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Plant Design

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MOBILE BIOMASS HTC-PROCESSING UNIT

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Abstract

Hydrothermal carbonization (HTC) is a process that produces coal-like hydrochar from wet biomass. As some of these biomass resources are in remote locations, in forests or fields far from industrial areas, transporting biomass with high moisture content to a central facility is not always possible or feasible.

A novel approach to this problem is to bring a processing unit to the raw material and process it on site. A mobile process can be shipped to different remote locations to access raw materials that are scarce available only temporarily. As hydrochar is hydrophobic, excess water can easily be expelled from the solid product through mechanical means without the need for drying. Thus excess water does not need to be transported.

A mobile HTC-processing unit was designed to fit into a 40-foot ISO-container that can be transported on the back of a truck. The processing unit uses 3 reactors to process biomass for two hours at a temperature of 220 °C and 3000 kPa pressure. Heat recovery for the process was designed using two flash tanks, a spiral heat exchanger and liquid recirculation. Separation of the liquid and solid products is performed with a filter press. The equipment was sized with the help of Computer Aided Design (CAD). Tinkercad was used to make a 3D model of the process. The major equipment costs were estimated using literature and cost correlations. Other costs of the process, such as installation of major equipment, electrical set up and equipment insulation, were obtained through cost factors based on the cost of major equipment. Thus the final price of the process was obtained as 1.2 million €. The economic feasibility of the process and the main process variables affecting it were studied using two business cases and a sensitivity analysis.

Keywords Hydrothermal carbonization, mobile, process design, equipment sizing, feasibility analysis, sensitivity analysis, biomass, hydrochar

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Tiivistelmä

Märkähiilto (Hydrothermal carbonization, HTC) on prosessi, joka tuottaa biohiiltä märästä biomassasta. Koska osa biomassaresursseista on syrjäisissä paikoissa, metsissä tai pelloilla kaukana teollisista alueista, ei kostean biomassan kuljetus keskitettyyn laitokseen ole usein mahdollista ja kannattavaa.

Uusi lähestymistapa tähän ongelmaan on prosessin tuominen raaka-aineen luo ja sen prosessointi paikan päällä. Siirrettävä prosessilaitos voidaan viedä paikkoihin, joissa prosessoitavia raaka-aineita on vähän tai niitä on saatavilla vain tilapäisesti. Lisäksi ylimääräinen vesi voidaan poistaa hydrofobisesta kiinteästä biohiilituotteesta mekaanisesti eikä tuotetta tarvitse erikseen kuivata. Tällöin ylimääräistä vettä ei tarvitse turhaan kuljettaa.

Tässä työssä kehitettiin siirrettävä HTC-prosessointiyksikkö 40 jalan ISO-konttiin, joka voidaan kuljettaa kuorma-autolla. Yksikössä biomassaa prosessoidaan kolmessa reaktorissa kahden tunnin ajan 220 °C lämpötilassa ja 3000 kPa paineessa. Lämmön talteenotto toteutetaan kahden paisunta-astian, spiraalilämmönsiirtimen ja prosessiveden kierrättämisen avulla. Kiinteä aine erotetaan puristimella. Laitteiden koot määritettiin tietokoneavusteisen suunnittelun (Computer Aided Design, CAD) avulla. Tinkercad-ohjelmistolla luotiin prosessin 3D-malli. Päälaitteiden hinnat arvioitiin kirjallisuuden ja hintakorrelaatioiden avulla. Pienemmät kustannukset, kuten laitteiden asennus, sähkötyöt ja laitteiden eristys, laskettiin hintakertoimilla pohjautuen päälaitteiden kustannuksiin. Prosessin hinnaksi saatiin 1,2 miljoonaa euroa. Prosessin taloudellista kannattavuutta ja tärkeimpiä siihen vaikuttavia tekijöitä tutkittiin kahden liiketoimintamallin sekä herkkyysanalyysin keinoin.

Avainsanat Hydrothermal carbonization, märkähiilto, mobiili, prosessikehitys, laitemitoitus, kannattavuuslaskelma, herkkyysanalyysi, biomassa, hydrochar

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Sammandrag

Hydrothermal carbonization (HTC) är en process som producerar biokol från våt biomassa. Eftersom en stor del av dessa biomassaresurser är vid avlägsna platser, i skogar eller fält långt från industriella områden, är transportering av biomassa med hög fukthalt till en central facilitet inte alltid möjligt eller lönsamt.

En ny infallsvinkel till detta problem är att föra processen tillråmaterialet och behandla den på plats. En mobil process kan transporteras till olika avlägsna platser där råmaterial som är knappa eller tillgängliga bara temporärt. Då biokol är hydrofobt kan man lätt avlägsna överlopps vatten mekaniskt, utan behov av torkning. På detta sätt behövs inte överlopps vatten transporteras.

En mobil HTC-processenhet har utvecklats för att passa in i en 40 fots ISO-container som kan placeras på en lastbil. Enheten använder 3 reaktorer för att bearbeta biomassan under två timmar vid en temperatur på 220 °C och 3000 kPa tryck. Värmeåtervinningen tillämpas genom att använda två expansionsförångningskärl, en spiralvärmeväxlare och återcirkulering av process vätskan. Den fasta produkten separeras från vätska med hjälp av en press. Apparaturens storlek harestimerats med hjälp av datorstödd design (Computer Aided Design, CAD). Programmet Tinkercad användes för att skapa en 3D modell av processen. Kostnaden för huvudsakliga apparaturen uppskattades från litteratur och kostnadskorrelationer. Bikostnader, såsom installation av apparatur, el och apparaturens isolering, beräknades från kostfaktorer på basis av huvudsakliga apparaturens kostnad. Kostnaden för enheten är 1,2 miljoner euro. Den ekonomiska lönsamheten av processen samt de variabler som påverkar lönsamheten mest studerades genom två affärsmodeller och en känslighetsanalys.

Nyckelord Hydrothermal carbonization, mobil, processdesign, dimensionering av apparatur, lönsamhetsanalys, känslighetsanalys, biomassa, biokol

Contents

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Used symbols, abbreviations and definitions

η	Efficiency, dimensionless
μ	Fluid viscosity, Pas
ρ	Fluid density, kg/m^3
A	Heat exchanger surface area, m^2
C	Cost, €
c_p	Specific heat capacity, kJ/(kg·K)
d	Pipe diameter, m
F	Heat exchange correction factor, dimensionless
f_x	Price correction factor according to component x, dimensionless
g	Gravitational acceleration, m/s^2
H	Pump head, m
h	Height, m
i	Interest rate, dimensionless
k	Heat exchange coefficient, W/m^2
L	Pipe length, m
m	Fluid mass, kg
n	Year of project, a
P	Pressure, Pa
$\dot{Q}$	Thermal duty, kW
r	Pipe radius, m
Re	Reynolds number, dimensionless
T	Temperature, °C
t	Time, s
t_i	Insulation thickness, m
ΔT_m	Logarithmic mean temperature difference, °C
$\dot{V}$	Volumetric flow, m^3/s
v	Flow velocity, m/s
Biochar	Solid product from slow pyrolysis process
BSG	Brewer’s spent grain
CF	Cash flow
CSTR	Continuous stirred tank reactor
DCFROR	Discounted cash flow rate of return
HHV	Higher heating value

HMF	Hydroxymethyl furfural
HTC	Hydrothermal carbonization
Hydrochar	The solid product from the HTC process
IRR	Internal rate of return
NPV	Net present value
PFR	Plug flow reactor

Theoretical part

1 Introduction

1.1 Thesis introduction

The general concept of production in chemical industry is quite evident: a plant is built at a feasible location, raw material is brought to the plant and the product is shipped to its designated goal. What if we look past this arbitrary model? What if instead, we have a mobile plant that can be transported to where the raw material is? And what if the product is used on site, then shipping raw material and product back and forth does not seem so feasible anymore?

Dictionary.com defines the adjective “mobile” in seven different ways (Dictionary.com, 2007a), the first on the list being “capable of moving or being moved readily”. But two other definitions are eye catching: “changeable or changing easily in purpose” and “flowing freely, as a liquid”. The second definition makes one think of Taoism and Laozi’s teachings of how water in the river passes all the obstacles it meets, yet is still able to hone the rocks in the river. So mobile can be viewed both as being able to move and to be able to adapt and change easily, while still fulfilling its purpose. Likewise adaptable is defined as “susceptible of modification or adaptation” (Dictionary.com, 2007b). Adapting a process to different environments means adapting it to different raw materials and product grades. Also available utilities vary from location to location. According to the definitions of mobile and flexible, a mobile process should be adaptable and modifiable, with the same core process but different “Lego pieces” for different environments.

1.2 Thesis goals

The objective of this thesis is to investigate the concept of placing a hydrothermal carbonization process (HTC) into a container, thus making it mobile and flexible. This thesis looks into the design of the process as well as assessing the economics of the mobile process in different environments and using different raw materials and analyzing the sensitivity of variables in the process and their effect on the economic feasibility. The idea is to design a HTC process that can be brought to the raw material and be operated at remote locations. It will also look into the possibility of placing the processing unit next to another process, thus utilizing the waste feeds and utilities from the process and moving the mobile unit from process to process. This thesis is part of EU project Mobile Flip and will thus, in accordance with the project, focus on application of the process in Europe.

1.3 Thesis structure

Chapter 1 is the introduction to the thesis, Chapter 2 looks at the background of the HTC process in general, the reaction chemistry of HTC, raw materials, products, applications, Chapter 3 gives background information concerning flexible process design, laws and regulations affecting mobile processes and lists some mobile process in operation to date. Chapter 4 describes the premises for the process, methods used to model the process and sizing of the equipment. Chapter 5 is the feasibility analysis which studies the capital costs, operating costs and profitability of two proposed business cases for the process. Chapter 6 is the sensitivity analysis, where the effects of changing certain variables in the process are studied. Chapter 7 is discussion and conclusion of the thesis.

2 HTC overview

Hydrothermal carbonization, also known as wet torrefaction or wet pyrolysis, is a technology where biomass is treated within subcritical water under pressure to produce coal-like material referred as hydrochar in this thesis. HTC is known for more than a century, the first experiments date back to 1913, conducted by Friedrich Bergius who simulated the natural coalification of organic matter in the laboratory (Bergius, 1913). Bergius received the Nobel prize for his research in 1931. However, this technology did not receive a wide spread interest until at the turn of the century, when interest in artificial coal made from biomass using HTC and other processes like pyrolysis sparked due to several reasons. Potential future applications include soil amendment, energy, fuel cells, carbon sequestration, catalysts, adsorbents and hybrid materials (Titirici and Antonietti, 2010). However, the development of the utilization of hydrochar in value added applications is still at an early stage.

2.1 HTC reaction mechanisms

HTC is a set of simultaneous reactions that are not all fully known to date. The reactions occur in the presence of water under a temperature around 180-250 °C (Reza, 2013). The pressure in the reactor is autonomously generated, corresponding mainly with the vapor pressure of water at the reaction temperature, usually between 2000 and 3000 kPa (Funke and Ziegler, 2010). The pressure is also partly affected by gases formed in the reactions. At higher reaction temperatures water also serves better as a solvent and even non-polar compounds may dissolve better (Funke and Ziegler, 2010).Weakly acidic reaction conditions have shown to improve the reaction rate of HTC (Funke and Ziegler, 2010). While thermochemical reactions such as pyrolysis and torrefaction shun water, HTC requires water to work (Hoekman et al., 2011). This is a huge advantage as raw material drying can be very expensive, especially due to the high enthalpy of vaporization of water. The water functions both as the solvent and reactant. Through the HTC process, the energy density and coal content of the biomass is increased (Nizamuddin et al., 2015). The solid product is an easy to handle, hydrophobic and homogenous hydrochar.

In HTC, the mass ratio between hydrogen and carbon as well as oxygen and carbon decreases in the solid phase (Funke and Ziegler, 2010). The main reactions in HTC include hydrolysis, dehydration, decarboxylation, condensation, polymerization and aromatization (Libra, et al., 2011). However, the detailed reaction pathways are not well known and they highly depend on the type of feed (Funke, et al., 2010). The reaction pathways have been illustrated in Figure 1. Principally, The HTC process is based on the ability of water to break down ionic bonds in other substances. The ionic product, describing the concentrations of ions formed when water undergoes an acid-base reaction with itself, reaches maximum values in the temperature interval 227-327 °C (Reza, 2013). As water breaks down to hydroxide and hydroxonium that dissolve sugars in the raw material. In the HTC reactions with biomass as a raw material, hemicellulose is degraded and cellulose partly degraded. These components degrade when the glycosidic (esters and ethers) bonds between the sugar molecules in the component break through hydrolysis (Funke and Ziegler, 2010; Petrović et al., 2016). Dehydration occurs physically, when the increasingly hydrophobic biomass expels water and chemically, through the elimination of hydroxyl groups. The biomass' hydrophobicity increases as lignin crosslinks and hemicellulose is degraded in the HTC process (Acharjee et al., 2011).

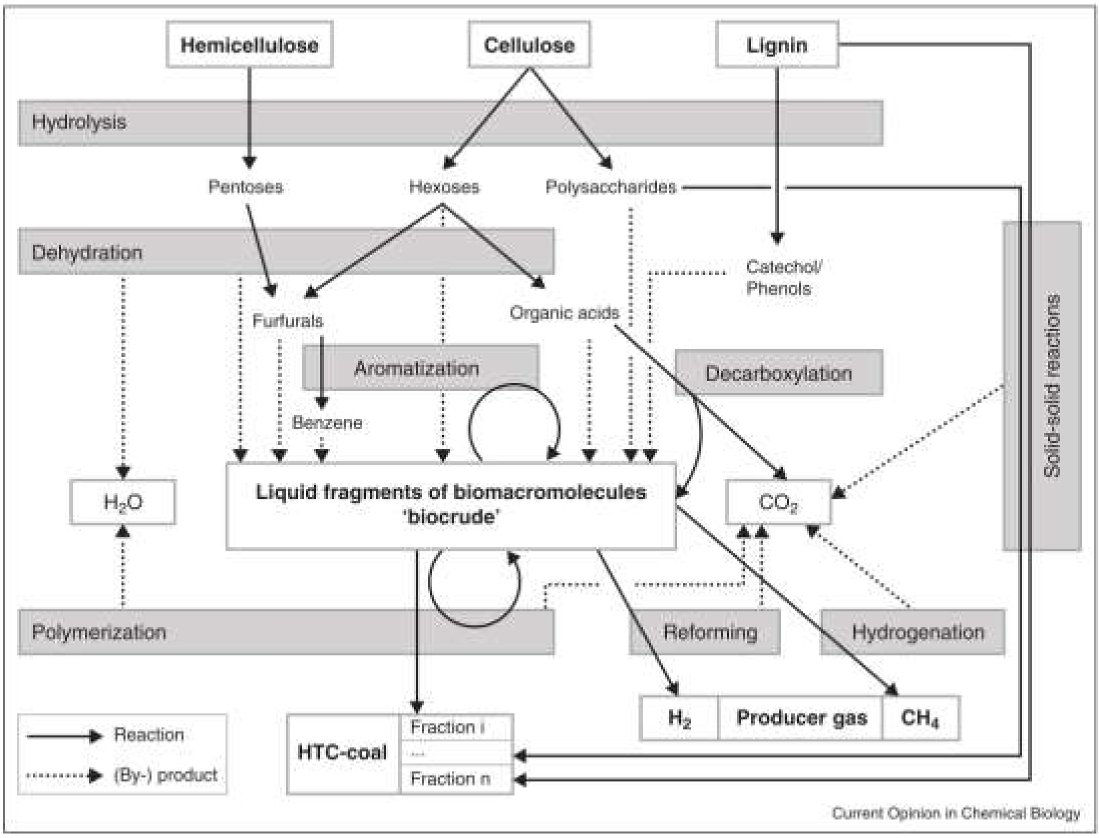

Figure 1. Reaction pathways of hydrothermal carbonization (Kruse et al., 2013)

Cellulose and hemicellulose form aromatic structures under hydrothermal conditions (Funke and Ziegler, 2010). Under acidic HTC conditions, pentose sugars break down mainly into furfural, while hexose sugars on the other hand produce hydroxymethyl furfural (HMF). The furfural intermediates then undergo polycondensation to form hydrochar, the solid product. Aromatization reactions are also presumed to take place during HTC (Funke and Ziegler, 2010). These crosslinked aromatic structures are one of the building blocks in hydrochar and can also be found in natural coal, explaining partly the similar characteristics. Decarboxylation is also part of the HTC reactions, forming carbon dioxide (Child, 2014).

Unlike cellulose and hemicellulose, lignin does not degrade effectively at the HTC reaction temperature 180 °C–250 °C, affecting the final structure of hydrochar (Wikberg et al., 2015b). Lignin is still an attractive option for HTC, as it has a high carbon content and lignin containing side streams are available in the pulp mills and possibly from the production of bioethanol in the future (Wikberg et al., 2015a, 2015b).

2.1.1 Raw materials

Raw materials that have been studied in HTC vary from simple sugars to more complex polymers like cellulose, hemicellulose or lignin to biomass. Examples of biomass that are waste from agriculture, such as unused parts of hay, straw and maize, waste from the food and beverage industry, residues from industrial vegetable oil processing, residues from industrial fruit and vegetable processing, wet forest industry residues and municipal sewage sludge. Organic raw material that is of low value and can be considered waste, preferably having a high moisture content (NIREX, 2013) are ideal for HTC because of the HTC reactions water requirement.

There are few things that limit which raw materials can be used in HTC. Particle size plays a sizable role when the rate of reaction is dependent on diffusion (Reza, 2013). Larger particle size in the reactor feed requires longer residence times and higher temperatures (Child, 2014). Reducing the raw material bodies to a standard size of a few millimeters can be done by shredding or grinding. Higher moisture content in the biomass is not a drawback, as HTC takes place in wet conditions. However, higher moisture content leads to higher heating costs. Different types of biomass contain different amounts of inorganic content. Inorganic compounds enter biomass from the soil in which the plant grows. There are less inorganics in wood-based raw material than in raw material from agriculture or sewage sludge. These inorganics reside in the ash, which can lead to problems in form of slagging, fouling and reduction in efficiency, should the hydrochar be burned (Reza et al., 2013).

A table of raw materials from different sources are shown in Appendix I. The table also lists reaction time and temperature and solid yield as well as ash content and higher heating value where the information is available. The table is divided into three sections: pure saccharide raw material, biomass raw material and waste raw material.

2.2 HTC products

The HTC reactions produce a solid, liquid and gas product. The yields are typically 50-80 % solid product, 5-20 % liquid product and 2-20 % gas product (Child, 2014).

2.2.1 Solid product

The solid product of HTC is hydrochar, a homogenous solid composing of micro-sized carbon spheres and oxygen-containing functional groups (Sevilla and Fuertes, 2009; Titirici and Antonietti, 2010). Hydrochar has higher C content and lower H/C and O/C ratios than the initial biomass. The chemical and physical characteristics of the hydrochar depend on the raw material and the process conditions employed. The carbon spheres size vary depending on the raw material used (Hu et al., 2010). These spheres are hydrophilic at the surface, due to the oxygen containing functional groups, and hydrophobic at the core. Hydrochar has a lower moisture content than the original biomass and is therefore less likely to decay in storage (Acharjee et al., 2011).

2.2.2 Liquid fraction

The process water is the liquid fraction that remains after the filtration of the hydrochar. The liquid fraction contains is a variety of organic degradation products and inorganic compounds dissolved into the process water. A large portion of the organic compounds are formic acid, acetic acid, phenols and furfurals (Reza et al., 2014). The total organic content varies from roughly 8 g/dm^3 to 17 g/dm^3 (Reza et al., 2014b). Some researchers suggest using the organic compounds in the liquid phase for chemical production (Xiao et al., 2012). Reza et al. (2016a) suggest wet air oxidation of the liquid phase, reducing the total organic content by 60 % (Reza et al., 2016a). The research concerning the use or treatment of the liquid fraction is at an early stage, yet could show a future possibility for cleaning the liquid product for discharge into the environment. Wet air oxidation is self-sustained and could possibly release energy from the wastes (Bhargava et al., 2006).

2.2.3 Gaseous fraction

About 10 % of the biomass converts into gas in HTC (Lynam et al., 2011). As a result of the decarboxylation reactions in the process, a large portion of the gaseous product is carbon dioxide. Other compounds in the gas phase include carbon monoxide, hydrogen and methane (Child, 2014). The rate of carboxylation increases with increasing temperature, yielding a higher amount of gases at higher HTC temperatures.

2.3 Use of hydrochar

No large scale market for hydrochar exists to date, leaving the prices of the product to be estimated from the price of biochar from pyrolysis and other hydrothermal processes, though characteristics of the char may vary. For example, biochar from pyrolysis has a larger surface area than hydrochar. The final price depends on the application and the price of biochar has been stated to vary from $0.09/kg in the Philippines to $8.85/kg in the UK, with a mean of $2.65/kg (Ahmed et al., 2016). It is unclear for which application these prices are or from what source. Jirka and Tomlinson, (2014) reported prices ranging from $0.08/kg to $13.48/kg, with the primary source being woody biomass and the foremost application being high-end retail, such as home gardening. 44 % of the biochar produced is made by pyrolysis and only 2 % by hydrothermal carbonization (Jirka and Tomlinson, 2014). The companies selling hydrochar today deal in small amounts (Jirka and Tomlinson, 2014) and an accurate estimate of large scale production revenues cannot be made from these prices. Furthermore, a portion of the companies selling hydrochar today sell them as blends, combined with e.g. compost, clay or soil microorganisms (Jirka and Tomlinson, 2014). The European companies stand for 31 % of the global biochar production and almost half of the biochar is produced in the USA (Jirka and Tomlinson, 2014).

2.3.1 Soil amendment/Fertilizer

Hydrochar can be added to soil in order to improve its functions and reduce greenhouse gas emissions from biomass. The effects of HTC hydrochar as soil amendment has not been well studied and the role of hydrochar as soil amendment depends on the raw material used and reaction conditions, as these affect the composition of hydrochar. Andert and Mumme, (2015) showed that using hydrochar as soil amendment causes changes in microbial activity, presumably due to the easily degradable carbon in the hydrochar. This is in accordance with Rillig *et al.*, (2010), who found hydrochar to increase fungus activity. Some studies suggest that the use of it in the ground might reduce plant growth rate, at least on a short term study (Rillig et al., 2010). This could possibly be due to unknown phytotoxic compounds (Busch et al., 2013).

Hydrochar as fertilizer means releasing nutrients from the hydrochar to the ground. Funke (2015) and Reza *et al.* (2016) stated that ammonia and phosphate is absorbed in the hydrochar in the HTC process and therefore opens the possibility of a slow releasing fertilizer. The liquid product has also been proposed to be used as fertilizer, based on its high phosphorous and nitrogen content (TerraNova, n.d.; Hernandez, 2011).

2.3.2 Renewable energy

Hydrothermal carbonization increases the energy density of the biomass used as raw material. Reza (2013) reported an increase in higher heating value (HHV) of 54 % for corn stover, 37 % for switch grass, 26 % for miscanthus and 21 % for rice hulls.

The higher heating value for hydrochar is around 20-29.5 MJ/kg depending on the elemental composition of the char (Yan et al., 2014; Nizamuddin et al., 2015; Reza et al., 2015; Saqib et al., 2015). Hydrochar can be burned or co-fired directly as fuel after drying. For easier handling or transportation, the hydrochar can be pelletized. Pellets made from HTC hydrochar show excellent binding capabilities and are durable and hydrophobic, increasing storage time. No binding agent is required for the pelletizing of hydrochar (Buttmann, 2011; Yan et al., 2014).

Hydrochar has been proposed for use in energy storage (Wei et al., 2011; Kruse et al., 2013). Hydrochar from hydrothermal carbonization could be used in fuel cells, supercapacitors, battery anodes or cathodes due to characteristics such as high energy density, surface are, portability and stability (Reza et al., 2014). Through activation at a temperature over 600 °C, hydrochar shows superior capacitance over graphite and activated carbon (Tang et al., 2012).

2.3.3 Other uses

Other applications for hydrochars include activated carbons, catalysts, electrodes among others. The large surface area and porosity of certain hydrochars make it an attractive choice as catalysts or adsorbents. The capture of carbon dioxide or methane via hydrochar has also shown promising results (Falco et al., 2013). Hydrochar has been shown to be a promising adsorber of liquids and heavy metals (Demir-Cakan et al., 2009; Liu and Zhang, 2009; Liu et al., 2010; Chen et al., 2011), organic pollutants (Sun et al., 2011; Unur, 2013)and uranium (Kumar et al., 2011; Zhang et al., 2013). This would allow hydrochar to be used in water filtration applications.

Hydrochar has also been further processed to produce materials for carbon aerogels, nanomaterials, sensors and electrodes (Titirici, 2013). One post treatment method is activation of the char to further increase its surface area.

However, the development of the utilization of hydrochar in value added applications of hydrochar is still at an early stage.

2.4 HTC Process

A generalized block diagram of the HTC process can be seen in Figure 2. The raw material is pretreated usually by grinding if necessary. The pretreatment allows for easier handling and pumping of the feed, avoiding blockage (Buttmann, 2011) and a more homogenous reaction, speeding up the hydrolysis in the HTC reactions (Child, 2014). The feed is then mixed with fresh or recycled process water to obtain a desired concentration, preheated and fed in to the reactor. At this point catalysts or other additives are added, if added. The used reaction times are typically from 2 hours to 8 hours (Reza, 2013) at 180 – 250 °C (Wikberg et al., 2015a). The liquid and solid fractions are separated through mechanical dewatering. The hydrochar is then dried and stored. The reactor outlet usually preheats the reactor feed.

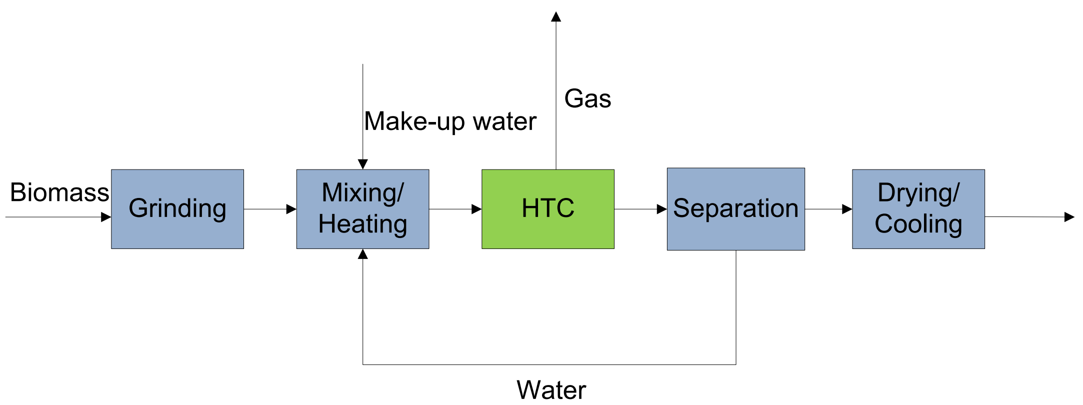

Figure 2. Block diagram of HTC process

The advantages of using HTC are the following: the raw material does not require drying, the process is flexible for raw materials, the hydrochar product is hydrophobic and therefore easy to handle and is an easily scalable technology (Titirici, 2013). HTC is a CO_2 neutral process (Heilmann et al., 2010).

The main focus when designing a HTC process should be on energy efficiency, making the process as economically feasible as possible. As the size of heat exchange-area is limited in a small scale process, it is efficient to use direct heat exchange (Bruins and Sanders, 2012), such as recycling the liquid product in HTC or flash tanks, over traditional heat exchangers. In a small scale process like this, minimizing investment costs is also crucial. Yield and energy densification of hydrochar is largely dependent on temperature. While carbon content and the calorific value of hydrochar increase with increasing HTC temperature, the solid yield drops (Zhao et al., 2014; Mäkelä et al., 2015), reducing energy recovery. An increase in gaseous product can also be noted at higher temperature (Stemann et al., 2013a). Finding the right balance between yield and calorific value is important if the hydrochar should be used for energy purposes. If the application of hydrochar is other than energy, maximum yield should be strived for.

A batch reactor is the simplest and cheapest reactor type. A batch reactor would be easier to handle than a continuous reactor, as feeding can be done by various systems and biomass can be easily varied (Titirici, 2013). Continuous reactors have shown to be more energy efficient than batch reactors (Seifert et al., 2012), increasing heat recovery. The capacity of a continuous reactor is also much higher than that of a batch reactor and there is less down time, improving the economics of the process. A continuous reactor would require a feed system that can work against the pressure in the reactor, making it more complex than the batch reactor. Another option would be to use a semi continuous batch system, where heat from one reactor is used to heat the next batch. However, several smaller reactors are more expensive than one large reactor (Titirici, 2013). The benefits and drawbacks of the reactor types are shown in Table 1.

Table 1. Comparison of different reactor types for the HTC process

Measure	Batch	Continuous	Semi continuous
Investment	Low	Medium	High
Complexity	Low	High	Medium
Flexibility	High	Low	Medium
Economics	Low	High	Medium
Capacity	Low	High	Medium
Heat recovery	Low	High	Medium

Some of the HTC reactions are mildly exothermic such as dehydration and carboxylation being the major reactions taking place (Funke and Ziegler, 2010). No heating of the reactor is usually required after the initial heating, as the reaction sustains itself (Unur, 2013). However, this might be the case for lab scale experiments but a large scale reactor will require sufficient insulation to uphold reactor conditions.

The hydrochar from the reactor is a brown to black liquid with some sediments of hydrochar. This char requires separation of the liquid phase, preferably through filtration. Hydrochar can be dewatered mechanically to a dry matter content of 57 % – 68 % (Ramke et al., 2010). This value can be raised further by recirculating the liquid product (Stemann et al., 2013b)

Recirculation of the liquid phase has shown to improve the higher heating value (HHV) of the hydrochar (Child, 2014) and recirculating the water could be a very interesting option for a mobile process that is in a remote location. Funke and Ziegler (2010) predict that recirculation may lead to higher ash content in the hydrochar. At the same time, valuable organic acids increase in concentration and the total amount of wastewater is reduced (Stemann et al., 2013b). Stemann *et al.* (2013b) showed that these acids reach equilibrium after 5 cycles. Recirculation of the hot liquid product into the cool raw material improves heat recovery.

2.5 Best available technology

HTC processing plants have yet to reach industrial scale, the largest demo plants being on pilot scale, processing around a few thousand tons of biomass annually. The European companies active in HTC process development are largely concentrated on dewatering of wet organic waste (i.e. sludge) to hydrochar for energy recovery. Most of the companies stated here are German. Many of the companies do also research for other biomasses and applications even though the economic viability is based on sludge treatment for energy.

SunCoal® industries have a HTC process called CarboREN. They have patented their process and call their hydrochar SunCoal, which they sell in form of powder, granules or pellets. The biomass (typically straw, wood chips, rice husks or waste from fruit juice or palm oil mills (SunCoal industries, 2016a)) is crushed to a particle size of 60 mm before being separated from impurities such as stones. The feed is mixed with low pressure steam to be heated up and pressurized, before being fed into the continuous reactor. After the reaction the reactor outlet is depressurized in atmospheric conditions. SunCoal industries use a mechanical press as the dewatering method, which gives a hydrochar cake. This cake is then dried in a fluidized bed to give hydrochar in powder form. (SunCoal industries, 2016b)

AVA-CO2 is a German company with a plant has a capacity to process 8400 tons of biomass annually (BusinessWire, 2010). AVA-CO2 uses a multi batch process of modular design (AVA-CO2, 2016). This allows for one reactor to be filled while another is emptied, with heat exchange between them. The reactor operates at 220 °C (BusinessWire, 2010). Flash tanks are used to cool the product and conserve the heat released. Steam is used to heat the feed. Recycling of the liquid phase leads to a carbon recovery of 95 % (Kläusli, 2014). The process allows for the recovery of phosphorus (Kläusli, 2014). AVA Biochem is also part of the AVA group. They recover 5-HMF and 2,5-furandicarboxylic acid (FDCA) from the liquid fraction, through a modified version of the HTC process (Sheridan and Kläusli, 2014).

The TerraNova Ultra process is the most interesting for this thesis, as it is a small scale HTC process that can fit into a container. TerraNova uses a high pressure sludge pump to pump sewage sludge through the heat exchanger to the reactor. The typical moisture content of the input is 70-95 % (TerraNova, n.d.). The reaction temperature is around 200 °C and reaction time around 3 h (TerraNova, n.d.). The reactor heating uses oil as a medium, which is also used in the heat exchanger system. The product is used to preheat the input through a system of tubes with oil as an intermediate. The tubes before and after the CSTR reactor function as tubular reactors, insuring that all sludge reacts (Titirici, 2013). A filter press is used to dewater the hydrochar to a moisture content of 30-35 % (TerraNova, n.d.). The process consumes 72 kWh of heat and 14 kWh of electricity per ton sewage sludge (TerraNova, n.d.). The capacity of the mobile facility is 1500 tons of sewage sludge (dry matter) annually, corresponding municipal waste from 10000 people (TerraNova, n.d.). TerraNova estimates the investment costs for the facility to be under 1.3 million euros (TerraNova, n.d.).

Antaco uses a continuous tube reactor where the water content decreases along the tube. It takes about 4-10 hours for the char to move along the tube reactor. There are no moving parts within the reactor. This process uses about 12-15 % of the energy produced. (Antaco, 2016)

Ingelia SL is a Spanish company that has a HTC plant in Valencia (Hernandez, 2011). It is a continuous plant that uses biomass as a raw material. Commissioned in 2010, the plant is of modular design, meaning flexible operation and the possibility for easy increase in capacity. With an annual capacity of 1100 tons, the plant tests different types of biomass and reaction conditions (Hernandez, 2011).

CarbonSolutions have a HTC process called CS-HTC90™ which can produce hydrochar in only 90 minutes (carbonSolutions, 2013). The process uses two reactors, one of which is countercurrent with recycled steam.

Eurosolid claims to have the first industrial HTC facility to process biomass from plants efficiently and profitably (Eurosolid Group, 2012). They claim an annual production of 8000 tons hydrochar. NIREX is a partner of AVA-CO2 (NIREX, 2013). Artec Biotechnologie GmbH has patented a horizontal HTC process (Zenthöfer, 2015). Loritus has been working on HTC since 2006 and have patented designs that claim to be more energy efficient, easily scalable and give better products than other HTC processes. (Loritus, 2015)

3 Small scale process design

This chapter looks into the theory of small scale process design, the laws that affect mobile processes and gives examples of available mobile processes.

3.1 Flexible process design

As the chemical industry is getting more globalized, Europe is falling behind and must develop new production technologies to stay competitive. Markets in Latin America, the Middle East and Asia are growing very quickly with cheaper labor and raw materials compared to Europe. To stay competitive, new products need to be developed constantly leading to quickly changing markets. This requires new thinking in the process design. Large capacity plants that take years to build will not suffice anymore. To respond to these quickly changing markets, flexible processes are required. Here we will look at some the approaches on flexible process design. (Lier and Grünewald, 2011)

Ideally a mobile process is flexible in capacity, products, feed, innovation and of course location (van Kranenburg et al., n.d.). Today, there is no process that is flexible in all these regards, as investment cost would be too high with current technology (van Kranenburg et al., n.d.). Capacity flexibility means the ability to change the production capacity of the process according to product demand and raw material and utility availability. Product and feed flexibility means the ability to easily switch to another product or raw material, respectively. Innovation flexibility means making the process adaptable to R&D, easily modifiable to try out new products or process settings. A diagram of the different flexibilities can be seen in Figure 3.

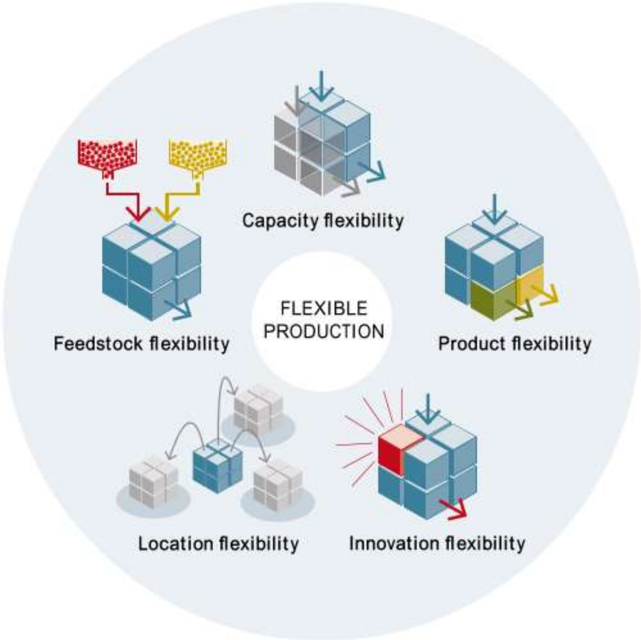

Figure 3. Flexible production concepts (van Kranenburg et al., n.d.)

The implementation of small scale flexible plants allows for the process to be brought to the market faster and thereby fast response to customer demand, especially to changing or emerging markets (van Kranenburg et al., n.d.). Small scale production also lessens the capital costs and economic risk concerning the installation of a new plant. A small scale production close to the market lowers transportation costs and brings the company and customer closer (van Kranenburg et al., n.d.). With small scale on site facilities, raw material that degrades quickly can be preprocessed to a more stable form (through e.g. drying) and stored at the farm, to be transported to the centralized process when needed (Bruins and Sanders, 2012).

St1 is a Finnish energy company. Their Etanolix® process produces bioethanol from industrial waste. The idea is the use of decentralized small scale plants close to the raw material, which is processed into a form that is easier to transport. One facility can produce 5000 m^3 99.8% bioethanol annually (St1, 2016a). As a side stream of the process, high protein animal fodder is obtained. The facility can operate independently or be integrated into another process. Waste streams that can be utilized by the Etanolix®-process are high in sugars, such as bakery residues, retail waste (surplus bread etc.), brewery waste and residues from beverage or confectionary production (St1, 2016b). The Etanolix®-process utilizes excess heat from the process it is integrated to or renewable fuels as energy.

Small scale plants should not be confused with pilot plants, whose purpose is to provide a necessary step between laboratory or bench scale experiments and industrial large scale facilities (Justin, 2014). Pilot facilities are used to obtain information on design, operation and products of the process that can be utilized in the final, large scale facility (Mannan, 2012). The pilot plant gives information about yield, mass and energy balance, pressure drops and mixing (Mannan, 2012). The difference in designing a small scale plant and a pilot plant is that the pilot plant is just an intermediate step in the upscaling process.

Modularity in process design means dividing up parts of the process so that they can be removed or added, depending on the situation, thus making the process more flexible. Parts can be connected in parallel to increase capacity or in series to change unit operations (van Kranenburg et al., n.d.). Nestlé has a blueprint for a modular facility (Eagle, 2014) for simple processes like packaging and mixing dry goods. Different modules are power generators or staff quarters. This facility can be expanded, transformed or moved easily. Envirochemie offers modular water treatment facilities built into containers called EnviModul (Envirochemie, 2014). Projects such as F^3 Factory and COPIRIDE also work towards developing modular chemical plants (F3 Factory, 2014; COPIRIDE, 2016).

Seifert *et al.* (2012) propose a process design concept, where the question is not finding the optimal equipment sizes but fitting the process into different standard sized modules. The capacity of a module would be the module design space and information about modules would be stored in a module database (Seifert et al., 2012). The range of optimization in the module design space is smaller than that of a plant. Optimization of a modular plant requires finding the optimal combination of modules.

Modular plants would bring additional savings through shortened lead time and adaptability to demand, as the capacity can be changed through shutting off or turning on some modules (Lier and Grünewald, 2011). Time to market is also shortened through easier development, faster transportation of the modules and construction of the plant (Lier et al., 2016). Modular plants would also improve testing and decrease downtime, as the plant can continue to operate while one module is tested (Lier and Grünewald, 2011).

A disadvantage using a modular small scale process, contrary to a large scale process, is the loss of economies of scale and through it higher investment costs (Lier et al., 2016). Operational costs may also be higher, as the modules are not specifically designed for a process, leading to higher raw material and energy consumption. An increase in capacity requires more personnel compared to an increase in a high capacity plant.

3.2 Laws & legislations

When applying a mobile process, one must consider the laws and legislations that apply to it. The purpose of this chapter was to find laws or legislations that apply to a mobile facility specifically, more specifically a mobile HTC process. But as none were found, this chapter will take up laws for chemical facilities and laws that affect road transport of containers and the transport of chemicals that apply to the mobile HTC processing unit. Also whenever designing a chemical facility, one must consider the emissions into air, water and land. However, HTC is a process with few hazardous chemicals in the liquid fraction. This chapter will take up directives on an EU scale but also showcase laws specifically for Finland as an example of possible national laws.

This thesis will not look into the laws that affect the products of HTC. When dealing with chemical products in Europe, the compound produced falls under the EU regulation of registration, evaluation, authorization and restriction of chemicals (REACH) (European Commission, 2006). Hydrochar used as soil amendment also falls under the EU regulation on fertilizers (European Commission, 2003a).

3.2.1 Process safety directives

Work safety in a chemical plant falls under the directives on safety and health at work (European Commission, 1989). The directives cover safety in the following areas: workplaces, equipment, signs, personal protective equipment, exposure to chemical agents and chemical safety, exposure to physical hazards, exposure to biological agents, provisions on workload, ergonomical and psychosocial risks and sector specific and worker related provisions. The legislation guarantees minimum safety and health requirements throughout Europe, while national laws can be stricter.

Before building a chemical facility, a Finnish company has to get approval by the Finnish Safety and Chemicals Agency (Tukes). The application sent to Tukes should include information about the company's practice, risk control and how the construction of the facility is executed (Tukes, 2015a). Before start up, Tukes will inspect the facility.

Pressure equipment such as the HTC reactor need to be registered with Tukes before taken into use (Kauppa- ja teollisuusministeriö, 1999). Small scale equipment doesn't need to be registered, e.g. an autoclave whose product of pressure and volume is less than 20000 kPa · L (Kauppa- ja teollisuusministeriö, 1999). Registration information should include the equipment's technical data, manufacturer, importer, owner, end user, overseer and location. The location of the pressure equipment in the plant should be selected so that in the event of an accident minimal danger is proposed and so that the equipment is easy to inspect and maintain. The inside of the equipment should be inspected every 4 years. These inspections can be replaced with thorough follow-up if access to the pressure equipment is difficult due to constructional reasons.

Leaks from piping usually occur at connections to other equipment. The person making these connections, e.g. a welder, must be properly qualified (Tukes, 2015b). When designing piping for a mobile plant, one should make sure that the pipes can handle the conditions of road transport e.g. shaking. The correct material should be chosen for the right circumstances, according to piping standard SFS-EN 13480 (Tukes, 2015b). Before starting the facility, the piping is tested with a pressure test. Liquid is pumped through the pipes at 1.43 times the maximum operating pressure (Tukes, 2015b).

When constructing a chemical facility, special consideration must be put into the location of the plant, to minimize risk should an accident occur. This of course applies to a mobile facility as well. The placement of the mobile process on site should be chosen taking into account distance from housing and people, e.g. a farm or public roads. For the HTC process, the risk is little as the process is simple.

3.2.2 Directives for transport

Before a container can be used for road transport, it has to be checked for proper structural integrity and equipment in accordance with the Finnish container law (Tukes, 1998a). Tukes appoints an inspectorate to approve the container. If the container is seen as unfit, its use can be denied. This also applies for containers that have been modified for a specific purpose (Tukes, 1998b), e.g. a hatch to allow for easy biomass feeding. All containers should have a visible safety chart with the following information: identification code, date of manufacture, maximum weight, and maximum weight when stacked. The container should be inspected every 30 months (Tukes, 1998b).

The size or weight of a container can bring its own limitations to where it can be transported. The most common containers are the 6 m and 12 m standard shipping containers. Their dimensions are listed in Table 2. The European Union has a directive (2015/719) on the largest allowed dimensions and weights for road vehicles (European Commission, 1996, 2015). These limits vary depending on the type of vehicle. Maximum dimensions for a normal transports are (l·w·h) 12m·2.6m·4m (European Commission, 1996). The largest allowed height for road transport in Finland is 4.4 m (Pöyskö et al., 2014).

Some bridges or tunnels might be impassable, as they have their own weight or height restrictions. Low hanging electrical lines on smaller roads also limit vehicle height. Other limitations for road transport derive from poor winter up keep in northern countries and road damage caused by e.g. frost.

Table 2. Measurements of standard shipping containers (Mirasela Oy, 2011; Finncontainers Oy, 2016).

Container	**6 m**	**12 m**
Weight	2300 kg	3750 kg
Volume	33.2 m^3	67.7 m^3
Internal length	5.9 m	12.03 m
Internal height	2.39 m	2.39 m
Internal width	2.35 m	2.35 m
External length	6.05 m	12.2 m
External height	2.59 m	2.59 m
External width	2.44 m	2.44 m

When using road transport, one must account for vehicle tax. The tax varies depending on the carbon dioxide emissions of the vehicle. If no carbon dioxide emissions for the vehicle have been reported by the manufacturer, the weight of the vehicle will be used to calculate the tax. For heavy traffic not using gasoline as fuel, the tax is calculated in accordance with the number of axles on the vehicle and if the vehicle has a trailer or not (Trafi, 2016). The rates are calculated per day for the total weight of the vehicle (for every starting 100 kg).

3.2.3 Directives for emissions and energy taxation

A hydrothermal carbonization process falls under an industrial emission directive. According to EU directive 2010/75/EU on industrial emissions and Finnish environmental protection act (27.6.2014/527) one must know how their operations affect the environment, the risks they present and how to minimize them (European Commission, 2010; Finlex, 2014). This means knowing the effect of the different product streams from HTC on the environment and knowing the effects of hydrochar as soil amendment. For the liquid phase to be discharged on site, its composition and effects on the soil quality or ground water must be known. If no useful components can be found in the liquid phase, it should be treated as waste in accordance with the Finnish waste act (17.6.2011/646) (Finlex, 2011). According to the Finnish environmental protection act and the EU directive 2010/75/EU on industrial emissions, the best available technology should be used and energy consumption should be minimized, implement waste prevention and management and strive to minimize accidents. (European Commission, 2010; Finlex, 2014).

If the hydrochar is produced for energy purposes, the process falls under the energy taxation directive (European Commission, 2003b). The directive defines a minimum tax based on energy production quantities. Production of biofuels can receive tax reductions, as long as the producer is not considered to be overcompensated.

The European Regulation on Registration, Evaluation, Authorization and Restriction of Chemicals (REACH) implies that a producer of chemicals must know and manage the risks connected with their product (European Chemicals Agency, 2016). With a production over 1 ton annually, the producer must be registered with the European chemicals agency.

3.3 Examples of mobile processes

Very little research or patents can be found concerning mobile chemical processes. Much of the information in this section was found from various companies' homepages, with limited technical information. As processes vary in size, this thesis will discuss those that are in scale between truck and large container. As for processes themselves, the thesis examines processes that are related to biomass, chemistry or fuel and contain at least some large equipment. This excludes the "mobile processing units" concerning slaughter or poultry, which basically are trucks in which the animal is slaughtered and the process being a guy with a knife.

3.3.1 Mobile pelletizing

Michigan state university has a mobile pelletizing unit, which is shown in Figure 4. This unit uses air to cool the pellets as they come out of the ring die mill (Michigan State University, 2015a). There is no mention of the capacity of the pelletizing unit.

Amisy, ABC and Azeus machinery all sell small scale mobile pellet plants witch capacities ranging from 100 kg biomass per hour to 500 kg/h (Amisy, 2013; ABC machinery, 2015; Azeus, 2016).

PCM Green energy have a mobile pelletizing plant that fits into a ISO standard container, the MPA1000 (PCM Green Energy, 2016). Using a diesel engine as a power source, the plant has a capacity of 1000 kg/h. Bales of straw are lifted to the top of the container and dropped into a shredder. The pellets can be discharged directly into bags or trailers. The plant is fully automated.

Mobile pelletizing could be used on site together with the mobile HTC processing unit, as no additives are required when making pellets of hydrochar.

Figure 4. Mobile pelletizing unit by Michigan state university (Michigan State University, 2015a)

3.3.2 Water treatment

GE Power provides mobile water treatment solutions (GE Power, 2016), e.g. reverse osmosis, demineralization, water softening, deoxygenation, deionization and filtration. GE power markets that they can deploy water treatment containers in two hours plus travel time, even in the middle of the night. GE Power can treat water up to 454 m^3/h (GE Power & Water, 2011). The system consists of several tanks in a container, with contents depending on the treatment operation.

GE Power also offers outsourcing options called Build, Own, Operate (BOO) contracts (GE Power, 2016). They market this as “letting you focus on your core operations while GE runs your water treatment system”. For large industrial operations there is a possibility for a joint venture where GE has an equity stake. This means no capital investment for the buyer, as GE parks their containers at the plant for as long as needed. With “largest and most comprehensive fleet of mobile water treatment equipment” as back up, GE can guarantee water quality and quantity. GE has 600 mobile units.

GE is just one of the companies that offer mobile water treatment solutions, as en-rionmental-expert.com lists 168 companies connected to mobile water treatment (Environmental Xprt, 2015), but as mobile water treatment is only lightly connected to HTC, this thesis will not look further into the matter.

3.3.3 Agriculture

Cassava is a root grown in Africa. Cassava does not have a season and grows all year. Due to the root degrading on the way to food processing facilities in urban areas, farmers are unable to generate income from cassava (DADTCO, n.d.). The Dutch Agricultural Development & Trading Company (DADTCO) designed an Autonomous Mobile Processing Unit (AMPU) to be taken to the cassava farmers. AMPU is stored in a container and processes the cassava onto a cake that can be stored for up to two years. The AMPU can handle 5 tons of cassava per hour (DADTCO, 2016). The roots are washed peeled and grated, after which mechanical drying removes 50% of the water (Dixon, 2015). The water extracted by the process, including minerals, is sprinkled back into the land, improving soil fertility (Bruins and Sanders, 2012).

The cassava cakes are used for the Impala beer, brewed by SABMiller. The cassava cake could also be used to make flour, lowering expensive wheat import costs in Africa (DADTCO, n.d.), or cassava based starch (Dixon, 2015). The AMPU is moved to different areas depending on harvesting schedules. This is important considering that cassavas degrade in two days (DADTCO, n.d.). The AMPU usually remains at site for 3-4 months (DADTCO, 2016).

Alvan Blanch offer mobile crop dyers. These dryers improve fuel efficiency by recirculating the hot drying air and have sound isolation to reduce noise (Alvan Blanch, n.d.). Instead of high temperature the system uses high airflow, making it suitable for crops that require lower temperature e.g. malting barley and seed grain. The mobile dryers have a capacity of 25 tons per hour (Alvan Blanch, n.d.).

Alvan Blanch also offers mobile fruit processing solutions (Alvan Blanch, 2012).These mobile systems allow collection of fruit from large rural areas in Africa that otherwise go commercially unexploited. A mill pulverizes the fruit and a mechanical press squeezes out the juice. The juice is blended with sugar and/or water in blending tanks whereafter it is pasteurized and bottled. Multi-stage filtration for water allows local water to be added into the juice.

3.3.4 PCB

Polychlorinated Biphenyls (PCBs) are used as dielectric liquid in transformers, heat transfer liquid, hydraulic fluid, lubricant and additives in pesticides and paints (Gonzalez, 2004). Kinetrics is one of many companies to have mobile processes for PCB removal (Kinetrics, 2011). These mobile units can handle 15 m^3 of transformer oil per day (Vijgen and McDowall, n.d.). The PCBs are degraded by adding an alkali reagent to the reactor, where the PCBs are destroyed irreversibly (Vijgen and McDowall, n.d.).

In Canada, mobile PCB treatment units are installed into 12 m containers. Smaller models operate batch reactors of 3000 liters and larger models apply a continuous flow system (Vijgen and McDowall, n.d.). The required power for heating the oil is about 220 kW (Vijgen and McDowall, n.d.). The units are run by two technicians. As the reagent degrades in contact with water, precautions must be taken to prevent said contact with e.g. rain water.

3.3.5 Energy

GE's TM2500+ is a mobile power generator offering up to 31 MW of power (GE Power & Water, 2014). It uses gas or distillate as fuel or a combination of these and can be commissioned in 11 days. The TM2500+ is mounted on a 2-trailer assembly. Several units can be connected if need be. The TM2500+ could be a viable option for electricity generation for the mobile HTC process.

Michigan state university has a small scale mobile biodiesel unit for demonstration purposes. This unit can produce a batch of 303 dm^3 biodiesel (Michigan State University, 2015b). This process is mounted in the back of a 5-meter cargo truck. The reaction takes 2.5 hours and up to 48 hours is required for glycerin to settle in the tank (Michigan State University, 2015b). The unit does not however contain any means for oil extraction or cleaning. Greg Mowry has also developed a mobile bio-diesel processing unit (Mowry, 2011). It can be attached to the back of a pick-up truck and has an annual capacity of 15000 dm^3 biodiesel, operating at 13-40 MPa and 300-400 °C.

Similarly to HTC, mobile fast pyrolysis aims to increase the energy density of bio-mass on site for more feasible transportation (Badger and Fransham, 2006). Bio oil produced from fast pyrolysis can easily be transported in tanker truck. Also syngas or char can be produced (Coleman and Mcdonald, n.d.). Mainstream engineering have a mobile fast pyrolysis process for military and commercial use (Amundsen and Yelvington, 2015). The causes for military application are the dangers of trans-porting fuel in warzones and the hazards of burning waste near a camp. The process unit is mounted in a Tricon container, which dimensions are 2.44x2.44x1.98 meters (Amundsen and Yelvington, 2015). They are usually deployed with another Tricon container which is for waste pretreatment. The pilot reactor has a capacity of 1 ton/day (Amundsen and Yelvington, 2015).

3.3.6 Other mobile processes

Mossé *et al.* (2012) developed a mobile plasma plant for waste treatment. The idea is that accumulation of waste in processes is periodic and therefore the leasing of a mobile treatment unit is the most viable option for customers. The unit can use power from the grid or from diesel generators. The temperature of the plasma jet is 3500-5000 °C and the combustion of waste occurs at 1200-1800 °C, depending on the wastes composition. At a reaction time of 2 seconds, the waste is 99.9 % decomposed (Mossé et al., 2012). Water or air can be used to cool the waste after the plasma.

In 1973, Canfield and Nafziger filed a patent for a mobile milk processing system. Using the transport vehicle as a power source, the processing unit could pasteurize the milk on site. Thus, the farmer could process the milk himself. Building on the same concept other patents have been filed: a mobile pasteurizing unit (Dumm, 2001), a mobile cheese and milk production unit (Thuli et al., 2007) and a milk concentration system through reverse osmosis (Smith, 2010).

Axiom and Environmental Drilling Solutions (EDS) offer mobile shale shakers that separate solids from expensive process fluids in drilling operations (Mazerov, 2010; Axiom, 2013). The Mobile Process Units can be transported on all roads.

SINTEF has a mobile unit for processing fish rest into high quality omega-3 fish oils and protein hydrolyzation (SINTEF, n.d.). The mobile SeaLab has a reactor of 800 dm^3 and capacity of 500 kg raw material per hour (Storrø, n.d.).

Applied part

4 HTC-processing unit design

The basis of the HTC process was explained in chapter 2.4. This chapter will focus on the design of the mobile HTC unit. The design part will limit itself to the process within the confinements of the container. Possible pretreatment of the biomass or further drying is not considered in the design.

4.1 Design specifications

4.1.1 Raw materials

The raw materials studied in this thesis are brewers spent grain (BSG) and wheat straw. In accordance with the project, this thesis will focus on raw materials in Europe.

Brewer’s spent grain is a by-product from beer production, from the mashing process. In the mashing process, sugars, proteins and nutrients are extracted from the grain and the leftover is BSG (Witkiewicz, n.d.). The grains used can be for example barley, maize or wheat. Typically, brewer’s spent grain is used as cattle feed. The problems with this application is incomplete degradation in the cattle’s digestive system and cyclic generation (Poerschmann et al., 2015). Thus, BSG is not a profitable product. With a mobile HTC unit, BSG could be collected from microbreweries and utilized on site. The largest facility in Europe, Obolon, produces 120 million dm^3 of beer annually (Obolon, n.d.). Some of Europe’s breweries are listed in Appendix 2. Typically, for every hectoliter of beer produced, 15-20 kg of BSG is formed, which has a dry matter content of 20-30 %.

Europe is the largest producer of wheat. The wheat production in Europe is 132 billion kg annually, which leads to 48 billion kg of residues. Roughly 170 million kg of this is wheat straw (Montane et al., 1998). The number of farms specialized on wheat production lies between 19300 and 113000, depending on the year (European Commission - Agriculture and Rural Development, 2013). The farm sizes vary roughly from 20 ha to 200 ha, with the average being 100 ha and 50 % of the area used to grow wheat. The average yield for specialized wheat farms is 5-5.7 ton/ha (European Commission - Agriculture and Rural Development, 2013). For every kilogram of wheat produced, 1.3 kg of straw is formed (Montane et al., 1998). The moisture content of wheat straw is 8-23 % (O'Dogherty et al., 1995a).

While addition of acids has been suggested to catalyze the HTC reaction (Lynam et al., 2011; Stemann et al., 2013b), the use of these would increase the operating costs, therefore diminishing the feasibility of the process. If the process water is recycled, organic acids formed in the process are recycled to the feed, thus catalyzing the reaction. The acids are in this work presumed to catalyze the reaction enough to produce a sufficient quality product.

4.1.2 Methods

The process was simulated with Balas process simulation software (http://balas.vtt.fi/) in conjuncture to Microsoft Visio diagramming software to achieve the mass and energy balances. The versions used were 3.2 and 2007, respectively.

The 3D model, to better give an understanding of equipment size, was designed with Tinkercad, a free online CAD editor (www.tinkercad.com). While Tinkercad is preliminarily designed for 3D printing of objects, it was sufficient in measuring the equipment sizes and helping the author picture the design in the limited space. The 3D model is presented in Figure 5.

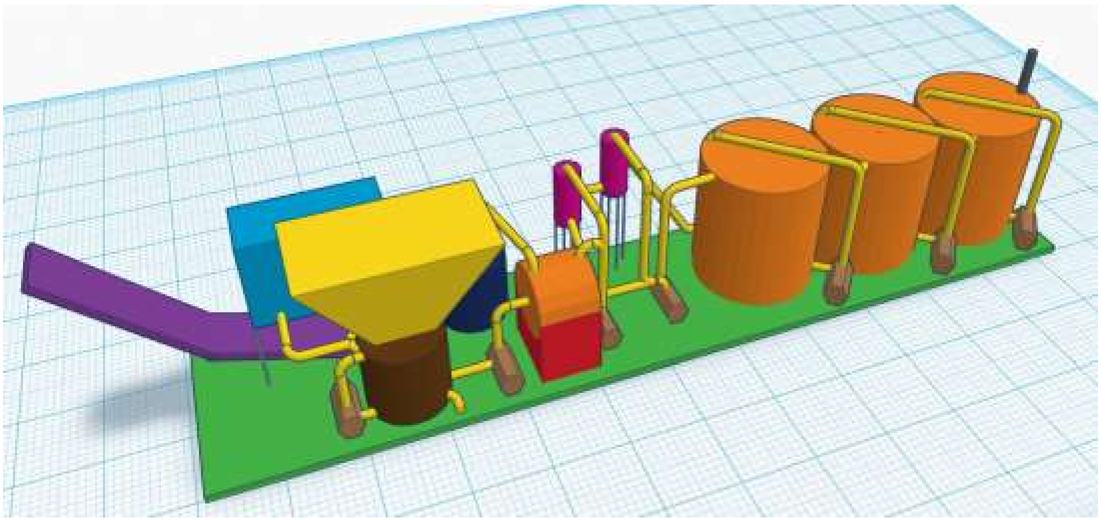

Figure 5. 3D model of the process. From left: conveyor (purple), filter press (blue), mixer (yellow and brown), buffer tank (dark blue), heat exchanger (orange and red), flash tanks (pink) and reactors (orange). Pumps are light brown and piping yellow. The exhaust pipe in reactor three is black.

Because of the space limit of a mobile process, the approach in this thesis was iterative: simulating the process, sizing the equipment, modelling the process with Tinkercad and adjusting the simulation model again to fit the space limits. This process was repeated several times. The goal was to achieve maximum capacity within the confinements of the mobile process, regardless of the availability of the raw materials. The HTC process is focused around the reactor, without any solvent besides water and relatively easy solid-liquid separation. The key factor in making the process feasible is sufficient energy recovery.

4.1.3 Process premises

The process was modelled for a feed of brewers spent grain. The solid content of the process feed was assumed to be 25 % in accordance with the BSG dry matter. The feed to the reactor was assumed to have a dry matter content of 10 %, for easy handling in the reactors. The amount of liquid recycled in the process is calculated accordingly.

The reaction conditions were assumed to be 220 °C, 3000 kPa and a reaction time of 2 hours. The conditions were presumed to be adequate to produce high quality hydrochar, both for energy and further processing. Assumptions were made because limited amount of data is found on the correlation of process conditions and product quality for the selected raw materials. HTC of biomass is a very complex process with very varying results depending on the raw materials and process conditions implemented. When further research on the subject has been conducted, the process can be adjusted accordingly.

The final product was assumed to have a moisture content of 30 %. A high moisture content would lead to unnecessary costs, due to the transport of water. If the product is too dry, transport is hampered by dust formation. A slightly moist product is a sticky mass and can be transported in an open flatbed. A 30 % moisture can be achieved with a filter press (Klima et al., 2012) or other equally efficient mechanical drying methods. A concept would be to perform further drying at a central location, processing products from several mobile HTC containers. Depending on the application of the hydrochar, further drying might not be required. 30 % moisture might be sufficient for e.g. soil amendment purposes.

Reactor energy loss was assumed to be 8 kW, in accordance with (Stemann et al., 2013a), which had similar reactor conditions. The energy losses in the pipe were assumed to be small and are not accounted for in the simulation model. The energy loss of the buffer tank, mixing tank and flash separators were calculated using the surface area and energy loss factors given in (Pentair Thermal Management Solutions, 2013). The temperature in the container is assumed to be 25 °C. The energy loss was assumed to be small for all tanks, 1 kW.

Softwood was used to model biomass, coal with low sulfur to model hydrochar, acetic acid to model the liquid product and carbon dioxide to model the gaseous product. The conversion was assumed to be 100 %. The yields were assumed to be the following: hydrochar 50 %, liquid product 35 % (and gaseous product 15 %.

Appendix 3 shows a list of all assumptions made in the process design and the feasibility analysis.

4.2 Process description

The model of the mobile process appears in Figure 6. The equipment is listed in Table 3. The biomass gets mixed with recycled liquid in the mixer FA-1, whilst the wastewater is fed through the mixers mantle, recovering energy. The mixing is done with pump GA-1. The slurry is then pumped with pump GA-2 through the heat exchanger EA-1 whereafter steam from flash tank FA-3 is inserted. The pressure is further raised with pump GA-3 and steam is inserted from tank FA-2. The pressure is raised to reactor pressure, 3000 kPa, with pump GA-4. The temperature is elevated through electric resistance EA-2 to reactor temperature, 220 °C. The biomass reacts in the three reactors DC-1, DC-2 and DC-3, while being mixed with pumps GA-5, GA-6 and GA-7. These pumps also transport part of the slurry to the next reactor. The mean residence time is 2 hours. Gas product is removed from reactor DC-3. The slurry then travels to tank FA-2 where steam is removed when the pressure drops. The pressure is further dropped in FA-3. After this the product travels through EA-1 to the buffer tank FA-4. GA-8 mixes the product. When the filter press HA-1 is empty the valve between it and the buffer tank turns, redirecting flow to HA-1. HA-1 then separates the solid product from the liquid, which is recycled partly to tank FA-1 and partly removed through the mantle of FA-1.

Stream	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Total flow kg/s	0.45238	0.97159	0.97159	1.12438	1.14646	1.10407	0.02208	1.0821	0.1528	0.9293	0.9293	0.08049	0.84881	0.3296	0.51921
Temperature C	25	62.5972	70.3996	111.678	203.131	217.432	209.877	209.877	138.884	138.884	90	89.35	89.35	89.35	89.35
Pressure kPa	101	101	150	150	1800	2894	2000	2000	350	350	101	101	101	101	101
Dry content %	25	11.7278	11.7278	10.1357	10.0073	5.50988	0	5.38705	0	6.21541	6.21541	70.0291	0.16384	0.16384	0.16384
Biomass flow kg/s	0.11309	0.11309	0.11309	0.11309	0.11309	0	0	0	0	0	0	0	0	0	0
Water flow kg/s	0.33928	0.85764	0.85764	1.0099	1.02945	1.04335	0.01954	1.0238	0.15226	0.87154	0.87154	0.02412	0.84742	0.32906	0.51836
CO2 flow kg/s	0	9E-07	9E-07	0.00046	0.00299	0.00299	0.00253	0.00046	0.00046	1.5E-06	1.5E-06	4.2E-08	1.5E-06	5.7E-07	9E-07
Acid flow kg/s	0	0.0005	0.0005	0.00058	0.00059	0.00093	8.8E-06	0.00092	7.8E-05	0.00084	0.00084	2.3E-05	0.00082	0.00032	0.0005
Hydrochar flow kg/s	0	0.00035	0.00035	0.00035	0.00035	0.05691	0	0.05691	0	0.05691	0.05691	0.05635	0.00057	0.00022	0.00035

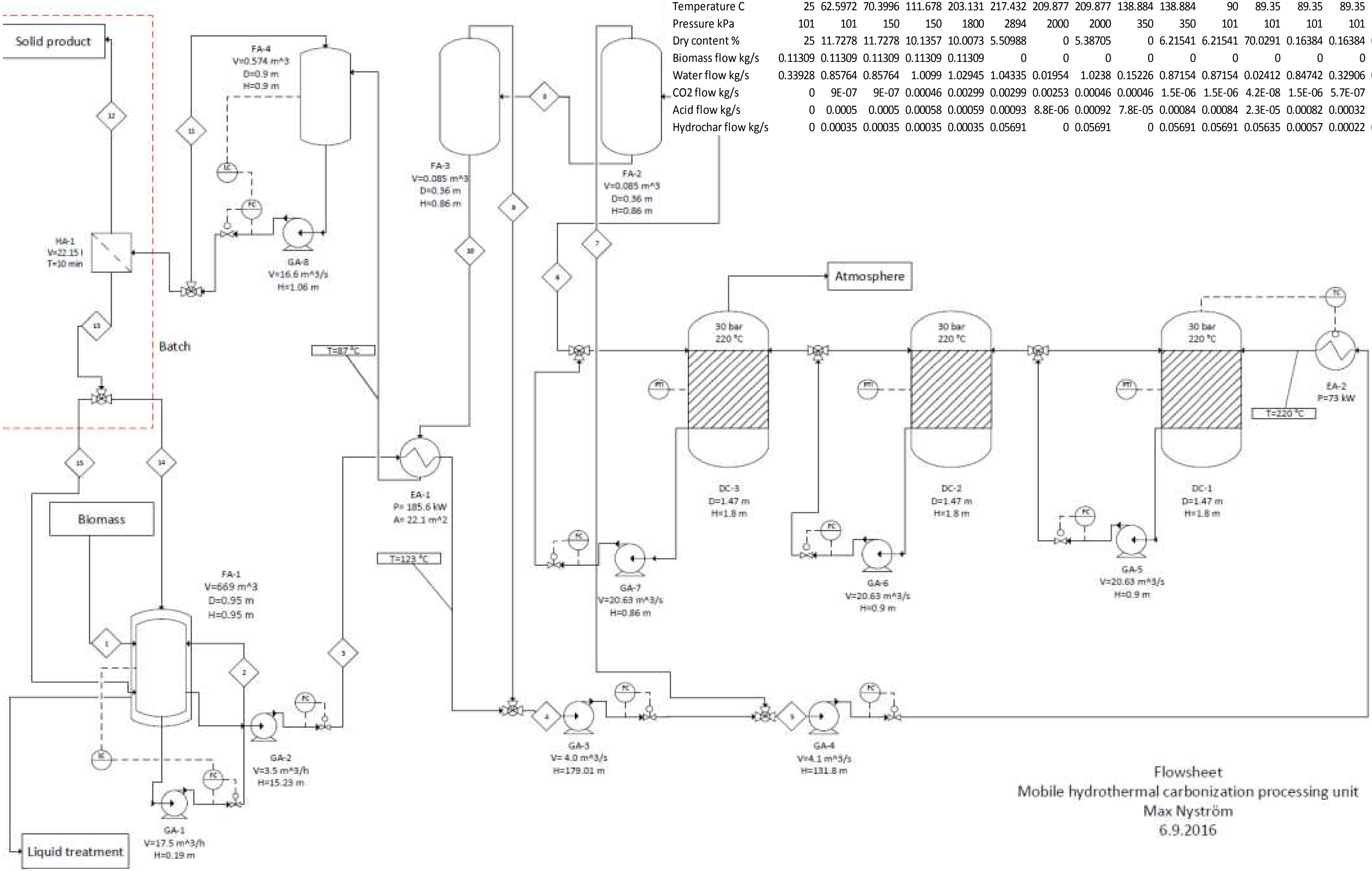

Figure 6. Process flow diagram and stream chart.

Table 3. Equipment numbers for major equipment in the process

Equipment	Description
FA-1	Mixer tank
FA-2	Flash tank 1
FA-3	Flash tank 2
FA-4	Buffer tank
EA-1	Heat exchanger
EA-2	Electric heating
HA-1	Filter press
DC-1	Reactor 1
DC-2	Reactor 2
DC-3	Reactor 3
GA-1	Pump 1
GA-2	Pump 2
GA-3	Pump 3
GA-4	Pump 4
GA-5	Pump 5
GA-6	Pump 6
GA-7	Pump 7
GA-8	Pump 8

4.3 Equipment selection & sizing

The sizing was done simultaneously with the simulation and 3D design. The size of the equipment was readjusted and the capacity of the process was changed accordingly.

4.3.1 Shell

When designing this process, the major limiting factor is the boundaries brought by the shell of the process, namely the shipping container. While this is not an absolute restriction, the utilization of a standardized container as the shell, in contrast to a custom casing, could substantially lower the investment costs. Standardized containers are also approved for all roads. The measurements for a standardized 40-foot ISO-shipping container can be found in Table 2. The internal volume of this container is 67.6 m^3. For easy access, e.g. for maintenance, a container with removable walls would be the optimal solution. Alternatively, one with several doors would suffice. As space is limited and the equipment takes up most of the container, access from several directions is important for maintenance purposes.

4.3.2 Reactors DC-1 DC2 DC-3

For a process with limited space, a plug flow or tubular reactor (PFR) is most optimal. PFRs demand less space than continuous stirred tank reactors (CSTR) (Wittrup, 2007). The temperature may be harder to control in a PFR than a CSTR, but temperature fluctuations are not critical in the production of hydrochar. The product is also harder to standardize, due to irregularities in process flow (Buttmann and Bottinger, 2009). For applications such as energy and soil amendment, these fluctuations in the product are not critical. A disadvantage with tubular reactors is the pressure drop in large volume reactors. The decisive factor when choosing a reactor for slurries is however the critical velocity. When travelling in a pipe, the slurry must move over a certain velocity to prevent the particles from settling (Total pump group, 2010). For coarse particles like shredded biomass (roughly 1 mm diameter particles) the critical velocity is around 0.62 m/s (Quiroga et al., 2013). This value was simulated for 10 vol-% sugarcane bagasse suspension, and is assumed to correlate with the biomass slurry in the HTC process. With a 2-hour reaction time, the tubular reactor would need to be 2200 m long. Therefore, it can be conducted that a tubular reactor is not feasible with the assumed reaction time. Because of the rectangular shape of the container, three CSTRs were selected, placed in series. This allows for maximum capacity, but also requires several pumps.

The size of the reactors was determined in conjunction with the Balas-simulation and the size of the other equipment. From the simulation it was determined that the flow through the reactors is 1.14 kg/s. In lack of better data, the density of the slurry is assumed to be 1000 kg/m^3, based on a simple test done with BSG at VTT. With a 2-hour reaction time and 90 % fill rate, the required volume of the reactor is 9.17 m^3. Keeping the height of the reactors limited to 1.8 meters, the required internal diameter of the reactors is 1.47 m. The required wall thickness is calculated from Figure 7. While the material used for the chart in Figure 7 is carbon steel, it can be assumed that the wall thickness requirement is the same for stainless steel. The values used are 155 cm and 3000 kPa, resulting in a wall thickness of 1.82 cm. The insulation thickness of the reactors is assumed to be 5 cm, resulting in a total diameter of 1.61 m and height of 1.94 m. The insulation thickness was estimated from the equation

$$t_i = \left(\frac{T}{150}\right) - 1 \qquad (1)$$

where t_i is insulation thickness in inches and T temperature in Fahrenheit (Nauman, 1987).

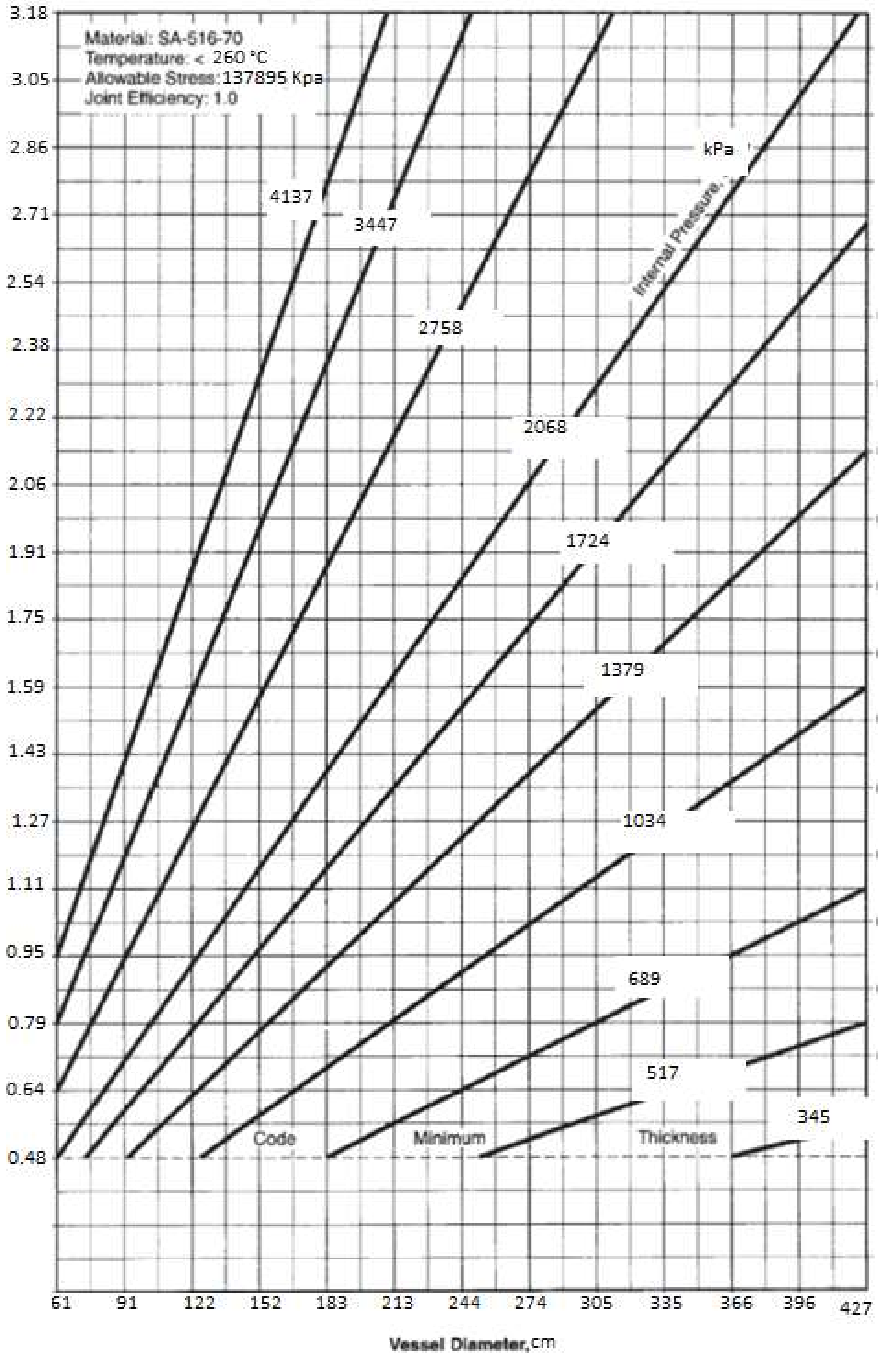

Figure 7. Pressure vessel wall thickness (Engineers Edge, 2016). The y-axis represents vessel wall thickness.

4.3.3 Heat exchanger EA-1

When dealing with slurries or other fouling feeds, spiral heat exchangers are the best choice. The curved path of the liquid creates turbulence that washes away depositions, making the spiral heat exchanger self-cleaning (Bengtson, 2010). The heat exchanger can also be cleaned easily by opening the sides. They are also compact, offer high overall heat transfer coefficient and low pressure drops (Bengtson, 2010). The spiral heat exchanger in this process is countercurrent flow, enabling high heat exchange efficiency.

The heat exchange coefficient is assumed to be 550 W/(m^2K), which was selected based on the 700 W/(m^2K) typical heat transfer coefficient of spiral heat exchangers for liquid-liquid heat exchange (The Engineering Toolbox, 2016), but reduced to account for the solids in the liquid. The assumption was made because more accurate data was not available. The process can later be modified when more accurate coefficients concerning the heat transfer of biomass and hydrochar can be found.

The thermal duty of the heat exchanger was calculated by Balas to be 214 kW. The simulator also gives the outlet temperatures. The required surface area of the heat exchanger can be calculated from the equation

$$\dot{\mathrm{Q}} = \mathrm{kA}\Delta\mathrm{T_m}\mathrm{F} \tag{2}$$

where $\dot{\mathrm{Q}}$ is the thermal duty of the heat exchanger, k the heat exchange coefficient, A the surface area of the heat exchanger, F the correction factor and $\Delta\mathrm{T_m}$ the logarithmic mean temperature difference (LMTD). $\Delta\mathrm{T_m}$ can be calculated from

$$\Delta\mathrm{T_m} = \frac{\Delta\mathrm{T_A} - \Delta\mathrm{T_B}}{\ln\left(\frac{\Delta\mathrm{T_A}}{\Delta\mathrm{T_B}}\right)} \tag{3}$$

where $\Delta\mathrm{T_A}$ is the temperature difference between the streams on one side of the heat exchanger and $\Delta\mathrm{T_B}$ the temperature difference on the other side. In a countercurrent heat exchanger, this means the temperature difference between the cold inlet and the hot outlet as well as the cold outlet and hot inlet. The LMTD is 16.8 °C in this heat exchanger. The correction factor for an ideal countercurrent heat exchanger is 1. Spiral heat exchangers are close to ideal countercurrent heat exchangers and therefore the correction factor is assumed to be 0.95. The heat exchanger area was calculated to be 22.1 m^2. The dimensions of the exchanger were estimated from Alfa Laval's spiral heat exchanger brochure. The body has a diameter of 1000 mm, length of 800 mm and the height with a stand is 1500 mm (Alfa Laval, 2015).

4.3.4 Flash tanks FA-2 FA-3

The flash tanks chosen were flash separators constructed by the Penn separator corporation, as they are smaller than regular flash tanks (Penn separator corporation, 2016). This is possible due to the tangential inlet and cyclonic flow through the separator. Thus, the steam formed does not pick up any water on its way out. It is assumed here that these flash separators work with slurries flawlessly. Therefore, the sizing was done with the Penn separator corporations own sizing charts, seen in
Table 4, which calculates the amount of steam flashed in the separator. Then the separator size is obtained from Table 5. The inlet pressure to the first flash tank is 2895 kPa and the flash tank pressure 2000 kPa. From
Table 4 we can see that the amount of flash steam is 3.6 %. The inlet feed is 3636 kg/h and the flashed steam 131 kg/h. Using Table 5 we can calculate the size of the flash separator, resulting in 0.86 m high and 0.36 m in diameter. The inlet pressure for the second tank is 2000 kPa and the tank pressure is 352 kPa, resulting in a flash amount of 12.1 % or 448 kg/h. Using Table 5we find that a flash tank of the same dimensions is required, but requires a larger steam vent (10 cm compared to 5 cm in the first flash tank).

Table 4. Penn flash separator flash steam percentage (Penn separator corporation, 2016)

Tank pressure [kPa]	**Feed pressure [kPa]**												
	4137	**3447**	**3103**	**2758**	**2413**	**2068**	**1724**	**1379**	**1034**	**689**	**517**	**345**	**172**
3447	2,80												
2758	5,80	3,10	1,30										
2413	7,50	4,80	3,00	1,80									
2068	9,20	6,60	4,80	3,60	1,90								
1724	11,10	8,50	6,80	5,60	3,90	2,10							
1551	12,20	9,70	8,00	6,80	5,10	3,30	1,20						
1379	13,30	10,80	9,10	7,90	6,20	4,40	2,40						
1207	14,40	11,90	10,30	9,10	7,40	5,70	3,70	1,30					
1034	15,70	13,20	11,60	10,40	8,80	7,00	5,00	2,70					
862	17,10	14,60	13,00	11,90	10,30	8,50	6,60	4,30	1,60				
689	18,60	16,20	14,60	13,50	11,90	10,20	8,30	6,00	3,40				
517	20,30	18,00	16,40	15,30	13,70	12,10	10,20	7,90	5,40	2,00			
345	22,60	20,30	18,80	17,70	16,10	14,50	12,60	10,40	7,90	4,60	2,60		
172	25,40	23,10	21,60	20,60	19,10	17,50	15,60	13,50	11,00	7,80	5,90	3,30	
103	26,90	24,70	23,20	22,10	20,60	19,00	17,20	15,10	12,70	9,50	7,60	5,10	1,80
69	27,80	25,60	24,20	23,10	21,60	20,10	18,30	16,20	13,80	10,60	8,70	6,20	2,90
34	28,90	26,70	25,20	24,20	22,70	21,10	19,40	17,30	14,90	11,80	9,90	7,40	4,20
0	30,20	28,00	26,60	25,60	24,10	22,60	20,80	18,80	16,40	13,30	11,40	9,00	5,80

Table 5. Penn flash separator sizing chart (Penn separator corporation, 2016). The chart gives the height, diameter, condensate drain and steam vent size in cm.

		Flash steam, [kg/h]												
Flash separator feed [1000 kg/h]		**45,3592**	**226,796**	**317,514**	**453,592**	**907,184**	**1360,78**	**1814,37**	**3628,74**	**5443,1**	**8164,66**	**9525,43**	**11339,8**	**15875,7**
45,3592	**Height**	274,32	274,32	274,32	274,32	274,32	274,32	274,32	274,32	274,32	274,32	274,32	274,32	274,32
	Dia.	76,2	76,2	76,2	76,2	76,2	76,2	76,2	76,2	76,2	76,2	76,2	91,44	91,44
	Drain	12,7	12,7	12,7	12,7	12,7	12,7	12,7	12,7	12,7	12,7	12,7	12,7	10,16
	Vent	5,08	5,08	6,35	7,62	10,16	12,7	15,24	20,32	25,4	30,48	35,56	40,64	45,72
40,8233	**Height**	243,84	243,84	243,84	243,84	243,84	243,84	243,84	243,84	243,84	243,84	243,84	243,84	243,84
	Dia.	76,2	76,2	76,2	76,2	76,2	76,2	76,2	76,2	76,2	76,2	76,2	91,44	91,44
	Drain	12,7	12,7	12,7	12,7	12,7	12,7	12,7	12,7	12,7	12,7	10,16	10,16	10,16
	Vent	5,08	5,08	6,35	7,62	10,16	12,7	15,24	20,32	25,4	30,48	35,56	40,64	45,72
36,2874	**Height**	243,84	243,84	243,84	243,84	243,84	243,84	243,84	243,84	243,84	243,84	243,84	243,84	243,84
	Dia.	60,96	60,96	60,96	60,96	60,96	60,96	60,96	60,96	60,96	60,96	76,2	91,44	91,44
	Drain	12,7	12,7	12,7	12,7	12,7	12,7	12,7	12,7	10,16	10,16	10,16	10,16	10,16
	Vent	5,08	5,08	6,35	7,62	10,16	12,7	15,24	20,32	25,4	30,48	35,56	40,64	45,72
31,7514	**Height**	182,88	182,88	182,88	182,88	182,88	182,88	182,88	182,88	182,88	182,88	182,88	182,88	182,88
	Dia.	60,96	60,96	60,96	60,96	60,96	60,96	60,96	60,96	60,96	60,96	76,2	91,44	91,44
	Drain	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16
	Vent	5,08	5,08	6,35	7,62	10,16	12,7	15,24	20,32	25,4	30,48	35,56	40,64	45,72
27,2155	**Height**	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24	
	Dia.	60,96	60,96	60,96	60,96	60,96	60,96	60,96	60,96	60,96	60,96	76,2	91,44	
	Drain	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	
	Vent	5,08	5,08	6,35	7,62	10,16	12,7	15,24	20,32	25,4	30,48	35,56	40,64	
22,6796	**Height**	182,88	182,88	182,88	182,88	182,88	182,88	182,88	182,88	182,88	182,88	182,88	182,88	
	Dia.	50,8	50,8	50,8	50,8	50,8	50,8	50,8	50,8	50,8	60,96	76,2	91,44	
	Drain	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	7,62	7,62	
	Vent	5,08	5,08	6,35	7,62	10,16	12,7	15,24	20,32	25,4	30,48	35,56	40,64	
18,1437	**Height**	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24		
	Dia.	50,8	50,8	50,8	50,8	50,8	50,8	50,8	50,8	50,8	60,96	76,2		
	Drain	10,16	10,16	10,16	10,16	10,16	10,16	10,16	10,16	7,62	7,62	7,62		
	Vent	5,08	5,08	6,35	7,62	10,16	12,7	15,24	20,32	25,4	30,48	35,56		
13,6078	**Height**	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24			
	Dia.	45,72	45,72	45,72	45,72	45,72	45,72	45,72	45,72	50,8	60,96			
	Drain	7,62	7,62	7,62	7,62	7,62	7,62	7,62	7,62	7,62	6,35			
	Vent	5,08	5,08	6,35	7,62	10,16	12,7	15,24	20,32	25,4	30,48			
9,07184	**Height**	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24	142,24				
	Dia.	35,56	35,56	35,56	40,64	40,64	40,64	40,64	40,64	50,8				
	Drain	6,35	6,35	6,35	6,35	6,35	6,35	6,35	6,35	6,35				
	Vent	5,08	5,08	6,35	7,62	10,16	12,7	15,24	20,32	25,4				
4,53592	**Height**	86,36	86,36	86,36	86,36	86,36	86,36	86,36	142,24					
	Dia.	35,56	35,56	35,56	35,56	35,56	35,56	35,56	40,64					
	Drain	5,08	5,08	5,08	5,08	5,08	5,08	5,08	3,81					
	Vent	5,08	5,08	6,35	7,62	10,16	12,7	15,24	20,32					
2,26796	**Height**	86,36	86,36	86,36	86,36	86,36	86,36	86,36						
	Dia.	25,4	25,4	25,4	25,4	35,56	35,56	35,56						
	Drain	3,81	3,81	3,81	3,81	3,81	3,81	3,175						
	Vent	5,08	5,08	6,35	7,62	10,16	12,7	15,24						
1,36078	**Height**	86,36	86,36	86,36	86,36	86,36								
	Dia.	25,4	25,4	25,4	25,4	35,56								
	Drain	3,175	3,175	3,175	3,175	3,175								
	Vent	5,08	5,08	6,35	7,62	10,16								
0,90718	**Height**	86,36	86,36	86,36	86,36									
	Dia.	20,32	20,32	20,32	25,4									
	Drain	3,175	3,175	2,54	2,54									
	Vent	5,08	5,08	6,35	7,62									
0,45359	**Height**	86,36	86,36	86,36										
	Dia.	15,24	20,32	20,32										
	Drain	2,54	2,54	2,54										
	Vent	5,08	5,08	6,35										

4.3.5 Mixer FA-1

The sizing of the mixing tank is a big uncertainty in this thesis, as no literature on tank sizing can be found. Most of the material concerning mixing handles the choice of agitators. However, since space is the limiting factor for this HTC process, the size of the mixing tank is a decisive factor. As the mantle of the mixer serves as the heat exchange area, the size needs to be calculated in conjunction with the heat exchange.

The heat exchange mantle was calculated with the assumption that the internals of the mixer are perfectly mixed and at the temperature of 62 °C (calculated with the Balas simulation). We assume that the jacket of the mixer further heats the system from this temperature, even though in reality the heating and mixing occurs simultaneously. The heat exchange coefficient was assumed to be 550 W/(m^{2}K), in lack of better data. The correction factor is assumed to be 0.5 as the jacketed vessel is far from an ideal countercurrent setup. The required heat exchange area is 10.5 m^2 according to equation (2). Balas gives the thermal duty of 30.05 kW for the heat exchange. The thermal duty inside the tank can be calculated from

$$Q = \frac{mc_p\Delta T}{t}. \tag{4}$$

where c_p is the specific heat capacity, ΔT the change in temperature, m is the mass in the mixer and t the residence time. As the mass in the tank is a function of the mass flow and residence time, which is equal to the heating time, equation (4) is independent of the amount of slurry in the tank. The area of the tank cannot match the heat exchange area within the confinements that the container and filter press present. Thus the area is increased by a central coil. The tank is sized so that the height and diameter are both 95 cm and the volume 673 dm^3.

4.3.6 Solid-liquid separation HA-1

The solid product needs to reach a moisture content of 30 %. Filter presses can reach this moisture content (Durco Filters, 2016), can handle viscous slurries (Towler and Sinnott, 2013) and are relatively small (long but not wide or high). While a centrifuge can be operated continuously, they cannot produce cakes with solid contents up to 70% (Porteous, 1998). Filter presses are batch operated and therefore a buffer tank is required.

The feed to the filter press is 0.87 kg/s of which the solid content is 6 %. It is assumed that the filter press can process a batch in ten minutes, meaning 144 cycles a day. Using these values, the required filter press volume can be calculated with the online filter press calculator (http://www.chamberfilterpress.net/en/online-calculator-filterpress.php), giving the volume of 27.5 dm^3. A 28 dm^3 filter press has the dimensions 1.25m·0.8m·1.32m (H·W·L) (SERFILCO, 2002). With 6 chambers and filter area of 470·470mm, the total press filter area is 2.65 m^2. With the assumed 10-minute processing time and 0.96 kg/s feed, the buffer tank needs a 526 dm^3 capacity. The buffer tank has a 90 cm diameter and height.

4.3.7 Materials

As the acids that are formed in the HTC reactions are recycled in the process, the materials of the equipment need to be resistant to corrosion. This rules out using carbon steel as material, as it corrodes easily. Stainless steel type 304 contains chromium and nickel that gives good resistance against corrosion. Stainless steel can resist acetic acid concentrations of up to 10 % at 100 °C (Towler and Sinnott, 2013), which implies that it would be an optimal choice as acetic acid is the foremost component found in the liquid product (Stemann et al., 2013b). Stainless steel 304 was thus selected as the material for pipes and equipment.

4.3.8 Controls

The instrumentation of the process can be seen in Figure 6. Both tanks and all three reactors have liquid level control through valves. The mixing tanks level indicator is coupled to a valve after the junction of waste and recycled liquid. Excess liquid levels close the valve, leading the liquid product out of the process. The reactors are also equipped with pressure release valves and burst discs leading out of the container. As the reaction is not very dependent on temperature, minor fluctuations in temperature are allowable. Only temperature controls in the reactors coupled to the release valves and electrical heating system are required to prevent the temperature from rising uncontrollably. However, the risk for this is slim, as the HTC reactions are only mildly exothermic. The valves in the process should be suitable for slurries, providing unobstructed and streamlined flow to prevent particles from gathering. Saunders valves, pinch valves and lined butterfly valves are examples of good valve types for slurries (Tandon, 2007). Saunders valves can operate at higher pressures than pinch valves and is thus more suitable for the process in question. The pipe between the buffer tank and filter press has a control valve. It is closed except when filling the filter press, recirculating the liquid in the tank and preventing solid retention. When the filter press is to be filled, the valve opens and the slurry is redirected.

4.3.9 Piping & pumps GA-1 to GA-8

The most economically feasible pipe diameter can be estimated from

$$\mathrm{d_{i,o}} = 3.2\left(\frac{\mathrm{m}}{\rho}\right)^{0.5}, \tag{5}$$

where m is the mass flow rate and ρ the fluid density (Towler and Sinnott, 2013). In this case the density is assumed 1000 g/l and the mass flow is 1.14 kg/s. Using equation (5) we get the optimal pipe diameter 108 mm. This diameter would lead to a flow velocity of 0.12 m/s, below the critical velocity of 0.62 m/s. For a slurry with a density of 800 g/l the maximum velocity in the pipes should not exceed 3 m/s to minimize erosion (Simpson, 1968). A pipe diameter of 40 mm gives a velocity above 0.73 m/s for all slurry streams in the process, above the critical velocity but below the maximum velocity. The exception is the pipes connecting to the mixing pumps, where the flow is higher to ensure better mixing. Therefore, the pipe diameter for these pipes was selected as 90 mm. All diameters give room for the large particles in the slurry, preventing clogging.

There are 8 pumps in the process. GA-1 is the slurry pump around for the mixing tank, GA-2 pumps the feed through the heat exchanger and against the pressure of the second flash tank. GA-3 works against the pressure of the first flash tank. GA-4 pushes the slurry to the reactor while GA-5, GA-6 and GA-7 act as the mixing pumps for the three reactors. GA-8 is the mixer pump in the buffer tank.

Reynolds number is a dimensionless quantity that can help predict the flow patterns and is calculated from

$$\mathrm{Re} = \frac{\rho v d}{\mu}, \tag{6}$$

where μ is the viscosity of the fluid, d the inside pipe diameter and v the fluid flow velocity. The viscosity of the biomass slurry is assumed to be 40 Pa·s, the same as 10 wt-% 20-mesh corn stover slurry in the experiments conducted by Viamajala *et al.* (2009). Thus the Reynolds number is 0.7, meaning the flow is Stokes flow, also known as creeping flow.

The calculation of the dynamic pressure drop in the pipes is extremely rough, as Reynolds number used is dependent on both density and viscosity of the slurry. Calculations will however be made with assumptions on density and viscosity. The calculations can then be upgraded when better data is available. The pressure drop in the pipe for laminar flow can be calculated trough the Hagen–Poiseuille equation

$$\Delta P = \frac{\mu L \dot{V}}{\pi d^4}, \quad (7)$$

where μ is the viscosity, L the pipe length, $\dot{V}$ the volumetric flow rate and d the pipe diameter.

The pipe lengths were calculated from the 3D model. The pump system's connections are seen in Figure 6. The pipe lengths and fittings, and their equal pipe lengths as well as the pressure drops are listed in Table 6 for all the pumps. The pressure drop due to pipe friction was calculated with equation (7). The heat exchanger also adds a pressure drop of about 100 kPa (Golam and Oinas, 2016), affecting pump 2.

The static pressure is calculated from the difference in liquid head and pressure. The head differences are shown in Table 7. The hydrostatic pressure is calculated from

$$\Delta P_h = \rho g h, \quad (8)$$

where ρ is the density of the liquid, g the gravitational acceleration 9.81 $\frac{m}{s^2}$ and h the head difference.

The total pump head is calculated from

$$H = \frac{\Delta P_{total}}{\rho g}, \quad (9)$$

where ΔP_{total} is the sum of static and dynamic pressure drops. The pressure difference, hydrostatic pressure drop, dynamic pressure drop and pump head are listed in Table 8.

Table 6. Pipe lengths, valves, fittings and pressure drops for the pipes after each pump in the process.

Pump	**Pipe length [m]**	**Fittings**	**Equal pipe length [m]**	**Total length [m]**	ΔP_{dyn} **[kPa]**
GA-1	0.5	Tank exit Valve 90° curve Tank entry	25 18 30 50	123.5	0.77
GA-2	2.45	Tank exit Valve 90° curve 90° curve 90° curve T-crossing	25 18 30 30 30 60	177.45	1.10
GA-3	0.61	90° curve Valve T-crossing	30 18 60	120.61	0.75
GA-4	2.1	90° curve Valve Tank entry	30 18 50	100.1	0.62
GA-5	3.79	Tank exit Valve T-crossing 90° curve 90° curve Tank entry	25 18 60 30 30 50	183	1.14
GA-6	3.79	Tank exit Valve T-crossing 90° curve 90° curve Tank entry	25 18 60 30 30 50	183	1.14
GA-7	3.79	Tank exit Valve 90° curve 90° curve Tank entry	25 18 30 30 50	123	0.76
GA-8	1.77	Tank exit Valve T-crossing 90° curve	25 18 60 30	133	0.83

Table 7. Head difference for the pumps

Pump	From	To	h [m]
GA-1	Mixer	Mixer	0.12
GA-2	Mixer	Pump 2	-1.04
GA-3	Pump 2	Pump 3	0.00
GA-4	Pump 3	Reactor 1	1.61
GA-5	Reactor 1	Reactor 1/2	0.78
GA-6	Reactor 2	Reactor 2/3	0.78
GA-7	Reactor 3	Reactor 3	0.78
GA-8	Buffer tank	Filter press	0.98

Table 8. Pressure difference, static, dynamic and total pressure drop as well as head for the pumps

Pump	ΔP [kPa]	ΔP_h [kPa]	ΔP_{dyn} [kPa]	ΔP_{total} [kPa]	H [m]
GA-1	0	1.07	0.77	1.84	0.20
GA-2	49	-9.63	1.10	140.48	15.23
GA-3	1650	0.00	0.75	1650.75	179.01
GA-4	1200	14.86	0.62	1215.49	131.81
GA-5	0	7.19	1.14	8.32	0.90
GA-6	0	7.19	1.14	8.32	0.90
GA-7	0	7.19	0.76	7.95	0.86
GA-8	0	9.03	0.83	9.85	1.07

As the flowrate in the process is low, a centrifugal pump is not the best option, as these tend to pump fluids at rates of more than 1000 m^3/h (Towler and Sinnott, 2013). The type of pump best suited for this process is a rotary positive displacement pump, as these can handle even coarse particles in slurries and highly viscous liquids such as asphalt and the high temperatures of the HTC process. A Hidromechanika CSN-101 was selected for the pump dimensions, which are 232·232·1065 mm (w·h·l) (Hidromechanika, 2016).

4.3.10 Belt

For easy transport of the solid product, a belt transporter is required. The belt collects the solid product from under the filter press and transports it to a collection vehicle. The belt feeder should fit inside the container and under the filter press. Thus it’s width is most optimal as 0.74m. The belt extends to the wall closest to the filter press and further up towards the roof, increasing its length. The length is calculated from the 3D model and is 4.65 m.

5 Economic feasibility analysis

This chapter will look into the cost of the process itself, the cost of running the process and its economic feasibility. Presented is the economic feasibility for two different business cases.

5.1 Business cases

Two business cases will be addressed in this thesis. The maximum production rate of the HTC-processing unit is 288 kg/h.

Case 1 consists of a HTC processing unit next to a midsized brewery in an urban area. The raw material is brewer's spent grain (BSG). The raw material is of no cost as this is an easy way for the brewer to dispose of the waste. Since microbreweries typically have small batches, keeping the container on site for a longer time is not feasible, the brewer's spent grain should instead be collected and stored on site until the container arrives to process it. A medium sized brewery produces 20-30 tons of BSG per month. The processing unit operates until the raw material is depleted and is then transported to the next brewery. The electricity can be taken from the grid and water is also readily available. An additional belt conveyor has to be installed above the container for smooth feeding of the raw material. If we assume that the processing unit operates together with 32 breweries of medium scale, the available biomass would be roughly 9600 tons annually, assuming the breweries and the processing unit operate 12 months a year. The BSG from one brewery could be processed in 19 hours, accounting start up time. Thus, the annual operating time is 5895 hours.

Case 2 comprises the processing of wheat straw at a remote location. Ideally, the straw should have been previously collected and gathered, preferably into a hopper system under which the container can park. Alternatively, a tractor is needed to feed the raw material to the top of the container. It is assumed that the processing unit operates in southern Europe and functions for almost 12 months each year. Electricity must be generated on site, by e.g. a diesel generator. It is preferable that the processing unit operates without shutting down until the biomass is depleted, as restarting the process is very energy consuming. The container could be connected to five large farms that each have a 200 ha area of which 50 % is wheat. The annual available biomass is thus 3705 tons. The wheat straw from one farm could be processed in roughly 72 days. The annual operating time is thus 8450 hours.

In both cases the product is assumed to be taken to a central plant for drying and/or packing.

5.2 Capital costs

Typically, when the capital cost of a process is estimated, both inside and outside battery limits are considered. As this process is small and is confined to the container, some modifications have to be made to the traditional cost estimations of a chemical plant.

5.2.1 Major equipment costs

The costs of the major equipment make up the bulk of the process capital costs. The costs are converted to Euro from dollar using the exchange at time of writing, \$1=0.902 € (XE, 2016). Most of the equipment is assumed to be available on the U.S Gulf Coast and thus the price of equipment in Europe can be estimated by multiplying the cost by 1.13 (average from Germany, Italy and France (Towler and Sinnott, 2013)). Most cost data found was from 1989. To update the costs to today's values we use the chemical engineering plant cost index for 2016 (556.8), 2014 (573.6) and 1989 (355.4). Table 9 lists both the U.S. Gulf Coast price in U.S. dollars at a certain year and the average European price in € today.

The Penn flash separators cost was found as an example of cost savings on the Penn separator company's website. The size of the separator in the example happened to be the same as used in the process, so no size conversion estimations were needed. The separators are presumed to be sold on the Gulf coast and the price up to date.

The cost of the heat exchanger was estimated using Figure 8. The area of the heat exchanger in square feet is 22.1 m^2.

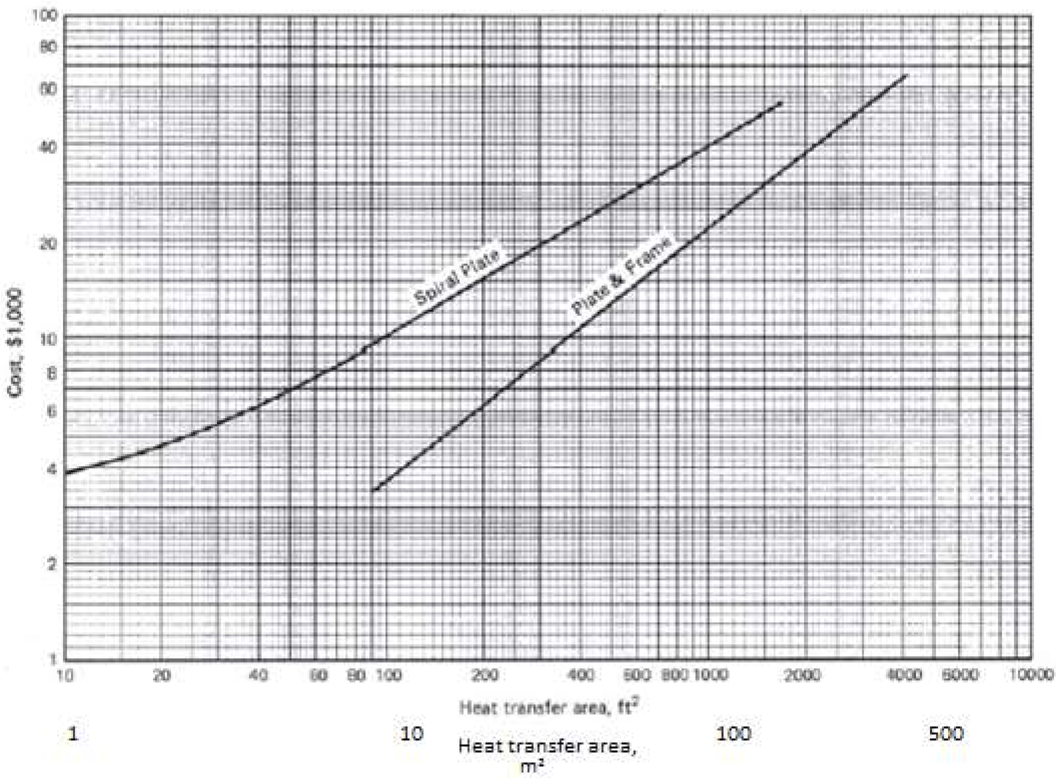

Figure 8. Spiral plate and plate & frame heat exchangers cost correlation (Garrett, 1989).

The mixing tank cost was estimated with Matches process equipment cost estimations (Matches, 2016). The estimations were for a large water heater of 10.5 m^2, stainless steel 304 and 1034 kPa pressure rating.

The cost of the filter press was calculated for a plate and frame filter in Figure 9. The area of the filter in our selected press is 2.6 m^2.

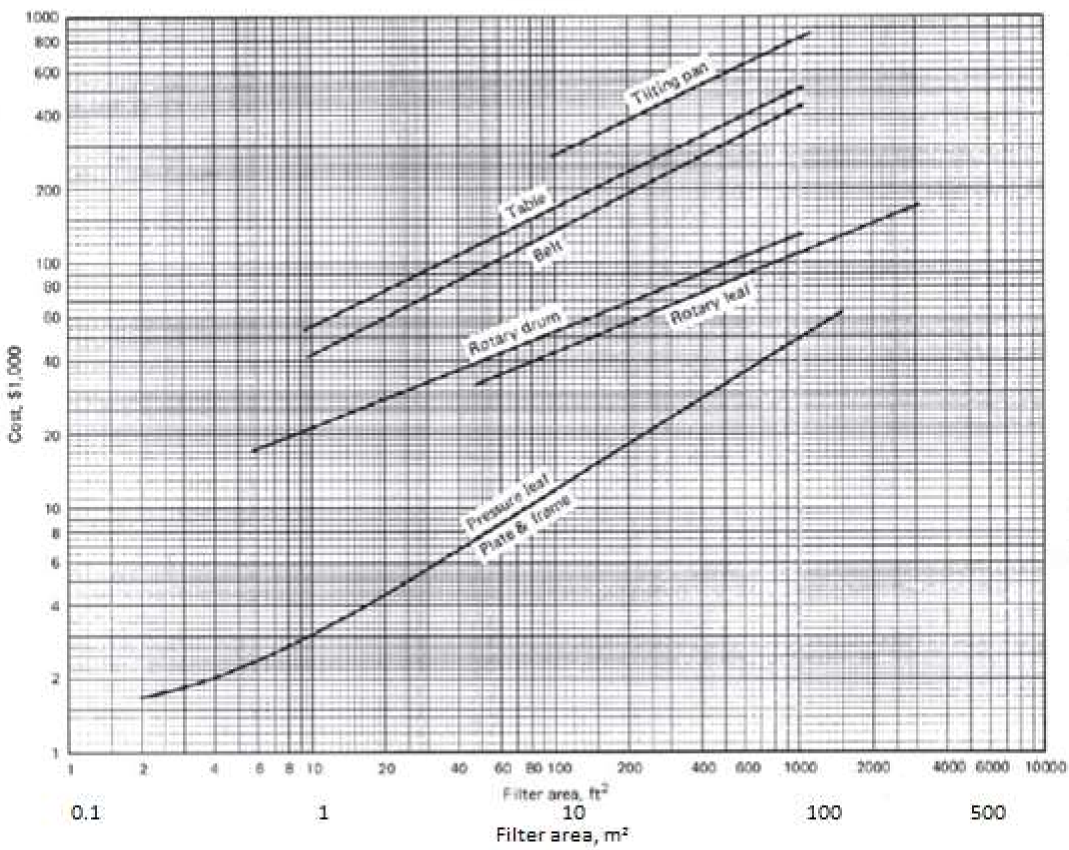

Figure 9. Filter cost correlation (Garrett, 1989).

The cost of the pumps in the process were received from the Hidromechanika price list (Hidromechanika, 2014) for the CSN-101 model. The CSN-600 model was used for the mixing pumps, due to a higher flowrate. The price is in Europe and from the year 2014.

The buffer tank cost was estimated using Figure 10 for dished heads tanks. The capacity is 538 dm^3.

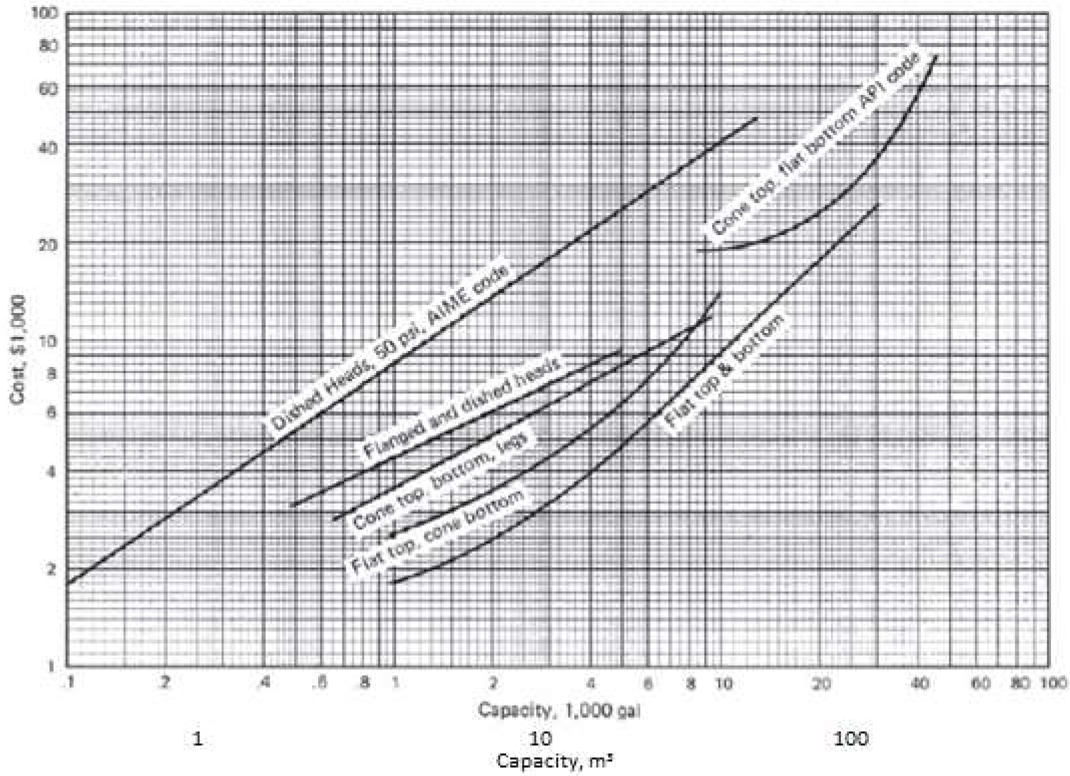

Figure 10. Cost correlations for small tanks(Garrett, 1989).

The reactors costs were calculated using Figure 11 along the extrapolated 3000 kPa line. As the correlation is for mild steel, the price must be multiplied with a factor of 1.7 to receive the price in stainless steel (Garrett, 1989). There are 3 reactors and each has a volume of 3418 dm^3.

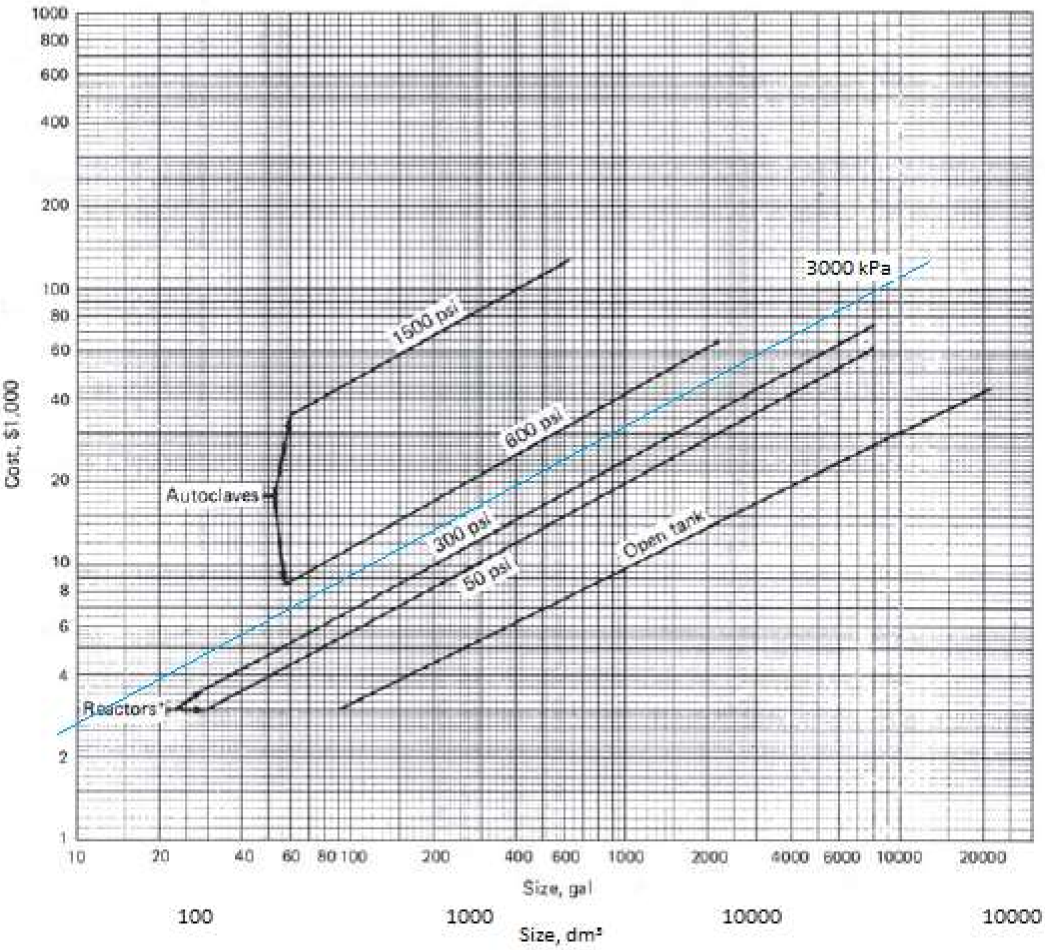

Figure 11. Reactors cost correlations (Garrett, 1989). Blue line is extrapolated value for 3000 kPa.

The cost of the belt conveyor was calculated from Figure 12 for a belt conveyor. The length is 4.66 m, the width is 74 cm. As the width differs from the width in the figure (16 in= 40.64 cm), it must be multiplied with the sizing factor

$\left(\frac{\text{width}}{40.64\text{ cm}}\right)^{0.6} = \left(\frac{73.99\text{ cm}}{40.64\text{ cm}}\right)^{0.6} = 1.433$ (Garrett, 1989).

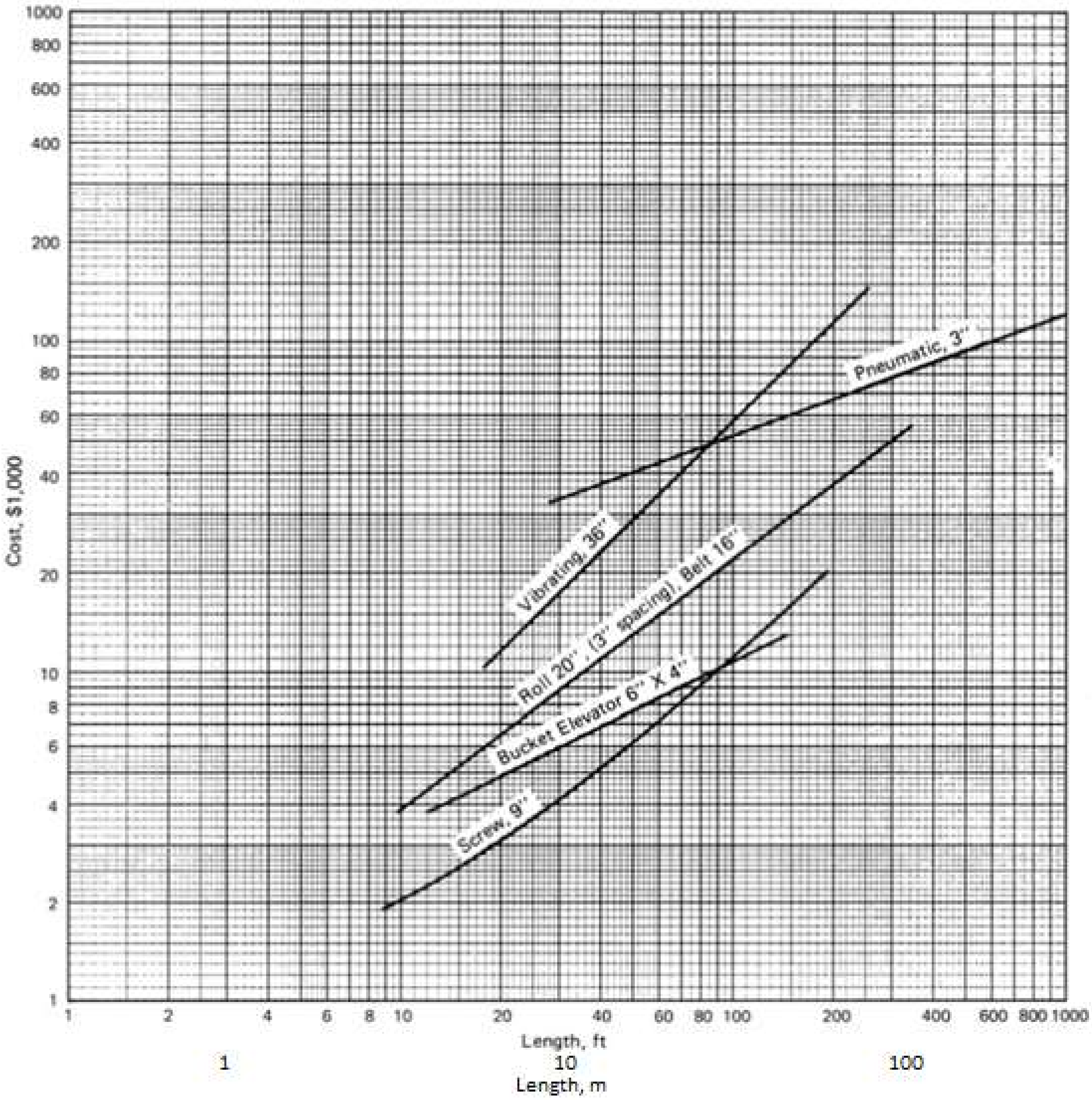

Figure 12. Cost correlation for conveyors (Garrett, 1989).

Table 9. Cost of major equipment and sum. Cost 1 stands for prices on the U.S. Gulf Coast in 1989. Cost 2 stands for equipment in Europe in 2016.

Item	**Cost 1**	**Cost 2**
Flash separator 1[2016]	$2 135	2 176 €
Flash separator 2[2016]	$2 135	2 176 €
Heat exchanger	$16 000	25 550 €
Mixing tank[2014]	$9 300	9 201 €
Filter press	$5 500	8 783 €
Pump 1[2014, Europe]	8 524 €	8 274 €
Pump 2[2014, Europe]	2 059 €	1 999 €
Pump 3[2014, Europe]	2 059 €	1 999 €
Pump 4[2014, Europe]	8 524 €	8 274 €
Pump 5[2014, Europe]	8 524 €	8 274 €
Pump 6[2014, Europe]	8 524 €	8 274 €
Pump 7[2014, Europe]	8 524 €	8 274 €
Pump 8[2014, Europe]	8 524 €	8 274 €
Buffer Tank	$2 300	3 673 €
Belt	$5 200	8 304 €
Reactors	$153 000	244 320 €
Equipment total	**$250 832**	**357 826 €**

[2016] Cost based on 2016 index.
[2014] cost based on 2014 index.
[Europe] cost based on European location.

5.2.2 Total process cost

The total ISBL cost of the plant is calculated according to

$$C_{ISBL} = \sum_{i=1}^{i=M} C_{e,i,SS} \left[\left(1 + f_p\right) + \frac{(f_{er}+f_{el}+f_i+f_c+f_l)}{f_{m,SS}} \right], \quad (10)$$

where M is the number of major equipment, $C_{e,i,SS}$ the equipment price in stainless steel 304, f_p piping cost factor, f_{er} equipment installation cost factor, f_{el} electrical cost factor, f_i cost factor for instrumentation and control, f_c civil cost factor, f_l lagging, insulation and paint cost factor and $f_{m,SS}$ the material factor for stainless steel 304. The factors for solid-liquid processes are listed in Table 10. These factors are given in (Towler and Sinnott, 2013). The total cost of the ISBL plant is thus 875 297 €.

Table 10. Factors of different parts of the process used in eq. 11 and 12.

Installation f_{er}	0.5
Piping f_p	0.6
Electrical f_{el}	0.2
Instrumentation & control f_i	0.3
Civil f_c	0.3
Lagging, insulation and paint f_l	0.1
Material $f_{m,SS}$	1.3
Design and engineering f_{de}	0.25
Contingency f_{co}	0.1

The cost of the entire process can be calculated from

$$C = C_{ISBL} \cdot (1 + f_{de} + f_{co}) + C_{container}, \quad (11)$$

where f_{de} is the cost factor for design and engineering, f_{co} the cost factor for contingency and $C_{container}$ the price of the container, roughly 1800 € for a used 40-foot ISO-container. The total cost of the process is thus 1 183 500 €.

5.2.3 Peripherals

The equipment listed here is not required for the process in itself, but are required depending on the business case considered. In case 1 the raw material does not need pretreatment but will most likely need a belt conveyor to feed the BSG to the top of the container. A belt conveyor of equal specifications as the one inside the processing unit will suffice. This adds 8300 € to the investment cost.

In case 2 the wheat straw needs to be shredded to smaller particle size. The cost of the crusher was estimated using a cost correlation from (Towler and Sinnott, 2013): Figure 13 for a jet mill working at a capacity of 0.61 t/h. This gives us the cost of 43000 €. A generator is also required to produce electricity for the process. The cost of a 100 kW diesel generator is estimated from Figure 14 for a small generator generating 100 kW electricity. The price is 60700 €.

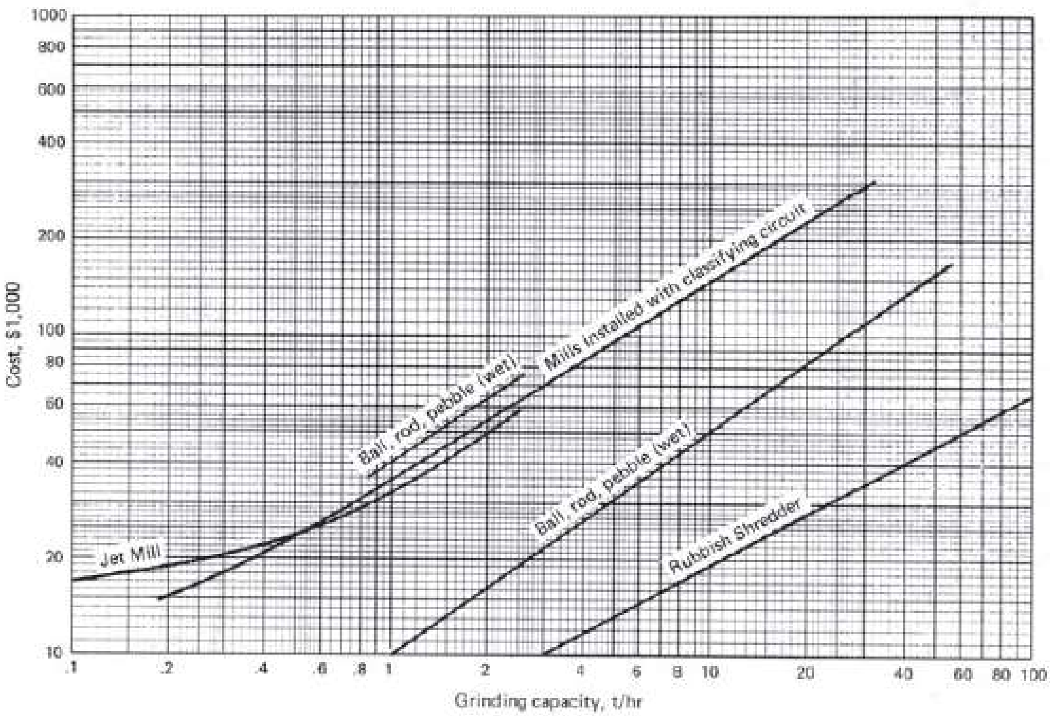

Figure 13. Cost correlation for small mills (Garrett, 1989).

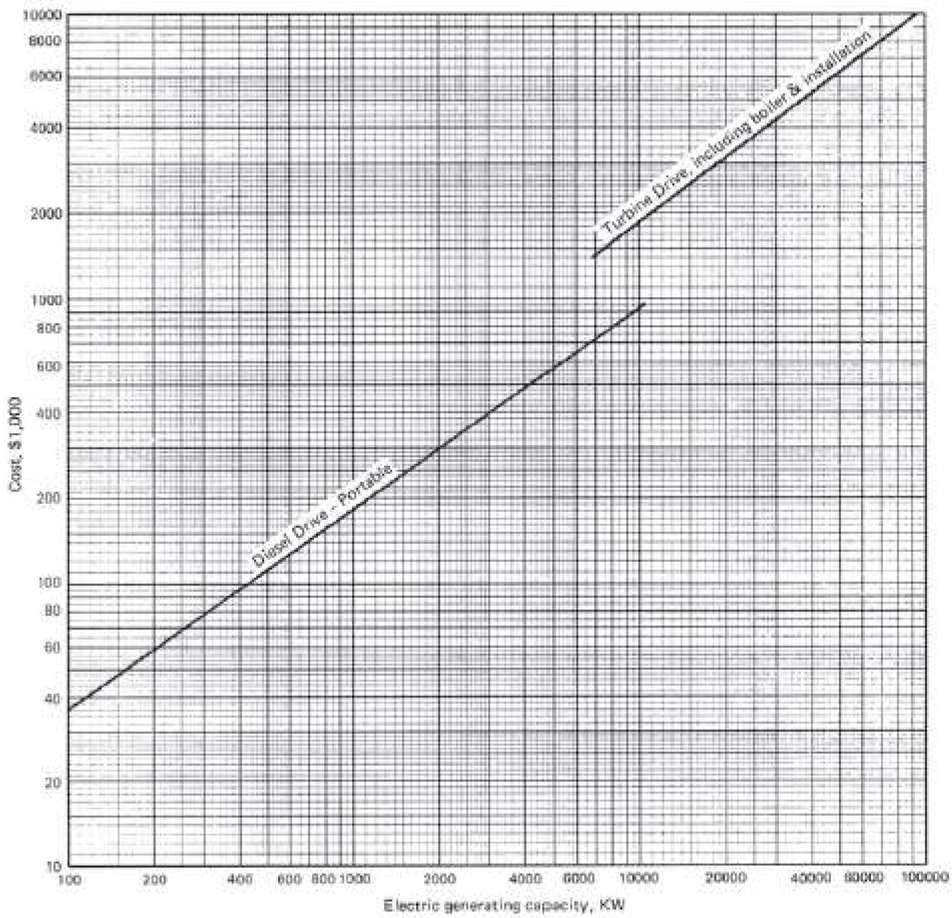

Figure 14. Cost correlation for generators (Garrett, 1989).

5.3 Operating costs

The raw material for this process is assumed to be a waste and therefore viewed to be without cost. This is because brewers rather give away BSG than pay for waste treatment. The same applies for wheat straw. Costs for the process include electricity, water, transport, maintenance and repairs, salaries, insurance and waste treatment costs (liquid and gas phase).

5.3.1 Water

Process water is recycled in the process. HTC also produces water as a byproduct. Therefore, no additional water is required during operation for case 1. However, water is required during start up. To prevent the pumps from running dry, it can be assumed that all the tanks and reactors in the process are full of water during start up. However, it can also be assumed that during shut down of the process, all of the liquid phase gets recycled. This can be done after the last biomass before shut down is fed into the process. With a 2-hour reaction time, the recycled water amounts to 5400 dm^3. The combined volume of the tanks is 11500 dm^3, so with each shut down 6100 l of water needs to be fed into the process. With a rough price assumption of 0.45 € per ton (Towler and Sinnott, 2013), the total price is 2.45 € per start up. In case 1 the amount of startups is 384, the total price thus 942 € annually. For case 2 the amount of startups is only 5, leading the annual price to be 12 €.

The moisture content for wheat straw can be as low as 8 % (O'Dogherty et al., 1995b; Mani et al., 2006). For easy handling of the slurry in the process, the dry matter content in the reactor feed is presumed to be 10 %. This requires the biomass feed to have a moisture content of at least 9 %. In the worst case scenario, water must be added. The amount of water required is 16 tons annually. The price 128 €/a, not considering the investment costs of the water transport and storage.

It can be concluded that water costs have very little effect on the overall feasibility of the process.

5.3.2 Electricity

The source of electricity is dependent on the process location. If the process is connected to an industrial installation, e.g. brewery, the electricity can be taken from the grid. A remote location requires a generator. The major equipment that uses electricity in the process are the pumps, the filter press, the heating of the feed to reaction temperature (pipe) and the belt conveyor.

The power consumption of the reactor feed heating is given from the Balas model, 80 kW for case 1 and 71 kW for case 2. It is assumed to be 100 % efficient (U.S. Department of energy, 2016).

The filter press can be assumed to consume 25 kWh/ton (Huber Technology, 2016). Multiplying this with the filter press feed 0.29 t/h gives us the power requirement of 7.2 kW.

The power required for the belt conveyor can be calculated from

$$P = \frac{C \cdot f \cdot L}{367} \cdot (3.6 \cdot G_m \cdot V + Q_t) + \frac{Q_t \cdot H}{367}, \tag{12}$$

Where C is the frictional resistance in the belts, f friction in the conveyor pulleys, L length of conveyor, G_m weight of belt and rotating parts in the conveyor, V the velocity of the conveyor, Q_t the capacity of the conveyor and H the conveyors lift height (Rulmeca, 2003). For our conveyor L=4.65 m, Q_t=0.2 t/h and H=2.6 m. C is given in (Rulmeca, 2003) for a 5 m conveyor as 6.6, f as 0.025 and G_m as 56 for a 800 mm wide conveyor. The velocity of the conveyor can be chosen freely. With the operation time of the filter press assumed to be 10 minutes, the solid product should move the length of the conveyor in that time. Thus the minimum velocity is 0.008 m/s and the power requirement according to equation (12) 2 W.

The power requirement of the pumps can be calculated from the equation

$$P = \frac{\rho \cdot g \cdot \dot{V} \cdot H}{\eta_p \cdot \eta_m}, \tag{13}$$

Where ρ is the density of the pumped liquid, g the gravitational acceleration, $\dot{V}$ the volumetric flow and H the pump head (ENCE engineering, 2016). In equation (13) η_p and η_m are the efficiencies of the pumps and motors and assumed to be 0.88 (Moore, 2009) and 0.9, respectively (Foray, 2014). The power demands are presented in Table 11.

Table 11. Required power for the pumps in the process.

Pump	Required power [W]
Pump 1	11.99
Pump 2	183.33
Pump 3	2 493.25
Pump 4	1 871.79
Pump 5	640.78
Pump 6	640.78
Pump 7	612.08
Pump 8	60.92

The electricity required for heating the process water up to reaction conditions during startup can be assumed to be equal to heating up the volume of the process tanks from ambient (25 °C) to reactor entry conditions (203 °C). The volume of the tanks is, as described in chapter 5.3.1, 11500 dm^3. Assuming the heat capacity of 4.32 kJ/kg·K and the startup time to be 4 hours for case 1 and 22.6 hours for case 2, we get according to equation (4) a heat requirement of 2468 kWh for case 1 and 2393 kWh for case 2.

The price of electricity for Europe is on average 0.0879 €/kWh (Eurostat, 2016). Assuming a yearly operation time of 5895 hours for case 1 the price for powering the process is 36 291 € annually. With startup heating this amounts to 92 641 €/a.

Case 2 uses diesel to power the generator. A 100 kW generator running at 75 % capacity consumes about 24 dm^3/h diesel (Diesel service & supply, 2016). The average price of diesel is 0.92 €/dm^3 (Fuel Prices Europe, 2016). With an annual operation time of 8450 hours, the price of electricity becomes 212 492 €/a, including startups.

5.3.3 Waste treatment

The liquid phase contains acids and other organic compounds and is therefore treated as a waste. The liquid phase could potentially be used in other applications, but because of lack of studies regarding the liquid product and its applications, it is in this work assumed to have no potential value. The gas phase consists of mostly carbon dioxide and is assumed to be pure carbon dioxide in this work.

The liquid phase is not severely toxic and can therefore be sent to a wastewater treatment plant for neutralization. The cost of the treatment can be assumed to be 1.35 € per ton (Towler and Sinnott, 2013). For case 1, the waste flow is 6995 ton/a, the price for treatment 9464 €/a. For case 2, the liquid waste is small because almost everything gets recycled. With a flow of 256 ton/a, the price is 346 €/a.

The amount of CO_2 produced annually is 673 ton/a for case 1 and 1273 ton/a for case 2. As the raw material is a bio-based waste, it can be assumed that there are no costs for gas emissions.

5.3.4 Transport

The weight of the processing unit is assumed to be 13 tons based on the weight of the container and major equipment. The solid product can be transported with a single tipper truck with 14.5 ton capacity (Reference.com, 2016). This truck can also pull the processing unit on a trailer.

In case 1, the processing unit is transported to a brewery where it operates until the biomass resource is depleted, after which it moves to the next brewery. The processing unit operates at 8 breweries per week. The total amount of trips per year is thus 435. The assumed travel distance to each brewery is assumed to be 25 km. Assuming the average cost of 0.93 €/km (Torrey and Murray, 2014), we get the annual price of 10200 € annually.

Concerning case 2, the processing unit is transported to the first field, where it operates until the biomass is depleted. Produced hydrochar is transported when the tipper truck is full. After the biomass is depleted the processing unit is transported to the next field and repeats the procedure, whereafter it is transported back to storage. The assumed travel distance to each field is 100 km. The amount of product produced is 0.3 t/h or 1461 t/a. The amount of trips is thus 337 plus 5 unit transports. The total price for transport in case 2 is thus 32 100 €, assuming the same price per km as in case 1.

5.3.5 Fixed operating costs

The process is highly automated, requiring only one operator that ideally also functions as the transporter of the process. For case 1, one employee is sufficient. But for case 2, additional workers are required to handle the feeding of biomass. It is assumed that no more than 2 workers per shift is required. The processing time per brewery is 19 hours including startup in case 1. 20 shifts are required per week, with each shift lasting 8 hours. In case 2, a 3 shift rotation is advisable until the biomass is depleted, extending to the weekend. We assume that the cost of labor is 29.50 €/h (including non-wage costs such as benefits) (Eurostat, 2015). The process runs practically 357 days straight in case 2, including five days for startups. Sundays and nightshifts (00-06) are assumed as double salary, evenings (18-22) 1.5 times normal rates. This leads to a weekly cost of 6 700 € and an annual cost of 322 800 € in case 1. For case 2 the corresponding values are 13 300 € and 567 200 €. A summary of the weekly costs for both cases are presented in Tables 12 and 13. The salaries of drivers are accounted for in the transport costs.

Table 12. Weekly salaries in € for case 1.

Day of Week	M	T	W	T	F	S	S
shift 1 6-14	236	236	236	236	236	236	442.5
Shift 2 14-22	295	295	295	295	295	295	442.5
Shift 3 22-06	442.5	442.5	442.5	442.5	442.5	442.5	0

Table 13. Summary of weekly salaries in € for case 2. Each shift has 2 workers.

Day of Week	M	T	W	T	F	S	S
shift 1 8-16	472	472	472	472	472	472	944
Shift 2 16-24	590	590	590	590	590	590	944
Shift 3 24-8	767	708	708	708	708	944	472

The cost of maintenance can be viewed as a percentage of the ISBL capital costs, typically between 3-5 % (Towler and Sinnott, 2013). We assume here that it is 5 %, since the process handles slurries that wear the equipment quickly. The cost maintenance is thus 43 800 €/a.

The cost of insurance is assumed to be 1% of the fixed capital costs (Towler and Sinnott, 2013). The cost is thus 11 800 €/a.

Because of the scale of the process, costs from management, research and development and marketing are not accounted for. These could be estimated when evaluating the feasibility of a company running several mobile biomass HTC processing units.

5.4 Profitability

5.4.1 Revenue

The main product of the process, the solid hydrochar, is the sole sellable product. Gas and liquid products are assumed to be waste. The applications for hydrochar were presented in chapter 2.2. In case 2 the product output is 2435 ton/a (70 % dry content). The same value is 1708 ton/a in case 1. For completely dry product the values are 1196 ton/a for case 1 and 1705 ton/a for case 2.

Because there is virtually no statistics on HTC hydrochar price, prices are assumed to correlate with biochar from pyrolysis. Pyrolysis was chosen as the reference as the process is similar to HTC. Prices of biochar is listed in chapter 2.2.1. In this thesis we will look at the prices range 0.6 €/kg and 0.7 €/kg, based on assumptions made for large scale biochar production. These prices are for dry biochar. The cost of drying and packing is assumed to be part of the price.

5.4.2 Summary of operational costs

The sum of operational costs is 522 600 €/a for case 1 and 977 800 € €/a for case 2. This is without taking into account the costs of drying or packing the product or the raw material containers, as these will be viewed in the price of product and raw material in the sensitivity analysis. A comparison of the operating costs can be seen in Figure 15 and Table 14. The largest contributors to operating costs are electricity and labor.

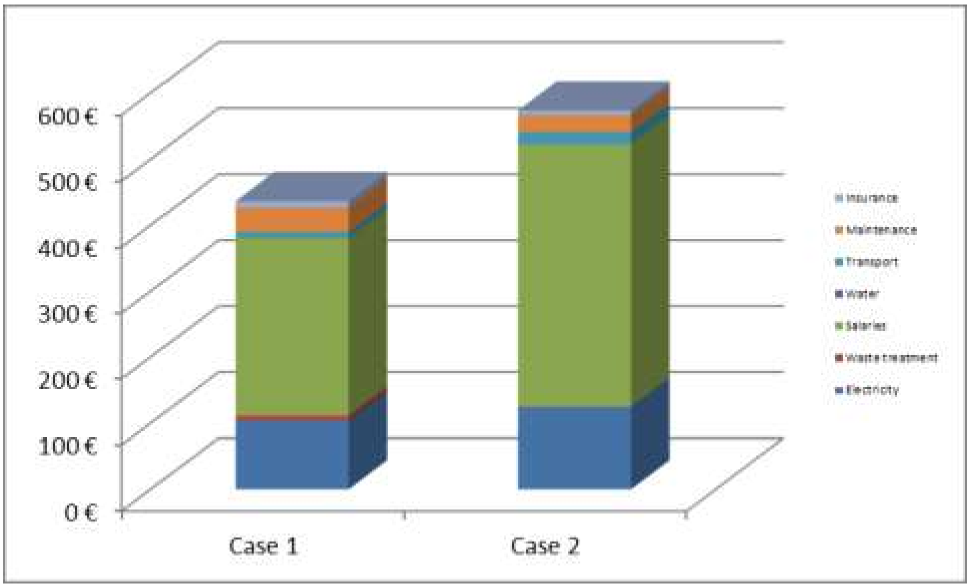

Figure 15. Comparison of operational costs for the business cases. The cost is for a ton of dry product.

Table 14. Comparison of operational costs. Costs are annual.

	Case 1	Case 2
Electricity	123 500 €	212 500 €
Waste Treatment	9 500 €	400 €
Salaries	322 900 €	677 000 €
Water	1000 €	300 €
Transport	10 200 €	32 100 €
Maintenance	43 800 €	43 800 €
Insurance	11 800 €	11 800 €
Total	522 600 €	977 800 €

5.4.3 Working capital

Working capital is “money that is tied up in sustaining plant operation” (Towler and Sinnott, 2013). The different parts that make up the working capital, their cost and how they are calculated are shown in Table 15. The working capital is added to the fixed capital investment costs at the beginning of the project. If the process is put out of function, the working capital is released. Because the process in case 2 operates continuously, the working capital is much higher than for case 1.

Table 15. Working capital for the business cases.

	Calculation basis	Cost case 1	Cost case 2
Value of product inventory	Two-week operating cost	17 451 €	38 892 €
Cash on hand	One-week operating cost	8 725 €	19 446 €
Accounts receivable	One-month operating cost	34 902 €	77 784 €
Accounts payable	One-month operating costs	34 902 €	77 784 €
Spare parts inventory	1 % of investment costs	11 835 €	11 835 €
Sum		107 814 €	225 740 €

By adding the fixed capital investment costs and working capital, we get the total capital investment of 1 300 000 € for case 1 and 1 500 000€ for case 2.

5.4.4 Simple economic analysis

The payback time can be roughly calculated from

$$\text{payback time} = \frac{\text{total investment}}{\text{pretax average annual cash flow}}, \tag{14}$$

where the cash flow is operating costs subtracted from the revenue (Towler and Sinnott, 2013). We can analyze the payback time for both prices. The results are seen in Table 16.

The pretax return on investment can be calculated according to

$$\text{pretax ROI} = \frac{\text{pretax cash flow}}{\text{total investment}} \cdot 100\%, \text{(Towler and Sinnott, 2013).} \quad (15)$$

The pretax ROIs for the different cases are listed in Table 16.

Table 16. Payback times and pretax return on investment for the business cases with different estimates on hydrochar prices

Case	**1**	**2**
Total investment cost, €	1 300 000 €	1 500 000 €
Total operating cost, €/a	523 000 €	978 000 €
Revenue 0.60 €/kg, €/a	717 000 €	1 023 000 €
Revenue 0.70 €/kg, €/a	837 000 €	1 193 000 €
Cash flow 0.60 €/kg, €/a	195 000 €	45 000 €
Cash flow 0.70 €/kg, €/a	314 000 €	216 000 €
Payback time 0.60 €/kg, a	7	34
Payback time 0.70 €/kg, a	4	7
Pretax ROI, 0.60 €/kg	15 %	3 %
Pretax ROI, 0.70 €/kg	24 %	14 %

5.4.5 Discounted cash flow rate of return

The annual cash streams are listed in Table 18 for case 1 with a product price of 0.6 €/kg. Values for case 2 or for a hydrochar price of 0.7 €/kg are not shown, but are calculated by the same methods.

Depreciation charges are tax allowances as a result of “wear and tear” (Towler and Sinnott, 2013). Here we assume that the depreciation charges are the cost of fixed capital investment divided over ten years. The charges are added to the cash flow after tax. The salvage value of the plant is assumed to be 0, as the plant continues to operate after ten years. The taxation of profit is 20 % of the previous calendar years profits, according to Finnish taxation principles (Verohallinto, 2015).

The value of a projects cash flow can be readjusted to a value today by dividing it with the discount factor,

$$(1 + i)^n, \tag{16}$$

Where i is the interest rate and n the project year that is to be discounted.

Assuming a project life of 15 years, we can calculate the discounted cash flow rate of return (DCFROR), also known as internal rate of return (IRR). The DCFROR measures the maximum interest rate a project can give while still breaking even at the end of the project life span. DCFROR can be calculated from the equation for net present value

$$\mathrm{NPV} = \sum_{n=1}^{n=t} \frac{CF_n}{(1+i')^n}, \tag{17}$$

where t is the project life in years, CF_n the cash flow in year n, and i' the DCFROR (Towler and Sinnott, 2013). DCFROR is achieved by assuming the NPV is zero at the end of the project life span. DCFROR-values are shown in Table 17. Figure 16 shows the NPV over the project years.

Table 17 - Discounted cash flow rate of return for both business cases

Case	DCFROR
Case 1, 0.60 €/kg	11 %
Case 2, 0.60 €/kg	0 %
Case 1, 0.70 €/kg	20 %
Case 2, 0.70 €/kg	10 %

Table 18. Cash flow forecasts for case 1 with a hydrochar price of 0.6 €/kg.

Year	Revenue	Operating costs	Gross profit	Depreciation	Taxable in-come	Taxes	Cash flow	Discount Factor	Present value of CF
0	0 €	0 €	0 €	0 €	0 €	0 €	-1 299 569 €	1.00	-1 299 569 €
1	717 435 €	522 438 €	194 998 €	119 178 €	75 819 €	0 €	194 998 €	0.90	175 534 €
2	717 435 €	522 438 €	194 998 €	119 178 €	75 819 €	15 164 €	179 834 €	0.81	145 725 €
3	717 435 €	522 438 €	194 998 €	119 178 €	75 819 €	15 164 €	179 834 €	0.73	131 179 €
4	717 435 €	522 438 €	194 998 €	119 178 €	75 819 €	15 164 €	179 834 €	0.66	118 085 €
5	717 435 €	522 438 €	194 998 €	119 178 €	75 819 €	15 164 €	179 834 €	0.59	106 298 €
6	717 435 €	522 438 €	194 998 €	119 178 €	75 819 €	15 164 €	179 834 €	0.53	95 688 €
7	717 435 €	522 438 €	194 998 €	119 178 €	75 819 €	15 164 €	179 834 €	0.48	86 137 €
8	717 435 €	522 438 €	194 998 €	119 178 €	75 819 €	15 164 €	179 834 €	0.43	77 539 €
9	717 435 €	522 438 €	194 998 €	119 178 €	75 819 €	15 164 €	179 834 €	0.39	69 799 €
10	717 435 €	522 438 €	194 998 €	119 178 €	75 819 €	15 164 €	179 834 €	0.35	62 832 €
11	717 435 €	522 438 €	194 998 €	0 €	194 998 €	15 164 €	179 834 €	0.31	56 561 €
12	717 435 €	522 438 €	194 998 €	0 €	194 998 €	39 000 €	155 998 €	0.28	44 166 €
13	717 435 €	522 438 €	194 998 €	0 €	194 998 €	39 000 €	155 998 €	0.25	39 758 €
14	717 435 €	522 438 €	194 998 €	0 €	194 998 €	39 000 €	155 998 €	0.23	35 789 €
15	717 435 €	522 438 €	194 998 €	0 €	194 998 €	39 000 €	263 784 €	0.21	54 477 €
						Interest rate	**11.09 %**	**NPV**	**0 €**

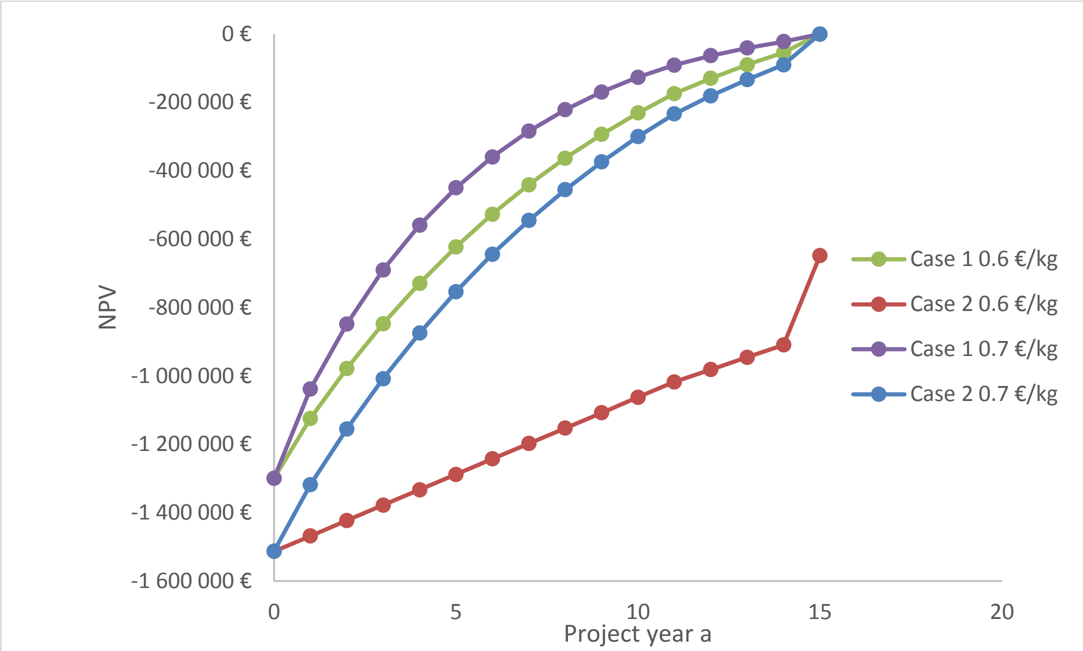

Figure 16. Net present values of both business cases for hydrochar prices 0.6 €/kg and 0.7 €/kg.

6 Sensitivity analysis

The business cases proposed in the previous chapter are only two of many possible cases. The values are assumed but give no room for variation. Sensitivity analysis gives an image of the effect of variation in the process conditions. It gives a better look at how changes in the process or business cases can change the feasibility. In this chapter we look at how changing different variables in the process affects the operating costs and revenue as well as how changing variables in the operating costs, capital costs and revenues affect the overall feasibility. The price of hydrochar is assumed to be 0.6 €/kg.

6.1 Process variables

Variations in the process model are viewed as to have an effect on the capital and operating costs of the process. We will not look as how these affect the overall feasibility of different business cases. The results are presented in Figures 17 and 18.

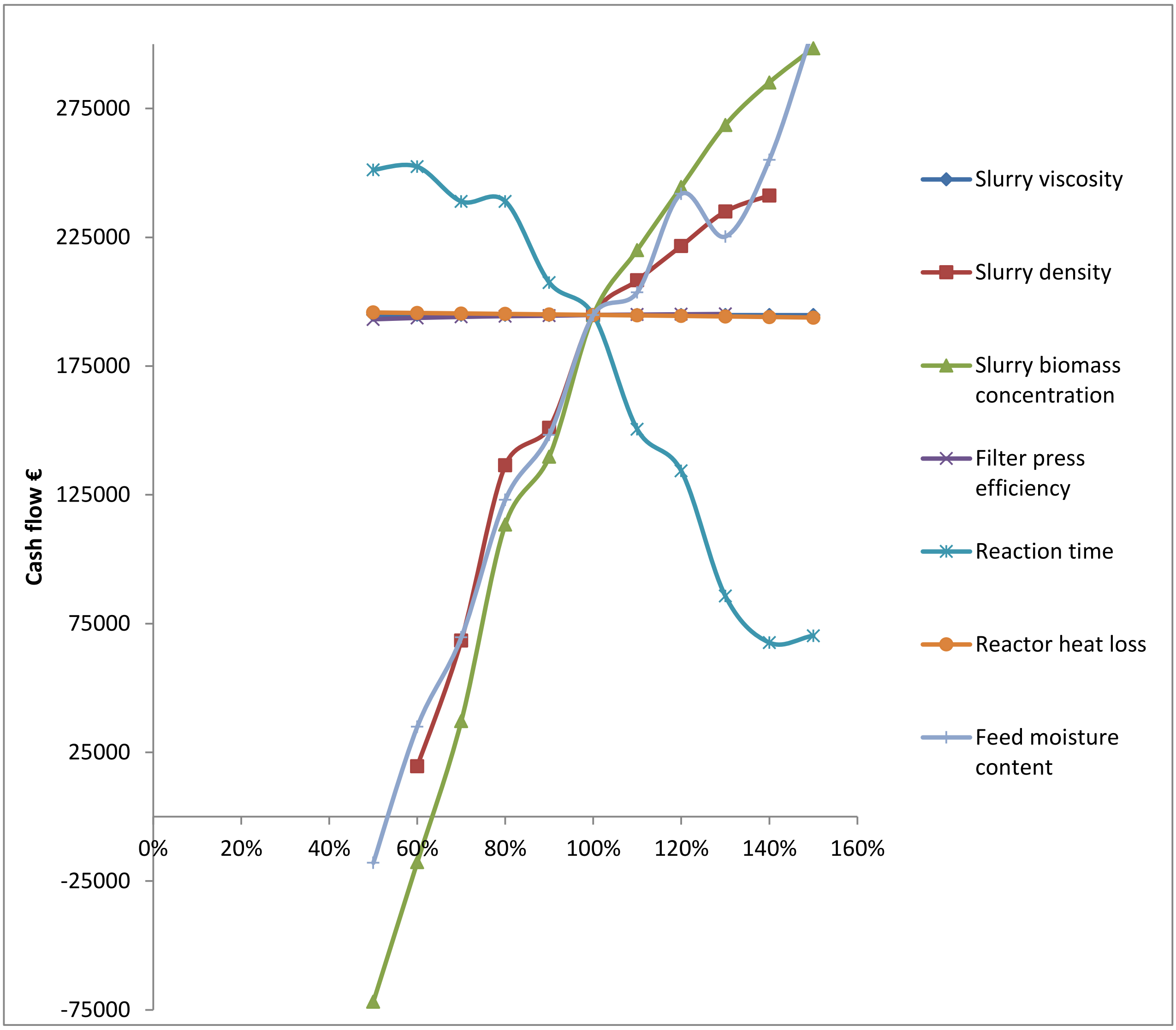

Figure 17. Sensitivity analysis for process variables in case 1.

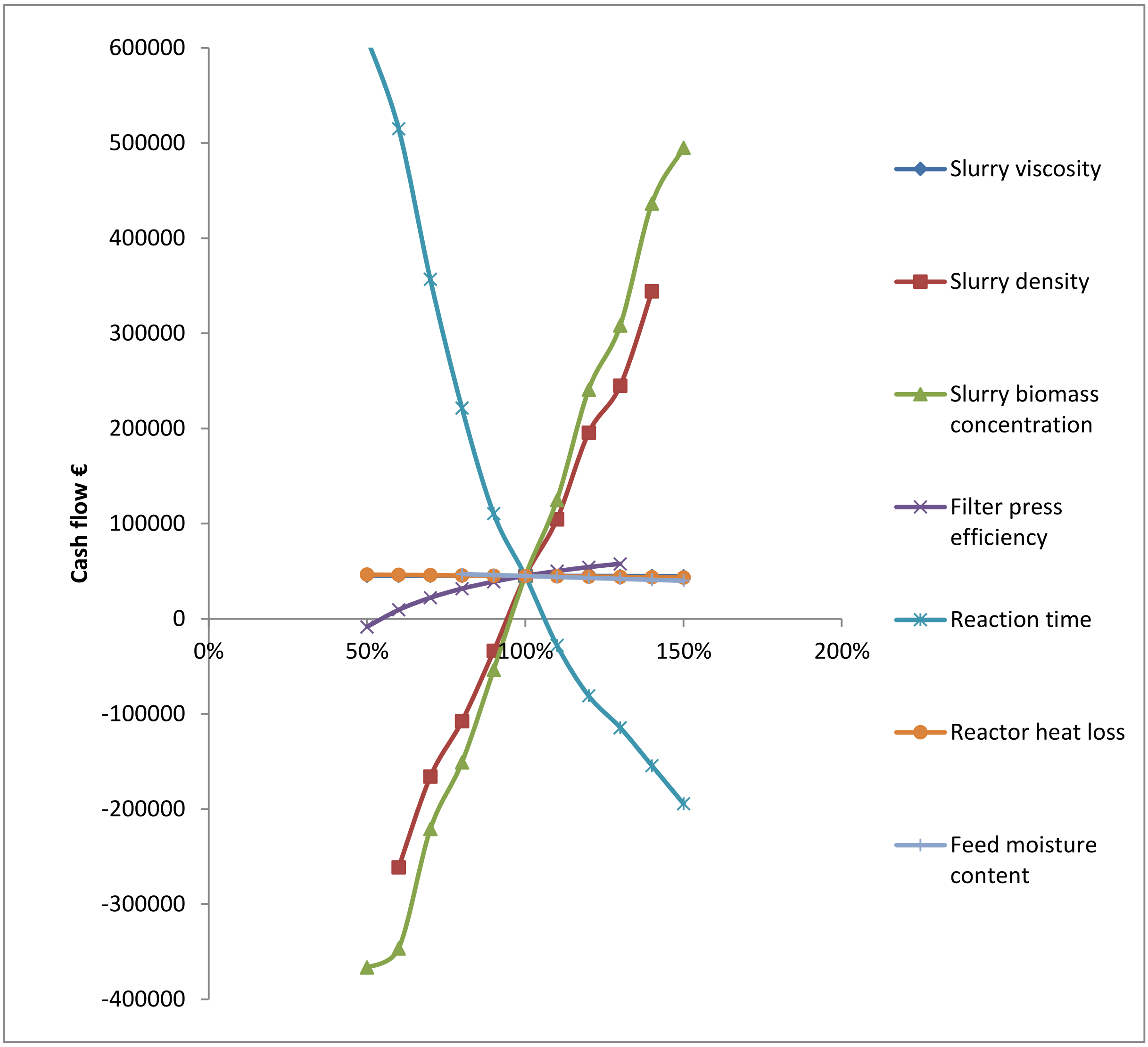

Figure 18. Sensitivity analysis for process variables in case 2.

6.1.1 Viscosity

One of the major uncertainties in the process model is the viscosity of the biomass or hydrochar slurry. This is because the viscosity is not measured in any previous works and the assumption of 40 Pa·s is based on slurries of corn stover (Viamajala et al., 2009).

6.1.2 Density

Density has a large impact on the process economics, as lower densities lead to a smaller product output, as the size of the reactors remain unchanged. The number of breweries or farms processed is dependent on the product output, as is the work hours and thus salaries. It can be seen from the figures that case 2 is much more sensitive to density change. This is due to the longer annual operating time and continuous operation. It can be conducted that a higher density gives better feasibility, as more biomass can be processed at a given time.

6.1.3 Dry matter

Like density, the dry matter content of the feed to the reactor affects the product output rate. Thus it has a major effect on the feasibility of the process. It can be concluded that a higher dry matter in the reactor feed gives better feasibility, as more biomass can be processed at a given time.

6.1.4 Filter press efficiency

The filter press is assumed to produce a solid product with 30 % moisture. This analysis studies how changes in the solid product moisture affects cash flow. The size of the filter press is assumed to remain unchanged. It can be deduced that a higher dry matter in the cake gives better feasibility, as the amount of solid product increases. This affects particularly transportation costs.

6.1.5 Reaction time

The reaction time was an assumption made, as no clear correlation between reaction severity and product quality could be found. The reaction time affects how much feed can be processed in the reactors. Much like density, case 2 is more sensitive to changes in reaction time. It can be concluded that a shorter processing times increase feasibility. Here it is assumed that the price of hydrochar remains unchanged though the quality of the products most certainly varies and thus there are bound to be variations in the price.

6.1.6 Reactor energy loss

The energy losses were assumed to be 8 kW for the reactors together and 1 kW for the mixing tank and buffer tank. This analysis shows changes in reactor energy loss. A higher energy loss leads to lower feasibility.

6.1.7 Feed moisture

The moisture content of the feed is not necessarily constant. BSG dry matter content varies between 20-30 % and wheat straw has a moisture content between 8-23 %. It can be reckoned that a higher dry matter in the feed leads to better feasibility. This is due to smaller heating costs as a result of lower moisture content.

6.2 Business case variables

This chapter looks at the variables in the various business cases and how they affect feasibility in each case. The annual cash flow is the measure of feasibility, except when analyzing the capital investment and working capital. For these we study the DCFROR as a measure of sensitivity. The sensitivity analysis for the cost of water, electricity, waste, transport and labor are presented in Figure 19 for case 1 and in Figure 20 for case 2, together with the sensitivity analysis for the final product price. Sensitivity analysis for raw-material cost and carbon taxation are shown in separate figures, because the base values are assumed as nil euro.

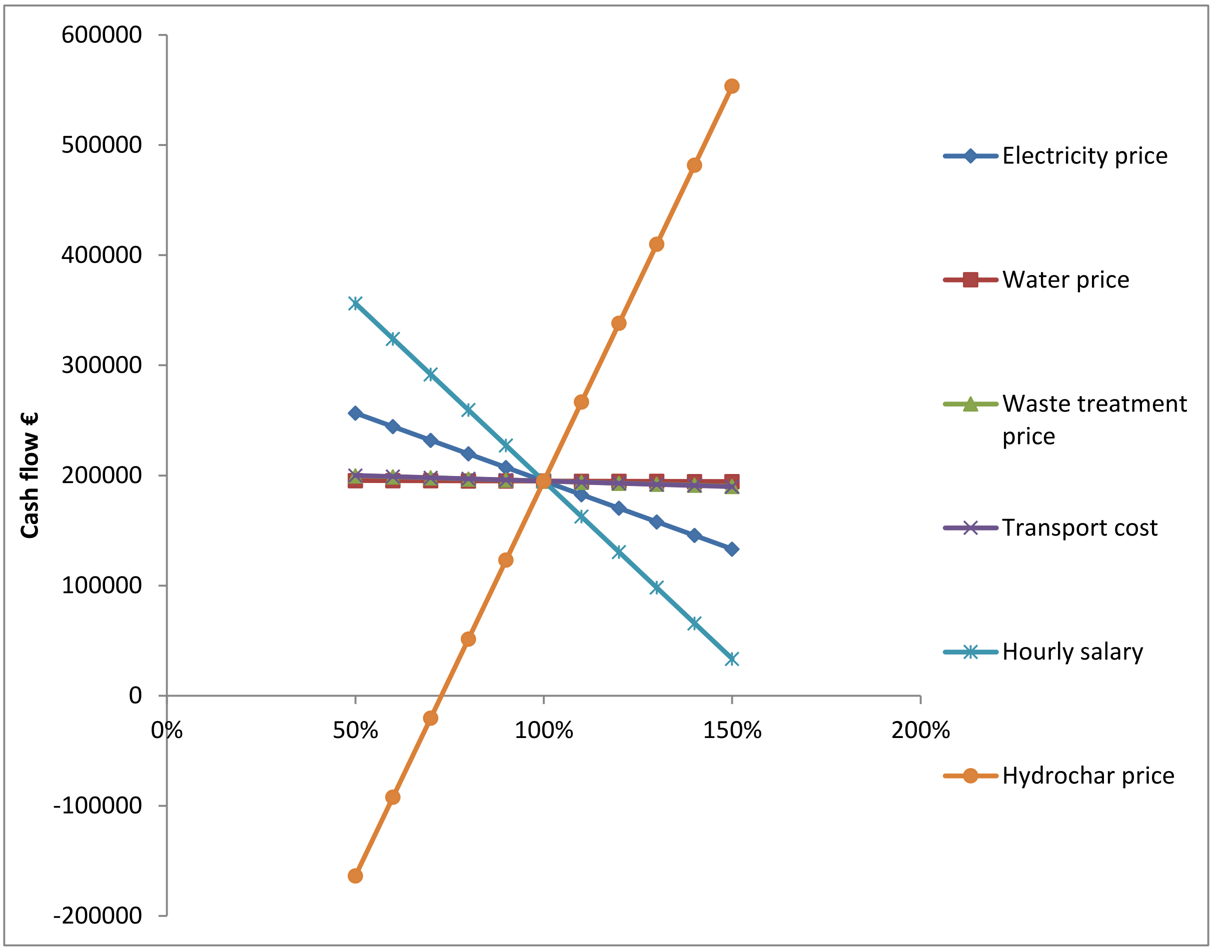

Figure 19. Sensitivity analysis for some business variables for case 1.

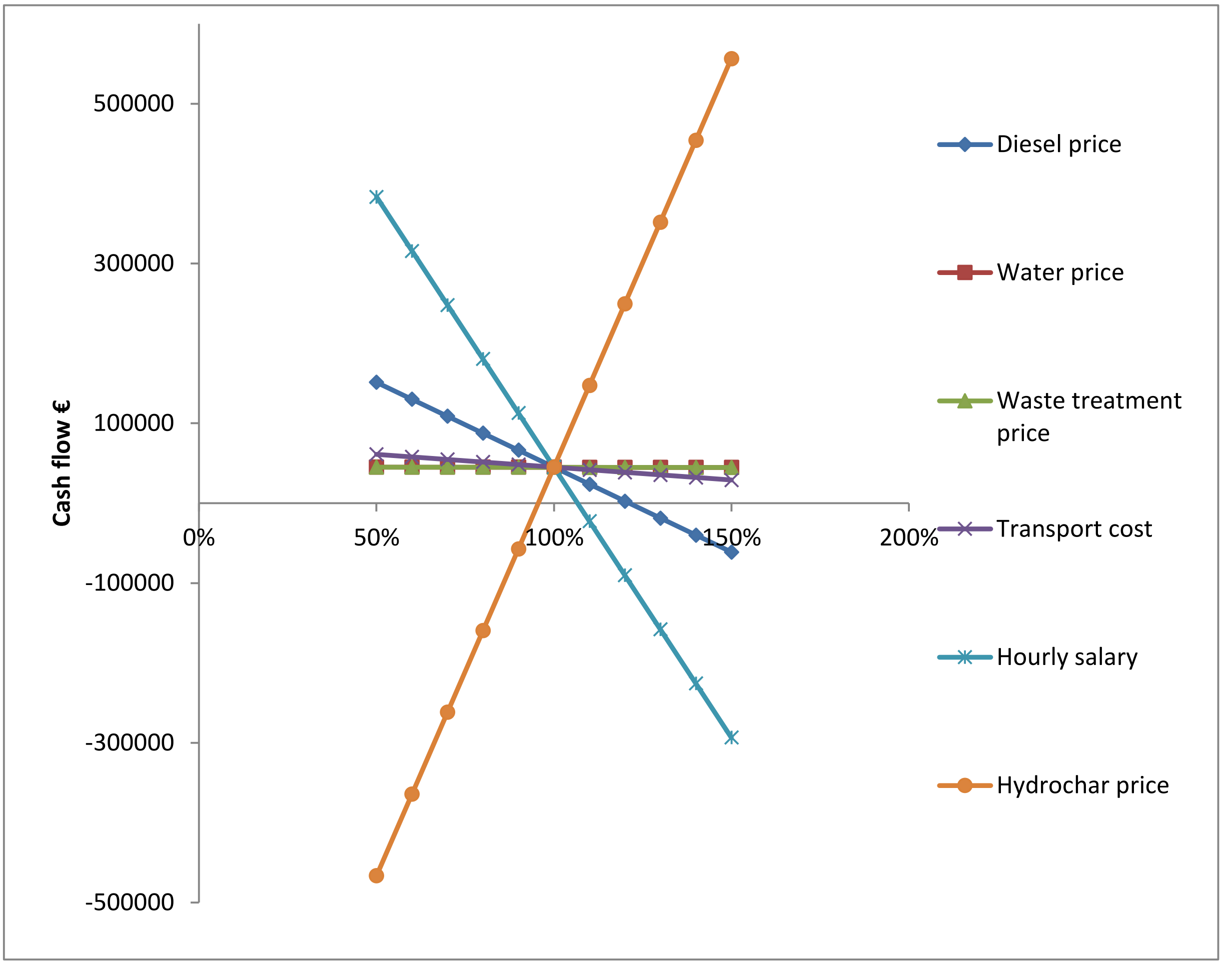

Figure 20. Sensitivity analysis for some business variables for case 2.

6.2.1 Raw material price

We have assumed that the price of the raw-material is nil, as they are wastes. However, this might not be always true, as brewers and farmers may require compensation for their product. This becomes topical if there is a rivalling company interested in buying the material. The price of the raw material can also be assumed to include the storage tanks, not addressed earlier. Because the raw-material is categorized as waste, there is a possibility that the producer of the is willing to pay for removal of the waste. Thus also negative values of the raw material price are examined. The negative value can also account for possible financial aid received from a national level for running a green process. The results for the sensitivity analysis are shown in Figure 21.

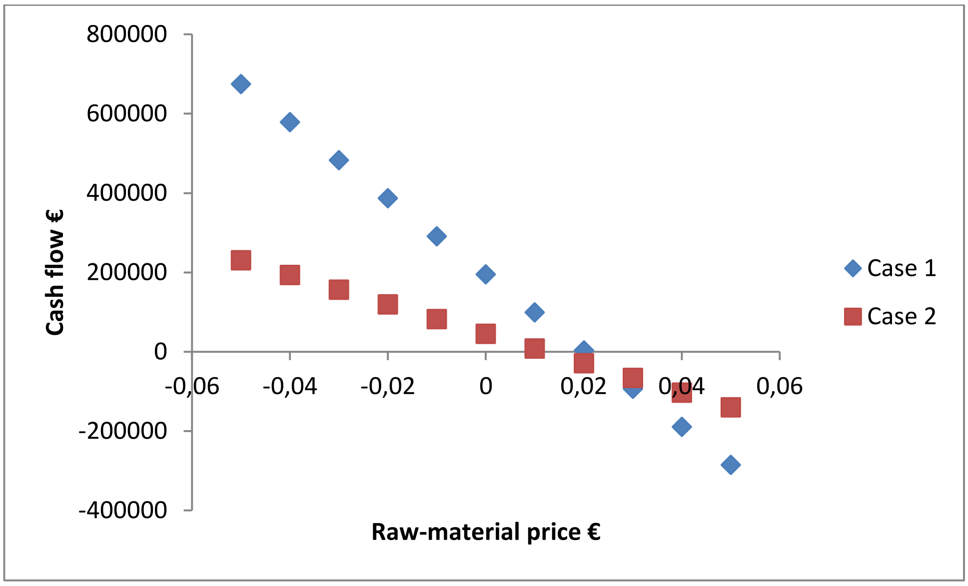

Figure 21. Raw material price sensitivity analysis.

6.2.2 Water cost

The price of water was assumed based on a reference value. One aspect affecting water price in case 2 is the cost of a water container. In case 1, no container is needed. The results are presented in Figures 19 and 20.

It can be conducted that water price has very little impact on the overall feasibility.

6.2.3 Cost of electricity

Variations in the cost of electricity and the effect on feasibility for case 1 is shown in Figure 19. The price of electricity has a high sensitivity as electricity is a large part of the operational costs.

6.2.4 Cost of diesel

In case 2 we use a generator to produce the required electricity. The price of diesel assumed was the average price for diesel in Europe at the time of writing. Fluctuations in fuel prices are not uncommon, thus we study the effect on feasibility. The results are shown in Figure 20.

6.2.5 Cost of water treatment

The cost of water treatment was assumed based on a reference value. Fluctuations in treatment price can occur depending on the location of the HTC processing unit. Storage of the liquid phase can also be included in the treatment price. It can be seen that case 1 is much more sensitive, due to more liquid recycling in the process.

6.2.6 Carbon dioxide taxation

It was assumed that there is no taxation on the carbon dioxide released as the process uses waste biomass as a raw material. In case taxation is applied, it is good practice to study the effects of the costs. It can be seen that case 2 is more sensitive due to higher carbon dioxide emissions.

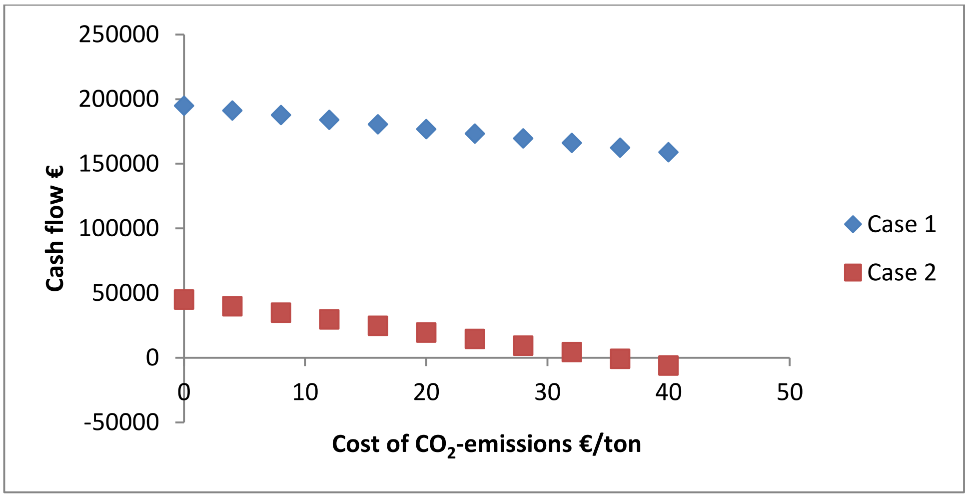

Figure 22. Sensitivity analysis for carbon tax.

6.2.7 Cost of transport

The price of transport was assumed based on a reference. Changes are bound to happen in the price of transportation, due to e.g. different pricing for different transport companies or varying gasoline prices.

6.2.8 Cost of labor

The salaries were assumed based on a reference. Costs may vary due to e.g. variations in the skill of labor or demands from labor unions. Case 2 is much more sensitive due to double amount of workers compared to case 1.

6.2.9 Hydrochar price

As a large scale hydrochar market is virtually non-existent, a sensitivity on the final price of the product was performed. The results are shown in Figures 19 and 20.

6.2.9 Working capital

For analysis of the working capital, we study changes in the DCFROR. Note that the sensitivity analysis for case 2 is done with the assumed product price of 0.7 €/kg, because DCFROR does not reach positive values with a product price of 0.6 €/kg.

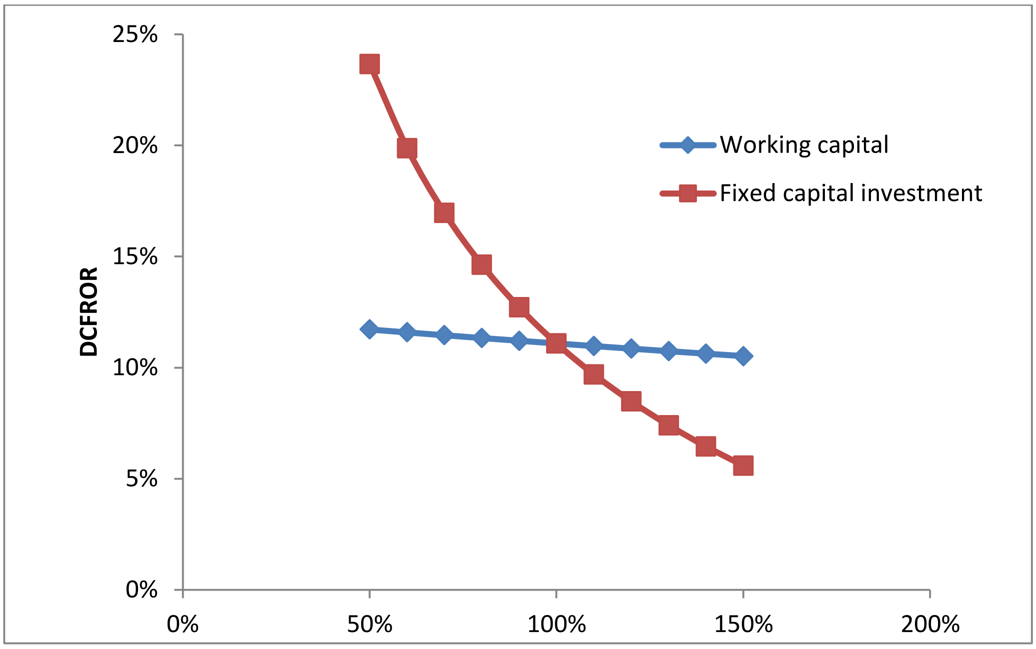

Figure 23. Sensitivity analysis for working capital and fixed capital investment for case 1 with discounted cash flow rate of return as a measure of feasibility.

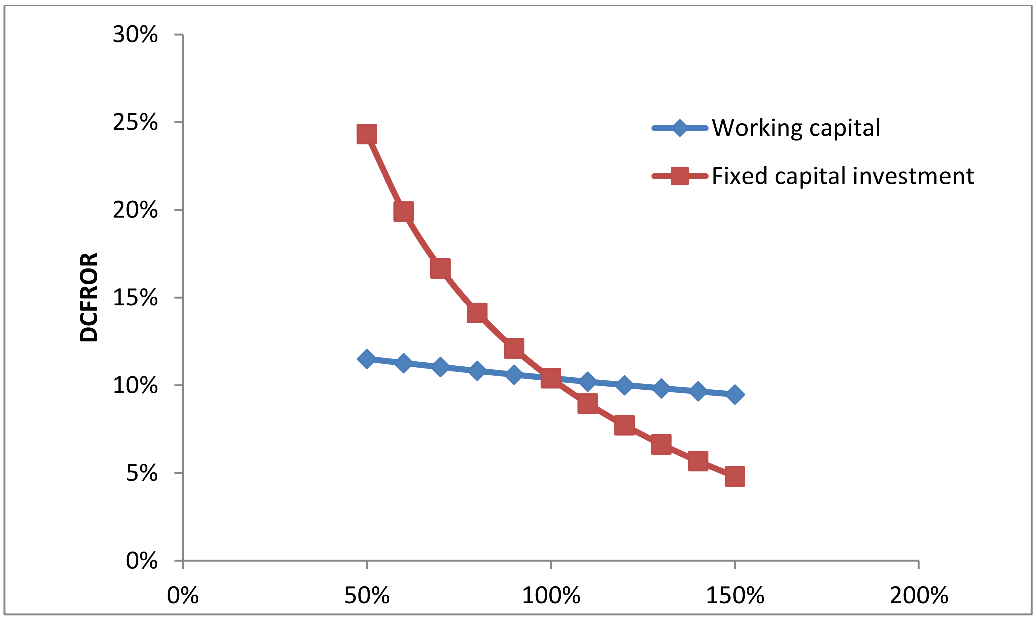

Figure 24. Sensitivity analysis for working capital and fixed capital investment for case 2 with discounted cash flow rate of return as a measure of feasibility. Note that the analysis is done for a product price of 0.7 €/kg.

6.2.10 Capital investment

Changes in capital investment are due to error or changes in the initial assumptions made in this thesis. The results of the sensitivity analysis are presented in Figures 23 and 24. Just as with the working capital sensitivity analysis, it should be noted that the sensitivity analysis for case 2 is done with the assumed product price of 0.7 €/kg.

7 Conclusions

A design for a mobile HTC-processing unit has been presented. It can be stated that the HTC process can be made mobile by placing it in a 12 m shipping container, with 3 CSTR reactors processing biomass at 220 °C and 3000 kPa for 2 hours. The maximum product output is 288 kg/h. Heat is recovered with two flash tanks and one spiral heat exchanger. The solid product is separated from the liquid by a filter press. The process liquid is recycled and mixed with the biomass feed. The equipment was sized with the help of a 3D model and the price of the process was estimated. Using price correlations, the capital cost of the process was estimated to be 1.2 million €. It can be disclosed that the limiting factor for the process is the space presented by the container. As the process is mobile the size cannot be expanded freely. If the capacity is to be increased, the reaction time has to be shortened.

Two business cases were used to measure the economic feasibility of the processing unit. The first case studied a processing unit connected to a medium sized brewery in an urban area, receiving utilities from the brewery. The raw material for this case is brewer's spent grain, a waste from the brewing industry. The second case studied the processing unit utilizing wheat straw from large farms in southern Europe. As the unit is at a remote location, a generator is required to produce energy. Wheat straw is also considered a waste. In both cases the annual operating time was maximized, taking into consideration transport and start up times. As no large scale market for hydrochar exists, the price was assumed to be 0.6 €/kg. It can be stated that at the assumed price and a project life of 15 years, the process is not feasible. Case 1 gives an 11 % return on investment and case 2 cannot produce any return in the given time. It should however be established, that these are only two of the possible business cases for the HTC-processing unit. Other cases might have a much higher feasibility.

Many assumptions were made for the design and business cases and variations in these are bound to occur, should the process be implemented. Thus, a sensitivity analysis was conducted for the process variables and business case variables. The process variables that affect the feasibility most are reactor feed concentration and density and reaction time, as these affect the amount of raw material that can be processed. Of the business variables, the most sensitive is product price. Salaries and electricity costs are also sensitive, as these make up the bulk of the operating costs.

7.1 Recommendations for future research

As the business cases studied in this work proved to be unfeasible, future research on different business cases should be conducted. HTC is a flexible process regarding raw material, resulting in a plethora of possible business cases. Applications for hydrochar, especially for soil amendment, also require further studies. As electricity and salaries are the largest operating costs, future research could look into ways to reduce these. This could be done e.g. by burning part of the char for energy, by further automating to process or by using a business case where the processing unit is rented to farmers that use the product to improve their own soil.

As many of the variables were assumed in this work, more correct data could aid future assessments. Data on biomass slurries is severely lacking, particularly for density, viscosity, heat capacity and heat exchange coefficients. No correlation between process variables and product quality exist to date. These could greatly assist in further studies concerning the feasibility of HTC

One design aspect that could greatly improve the economic feasibility of the processing unit is wet air oxidation of the liquid fraction. If the oxidation could release energy to the process, operational costs would be reduced both for electricity and water treatment.

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Appendix 1 – Table of HTC reactions from different sources

Raw-material	Temperature (°C)	Time (h)	Dry matter Content (%) raw material	HTC yield (dry wt-%)	Ash (dry wt-%)	HHV (MJ/kg)	Source
Saccharides							
sucrose	190	4.5	0.5 [b]	23			(Sevilla and Fuertes, 2009)
glucose	170			1.5			
	180			5.1			
	190			9.4			
	210			28			
	230			36			
	170	15		6			
	180			15			
	190	4.5	1 [b]	26			
	230	1		31			
	240			43			
	240	0.5		37			
Starch	180	4.5	0.5 [b]	7.1			
			0.25 [b]	5.1			
	200			25			
	200		0.1 [b]	15			
Cellulose	170	4		94.98		17	(Álvarez-Murillo et al., 2016)
	170	6.8		61.5		17.8	
	180	5.3		92.4		17	
	180	11.8		45.3		20.4	
	210	7.6		43.2		25	

Raw-material	Temper-ature (°C)	Time (h)	Dry matter Con-tent (%) raw mate-rial	HTC yield (dry wt-%)	Ash (dry wt-%)	HHV (MJ/kg)	Source
Cellulose	210	11.8		44.4		27.1	(Álvarez-Murillo et al., 2016)
	235	2		99.5		17	
	235	9.8		42.7		25.1	
	245	2.6		92.3		17.1	
	245	3.8		39.6		21.7	
	245	6.8		44.1		28.2	
Biomass							
Empty fruit bunch (palm oil)	220	4	17	56,6	3,00	27,43	(Stemann et al., 2013a)
Empty fruit bunch (palm oil) – process water recycled	220	4	17	58,6	3,62	28,03	(Stemann et al., 2013a)
Poplar	180	8		89.9	2.2		(Wiedner et al., 2013)
	210			66.9	2.7		
	230			51.9	3.4		
Olive Residues	180			75.4	4.3		
	210			63.3	4.4		
	230			49	4.3		
Wheat straw	180			80.1	4.3		
	210			54.6	2.8		
	230			53.7	2.8		

Raw-material	Temper-ature (°C)	Time (h)	Dry matter Con-tent (%) raw mate-rial	HTC yield (dry wt-%)	Ash (dry wt-%)	HHV (MJ/kg)	Source
Orange peel	180-230	6-8		37.3	3.2-6.8	24.5	(Burguete et al., 2015)
Pepper farm residue	180-230	6-8		53.6	11.4-19.7	19.6	
Microalgae	190	0.5	5	28.4			(Heilmann et al., 2010)
		0.5	25	45.7			
		2	5	29.3			
		2	25	42.9			
	200	1.25	15	39.3			
		1.25		39.0			
		1.25		37.4			
		1.25		38.1			
	210	0.5	5	27.9			
		0.5	25	42.1			
		2	5	25.3			
		2	25	38.8			
Grape pomace	180	1	17	86	3.67	24.43	(Petrović et al., 2016)
	200	1		78	3.55	25.72	
	220	1		66	6.75	26.13	
Wheat straw	180	2	17	59.6	19.3	22.2	(Reza et al., 2015)

Raw-material	Temper-ature (°C)	Time (h)	Dry matter Con-tent (%) raw mate-rial	HTC yield (dry wt-%)	Ash (dry wt-%)	HHV (MJ/kg)	Source
Wheat straw	180	4	17	74.6	29.4	20.9	(Reza et al., 2015)
		6		69.8	26.4	20.7	
		8		73.6	25.1	21.7	
	200	2		58.7	19.8	21.7	
		4		63.6	24.7	24	
		6		63.5	24.2	22	
		8		57	21.6	22.9	
	220	2		55.3	21.4	23.4	
		4		48.8	19.4	25.5	
		6		57.9	26.8	26.3	
		8		53.3	24.6	27.7	
	240	2		56.8	28.6	27.3	
		4		51.6	25.8	28.6	
		6		49.3	24.7	29.5	
		8		50.5	25.7	29.5	
	260	2		45.9	20.3	23.3	
		4		51.1	25	27.6	
		6		57.6	32.6	26.3	
		8		53.3	27.8	29.1	
Dry leaves	200	0.5	17	70.98	11.08	16.81	(Saqib et al., 2015)
	210			63.13	13.64	17.66	
	220			61.12	15.89	18.98	
	230			63.43	21.04	18.1	

Raw-material	**Temper-ature (°C)**	**Time (h)**	**Dry matter Con-tent (%) raw mate-rial**	**HTC yield (dry wt-%)**	**Ash (dry wt-%)**	**HHV (MJ/kg)**	**Source**
Dry leaves	240	0.5	17	67.41	11.08	19.98	(Saqib et al., 2015)
	250			57.39	19.19	18.29	
Tahoe mix	155	0.5	11	78.82			(Hoekman et al., 2012)
	175			73.4			
	195			68.7			
	215			63.68			
	235			48.53			
	255			51.76			
	275			50.1			
Tahoe mix	295	0.5	11	78.82			(Hoekman et al., 2012)
Loblolly pine	175			77.71			
	200			70.07			
	215			72.43			
	235			63			
	255			50.13			
	275			48.18			
	295			52.37			
Pinyon/Juniper	175			77.38			
	215			71.54			
	235			62.73			
	255			50.58			
	275			49.63			

Raw-material	Temper-ature (°C)	Time (h)	Dry matter Con-tent (%) raw mate-rial	HTC yield (dry wt-%)	Ash (dry wt-%)	HHV (MJ/kg)	Source
Pinyon/Juniper	295	0.5	11	48.65			(Hoekman et al., 2012)
Sugarcane bagasse	175			69.63			
	215			63.77			
	235			59.09			
	255			45.01			
	275			44.57			
	295			42.75			
Corn stover	175			67.03			
	215			56.62			
	235			56.41			
	255			40			
	275			42.39			
	295			37.98			
Rice hulls	175			78.48			
	215			72.57			
	235			64.33			
	255			55.44			
	275			54.55			
	295			52.23			
Waste							
Stabilized sewage sludge	205	7	9	0,62		11,35	(Escala et al., 2013)

Raw-material	**Temper-ature (°C)**	**Time (h)**	**Dry matter Con-tent (%) raw mate-rial**	**HTC yield (dry wt-%)**	**Ash (dry wt-%)**	**HHV (MJ/kg)**	**Source**
Sewage sludge	205	7	25[b]	0,60		18,78	(Escala et al., 2013)
Paper waste	250	20	20	29.2	24.2	23.9	(Berge et al., 2011)
Food waste	250	20	20	43.8	11.2	29.1	
mixed municipal solid waste	250	20	20	63.2	46	20	
Anaerobic digestion waste	250	20	20	47.1	55.8	13.7	

[a] citric acid catalyst, [b] calculated from the raw-materials moisture

[b] concentration mol/l

Appendix 2 – List of brewery capacities in Europe

Brewery	Country	Capacity (100l/a)	Source
Stadin panimo	Finland	900	(Stadin Panimo, 2016)
Nokian panimo		25000	(Nokian Panimo, 2013)
Stallhagen		6820	(Stallhagen, 2016)
Koskipanimo		2000	(Tampereen panimoravintola Oy, 2013)
Saimaan Juomatehdas		7000	(Saimaan Juomatehdas, 2016)
Maku		3000	Juhani Repo[a]
Ruosniemen panimo		2080	Juho-Matti Karpale[a]
Suomenlinnan Panimo		10000	Jussi Heikkilä[a]
Huvila		250	(European Beer Guide, 2016)
Olvi		55000	
Ægir Bryggeri	Norway	2000	
Grans Bryggeri		80000	
Hansa Bryggeri		600000	
Lillehammer Bryggeri AS		500	
Nøgne		80000	
Nordkapp Mikrobryggeri		400	
Ringnes E.C.Dahls		500000	
Trondhjem Mikrob.		1200	
Einstök	Iceland	120000	info@einstokbeer.com[a]

Brewery	Country	Capacity (100l/a)	Source
Bredlunds Bryghus	Denmark	1000	(European Beer Guide, 2016)
Carlsberg Bryggerierne A/S		2.5 million	
Herslev Bryghus		1500	
Holbæk Bryghus		750	
Bryghuset Kragelund		800	
Rise Bryggeri A/S		240	
Falkenberg		1,3 million	
Jämtlands Bryggeri AB	Sweden	3500	
Kopparbergs Bryggeri AB		300000	
Nynäshamns Ångbryggeri	Sweden	720	
Oppigårds Bryggeri		3500	
Spendrups Bryggeri AB		1,24 million	
AB Åbro Bryggeri		300000	
Brasserie Battin	Luxembourg	12000	

Brewery	Country	Capacity (100l/a)	Source
Restaurant Beierhaascht	Luxembourg	1200	(European Beer Guide, 2016)
Brasserie Bofferding		168000	
Cornelyshaff		800	
Brasserie Simon		23000	
Franciscan Well		2500	
Great Northern Brewery	Ireland	1 million	
Arthur Guinness Son & Co.		4 million	
Murphy Brewery Ireland Ltd.		1,03 million	
The Porterhouse Brewing Co.		2500	
E. Smithwick and Sons Ltd.		1,2 million	
Whitewater Brewing Co.		4500	
Sociedade Central De Cervejas	Portugal	4,8 million	
Cervejeira Lusitana		12000	

Brewery	Country	Capacity (100l/a)	Source
Uniao Cervejeira	Spain	2,4 million	(European Beer Guide, 2016)
Calvin's Beer		1000	
Damm		5 million	
Estrella de Levante		1 million	
Heineken España		900000	
Llúpols i Llevats S.L.		250	
La Zaragozana S.A.		500000	

[a] Information gathered through email correspondence

Appendix 3 – Assumptions made in the design and feasibility analysis

Variable	Value
Biomass dry matter content	25 wt-%
Reactor feed dry matter content	10 wt-%
Reactor temperature	220 °C
Reactor pressure	3000 kPa
Reaction time	2 h
Solid product moisture content	30 wt-%
Conversion	100 %
Solid yield	50 %
Liquid yield	35 %
Gaseous yield	15 %
Density	1000 kg/m^3
Reactor insulation thickness	5 cm
Heat exchange coefficient	550 $W/(m^2K)$
Heat exchanger correction factor	0.95
Flash 1 pressure	290 psi
Flash 2 pressure	51 psi
Mixer correction factor	0.5
Mixer tank volume	673 l
Filter press operation time	10 min
Equipment and pipe material	Stainless steel 304
Pipe diameter	40 mm
Pipe diameter for mixing	90 mm
Heat exchanger pressure drop	100 kPa
Equal pipe length, 45 ° curve	15 m
Equal pipe length, 90 ° curve	30 m
Equal pipe length, valve	18 m
Equal pipe length, tank entry	50 m

Variable	Value
Equal pipe length, tank exit	25 m
Equal pipe length, T-crossing	60 m
Breweries processed per week	8
Wheat fields processed per year	5
Brewery size, production	1250 hl/month
Wheat field size, area of wheat	100 ha
Dollar-euro exchanger rate, 2016	0.902 €
Dollar-euro exchanger rate, 2003	0.883 €
Gulf coast-Europe cost multiplier	1.13
chemical engineering plant cost index 2016	556.8
chemical engineering plant cost index 2014	573.6
chemical engineering plant cost index 1989	355.4
Belt conveyor width	0.74 m
Belt conveyor lift height	2.6 m
Cost factor, Installation	0.5
Cost factor, Piping	0.6
Cost factor, Electrical	0.2
Cost factor, Instrumentation & control	0.3
Cost factor, Civil	0.3
Cost factor, Lagging, insulation and paint	0.1
Cost factor, Material	1.3
Cost factor, Design and engineering	0.25
Cost factor, Contingency	0.1
Price 40-foot ISO-container, used	1800 €
Raw material price	0 €/kg
Hydrochar price	0.6 €/kg
Water price	0.45 €/ton
Electricity price	0.0879 €/kWh
Diesel price	0.92 €/l

Variable	Value
Reactor feed heating efficiency	100 %
Pump efficiency	88 %
Pump motor efficiency	90 %
Filter press power consumption	25 kWh/ton
Belt conveyor, conveyor pulley friction	0.025
Belt conveyor, weight of rotating parts	56
Belt conveyor, frictional resistance in belts	56
Belt conveyor, belt velocity	0.008 m/s
Water treatment price	1.35 €/ton
Carbon tax	0 €
Transport capacity	14.5 ton
Transport price	0.93 €/km
Labor cost	29.5 €/h
Maintenance cost, percentage of ISBL capital costs	5%
Insurance, percentage of capital costs	1 %
Working capital, Value of product inventory, weeks operational cost	2
Working capital, Cash on hand, weeks operational cost	1
Working capital, Accounts receivable, weeks operational cost	4
Working capital, Accounts payable, weeks operational cost	4
Working capital, Spare parts inventory, percentage of investment costs	1 %