, parking areas, recreational facilities.
    - Sinnot (1999) reports the contribution of this element as 5% of the EPC.
    - Peters and Timmerhaus (1991) reports the contribution of this element as 2-5 % of total capital investment.
    - For the calculations the contribution of this element to cost is taken as 5% of the EPC.
8. Service facilities:
    a. Utilities: steam, power, water, refrigeration, compressed air, fuel, waste disposal.
    b. Facilities: boilers, wells, river intake, water treatment, cooling towers, water storage, electric substation, refrigeration plant, air plant, fuel storage, waste disposal plant, environmental controls, fire protection
    c. Nonprocess equipment: office furniture and equipment, cafeteria equipment, safety and medical equipment, shop equipment, yard and material handling equipment, laboratory equipment etc.
    d. Distribution and packaging: raw material and product storages and handling equipment, product packaging equipment, blending facilities loading station.

    - Sinnot (1999) reports the contribution of this element as 65% of the EPC.
    - Peters and Timmerhaus (1991) report the contribution of this element as 55% of EPC for normal solid fluid processing plant.
    - The investments on steam, electricity, cooling tower and the major storage facilities are already included in equipment costs, and there aren't any major water treatment and waste treatment facility. Since these are the major contributors with comparably high prices, the contribution of this element in the investment is taken as 20% of the EPC.
9. Land: property cost.
    - Taken as 5% of EPC.

Indirect Costs:

1. Design and Engineering:
    - Sinnot (1999) reports the contribution of this element as 25% of the direct expenses.
    - Peters and Timmerhaus (1991) reports the contribution of this element as 8% of direct expenses.

2. Construction Expenses:
   - Sinnot (1999) reports the contribution of this element as 5% of the direct expenses.
   - Peters and Timmerhaus (1991) reports the contribution of this element as 2-8 % of direct plant cost.
3. Contingency: To cover for the unforeseen circumstances
   - Sinnot (1999) reports the contribution of this element as 10% of the direct expenses.
   - Peters and Timmerhaus (1991) reports the contribution of this element as 5-15 % of direct costs.

Since the sugar ethanol plants are state of the art well known processes and there are engineering firms in Brazil that construct package plants on demand, the engineering and construction fees are taken relatively lover then the stated margins and each contribute to the fixed capital investment 8% of the direct expenses. Due to the well known process structure, detailed factoring of the equipment design and economical analysis the contingency taken as 5% in the cost calculations.

Working Capital:

The working capital is defined as the amount of money invested on

1. Raw materials and supplies carried in stock: one moths of supply
2. Finished product in stock : one months of finished product storage
3. Accounts receivable: production cost of one months of operation
4. Accounts payable.
5. Taxes payable.

The working capital and start up expenses are assumed as 15% of total direct and indirect expenses (Sinnot, 1999).

A.10.11. Operational Costs

A.10.11.1. Fixed Operational Costs

Labour Cost

The labour requirement of the plant is determined based on the information given by Brown (2003) in Table A.10.18. The labor requirements for the individual equipments are given in Table A.10.19. The labour requirement is obtained as 18.3 operators /shift.

Table A.10.18 Operator requirements of the equipments

Equipment	Operator Per unit Per Shift
Boilers	1
Cooling tower	1
Electric generating plants	3
Waste water treatment plants	2
Conveyors	0.2
Crushers, mills grinders	0.5
Evaporators	0.3
Furnaces	0.5
Blowers and compressors	0.1-0.2
Heat exchangers	0.1
Mixers	0.3
Reactors	0.5
Clarifiers-thickeners	0.2
Centrifugal separators and filters	0.05-0.2
Rotary and belt filters	0.1
Rotary drier*	0.5
Mechanical crystallizer*	1/6
Water treatment plants	2

*Peters and Timmerhaus (1991)

Table A.10.19 Operator requirements of the equipments (2005)

Equipment	**Labour**	**Equipment**	**Labour**
Mill & Clarification		Sugar Carrier	0.20
Conveyors	0.40	**Fermentation**	
Mills	2.50	Juice Coolers	0.20
Furnaces	1.00	Fermentors	1.00
Sulphitation tower	0.50	Centrifuges	0.10
Lime milk prep. tanks	0.30	Yeast sterilization tank	0.30
Liming tanks	0.60	**Ethanol Recovery**	
Juice heaters	0.20	Columns	0.30
Clarifiers	0.20	Condensers	0.30
Rotary filter	0.10	Reboilers	0.20
Evaporation, Crystallization & Drying		Heat Exchangers	0.30
Evaporators	1.50	**Cogeneration**	
Pans	0.50	Rotary driers	0.50
Centrifuges	0.10	Boiler	1.00
Crystallizers	0.34	Turbine generator	3.00
Reheaters	0.40	Cooling tower	1.00
Air heater	0.10	Pumps	1.00
Rotary drier	0.10	**Total (1 shift)**	**18.30**

The average wage of ethanol industry workers is given as \$446/month (February 2004) including the benefits (Report, The facts about the Brazilian ethanol industry, IRef-17). Since the salary is given in US\$, US inflation rate of 2.47% is used and the 2005 salaries are obtained as \$466/month (\$2,796/ year for 6 months of operation).

A.10.12. Cash Flow Analysis

A.10.12.1. Gross Cash Flow (GCF)

The net cash flow in year *i* is calculated as follows,

$$GCF_i = revenue_i - (fixed + variable + misc.)_i \qquad \text{Eqn .A.10.7}$$

A.10.12.2. Cumulative Gross Cash Flow (CGCF)

The cumulative gross cash flow is calculated using,

$$CGCF = \sum_{0}^{13} GCF_i \qquad \text{Eqn A.10.8}$$

Year 0 is the time of decision to invest the plant and year 4 is the time when the plant starts operation

A.10.12.3. Net Cash Flow (NCF)

The net cash flow in year *i* is calculated using.

$$NCF_i = (GCF_i - Depreciation) \times (1 - r_{tax}) \qquad \text{Eqn A.10.9}$$

Where, r_{tax} is the tax rate

A.10.12.4. Cumulative Net Cash Flow (CNCF)

The cumulative net cash flow is calculated using.

$$CNCF = \sum_{0}^{13} NCF_i \qquad \text{Eqn A.10.10}$$

A.10.12.5. Discounted Cash Flow (DCF)

The discounted cash flow is calculated using.

$$DCF_i = \frac{NCF_i}{(1+r_{dis})^i} \qquad \text{Eqn A.10.11}$$

Where, r_{dis} is the discount rate

A.10.12.6. Net Present Value (NPV)

$$NPV = \sum_{0}^{13} \frac{NCF_i}{(1+r_{dis})^i} \qquad \text{Eqn A.10.12}$$

A.10.12.7. Discounted Cash Flow Rate of Return (DCFROR)

DCFROR is the interest rate which gives zero NPV. In other words, DCFROR is the maximum interest rate which gives profitable investment.

$$DCFROR = i \; given \; that \sum_{0}^{13} \frac{NCF_i}{(1+r_{dis})^i} = 0 \qquad \text{Eqn A.10.13}$$

A.11. Economical Evaluation: 2015 Plant

The sources used during the equipment purchase cost estimations are Sinnott R.K. (1999), Peters and Timmerhaus (1991), Super Pro Designer 5.1 cost analysis tool and Matche online database (IRef-4). Sinnot (1999) describes the purchase costs for the equipments can be calculated using the equations given in Appendix A10,

The prices referring to Peters and Timmerhaus (1991) are from late year 1990, the prices taken from Matches (IRef-4) are from year 2003, the prices taken from super pro designer are adjusted to 2005, and that taken from Sinnot, (1999) are from 1998. Therefore, during the investment cost calculations the prices will be converted to current values using the US inflation rate.

The prices of the pumps are determined based on centrifugal pumps by using Super Pro Designer database. The pumps have one spare, therefore the costs are doubled during the calculations.

The equipment that are same in 2005 and 2015 plant are not shown in this chapter and their pricing is as given in Appendix A.10.

A.11.1. Equipments Manipulated in 2015 Plant

A.11.1.1. Lime Kiln (F-101)

The cost of the lime kiln is determined based on the process furnaces. The heat duty of the kiln is 1,957.1 kW. The coefficients for process furnaces to be used in Eqn. A.10.1 are as given in Table A.10.2 and the cost is calculated as $123,266.4 (Sinnot, 1999).

Blower
The capacity of the blower is 75.96 m^3/min. The cost of centrifugal fan of this capacity is reported as $530 (tube-axial fan) (Peters and Timmerhaus, 1991).

A.11.1.2. Lime Milk Preparation Vessels (V-101)

Vessels
The capacity of the vessels is 20.5 m^3. The cost of mixing tanks for this capacity is given as $40,800 including the drive for the agitator. (Peters and Timmerhaus, 1991).

Cooling
To account for the cooling the cost of cooling coils is assumed to be the half the cost of double pipe heat exchanger with the same area. The cost of the stainless steal double pipe heat exchanger with 18.7 m^2 heat exchanging area is $2,300 ,therefore, the cooling system is assumed to cost $1,150 for each vessel (Peters and Timmerhaus, 1991).

A.11.1.3. Juice Heaters (E-101&E-102)

The juice heaters have areas of 527 and 328 m^2 respectively for E-101 and E-102. The costs are estimated based on shell and tube heat exchangers. The cost of is determined based on stainless steal tubes (SS-304) shell as $290,000 and $202,000 respectively (Sinnot, 1999).

A.11.1.4. Molasses Reheaters (E-209, E-210, E-211&E-212)

The Molasses Reheaters have areas of 421, 178, 207 and 78 m^2 respectively for E-209, E-210, E-211 and E-212. The costs are estimated based on shell and tube heat exchangers. The cost of is determined based on carbon steal tubes steel shell as \$61,000, \$40,000, \$43,500 and \$32,800 respectively (Matches, IRef-4).

A.11.2.Fermentation

A.11.2.1. Juice Cooler (E-301)

The area of the cooler is 2,144 m^2. The cost is estimated based on floating head shell and tube heat exchangers. Since the area of the exchanger is outside the cost model the cost is determined by using the half of the equipment area and multiplying the obtained cost by two. The cost of exchanger of this type with area 1,072 m^2 is determined based on carbon steal shell and tubes as \$135,000 (Sinnot, 1999). Therefore, the total cost of equipment is \$270,000.

A.11.2.2. Fermentors (BR-301)

The volumes of the vessels are 1,481 m^3 each. The costing is carried out based on ordinary process tanks. Brown (2003), reports the costs of vertical process vessels in terms of Equation A.10.4. The base capacity for a fermentor is given as 177 m^3 and the base cost is given as \$176,000 for stainless steel fermentor, the sizing exponent n=0.71 using this values the cost of one vessel is obtained as \$795,527

Cooling

The cooling equipment cost is calculated in using the same methodology with Appendix A.10.3.4. The cooling area is 150.5 m^2. The outer diameter of the tubes is taken as 5 cm. The required pipe lengths to supply the necessary areas are obtained as 958 m. Using the pipe cost of \$6.9/m obtained in Appendix A.10.3.2, the piping cost is obtained as \$6,611. The cost of cooling installation is assumed to be twice the cost of the piping. For Stainless steel material, the costs are to be multiplied by 2. Thus, the total cost of cooling installation is obtained as \$26,444.

A.11.2.3. Centrifuges (S-301&S-302)

The diameters of the centrifuges are 0.61 m. Using the same costing methodology with Appendix A.10.4.3 the cost of each centrifuges is obtained as \$43,926.

A.11.2.4. Yeast Sterilization Tank (V-301)

The volume of the sterilization tank is 284 m^3. The costing calculations are carried out based on agitated vessel cost reported by Peters and Timmerhaus (1991). And the cost is obtained as \$185,000 for stainless steal mixing vessel including the motor and agitator.

A.11.2.5. Columns (C-301, C-302, C-303, C-304)

The costs of the columns are determined based on vertical pressure vessels (Sinnot, 1999). The prices are reported based on the height and diameter of the vessels.

C-301

Column: Column diameter is 3.3 and the column height is 30m the corresponding carbon steal column cost is obtained as \$160,000. The material factor is 2 for stainless steal material. The cost for supports and skirt is assumed to be 10% of the total cost. Thus, the cost of column is \$352,000.

Packing: The volume of the packing is 257 m^3. The cost of polypropylene pall ring packing is \$400/m^3. The corresponding packing cost is obtained as \$102,636.

C-302

Column: The column has 2 different diameter sections (See Table 9.26). Upper section has diameter of 3.7 m and height of 4 m. And the lower section has diameter of 5.7 m and height of 15 m. Therefore, the cost of the column is assumed as the total cost of two different columns with specified diameters and heights. The cost of carbon steel column with 5.7 m diameter and 14 m height is obtained as $150,000. And the cost of carbon steel column with 3.7 m diameter and 4 m height is obtained as $24,000. The total cost of the column is obtained as $174,000 (Sinnot,1999). The factor for stainless steel is 2 and the factor for the supports and skirts is 1.1. Thus, the total cost of column is obtained as $382,800.

Trays: The costing is carried out based on valve plates based on their diameters. The cost of carbon steel valve tray with 3.7 m diameter is obtained as $3,100 and that of 5.7 m is obtained as $9,100 per plate. The material factors for stainless steal is 1.7. The costs of the stainless steel valve plates with diameter 3.7 and 5.7 m is obtained as $5,270 and $15,470 respectively (Sinnot, 1999). The cost of plates with larger diameter than the upper margin of the cost model is obtained by extrapolation.

C-303

The column has 5 different diameter sections (See Table 9.30). The sections their diameter number of trays in each section and height of each section is given in Table A.11.1. The costing of the column is carried out in similar manner with Appendix A.10.4.5. and the. The total cost of stainless steel valve trays is obtained as $427,550 and the cost of stainless steel column is obtained as $444,307.

Table A.11.1 The column cost calculations for C-303 (Prices are given for carbon steel)

Section	Trays	Number of trays	Diameter (m)	Height (m)	Tray Cost ($/tray)	Column Cost ($)	Head %	Section Cost ($)
1	2-6	5	6.8	5	12,000	69,000	25.4	51,474
2	7-12	6	6.6	6	10,500	78,000	43.2	44,304
3	13-17	5	6.2	5	10,100	59,000	47.3	31,093
4	18-24	7	5.8	7	9,100	71,000	34.3	46,647
5	25-35	11	2.4	11	1,300	30,000	5.2	28,440
Total	2-35	34	-	-	251,500			201,958

C-304

Column: Column diameter is 3.2 and the column height is 8.4 m. The corresponding carbon steal column cost is obtained as $43,000. The cost for supports and skirt is assumed to be 10% of the total cost. Thus, the cost of one column is $47,300 (Sinnot,1999).

Packing: The packing weight is 49,300 kg. Benson, (2003), reports the cost of molecular sieve zeolites as $10.3. The packing cost for one column is $507,790.

Cooling/Heating: The cooling/heating equipment cost is calculated in using the same methodology with Appendix A.10.3.4. The cooling area is 436 m^2. The outer diameter of the tubes is taken as 5 cm. The required pipe lengths to supply the necessary areas are obtained as 2,775 m. Using the pipe cost of $6.9/m obtained in Appendix A.10.3.2, the piping cost is obtained as $19,152. The cost of cooling installation is assumed to be twice the cost of the piping. Thus, the total cost of cooling installation is obtained as $38,304.

A.11.2.6. Beer Column feed Heater (E-302)

The area of the heat exchanger is 2,696 m^2. The cost is estimated based on floating head shell and tube heat exchangers. Since the area of the exchanger is outside the cost model the cost is determined by using one third of the equipment area and multiplying the obtained cost by three. The cost of exchanger of this type with area 899 m^2 is determined based on carbon steal shell and stainless steel tubes as $340,000 (Sinnot, 1999). Therefore, the total cost of equipment is $1,020,000.

A.11.2.7. Condensers (E-304, E-306, E-309)

The costs of the condensers are determined based on floating head shell and tube heat exchangers. Equation A.10.4 is used to estimate the costs of the equipments with areas larger that the cost model. Perry and Green (1997) reports the exponent for Equation A.10.4 for heat exchanger cost calculation as 0.59.

Beer Column Condenser (E-304)
The heat exchanging area for the condenser is 1,502 m^2. Since the area of the condenser is out side the upper margin of the cost model, the condenser is priced based on Equation A.10.4. The cost of shell and tube heat exchanger with SS shell and tubes with 1,000 m^2 area is obtained as $435,000 (Sinnot, 1999). Therefore, the cost of the condenser is $553,000.

Rectification Column Condenser (E-306)
The heat exchanging area for the condenser is 4,951 m^2. Since the area is larger than the upper margin of the costing model given by Sinnot (1999), the condenser is priced based on Equation A.10.4. The material is carbon steel. The cost of shell and tube heat exchanger with CS shell and tubes with 1,000 m^2 area is obtained as $125,000 (Sinnot, 1999). Therefore, the cost of the rectification column condenser is obtained as $321,302.

Molecular Sieve Column Condenser (E-309)
The heat exchanging area for the condenser is 565 m^2. The cost of shell and tube heat exchanger with CS shell and tubes with given area is obtained as $87,000 (Sinnot, 1999).

A.11.2.8. Reboilers (E-303 & E-305)

The costs of the reboilers are determined based on floating head shell and tube heat exchangers. Equation A.10.4 is used to estimate the costs of the equipments with areas larger that the cost model. Perry and Green (1997) reports the exponent for Equation A.10.4 for heat exchanger cost calculation as 0.59.

Beer Column Reboiler (E-303)
The heat exchanging area for the reboiler is 4,168 m^2. Since the area is larger than the upper margin of the costing model given by Sinnot (1999), the reboiler will be priced based on is priced based on Equation A.10.4. The material is stainless steel. The cost of shell and tube heat exchanger with SS shell and tubes with 1,000 m^2 area is obtained as $435,000 (Sinnot, 1999). Therefore, the cost of the rectification column condenser is obtained as $1,009,830.

Rectification Column Reboiler (E-306)
The heat exchanging area for the reboiler is 653 m^2. The material is carbon steel. The cost of shell and tube heat exchanger with CS shell and tubes with specified area is obtained as $93,000(Sinnot, 1999).

A.11.2.9. Heat Exchangers (E-307&E-308)

The cost of E-307 is determined based on floating head shell and tube heat exchanger with CS material. The heat exchanging area of E-307 is 82 m^2 and the corresponding cost is $25,000.

The cost of E-308 is determined based on U tube shell and tube heat exchanger with CS material. The heat exchanging area of E-308 is 893 m^2 and the corresponding cost for CS floating head heat exchangers is $112,000. the correction factor for U tube exchangers is 0.85. Consequently, the cost of E-308 is obtained as $95,200.

A.11.2.10. Ethanol Storage Tanks

The ethanol storage tanks are sized based on one week residence time. The ethanol volumetric flow rate is 128 m^3/h which corresponds to 21,544 m^3/week. The cost of storage tanks are expressed using Equation A.10.1. The total number of storage Tanks are taken as 4 with 5,400 m^3 volume. The coefficients for floating roof tanks are given as *C*=$2,900 and *n*=0.55. The corresponding cost for each tank is obtained as $300,574 (Sinnot,1999).

A.11.3. Cogeneration

A.11.3.1. Rotary Driers (D-101)

The cost of the rotary driers is calculated using the same methodology with Appendix A.10.3.7. and the results are given in Table A.11.2.

Table A.11.2 The costs of rotary drier.

	Capacity	Unit	Base Capacity		Base Cost ($)	n	Price($)
Drum	485	m^2	9.3	1	73,000	0.45	432,602
Engine	188	kW	7.5	1	12,300	0.56	74,713

A.11.3.2. Steam Boilers (F-103)

In 2015 plant, two different qualities of steam is produced in the plant one is 86 bars and 525 ºC and the other is 2.75 bar and 131 ºC. Therefore, two different boilers will be operated. One of them is producing 145 ton/h of steam at 86 bars and the other is producing 512 ton/h at 2.75 bar. The cost of the boiler producing 145 ton steam/h (86 bars) is obtained as $2,900,000 (Peters and Timmerhaus). The price includes complete boiler, feed water deaerator, boiler feed pumps, chemical injection system, stack and shop assembly labor.

The cost of second boiler is determined based on the correlations given by Peters and Timmerhaus (1991). The maximum boiler capacity for low pressure steam in the model is 454 ton/h with cost of $1.5 million . Therefore, the costing of the boiler second boiler is carried out based Equation A.10.4. Sinnot (1999) reports the sizing exponent for the boilers as 0.8. Inserting the base capacity and cost in Equation A.10.4 yields the cost of boiler producing 512 ton/h steam as $1,651,440.

A.11.3.3. Economizer (F-103)

The cost of the economizer is determined based on shell and tube heat exchangers. The area of the exchanger is 14,455 m^2. The model given by Sinnot (1999) predicts the cost of shell and tube heat exchangers with maximum size of 1,000 m^2. The cost of SS shell and tube heat exchanger with 1,000 m^2 area is obtained as $310,000. Perry and Green (1997) reports the exponent for Equation A.10.4 for heat exchanger cost calculation as 0.59. Using the values for 1,000 m^2 heat exchanger the cost of economizer is obtained as $1,498,900.

A.11.3.4. Turbine Generators

The cost of turbine generator is obtained using the costs and capacities given in NREL report (Adan *et al.*, 2002). The electricity generation capacity of their plant is 30.3 MW and the sizing exponent for the turbines is given as 0.71. The cost of the turbine used in their study is reported as $7,422,186. Using the base capacity and the cost of turbine generating 13.52 MW electricity is obtained as $4,185,053 (2002 prices).

A.11.3.5. Chimney

The height of the chimney is 40 m and the diameter is 11.2 m. To estimate the cost the wall thickness is taken as 0.4 meter on average which yields an external diameter of 12 m. The chimney is assumed to be a hollow cylinder and the volume of the wall of the chimney is obtained as the difference of the volume of the cylinder with 12 m diameter with that of 11.2 m diameter. The side wall volume of the chimney is obtained as 583 m^3 using,

$$Volume\,of\;Chimney\,Wall = \frac{\pi D_{ext}^2 H}{4} - \frac{\pi D_{int}^2 H}{4}$$

The costing is determined based on reinforced concrete wall costs. The costs of the reinforced concrete walls with 0.3 meter thickness is given as $87.6/ m^2 which gives the cost as $292.1/$m^3$ including the labor for the construction (Peters and Timmerhaus, 1991). Using this value the cost of chimney is obtained as $170,317.

A.11.3.6. Cooling Tower

The cost of cooling tower is calculated based on the capacity of 15,000 m^3 water/h capacity. Peters and Timmerhaus (1991) reports the cost of cooling tower with this capacity as $800,000.

A.11.4. Pretreatment and Hydrolysis

A.11.4.1. Pretreatment Reactor (R-401)

The reactor set is composed of,

- chip bin,
- steaming vessel,
- two screws one for transporting fibres from chip bin to steaming vessel and the other is in the steaming vessel to move the fibres,
- high pressure pump (will be given in pumps section)
- reactor operating at 14 bar.
- collection flash tank (V-401)

Chip Bin

Cost of chip bin is determined based on 15 minute residence time for bagasse fibres. The total volumetric flow rate of the bagasse is calculated as 1.78 m^3/min in Appendix A.9.1. based on this information the volume of the chip bin is 26.7 m^3. the cost is determined based on vertical cylinder tank. The cost is determined using Equation A.10.4. The cost of cylindrical vertical tank of volume 3.8 m^3 is given as $3,300 (Perry and Green, 1997). The same source reports the sizing exponent as 0.3. Using this information, the cost of the chip bin is obtained as $5,923.

Steaming Vessel

Heating with direct contact with steam does not require long residence times. Therefore, the residence time in the steaming vessel is taken as 5 minutes and the necessary volume is obtained as 8.9 m^3. The cost of the steaming vessel is obtained as $4,260 for carbon steel this cost is multiplied by 2 for SS material and the cost of the equipment is obtained as $8,520, using the same methodology with chip bin based on cylindrical vessels. To determine the required screw conveyor length the vessel diameter is taken as 1 m and the total length is obtained as 11.4 m.

The Screw Conveyors

The capacity of the conveyors is 5.7 ton/min. The screw diameter to manage this capacity is given as 0.4 m (Peters and Timmerhaus, 1991). The length of the first conveyor between the chip bin and the steaming vessel is taken as 2 m and the cost is obtained as $8,000. And the cost of the second conveyor is $17,000. The power consumption is 2.24 kW each.

Pretreatment reactor (R-401)

The volume of the reactor is 22 m^3. The reactor is priced based on jacketed reactors. Equation A.10.1 is used for the cost determination and the coefficients are given as, C= $2,400 and the sizing exponent n=0.6. Using this information the cost of the reactor is obtained as $123,957. To account for the continuity of the operation in case of clogging the number of reactors is taken as 2.

Collecting Flash Tank (V-401)

The volume of the flash vessel is 106 m^3 with diameter of 4.2 and height of 8 m. The cost is determined based on vertical pressure vessels as $54,000 for CS material. The material factor for SS is 2. Therefore, the total cost of the vessel is obtained as $108,000.

A.11.4.2. Screw Presses (S-401&S-404)

The cost of screw press of capacity 1.8 t/h is given as $121,517 (IRef-16) including the screw press and the filtration unit. The cake capacity of the screw presses S-401&404 are 204 and 102 ton/h respectively. The sizing exponent is taken as 0.46 based on screw conveyors (Perry and Green, 1997) and the costs are obtained as $2,330,222 and $1,694,040. The material factor for stainless steel is assumed as 1.44 based on the calculations carried out in Appendix A.10.4.3. Thus the total costs are obtained as $3,356,000 and $2,440,000 respectively for S-401 and S-404.

A.11.4.3. Heat Exchangers (E-401&E-402)

The areas of the exchangers are 384 and 13 m^2 respectively for E-401 and 402. The costs of the exchangers are determined based on floating head shell and tube heat exchangers. The material for exchangers is CS shell and SS tubes. The corresponding costs of equipments are obtained as $200,000 and $10,500 respectively.

A.11.4.4. Overliming and Conditioning Vessels (E-401&E-402)

The volume of the vessels is 260 m^3. The cost of the vessels is determined using the correlation given by Peters and Timmerhaus (1991) for agitated vessels. The cost of agitated vessel of specified volume is obtained as $170,000 including the drive for the agitator.

A.11.4.5. Cellulose Hydrolysis Vessels (BR-401)

The cost of the cellulose hydrolysis vessels is determined based on vessel and agitator of given power.

Vessels

The volumes of the vessels are 2,331 m^3 each. The costing is carried out based on ordinary process tanks. Brown (2003), reports the costs of vertical process vessels in terms of Equation A.10.4. The base capacity for a fermentor is given as 177 m^3 and the base cost is given as $176,000 for stainless steel fermentor, the sizing exponent n=0.71 using this values the cost of one vessel is obtained as $1,097,505

Agitators

Brown (2003), reports the cost of agitator of 87.3 kW as $133,400. Same source reports the sizing exponent for Equation A.10.4. as 0.5. using this values the cost of 42.2 kW agitator is obtained as $92,748.

A.11.4.6. Clarifiers (S-402)

The cost of the clarifiers is identical with the costs calculated in Appendix A.10.2.8.

A.11.4.7. Rotary Vacuum Filter (S-403)

The costs for the rotary vacuum filters are determined using the formula given in Equation A.10.4. Perry and Green, (1997) reports the base capacity, base cost and the exponent as 9.3 m^2 and $63,300 and 0.48 respectively. Using these values and filter area of one filter as 114 m^2, the cost of one filter equipment is obtained as $210,788 including the motor.

A.11.4.8. Fiber Conveyors

The belt conveyors are required to carry the fibres from Screw presses (S-401&S-404) to cellulose hydrolysis vessels (BR-401) and fibre drying (D-401). Since the configuration of the factory is not known the it is assumed that for each operation one conveyor of length 50m will be operated. The capacities of the conveyors are 204.24 and 99.54 ton/h. belt widths are taken as 0.75m for the former one and 0.5 m for the latter one to meet the capacity. The speed of the conveyors is taken as 45 m/s (Peters and Timmerhaus, 1991). The corresponding costs are obtained as $80,000 and $60,000 (Peters and Timmerhaus, 1991).

A.11.5. Waste Treatment

A.11.5.1. Waste Water Treatment Plant

The streams that requires to be treated in waste water treatment plant are <334>, <405> and <429>. And their total flow is 126.7 ton/h. The waste treatment plant is priced based on activated sludge system. Peters and Timmerhaus (1991) reports the cost of waste treatment plants in volumetric waste water flow. For the calculations it is assumed that the volumetric flow is 126.7 m^3/h and the corresponding cost for waste treatment plant is obtained as $1,800,000.

A.11.5.2. Waste Acid Solution Neutralization

This system is composed of agitated vessel and, clarifier and a rotary filter to remove the gypsum crystals. The system has similar structure with V-403, S-402 and S-403 system only the capacity is different. The total cost of that system is $698,000 with processing capacity of 527.6 ton/h. The capacity of the waste acid solution is 1,224 ton/h. The cost is calculated as $1,156,488 using Equation A.10.4 with sizing exponent of 0.6.

A.11.6.Pumps

The pumps that have different specifications than 2005 plant are listed in Table A.11.3. The detailed design is not carried out. But, to have an estimate on power requirements and costs, the pumps are designed in SuperPro designer using the same amount of water as the pumped media. The effect of the density and viscosity differences are implemented by manipulating the head requirements intuitively.

Table A.11.3 The power requirements and costs of the pumps
(The rest of the equipments are same with 2005 design)

Pump	Number	Material	Capacity (m^3/s)	Head (m)	Power (kW)	Cost ($/unit)
P-301	2	CS	0.031	30	13.09	16,153
P-302	2	SS	0.468	2	13.44	44,123
P-303	2	SS	0.385	2	11.06	40,805
P-304	2	SS	0.266	17	64.94	82,849
P-305	2	SS	0.021	20	6.03	30,352
P-306	2	SS	0.090	2	2.59	21,084
P-307	2	SS	0.221	20	63.48	82,096
P-308	2	SS	0.052	32	23.90	55,541
P-309	2	CS	0.157	32	72.15	31,972
P-310	2	CS	0.036	20	10.34	14,699
P-311	2	SS	0.119	13	22.22	53,945
P-312	2	SS	0.119	19	32.47	62,788
K-301	2	SS	12.638	4	632.64	32,541
K-302	2	SS	5.316	4	266.11	17,877
K-303	2	SS	81.440	1	1121.6	131,614
K-304	2	CS	22.031	1	303.43	18,285
K-305	2	CS	22.031	3	852.27	18,285
P-401	2	SS	0.168	140	482.56	184,785
P-402	2	SS	0.135	4	7.76	35,409
P-403	2	SS	0.147	15	31.67	62,161
P-404	2	CS	0.032	2	0.92	5,001
P-405	2	CS	0.147	15	31.67	23,000
P-406	2	CS	0.147	10	21.11	19,556
P-407	2	CS	0.003	3	0.13	2,596
P-408	2	SS	0.145	21	43.73	70,728
P-409	2	SS	0.002	21	0.60	11,277
P-410/A*	12	SS	0.139	21	41.92	69,542
P-410/B*	4	SS	0.139	2	3.99	25,417
P-411	2	SS	0.154	20	44.23	71,050
P-412	2	SS	0.009	3	0.39	9,325

* A is for the pumps between two sequential vessels B is for the pumps in pumping the hydrolyzate from final vessel to screw presses.

A.11.7.Equipment Purchase Costs and Installed Equipment Costs (IEC)

The equipments, prices of which are different from 2005 plant, are given in this section. The remaining equipments can be found in Appendix A.10.9.

Table A.11.4 Equipment purchase and installed equipment costs for clarification unit (2015)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
Lime kiln	F-101	SS	$123,266	1998	$174,788	1.40	$244,703
Lime kiln blower	F-101		$530	1990	$764	1.30	$994
Lime milk prep. tanks and agit.	V-101	SS	$81,600	1990	$117,663	1.40	$164,729
V-101 cooling			$2,300	1990	$3,316	2.10	$6,965
Sugar branch juice heater	E-101	SS/CS	$290,000	2003	$344,013	2.10	$722,428
Ethanol branch juice heater	E-102	SS/CS	$202,000	2003	$239,623	2.10	$503,208
Clarification unit sum			$3,922,316		$4,646,813		$7,434,777

Table A.11.5 Equipment purchase and installed equipment costs for crystallization and drying unit (2015)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
Reheater	E-209	CS	$61,000.00	1998	$72,361	2.1	$151,959
Reheater	E-210	CS	$40,000.00	1998	$47,450	2.1	$99,645
Reheater	E-211	CS	$65,500.00	1998	$77,700	2.1	$163,169
Reheater	E-212	CS	$32,800.00	1998	$38,909	2.1	$81,709
Crystallization and drying unit sum			$5,216,410		$6,940,825	2.02	$10,018,871

TableA.11.6 Equipment purchase and installed equipment costs for fermentation unit (2015)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
Juice cooler	E-301	CS	$270,000	1998	$320,288	2.1	$672,605
Product cooler	E-302	CS	$1,020,000	2003	$1,071,010	2.1	$2,249,122
Fermentors	V-301	SS	$11,137,378	1998	$13,211,743	1.2	$15,854,091
Fermentor cooling		SS	$370,216	1990	$533,834	1.2	$640,601
1st centrifuge	S-301	SS	$966,367	1998	$1,146,355	1.4	$1,604,897
2nd centrifuge	S-302	SS	$1,229,921	1998	$1,458,997	1.4	$2,042,596
Yeast sterilization tank	V-302	SS	$185,000	1990	$266,761	1.4	$373,466
Scrubber column	C-301	SS	$352,000	1998	$417,561	2.1	$876,878
Scrubber column packing	C-301	Plastic	$102,636	1998	$121,752	2.1	$255,680
Fermentation unit sum			$15,633,518		$18,548,302		$24,569,935

Table A.11.7 Equipment purchase and installed equipment costs for ethanol purification unit (2015)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
Beer column	C-302	SS	$382,800	1998	$454,097	2.1	$953,605
Beer column trays	C-302	SS	$253,130	1998	$300,276	2.1	$630,580
Rectification column	C-303	SS	$444,307	1998	$527,060	2.1	$1,106,827
Rectification column trays	C-303	SS	$427,550	1998	$507,182	2.1	$1,065,083
Molecular sieve column	C-304	CS	$94,600	1998	$112,219	2.1	$235,661
Molecular sieve packing	C-304	Zeolite	$1,015,580	2002	$1,092,709	2.1	$2,294,688
Molecular sieve cooling	C-304	CS	$76,608	1990	$110,465	2.1	$231,977
Beer column condenser	E-304	SS	$553,000	1998	$655,998	2.1	$1,377,595
Rectification col. condenser	E-306	CS	$321,302	1998	$381,145	2.1	$800,405
Nitrogen dehumidifier	E-309	CS	$87,000	1998	$103,204	2.1	$216,728
Beer column reboiler	E-303	SS	$1,009,830	1998	$1,197,913	2.1	$2,515,618
Rectification col. reboiler	E-305	CS	$93,000	1998	$110,321	2.1	$231,675
Molecular sieve feed cooler	E-307	CS	$25,000	1998	$29,656	2.1	$62,278
Nitrogen heater	E-308	CS	$95,200	1998	$112,931	2.1	$237,156
Ethanol storage tanks		CS	$1,310,010	1998	$1,554,002	1.4	$2,175,603
Pumps	P-301	CS	$32,306	2005	$22,303	2.8	$62,448
Pumps	P-302	SS	$88,246	2005	$60,078	2	$120,156
Pumps	P-303	SS	$81,610	2005	$58,408	2	$116,816
Pumps	P-304	SS	$165,698	2005	$122,498	2	$244,996
Pumps	P-305	SS	$60,704	2005	$52,526	2	$105,052
Pumps	P-306	SS	$42,168	2005	$33,014	2	$66,028
Pumps	P-307	SS	$118,192	2005	$118,192	2	$236,384
Pumps	P-308	SS	$59,442	2005	$59,442	2	$118,884
Pumps	P-309	CS	$47,822	2005	$47,822	2.8	$133,901
Pumps	P-310	CS	$19,435	2005	$19,435	2.8	$54,417
Pumps	P-311	SS	$83,464	2005	$83,464	2	$166,928
Pumps	P-312	SS	$95,104	2005	$95,104	2	$190,208
Pumps	K-301	SS	$34,118	2005	$34,118	1.3	$44,354
Pumps	K-302	SS	$26,482	2005	$26,482	1.3	$34,426
Pumps	K-303	SS	$84,475	2005	$84,475	1.3	$109,818
Pumps	K-304	CS	$21,722	2005	$21,722	1.3	$28,239
Pumps	K-305	CS	$21,722	2005	$21,722	1.3	$28,239
Ethanol purification unit sum			$8,209,986		$8,209,986		$15,996,772

Table A.11.8 Equipment purchase and installed equipment costs for pretreatment and hydrolysis unit (2015)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
Chip bin	R-401	CS	$5,933	1996	$7,390	1.4	$10,346
Steaming vessel	R-401	SS	$4,260	1996	$5,306	1.2	$6,367
Screw conveyor	R-401	SS	$8,000	1990	$11,536	1.3	$14,996
Screw conveyor	R-402	SS	$17,000	1990	$24,513	1.3	$31,867
Pretreatment reactor	R-401	SS	$247,914	1998	$294,089	1.7	$499,951
Collection flash tank	V-401	SS	$108,000	1998	$128,115	1.2	$153,738
Screw press	S-401	SS	$3,355,520	2005	$3,355,520	1.4	$4,697,728
Screw press	S-401	SS	$2,439,418	2005	$2,439,418	1.4	$3,415,185
Hydrolysate cooler	E-401	CS/SS	$200,000	1998	$237,251	2.1	$498,226
Hydrolysate heater	E-402	CS/SS	$20,500	1998	$24,318	2.1	$51,068
Overliming vessel	V-402	SS	$170,000	1990	$245,132	1.2	$294,158
Conditioning vessel	V-403	SS	$170,000	1990	$245,132	1.2	$294,158
Cellulose hydrolysis vessel	BR-401	SS	$8,780,040	2003	$9,219,131	1.2	$11,062,957
Cellulose hydrolysis agit.	BR-401	SS	$741,984	2003	$779,091	1.25	$973,863
Clarifiers	S-402	SS	$317,260	2002	$341,354	1.5	$512,032
Rotary vacuum filter	S-403		$210,788	1996	$262,553	1.4	$367,574
Fibre conveyors			$80,000	1990	$115,356	1.3	$149,963
Fibre conveyors			$60,000	1990	$86,517	1.3	$112,472
Pumps	P-401	SS	$369,570	2005	$369,570	2	$739,140
Pumps	P-402	SS	$70,818	2005	$70,818	2	$141,636
Pumps	P-403	SS	$124,322	2005	$124,322	2	$248,644
Pumps	P-404	CS	$10,002	2005	$10,002	2.8	$28,006
Pumps	P-405	CS	$46,000	2005	$46,000	2.8	$128,800
Pumps	P-406	CS	$39,112	2005	$39,112	2.8	$109,514
Pumps	P-407	CS	$5,192	2005	$5,192	2.8	$14,538
Pumps	P-408	SS	$141,456	2005	$141,456	2	$282,912
Pumps	P-409	SS	$22,554	2005	$22,554	2	$45,108
Pumps	P-410-A	SS	$973,588	2005	$973,588	2	$1,947,176
Pumps	P-410-B	SS	$50,834	2005	$50,834	2	$101,668
Pumps	P-411	SS	$142,100	2005	$142,100	2	$284,200
Pumps	P-412	SS	$18,650	2005	$18,650	2	$37,300
Pretreatment and hydrolysis unit sum			$18,950,815		$19,835,919		$27,255,291

Table A.11.9 Equipment purchase and installed equipment costs for cogeneration and aux. process equipments (2015)

Equipment	Name	Material	Equipment Price	Base year	2005 Price	EIF	IEC
Rotary driers	D-101	CS	$1,297,806	1996	$1,616,517	1.4	$2,263,124
Rotary drum motors	D-101		$224,139	1996	$279,182	1.4	$279,182
Evaporator	E-403	SS	$374,247	1998	$443,951	2.1	$443,951
Boiler	F-401	CS	$2,900,000	1990	$4,181,663	1.3	$4,181,663
Boiler	F-402	CS	$1,651,440	1990	$2,381,299	1.3	$2,381,299
Economizer	F-103	SS	$1,180,448	1998	$1,400,309	2.1	$1,400,309
Turbine generator	F-103		$4,185,053	2002	$4,502,888	1.5	$4,502,888
Chimney			$170,317	1990	$245,589	1	$245,589
Waste water treatment			$1,800,000	1990	$2,595,515	1	$2,595,515
Waste acid neutralization			$1,156,488	1996	$1,440,495	1.3	$1,440,495
Cooling tower			$800,000	1990	$1,153,562	1.2	$1,153,562
Cogeneration and auxiliary equipment sum			$15,739,938		$20,240,972		$20,887,579

A.11.8. Operational Costs

A.11.8.1. Fixed Operational Costs

Labour Cost

The labour requirement of the plant is determined based on the information given by Brown (2003) in Table A.10.18. The labour requirements in addition to 2005 equipments are given in Table A.11.10. The labour requirements for the remaining equipments can be found in Table A.10.19. The labour requirement is obtained as 23 operators/shift. The cost of one operator is obtained as $2,796/year (See Appendix A.10.11.1)

Table A.11.10 Operator requirements of the equipments (2015)

Pretreatment and Hydrolysis		
Conveyors	R-401	0.40
Pretreatment reactor	R-401	0.50
Screw presses	S-401	0.40
Heat exchangers	E-401	0.20
Overliming vessels	V-402	0.30
Cellulose hydrolysis vessel	BR-401	0.50
Clarifiers	S-402	0.20
Cogeneration		
Rotary driers	D-101	0.50
Evaporator	E-403	0.25
Boilers	F-401	2.00
Turbine generator	F-103	3.00
Waste water treatment		2.00
Cooling tower		1.00
2015 Total		**23.00**

A.12. Equipment Summaries and Specification Sheets

A.12.1. Equipment Summaries

REACTORS & VESSELS – SUMMARY

EQUIPMENT NR. : **NAME :**	**BR-301** **Fermentors**	**BR-401** **Cellulase Hydrolysis**	**R-401** **Pretreatment Reactor**	**V-101** **Lime Milk Preparation**
Pressure [bara] :	1.013	1.013	14	1
Temp. [^{o}C] :	30	65	190	90
Volume [m^3] :	1,481	2,331	22	20.5
Diameter [m] :	10.0	12.0	-	2.1
L or H [m] :	18.8	20.6	-	5.9
Liquid Filling [%] :	90	90	90	90
Number				
- Series :	7	4	-	-
- Parallel :	2	2	2	2
Materials of Construction	SS304	SS304	SS304	SS304
Agitation power [kW]	-	339.2	-	20.5
Cooling/Heating	Coils	-	-	Jacket
EQUIPMENT NR. : **NAME :**	**V-102** **Sugar Brach Liming Tank**	**V-103** **Sugar Brach Liming Tank**	**V-104** **Sugar Branch Flash Tank**	**V-105** **Ethanol Branch Flash Tank**
Pressure [bara] :	1.013	1.013	1.013	1.013
Temp. [^{o}C] :	44	45.2	100	100
Volume [m^3] :	400	216	23.9	13.2
Diameter [m] :	5.7	4.7	3.9	2.9
L or H [m] :	17.1	14.1	2.0	2.0
Liquid Filling [%] :	90	90	-	-
Number				
- Series :	1	1	1	1
- Parallel :	-	-	-	-
Materials of Construction	SS304	SS304	CS	CS
Agitation power [kW]	364	196	-	-
Cooling/Heating	-	-	-	-

Remarks:
(1) SS = Stainless Steel; CS = Carbon Steel

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REACTORS & VESSELS – SUMMARY

EQUIPMENT NR. : **NAME :**	**V-201** **1st Crystallizers**	**V-202** **2nd Crystallizers**	**V-301** **Yeast Acidification Tanks**	**V-401** **Pretreatment Collection Tank**
Pressure [bara] :	1.013	1.013	1.013	1.013
Temp. [°C] :	40	40	30	100
Volume [m^3] :	137.2	122.3	312.0	106.0
Diameter [m] :	3.5	3.5	5	4.2
L or H [m] :	12.9	11.5	15.9	8.0
Liquid Filling [%] :	90	90	90	-
Number				
- Series :	6	6	1	1
- Parallel :	-	-	-	-
Materials of Construction	CS	CS	SS316	SS316
Agitation power [kW]	165.6	148.2	426	-
Cooling/Heating	Coils	Coils	-	-
EQUIPMENT NR. : **NAME :**	**V-402** **Overliming Vessel**	**V-403** **Conditioning Vessel**		
Pressure [bara] :	1.013	1.013		
Temp. [°C] :	50	50		
Volume [m^3] :	276	276		
Diameter [m] :	5.0	5.0		
L or H [m] :	14.0	14.0		
Liquid Filling [%] :	90	90		
Number				
- Series :	1	1		
- Parallel :	-	-		
Materials of Construction				
Agitation power [kW]				
Cooling/Heating				

Remarks:
(1) SS = Stainless Steel; CS = Carbon Steel

COLUMNS – SUMMARY

EQUIPMENT NR. : NAME :	C-101 Sulphitation Column	C-301 Ethanol Scrubber	C-302 Beer Column	C-303 Rectification Column	C-304 Molecular Sieve Ads. Colm.
Pressure [bara] :	1.013	1.013	1.013	1.013	1.013
Temp. [°C] :	44-713	45-52	20-100	77.4-100	60-200
Volume [m^3] :					
Diameter [m] :	2.8	3.3	3.17-5.52	2.2-6.8	3.2
L or H [m] :	2.3	30.0	20	35	8.4
Internals					
- Tray Type :	n.a.	n.a.	Valve	Valve	n.a.
- Stage Number :	3	40	20	35	n.a.
- Fixed Packing					
Type :	Rasching rings	Pall ring	n.a.	n.a	3Å Zeolite
Shape :	50 mm	50 mm	n.a.	n.a.	Cylindirical
- Catalyst	n.a.	n.a.	n.a.	n.a.	n.a.
Type :	n.a.	n.a.	n.a.	n.a.	n.a.
Shape :					
-					
-					
-					
Number					
-Series :	1	1	1	1	1
-Parallel :	-	-	-	-	2
Materials of Construction *(1)* **:**	Packing: Ceramic Column: SS304	Polypropylene SS304	Trays: SS304 Colm.:SS304	Trays: CS Colm,: CS	Pack.: Zeolite Colm.: CS

Remarks:
(1) SS = Stainless Steel; CS = Carbon Steel

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HEAT EXCHANGERS, COOLERS & HEATERS - SUMMARY

EQUIPMENT NR : **NAME :**	**E-101** **Sugar Branch Juice Heater**	**E-102** **Ethanol Branch Juice Heater**	**E-201** **1st Effect Evaporator**	**E-202** **2nd Effect Evaporator**	**E-203** **3rd Effect Evaporator**
Substance - **Tube :** - **Shell (1) :**	Stream <108> Stream <436>	Stream <116> Stream <436>	Stream <114> L.P.S.	1st effect vap. 1st effect syr	2nd effect vap. 2nd effect syr.
Duty [kW] :	47,257	25,265	-	-	-
Heat Exchange Area $[m^2]$ **:**	527	283	1,879	1,348.5	1,561.5
Number - **Series :** - **Parallel :**	1 -	1 -	1 2	1 2	1 2
Pressure [bar] - **Tube :** - **Shell :**	1.013 1.013	1.013 1.013	2.75 2.32	2.32 1.89	1.89 1.46
Temperature In/Out [ºC] - **Tube :** - **Shell :**	45.2/103 125/125	44/103 125/125	130.7/130.7 100/125	125/125 118.6/118.6	118.6/118.6 110.7/110.7
Materials of Construction (2) :	Tubes: SS304 Shell : SS304	Tubes: SS304 Shell : SS304	Tubes: CS Shell :CS	Tubes: CS Shell :CS	Tubes: CS Shell :CS
Other :					
EQUIPMENT NR : **NAME :**	**E-204** **4th Effect Evaporator**	**E-205** **5th Effect Evaporator**	**E-206** **6th Effect Evaporator**	**E-207** **1st Vacuum Pan**	**E-208** **2nd Vacuum Pan**
Substance - **Tube :** - **Shell (1) :**	3rd effect vap. 3rd effect syr.	4th effect vap. 4th effect syr.	5th effect vap. 5th effect syr.	Stream <202> L.P.S.	Stream <214> L.P.S.
Duty [kW] :	-	-	-	16,276	5,324
Heat Exchange Area $[m^2]$ **:**	1,836.5	2,229	2,991	1,566.5	891.5
Number - **Series :** - **Parallel :**	1 2	1 2	1 2	1 -	1 -
Pressure [bar] - **Tube :** - **Shell :**	1.46 1.03	1.03 0.60	0.60 0.17	2.75 0.38	2.75 0.38
Temperature In/Out [ºC] - **Tube :** - **Shell :**	110.7/110.7 100.5/100.5	100.5/100.5 86/86	86/86 56.3/56.3	130.6/130.6 75/75	130.6/130.6 75/75
Materials of Construction (2) :	Tubes: CS Shell :CS	Tubes: CS Shell :CS	Tubes: CS Shell :CS	Tubes: CS Shell :CS	Tubes: CS Shell :CS
Other :					
Remarks	(1)C.W. = cooling water H.W. =saturated water, L.P.S.= low pressure steam Ch.W.= chilled water Vap.=vapour, syr.= syrup (2) SS = stainless steel; CS = carbon steel				

HEAT EXCHANGERS, COOLERS & HEATERS - SUMMARY

EQUIPMENT NR : **NAME :**	**E-209** **Juice Heater**	**E-210** **Juice Heater**	**E-211** **Juice heater**	**E-212** **Juice Heater**	**E-213** **Air Heater**
Substance - **Tube :** - **Shell (1) :**	Stream <201> H.W.	Stream <210> H.W.	Stream <212> H.W.	Stream <217> H.W.	Stream <223> L.P.S.
Duty [kW] :	2,008	1,100	995.5	484	593
Heat Exchange Area [m^2] :	421	178	207	78	41
Number - **Series :** - **Parallel :**	1 -	1 -	1 -	1 -	1 -
Pressure [bar] - **Tube :** - **Shell :**	1.013 2.75	1.013 2.75	1.013 2.75	1.013 2.75	1.013 2.75
Temperature In/Out [ºC] - **Tube :** - **Shell :**	56/75 130.6/120	40/55 130.6/120	55/75 130.6/120	40/55 130.6/120	30/80 130.7/130.7
Materials of Construction (2) :	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS
Other :					

EQUIPMENT NR : **NAME :**	**E-301** **Juice Cooler**	**E-302** **Distillation Feed Heater**	**E-303** **Beer Colm. Reboiler**	**E-304** **Beer Colm. Condenser**	**E-305** **Rectification Colm. Reboiler**
Substance - **Tube :** - **Shell (1) :**	Stream <301> C.W.	Stream <308> Stream <319>	Stream<311> L.P.S.	Stream <312> Ch.W.	Stream <318> L.P.S.
Duty [kW] :	53,726	58,133	165,553	36,738	25,729
Heat Exchange Area [m^2] :	2,144	2,696	4,168	1,502	653.4
Number - **Series :** - **Parallel :**	1 -	1 -	1 -	1 -	1 -
Pressure [bar] - **Tube :** - **Shell :**	1.013 1.013	1.013 1.013	1.013 2.75	1.013 1.013	1.013 2.75
Temperature In/Out [ºC] - **Tube :** - **Shell :**	79.5/30 25/40	30/80 100/40.8	100/100 130.7/130.7	78/20 10/15	100/100 130.7/130.7
Materials of Construction (2) :	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : SS304 Shell : CS	Tubes : SS304 Shell : CS	Tubes : CS Shell : CS
Other :					
Remarks	(1)C.W. = cooling water H.W. =saturated water, L.P.S.= low pressure steam Ch.W.= chilled water Vap.=vapour, syr.= syrup (2) SS = stainless steel; CS = carbon steel				

HEAT EXCHANGERS, COOLERS & HEATERS - SUMMARY

EQUIPMENT NR : **NAME :**	**E-306** **Rectification Colm. Cond.**	**E-307** **Molec. Sieve Feed Cooler**	**E-308** **Nitrogen Heat Exch.**	**E-309** **Water Condenser**	**E-310** **Product Cooler**
Substance - **Tube :** - **Shell (1) :**	 Stream <314> C.W.	 Stream <314> C.W.	 Stream <321> Stream <324>	 Stream <321> Refrigerant	 Stream <316> C.W.
Duty [kW] :	135,558	1,603	6,158	1,977	2,340
Heat Exchange Area [m^2] :	4,951.4	81.8	893	565	433.3
Number - **Series :** - **Parallel :**	 1 -	 1 -	 1 -	 1 -	 1 -
Pressure [bar] - **Tube :** - **Shell :**	 1.013 1.013	 1.013 1.013	 1.013 1.013	 1.013 1.013	 1.013 1.013
Temperature In/Out [ºC] - **Tube :** - **Shell :**	 78.2/77.5 25/60	 77.5/60 25/40	 5/180 200/30	 30/5 0/0	 60/30 25/40
Materials of Construction (2) :	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS	Tubes : CS Shell : CS
Other :					
EQUIPMENT NR : **NAME :**	**E-401** **Hydrolyzate Cooler**	**E-402** **Hydrolyzate Heater**	**E-403** **Vinasse Evaporator**		
Substance - **Tube :** - **Shell (1) :**	 Stream <407> C.W.	 Stream <419> L.P.S.	 L.P.S. Stream <311>		
Duty [kW] :	22,913	1,473	19,375		
Heat Exchange Area [m^2] :	384	13	714,4		
Number - **Series :** - **Parallel :**	 1 -	 1 -	 1 -		
Pressure [bar] - **Tube :** - **Shell :**	 1.013 1.013	 1.013 2.75	 1.013 1.013		
Temperature In/Out [ºC] - **Tube :** - **Shell :**	 84.4/50 25/40	 50/54.4 130.7/130.7	 100/103 130.7/130.7		
Materials of Construction (2) :	Tubes : SS304 Shell : CS	Tubes : SS304 Shell : CS	Tubes : SS304 Shell : SS304		
Other :					
Remarks	(1)C.W. = cooling water H.W. =saturated water, L.P.S.= low pressure steam Ch.W.= chilled water Vap.=vapour, syr.= syrup (2) SS = stainless steel; CS = carbon steel				

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A.12.2. Equipment Specification Sheets

FERMENTOR – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**BR-301**	**In Series**	**7**
NAME	:	**Ethanol Fermentors**	**In Parallel**	**2**
General Data				
Service	:	~~Buffer~~ / ~~Storage~~ / ~~Separation~~/ Reaction		
Type	:	Fermentor		
Position	:	~~Horizontal~~ /Vertical		
Internals	:	~~Demister~~ / ~~Plate~~ / Coil / ~~Gas distributor~~ / ~~Jacket~~ / ~~Filter~~		
~~Heating~~/Cooling				
Medium		Water		
- **Flow rate**	**[kg/s]**	426.3		
- **Pressure**	**[bar]**	1.013		
- **Temperature**	**[°C]**	30		
Vessel Diameter	**[m]**	10		
Vessel Height	**[m]**	18.8		
Vessel Total Volume	**[m^3]**	1,481		
Vessel Material		Stainless steel		

PROCES CONDITIONS

Stream Details		**<302> (IN)**		**<333> (IN)**		**<303>(OUT)**		**<304> (OUT)**	
Temperature	**[°C]**	30		30		30		30	
Pressure	**[bar]**	1.013		1.013		1.013		1.013	
Density	**[kg/m^3]**	1,067		1,131		1.73		1,087	
Mass Flow	**[kg/s]**	290.7		134.9		21.9		403.4	
Mole Flow	**[kmol/s]**	13.1		7.08		0.51		20.7	
Composition		Wt%	Mol%	Wt%	Mol%	Wt%	Mol%	Wt%	Mol%
Acetic Acid		0.287	0.106			0.009	0.007	0.273	0.089
Ash		0.551	0.305					0.397	0.194
Yeast				20.000	15.478			7.102	5.638
CO$_2$						96.530	94.332	1.173	0.521
Cellulase		0.001	0.000					trace	trace
Cellulose		0.003	0.000					0.002	trace
Ethanol						1.799	1.680	6.443	2.732
Furfural		0.035	0.008			trace	trace	0.025	0.005
Glycerol								0.310	0.066
Hemicellulose		0.002	0.000					0.002	trace
Hexoses		6.749	0.829						
HMF		0.006	0.001			trace	trace	0.005	0.001
Impurity		1.739	1.161					1.194	0.703
Isoamylalcohol						trace	trace	0.044	0.010
Lactic Acid								0.018	0.004
Lignin		0.019	0.002					0.014	0.002
Lignin Monomers		0.132	0.032					0.095	0.020
Pentoses		3.569	0.526						
Succinic Acid								0.032	0.005
Sucrose		8.529	0.551						
H$_2$SO$_4$		0.000	0.000					trace	trace
Sulphur									
Water		78.664	96.583	80.000	84.522	1.671	3.988	83.138	90.098
Xylitol								0.004	0.001

Remarks

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BIOREACTOR – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**BR-401**	**In Series**	**4**
NAME	:	**Cellulose hydrolysis bioreactor**	**In parallel**	**2**

General Data

Service	:	~~Buffer~~ / ~~Storage~~ / ~~Separation~~/ Reaction
Type	:	Bioreactor
Position	:	~~Horizontal~~ /Vertical
Internals	:	~~Demister~~ / ~~Plate~~ / ~~Coil~~ / ~~Gas distributor~~ / ~~Jacket~~ / ~~Filter~~ / Agitator

~~Heating~~/Cooling

Medium		Water/fibre
- **Flow rate**	**[kg/s]**	145.4
- **Pressure**	**[bar]**	1.013
- **Temperature**	**[°C]**	65
Vessel Diameter	**[m]**	12
Vessel Height	**[m]**	20.6
Vessel Total Volume	**[m^3]**	2,331
Vessel Material		Stainless steel

PROCES CONDITIONS

Stream Details		**<420> (IN)**		**<421> (IN)**		**<437>(IN)**		**<422> (OUT)**	
Temperature	**[°C]**	54.4		84.4		30		65	
Pressure	**[bar]**	1.013		1.013		1.013		1.013	
Density	**[kg/m^3]**	1,066		1,600		1,600		1,050	
Mass Flow	**[kg/s]**	88.5		56.5		0.4		145.4	
Mole Flow	**[kmol/s]**	4.2		1.4		trace		5.6	
Composition		Wt%	Mol%	Wt%	Mol%	Wt%	Mol%	Wt%	Mol%
Acetic Acid		0.722	0.251	0.344	0.231			0.573	0.248
Ash		1.593	0.832	0.760	0.768			1.265	0.823
Cellulase						100.00	100.00	0.252	0.013
Cellulose		trace	trace	30.827	3.845			1.199	0.096
Furfural		0.086	0.019	0.046	0.019			0.070	0.019
Hemicellulose				2.499	0.382			0.971	0.096
Hexoses		1.894	0.220	0.903	0.203			13.490	1.949
HMF		0.016	0.003	0.008	0.003			0.013	0.003
Impurity		1.583	0.997	0.752	0.917			1.256	0.986
Lignin		trace	trace	19.421	4.500			7.551	1.127
Lignin Monomers		0.318	0.073	0.180	0.080			0.264	0.075
Pentoses		8.984	1.251	4.282	1.153			7.133	1.237
Sucrose		2.310	0.141	1.078	0.127			1.825	0.139
H_2SO_4		trace	trace					trace	trace
Water		83.215	96.465	39.245	88.003			64.711	93.438

Remarks

REACTOR – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**R-401**	**In Series**	**1**
NAME	**:**	**Pretreatment Reactor**	**In parallel**	**2**

General Data

Service	**:**	~~Buffer~~ / ~~Storage~~ / ~~Separation~~/ Reaction
Type	**:**	Plug Flow Reactor
Position	**:**	~~Horizontal~~ /Vertical
Internals **:**		~~Demister~~ / ~~Plate~~ / ~~Coil~~ / ~~Gas distributor~~ / ~~Jacket~~ / ~~Filter~~ / Steam Injection

~~Heating~~/Cooling

Medium		Water/fibre
Flow rate	**[kg/s]**	145.4
Pressure	**[bar]**	14
Temperature	**[°C]**	190
Vessel Diameter	**[m]**	-
Vessel Height	**[m]**	-
Vessel Total Volume	**[m^3]**	22
Vessel Material		Stainless steel

PROCES CONDITIONS

Stream Details		**<403> (IN)**		**<404> (IN)**		**<405>(OUT)**		**<406> (OUT)**	
Temperature	**[°C]**	275		100		100		100	
Pressure	**[bar]**	25		14		1.013		1.013	
Density	**[kg/m^3]**	-		1,003		0.61		917	
Mass Flow	**[kg/s]**	20.2		148.1		24.3		143.0	
Mole Flow	**[kmol/s]**	1.1		5.7		1.3		5.5	
Composition		**Wt%**	**Mol%**	**Wt%**	**Mol%**	**Wt%**	**Mol%**	**Wt%**	**Mol%**
Acetic Acid						0.515	0.157	0.583	0.253
Ash								1.289	0.840
Bagasse				31.127	3.159				
Cellulose								12.248	0.985
Furfural						1.522	0.290	0.093	0.025
Hemicellulose								0.993	0.098
Hexoses								1.529	0.221
HMF						0.279	0.040	0.017	0.004
Impurity				0.578	0.456			1.268	0.997
Lignin								7.716	1.153
Lignin Monomers								0.406	0.116
Pentoses								7.253	1.260
Sucrose				1.684	0.129			1.744	0.133
H_2SO_4				0.404	0.108			0.418	0.111
Water		100.00	100.00	66.207	96.148	98.200	99.670	65.024	94.055

Remarks

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ABSORBTION COLUMN – SPECIFICATION SHEET

EQUIPMENT NUMBER : **C-301**
NAME : **Ethanol Scrubber**

General Data

Service : - ~~distillation~~/ ~~extraction~~/ absorption / ———
Column Type : - packed / ~~tray~~ / ~~spray~~ / ———
Tray Type : - ~~cap~~ / ~~sieve~~ / ~~valve~~ / none—
Tray Number *(1)*
- **Theoretical** : -
- **Actual** : 40
- **Feed (actual)** : 40

Stage Distance (HETP) [m] : 0.75 Packing Material : Polypropylene *(2)*
Column Diameter [m] : 3.3 Column Material : SS304 *(2)*
Column Height [m] : 30
Heating : - none / ~~open steam~~ / ~~reboiler~~ / ————

Process Conditions

Stream Details	**<303> (IN)**		**<312> (IN)**		**<326> (IN)**		**<307> (OUT)**		**<327> (OUT)**	
Temp. [°C]	30		20		30		51.5		45.2	
Pressure [bara]	1.013		1.013		1.013		1.013		1.013	
Density [kg/m^3]	1.73		1.61		998		947		1.59	
Mass Flow [kg/s]	21.8		8.6		30.4		33.9			
Mole Flow [kmol/s]	0.51		0.22		1.69		1.77			
Composition	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Acetic Acid	0.009	0.007	0.035	0.023			0.015	0.005		
CO_2	96.530	94.332	59.341	52.235			1.040	0.453	95.870	90.490
Ethanol	1.799	1.680	30.261	25.451			8.801	3.662	0.007	0.007
Furfural	trace	trace	0.023	0.009			0.006	0.001		
HMF	trace	trace	trace	trace			trace	trace		
Isoamylalcohol	trace	trace	trace	trace			trace	trace		
Lactic Acid			trace	trace			trace	trace		
Water	1.671	3.988	10.375	22.304	100.00	100.00	90.154	95.884	4.123	9.504

Column Internals

Trays *Not applicable*

Packing
Type : Pall rings
Material : Polypropylene
Volume [m^3] : 257
Length [m] : 0.05
Width [m] : 0.05
Height [m] : 0.05

Remarks:
Stage numbering from top to bottom.
SS = Stainless Steel; CS = Carbon Steel.

DISTILLATION COLUMN – SPECIFICATION SHEET

EQUIPMENT NUMBER : C-302
NAME : Bear Column

General Data

Service : - distillation/ ~~extraction~~/ ~~absorption~~/ ———
Column Type : - ~~packed~~ / tray / ~~spray~~ / ———
Tray Type : - ~~cap~~ / ~~sieve~~ / valve / ~~none~~
Tray Number *(1)*
- **Theoretical** : -
- **Actual** : 20
- **Feed (actual)** : 4
- **Side flow** : 6

Stage Distance (HETP) [m] : 1 — **Tray Material** : **SS304** *(2)*
Column Diameter [m] : 3.2-5.5 — **Column Material** : **SS304** *(2)*
Column Height [m] : 20
Heating : - ~~none~~ / ~~open steam~~ / reboiler / ——— *(3)*

Process Conditions

Stream Details	**<309> (IN)**		**<310> (OUT)**		**<311> (OUT)**		**<312> (OUT)**		**<313> (OUT)**	
Temp. [°C]	80		92.1		100		20		100	
Pressure [bara]	1.013		1.013		1.013		1.013		1.013	
Density [kg/m^3]	900		0.80		925.6		1.61		918.5	
Mass Flow [kg/s]	294.1		65.1		18.5		8.6		201.9	
Mole Flow [kmol/s]	14.9		2.71		0.82		0.22		11.15	
Composition	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Acetic Acid	0.376	0.124	0.324	0.130	0.205	0.078	0.035	0.023	0.423	0.128
Ash	0.545	0.269			8.648	4.907				
CO_2	1.729	0.775	0.008	0.004			59.341	52.235		
Cellulase	0.001	0.000			0.011	0.000				
Cellulose	0.003	0.000			0.049	0.003				
Ethanol	9.852	4.221	40.536	21.109	trace	trace	30.261	25.451	0.002	0.001
Furfural	0.035	0.007	0.157	0.039	trace	trace	0.023	0.009	trace	trace
Glycerol	0.426	0.091	trace	trace	1.997	0.492			0.437	0.086
Hemicellulose	0.002	0.000			0.038	0.003				
HMF	0.006	0.001	0.029	0.006	trace	trace	trace	trace	trace	trace
Impurity	1.638	0.975			26.002	17.804				
Isoamylalcohol	0.061	0.015	0.048	0.014	0.032	0.009	trace	trace	0.070	0.015
Lactic Acid	0.025	0.006	trace	trace	0.113	0.029	trace	trace	0.027	0.005
Lignin	0.019	0.002			0.297	0.039				
Lignin Monomers	0.130	0.028			2.067	0.516				
Succinic Acid	0.044	0.007	trace	trace	0.205	0.039			0.046	0.007
H_2SO_4	0.000	0.000			0.001	0.000				
Water	85.479	93.602	59.223	78.828	60.453	76.145	10.375	22.304	99.419	99.886
Xylitol	0.005	0.001			0.086	0.013				

Column Internals

Trays (Not available)
Number of ~~caps~~ / valves / ——— : …
Active Tray Area [m^2] : …
Weir Length [mm] : …
Diameter of chute pipe / hole / ———[mm] : …

Packing *Not Applicable*
Type :
Material :
Volume [m^3] :
Length [m] :
Width [m] :
Height [m] :

Remarks:
(1) Tray numbering from top to bottom.
(2) SS = Stainless Steel; CS = Carbon Steel.
(3) Reboiler is E-303; operates with LP steam.

DISTILLATION COLUMN – SPECIFICATION SHEET

EQUIPMENT NUMBER : C-303
NAME : Rectification Column

General Data

Service : - distillation/ ~~extraction~~/ ~~absorption~~/ ———
Column Type : - ~~packed~~ / tray / ~~spray~~ / ———
Tray Type : - ~~cap~~ / ~~sieve~~ / valve / ~~none~~—
Tray Number *(1)*
- **Theoretical** : -
- **Actual** : 35
- **Feed (actual)** : 25

Stage Distance (HETP) [m] : 1 **Tray Material** : **CS** *(2)*
Column Diameter [m] : 2.2-6.8 **Column Material** : **CS** *(2)*
Column Height [m] : 20
Heating : - ~~none~~ / ~~open steam~~ / reboiler / ———— *(3)*

Process Conditions

Stream Details	**<310> (IN)**		**<314> (OUT)**		**<318> (OUT)**					
Temp. [°C]	92.1		77.5		100					
Pressure [bara]	1.013		1.013		1.013					
Density [kg/m^3]	0.80		725		918					
Mass Flow [kg/s]	65.1		28.5		36.6					
Mole Flow [kmol/s]	2.71		0.69		2.02					
Composition	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Acetic Acid	0.324	0.130	trace	trace	0.577	0.174				
CO_2	0.008	0.004	0.018	0.016						
Ethanol	40.536	21.109	92.504	82.862	0.008	0.003				
Furfural	0.157	0.039	trace	trace	0.279	0.053				
Glycerol	trace	trace			trace	trace				
HMF	0.029	0.006	trace	trace	0.052	0.007				
Isoamylalcohol	0.048	0.014			0.085	0.019				
Lactic Acid	trace	trace			trace	trace				
Succinic Acid	trace	trace			trace	trace				
Water	59.223	78.828	7.478	17.122	99.576	99.918				

Column Internals

Trays (Not available)
Number of ~~caps~~ / valves / ———— : …
Active Tray Area [m^2] : …
Weir Length [mm] : …
Diameter of chute pipe / hole / ————[mm] : …

Packing *Not Applicable*
Type :
Material :
Volume [m^3] :
Length [m] :
Width [m] :
Height [m] :

Remarks:
(1) Tray numbering from top to bottom.
(2) SS = Stainless Steel; CS = Carbon Steel.
(3) Reboiler is E-305; operates with LP steam.

ADSORPTION COLUMN – SPECIFICATION SHEET

EQUIPMENT NUMBER : C-304
NAME : Molecular sieve adsorption column

General Data

Service : - ~~distillation~~/ ~~extraction~~/ adsorption /
Column Type : - packed / ~~tray~~ / ~~spray~~ /
Tray Type : - ~~cap~~/ ~~sieve~~ / ~~valve~~ / none
Tray Number *(1)*
- **Theoretical** : -
- **Actual** : -
- **Feed (actual)** : -

Stage Distance (HETP) [m] : - **Packing Material** : 3Å Zeolite
Column Diameter [m] : 3.2 **Column Material** : CS *(2)*
Column Height [m] : 8.4
Heating : - ~~none~~ / ~~open steam~~ / ~~reboiler~~ / Coils

Process Conditions

Stream Details	**<314> (IN)**		**<325> (IN)**		**<316> (OUT)**		**<320> (OUT)**			
Temp. [°C]	77.5		180		60		200			
Pressure [bara]	1.013		1.013		1.031		1.013			
Density [kg/m^3]	725		1.31		725		1.34			
Mass Flow [kg/s]	28.5		23.1		26.4		25.2			
Mole Flow [kmol/s]	0.69		0.83		0.58		0.95			
Composition	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**	**mol%**	**wt%**
Acetic Acid	trace	trace			trace	trace				
CO_2	0.018	0.016			0.019	0.020				
Ethanol	92.504	82.862			99.890	99.748	0.012	0.007		
Furfural	trace	trace			trace	trace				
HMF	trace	trace			trace	trace				
N_2			99.498	99.222			91.172	86.928		
Water	7.478	17.122	0.502	0.778	0.091	0.232	8.816	13.065		

Column Internals

Trays *Not applicable*

Packing
Type : 3Å Zeolite
Material : Polypropylene
Volume [m^3] : 257
Length [m] : 0.05
Width [m] : 0.05
Height [m] : 0.05

Remarks:
(1) Stage numbering from top to bottom.
(2) SS = Stainless Steel; CS = Carbon Steel.

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EVAPORATOR – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**E-201**	**In Series**	:	1
NAME	:	**First effect juice evaporator**	**In Parallel**	:	2

General Data

Service - ~~Heat Exchanger~~ - Vaporizer - ~~Cooler~~ - ~~Reboiler~~ - ~~Condenser~~

Type :
- ~~Fixed Tube Sheets~~
- ~~Floating Head~~
- ~~Hair Pin~~
- ~~Double Tube~~
- ~~Plate Heat Exchanger~~
- Finned Tubes
- ~~Thermosyphon~~
- Calandria

Position :
- ~~Horizontal~~
- Vertical

Capacity [kW]	:	31,944	
Heat Exchange Area [m^2]	:	1,879	**(Calc.)**
Overall Heat Transfer Coefficient [W/m^2·°C]	:	3,233	**(Approx.)**
Log. Mean Temperature Diff. (LMTD) [°C]	:	14.85	
Passes Tube Side	:	1	
Passes Shell Side	:	1	
Correction Factor LMTD (min. 0.75)	:	1	
Corrected LMTD [°C]	:	14.85	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Stream <114>	Low pressure steam
Mass Stream	**[kg/s]**	:	98.6	14.7
Mass Stream to				
- Evaporate	**[kg/s]**	:	10.1	-
- ~~Condense~~	**[kg/s]**	:	-	14.7
Average Specific Heat	**[kJ/kg·°C]**	:	-	-
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	2,171
Temperature IN	**[°C]**	:	100	130.7
Temperature OUT	**[°C]**	:	125	130.7
Pressure	**[bara]**	:	2.3	2.75
Material		:	CS	CS

Remarks:

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EVAPORATOR – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	E-202	**In Series**	**:**	1
NAME	**:**	Second effect juice evaporator	**In Parallel**	**:**	2

General Data

Service	**:**	**- ~~Heat Exchanger~~**	**- Vaporizer - ~~Cooler~~ - ~~Reboiler~~ - ~~Condenser~~**
Type	**:**	**- ~~Fixed Tube Sheets~~** **- ~~Floating Head~~** **- ~~Hair Pin~~** **- ~~Double Tube~~**	**- ~~Plate Heat Exchanger~~** **- Finned Tubes** **- ~~Thermosyphon~~** **- Calandria**
Position	**:**	**- ~~Horizontal~~** **- Vertical**	

Capacity [kW]	**:**	31,944
Heat Exchange Area [m^2]	**:**	1,348.5 **(Calc.)**
Overall Heat Transfer Coefficient [W/m^2·°C]	**:**	2,683 **(Approx.)**
Log. Mean Temperature Diff. (LMTD) [°C]	**:**	6.4
Passes Tube Side	**:**	1
Passes Shell Side	**:**	1
Correction Factor LMTD (min. 0.75)	**:**	1
Corrected LMTD [°C]	**:**	6.4

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Juice from 1st effect	1st effect vapor
Mass Stream	**[kg/s]**	**:**	88.5	10.1
Mass Stream to				
- Evaporate	**[kg/s]**	**:**	11.07	-
- ~~Condense~~	**[kg/s]**	**:**	-	10.1
Average Specific Heat	**[kJ/kg·°C]**	**:**	-	-
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[°C]**	**:**	118.6	125
Temperature OUT	**[°C]**	**:**	118.6	125
Pressure	**[bara]**	**:**	1.89	2.3
Material		**:**	CS	CS

Remarks:

EVAPORATOR – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-203**	**In Series**	**:**	**1**
NAME	**:**	**Third effect juice evaporator**	**In Parallel**	**:**	**2**

General Data

Service **- ~~Heat Exchanger~~ - Vaporizer - ~~Cooler~~ - ~~Reboiler~~ - ~~Condenser~~**

Type :
- **~~Fixed Tube Sheets~~**
- **~~Floating Head~~**
- **~~Hair Pin~~**
- **~~Double Tube~~**
- **~~Plate Heat Exchanger~~**
- **Finned Tubes**
- **~~Thermosyphon~~**
- **Calandria**

Position :
- **~~Horizontal~~**
- **Vertical**

Capacity [kW]	**:**	31,944
Heat Exchange Area [m^2]	**:**	1,561.5 **(Calc.)**
Overall Heat Transfer Coefficient [W/m^2·^{o}C]	**:**	2,120 **(Approx.)**
Log. Mean Temperature Diff. (LMTD) [^{o}C]	**:**	7.9
Passes Tube Side	**:**	1
Passes Shell Side	**:**	1
Correction Factor LMTD (min. 0.75)	**:**	1
Corrected LMTD [^{o}C]	**:**	7.9

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Juice from 2nd effect	2nd effect vapor
Mass Stream	**[kg/s]**	**:**	77.5	11.07
Mass Stream to				
- Evaporate	**[kg/s]**	**:**	12.0	-
- ~~Condense~~	**[kg/s]**	**:**	-	11.07
Average Specific Heat	**[kJ/kg·^{o}C]**	**:**	-	-
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^{o}C]**	**:**	110.7	118.6
Temperature OUT	**[^{o}C]**	**:**	110.7	118.6
Pressure	**[bara]**	**:**	1.46	1.89
Material		**:**	CS	CS

Remarks:

EVAPORATOR – SPECIFICATION SHEET

EQUIPMENT NUMBER : **E-204** **In Series** : **1**
NAME : **Fourth effect juice evaporator** **In Parallel** : **2**

General Data

Service - ~~Heat Exchanger~~ - Vaporizer - ~~Cooler~~ - ~~Reboiler~~ - ~~Condenser~~

Type :
- ~~Fixed Tube Sheets~~
- ~~Floating Head~~
- ~~Hair Pin~~
- ~~Double Tube~~
- ~~Plate Heat Exchanger~~
- Finned Tubes
- ~~Thermosyphon~~
- Calandria

Position :
- ~~Horizontal~~
- Vertical

Capacity [kW]	:	31,944
Heat Exchange Area [m^2]	:	1,836.5 **(Calc.)**
Overall Heat Transfer Coefficient [W/m^2·^{o}C]	:	1,549 **(Approx.)**
Log. Mean Temperature Diff. (LMTD) [^{o}C]	:	10.2
Passes Tube Side	:	1
Passes Shell Side	:	1
Correction Factor LMTD (min. 0.75)	:	1
Corrected LMTD [^{o}C]	:	10.2

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Juice from 3rd effect	3rd effect vapour
Mass Stream	**[kg/s]**	:	65.47	12.0
Mass Stream to				
- Evaporate	**[kg/s]**	:	12.95	-
- ~~Condense~~	**[kg/s]**	:	-	12.0
Average Specific Heat	**[kJ/kg·^{o}C]**	:	-	-
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^{o}C]**	:	100.5	110.7
Temperature OUT	**[^{o}C]**	:	100.5	110.7
Pressure	**[bara]**	:	1.028	1.46
Material		:	CS	CS

Remarks:

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EVAPORATOR – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**E-205**	**In Series**	:	**1**
NAME	:	**Fourth effect juice evaporator**	**In Parallel**	:	**2**

General Data

Service - ~~**Heat Exchanger**~~ - **Vaporizer** - ~~**Cooler**~~ - ~~**Reboiler**~~ - ~~**Condenser**~~

Type :
- ~~**Fixed Tube Sheets**~~
- ~~**Floating Head**~~
- ~~**Hair Pin**~~
- ~~**Double Tube**~~
- ~~**Plate Heat Exchanger**~~
- **Finned Tubes**
- ~~**Thermosyphon**~~
- **Calandria**

Position :
- ~~**Horizontal**~~
- **Vertical**

Capacity [kW]	:	31,944
Heat Exchange Area [m^2]	:	2,229 **(Calc.)**
Overall Heat Transfer Coefficient [W/m^2·^{o}C]	:	979 **(Approx.)**
Log. Mean Temperature Diff. (LMTD)[^{o}C]	:	14.5
Passes Tube Side	:	1
Passes Shell Side	:	1
Correction Factor LMTD (min. 0.75)	:	1
Corrected LMTD [^{o}C]	:	14.5

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Juice from 4th effect	4th effect vapour
Mass Stream	**[kg/s]**	:	52.53	12.95
Mass Stream to				
- Evaporate	**[kg/s]**	:	13.94	-
- ~~Condense~~	**[kg/s]**	:	-	12.95
Average Specific Heat	**[kJ/kg·^{o}C]**	:	-	-
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^{o}C]**	:	85.98	100.5
Temperature OUT	**[^{o}C]**	:	85.98	100.5
Pressure	**[bara]**	:	0.59	1.028
Material		:	CS	CS

Remarks:

Çağrı Efe	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse
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EVAPORATOR – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-206**	**In Series**	**:**	**1**
NAME	**:**	**Sixth effect juice evaporator**	**In Parallel**	**:**	**2**

General Data

Service - ~~**Heat Exchanger**~~ - **Vaporizer** - ~~**Cooler**~~ - ~~**Reboiler**~~ - ~~**Condenser**~~

Type	**:**	- ~~**Fixed Tube Sheets**~~	- ~~**Plate Heat Exchanger**~~
		- ~~**Floating Head**~~	- **Finned Tubes**
		- ~~**Hair Pin**~~	- ~~**Thermosyphon**~~
		- ~~**Double Tube**~~	- **Calandria**

Position	**:**	- ~~**Horizontal**~~
		- **Vertical**

Capacity [kW]	**:**	31,944	
Heat Exchange Area [m^2]	**:**	2,991	**(Calc.)**
Overall Heat Transfer Coefficient [W/m^2·°C]	**:**	409	**(Approx.)**
Log. Mean Temperature Diff. (LMTD) [°C]	**:**	29.7	
Passes Tube Side	**:**	1	
Passes Shell Side	**:**	1	
Correction Factor LMTD (min. 0.75)	**:**	1	
Corrected LMTD [°C]	**:**	29.7	

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Juice from 5th effect	5th effect vapour
Mass Stream	**[kg/s]**	**:**	38.6	13.94
Mass Stream to				
- Evaporate	**[kg/s]**	**:**	15.1	-
- ~~Condense~~	**[kg/s]**	**:**	-	13.94
Average Specific Heat	**[kJ/kg·°C]**	**:**	-	-
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[°C]**	**:**	56.3	85.98
Temperature OUT	**[°C]**	**:**	56.3	85.98
Pressure	**[bara]**	**:**	0.17	0.59
Material		**:**	CS	CS

Remarks:

Çağrı Efe	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse
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VACUUM PAN – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**E-207**	**In Series**	:	**1**
NAME	:	**First vacuum pan**	**In Parallel**	:	**-**

General Data

Service		- ~~**Heat Exchanger**~~ - **Vaporizer** - ~~**Cooler**~~ - ~~**Reboiler**~~ - ~~**Condenser**~~
Type	:	- ~~**Fixed Tube Sheets**~~ - ~~**Floating Head**~~ - ~~**Hair Pin**~~ - ~~**Double Tube**~~ - ~~**Plate Heat Exchanger**~~ - **Finned Tubes** - ~~**Thermosyphon**~~ - **Calandria**
Position	:	- ~~**Horizontal**~~ - **Vertical**

Capacity [kW]	:	16,276
Heat Exchange Area [m^2]	:	1,566.5 **(Calc.)**
Overall Heat Transfer Coefficient [W/m^2·^{o}C]	:	-
Log. Mean Temperature Diff. (LMTD) [^{o}C]	:	55.7
Passes Tube Side	:	1
Passes Shell Side	:	1
Correction Factor LMTD (min. 0.75)	:	1
Corrected LMTD [^{o}C]	:	55.7

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Stream <202>	Low pressure steam
Mass Stream	**[kg/s]**	:	46.3	7.5
Mass Stream to				
- Evaporate	**[kg/s]**	:	7.0	-
- ~~**Condense**~~	**[kg/s]**	:	-	7.5
Average Specific Heat	**[kJ/kg·^{o}C]**	:	-	-
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^{o}C]**	:	75	130.7
Temperature OUT	**[^{o}C]**	:	75	130.7
Pressure	**[bara]**	:	0.38	2.75
Material		:	CS	CS

Remarks:

Çağrı Efe	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse
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VACUUM PAN – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**E-208**	**In Series**	:	**1**
NAME	:	**Second vacuum pan**	**In Parallel**	:	**-**

General Data

Service - ~~**Heat Exchanger**~~ - **Vaporizer** - ~~**Cooler**~~ - ~~**Reboiler**~~ - ~~**Condenser**~~

Type	:	- ~~**Fixed Tube Sheets**~~	- ~~**Plate Heat Exchanger**~~
		- ~~**Floating Head**~~	- **Finned Tubes**
		- ~~**Hair Pin**~~	- ~~**Thermosyphon**~~
		- ~~**Double Tube**~~	- **Calandria**

Position : - ~~**Horizontal**~~
- **Vertical**

Capacity [kW]	:	9,286
Heat Exchange Area [m^2]	:	891.5 **(Calc.)**
Overall Heat Transfer Coefficient [W/m^2·^{o}C]	:	-
Log. Mean Temperature Diff. (LMTD) [^{o}C]	:	55.7
Passes Tube Side	:	1
Passes Shell Side	:	1
Correction Factor LMTD (min. 0.75)	:	1
Corrected LMTD [^{o}C]	:	55.7

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Stream <214>	Low pressure steam
Mass Stream	**[kg/s]**	:	21.5	4.3
Mass Stream to				
- Evaporate	**[kg/s]**	:	4	-
- ~~Condense~~	**[kg/s]**	:	-	4.3
Average Specific Heat	**[kJ/kg·^{o}C]**	:	-	-
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^{o}C]**	:	75	130.7
Temperature OUT	**[^{o}C]**	:	75	130.7
Pressure	**[bara]**	:	0.38	2.75
Material		:	CS	CS

Remarks:

Çağrı Efe	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse
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REBOILER – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**E-303**	**In Series**	:	**1**
NAME	:	**Beer column reboiler**	**In Parallel**	:	**-**

General Data

Service - ~~Heat Exchanger~~ - ~~Vaporizer~~ - ~~Cooler~~ - **Reboiler** - ~~Condenser~~

Type :
- ~~Fixed Tube Sheets~~
- ~~Floating Head~~
- ~~Hair Pin~~
- ~~Double Tube~~
- ~~Plate Heat Exchanger~~
- ~~Finned Tubes~~
- ~~Thermosyphon~~
- **Kettle**

Position :
- **Horizontal**
- ~~Vertical~~

Capacity [kW]	:	165,553
Heat Exchange Area [m^2]	:	4,168 **(Calc.)**
Overall Heat Transfer Coefficient [W/m^2·^{o}C]	:	1,281(Calc.)
Log. Mean Temperature Diff. (LMTD) [^{o}C]	:	30.7
Passes Tube Side	:	**-**
Passes Shell Side	:	-
Correction Factor LMTD (min. 0.75)	:	1
Corrected LMTD [^{o}C]	:	30.7

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Low pressure steam	Stream <311>
Mass Stream	**[kg/s]**	:	76.25	73.45
Mass Stream to				
- Evaporate	**[kg/s]**	:	-	73.45
- ~~Condense~~	**[kg/s]**	:	76.25	-
Average Specific Heat	**[kJ/kg·^{o}C]**	:	-	-
Heat of Evap. / Condensation	**[kJ/kg]**	:	2,171	-
Temperature IN	**[^{o}C]**	:	130.7	100
Temperature OUT	**[^{o}C]**	:	130.7	100
Pressure	**[bara]**	:	2.75	1.013
Material		:	CS	SS304

Remarks:

Çağrı Efe	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse
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REBOILER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-305**	**In Series**	**:**	**1**
NAME	**:**	**Rectification column reboiler**	**In Parallel**	**:**	**-**

General Data

Service - ~~**Heat Exchanger**~~ - ~~**Vaporizer**~~ - ~~**Cooler**~~ - **Reboiler** - ~~**Condenser**~~

Type :
- ~~**Fixed Tube Sheets**~~
- ~~**Floating Head**~~
- ~~**Hair Pin**~~
- ~~**Double Tube**~~
- ~~**Plate Heat Exchanger**~~
- ~~**Finned Tubes**~~
- ~~**Thermosyphon**~~
- **Kettle**

Position :
- **Horizontal**
- ~~**Vertical**~~

Capacity [kW]	**:**	25,729
Heat Exchange Area [m^2]	**:**	653.4 **(Calc.)**
Overall Heat Transfer Coefficient [W/m^2·°C]	**:**	1,281 **(Calc.)**
Log. Mean Temperature Diff. (LMTD) [°C]	**:**	30.7
Passes Tube Side	**:**	-
Passes Shell Side	**:**	-
Correction Factor LMTD (min. 0.75)	**:**	1
Corrected LMTD [°C]	**:**	30.7

Process Conditions

			Shell Side	**Tube Side**
Medium		**:**	Low pressure steam	Stream <311>
Mass Stream	**[kg/s]**	**:**	12.81	12.07
Mass Stream to				
- Evaporate	**[kg/s]**	**:**	-	12.07
- ~~Condense~~	**[kg/s]**	**:**	12.81	-
Average Specific Heat	**[kJ/kg·°C]**	**:**	-	-
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	2,171	-
Temperature IN	**[°C]**	**:**	130.7	100
Temperature OUT	**[°C]**	**:**	130.7	100
Pressure	**[bara]**	**:**	2.75	1.013
Material		**:**	CS	SS304

Remarks:

CONDENSER – SPECIFICATION SHEET

EQUIPMENT NUMBER	**:**	**E-304**	**In Series**	**:**	1
NAME	**:**	**Beer column condenser**	**In Parallel**	**:**	-

General Data

Service		**- ~~Heat Exchanger~~ - ~~Vaporizer~~ - ~~Cooler~~ - ~~Reboiler~~ - Condenser**
Type	**:**	**- ~~Fixed Tube Sheets~~ - ~~Plate Heat Exchanger~~** **- ~~Floating Head~~ - ~~Finned Tubes~~** **- ~~Hair Pin~~ - ~~Thermosyphon~~** **- ~~Double Tube~~ - Kettle**
Position	**:**	**- Horizontal** **- ~~Vertical~~**

Capacity [kW]	**:**	36,738	
Heat Exchange Area [m^2]	**:**	1,502	**(Calc.)**
Overall Heat Transfer Coefficient [W/m^2·^{o}C]	**:**	1,229	**(Calc.)**
Log. Mean Temperature Diff. (LMTD) [^{o}C]	**:**	19.9	
Passes Tube Side	**:**	-	
Passes Shell Side	**:**	-	
Correction Factor LMTD (min. 0.75)	**:**	1	
Corrected LMTD [^{o}C]	**:**	19.9	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Chilled water	Stream <311>
Mass Stream	**[kg/s]**	**:**	293.4	26.61
Mass Stream to				
- Evaporate	**[kg/s]**	**:**	-	-
- ~~Condense~~	**[kg/s]**	**:**	-	26.61
Average Specific Heat	**[kJ/kg·^{o}C]**	**:**	4.179	-
Heat of Evap. / Condensation	**[kJ/kg]**	**:**	-	-
Temperature IN	**[^{o}C]**	**:**	10	76.3
Temperature OUT	**[^{o}C]**	**:**	40	20
Pressure	**[bara]**	**:**	1.013	1.013
Material		**:**	CS	SS304

Remarks:

Çağrı Efe	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse
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CONDENSER – SPECIFICATION SHEET

EQUIPMENT NUMBER	: **E-306**		**In Series**	: 1
NAME	: **Rectification column condenser**		**In Parallel**	: -

General Data

Service - ~~**Heat Exchanger**~~ - ~~**Vaporizer**~~ - ~~**Cooler**~~ - ~~**Reboiler**~~ - **Condenser**

Type :
- ~~**Fixed Tube Sheets**~~
- ~~**Floating Head**~~
- ~~**Hair Pin**~~
- ~~**Double Tube**~~
- ~~**Plate Heat Exchanger**~~
- ~~**Finned Tubes**~~
- ~~**Thermosyphon**~~
- **Kettle**

Position :
- **Horizontal**
- ~~**Vertical**~~

Capacity [kW] : 135,558
Heat Exchange Area [m^2] : 4,951.4 **(Calc.)**
Overall Heat Transfer Coefficient [W/m^2·^{o}C] : 845 **(Calc.)**
Log. Mean Temperature Diff. (LMTD) [^{o}C] : 32.4

Passes Tube Side : -
Passes Shell Side : -

Correction Factor LMTD (min. 0.75) : 1
Corrected LMTD [^{o}C] : 32.4

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Chilled water	Stream <311>
Mass Stream	**[kg/s]**	:	930.1	113.6
Mass Stream to				
- Evaporate	**[kg/s]**	:	-	-
- ~~Condense~~	**[kg/s]**	:	-	113.6
Average Specific Heat	**[kJ/kg·^{o}C]**	:	4.184	-
Heat of Evap. / Condensation	**[kJ/kg]**	:	-	-
Temperature IN	**[^{o}C]**	:	25	78.2
Temperature OUT	**[^{o}C]**	:	60	77.5
Pressure	**[bara]**	:	1.013	1.013
Material		:	CS	SS304

Remarks:

Çağrı Efe	Technical and Economical Feasibility of Production of Ethanol from Sugar Cane and Sugar Cane Bagasse
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