POLITECNICO DI MILANO

Scoula di Ingegneria Industriale e Informazione

Master of Science in Chemical Engineering

Department of Chemistry, Materials and Chemical Engineering "Giulio Natta"

Bioethanol Production from Fruit Waste: Potential Development in Mexico

Supervisor: Prof. Giulia Luisa Bozzano

Master thesis by:

Ricardo Hugo Blanco Mendoza

Student ID: 904599

Academic year 2020/2021

Contents

List of Figures

List of Tables

Abstract

The growing demand for energy in transportation, heating, and industrial processes sectors, together with the increase of the emission to the environment due to the current linear economy, bad management, and lack of disposals of residues are becoming one of the biggest challenges around the world. Therefore, the interest for reused waste has increased considerably worldwide. The so-called biorefinery of waste is an attractive technology in the countries with a huge amount of organic residues to produce biofuels and value-added chemicals. (Cherubini 2010; Hahn-Hägerdal et al. 2006).

Fruit production is an outstanding activity in Mexico. Hence, the aim of this M.Sc. thesis is to analyze the potential of Mexico in the production of bioethanol from fruit waste throughout an economic and technical assessment of a bioethanol plant. The conceptual design of the bioethanol plant is based on the processing of citrus waste (orange, lime, mandarin), banana and mango waste, and the proposed technology is already used at the industrial level.

The technologies for the pretreatment stage are Grinding process and Steam Explosion. From the top of the Steam Explosion the vapor is sent to the decantation process to separate the limonene from the water, and from the bottom of the Steam Explosion the liquid part is sent to the Belt filter press to separate the solid content from the liquid. Then, the liquid is sent to the Simultaneous Saccharification and Fermentation process to produce the bioethanol. Finally, the bioethanol is sent to the Bioethanol Purification stage where there are Beer and Rectification Columns, and the Ethanol Dehydration (Molecular Sieves) process to obtain an almost pure bioethanol.

To evaluate the feasibility of this Bioethanol plant, 3 scenarios are analysis, Worst-Case, Standard-Case and Best-Case scenarios. The Worst-Case scenario has a

recover of the investment in the year 12 and an Internal Rate of Return of 11.7%. For the Standard-Case scenario, the return of the investment is around the year 8 and an Internal Rate of Return of 17.5%. While the Best-Case scenario has a return of the investment in the year 5 and an Internal Rate of Return of 27.1%.

The increment of final price and the annual production of the limonene increases the Net Income of the plant. Thus, the final price of the bioethanol can be decreased to be competitive with the market price of the ethanol and MTBE. If all the fruit waste used in the industry in Mexico is processed, there is a potential production of 88,750,000 L/y of Bioethanol and 11,398,000L/y of Limonene based on the Standard-Case scenario.

Sommario

La crescente domanda di energia nei settori dei trasporti, del riscaldamento e dei processi industriali, insieme all'aumento delle emissioni nell'ambiente che risultano dall'attuale economia lineare, la cattiva gestione e la mancanza di smaltimento dei rifiuti, stanno diventando una delle maggiori sfide in tutto il mondo. Pertanto, l'interesse per il riutilizzo dei rifiuti è aumentato considerevolmente in tutto il mondo. La cosiddetta bioraffinazione dei rifiuti è una tecnologia interessante nei paesi con una grande quantità di rifiuti organici per produrre biocarburanti e prodotti chimici a valore aggiunto. (Cherubini 2010; Hahn-Hägerdal et al. 2006).

La produzione di frutta è un'attività importante in Messico. Quindi, l'obiettivo di questa tesi di Laurea Magistrale consiste nell'analizzare le potenzialità del Messico nella produzione di bioetanolo dai residui della frutta, attraverso una valutazione economica e tecnica di un impianto di bioetanolo. Il progetto concettuale dell'impianto a bioetanolo si basa sulla lavorazione dei residui di agrumi (arancia, lime, mandarino), banana e mango; inoltre, la tecnologia proposta è già utilizzata a livello industriale.

Le tecnologie per la fase di pretrattamento sono *Grinding process* e *Steam Explosion*. L'effluente gassoso della *Steam Explosion* viene inviato al processo di decantazione per separare il limonene dall'acqua, e l'effluente liquido viene inviato ad un filtro a nastro per separare il contenuto solido.in In seguito, il liquido viene inviato al processo di saccarificazione e fermentazione simultanea per produrre il bioetanolo. Finalmente, il bioetanolo viene inviato alla fase di purificazione dove sono presenti le *Beer and Rectification Columns* e il processo di disidratazione dell'etanolo (setacci molecolari) per ottenere un bioetanolo quasi puro.

Per valutare la fattibilità di questo impianto a Bioetanolo, sono 3 gli scenari di analisi, scenari *Worst-Case*, *Standard-Case* e *Best-Case*. Lo scenario peggiore prevede un recupero dell'investimento nell'anno 12 e un tasso di rendimento interno dell'11,7%. Per lo scenario *Standard-Case*, il ritorno dell'investimento si raggiunge dopo circa 8 anni e ha un tasso di rendimento interno del 17,5%. Mentre lo scenario Best-Case ha un ritorno dell'investimento a 5 anni e un tasso di rendimento interno del 27,1.

L'incremento del prezzo finale e la produzione annuale del limonene aumentano l'utile netto della pianta. Pertanto, il prezzo finale del bioetanolo può essere ridotto per essere competitivo con il prezzo di mercato dell'etanolo e dell'MTBE. Se tutti i rifiuti della frutta utilizzati nell'industria in Messico venissero trattati, si potrebbe avere una produzione potenziale di 88.750.000 L/anno di bioetanolo e 11.398.000 L/anno di limonene in base allo scenario Standard.

1. State of the Art

The growing demand for energy in transportation, heating and industrial processes sectors, is becoming one of the biggest challenges around the world due to the huge amount of energy required (Hahn-Hägerdal et al. 2006). The transportation sector itself represents around the 65% of the total oil consumption (IEA 2017), this percentage has increased rapidly during the last 2 decades together with the pollution emitted to the environment (Aditiya et al. 2016).

Additionally, the emissions to the environment have increased considerably due to the current linear economy ("take, make and dispose"), a bad management, and lack of disposals of residues that generate a huge amount of solid and food supply chain waste (Maina, Kachrimanidou, and Koutinas 2017). As a possible solution for the actual linear economy, two concepts are introduced: circular economy and bioeconomy. The circular economy principle aims for "the elimination of waste" by extraction and recovery of the high value-added compounds (Ellen MacArthur Foundation (EMAF) 2013), while the bioeconomy approach searches for "the sustainable production and conversion of biomass, for a range of food, health, fiber and industrial products and energy" (Albrecht et al. 2010).

The interest of reused waste has increased considerably since it is possible to produce fuels and value-added chemicals. Currently worldwide, the most attractive and exploited technologies to perform that are the so-called biorefinery technologies. These are a variety of processes to convert the organic waste with different composition and type into biofuels, power, and chemicals (Cherubini et al. 2010). The potential of waste refining is mainly the reduction of greenhouse gas emissions and the dependence on non-renewable resources to obtain the same products (Lynd et al. 2005).

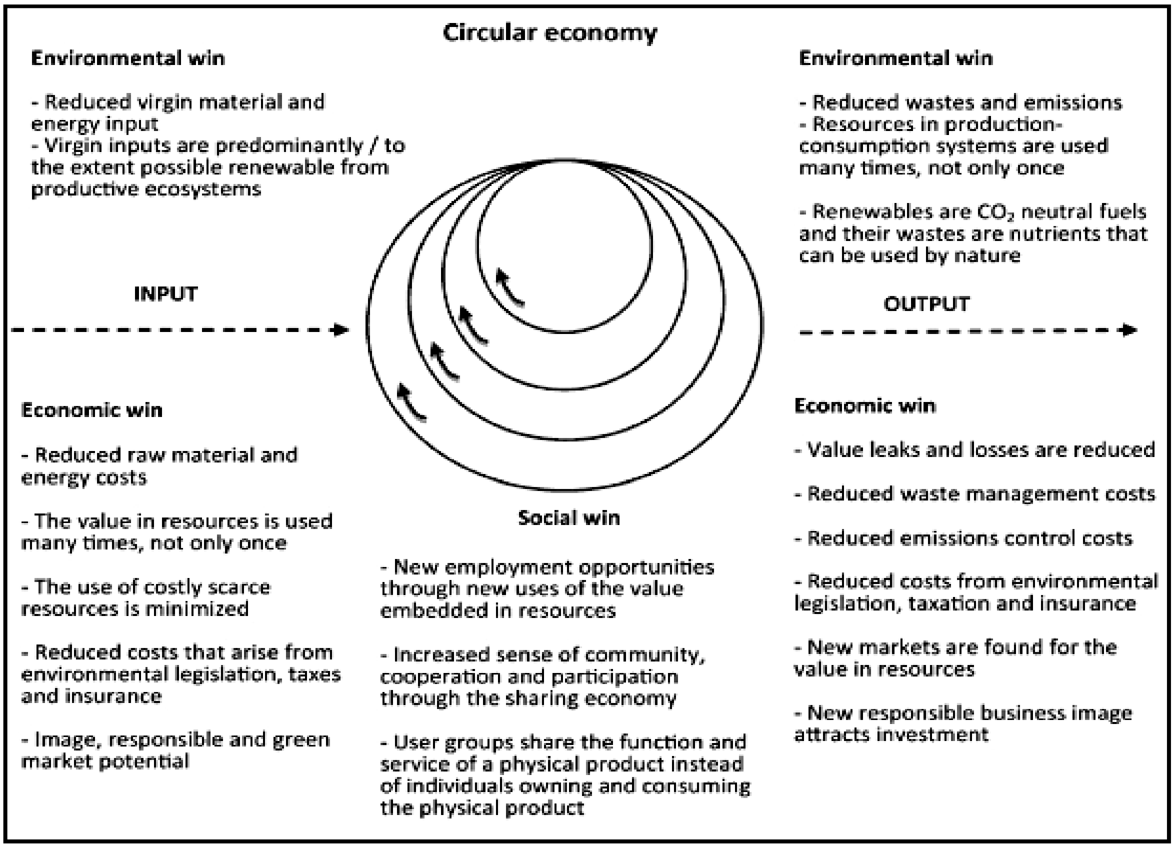

Figure 1. Circular economy for sustainable development. The win-win-win potential of circular economy (Korhonen, Honkasalo, and Seppälä 2018).

Therefore, in order to reduce the emissions generated by fossil fuels a transition to biofuels is the primary necessity worldwide. The biofuel production from lignocellulosic biomass (known as second generation biofuels) seems to be the most promising solution for this transition (Anuj K. Chandel et al. 2019). There are two well-known products, the biodiesel which is used to substitute the fossil diesel, and the bioethanol which can be used as a substitute, or be blended with, gasoline. Additionally, compared to the first-generation biofuels, the lignocellulosic biomass is cheaper, more abundant, and renewable and specially, does not compete directly with the food production. Thus, the reduction of waste is possible (Haghighi Mood et al. 2013).

The fruit commercialization and processing do not only represent food sources for the planet, but at the same time, cause environmental problems due to the fact that the residues constitute about 50% of the total fruit production (peels, seeds, and pulps), and the lack of the residues disposals worldwide increases the problem (Plazzotta, Manzocco, and Nicoli 2017; Sánchez Orozco et al. 2014). Therefore, different alternatives and processes for the transformation of fruit waste into valuable products have been proposed in the literature based on the concepts of circular economy and bioeconomy.

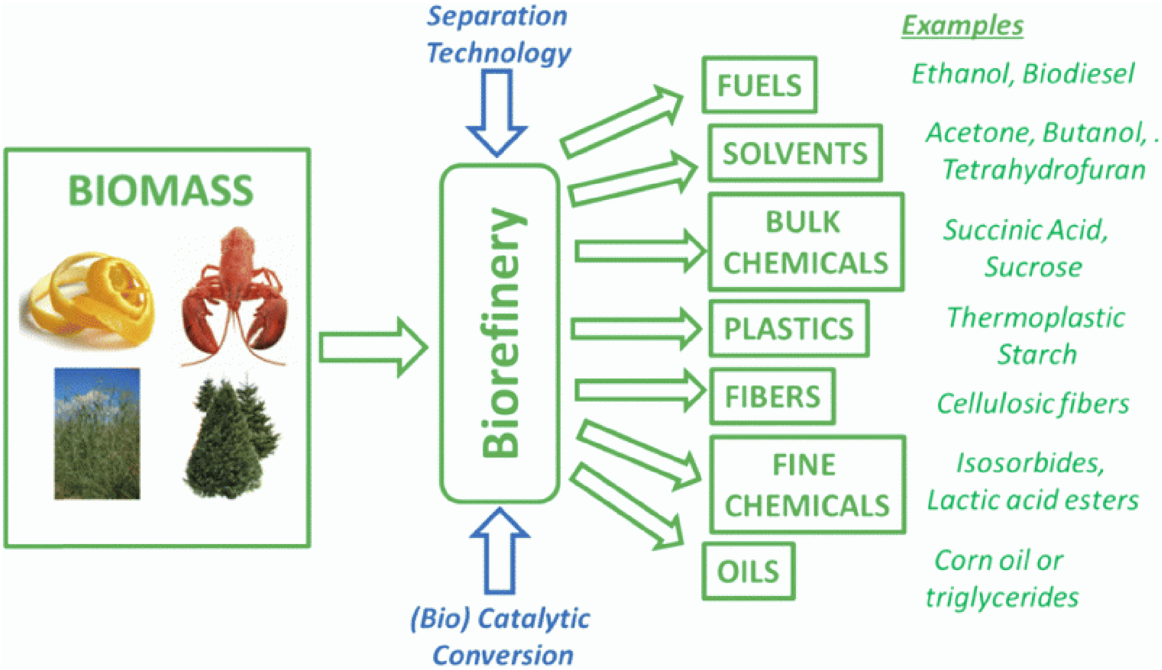

Figure 2. Biorefinery scheme (Nizami et al. 2017).

Zhou et al. (2008) developed a new process to produce bioethanol from orange peel waste (OPW), improving the recovery of D-limonene. The pretreatment used was steam explosion to separate the D-limonene from the OPW. Then, the OPW pretreated was cooled to 38°C for a subsequent simultaneous saccharification and fermentation (SSF). The bioethanol production based on 1ton fresh OPW was 41.6L, 7.33L of limonene and 3.4L of methanol.

Boluda-Aguilar et al. (2010) and Boluda-Aguilar (2013) studied the bioethanol production from mandarin peel wastes (MPW) and lemon peel waste (LPW). In both cases, the pretreatment was steam explosion to recover the D-limonene and reduce its concentration >0.05% v/v. Then, the citrus waste was cooled down for SSF. The bioethanol production for both cases was around 60L/1ton of fresh citrus waste, 6l of D-limonene and 36kg of galacturonic acid.

Pourbafrani et al. (2010) studied a new process of bioethanol production from OPW. The OPW was pretreated with dilute-acid explosion process to hydrolyze it and reduce the concentration of the D-limonene. The resultant slurry was then centrifuged, and the liquid part was fermented to bioethanol. The bioethanol production was 40L per ton of OPW and 8.9L D-limonene.

In 2010, a Spanish company call Citritecno S.L. started to operate a plant with a capacity of 25 ton/h of orange residues. They reported problems in the D-limonene recovery. The process used to recover the D-limonene was pressing but the efficiency of this system was low, and the essential oils affected the fermentation of the orange residues. Therefore, they produced low amount of bioethanol. Additionally, they could not produce more bioethanol due to the lack of supply, the main supplier stopped providing the orange residues (Citrotecno 2013).

Apart from the citrus waste, there are some studies using other fruits to produce bioethanol. For example, Kounbesiou et al. (2011) studied the use of mango residues for the production bioethanol. The pretreatment of the mango residues was hydrolysis at 55°C using *Bacillus licheniformis* for 5h and a pH of 4.5. Then, SSF process was carried out to final obtain a maximum bioethanol production of 21.75 g/L.

Sharma et al. (2007) studied the bioethanol production using kinnow waste and banana peels. The kinnow waste was previously pretreated by steam explosion to remove the essential oils. Then, both waste were mixed in a ratio of 4:6 (kinnow waste:banana peel) for the SSF. The maximum bioethanol production was 26.84g/L.

Oberoi et al. (2011) evaluate the bioethanol production from banana peels (BP). The BP were subjected to autoclave-sterilization before the SSF process. The bioethanol production was 28.2g/L. Gebregergs et al. (2016) also evaluate the bioethanol production from BP. The pretreatment was acid hydrolysis, and they concluded that higher levels of acid concentration and time, the production of bioethanol decreased. After the acid hydrolysis, the SSF process was carried out, obtaining a maximum yield of 45% ethanol.

Additionally, the compositional analysis of the orange peel, orange bagasse, banana peel, and mango peel is shown in table 1 (Sánchez Orozco et al. 2014), and the information found about the bioethanol production from fruit waste and the main processes is presented in table 2.

Table 1. Compositional analysis of fruit waste on dry matter basis.

Component	Fruit waste, wt%			
	Orange bagasse	Orange peel	Banana peel	Mango peel
Hemicellulose	26.45	14.46	25.52	14.51
Cellulose	9.93	11.93	11.45	9.19
Lignin	2.81	2.17	9.82	4.25
Protein	7.59	5.97	5.62	7.00

Source: Sánchez Orozco et al. 2014

All of this information shows that it is possible to produce value-added products from fruit waste. The drawback is the necessity to have a considerable amount of fruit waste in order to have a sustainable production. The high potential to develop this technology in an industrial level is in those countries with high production of fruits (Sánchez Orozco et al. 2014; W. Zhou, Widmer, and Grohmann 2007). Therefore, Mexico could have an enormuos potential.

Fruit production is an outstanding activity in Mexico. This is highlighted by the places Mexico occupied in the worldwide ranking of fruit producers. For example, Mexico occupied the first place in the avocado production, the second place in lime production, and the fifth place in the mango and orange production. The annual production of orange, lime, banana and mango are approximately 4.6, 2.42, 2.23 and 1.89 million tons respectively and are the most important fruit products in Mexico (SAGARPA 2017).

Hence, the aim of this M.Sc. thesis is to analyze the potential of Mexico in the production of bioethanol from fruit waste throughout an economic and technical assessment of a bioethanol plant. The conceptual design of the bioethanol plant is based on the processing of citrus waste (orange, lime, mandarin), banana and mango waste, and the proposed technology is already used at the industrial level.

There is not information about the conceptual design and economic evaluation using this fruit waste (citrus, banana and mango waste), neither corresponding to the potential ethanol production in Mexico using it. Therefore, this master thesis evaluates a new innovative approach for bioethanol production and value-added byproducts.

Table 2. Production of bioethanol by fruit processing waste.

Substrate	Pretreatment	Fermentation	Bioethanol production	References
Orange peel waste	Milling and steam explosion	SSF process. The enzymes were cellulase, pectinase, glucosidase and saccharomyces cerevisiae.	The maximum yield was 4.05% w/v in 18h, producing 41.6L of bioethanol and 3.4L of methanol.	(Weiyang Zhou, Widmer, and Grohmann 2008)
Mandarin peel waste	Milling and steam explosion	SSF process. The enzymes were cellulase, pectinase, glucosidase and saccharomyces cerevisiae.	The bioethanol production was 59.26L and 6L of limonene per ton of wet mandarin peel waste.	(Boluda-Aguilar et al. 2010a)
Orange waste	Milling and acid hydrolysis	The enzyme for the fermentation process was saccharomyces cerevisiae.	The bioethanol production was 39.64L of bioethanol and 8.9L of limonene per ton of wet orange waste.	(Pourbafrani et al. 2010)
Lemon peel waste	Milling and steam explosion.	SSF process. The enzymes were cellulase, pectinase, glucosidase and saccharomyces cerevisiae.	The bioethanol production was 63L, 36kg of galacturonic acid and 6L of limonene per ton of wet mandarin peel waste.	(Boluda-Aguilar and López-Gómez 2013)

Banana peel and kinnow waste	Milling and dried. Steam explosion was used for the extraction of essential oils of kinnow waste.	The fruit waste was mixed in a ratio 4:6 (kinnow waste:banana peel) and fermented with saccharomyces cerevisiae.	The maximum bioethanol production was 23.79g/L	(Sharma et al. 2007)
Banana peel	Milling, acid hydrolysis and autoclaved	SSF process. The enzymes were cellulase, pectinase, and saccharomyces cerevisiae.	The ethanol production was 28.2 g/L	(Oberoi et al. 2011)
Banana peel	Milling and dried. Acid hydrolysis	The enzyme for the fermentation process was saccharomyces cerevisiae.	The bioethanol yield was 45.1%	(Gebregergs, Gebresemati, and Sahu 2016)
Banana Peel	Milling, dried and acid hydrolysis	Glucose-yeast broth	The bioethanol production was 157.07 mg/ml	(Hamzah, Alias, and Ahmad 2019)
Mango residues	Milling. Hydrolysis of mango by Bacillus licheniformis.	Saccharomyces cerevisiae enzymes used for the fermentation process	The bioethanol production was 21.75g/L	(Kounbesiou et al. 2011)

Mango and banana waste	Milling and acid hydrolysis.	The enzyme for the fermentation process was saccharomyces cerevisiae.	The bioethanol yield was 35.86% for mixed fruit, 28.45% banana pulp and 26.5% mango pulp, 13.84% for banana peel and 9.68% for mango peel.	(Arumugam and Manikandan 2011)

SSF: simultaneous saccharification and fermentatio

2. Valorization of Fruit Waste

The definition of waste management is "the collection, transport, recovery, and disposal of waste, including the supervision of such operations" and the waste management system consists of "the whole set of activities related to handling, disposing or recycling waste materials". The strategies of the waste management can be classified with respect to the final disposition of waste and ordered according to their priority: minimization and prevention of waste generation, recycling and reuse, energy recovery and landfilling (Plazzotta, Manzocco, and Nicoli 2017).

Figure 3. Fruit waste in municipal dumps.

The fruit waste generation depends on the fruit processing industries and their fruit processing technology (e.g., drying/dehydration, pressing), that generates solids (e.g., pomace, pulp, peels, seeds and stems), or liquids (e.g., juices, wash water and cleaning chemicals). These by-products are highly fermentable and perishable because of moisture (80-90%), soluble sugars (6-64%) and crude protein (10-24%) contents. They can also retain bioactive compounds, for instance, phenolic compounds, dietary fiber, essential oils, enzymes and organic acids, with a high commercial value in the food, pharmaceutical and biofuel industries (Mirabella, Castellani, and Sala 2014).

2.1. Lignocellulosic biomass pretreatment

Recently, the interest in lignocellulosic biomass has increased as an attractive feedstock to produce biofuels and other chemicals due to the low price and high availability. However, the resistance of the plant cell wall to deconstruction makes their utilization a challenge. This is because the hydrolyzation of the cellulose present in the lignocellulosic biomass has low effectiveness, and less than 20% of yields are normally obtained. Factors that contribute considerably to the low effectiveness of this process are the cellulose crystallinity, the protection exerted by hemicellulose and lignin fraction, and the low accessibility to cellulose. Thus, a pretreatment step is required to overcome these factors (Mussatto 2016).

The pretreatment plays a crucial role for the transformation of the lignocellulosic biomass into valuable products. In general, this process allows the breaking of the cell wall physical barriers, removing the lignin and hemicellulose fractions, reducing the cellulose crystallinity, swelling pores in the biomass structure, and increasing the accessible surface area. Therefore, the cellulose fibers are more accessible for the enzymes since the enzymatic hydrolysis is limited by the pore size, so the increase in the material porosity can considerably improve the hydrolysis yield. The lignin acts as a physical barrier, so the removal of the lignin is also a crucial step because it prevents the hydrolyzation of the digestible parts (Mussatto 2016; Srivastava et al. 2015).

Different pretreatment alternatives have been proposed to change the physical and chemical structure of the lignocellulosic biomass in order to avoid the sugar degradation by efficiently separating hemicellulose and lignin fractions. They can be classified into four categories: physical, chemical, physicochemical, and biological. For example, mechanical pretreatment, pyrolysis, pretreatment using liquid hot water (hydrothermal), steam (steam explosion), biological pretreatment using microbe or enzymes, etc. (Naik et al. 2010) .

2.1.1. Physical pretreatment

The main objective of a physical pretreatment is to increase the surface area and reduce particle size of the lignocellulosic biomass, to decrease the degree of polymerization and decrystallization of cell wall. Some examples of physical pretreatment are milling, grinding, chipping, freezing, among others. The mechanical pretreatment can improve by 5-25% the hydrolysis product and boost the hydrolysis rate by 23-59%, depending on the milling techniques. The main drawback of this pretreatment is the high energy consumption, and this depends on the desired particle size (Aditiya et al. 2016; Haghighi Mood et al. 2013).

2.1.2. Chemical pretreatment

The chemical pretreatment consists in the addition of a specific chemical substances to the lignocellulosic biomass. The purpose of this pretreatment is to convert the lignocellulosic material into a more suitable form for the hydrolysis process, the chemical substances degrade the cell walls and the complex carbohydrate chain through direct chemical reaction that requires less energy (G. Kumar et al. 2019; Mussatto 2016).

The most common chemical pretreatment is with the use of acids, for instance, hydrochloric acid, nitric acid, and sulphuric acid. In general, the pretreatment can be carried out in two manners: (1) concentrated acid pretreatment and (2) dilute acid pretreatment. The first pretreatment requires shorter time and moderate temperature to yield the sugar monomers, but in practice, the drawback is the formation of inhibitors, easily corrodes the equipment and possible degradation of sugar monomers. The second pretreatment behavior is different from the first one, it can be performed in different paths: (1) batch process with low temperature (less than 160 °C) and high substrates loadings (10-40% w/w), and (2) continuous process with high temperature (more than 160 °C) and low substrate loadings (5-10% w/w) (Aditiya et al. 2016).

Alkaline pretreatment is also known for causing chemical swelling of fibrous cellulose, in which, saponification and salvation reactions occur which lead to the disruption the crosslinks between hemicelluloses, increasing the porosity of biomass. Comparatively, alkaline pretreatment is operated at lower temperatures and does not require complex reactors. However, the major drawbacks are long residence time (from hours to days) and the need for neutralization of the pretreated slurry. Sodium hydroxide (NaOH), potassium hydroxide (KOH), calcium hydroxide (CaOH2), and ammonia are mostly used in this pretreatment method (Haghighi Mood et al. 2013).

Wet Oxidation utilizes oxygen as an oxidizer for compounds dissolved in water. It uses to disrupt the hemicellulose and lignin in the lignocellulosic biomass. There are two reactions that occur during this process: (1) a low temperature hydrolysis reaction and (2) a high temperature oxidation reaction. Normally, the procedure for wet oxidation consists of drying and milling lignocellulosic biomass to obtain particles that are 2 mm in length, to which water is added at a ratio of 1 L to 6 g biomass. Na_2CO_3 is usually used and introduced to the mixture to reduce the formation of byproducts. Air is pumped into the vessel to reach a pressure of 12 bar, the working temperature is 195 °C for a range of 10 to 20 minutes (Brodeur et al. 2011; G. Kumar et al. 2019; Mussatto 2016).

Organo-solvent dissolves lignin in the lignocellulosic biomass. This method is carried out by mixing organic solvent with inorganic acid as the catalyst, such as HCl or H_2SO_4. The typical organic solvents, for instance ethanol, methanol, phenol, acetone, triethylene glycol and ethylene glycol, are combustible, this pre-treatment method must be performed under extreme safety care. Although its effectiveness in producing high quality bioethanol, this method is industrially impractical due to high solvents cost (Aditiya et al. 2016).

2.1.3. Physicochemical pretreatment

Steam Explosion is widely used as a pretreatment option in order to break the structure of the lignocellulosic biomass by applying a combination of both chemical and physical techniques. The pretreatment consists in subject the biomass to high pressure and temperature for a short residence time after which it rapidly depressurizes, disrupting the structure of the cell wall. The disruption of the cell wall increases the accessibility of the cellulose to the enzymes during hydrolysis. The major parameter that affects the steam explosion efficiency is the particle size, and the literature shows that relatively large particle sizes are able to yield maximum sugar concentrations. The operating temperatures ranging from 160-270 °C for several seconds to few minutes (Brodeur et al. 2011; Haghighi Mood et al. 2013).

Acid catalysts is normally used in the steam explosion process in dilute quantities in order to improve hemicellulose hydrolysis during the pretreatment and cellulose digestibility further on in the process. Dilute acids, such as H_2SO_4 or SO_2, can decrease residence time and temperature of current operating process. By decreasing the residence time and temperature with the addition of acid catalyst, the inhibitory compounds are considerably reduced, the hemicellulose removal is almost completed, and the further hydrolysis is improved (Brodeur et al. 2011; Mussatto 2016).

Ammonia Fiber Explosion subjects the biomass to liquid anhydrous ammonia under high pressures and moderate temperatures with a rapidly release of pressure. The moderate temperatures (range from 60°C to 100°C) are significantly less than steam explosion process, so less energy use and overall costs associated with the process. The degree of disruption to the lignocellulosic biomass depends on the temperature, this directly affects the rapidness of the ammonia vaporization within the reactor during depressurization. The residence time can vary from low (5 to 10 min) to moderate (30 min) time depending on the degree of saturation needed for the type of biomass (Brodeur et al. 2011; Haghighi Mood et al. 2013; Mussatto 2016).

2.1.4. Biological pretreatment

Biological pretreatment uses microorganisms like brown, white and soft rot fungi for degradation of lignin and hemicelluloses from the lignocellulosic biomass, using white rot fungi that can degrade lignin seems the most promising alternative since they consume less energy and less damage to the environment. The byproducts produced during biological pretreatment do not normally inhibit subsequent hydrolysis since the pretreatment is carried out at mild conditions. The white rot fungus helps in delignification which in turn improves the enzymatic saccharification rate. Currently there is a need for unique consortia for biological pretreatment (Mussatto 2016; Sindhu, Binod, and Pandey 2016).

Effective biodegradation of lignocellulosic biomass takes place by synergistic action of microbial consortium including various bacteria and fungi. There are several advantages of using microbial consortium which include increase adaptability, improving productivity, improving enzymatic saccharification efficiency, control of pH during sugar utilization and increasing substrate utilization. The drawback of this technology is that at large scale operation it leads to high operational costs since pretreatment needs to be carried out in sterile conditions. Additionally, the process is too slow and is not recommended for industrial purposes (Sindhu, Binod, and Pandey 2016).

Table 3. Comparison of the different pretreatment for lignocellulosic biomass.

Pretreatment	Yield			Equipment Cost	Advantages	Disadvantages
	Sugar	Inhibitor	By-product			
Physical pretreatment						
Mechanical	Low	-	-	High	Reduce cellulose crystallinity	High power consumption
Chemical pretreatment						
Acids	High	High	High	High	Alter lignin structure via hydrolysis	Hazardous, toxic, corrosive
Alkaline	High	Low	High	Low	Remove lignin and hemicellulose	Long residence time
Wet oxidation	Low/high	-	Low/high	Low	Cellulose decristallization and non-toxic	-
Organo-solvent	High	High	High	High	Hydrolyze lignin and hemicellulose	Solvents are drained, evaporated, condensed & reused

Physicochemical pretreatment						
Steam explosion	High	High	Low	High	Remove hemicellulose	Incomplete destruction of lignin carbohydrate matrix
Ammonia fiber explosion	High	Low	-	Low	Remove lignin	Not efficient, high lignin content
Biological pretreatment is not recommended to use in the industrial level, high operating cost.						

Source: G. Kumar et al. 2019

2.2. Lignocellulosic biomass conversion

The selection of the conversion process and the form in which the energy is required is determined by the biomass source. Therefore, biomass conversion into energy is carried out in different ways. The use of biomass to produce energy is in two different forms; (1) biomass can be directly burned or (2) converted into liquid fuels to subsequently burn as fuel, obtaining three different types of products: electricity/heat energy, transport fuel, and chemical feedstock (McKendry 2002).

Direct combustion is widely used to convert biomass to heat and/or steam for cooking, space heating, and industrial processes, or electricity generation. Small scale applications, for instance, the domestic used can be inefficient, with heat transfer losses up to 90%. For large-scale uses, the biomass is combusted in furnaces and boilers to produce process heat or steam for a steam turbine generator. The potential use of direct combustion is by combining the heat and electricity production systems (or cogeneration systems), improving the economics of the process (Demirbaş 2001).

The liquid fuels from biomass are obtained through thermochemical and biochemical processes. The main differences between them are the working temperatures. For thermochemical processes, the working temperature is much higher than biochemical processes, they may also require the use of a catalyst, and in general, the thermochemical processes are faster than the biochemical ones (Gollakota, Kishore, and Gu 2018).

2.2.1. Thermochemical processes

The thermochemical processes are divided into gasification, super critical fluid, pyrolysis, and liquefaction. The difference between the last two processes is that the feedstock macromolecule compounds of the liquefaction are decomposed into fragments of light molecules in the presence of a catalyst, while pyrolysis does not require a catalyst, and the light decomposed fragments are converted to oily compounds through homogeneous reactions in the gas phase (Demirbaş 2001).

Pyrolysis is the thermochemical process that converts biomass into liquid, charcoal, and non-condensable gases by heating the biomass to about 750 K in anaerobic conditions. Pyrolysis produces energy fuels with high fuel-to-feed ratios, making it the most efficient process for biomass conversion, having a high potential to compete directly with fossil fuel resources. If the purpose is to maximize the yield of the liquid products, a low temperature, high heating rate, and short gas residence time process is required. For high chart production, a low temperature, low heating rate process is required, and among other operating conditions.

Fast pyrolysis is characterized by high heating rates, very short vapor residence time, moderate and careful control of the temperature, with rapid cooling or quenching of the vapors. The feedstock is prepared as small particles sizes and heat transfer rates to the particle between 600 and 1000 W/cm^2. To avoid the formation of char a residence time on the order of seconds is required. The temperature is in the range of 450-550 °C depending on the characteristics of the lignocellulosic biomass (Demirbaş 2001; Ibarra-Gonzalez and Rong 2019; Mussatto 2016).

Gasification is a form of pyrolysis, which is performed at high temperatures to optimize gas production. This gas is a mixture of carbon monoxide, hydrogen, and methane, additionally, carbon dioxide and nitrogen. The gas is more versatile than the original solid biomass, and can be burnt to produces heat and steam, or used to produce electricity in gas turbines. It is the latest generation of biomass energy conversion process and improves efficiency and reduces the investment costs of biomass electricity generation. To produce transportation fuels from lignocellulosic biomass via gasification, the main steps that take place include syngas generation followed by syn-gas cleanup and Fischer-Tropsch synthesis in a gasifier (Mussatto 2016).

Gasifiers are designed to gasify almost any kind of organic feedstock, including many types of wood, agricultural residues, and municipal solid waste. In a typical atmospheric fluidized bed gasifier, the feed together with bed material, are fluidized by the gasifying agents, such as air, oxygen and/or steam, entering at the bottom of the bed. Heat is supplied to the gasifier either directly or indirectly to raise the gasification temperature to 600–1000 °C. Residences times for the gasification reaction are in the order of 3–4 s (Demirbaş 2001; Ibarra-Gonzalez and Rong 2019; Mussatto 2016).

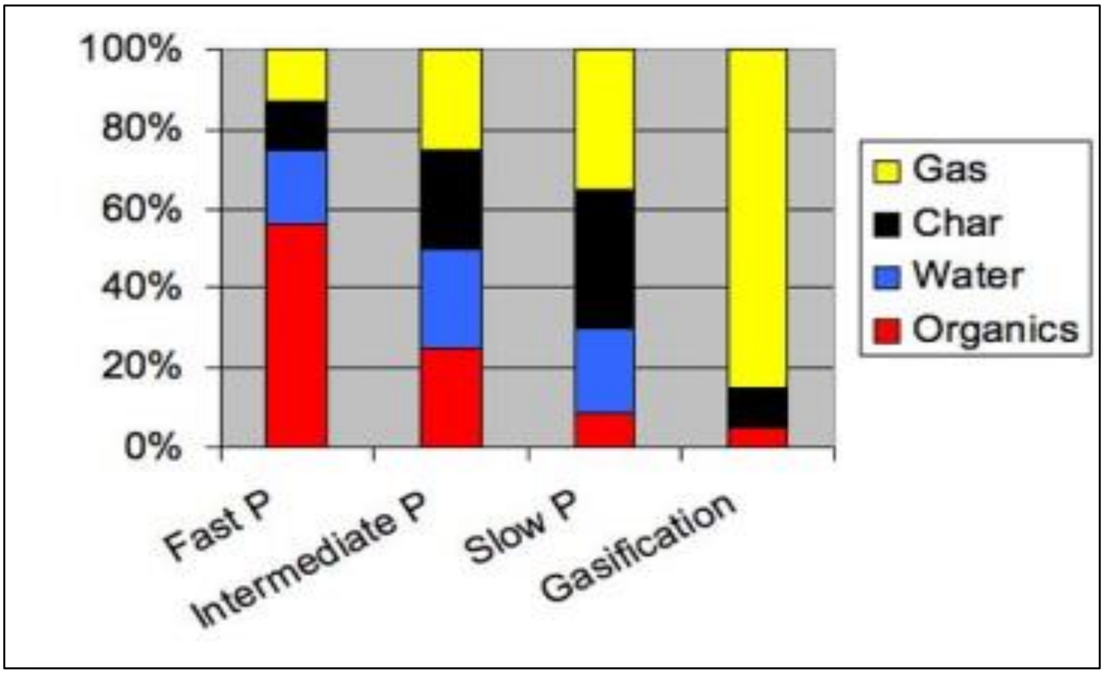

Figure 4. Comparison between Pyrolysis and Gasification products at different operating conditions (Mussatto 2016).

Liquefaction is a low-temperature, high-pressure thermochemical process that required the use of catalysts and hydrogen. Direct liquefaction, also called hydrothermal liquefaction is the thermochemical conversion of the biomass into liquid fuels by processing in a hot, pressurized water environment for sufficient time to break down the solid polymeric structure to mainly liquid components. The interest is low compare with the other technologies due to the complex and expensive reactors and fuel feeding systems (Ibarra-Gonzalez and Rong 2019).

Supercritical fluid extraction is the thermal disruption process of the lignocellulosic biomass in a range of temperatures of 250-400 °C under high pressure (4-5 MPa). A mixture of liquid and gas at equilibrium is heated and the thermal expansion causes the liquid to become less dense while the gas becomes denser as pressure increases. At the critical point, the densities of the two phases become identical and the distinction between them disappears. The unique properties at supercritical conditions, such as the strong dependence of the solubility of a material in a supercritical fluid to its density and good contact between oxidants and reactants, make supercritical fluid extraction ideal for separation and extraction of useful products and oxidation of organic materials. This technology, and compared to the others (Table 3), presents advantages like fast kinetics, higher biomass conversion, continuous operation, and without using a catalyst. The main drawback is the high temperature and pressure which increase the operating cost, and there is a lack of information on the literature that limits the use at an industrial level (Demirbaş 2001; Ibarra-Gonzalez and Rong 2019).

Table 4. Comparison of the thermochemical processes for biomass conversion.

Thermochemical technologies	Operation conditions/ requirements	Process description and reaction mechanisms	Technology feasibility	
			Advantages	**Disadvantages**
Gasification	Takes place at approximately 600–1000 °C in the presence of a controlled amount of oxidizing agent; residence time of 3–4 s; atmospheric pressure; drying necessary.	Syngas containing mainly H2, and CO is produced. Fuels production via gasification requires the syngas generation, followed by syngas cleanup, WGSR and Fischer–Tropsch synthesis.	Gasification is a mature technology and biomass gasification for generation of heat and power is already commercialized.	Gasification requires higher temperatures. Produces 20 wt% oil and 80 wt% gas, therefore it requires FT reaction to increase oil production.
Pyrolysis	Relatively high temperature (450–500 °C); a short residence time (~1 s); absence of oxygen; atmospheric pressure; drying necessary.	The light small molecules are converted to oily products through homogeneous reactions in the gas phase and rapid condensation.	High oil yield up to 80 wt% on dry feed; lower capital cost; Gasification produces heat and combustible gas, while pyrolysis produces bio-oil, gases and heat as principle products.	Poor fuel quality obtained.

Hydrothermal liquefaction	Lower temperature (300– 400 °C); long residence time (0.2–1.0 h.); high pressure (5– 20 MPa); drying unnecessary. Similar	Occurs in aqueous medium, which involves complex sequences of reactions (dehydration, decarboxylation, and hydrogenation of functional groups, etc.).	Better quality of bio-oil obtained (high heating value, low moisture content).	Relatively low oil yield (20 wt%–60 wt%); need high-pressure equipment, thus higher capital cost.
Supercritical fluid extraction	Similar conditions to HTL, 250– 400 °C, high pressure (4– 5 MPa); but longer residence times (20–240 min); requires solvent addition.	Thermal disruption of the lignocellulose or other organic materials for extraction of useful products and for oxidation of the organic materials.	Fast kinetics, higher biomass conversion, ease of continuous operation and elimination of the use of catalysts.	High temperature and pressure, which increase the operation costs. Bio-oil yield of 26 wt%–60 wt%.

Source: Ibarra-Gonzalez and Rong 2019.

2.2.2. Biochemical processes

The biochemical processes generally use microorganisms/enzymes to decompose biomass to produce biofuels or other added-value products in a hydrolyzation process. Through a fermentation reaction, the lignocellulosic biomass is converted into valuable biofuels, such as biohydrogen and bioethanol, and this conversion is eco-friendly and cost-effective. The biochemical processes of lignocellulosic biomass produce three valuable products cellulose (C6-sugars), hemicellulose (C5/C6-sugars), and lignin (G. Kumar et al. 2019).

Fermentation consists of the conversion of the C5-C6 sugars to bioethanol. The process is carried out in a batch, fed-batch, or continuous process. The selection of the most appropriate methods is based on the composition of lignocellulosic hydrolysate, kinetic properties of the microorganisms, and the process economics. The batch system is supplied with a limited amount of nutrients at the initial stage and inoculated with microorganisms to carry out the fermentation. For the industry level, fed-batch reactors are the most suitable to merge the benefits of batch and continuous processes, the major advantages are the increase in the concentration of the maximum viable cell, product accumulation, and prolong culture lifetime. In the case of the continuous process, the feed is continuously pumped with the culture medium, substrate, and necessary nutrients into an agitated vessel. Additionally, another mode of operation is by performing separately or simultaneously hydrolysis (Anuj Kumar Chandel et al. 2007; John, Muthukumar, and Arunagiri 2017).

Separate Hydrolysis and Fermentation (SHF), the enzymatic hydrolysis is performed in a separated step from the fermentation process. In the SHF process, the liquid flow from hydrolysis reactors enters into the fermentation reactor. The advantage of the SHF process is the capability to operate in optimal conditions of the two processes, for instance, the enzymatic hydrolysis works in the range of

318-323 K and the fermentation at 303 K. The disadvantages of the SHF process is the inhibition of the cellulase and β-glucosidase enzymes by glucose liberated during hydrolysis, which leads to poorer solid loadings and higher enzyme loadings to attain sensible yields (John, Muthukumar, and Arunagiri 2017; Wright, Wyman, and Grohmann 1988).

Simultaneous Saccharification and Fermentation (SSF) is an alternative option to produce bioethanol from lignocellulosic biomass. In SSF process, cellulase and β-glucosidase, and fermentable enzymes, such as *S. cerevisiae,* are in the same bioreactor to overcome the enzyme inhibition by keeping the lower concentration of cellobiose, glucose, and xylose. The operating conditions for the SSF process are at 37 °C and atmospheric pressure (Abril and Abril 2009; John, Muthukumar, and Arunagiri 2017).

Continuous Immobilized Fermentation is an efficient process due to the immobilization of the microorganisms for fermentation. The advantages are that the immobilized enzymes can be reused for further fermentation, reducing the contamination problems, enhanced fermentation productivity, shorter fermentation time, high cell density, easy downstream processing, and high tolerance to inhibitors (John, Muthukumar, and Arunagiri 2017).

2.3. Ethanol production

Ethanol production is increasing in recent years, and lignocellulosic biomass as a renewable energy source is receiving more attention around the world. The main producers of ethanol are Brazil and United States. In 2017, both countries represented 84% of the total ethanol production worldwide, while China only contributed with 3%. Most of the ethanol produced is utilized as fuel (around 73%) whereas 17% goes to the production of alcoholic beverages, and around 10% is used in the chemical industry. Many countries around the world have targeted the reduction of oil imports and the use of more sustainable fuels to reduce their CO_2 emissions. The great advantage of bioethanol is that the actual automotive engines do not need significant changes to use ethanol as fuel and the actual infrastructure of gasoline could be used for its distribution (Abril, 2009; Fang, Wu, and Xie 2019; Srivastava et al. 2015).

The controversial *first-generation biofuel* is a well-established industry, and the main way of bioethanol production in Brazil is by using sugarcane. There are other feedstocks used in the first generation, such as maize or edible oil seed, but like sugarcane, they compete directly with food production, being one of the reasons that the food prices increase. Therefore, another alternative has been proposed to the first-generation biofuels, the lignocellulosic feedstocks (Anuj K. Chandel et al. 2019; Naik et al. 2010).

The lignocellulosic biomass is the feedstock of the *second-generation biofuels* and it is the most abundant organic material on Earth. It can obtain from forests, agricultural and industrial residues, for instance, fruit waste. These feedstocks are generally cheaper or even costless than the conventional agricultural feedstocks and contribute to minimizing the negative effect that these residues have on the environment when they do not have correct disposal (Abril, 2009; Srivastava et al. 2015).

The most important feature to assess the feasibility of any country for bioethanol production is to evaluate the amount of fruit waste the country generates and then, how much of the waste is available to use, analyzing also whether the fruit residues are easily accessible or not. Therefore, Mexico is a potential option for bioethanol production from fruit waste due to the high amount of these residues generated in the fruit processing industries. The main fruits produced in Mexico are citrus fruits, such as orange, lime, mandarin, and banana and mango fruits. According to SAGARPA 2017, the annual production of millions of tons of these fruits were 4.600, 2.420, 0.285, 2.230, and 1.890 for orange, lime, mandarin, banana, and mango, respectively (Table 3).

The total amount of fruit residues used in the fruit processing industries it is possible to estimate since the values of the exportations are known. SAGARPA 2017 reported that the percentage of the fruit exported are 1.3%, 27.7%, 1.8%, 22.4%, and 18.9% for orange, lime, mandarin, banana, and mango, respectively, and approximately, 42% of the fruit is used in the processing industries in Mexico. Therefore, the million tons for processing industrial use are approximately 1.924, 0.735, 0.118, 0.727, and 0.643 for orange, lime, mandarin, banana, and mango, respectively.

Table 5. Annual production of fruits, percentage of exportations of each fruit, and the total million tons of fruits for local use of the different fruits considering in this project (SAGARPA 2017).

Fruits	**Annual production (million tons)**	**Percentage of exportations (%)**	**Total of fruit for local use (million tons)**
Orange	4.600	1.3	4.580
Lime	2.420	27.7	1.750
Mandarin	0.285	1.8	0.280
Banana	2.230	22.4	1.730
Mango	1.890	18.9	1.530
Total	11.425		9.87

According to the literature already mentioned in the last chapter (Table 2), the percentage of residues generated concerning the total weight of the fruit is approximately 50% for citrus waste, considering peel, seed, and pulp. In the case of bananas, the percentage of the residues is approximately 30% of the total weight, and mango waste is approximately 30% of the total weight, considering only the peel. Therefore, the total processing industrial residues are approximately 0.9620, 0.3675, 0.0590, 0.2181, and 0.1929 million tons for orange, lime, mandarin, banana, and mango respectively, for a potential total of 1.8 million tons of fruit waste to use in the bioethanol production (Table 4).

Table 6. Total fruits for industrial use, percentage of residues generated, and total fruit residues of the different fruits considering in this project.

Fruits	**Total fruits for industrial use (million tons)**	**Percentage of residues generated (%)**	**Total fruit residues (million tons)**
Orange	1.924	50*	0.9620
Lime	0.735	50*	0.3675
Mandarin	0.118	50*	0.0590
Banana	0.727	30	0.2181
Mango	0.643	30 peel/50 with the seed	0.1929 (0.3215 with the seed)
Total	4.147		1.8

*considering peel, seed, and pulp

3. Bioethanol Production from Fruit Waste

The most promissory strategy to maximize the value derived from lignocellulosic biomass is the so-called "biorefinery", it allows the production of multiple commercially valuable products, creating new bio-based industries, and represents a key step for more sustainable chemical industries. In the traditional oil refinery, petroleum is converted into fuels and chemicals. On the other hand, biorefineries use different kinds of renewable feedstocks, such as crops and agricultural residues, municipality waste, and fruit waste, among others, to produce value-added products (Nizami et al. 2017).

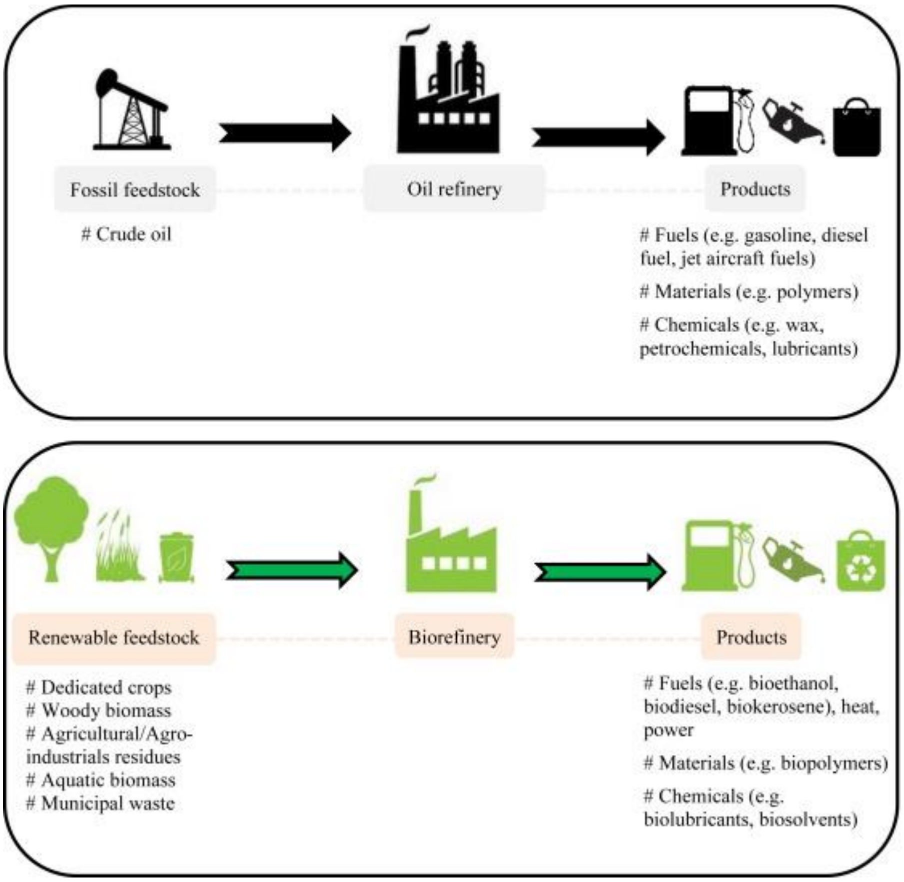

Figure 5. Comparison between oil refinery and biorefinery (Mussatto 2016).

To identify the most promising products with greater economic potential for a biorefinery, several framework factors should be taken into account, for instance, price, composition, and availability of the raw material, energy price, biomass pretreatment cost, market needs, the regulatory conditions, among others. However, the main challenge of the biorefinery is the adequate selection of the value-added products to have a sustainable production (Mussatto 2016; Satari and Karimi 2018; Zema et al. 2018).

3.1. Fruit waste to bioethanol process design

A complex process of biomass pretreatment is involved in the conventional bioethanol production to maximize the recovery of sugars from the hydrolysis of cellulose and hemicellulose, to ferment them into bioethanol (Petrova and Ivanova 2010).

Pretreatment should meet the following requirements: (a) improve the formation of sugars, (b) avoid degradation of carbohydrate, (c) avoid the formation of inhibitory byproducts to the subsequent hydrolysis and fermentation processes, and (d) be a sustainable process (P. Kumar et al. 2009). For the case of bioethanol production from fruit waste, two pretreatments are considered: grinding of the biomass, and then, steam explosion.

The effectiveness of enzymatic processes of the biorefinery is determined by the lignin content of the plant raw materials. In this context, the selection of feedstock with low lignin content is an important step for biomass conversion. Thus, the use of citrus waste, banana, and mango peel is suitable for bioethanol production since their lignin content is lower than 10% (Sánchez Orozco et al. 2014).

Previous to the steam explosion of the fruit waste, a mechanical pretreatment of this fruit is required. In this case, a grinding process. One of the advantages of mechanical pretreatment in biorefinery is the possibility of the simultaneous destruction of cell membranes. (P. Kumar et al. 2009).

For the case of this thesis, it is assumed that the fruit waste is ground in two steps, the citrus waste can be ground together (orange, lime, and mandarin) and then, the other fruit waste is ground (mango and banana peel). These last two fruit wastes can be mixed since any of them produce inhibitors for the next steps of bioethanol production (Arumugam and Manikandan 2011). While the citrus waste, the extraction of essential oils must be done since the essentials oils are inhibitors of the subsequent fermentation process (W. Zhou, Widmer, and Grohmann 2007).

After the mechanical pretreatment of the fruit waste, the pretreated biomass is sent to the steam explosion process. The steam explosion pretreatment has several advantages, for example, the use of hazardous chemicals is not needed, low capital investment, and low environmental impact. An important additional advantage is that all types of biomass can be loaded in the process and large chips can be used, reducing the energy required in the previous mechanical pretreatment. (Abril 2009; Mussatto 2016; Negro et al. 2016).

Steam explosion is operated at a temperature of 160°C for 5 minutes, with a sudden release of pressure from 6 bar to ambient pressure. In the case of citrus waste, the main role of the steam explosion is the removal of the content of the essential oil to lower than 0.05% (v/v). The effluent is cooled down by a condenser, and the essentials oils are recovered by decantation. Once the biomass is pretreated, it is sent to belt filter press to remove the non-soluble solids that can contain some fermentation inhibitors. (Boluda-Aguilar et al. 2010a; Weiyang Zhou, Widmer, and Grohmann 2008).

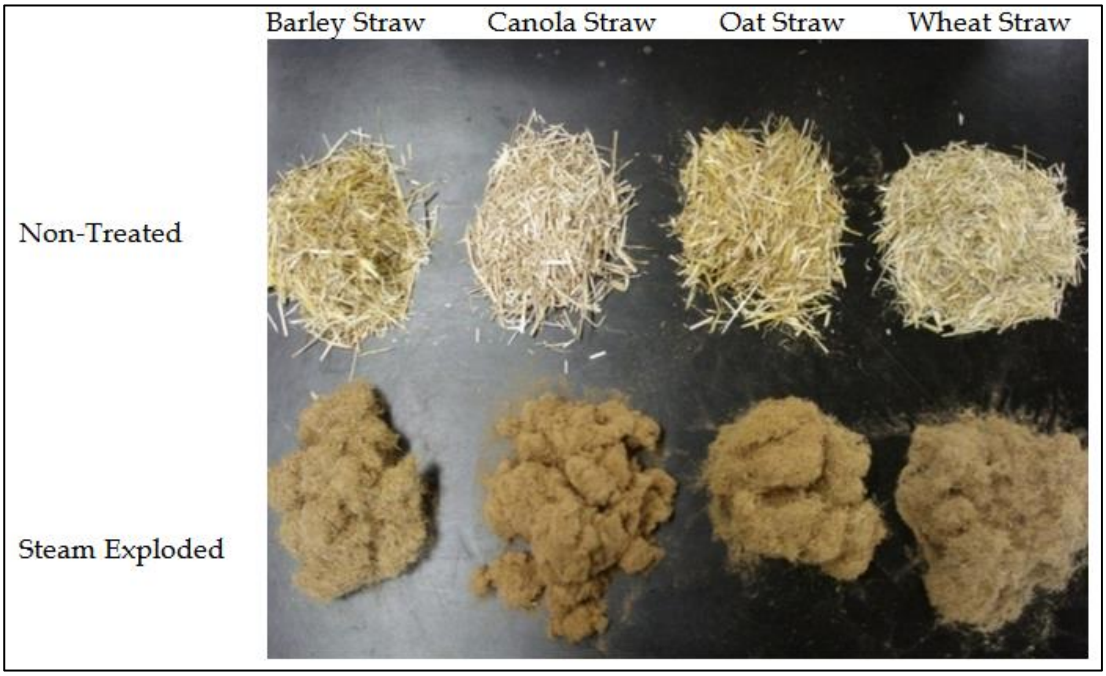

Figure 6. Comparison between non-treated biomass and steam-exploded biomass (Tabil, Adapa, and Kashaninejad 2016).

After the filtration, the pretreated biomass is sent to a SSF process. In this bioreactor of the SSF process, the cellulase enzymes and the fermenting microorganisms are added, and the microorganisms directly convert the glucose.

The ethanol yield in the SSF process is higher than the SHF, has a lower energy consumption and shorter processing time, thus, the investment costs are reduced. Additionally, the main benefits of the SSF process are that the end-product inhibition of the enzymatic hydrolysis is reduced compared to the SHF process, realizing high-solids fermentation, improved cellulose conversion rates, increased ethanol concentration, and reduced enzyme loadings (Haq et al. 2016; Petrova and Ivanova 2010; Wright, Wyman, and Grohmann 1988).

The SSF process conditions for this case are the following: (1) the pretreated citrus waste are sent to one bioreactor for the SSF process, and in another stream, the banana and mango peel are hydrolyzed and fermented in another bioreactor to avoid any kind of bioethanol production limitation due to the kind of fruit waste, (2) three enzymes for the hydrolysis process (pectinase, cellulase, and β-glucosidase), (3) the fermentation process is carried out with *Saccharomyces cerevisiae* yeast, and (4) the temperature is 37 °C at atmospheric pressure during 3 days. (Boluda-Aguilar et al. 2010b; Boluda-Aguilar and López-Gómez 2013; Lohrasbi et al. 2010a).

The outlet streams of the fermenters are sent to the ethanol purification process which consists of two beer columns, one rectification column, and the dehydration of the bioethanol process with molecular sieves. The main purpose of the beer columns is that they receive the beer outlet stream of the bioreactors that contain bioethanol, non-reacted sugars, pectin, solids, among others, and separate bioethanol-water mixture from the rest of the by-products formed in the fermentation process (Boluda-Aguilar et al. 2010b; Lohrasbi et al. 2010a; Pourbafrani et al. 2010). In this case, there are two beer columns because the first one receives the outlet stream of the citrus waste fermenter, and the second one receives the outlet stream of the banana and mango peel fermenter.

The top outlet mixture bioethanol/water composition is about 75-15 % (w/w) in the beer columns. Then, both top streams of the beer columns are mixed and sent to the rectification column. Due to the azeotropic mixture that ethanol and water form, the maximum ethanol composition that can be obtained is around 95.6% (w/w) at 75.15 °C and standard atmospheric pressure. The ethanol separation from water presents some challenges that need to be considered: (1) energy consumption, and (2) ethanol concentration. For this thesis case, the outlet ethanol concentration from the rectification column is about 92% (w/w) to reduce the energy consumption but at the same time obtained a good water separation from the ethanol at 78.2 °C and standard atmospheric pressure.

The azeotropic mixture of ethanol/water is impossible to separate with a traditional distillation column. Therefore, different alternatives have been studied for the dehydration of the ethanol, for example, adsorption, chemical dehydration, azeotropic distillation, membrane processes, among others, being the most attractive option the adsorption process due to the low energy consumption.

The adsorption process considering is with molecular sieves technology. This technology selectively adsorbs the water from the azeotropic mixture to obtain almost a pure ethanol product. Molecular sieves adsorb approximately 22% of the water weight and operate in a semi-continuous mode. The advantage of this process is that the bed with the molecular sieves after the saturation with water can be regenerated. The molecular sieves are normally Zeolites with a nominal pore size of 3A. They are regenerated using temperature swings with hot carrier gas at 175-260 °C for the 3A zeolites, for a bigger zeolite, the temperature is higher. A less energy consumption regeneration process can be with desorbing agents, for instance, methanol or acetone (Guizzetti and Nicora 2013; W. Zhou, Widmer, and Grohmann 2007). Figure 7 shows the process diagram considering in this thesis work.

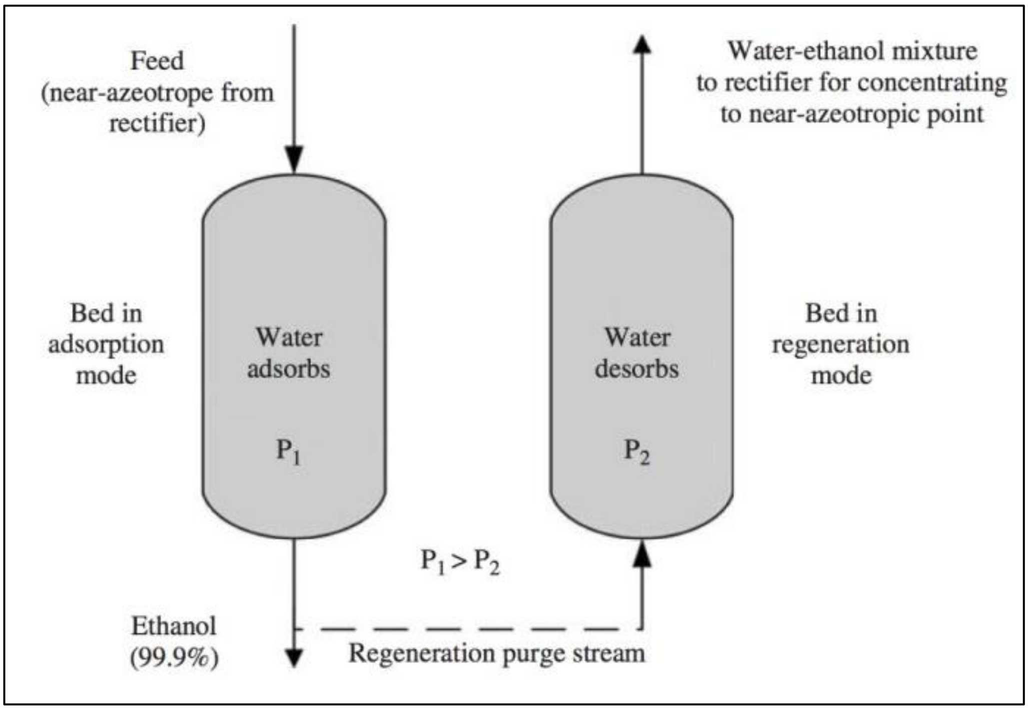

Figure 7. Molecular sieves columns (Guizzetti and Nicora 2013).

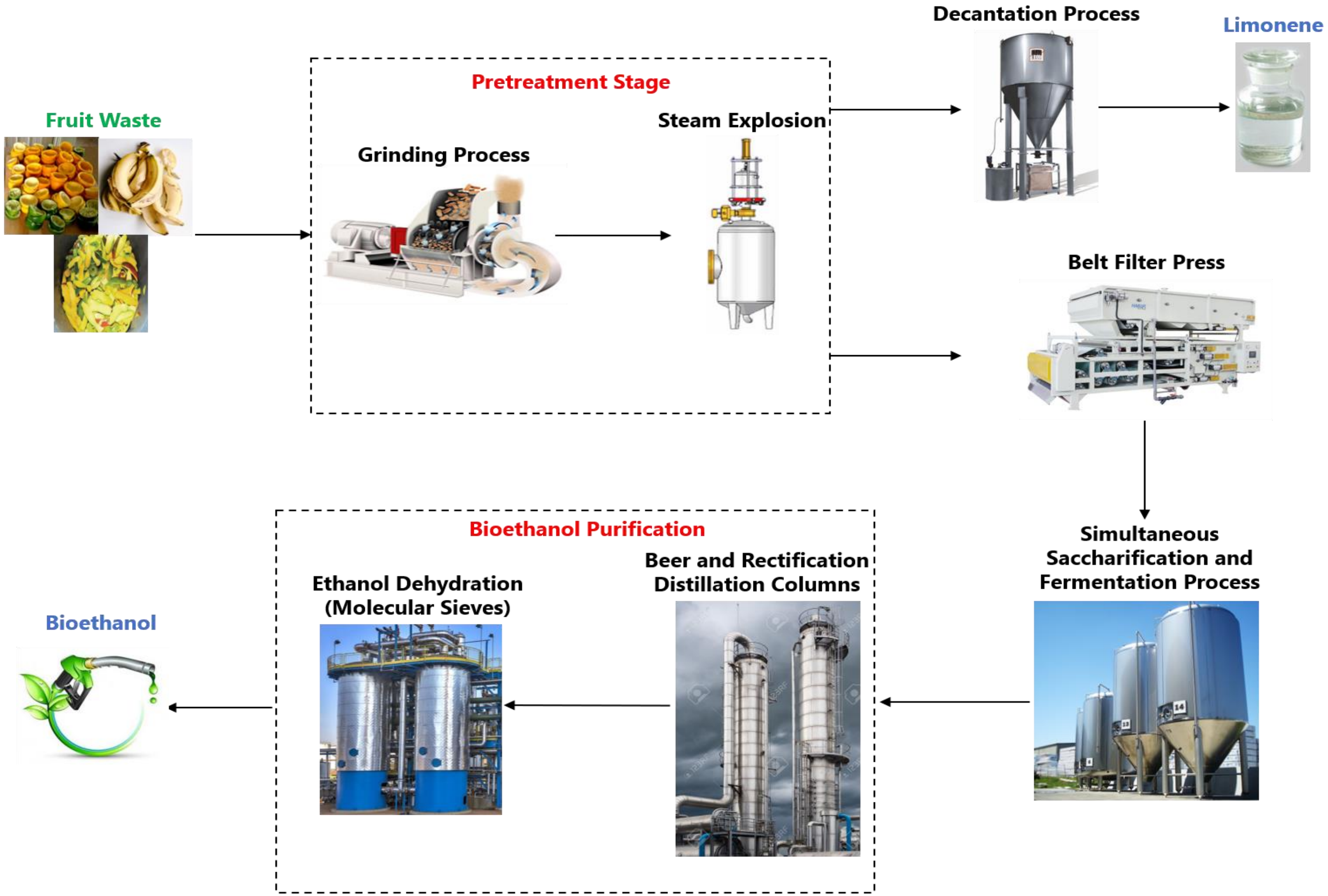

Figure 8. Fruit waste to bioethanol process diagram.

3.2. Process simulation

Fruit Waste to Bioethanol Process Simulation was carried out with UniSim software to estimate the rigorous material and energy balances calculations. The physical properties of the different lignocellulosic biomass components are not available in the standard UniSim database, so the properties of the cellulose, hemicellulose, and lignin were added and obtained from the literature found on different academic research websites (Boluda-Aguilar et al. 2010a; Lohrasbi et al. 2010b; Pathak, Mandavgane, and Kulkarni 2016).

The main unit operations for the steam explosion are considered for simulation; the pressure vessel working at 6 bar and 160°C, then, the atmospheric vessel where the vapor phase is condensed and sent to the decantation process to separate the limonene from the water, and the liquid part is sent to the filtration process. The following figure 8 shows the process simulation of the steam explosion pretreatment of the fruit waste.

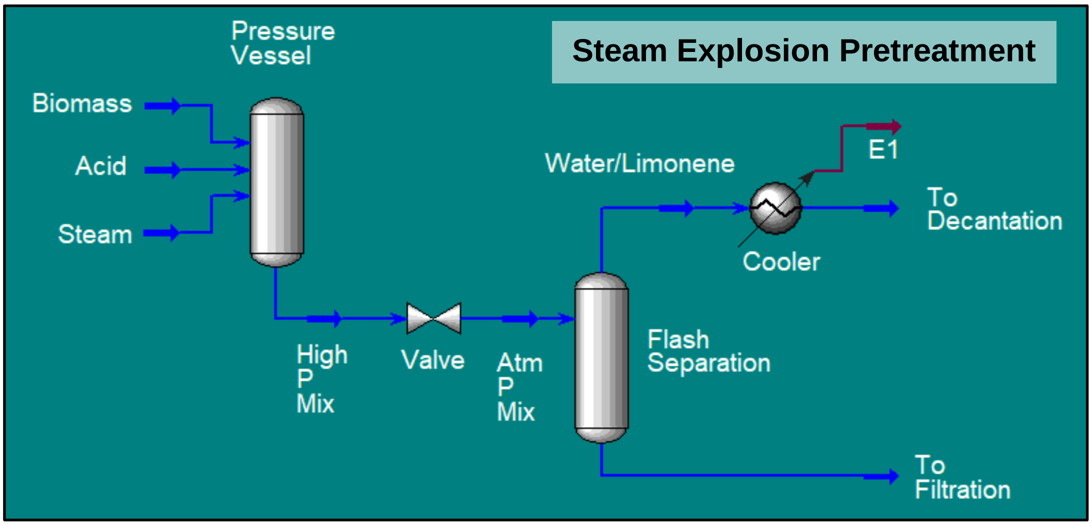

Figure 9. Steam explosion process simulation.

For the case of the SSF process, the main assumption is the amount of bioethanol that can be converted from 1 ton of wet fruit waste. For the case of simplicity, this assumption is based on the information found in the literature for the fruit waste

SSF process; additionally, this process is further explained in the economical assessment since different operating considerations that affect the bioethanol conversion but also the economic assessment is used, according to what was found on the literature.

After the SSF process, the outlet stream of the SSF reactor is sent to the bioethanol purification stage. The following figure 9 shows the process simulation of the bioethanol purification (Beer column and Rectification column), and table 10 shows the operating conditions of each distillation column.

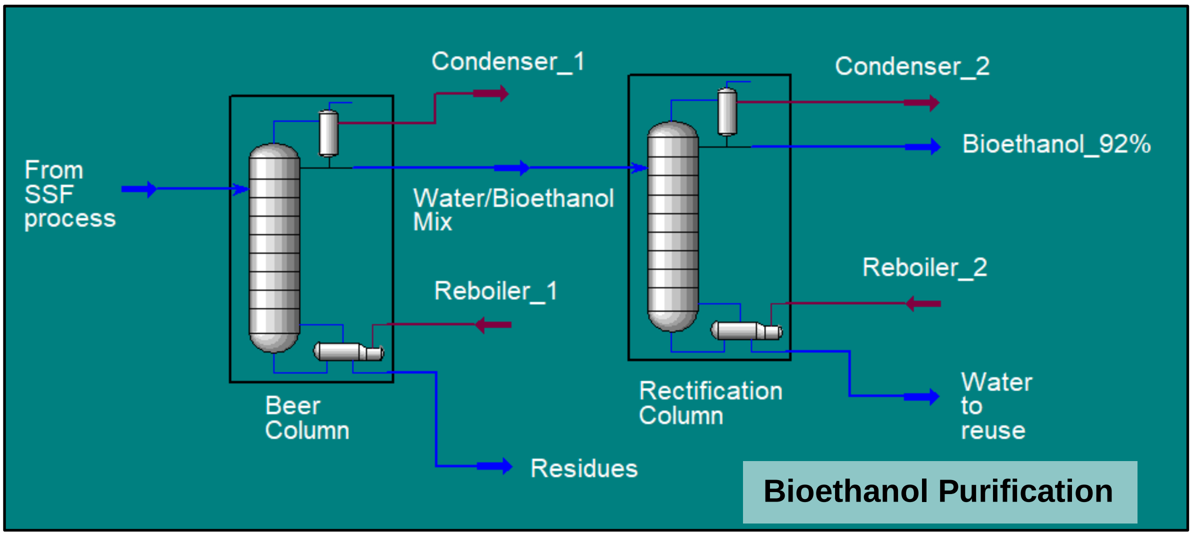

Figure 10. Bioethanol purification process simulation to obtain a bioethanol concentration of 92% w/w.

Figure 11. Operating conditions for the bioethanol purification process simulation.

Characteristics	**Beer Column**	**Rectification Column**
Number of trays	8	11
Bioethanol recovery	0.996	0.997
Bioethanol concentration	0.85	0.92
Reflux Ratio	2.1	2.0

4. Economic Assessment

Chemical engineering almost by definition involves the practical and economic combination of science, engineering, and mathematics to solve problems and make engineering decisions. The foundation of most estimates is the cost of individual pieces of equipment, and it is fundamental to considering costs and economics for engineering decisions and recommendations. Equipment cost estimates are required in industrial projects and are the first step in more detailed plant cost estimates. *Following books were considered for the development of this chapter: Beherans and Hawranek 1991; Garrett 1989; and Sinnott 2005*.

There two main contributions to take into account to estimate the total costs of an industrial plant: (1) the capital costs that are related to the construction of a new industrial plant or modifications to an existing plant, and (2) the operating costs that involve the daily operation of the industrial plant.

4.1. Classifications of capital cost estimates

The five classifications of capital cost estimates generally used in the process industries are the following:

1. *Order-of-Magnitude Estimate* (or *Ratio* or *Feasibility Estimate*). It relies on cost information for a complete process taken from already existing plants, and the available cost information is then adjusted through appropriate scaling factors for capacity and inflation, to provide the estimated capital cost. It normally requires only a Block Flow Diagram (BFD).
2. *Study Estimate* (or *Major Equipment* or *Factored Estimate*). A list of the major equipment is presented in the process, including pumps, compressors and turbines, columns and vessels, fired heaters, and heat exchangers. The approximate sizing is performed for each piece of

equipment and the approximate cost is determined. The total cost of equipment is then factored to obtain the estimated capital cost. It is based on the Process Flow Diagram (PFD) and costs from generalized charts.

3. *Preliminary Estimate* (or *Scope Estimate*). The level of accuracy of the equipment sizing needs to be more precise than the Study Estimate. An approximate layout of equipment is made with estimates of piping, instrumentation, and electrical requirements, considering also the utilities. It is based on PFD, vessel sketches for major equipment, preliminary plot plant, and elevation diagram.
4. *Definitive Estimate* (or *Project Control Estimate*). Preliminary specifications are required for all the equipment, utilities, instrumentation, electrical and off-sites. It is based on final PFD, vessel sketches, plot plant and elevation diagrams, utility balances, and a preliminary P&ID.
5. *Detailed Estimate* (or *Firm or Contractor's Estimate*). A complete engineering of the process and all related off-sites and utilities is needed, as well as vendor quotes for all expensive items. At the end of a detailed estimate, the plant is ready for construction. It is based on final PFD and P&ID, vessel sketches, utility balances, plot plan and elevation diagrams, piping isometrics.

The *Order-of-Magnitude Estimate* is based on the method by Hill (1956), the *Study Estimate* considered the method of Lang (1947-1948), and the *Preliminary Estimate* is based on the Guthrie method (1974).

4.2. Fixed and working capital.

Fixed capital cost is the total cost of the plant ready for start-up, it is a once-only cost, and it includes the cost of:

1. Design, and other engineering and construction supervision.

2. All items of equipment and their installation.
3. All piping, instrumentation, and control systems.
4. Buildings and structures.
5. Auxiliary facilities, for instance, utility land and civil engineering work.

Working capital is the additional investment needed to start the plant up and operate it to the point when the incomes are earned and includes the cost of:

1. Start-up
2. Initial catalyst charges.
3. Raw materials and intermediates in the process.
4. Finished product inventories.
5. Funds to cover outstanding accounts from customers.

The sum of the fixed and working capital is the total investment needed for a project, and at the end of the project, most of the working capital is recovered. For this thesis case, the percentage of the working capital is the same considered for petrochemical plants, 15% of the fixed capital.

4.3. Cost estimation

All cost-estimating methods use historical data and are themselves forecasts of future costs. The Chemical Engineering Plant Cost Index is an inflation indicator made specifically for the chemical industry to approximately estimate the cost of each piece of equipment to the actual date of estimation, and it is given by the following expression:

$$Cost\ in\ year\ A = Cost\ in\ year\ B * \frac{Cost\ Index\ in\ year\ A}{Cost\ Index\ in\ yearB} \quad (1)$$

The chemical industry indexes like *Marshall and Swift*, *Equipment Cost Index* and the *Chemical Engineering Plant Cost Index* are the most used. They should be used with caution and judgment, considering that the longer the period over which the correlation is made the less reliable the estimate.

Another common and simple cost estimating method is based on the capacity units of earlier projects to estimate the actual cost of the same equipment with a different capacity and is given by:

$$\frac{C_2}{C_1} = \left(\frac{S_2}{S_1}\right)^n \qquad (2)$$

where C_2 = capital cost of the project with capacity S_2

C_1 = capital cost of the project with capacity S_1

And the value of the index *n* refers to the well-known six-tenths rule and is normally taken as 0.6, the concept of economy of scale is introduced: the larger the equipment, the lower the cost of equipment per unit of capacity. In table 5 some typical values of cost exponent "*n*" and unit capacities are given, with an additional description of the equipment type, the capacity units, and the typical values of the cost exponent for each equipment.

Table 7. Typical values of cost exponents for a process equipment selection (Moioli 2019).

Equipment Type	**Range of Correlation**	**Capacity Units**	**Cost Exponent *n***
Reciprocating Compressor with motor drive	0.75 to 1490	kW	0.84
Heat Exchanger Shell&Tube carbon steel	1.9 to 1860	m^2	0.59
Vertical Tank carbon steel	0.4 to 76	m^3	0.30
Centrifugal Blower	0.24 to 71	std m^3/s	0.60
Jacketed kettle glass lined	0.2 to 3.8	m^3	0.48

4.4. Total capital cost

To make a *preliminary cost estimate,* the equipment module costing technique is normally used. The technique relates all costs to the purchased cost of equipment evaluated for some base conditions and then, accounts for deviations from these base conditions by the use of appropriate multiplying factors (Guthrie 1969, 1974; Navarrete 1995). They refer to the specific equipment type, system pressure, and materials of construction.

The definition of the *Bare Module Equipment Cost* is introduced, it is the sum of the direct and indirect costs related to the project, so it is calculated as the product between the *Purchased Cost* for base conditions and a *Bare Module Cost Factor:*

$$C_{BM} = C_P^0 F_{BM} \qquad (3)$$

where C_{BM} is the *Bare Module Equipment Cost* to account for direct and indirect costs for each equipment, F_{BM} is the *Bare Module Cost Factor* to account for the items and the specific materials of construction and operating pressure, and C_P^0 is the *Purchased Cost* for base conditions of each equipment, taking into account the material that is made, and pressure operating condition.

The Direct Costs cover the material needed for installation (piping, insulating material structural support, etc.) and the equipment cost, and it is calculated by the following expression:

$$C_P = C_P^0 F_M F_P \tag{4}$$

where C_P is the direct cost of equipment, F_P is the actual operating pressure, and F_M is the construction material. The Indirect Costs consider transport and insurance costs with taxes, labor costs for installation, and the engineering and supervision costs of the project.

Hence, the *Bare Module Cost* is calculated with the following expression:

$$C_{BM} = C_P^0 F_{BM} = C_P^0 (B_1 + B_2 F_M F_P) \tag{5}$$

where B_1 and B_2 are coefficients that depend on the equipment type and change every year.

The *Purchased Cost* of equipment is evaluated at the ambient pressure and the material that is made. In the case of this thesis, the assumptions for the material equipment are: Stainless Steel SS-316 for the hydrolysis reactors and the expansion vessels and Stainless Steel SS-304 for the other processes. The logarithmic expression to calculate the C_P^0 is the following:

$$\log_{10}(C_P^0) = K_1 + K_2 \log_{10}(A) + K_3[\log_{10}(A)]^2 \tag{6}$$

where A is the capacity or size of the equipment while the coefficients K_1, K_2 and K_3 depend on the equipment type and vary every year. The following table 6 provides the coefficients that are considered in this thesis for the year 2001.

Table 8. Equipment coefficients for the calculation of *Purchased Cost* in 2001 (Moioli 2019).

Equipment	Type	K_1	K_2	K_3	A
Heat Exchanger	Fixed tubes	4.3247	-0.3030	0.1634	Exchanger area [m^2]
	Kettle	4.4646	-0.5277	0.3955	
Columns (vessel)	With trays or packed	3.4974	0.4485	0.1074	Volume [m^3]
Trays	Sieve	2.9949	0.4465	0.3961	Cross sectional area [m^2]

The indexes to estimate equipment are considered for the year 2019 and are shown in the following table. The *Chemical Engineering Plant Cost Index (CEPCI)* is considered for this estimation.

Table 9. *Chemical Engineering Plant Cost Index* for the years 2001, 2009, and 2019 (CEPCI website).

CEPCI	2001	2009	2019
CE Index	394.4	532.3	607.5
Equipment	437.7	6319	740.0
Tanks/Heat Exchanger	365.7	587.0	649.8

The *Bare Module Cost Factor* (F_{BM}) can be calculated with the following equation:

$$F_{BM} = B_1 + B_2 F_M F_P \qquad (7)$$

where F_p is the pressure factor and F_M is the material factor, and the coefficients B_1 and B_2 depend on the equipment type and vary every year (table 8). For the case of Stainless Steel 304 and Stainless Steel 3016, the values of the F_M are 1.7 and 2.1, respectively.

Table 10. Equipment coefficients for the calculation of *Bare Module Cost Factor* in 2001 (Moioli 2019).

Equipment	Type	B_1	B_2
Heat Exchanger	Fixed tubes	1.63	1.66
	Kettle	1.63	1.66
Vertical tanks (columns/included)	/	2.25	1.82
Pumps	Centrifugal	1.89	1.35

For the calculation of the F_p the following equation is used for pressure between 5 and 140 bar, in the case of pressure less than 5 bar F_p=1.

$$\log_{10}(F_P) = 0.03881 - 0.11272 * \log_{10}(P) + 0.08183[\log_{10}(P)]^2 \quad (8)$$

For the calculation of the heat exchanger areas the following assumptions are considered:

- Condenser for the outlet stream of the steam explosion, U_{HE}=280 W/m^2*K
- For the condenser and the Kettle in the distillation process, $U_{Condenser}$=650 W/m^2*K and U_{Kettle}=1100 W/m^2*K

For the installation cost of perforated trays, the following equation is used.

$$C_{BM} = C_P^0 N F_{BM} F_q \quad (9)$$

$$\begin{cases} \log_{10}(F_q) = 0.4771 + 0.08561 * \log_{10}(N) - 0.3473[\log_{10}(N)]^2 & N < 20 \\ F_q = 1 & N \geq 20 \end{cases} \quad (10)$$

$$N = numbre\ of\ trays;\ F_{BM} = 1 \quad (11)$$

For the estimation cost of the fermentation tanks, the belt filter press and the Molecular sieves, the following equations are used (Guizzetti and Nicora 2013; McGivney and Kawamura 2008).

- Fermentation tanks:

  $$C_{Fermentors} = 66532 * Volume^{0.5},\ Volume\ [m^3] \tag{12}$$

- Belt filter press:

  $$C_{BFP} = 146.29 * X + 433972, \tag{13}$$

  $$where\ X = Machine\ Capacity\,\frac{Gal}{h}, \min 800\ and \max\ 53{,}000\,\frac{Gal}{h}$$

- Molecular sieves:

  The price of the zeolites is around \$1,500 per ton.

  $$C_{MS} = 2 * (Cost\ of\ bed + 56.171 * W^{0.878}),\ W\ [kg] \tag{14}$$

While for the grinding process cost is approximately 15% of the Fixed Capital Investment (FCI) and the energy used is approximately 10% of the total energy (Mussatto 2016).

After the calculation of the *Bare Module Cost,* the *Total Module* or *Grass Roots Cost* can be calculated. The *Total Module Cost* takes into account the cost to make some moderate expansions or alterations to an existing facility, and the *Grass Roots Cost* contemplate a completely new construction in an undeveloped land, a grass field.

For the *Total Module Cost* calculation, it is considered an 18% extra of the *Bare Module Cost,* for the additional cost of contingency and fees besides direct and indirect costs. In particular, the percentage of the contingency cost depends on the reliability of the cost data and the detail level of the available process flowsheet, so the assumption for this case is 15% of contingency and 3% of fees of the *Bare Module Cost* (equation 15).

$$C_{TM} = \sum_{i=1}^{n} C_{TM,i} = 1.18 \sum_{i=1}^{n} C_{BM,i} \qquad (15)$$

Finally, the calculation of the *Grass Roots Cost* is with the addition of the auxiliary facilities costs to the *Total Module Cost*. Normally, it is assumed equal to 50% of the *Bare Module Cost* for the base conditions ($F_{M,0}$= $F_{P,0}$=1) (equation 16).

$$C_{GR} = C_{TM} + 0.50 \sum_{i=1}^{n} C_{BM,i}^{0} \qquad (16)$$

Generally, the cost contributions are independent of the material of construction and the operating pressure and include the cost of site development, auxiliary buildings, off-sites, and utilities.

4.5. Cost of manufacturing.

The *Cost of Manufacturing* refers to the daily operation of the chemical plant, the estimation of manufacturing cost is evaluated before the feasibility assessment of the proposed process. The manufacturing cost is affected by some different factors, and they are classified into three main categories:

1. Direct manufacturing costs. They refer to the operating expenses and vary with the production rate.
2. Fixed manufacturing costs. They are not affected by the production rate at all and they include property taxes, insurance, and depreciation.
3. General expenses. They are rarely affected by the production rate and they include management, sale, financing, and research functions.

Therefore, the *Cost of Manufacturing* is given by the following expression.

Cost of Manufacturing = Direct Manufacturing Costs + Fixed Manufacturing Costs + General Expenses (17)

To estimate the *Cost of Manufacturing* the following costs are required (Table 9).

1. Fixed Capital Investment (FCI).
2. Cost of operating labor.
3. Cost of utilities.
4. Cost of waste treatment.
5. Cost of raw materials.

Table 11. Factors that affect the *Cost of Manufacturing* in chemical plant (Moioli 2019).

Factor	**Description of Factor**
1. Direct costs	**Factors that vary with the production rate**
a. Raw materials	Feedstock cost for the process
b. Waste treatment	Cost of waste treatment
c. Utilities	Cost of utility streams in the process
d. Operating labor	Cost of labor in the plant
e. Direct supervisory and clerical labor	Cost of administrative/engineering and support personnel
f. Maintenance and repairs	Cost of labor and materials associated with the maintenance
g. Operating supplies	Cost of miscellaneous supplies
h. Laboratory charges	Cost of routine and special laboratory tests required for product quality control and troubleshooting
i. Patents and royalties	Cost of using patented or licensed technology
2. Fixed costs	**Factor that are not affected by the production rate**

a. Depreciation	Legal operating expenses for tax purposes.
b. Local taxes and insurance	Cost of property taxes and liability insurance
c. Plant overhead costs	Cost of operations of auxiliary facilities supporting the manufacturing process
3. General expenses	**Cost of management level and administrative activities different from the manufacturing process**
a. Administration costs	Cost of salaries, buildings, and other related activities
b. Distribution and selling costs	Cost of sales and marketing
c. Research and development	Cost of research and development activities

For this thesis case, the assumption for insurance and maintenance costs is 7% of the FCI and for salaries, the assumption is three times less than a normal salary in the case of Europe or USA (Lohrasbi et al. 2010a).

4.6. Scenario analysis

The base plant design and economic assessment start with 120,000 tons per year of fruits waste from which 80% is citrus waste (orange, lime, and mandarin) and 20% is banana and mango peel. For this evaluation, three different scenarios are considered: (1) the worst-case scenario, (2) the standard-case scenario, and (3) the best-case scenario.

4.6.1. Worst-case scenario

The worst-case scenario contemplates a negative scenario in selling prices and production at low levels. The selling price of ethanol is assumed to be $0.55/per liter that is a little more expensive than the MTBE price ($0.52/ per liter, according to the U.S. Grains Council in 2019, https://grains.org/). While in the case of the limonene, the selling price is assumed to be $6.00/per liter that is lower than the minimum price of limonene reported on the literature ($6.66/per liter, Sun, Theodoropoulos, and Scrutton 2020).

The production of the bioethanol is assumed to be 50 L/ per ton of citrus waste, lower than the maximum amount of bioethanol produces reported in the literature (60 L/per ton of citrus waste), and 30 L/ per ton of mango and banana peel. In the case of limonene, the amount of limonene recovery is 7.5 L/per ton of citrus waste, being the maximum amount of limonene recovery of 9.3 L/ per ton of citrus waste reported on the literature (Boluda-Aguilar et al. 2010a; Lohrasbi et al. 2010a).

4.6.2. Standard-case scenario

The Standard-case scenario contemplates a more positive scenario in selling prices and the production in standard levels. The selling price of the ethanol is assumed to be $0.45/per liter that is cheaper than the MTBE price ($0.52/ per liter). While in the case of the limonene, the selling price is assumed to be $8.00/per liter that is higher than the minimum selling price of limonene reported on the literature ($6.66-$17/per liter, Sun, Theodoropoulos, and Scrutton 2020).

The production of the bioethanol is assumed to be 55 L/ per ton of citrus waste, which is close to the maximum amount of bioethanol produces reported on the literature (60 L/per ton of citrus waste, Boluda-Aguilar et al. 2010a; Lohrasbi et al. 2010a), and 30 L/ per ton of mango and banana peel. In the case of limonene, the amount of limonene recovery is 8.2 L/per ton of citrus waste, being the maximum amount of limonene recovery of 9.3 L/ per ton of citrus waste reported on the literature.

4.6.3. Best-case scenario

The Best-case scenario contemplates an ideal scenario in selling prices and the production at high levels. The selling price of the ethanol is assumed to be $0.40/per liter that is much cheaper than the MTBE price ($0.52/ per liter). While in the case of the limonene, the selling price is assumed to be $11.00/per liter that is around the middle of the selling price of limonene reported on the literature ($6.66-$17/per liter).

The production of the bioethanol is assumed to be 60 L/ per ton of citrus waste, that is the maximum amount of bioethanol produces reported in the literature (60 L/per ton of citrus waste), and 30 L/ per ton of mango and banana peel. In the case of limonene, the amount of limonene recovery is 8.9 L/per ton of citrus waste, being the maximum amount of limonene recovery of 9.3 L/ per ton of citrus waste reported on the literature.

For the evaluation of each scenario, 4 variables are considered: a) the bioethanol price, b) the limonene price, c) the bioethanol production (L per ton) of fruit waste and d) the limonene recovery (L per ton). The following table 12 summarized the assumptions of each scenario.

Table 12. Assumptions for the scenario analysis.

Assumption	**Worst-Case**	**Standard-Case**	**Best-Case**
Ethanol price	0.55 $/L	0.45 $/L	0.40 $/L
Limonene price	6.00 $/L	8.00 $/L	11.00 $/L
Bioethanol production from citrus waste	50 L/ton	55 L/ton	60 L/ton
Limonene recovery	7.5 L/ton	8.2 L/ton	8.9 L/ton
Bioethanol production from mango and banana peel	30 L/ton	30 L/ton	30 L/ton

5. Results

5.1. Process costs

As mentioned in the previous chapter, the base capacity of the plant is 120,000 tons per year of fruits waste from which 80% is citrus waste (orange, lime, and mandarin) and 20% is banana and mango peel. Based on figure 8, the cost of each process of this plant is discussed.

5.1.1. Grinding process

According to the literature, the cost of the grinding process is approximately 15% of the Fixed Capital Investment (FCI) and the energy used is approximately 10% of the total energy. Therefore, the cost of the grinding process is about \$1,775,335.50 (CEPCI Index 2019) and the operating cost is about \$13,588.00 per year.

5.1.2. Steam explosion

The steam explosion is designed for pretreatment of 15 ton/h of fruit waste, 12 ton/h for citrus waste, and 3 ton/h for banana and mango waste. The size of the pressure vessel for citrus waste is 2 m^3 and 1 m^3 for banana and mango waste, and the expansion vessel is 10 times larger than the pressure vessel. According to these sizes of pressure vessels, the first one is capable of pretreating 6 ton/h of fruit waste, so 2 pressure vessels are required to satisfy the 12 ton /h of citrus waste, while the second one is capable of pretreating 3 ton/h of fruit waste, so only one pressure vessel is required to satisfy the 3 ton/h of banana and mango waste.

The cost of each pressure vessel of 2 m^3 is \$47,371.08 and the expansion vessel is \$197,559.90, while the cost of each pressure vessel of 1 m^3 is \$33,944.45 and the expansion vessel is $121,979.40 (CEPCI Index 2019). The actual total cost of the full steam explosion process is $645,779.80 (CEPCI Index 2019).

The cost of the heat exchanger to cool down the water and limonene mixture is $143,668.16, the cost of the decanter is $95,345.73 and the cost of the storage vessel of limonene is $31,732.70 (CEPCI Index 2019). Moreover, the total operating cost of the process includes the cost of steam, sulfuric acid, electricity, cooling water that is around $646,944.00 per year.

5.1.3. Belt filter press

The cost of the belt filter press for the citrus waste is $ 1,077,893.11 and for banana and mango waste is $603,254.65 (CEPCI Index 2019), for a volume flow of 4,300 gal/h and 1,100 gal/h respectively.

The total cost of the full belt filter press process is $1,681,147.76 (CEPCI Index 2019). The operating cost only includes the price of the electricity, so the total operating cost of the belt filter press is about $22,140.00 per year.

5.1.4. Fermentation

The cost of each fermenter reactor is $1,093,232.32 (2009) with a size of 270 m^3 (total volume) for an operation of 72 h. The total number of fermenters required in this plant is 5, 4 fermenters are used for the fermentation of citrus waste and only one for banana and mango peel. The total cost of the full fermentation process is $6,401,265.34 (CEPCI Index 2019).

The operating cost includes the cost of the enzymes used and the price of the electricity. The total operating cost of the fermentation process is about $527,127.00 per year, and the enzymes account for 90% of the total operating cost.

5.1.5. Bioethanol purification

The cost of the first step, Beer column, is $153,995.00 (CEPCI Index 2019) with a total volume of 13.22 m^3 approximately, and 8 trays. Then, the cost of the Rectification column is $157,027.32 (CEPCI Index 2019) with a total volume of 13.63 m^3 approximately, and 11 trays. Finally, the cost of a molecular sieve system is $308,308.15 (CEPCI Index 2019), considering the amount of zeolites used in this process.

The actual total cost of the full bioethanol purification process is about $1,060,514.00 (CEPCI Index 2019). The operating cost includes the amount of steam, electricity, and cooling water required in this process. the total operating cost of the full bioethanol purification process is about $177,951.50 per year.

Table 13, figures 12 and 13 summarize the cost of each process. It is shown that the most expensive process for this Bioethanol plant is the fermentation process, as it represents more than 50% of the total FCI. Meanwhile, the operating cost of the Steam Explosion and Fermentation account for 85% of the total OPEX of the whole plant.

Table 13. The total cost of each process.

Process	***FCI***	***OPEX***	***Total***
Grinding	$ 1,775,335.48	$ 13,588.00	$ 1,788,923.48
Steam Explosion	$ 917,306.74	$ 646,944.00	$ 1,564,250.74
Belt Filter Press	$ 1,681,147.76	$ 22,140.00	$ 1,703,287.76
Fermentation	$ 6,401,265.34	$ 527,127.00	$ 6,928,392.34
Bioethanol Purification	$ 1,060,514.00	$ 177,951.50	$ 1,238,465.50
Total	**$11,835,569.32**	**$1,387,750.50**	**$13,223,319.82**

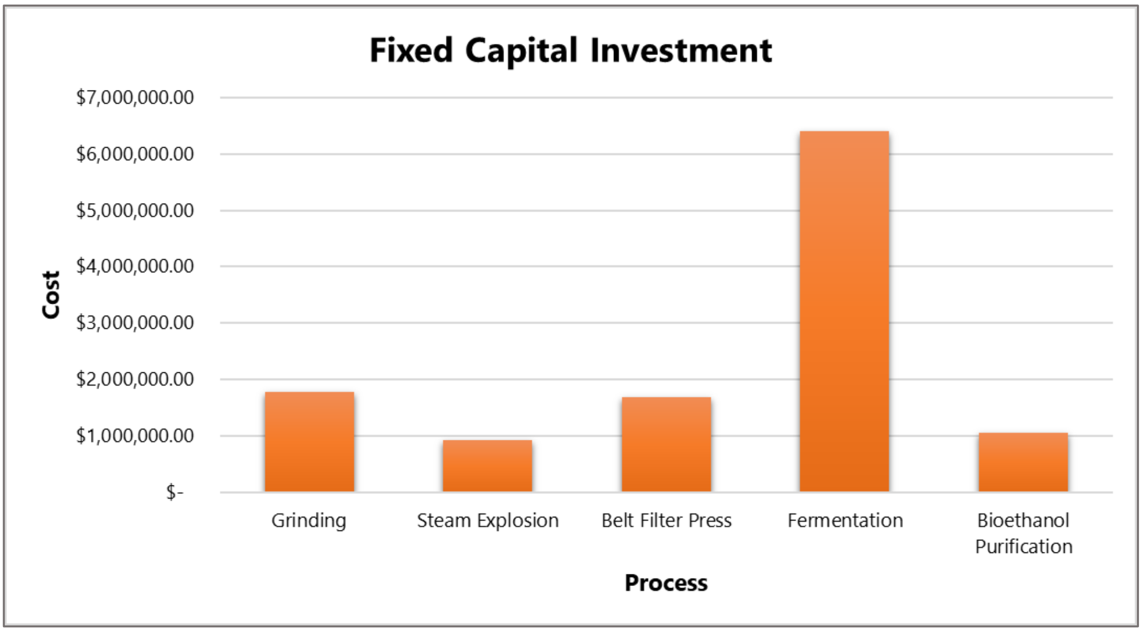

Figure 12. Fixed Capital Investment of each process.

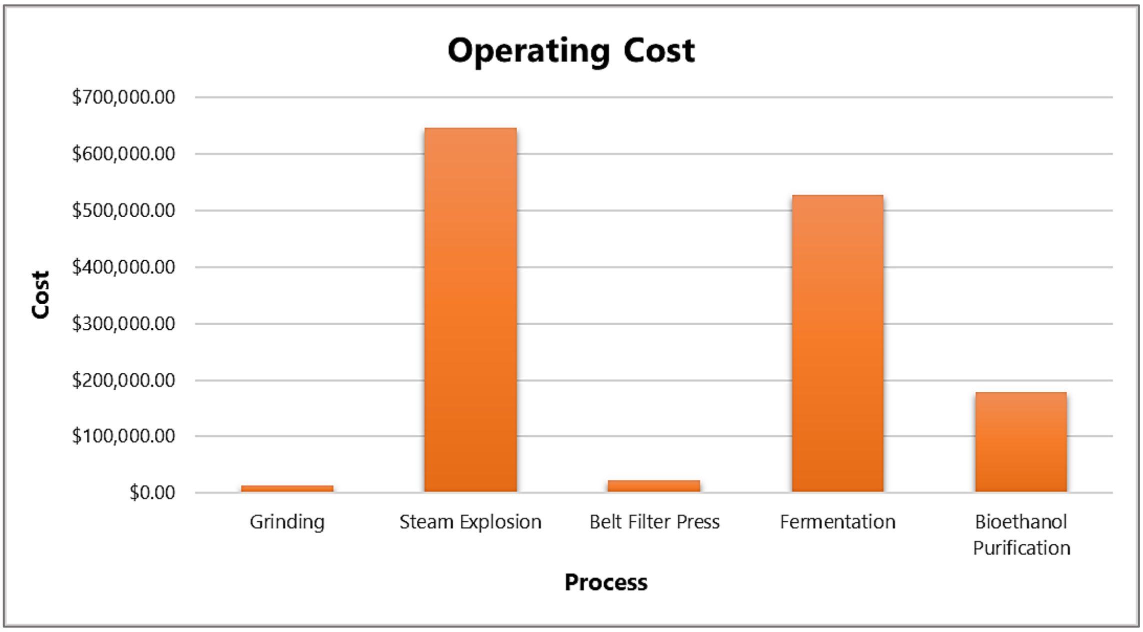

Figure 13. Operating Cost of each process.

The following table 14 shows the summary of the production cost of this Bioethanol plant.

Table 14. The production cost of the Bioethanol plant.

Variable Costs

1. Raw materials	$ 60,000 [USD/y]
2. Utilities	$ 1,387,800.00 [USD/y]
Fixed Costs	
3. Fixed capital investment	$ 11,835,570.00 [USD]
4. Maintenance	$ 710,140.00 [USD/y]
5. Insurance	$ 118,360.00 [USD/y]
6. Salaries	$ 401,000.00 [USD/y]
Total Operating Costs	$ 2,677,300.00 [USD/y]

As it was stated before, the salaries in Mexico are at least three times less than a normal salary in Europe or USA, so instead of being around $1,200,000.00 [USD/y], It is only around $400,000.00 [USD/y]. The cost of the raw materials is very low comparing with a petrochemical process. Therefore, the operating cost is reduced considerably every year.

5.2. Worst-Case scenario.

The economic analysis for the Worst-Case scenario is evaluated with the following information (Table 15) for a base capacity of 120,000 tons citrus waste per year:

Table 15. Economic Analysis information for the Worst-case scenario.

Information	**Quantity**
Total Fixed Capital Investment	$ 11,835,570.00 [$]
Total Module Cost	$ 13,965,980.00 [$]
Total Grass Roots	$ 20,048,850.00 [$]
Operating Cost	$ 2,617,300.00 [$/y]
Raw Materials	$ 60,000.00 [$/y]
Working Capital	$ 1,775,350.00 [$/y]
Depreciation	$ 1,336,590.00 [$/y]
Revenues	$ 6,620,400.00 [$/y]

The income statement (Table 16) is made by considering 30% of taxation (Mexico) and a discounted rate of 9%. The depreciation period is 15 years, and the construction time is 1 year. Additionally, it is assumed to have 90% of the total sales during the period, and the working hours per year are 8,000 hours.

Table 16. Income Statement of the Worst-Case scenario.

Year	**-1**	**0**	**0**	**1**	**2**	**3**	**4**	**5**
Revenues			0	$ 6,620,400.00	$ 6,620,400.00	$ 6,620,400.00	$ 6,620,400.00	$ 6,620,400.00
Expenses								
Investment			$20,048,850.00	0	0	0	0	0
Operation			0	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00
Feed			0	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00
WC			$ 1,775,350.00	0	0	0	0	0
Total			$ 1,824,200.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00
Profit			-$21,824,200.00	$ 3,943,100.00	$ 3,943,100.00	$ 3,943,100.00	$ 3,943,100.00	$ 3,943,100.00
Depreciation			0	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00
Net Income			-$21,824,200.00	$ 2,606,510.00	$ 2,606,510.00	$ 2,606,510.00	$ 2,606,510.00	$ 2,606,510.00
Before Taxes								
Taxes			0	$ 781,953.00	$ 781,953.00	$ 781,953.00	$ 781,953.00	$ 781,953.00
Net Income			-$21,824,200.00	$ 1,824,557.00	$ 1,824,557.00	$ 1,824,557.00	$ 1,824,557.00	$ 1,824,557.00
After Taxes								
Cash Flow			-$21,824,200.00	$ 3,161,147.00	$ 3,161,147.00	$ 3,161,147.00	$ 3,161,147.00	$ 3,161,147.00
Income Recovery								
(Discounted CF)	**0**	**-$20,048,850.00**	**-$21,824,200.00**	**-$18,924,065.14**	**-$16,263,390.95**	**-$13,822,405.46**	**-$11,582,969.23**	**-$9,528,440.58**

Year	**0**	**6**	**7**	**8**	**9**	**10**
Revenues	0	$ 6,620,400.00	$ 6,620,400.00	$ 6,620,400.00	$ 6,620,400.00	$ 6,620,400.00
Expenses						
Investment	$20,048,850.00	0	0	0	0	0
Operation	0	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00
Feed	0	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00
WC	$ 1,775,350.00	0	0	0	0	0
Total	$ 1,824,200.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00
Profit	-$21,824,200.00	$ 3,943,100.00	$ 3,943,100.00	$ 3,943,100.00	$ 3,943,100.00	$ 3,943,100.00
Depreciation	0	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00
Net Income Before Taxes	-$21,824,200.00	$ 2,606,510.00	$ 2,606,510.00	$ 2,606,510.00	$ 2,606,510.00	$ 2,606,510.00
Taxes	0	$ 781,953.00	$ 781,953.00	$ 781,953.00	$ 781,953.00	$ 781,953.00
Net Income After Taxes	-$21,824,200.00	$ 1,824,557.00	$ 1,824,557.00	$ 1,824,557.00	$ 1,824,557.00	$ 1,824,557.00
Cash Flow	-$21,824,200.00	$ 3,161,147.00	$ 3,161,147.00	$ 3,161,147.00	$ 3,161,147.00	$ 3,161,147.00
Income Recovery (Discounted CF)	**-$21,824,200.00**	**-$7,643,551.91**	**-$5,914,296.24**	**-$4,327,823.16**	**-$2,872,343.27**	**-$1,537,040.61**

Year	**0**	**11**	**12**	**13**	**14**	**15**
Revenues	0	$ 6,620,400.00	$ 6,620,400.00	$ 6,620,400.00	$ 6,620,400.00	$ 6,620,400.00
Expenses						
Investment	$20,048,850.00	0	0	0	0	0
Operation	0	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00
Feed	0	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00
WC	$ 1,775,350.00	0	0	0	0	0
Total	$ 1,824,200.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00
Profit	-$21,824,200.00	$ 3,943,100.00	$ 3,943,100.00	$ 3,943,100.00	$ 3,943,100.00	$ 3,943,100.00
Depreciation	0	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00
Net Income Before Taxes	-$21,824,200.00	$ 2,606,510.00	$ 2,606,510.00	$ 2,606,510.00	$ 2,606,510.00	$ 2,606,510.00
Taxes	0	$ 781,953.00	$ 781,953.00	$ 781,953.00	$ 781,953.00	$ 781,953.00
Net Income After Taxes	-$21,824,200.00	$ 1,824,557.00	$ 1,824,557.00	$ 1,824,557.00	$ 1,824,557.00	$ 1,824,557.00
Cash Flow	-$21,824,200.00	$ 3,161,147.00	$ 3,161,147.00	$ 3,161,147.00	$ 3,161,147.00	$ 3,161,147.00
Income Recovery (Discounted CF)	**-$21,824,200.00**	**-$311,992.30**	**$811,905.23**	**$1,843,003.88**	**$2,788,965.94**	**$3,656,821.05**

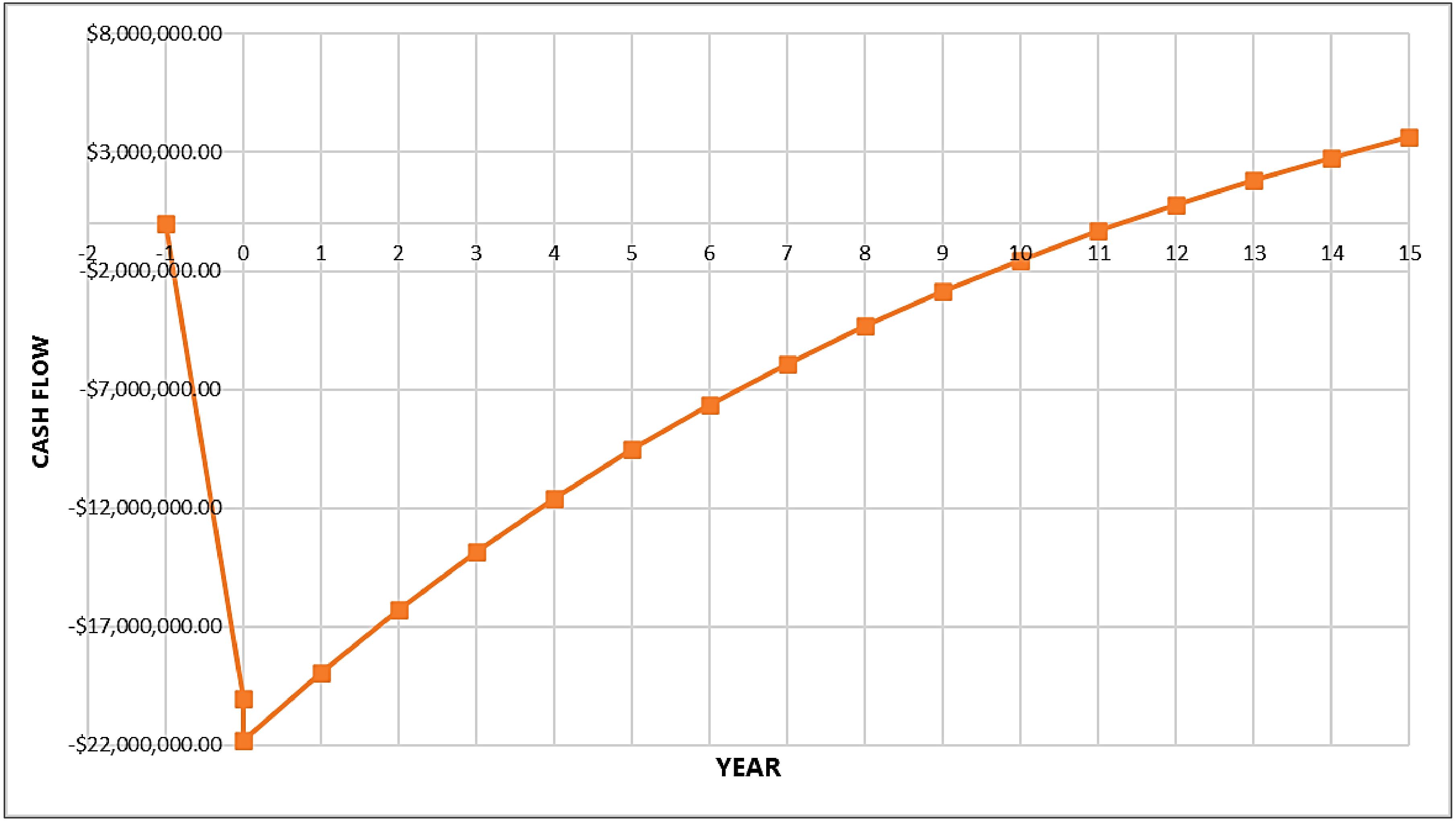

Figure 14. Cash Flow of the Worst-Case scenario.

The Internal Rate of Return (IRR) of the Worst-Case scenario is 11.7%, so this is the maximum discount rate that makes the Net Present Value (NPV) of all cash flows equal to zero in a discounted cash flow analysis (Figure 15).

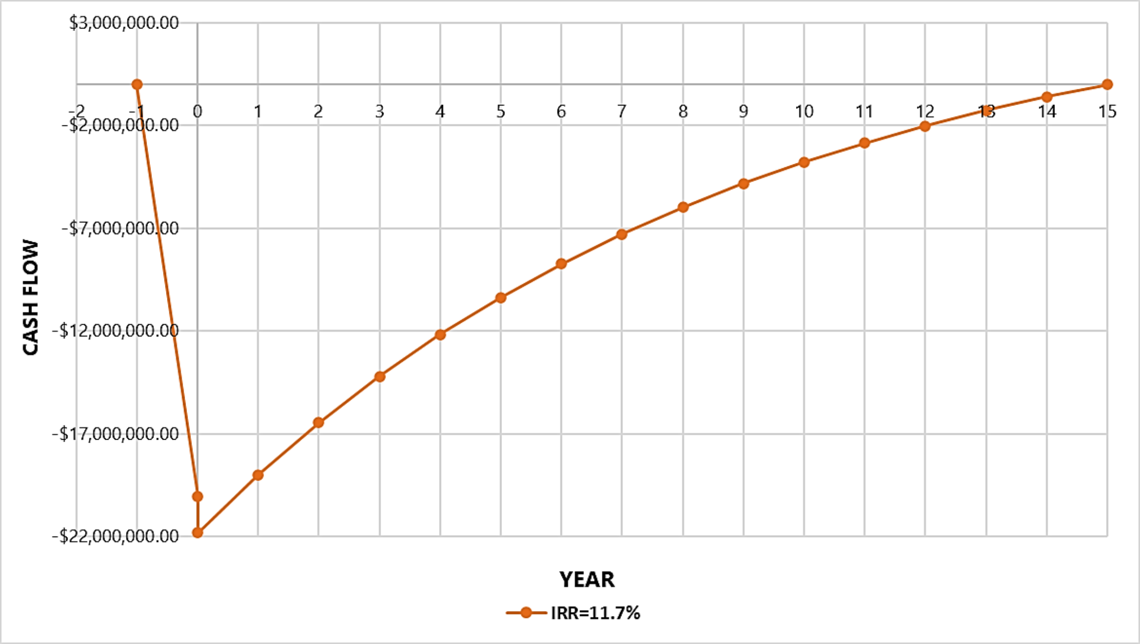

Figure 15. Internal Rate of Return of 11.7% for the Worst-Case scenario.

5.3. Standard-Case scenario.

The economic analysis for the Standard-Case scenario is evaluated with the following information (Table 17) for a base capacity of 120,000 tons citrus waste per year:

Table 17. Economic Analysis information for the Standard-Case scenario.

Information	**Quantity**
Total Fixed Capital Investment	$ 11,835,570.00 [$]
Total Module Cost	$ 13,965,980.00 [$]
Total Grass Roots	$ 20,048,850.00 [$]
Operating Cost	$ 2,617,300.00 [$/y]
Raw Materials	$ 60,000.00 [$/y]
Working Capital	$ 1,775,350.00 [$/y]
Depreciation	$ 1,336,590.00 [$/y]
Revenues	$ 8,097,840.00 [$/y]

The income statement (Table 18) is made by considering 30% of taxation (Mexico) and a discounted rate of 9%. The depreciation period is 15 years, and the construction time is 1 year. Additionally, it is assumed to have 90% of the total sales during the period, and the working hours per year are 8,000 hours.

Table 18. Income Statement of the Standard-Case scenario.

Year	**-1**	**0**	**0**	**1**	**2**	**3**	**4**	**5**
Revenues			0	$ 8,097,840.00	$ 8,097,840.00	$ 8,097,840.00	$ 8,097,840.00	$ 8,097,840.00
Expenses								
Investment			$20,048,850.00	0	0	0	0	0
Operation			0	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00
Feed			0	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00
WC			$ 1,775,350.00	0	0	0	0	0
Total			$ 1,824,200.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00
Profit			-$21,824,200.00	$ 5,420,540.00	$ 5,420,540.00	$ 5,420,540.00	$ 5,420,540.00	$ 5,420,540.00
Depreciation			0	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00
Net Income Before Taxes			-$21,824,200.00	$ 4,083,950.00	$ 4,083,950.00	$ 4,083,950.00	$ 4,083,950.00	$ 4,083,950.00
Taxes			0	$ 1,225,185.00	$ 1,225,185.00	$ 1,225,185.00	$ 1,225,185.00	$ 1,225,185.00
Net Income After Taxes			-$21,824,200.00	$ 2,858,765.00	$ 2,858,765.00	$ 2,858,765.00	$ 2,858,765.00	$ 2,858,765.00
Cash Flow			-$21,824,200.00	$ 4,195,355.00	$ 4,195,355.00	$ 4,195,355.00	$ 4,195,355.00	$ 4,195,355.00
Income Recovery (Discounted CF)	**0**	**-$20,048,850.00**	**-$21,824,200.00**	**-$17,975,250.46**	**-$14,444,104.09**	**-$11,204,520.27**	**-$8,232,425.02**	**-$5,503,732.12**

Year	**0**	**6**	**7**	**8**	**9**	**10**
Revenues	0	$ 8,097,840.00	$ 8,097,840.00	$ 8,097,840.00	$ 8,097,840.00	$ 8,097,840.00
Expenses						
Investment	$20,048,850.00	0	0	0	0	0
Operation	0	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00
Feed	0	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00
WC	$ 1,775,350.00	0	0	0	0	0
Total	$ 1,824,200.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00
Profit	-$21,824,200.00	$ 5,420,540.00	$ 5,420,540.00	$ 5,420,540.00	$ 5,420,540.00	$ 5,420,540.00
Depreciation	0	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00
Net Income	-$21,824,200.00	$ 4,083,950.00	$ 4,083,950.00	$ 4,083,950.00	$ 4,083,950.00	$ 4,083,950.00
Before Taxes						
Taxes	0	$ 1,225,185.00	$ 1,225,185.00	$ 1,225,185.00	$ 1,225,185.00	$ 1,225,185.00
Net Income	-$21,824,200.00	$ 2,858,765.00	$ 2,858,765.00	$ 2,858,765.00	$ 2,858,765.00	$ 2,858,765.00
After Taxes						
Cash Flow	-$21,824,200.00	$ 4,195,355.00	$ 4,195,355.00	$ 4,195,355.00	$ 4,195,355.00	$ 4,195,355.00
Income Recovery						
(Discounted CF)	**-$21,824,200.00**	**-$3,004,179.01**	**-$709,176.16**	**$1,396,331.05**	**$3,327,989.03**	**$5,100,152.32**

Year	**0**	**11**	**12**	**13**	**14**	**15**
Revenues	0	$ 8,097,840.00	$ 8,097,840.00	$ 8,097,840.00	$ 8,097,840.00	$ 8,097,840.00
Expenses						
Investment	$20,048,850.00	0	0	0	0	0
Operation	0	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00
Feed	0	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00
WC	$ 1,775,350.00	0	0	0	0	0
Total	$ 1,824,200.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00
Profit	-$21,824,200.00	$ 5,420,540.00	$ 5,420,540.00	$ 5,420,540.00	$ 5,420,540.00	$ 5,420,540.00
Depreciation	0	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00
Net Income Before Taxes	-$21,824,200.00	$ 4,083,950.00	$ 4,083,950.00	$ 4,083,950.00	$ 4,083,950.00	$ 4,083,950.00
Taxes	0	$ 1,225,185.00	$ 1,225,185.00	$ 1,225,185.00	$ 1,225,185.00	$ 1,225,185.00
Net Income After Taxes	-$21,824,200.00	$ 2,858,765.00	$ 2,858,765.00	$ 2,858,765.00	$ 2,858,765.00	$ 2,858,765.00
Cash Flow	-$21,824,200.00	$ 4,195,355.00	$ 4,195,355.00	$ 4,195,355.00	$ 4,195,355.00	$ 4,195,355.00
Income Recovery (Discounted CF)	**-$21,824,200.00**	**$6,725,992.30**	**$8,217,584.59**	**$9,586,019.81**	**$10,841,464.96**	**$11,993,249.51**

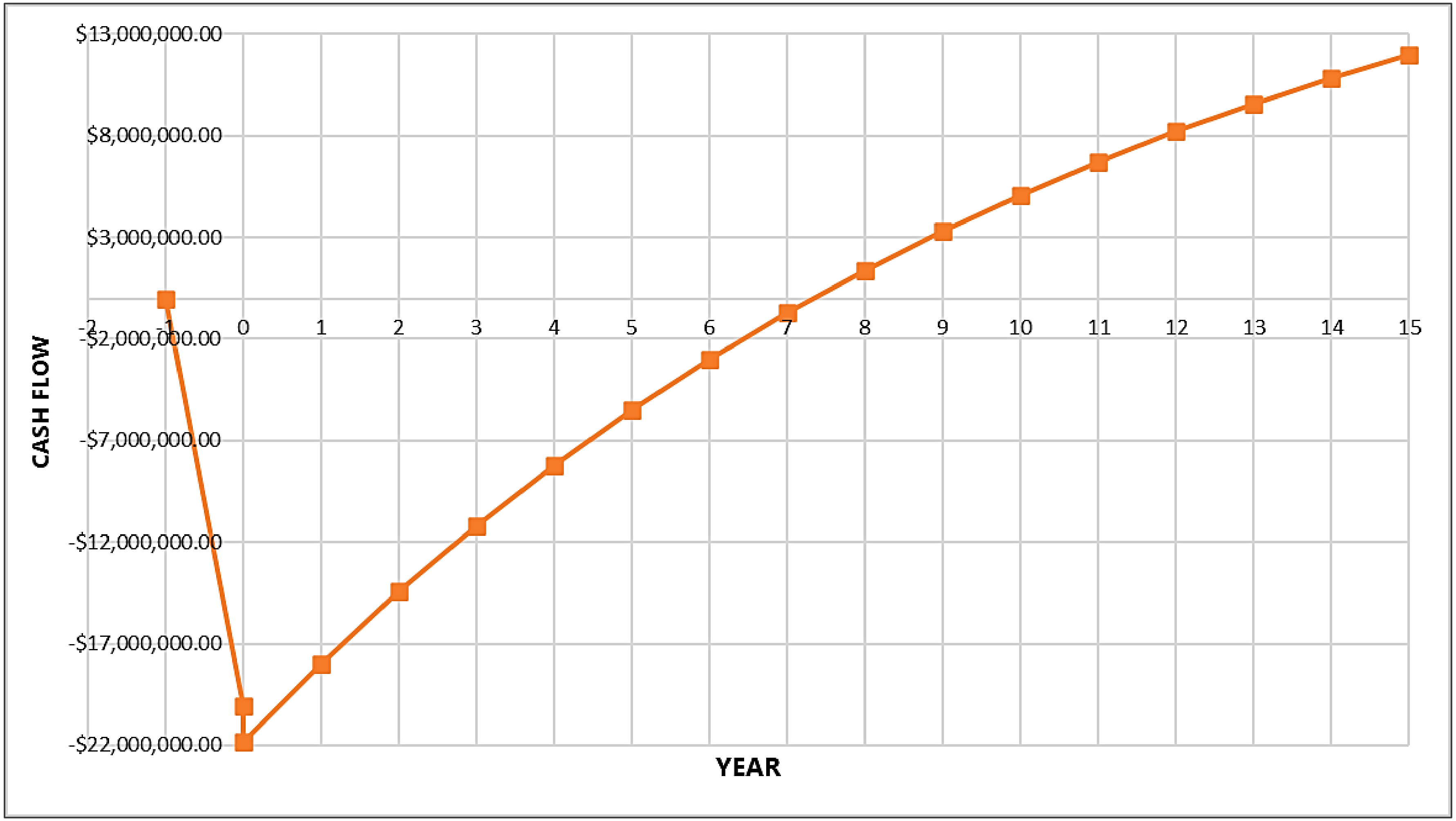

Figure 16. Cash Flow of the Standard-Case scenario.

The Internal Rate of Return (IRR) of the Standard-Case scenario is 17.5%, so this is the maximum discount rate that makes the Net Present Value (NPV) of all cash flows equal to zero in a discounted cash flow analysis (Figure 17).

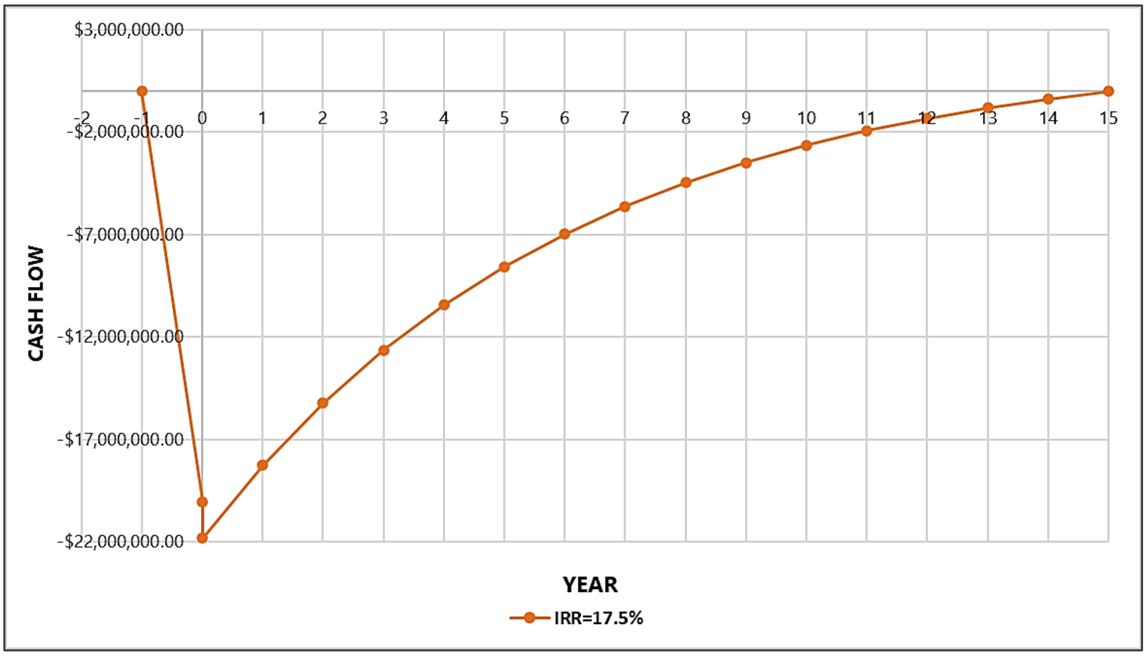

Figure 17. Internal Rate of Return of 17.5% for the Standard-Case scenario.

5.4. Best- Case Scenario.

The economic analysis for the Best-Case scenario is evaluated with the following information (Table 19) for a base capacity of 120,000 tons of citrus waste per year:

Table 19. Economic Analysis information for the Best-Case scenario.

Information	Quantity
Total Fixed Capital Investment	$ 11,835,570.00 [$]
Total Module Cost	$ 13,965,980.00 [$]
Total Grass Roots	$ 20,048,850.00 [$]
Operating Cost	$ 2,617,300.00 [$/y]
Raw Materials	$ 60,000.00 [$/y]
Working Capital	$ 1,775,350.00 [$/y]
Depreciation	$ 1,336,590.00 [$/y]
Revenues	$10,791,360.00 [$/y]

The income statement (Table 20) is made by considering 30% of taxation (Mexico) and a discounted rate of 9%. The depreciation period is 15 years, and the construction time is 1 year. Additionally, it is assumed to have 90% of the total sales during the period, and the working hours per year are 8,000 hours.

Table 20. Income Statement of the Best-Case scenario.

Year	**-1**	**0**	**0**	**1**	**2**	**3**	**4**	**5**
Revenues			0	$ 10,791,360.00	$ 10,791,360.00	$ 10,791,360.00	$ 10,791,360.00	$ 10,791,360.00
Expenses								
Investment			$20,048,850.00	0	0	0	0	0
Operation			0	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00
Feed			0	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00
WC			$ 1,775,350.00	0	0	0	0	0
Total			$ 1,824,200.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00
Profit			-$21,824,200.00	$ 8,114,060.00	$ 8,114,060.00	$ 8,114,060.00	$ 8,114,060.00	$ 8,114,060.00
Depreciation			0	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00
Net Income Before Taxes			-$21,824,200.00	$ 6,777,470.00	$ 6,777,470.00	$ 6,777,470.00	$ 6,777,470.00	$ 6,777,470.00
Taxes			0	$ 2,033,241.00	$ 2,033,241.00	$ 2,033,241.00	$ 2,033,241.00	$ 2,033,241.00
Net Income After Taxes			-$21,824,200.00	$ 4,744,229.00	$ 4,744,229.00	$ 4,744,229.00	$ 4,744,229.00	$ 4,744,229.00
Cash Flow			-$21,824,200.00	$ 6,080,819.00	$ 6,080,819.00	$ 6,080,819.00	$ 6,080,819.00	$ 6,080,819.00
Income Recovery (Discounted CF)	**0**	**-$20,048,850.00**	**-$21,824,200.00**	**-$16,245,466.97**	**-$11,127,363.28**	**-$6,431,855.30**	**-$2,124,049.82**	**$1,828,065.31**

Year	**0**	**6**	**7**	**8**	**9**	**10**
Revenues	0	$ 10,791,360.00	$ 10,791,360.00	$ 10,791,360.00	$ 10,791,360.00	$ 10,791,360.00
Expenses						
Investment	$20,048,850.00	0	0	0	0	0
Operation	0	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00
Feed	0	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00
WC	$ 1,775,350.00	0	0	0	0	0
Total	$ 1,824,200.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00
Profit	-$21,824,200.00	$ 8,114,060.00	$ 8,114,060.00	$ 8,114,060.00	$ 8,114,060.00	$ 8,114,060.00
Depreciation	0	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00
Net Income Before Taxes	-$21,824,200.00	$ 6,777,470.00	$ 6,777,470.00	$ 6,777,470.00	$ 6,777,470.00	$ 6,777,470.00
Taxes	0	$ 2,033,241.00	$ 2,033,241.00	$ 2,033,241.00	$ 2,033,241.00	$ 2,033,241.00
Net Income After Taxes	-$21,824,200.00	$ 4,744,229.00	$ 4,744,229.00	$ 4,744,229.00	$ 4,744,229.00	$ 4,744,229.00
Cash Flow	-$21,824,200.00	$ 6,080,819.00	$ 6,080,819.00	$ 6,080,819.00	$ 6,080,819.00	$ 6,080,819.00
Income Recovery (Discounted CF)	**-$21,824,200.00**	**$5,453,859.00**	**$8,780,275.23**	**$11,832,033.23**	**$14,631,811.22**	**$17,200,414.88**

Year	**0**	**11**	**12**	**13**	**14**	**15**
Revenues	0	$ 10,791,360.00	$ 10,791,360.00	$ 10,791,360.00	$ 10,791,360.00	$ 10,791,360.00
Expenses						
Investment	$20,048,850.00	0	0	0	0	0
Operation	0	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00	$ 2,617,300.00
Feed	0	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00	$ 60,000.00
WC	$ 1,775,350.00	0	0	0	0	0
Total	$ 1,824,200.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00	$ 2,677,300.00
Profit	-$21,824,200.00	$ 8,114,060.00	$ 8,114,060.00	$ 8,114,060.00	$ 8,114,060.00	$ 8,114,060.00
Depreciation	0	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00	$ 1,336,590.00
Net Income Before Taxes	-$21,824,200.00	$ 6,777,470.00	$ 6,777,470.00	$ 6,777,470.00	$ 6,777,470.00	$ 6,777,470.00
Taxes	0	$ 2,033,241.00	$ 2,033,241.00	$ 2,033,241.00	$ 2,033,241.00	$ 2,033,241.00
Net Income After Taxes	-$21,824,200.00	$ 4,744,229.00	$ 4,744,229.00	$ 4,744,229.00	$ 4,744,229.00	$ 4,744,229.00
Cash Flow	-$21,824,200.00	$ 6,080,819.00	$ 6,080,819.00	$ 6,080,819.00	$ 6,080,819.00	$ 6,080,819.00
Income Recovery (Discounted CF)	**-$21,824,200.00**	**$19,556,932.00**	**$21,718,874.32**	**$23,702,307.63**	**$25,521,971.22**	**$27,191,387.36**

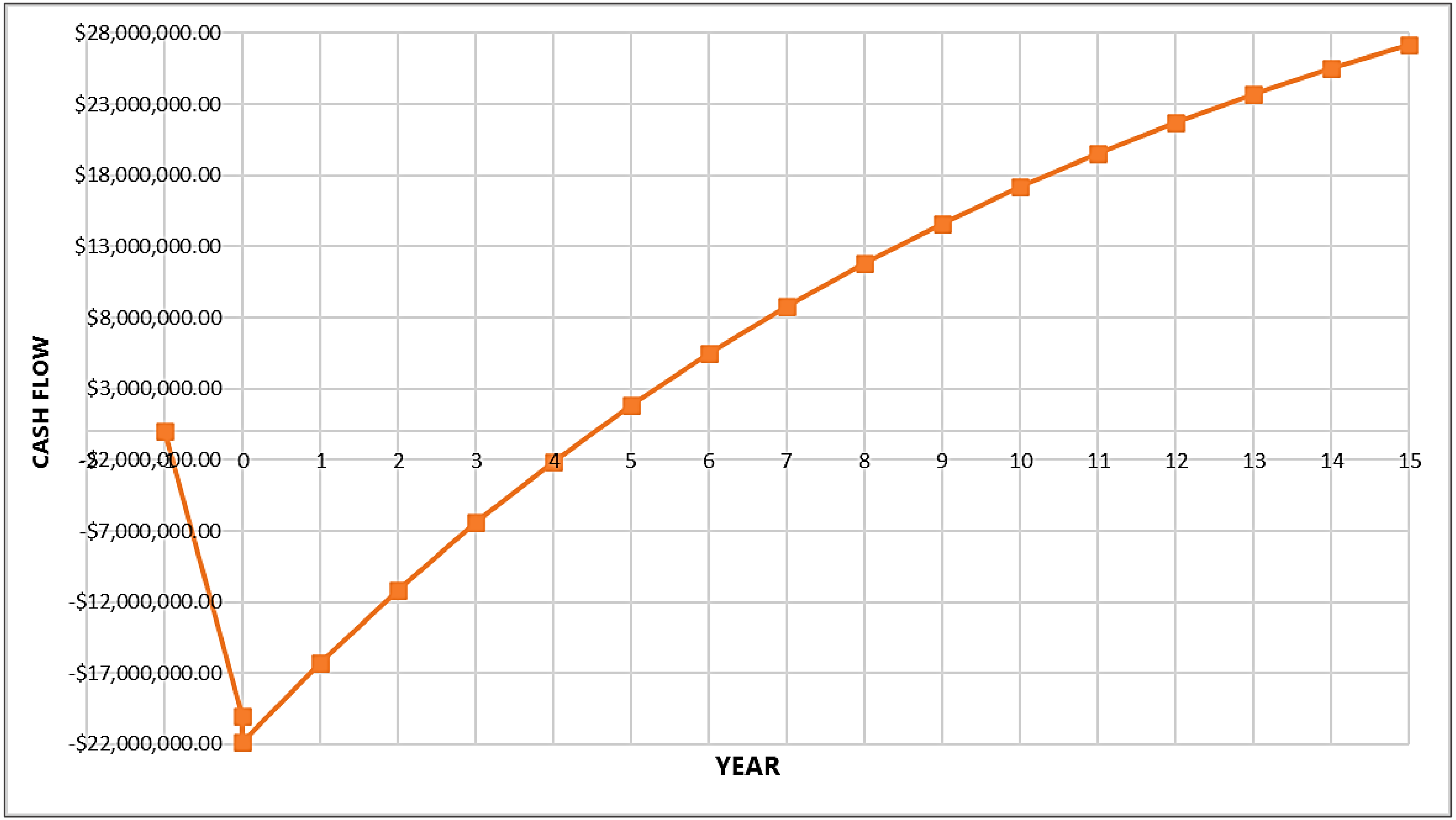

Figure 18. Cash Flow of the Best-Case scenario.

The Internal Rate of Return (IRR) of the Best-Case scenario is 27.1%, so this is the maximum discount rate that makes the Net Present Value (NPV) of all cash flows equal to zero in a discounted cash flow analysis (Figure 19).

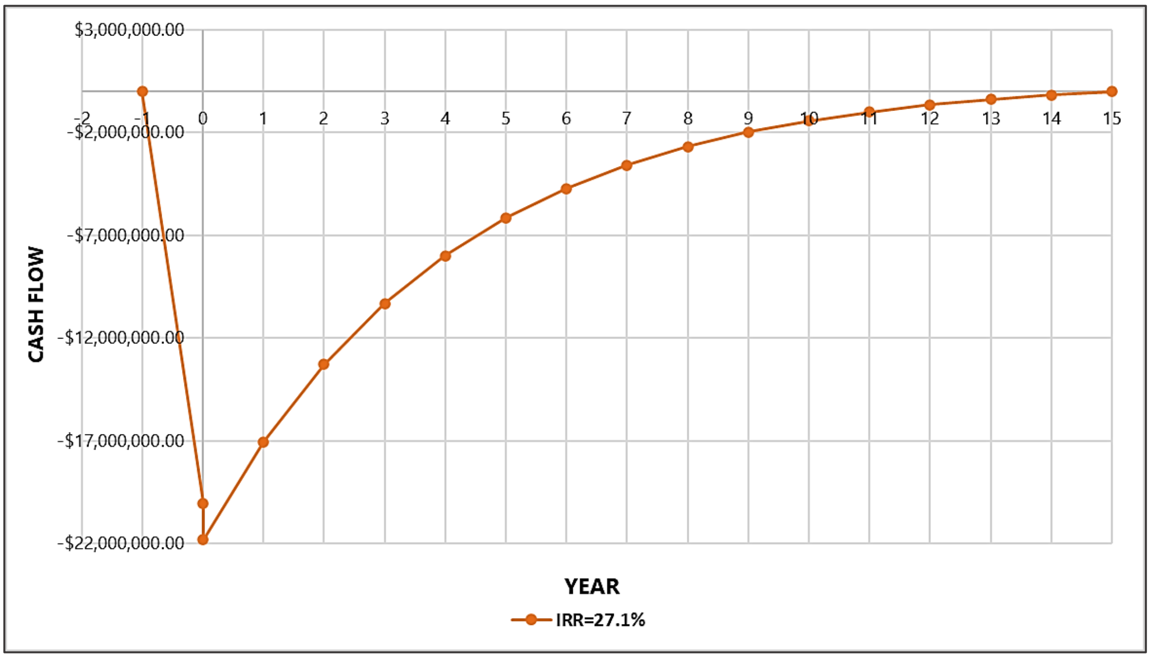

Figure 19. Internal Rate of Return of 27.1% for the Best-Case scenario.

It is clear that the final price and the annual production of the limonene increases the Revenues of the plant, thus the final price of the bioethanol can be decreased to be competitive with the market price of the ethanol and MTBE. The annual production of bioethanol and limonene for each case is shown in the following table 21.

Table 21. Annual production of bioethanol and limonene for each case.

Case Scenario	Bioethanol Production	Limonene Production
Worst-Case	5,520,000 [L/y]	720,000 [L/y]
Standard- Case	6,000,000 [L/y]	787,200 [L/y]
Best-Case	6,480,000 [L/y]	854,400 [L/y]

The economic analysis for each scenario is resumed in the following table 22.

Table 22. Summary of the economic analysis of each case.

Case Scenario	Revenues	Net Income	IRR
Worst-Case	$6,620,400.00	$1,824,557.00	11.7%
Standard- Case	$8,367,840.00	$2,858,765.00	17.5%
Best-Case	$10,791,360.00	$4,744,229.00	27.1%

5.5. The bioethanol production in Mexico.

According to Chapter 2 in table 5, Mexico has a total annual production of 9.87 million tons of fruit for local use, and 42% of this fruit is used in the processing industries (4.15 million tons approx.). Assuming that all the fruit residues used in the processing industries are collected and processed in value-added products, it is possible to estimate a potential maximum amount of bioethanol and limonene production, according to the 3 case scenarios taken into account in this work.

Based on the information in table 6, the total fruit residues generated by the industry are 1.8 million tons per year which 1.39 million tons correspond to citrus residues, and 0.41 million tons come from banana and mango residues. Then, the potential amount of bioethanol and limonene production assuming that all these residues are collected and processed is shown in the following table 23 and the approximation of the economic analysis for each scenario in table 24. Additionally, 15 plants of 120,000 tons/year are required to process these residues.

Table 23. Annual production of bioethanol and limonene for each case, assuming 1.8 million tons processing.

Case Scenario	Bioethanol Production	Limonene Production
Worst-Case	81,800,000 [L/y]	10,425,000 [L/y]
Standard- Case	88,750,000 [L/y]	11,398,000 [L/y]
Best-Case	95,700,000 [L/y]	12,371,000 [L/y]

Table 24. Approximation of the economic analysis of each case.

Case Scenario	Revenues	Net Income	IRR
Worst-Case	$107,540,000.00	$29,627,000.00	11.7%
Standard- Case	$131,121,000.00	$44,790,000.00	17.5%
Best-Case	$174,361,000.00	$76,649,000.00	27.1%

Conclusions

It is clear that Mexico, as one of the main producers of fruit around the world, has considerable potential for the biorefinery of fruit waste. The amount of fruit waste produced every year in Mexico is enough to satisfy the demand of fruit residues that a standard plant of 120 tons per year needs and has a constant production notwithstanding the season.

Based on the process diagram for this plant, the Fermentation process represents more than 60% of the Fixed Capital Investment due to the time needed to carry out the fermentation of the fruit waste, which is about 3 days, so big fermenters are needed to fulfill the amount of fruit residues processed. On the other hand, the Operating Cost comes mainly from two processes, the Steam Explosion, and the Fermentation. The first one because of the total amount of energy required, and the second one due to the price of the enzymes needed.

The evaluation of the 3 scenarios shows how the limonene price affects the feasibility of the plant since the higher the price and annual production of limonene, the higher the Net Income at the end of each year. Additionally, a higher price of limonene permits the decrease of the final price of the bioethanol for it to be competitive with the market price of MTBE.

The maximum amount of bioethanol production that can be reached from the processing of all the fruit waste for industrial use in the Best-Case scenario (95 million liters per year) is still far from the million liters of gasoline consumed per day in a country like Mexico. Thus, the focus of this technology must be the substitution of the MBTE in the petrochemical processes.

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