

POLITECNICO DI TORINO

Collegio di Ingegneria Chimica e dei Materiali

**Corso di Laurea Magistrale in Ingegneria Chimica e dei  
Processi Sostenibili**

TESI DI LAUREA MAGISTRALE

**Hydrothermal treatment of Vinasse**



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March 15, 2018



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# Riassunto

## 0.1 Introduzione

La limitata disponibilità di combustibili fossili rende necessaria la ricerca e lo sviluppo di alternative sostenibili. Diventa quindi indispensabile aumentare il consumo di energia ottenuta da fonti rinnovabili, penalizzando l'uso di combustibili fossili. Lo scopo di questo lavoro di tesi è quello di sviluppare un processo innovativo per l'ottenimento di combustibile liquido da quello che oggi è considerato un prodotto di scarto, ovvero le vinacce prodotte dalla lavorazione della canna da zucchero. Esse sono ottenute dal trattamento della melassa, un sottoprodotto della produzione dello zucchero. La melassa è sottoposta a un processo di fermentazione al termine del quale si ha la rimozione dell'etanolo per distillazione, le vinacce sono il residuo di questa distillazione. In figura 1 sono riportati gli step fondamentali che permettono la produzione delle vinacce a partire dalla canna da zucchero.

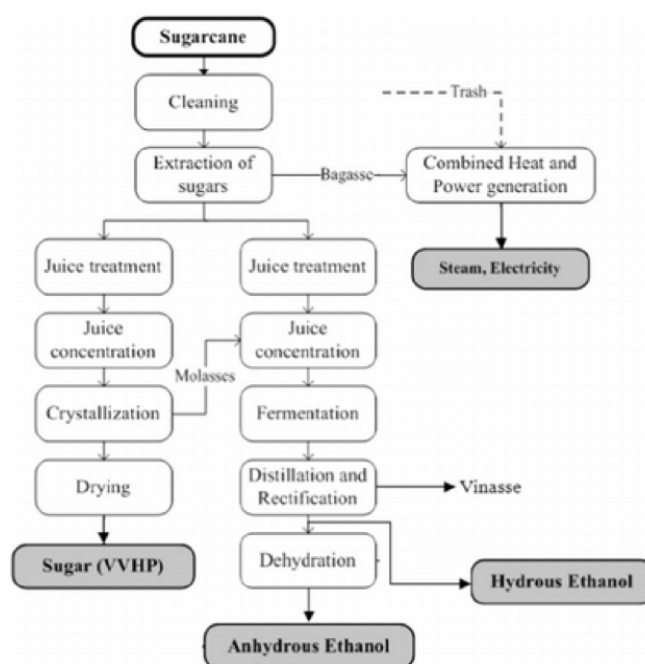


FIGURE 1: Step per la produzione delle vinacce [8].

Le vinacce derivanti dalla melassa sono costituite da acqua, composti organici e inorganici e la composizione varia a seconda degli impianti di trattamento della canna da zucchero per l'ottenimento di etanolo. Grazie all'alto contenuto organico hanno un buon potenziale come combustibile [5]. Inoltre, grazie all'elevato contenuto di umidità (che varia tra 40% e 50%) risultano una valida risorsa per i trattamenti idrotermici. Questi, infatti, sono condotti in presenza di acqua, diversamente dalla pirolisi, e di conseguenza non è necessario uno stadio di essiccamento del feed. I composti presenti in maggior quantità nelle vinacce sono acido lattico, acido acetico e glicerolo. Gli studi al momento condotti sulle vinacce sono principalmente basati su gassificazione convenzionale e combustione.

Tuttavia questi processi non si sono mostrati favorevoli a causa dell'elevato contenuto di umidità del feed. Per questo motivo, si è pensato di iniziare a valutare la gassificazione in acqua supercritica. La prima fase di questo lavoro di tesi è stato affiancato da esperimenti alla Åbo Akademi, in Turku, su *model compounds* per le vinacce. Quando l'acqua viene portata in condizioni supercritiche (al di sopra di 374°C e 221 bar) le sue proprietà cambiano in modo significativo. Si osserva una diminuzione della densità e della costante dielettrica. La caratteristica più interessante è però la bassa polarità, che la rende un grande solvente per le reazioni con feedstock ad alto contenuto organico. In condizioni supercritiche l'acqua ha una elevata reattività e le velocità di reazione sono più elevate in confronto ad altri processi, ad esempio la digestione anaerobica. Come risultato è possibile ottenere una buona conversione della biomassa. Il processo può essere condotto a temperature più basse (tra 374°C e 550°C) o più elevate (al di sopra di 550°C), e si avranno risultati diversi a seconda della temperatura di lavoro selezionata. Nel primo caso si osserva la formazione di un gas ricco in carbonio, quando usato un catalizzatore metallico o alcalino; nel secondo caso si ottiene un gas ad alto contenuto di idrogeno a causa della endotermicità della reazione di formazione dell'idrogeno. E' quindi necessario definire prima l'obiettivo del processo in modo da poter selezionare le ottimali condizioni di lavoro per raggiungerlo.

## 0.2 Obiettivi e metodi

La tesi ha l'obiettivo di sviluppare un processo innovativo per la produzione di combustibili liquidi dalle vinacce da canna da zucchero attraverso gassificazione in acqua supercritica. Si vuole valutare la possibilità che il processo sia un'alternativa sostenibile alle attuali soluzioni per la produzione di combustibili liquidi. Inoltre, si vuole studiare un nuovo modo di processare un sottoprodotto come le vinacce. Si vuole quindi:

- Sviluppare un processo innovativo e simulare l'impianto sul software Aspen Plus®V8.8
- Valutare la fattibilità del processo da un punto di vista economico, confrontandolo con altri processi di trattamento della biomassa per produzione di combustibile

I valori delle portate, solventi e reagenti sono state prima valutate su Microsoft Office Excel e poi implementate su Aspen Plus®. I metodi e le equazioni di stato selezionate per la simulazione sono quelli che sono in grado di predire al meglio l'evoluzione del sistema, e sono stati selezionati dopo uno studio di letteratura. Lo scenario è stato analizzato con il metodo della *pinch analysis* per determinare la richiesta di utility esterna e per valutare le efficienze termiche globali. Infine, il potenziale commerciale dell'impianto è stato calcolato tramite la determinazione del costo minimo di vendita e del valore attuale netto relativi alle condizioni di mercato.

## 0.3 Sviluppo del processo

Il progetto ha l'obiettivo di valutare la possibilità di trasformare le vinacce in combustibile liquido e analizzare la fattibilità del processo anche da un punto di vista economico.

La modellazione matematica del processo e la sua simulazione sono state effettuate con il simulatore di processo Aspen Plus. I modelli termodinamici usati sono stati scelti in base a quanto riportato in letteratura. Essi sono l'equazione di Peng Robinson Boston and Mathias (PR-BM), adatta a sistemi gassosi, l'ENRTL-RK per le apparecchiature contenenti soluzione acquose, e la BK10 quando si è in presenza di idrocarburi pesanti. A partire da un flowsheet vuoto sono stati valutati e simulati

i vari passaggi necessari per la produzione del combustibile. Il processo è stato costruito a blocchi valutando di volta in volta i parametri e definendo le condizioni di lavoro necessarie per raggiungere determinate rese ed efficienze. In figura 2 sono riportate le fasi fondamentali sviluppate, dove i numeri presenti sulle correnti fanno riferimento allo schema di processo riportato nell'appendice della tesi.

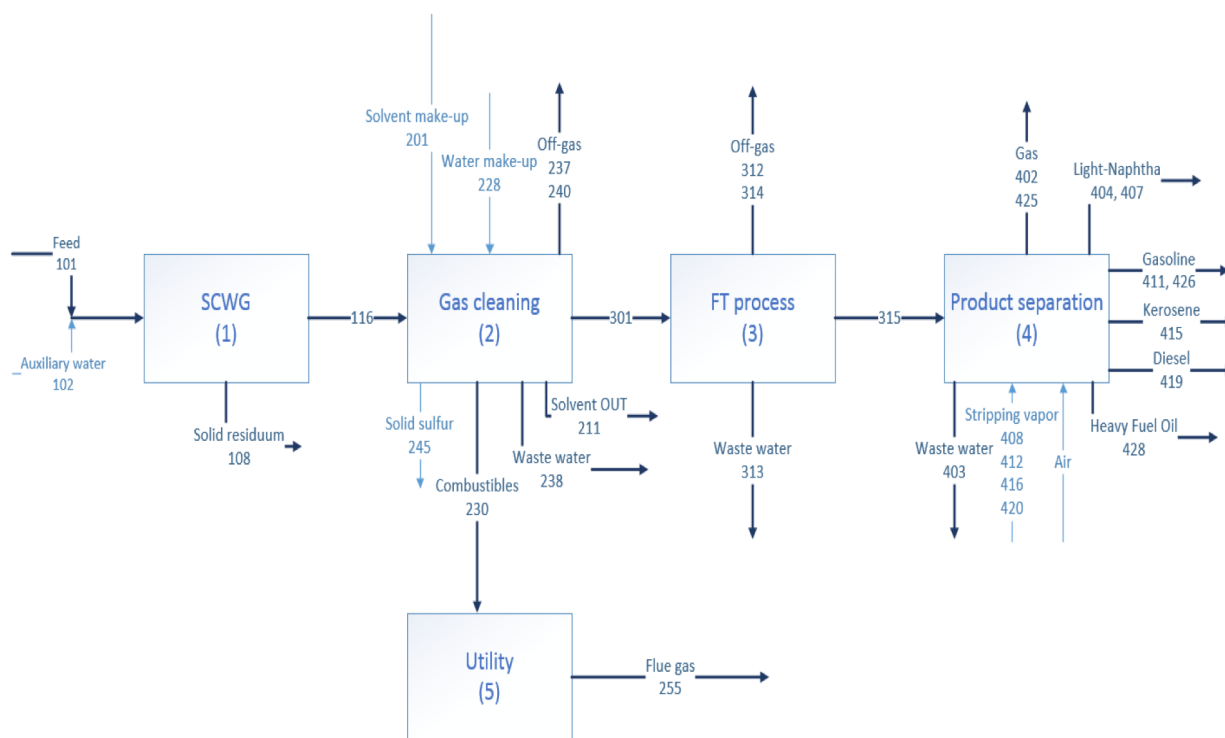


FIGURE 2: Diagramma a blocchi dell'impianto.

Come si può vedere dalla figura 2 i principali stadi sono:

1. Gassificazione in acqua supercritica (SCWG)
2. Purificazione del gas
3. Fischer Tropsch
4. Separazione dei prodotti
5. Utility

Ogni stadio è stato valutato tenendo in considerazione le condizioni a valle e a monte di esso, e quindi una volta posti i target sono state fatte analisi di sensitività e design specification per raggiungerli. La quantità di vinacce che arrivano al processo è stata scelta tramite studi di letteratura e pari a 230 tonnellate al giorno. Questo valore è stato ottenuto considerando che la quantità di vinacce ottenute dalla produzione di bioetanolo è circa 23 kg per litro di etanolo. Inoltre, è stato considerato che la capacità di un impianto di produzione di etanolo su scala industriale è di circa  $10 \text{ m}^3$  di etanolo al giorno [50]. Per simulare il comportamento delle vinacce, sono stati selezionati tre *model*

*compounds* (acido lattico, acido acetico e glicerolo) e le loro concentrazioni sono state valutate come media dei valori trovati in letteratura per le vinacce da canna da zucchero [21], [22], [23], [12]. Siccome acido lattico, acido acetico e glicerolo sono i componenti presenti in maggiore quantità ma non gli unici, un componente non convenzionale (CHOSN) è stato definito per bilanciare il contenuto organico, e la sua quantità è stata valutata tramite un calcolo iterativo. I componenti solidi (ceneri e contenuto organico) sono stati definiti come *non conventional* e da un punto di vista termodinamico sono stati simulati sulla base delle correlazioni di entalpia e densità del carbone, rispettivamente HCOALGEN e DCOALGEN. In tabella 1 è riportata la composizione dei componenti presenti nel feedstock per la simulazione.

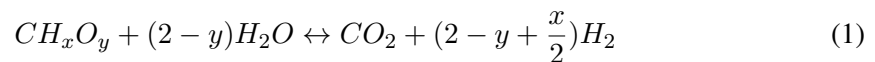
TABLE 1: Concentrazione dei componenti nel feedstock.

Compound	%wt
Glycerol	6.46
Lactic acid	14.44
Acetic acid	4.87
Ash	15.33
Moisture	49.16
$C_v H_w O_x N_y S_z$	9.75

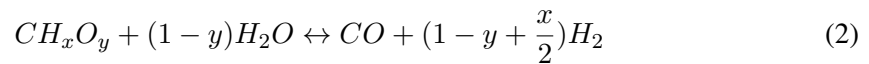
Come si può notare, il contenuto di acqua è piuttosto elevato e per questo sono preferibili trattamenti idrotermici, e quindi che richiedono acqua come solvente ausiliario, rispetto a quelli termici come la gassificazione convenzionale, infatti in questo caso sarebbe necessaria una grande quantità di energia per essiccare il feed.

### 1. Gassificazione in acqua supercritica (SCWG)

Questo stadio ha l'obiettivo di convertire la biomassa solida in un gas. Il processo è complesso, ma la conversione del feed in idrogeno può essere semplificata come:



Allo stesso tempo, però, possono avvenire delle reazioni secondarie quali steam reforming (equazione 2), water gas shift (equazione 3) e metanazione (equazione 4).



I parametri che influenzano il decorso della reazione sono temperatura, pressione e contenuto di solidi. La scelta delle ottimali condizioni di lavoro è fondamentale in quanto a diverse condizioni di temperatura si ottengono diverse concentrazioni di prodotti, dovuti alla endotermicità e esotermicità delle reazioni. Infatti, per ottenere un syngas ricco di idrogeno è necessario lavorare a temperature più elevate. A differenza dalla gassificazione convenzionale, con la gassificazione in acqua supercritica si ottiene un syngas con un maggior contenuto di  $CO_2$ ; questo

è causato sia dall'ossigeno presente nella biomassa che da quello presente nell'acqua ausiliaria che permette una migliore ossidazione.

Le condizioni di lavoro di questo step sono state selezionate in base agli studi sperimentali condotti sulla gassificazione in acqua supercritica di *model compounds* per le vinacce alla Åbo Akademi con l'obiettivo di produrre idrogeno. Quindi, è stata selezionata una temperatura pari a 600°C, una pressione pari a 250 bar ed un contenuto di solidi pari al 10% in modo da minimizzare la probabilità di ostruzione dei tubi di alimentazione. La fase di gassificazione in acqua supercritica è stata simulata in due step: un reattore RYIELD per decomporre il feedstock nei suoi elementi costitutivi, e un reattore RGIBBS per valutare la formazione dei prodotti. Questo design permette la simulazione di un'idrolisi, che permette la rottura dei legami chimici, e le successive reazioni 1, 2, 3, 4 [16].

E' stata assunta una conversione del 98% del carbonio nei prodotti, di conseguenza il restante 2% esce dall'impianto con le ceneri.

## 2. Purificazione del gas

Questo step è stato valutato a causa dell'elevato contenuto di zolfo presente nel feedstock. I catalizzatori presenti nelle fasi successive hanno una bassa tolleranza allo zolfo, di conseguenza deve essere rimosso. Il catalizzatore del processo di Fischer-Tropsch è senza dubbio quello che crea maggiori problematiche. Esso infatti ha una tolleranza molto limitata allo zolfo (minore di 2 ppm), e ciò ha richiesto l'inserimento e la valutazione di numerose apparecchiature per la rimozione dello zolfo. A seguito di un attento studio di letteratura, sono stati valutati i processi più adatti alla rimozione dell'  $H_2S$  che si forma nel reattore di gassificazione, e sono stati simulati. Insieme alla rimozione di  $H_2S$  è stato anche valutato un recupero di  $CO_2$ , per limitare la perdita di carbonio dell'impianto e massimizzare la resa in combustibile liquido, e una purificazione della corrente contenente idrogeno.

In figura 3 è riportato un diagramma a blocchi dello step di purificazione del syngas.

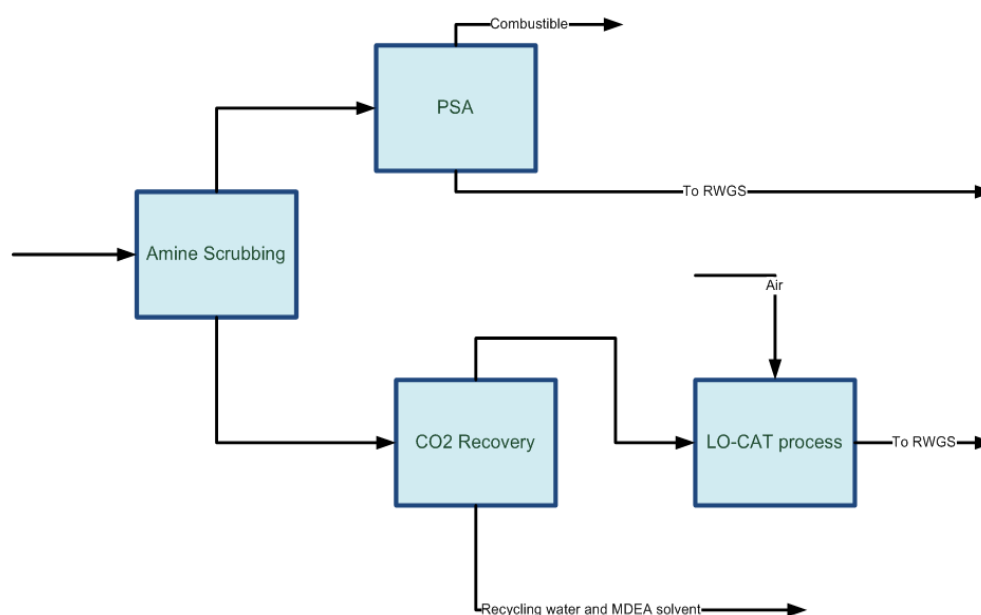


FIGURE 3: Diagramma a blocchi dello step di purificazione del gas.

- Amine scrubbing

Il processo avviene in un estrattore in cui è alimentata in cima alla colonna una soluzione al 30% di MDEA e 70% acqua, e sul fondo della colonna viene alimentato il gas da purificare. La MDEA permette la rimozione di 0,4 moli di gas acido per mole di ammina [38]. Una prima valutazione è stata fatta su Microsoft Office Excel considerando l'estrazione come ideale, e quindi una portata costante in ingresso e in uscita dall'estrattore. Con questa ipotesi è stato valutato tramite un metodo iterativo il numero di stadi necessari a ridurre il contenuto di  $H_2S$  sotto le 4 ppm. Il risultato è stato poi implementato su Aspen Plus<sup>®</sup>. Un'analisi di sensitività ha permesso la determinazione delle condizioni ottimali di lavoro e della portata di solvente necessaria.

- Pressure Swing Adsorption (PSA)

Come risultato dell'*amine scrubbing* si ottiene in coda una corrente che contiene principalmente  $H_2S$  e  $CO_2$ , e in testa un gas composto per l'80% da idrogeno. Questa fase è stata in realtà aggiunta dopo lo sviluppo del processo di Fischer-Tropsch. Esso, infatti, necessita di un reattore di reverse water gas shift (RWGS) a monte. Di conseguenza la PSA è stata inserita in modo da avere una corrente ad alta purezza di idrogeno da mandare al reattore di RWGS.

- Recupero della  $CO_2$

Questa sezione è stata valutata per due motivi: per recuperare il solvente contenente MDEA in modo tale da poterlo ricircolare e quindi ridurre il costo legato all'uso di solvente *fresh*, e per minimizzare le perdite di carbonio dell'impianto. La  $CO_2$  recuperata, infatti, viene alimentata al reattore di RWGS. Allo stesso tempo, tramite water scrubbing, si permette la formazione di una corrente ad alto contenuto di metano che viene usata poi come utility calda per l'impianto.

- Processo LO-CAT<sup>®</sup>

E' stato aggiunto per permettere un'ulteriore riduzione del contenuto di  $H_2S$  da inviare al reattore di Fischer-Tropsch. Una parte di  $H_2S$ , infatti, viene recuperata insieme alla  $CO_2$  a causa della loro solubilità in acqua. Con il processo LO-CAT<sup>®</sup> si permette la conversione del 99.99% di  $H_2S$  a zolfo solido tramite l'immissione di aria.

Lo sviluppo della fase di purificazione del gas è senza dubbio quello che ha causato il maggior numero di problema da un punto di vista di convergenza. Volendo limitare la perdita di carbonio, infatti, sono stati riciclati al *Recupero della  $CO_2$*  tutti gli off-gas del processo contenenti una larga percentuale di  $CO_2$ . Questo non solo ha causato problemi di convergenza, dovuti essenzialmente alla complessità del flowsheet, ma ha comportato una modifica delle concentrazioni delle correnti, e di conseguenza numerose analisi di sensitività e design specification sono state implementate per ottenere determinati target. Ad esempio, due design specification sono state utilizzate per controllare il corretto rapporto tra reagenti in ingresso sia al reattore di RWGS che al reattore di Fischer-Tropsch.

### 3. Fischer Tropsch (FT)

Il processo di Fischer-Tropsch deve avvenire con uno specifico rapporto dei reagenti in ingresso. Per questo motivo, a valle del reattore di FT è stato aggiunto un reattore di reverse water gas shift (RWGS).

- Reverse water gas shift (RWGS)

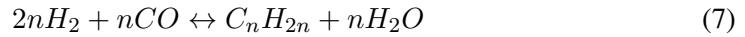
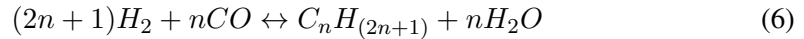
La reazione di reverse water gas shift (equazione 5) è debolmente endotermica, quindi favorita ad alte temperature.



Uno studio di letteratura ha permesso di determinare il migliore catalizzatore che permettesse una buona resistenza ad alte temperature ed un'elevata selettività verso la  $CO$ , e che quindi limitasse le reazioni secondarie. La composizione di equilibrio varia fortemente con la temperatura e con il rapporto dei reagenti. Infatti, per esempio, a basse temperature si ha la formazione di metanolo e metano. Il catalizzatore che è stato selezionato è il  $NiO/CeO_2$  che permette una conversione della  $CO_2$  del 40%, con una selettività verso la  $CO$  di circa 100%. Come già anticipato, la PSA permette di avere una corrente ad alta purezza di idrogeno in ingresso al reattore di RWGS. Infine, il rapporto  $H_2/CO_2$  necessario per il corretto funzionamento della reazione, è regolato tramite una design specification da uno split presente nello step di purificazione del gas.

- Fischer-Tropsch (FT)

Il reattore di FT permette la sintesi di un combustibile liquido (*crude*) dalla  $CO$ : la  $CO$  reagisce con l'idrogeno per produrre paraffine, olefine e acqua [59]:



Il corretto funzionamento della reazione si ha per un rapporto  $H_2/CO$  pari a 2. Per questo motivo è stata utilizzata una design specification per garantire il rapporto dividendo la corrente in uscita dalla PSA per alimentare parzialmente il reattore di RWGS e parzialmente quello di FT. Un parametro che condiziona fortemente la distribuzione di idrocarburi in uscita dal reattore di FT è il fattore di crescita della catena ( $\alpha$ ). Esso dipende dalla temperatura di lavoro e dalla concentrazione  $CO$  e  $H_2$  in ingresso e varia tra 0 e 1. Per questo studio è stato selezionato un valore di  $\alpha$  pari a 0,82 e una temperatura di lavoro di 220°. Il catalizzatore scelto è a base di ferro, che ha una maggiore tolleranza allo zolfo rispetto al catalizzatore a base di cobalto, con una conversione della  $CO$  pari al 72%.

#### 4. Separazione dei prodotti

La separazione dei prodotti è necessaria poichè la corrente ottenuta dal reattore di FT non può essere usata direttamente come combustibile. Essa infatti è costituita da acqua,  $CO_2$  e tutti gli idrocarburi da  $C_1$  a  $C_{30}$  che si sono formati. Si ha quindi una prima rimozione di acqua e off-gas tramite un separatore a tre fasi. Successivamente, la separazione avviene tramite tre stadi successivi:

- Separazione della light naphta:

Realizzata tramite una colonna con un condensatore totale e un ribollitore parziale. Il rapporto di riflusso e il rapporto distillato to feed sono stati valutati tramite analisi di sensitività per permettere la separazione dei componenti leggeri in testa e i componenti pesanti in coda.

- Produzione di gasolio, cherosene e diesel:

E' stata realizzata tramite una colonna atmosferica. Il frazionamento nei vari distillati è stato ottenuto cercando di adattare la colonna di distillazione del greggio da petrolio al crude ottenuto dal processo di FT. Le curve true boiling point (TBP) and the ASTM D86, che sono utilizzate per caratterizzare i vari tagli del greggio, sono state valutate tramite il tool *Stream Analysis* di Aspen Plus®. Il crude alimentato alla colonna è stato definito come blend dei diversi tagli, conoscendone la concentrazione. La colonna è stata modellata come una colonna a 25 stadi con una fornace, che serve a vaporizzare circa il 70% del feed, tre strippers e un pump around. Gli stripper permettono la distillazione dei vari tagli: gasolio, cherosene e diesel. Quello che rimane sul fondo ( $C_{22+}$ ), può essere ulteriormente trattato per massimizzare la produzione di combustibile, riducendo al minimo il residuo. Per questo è stato aggiunto lo step di fluid catalytic cracking.

- Fluid catalytic cracking (FCC):

Il processo permette il cracking degli idrocarburi pesanti in gas e gasolio. La conversione avviene in un reattore verticale con un tempo di contatto molto breve a cui viene alimentato del vapore insieme al residuo. Per la simulazione su Aspen Plus® è stato utilizzato un reattore RYIELD in cui sono state inserite le rese per ogni frazione date dalla letteratura [43]. Le frazioni sono state definite all'inizio del flowsheet come *pseudo-components*.

## 5. Utility

L'ultimo blocco riportato in figura 2 è stato inserito per cercare di sfruttare i combustibili presenti nelle correnti in uscita dal processo, in modo da minimizzare la necessità di una utility calda esterna. Infatti, sia dal water scrubbing che dal PSA si ottengono due correnti ad alto contenuto di metano che hanno un grosso potenziale come fonti di calore. Per controllare il rapporto combustibile-comburente è stata utilizzata una design specification, per avere una quantità di ossigeno nella corrente di uscita pari a circa 6% in peso.

## 0.4 Valutazioni energetiche e economiche

Una volta completato il design dell'impianto è stata fatta una valutazione di tipo energetico e una di tipo economico.

- Analisi energetica

Il processo è stato analizzato con il metodo della pinch analysis per valutare la richiesta di utility esterna e per determinare l'efficienza termica. L'analisi è stata effettuata con il tool Aspen Energy Analyzer che permette di determinare i target di utility richiesta. Attraverso Aspen Energy Analyzer è anche stato possibile ricavare le configurazioni ottimali per la rete di scambiatori di calore. Sono state condotte numerose iterazioni per cercare di raggiungere la rete ottimale di scambiatori di calore, fino a selezionare due configurazioni. Nella prima configurazione non è stato posto alcun limite sulla disposizione degli scambiatori, quindi permettendo lo scambio di calore tra le diverse sezioni. Nella seconda sono state poste delle restrizioni in modo tale da ridurre la complessità del processo da un punto di vista costruttivo e permettere lo scambio solo tra correnti appartenenti alla stessa sezione. Le due configurazioni sono riportate nell'appendice E. In entrambi i casi sono stati ottenuti un eccessivo numero di scambiatori (33 e 34 unità) ed inoltre alcune correnti attraversano un elevato numero di scambiatori, cosa poco



fattibile da un punto di vista pratico. Sarebbe dunque necessario ottimizzare e ridurre il numero di scambiatori.

- Analisi economica

L'analisi economica permette di valutare i profitti che si possono ottenere dal processo e la sua remuneratività. La valutazione è stata fatta con la metodologia riportata in ref. [72]. Permette di determinare il prezzo minimo di vendita prendendo in considerazione i costi capitali e operativi relativi a 20 anni di vita dell'impianto. Il *purchased equipment cost* è stato valutato in due modi diversi per apparecchiature convenzionali e non. Per le prime è stato utilizzato il tool Aspen Plus<sup>®</sup>Economic, mentre per le altre, per esempio i reattori di SCWG, FT e RWGS, i valori sono stati adattati da quelli di letteratura considerando un fattore di scala relativo alla capacità, e sono stati riportati all'anno relativo alla messa in moto dell'impianto (2017) tramite il Chemical Engineering Plan Cost Index (CEPCI). I costi operativi (O&M) sono divisi in fissi e variabili: i fissi includono stipendi degli impiegati e assicurazione, mentre i variabili dipendono dall'operatività dell'impianto, considerata pari a 8000 ore all'anno. Questi includono il costo di materie prime, feedstock, acqua e solvente di make-up, utilities e i costi relativi alla sostituzione dei catalizzatori. I costi O&M sono dettagliati in appendice F. In tabella 2 sono riportate le maggiori assunzioni effettuate per la valutazione economica.

Il valore del costo totale di investimento (TCI) è stato valutato tramite le assunzioni riportate in tabella 3. Una volta calcolato il TCI e i costi relativi per ogni anno è stato valutato il costo di produzione e quindi il minimo costo di vendita del prodotto (MSP).

TABLE 2: Assunzioni per la valutazione economica.

Economic assumptions	
Return on investment	10%
Project life time	20 years
Capacity factor	91%
O&M escalation rate	3% except for electricity (0,34%)
PEC/TCI ratio	0.3
- Case 1	29%
- Case 2	30%
Contingency	15% of Total direct cost, Engineering and supervision and Construction costs
Equity	100%
Average labor cost	36000 euros per year
Number of employees	30
Overhead & insurance	1,5% TCI
Straight line depreciation	
No salvage value	
No fuel cost	
All construction costs are escalated to 1 <sup>st</sup> year operation	

TABLE 3: Calcolo del TCI [72].

<b>Fixed Capital Investment</b>	
<b>Onsite costs</b>	
Purchased equipment cost	PEC
Piping	35 %PEC
Instrumentation and controls	12%PEC
Electrical equipment and material	13%PEC
<b>Offsite costs</b>	
Land	6% PEC
Civil, structural and architectural work	21%PEC
Service facilities	35%PEC
<b>Total direct costs (DC)</b>	Onsite + Offsite
<b>Indirect costs</b>	
Engineering and supervision	8%DC
Construction costs and contractors profit	15%DC
Contingency	15% of Direct+indirect costs
<b>Other outlays</b>	
Startup costs	$\frac{\text{Cost of fixed O\&M}}{\text{Month}} + \frac{\text{Cost of variable O\&M}}{\text{Month}} + 2\%(\text{FCI-Land})$
Working capital	$\frac{\text{Fuel cost}}{6} + \frac{\text{Cost of variable O\&M}}{\text{Month}} + \text{Cost of fixed O\&M} \cdot 1.25$
Costs of licensing, research and development	2%DC
<b>Total Capital Investment</b>	<b>TCI</b>

## 0.5 Risultati e osservazioni

Come è possibile notare in figura 4 il 14% circa del carbonio è recuperato nel prodotto quindi si ha una elevata perdita di carbonio. Tuttavia, il 42% circa del carbonio è utilizzato come combustibile (prevalentemente metano) e in questo modo soddisfa la richiesta energetica dell'impianto. Da un'analisi energetica, infatti, risulta che il processo è energeticamente autosufficiente e non richiede un'utility calda esterna. L'off-gas è costituito dal 41% circa del carbonio che attraversa il processo. La corrente è principalmente costituita da  $CO_2$  con un minor contenuto di acqua e  $H_2S$ . Il recupero di  $CO_2$ , che si è cercato di massimizzare, è limitato dalla presenza di  $H_2S$ , infatti le solubilità dei due in acqua sono diverse ma confrontabili. Di conseguenza maggiore è il recupero di  $CO_2$ , maggiore è la quantità di  $H_2S$  che finisce nella corrente alimentata al reattore di RWGS. Le analisi di sensitività condotte nel design dello step di purificazione del gas hanno permesso una notevole riduzione delle perdite di carbonio nell'acqua residua e nel solvente MDEA, che infatti escono dal processo con un'elevata purezza. Per questo motivo per la valutazione economica è stato considerato la possibilità di ricircolare sia il solvente che l'acqua con una sostituzione periodica di solventi freschi per la presenza di impurità.

Per quanto riguarda la distribuzione del prodotto, quello che si ottiene in maggior quantità è il cherosene (circa 40% della corrente liquida ottenuta con il processo di FT), seguito poi da gasolio (~28%) e diesel (~22%). Questa distribuzione è influenzata dalle condizioni di lavoro del reattore di FT e in modo particolare dal parametro  $\alpha$  che caratterizza la crescita della catena idrocarburica.

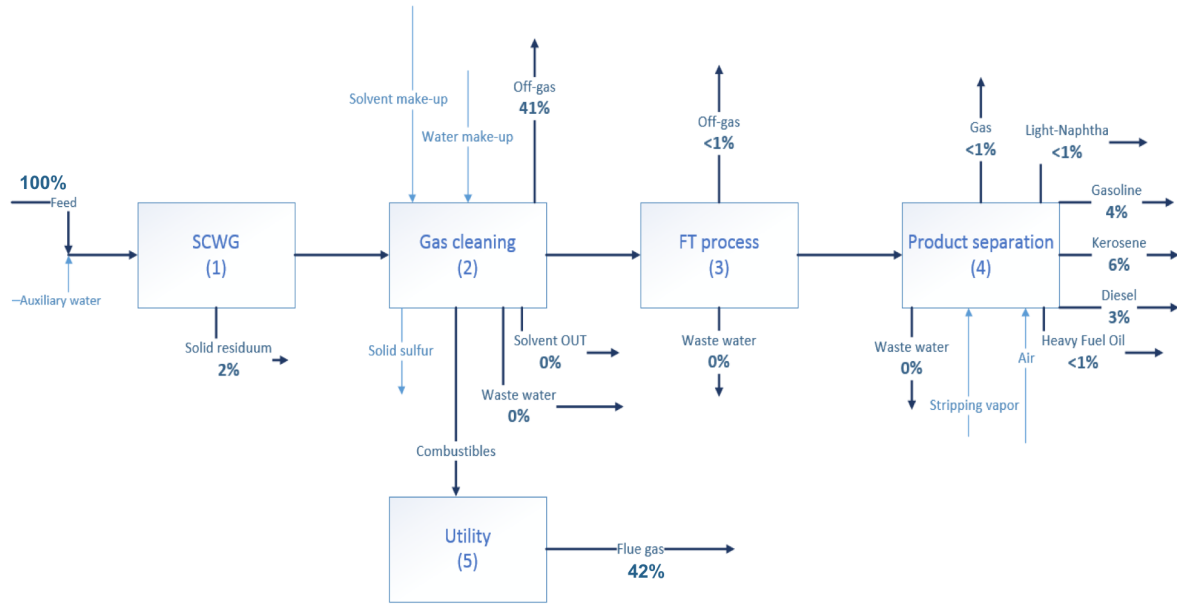


FIGURE 4: Partizione del carbonio nel processo.

Le curve TBP e ASTM D86 sono solitamente usate per caratterizzare i vari tagli ottenuti dal greggio. Per questo motivo è stato fatto un confronto tra le TBP dei tagli ottenuti da questo processo e quelle del greggio da petrolio trovate in letteratura [70]. Quello che si osserva da questi grafici è la percentuale di distillato che può essere estratto in funzione della temperatura. Tagli con idrocarburi pesanti hanno curve TBP più elevate, e quindi è necessaria una maggiore temperatura per estrarre la stessa quantità di distillato. La differenza che si nota tra le curve del processo e le curve tratte dalla letteratura è dovuta alla differente composizione tra il greggio da petrolio e quello relativo al processo in esame. Le curve da dati di letteratura sono più elevate, e questo è principalmente causato dalla elevata presenza di impurezze nelle correnti, per esempio zolfo e ossigeno, che condizionano la temperatura di ebollizione (si fa riferimento alle figure 4.4, 4.5, 4.6 presenti nel testo).

Per quanto riguarda le valutazioni di tipo energetico è stata fatta una *pinch analysis*, come già accennato precedentemente. La curva gran composta permette di valutare il massimo recupero di calore possibile e allo stesso tempo il minimo valore di richiesta di utility esterna. Sulla base di queste informazioni, è stato utilizzato Aspen Energy Analyzer per valutare la rete ottimale di scambiatori. Si ottengono due configurazioni che non necessitano un'utility esterna calda, grazie all'elevata quantità di combustibili (principalmente metano) ricavata dalla PSA e dal water scrubbing. Inoltre, per soddisfare la richiesta di utility fredda è sufficiente la cooling water e non risulta necessario l'uso di refrigerante. Con queste configurazioni si ha anche generazione di vapore ad alta pressione (HP steam) grazie alle elevate temperature del syngas, che permette di ridurre il rifornimento esterno di potenza e quindi i MW consumati dall'impianto. A seguito della determinazione della rete di scambiatori ottimale, è stata valutata l'efficienza termica del processo sia globale che in termini di *biomass to liquid* tramite le equazioni 8 e 9.

$$\eta_{BtL} = \frac{Q_{HHV,product}m_{product}}{Q_{HHV,biomass}m_{biomass}} = 17\% \quad (8)$$

$$\eta_{thermal} = \frac{q_{LHV,products} m_{product}}{q_{LHV,biomass} m_{biomass} + P} = 20.7\% \quad (9)$$

I valori ottenuti sono decisamente inferiori a quelli riportati in letteratura per altri processi di conversione della biomassa in combustibile liquido. Ad esempio, nei processi di HTL (hydrothermal liquefaction) l'efficienza globale termica del processo è circa 30% [81]. Nei processi di gassificazione in acqua supercritica, inoltre, le efficienze sono maggiori e comprese tra 40 e 50% (come riportato nella tabella 2.5 presente nel testo). Questi valori sono tuttavia riferiti alla sola gassificazione della biomassa, senza considerare ulteriore trattamento del gas per la sintesi di combustibili liquidi. Altri studi sulla conversione della biomassa a combustibile liquido tramite gassificazione convenzionale e Fischer-Tropsch hanno fornito efficienze di circa 38% [47]. Il valore, decisamente più elevato di quello riportato dall'equazione 9 può essere dovuto al fatto che la gassificazione in acqua supercritica, rispetto a quella convenzionale, avviene in condizioni di lavoro che richiedono un carico energetico decisamente superiore.

La valutazione economica ha permesso di determinare il prezzo minimo di vendita del prodotto, che risulta pari a 360 €/MWh (nel caso migliore dei due studiati). Questo valore risulta superiore a quello relativo ad altri processi di conversione della biomassa in combustibile liquido. Ad esempio, nei processi di HTL il minimo prezzo di vendita è di circa 190 €/MWh [44] e nei processi con FT dopo gassificazione convenzionale è intorno al 75 €/MWh [45]. Questa differenza è dovuta soprattutto dall'elevato costo relativo alle apparecchiature che lavorano in condizioni supercritiche. Inoltre, il processo per come è stato progettato presenta un elevato numero di scambiatori di calore che condizionano il costo totale di investimento. Il costo capitale, infatti, dovuto a acquisto e installazione delle apparecchiature, è senza dubbio quello che influisce maggiormente sul costo dell'impianto e di conseguenza sul prezzo minimo di vendita del prodotto. Ciò è visibile in figura 5.

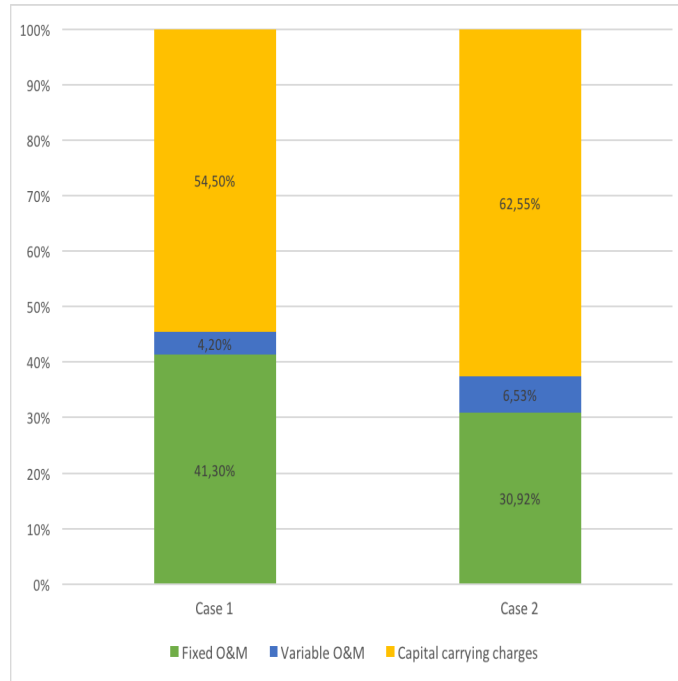


FIGURE 5: Contributi percentuali del costo capitale e dei costi operativi al prezzo minimo di vendita.

Il valore attuale netto dell'impianto è stato valutato sulla base dei prezzi relativi al 2016. In entrambe le configurazioni studiate gli introiti sono minori rispetto alle spese totali. Un'analisi di sensitività è stata condotta per capire la quantità di introiti necessari a pareggiare le uscite, e come risultato si è ottenuto che i primi dovrebbero essere circa sette volte maggiori rispetto alla situazione attuale. Solamente un'ottimizzazione del processo potrebbe portare alla competitività del prodotto sul mercato.

## 0.6 Conclusioni

La tesi propone lo sviluppo di un processo innovativo per la produzione di combustibile liquido dalle vinacce con l'obiettivo di trovare un'alternativa sostenibile ai processi attuali. Il progetto vuole analizzare la possibilità di utilizzare le vinacce ottenute dalla canna da zucchero come feedstock per la produzione di combustibile liquido, e di capirne lo sviluppo ottimale. Il processo è stato simulato su Aspen Plus® sulla base della quantità di vinacce prodotte da un impianto di produzione di bio-etanolo dalla canna da zucchero. Le vinacce, infatti, sono un sottoprodotto ottenuto dal trattamento della melassa, anch'essa un sottoprodotto della produzione di zucchero, tramite fermentazione e successiva distillazione. Per come è stato sviluppato il processo, le vinacce hanno un buon potenziale per la produzione di idrogeno attraverso la gassificazione in acqua supercritica. Tuttavia, la conversione del gas in liquido tramite Fischer-Tropsch è limitata dal basso recupero di carbonio nel prodotto (circa 14%). Questo limite è dovuto soprattutto all'elevato contenuto di zolfo nel feedstock che limita il recupero della  $CO_2$ . Per ottimizzare il processo si potrebbero valutare alternative per la rimozione dell' $H_2S$  in modo da permettere un maggior recupero della  $CO_2$ .

Il processo risulta autosufficiente in termini energetici, non è infatti necessaria alcuna utility esterna calda. Questa condizione è possibile grazie alla grande quantità di combustibili che vengono recuperati nello step di purificazione del gas. Infine, i costi relativi al processo lo rendono al momento non economicamente sostenibile, analogamente ad altri impianti di conversione della biomassa in combustibile liquido. In conclusione, il processo risulta fattibile e in futuro, grazie ad ulteriori ricerche ed ottimizzazioni, potrà diventare economicamente sostenibile.



## Chapter 1

# Introduction

In recent years, the growth of energy demand has focused the attention on sustainable alternatives, due to the limited availability of fossil fuels. This led to increasing the research on the field. As far as transportation liquid fuels are concerned, several biomass treatments have been studied, especially the hydrothermal ones due to the high moisture content of biomass.

This thesis has the aim of developing an innovative process for the vinasse treatment for the production of transportation fuels through supercritical water gasification. Furthermore, the project wants to evaluate the economic feasibility and profitability of the process compared to other bio-fuels processes. Vinasse is, as a matter of fact, a by-product obtained from the processing of sugarcane.

In the first chapter of the thesis is given an overview on the actual energy situation and the trends and objectives set by UE for future years. At first, the characteristics of the vinasse and its production paths are described. Due to the high moisture content, vinasse is a good feedstock for hydrothermal treatments. In this chapter, is thus given an overview on the main hydrothermal treatments that can be used to processing vinasse. At last, in the first chapter are described the objectives of the thesis and the methods.

The second chapter is a literature review on the vinasse composition and the current technologies that are used for its treatment. It also gives an overview on the characteristics of water when upgraded into supercritical conditions and the current process designs for supercritical water gasification and liquid fuel production. Finally, the literature review chapter gives an information of the technical analysis in literature, and the current economic status of hydrothermal plants. There is also a brief description of the Åbo Akademi experimental set up whose datas were used as starting point for this work.

In the third chapter are described the choices made during the process design and how those were simulated in the flowsheet. The process is divided in main steps: supercritical water gasification, gas cleaning, Fischer-Tropsch process, product separation and heat recovery. For each step are described and illustrated the main designing decisions. Moreover, sections 3.3 and 3.4 describe how the economic evaluation and the thermal analysis were made.

The last chapter shows the results obtained and it makes a comparison with other processes for the conversion of biomass into transportation liquid fuel. It also gives suggestions for process optimization and future researches.

## 1.1 Research background

### 1.1.1 Energy outlook

Nowadays, fossil fuels are the world's primary energy source. Although they have fueled the global economic development over the past century, fossil fuels are a limited source and they also can irreparably damage the environment. Fortunately, sustainable solutions, such as renewable energy, exist. In recent years, especially from 2010, renewable sources consumption has risen, although the largest growth was shown by natural gas, as shown in Figure 1.1.

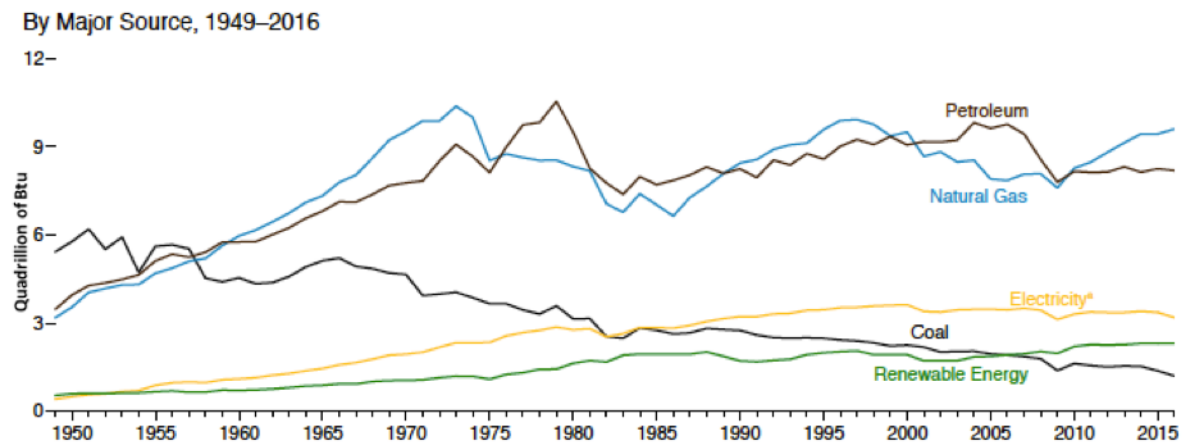


FIGURE 1.1: Energy consumed in the industrial sector [1].

If we focus on the EU members, we can say that the energy produced from renewable sources is higher in the Nordic countries and the highest trends are shown from Sweden and Finland. Concerning other countries, the targets expected for 2020 are still far to reach as shown in Figure 1.2.

In the Monthly energy review of July 2017 (U.S. Energy information Administration *Monthly Energy Review* [1]) it is reported the currently quadrillion of btu consumed in March 2017 for each source (Figure 1.3). It is possible to notice that the renewable energy consumption is still largely lower than the petroleum one [1].

The greenhouse gases (GHG) emitted by the treatment of fossil fuel contribute to climate change. With the use of renewable sources it is possible not only to have a sustainable feedstock, but the emission of pollutants can be largely reduced.

For these reasons, EU countries have agreed new targets and objectives with the aim of achieving a energy system with lower emissions. Thus, the European commission set new targets for 2030: a decrease of greenhouse gas emission of 40% compared to the levels in 1990 and also a share of renewable energy consumption at least 27% [3].

The term *biomass* is used to indicate any organic matter, derived from plants or animals, that can be converted into biofuels after treatment, and is therefore regarded as an energy source.

Biomass chemical compositions varies in terms of moisture, ash and organic content based on the feedstock source [4]. It is mostly composed of carbon, hydrogen and oxygen with a minor percentage of other elements such as nitrogen, phosphor, potassium and sulfur, which vary with the origin of the biomass (ultimate analysis). From a chemical point of view cellulose, hemicellulose and lignin are found to be the major components of biomass [5].



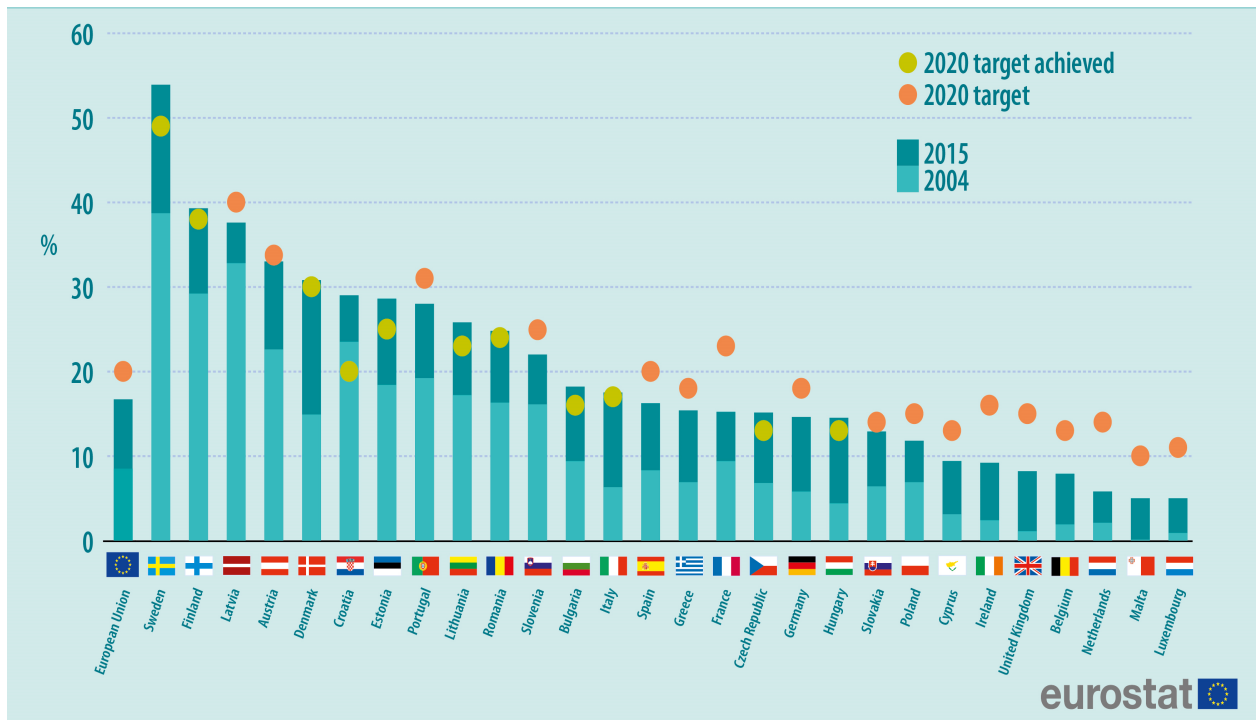


FIGURE 1.2: Share of energy from renewable sources in the EU member states [2].

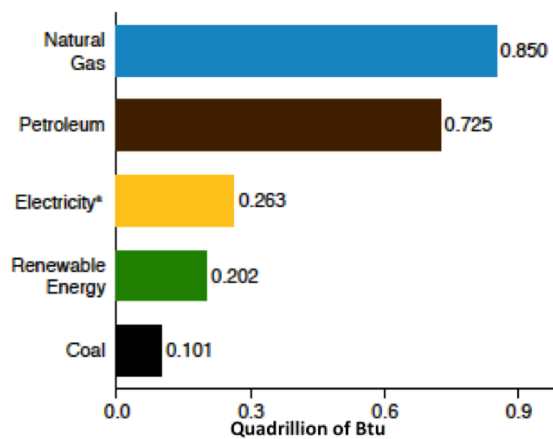


FIGURE 1.3: Energy consumed in the industrial sector for each type of source [1].

Cellulose is a polysaccharide with the molecular formula  $(C_6H_{10}O_5)_n$  containing a great number of glucose units linked together and it is the major constituent of the cell walls. Hemicellulose is a matrix of polysaccharides, such as arabinoxylans, that exist with cellulose in the plant cell walls. Combined with cellulose, it provides structural strength to the cell wall. Lignin is a complex polymer of phenolic compounds, insoluble in water and alcohol. Its composition varies from species to species.

Depending on the composition of these compounds and on the extractives, which happens to be one of the minor compound, the elemental analysis varies.

If properly treated, the biomass can be converted into various biofuels (e.g. syngas, bio-oil, bioethanol). Table 1.1 shows current technologies used to obtain energy from biomass and their

typical conditions and characteristics.

TABLE 1.1: Current technologies for biomass treating [6].

Process	Type of biomass	Conditions	Features
Combustion	Dry biomass	150-800°C Atmospheric pressure With air	Low energy efficiency (10-30%) Pollutant emissions
Pyrolysis	Dry biomass	350-550°C 10-50 bar Absence of air	Produced bio-oil and gases
Liquefaction	Wet biomass	250-450°C 50-200 bar Absence of air	Transforming biomass to liquid fuel biocrude Without a necessary drying process
Conventional gasification	Dry biomass	800-900°C Atmospheric pressure Limited oxygen	Partial oxidation reaction A gaseous mixture, mainly $H_2$ , $CO$ and $CO_2$
Anaerobic digestion	Wet biomass	Bacteria, in absence of oxygen, degrade the organic matter	Very slow process A relatively low efficiency It produces methane only, no hydrogen
SCWG	Wet biomass	300-650°C >221 bar Limited oxygen (oxygen coefficient 0-0.25)	Higher gas yields Hydrogen more than 50% on a mole basis, $CO_2$ about 33%, and others including $CH_4$ and low $CO$ Low level of residual chars and tars

Traditional gasification technologies have encountered several difficulties, for instance a low quality gas due to higher presence of impurities such as char and tar [6], and the low efficiency at high moisture content. It was demonstrated that through anaerobic digestion or hydrothermal gasification, efficiency does not decrease with moisture content [6].

### 1.1.2 Vinasse production: the bioethanol plant

Sugarcane production is mostly based on the world demand for sugar, in fact almost the 80% of sugar globally produced comes from sugarcane [7]. The sugar extraction and refining is a process that leads to the recovery of molasses as a byproduct. It is a circular path of washing and heating the cane (and beets also) with hot water. It has a great potential for the production of ethanol through fermentation and distillation. The residue obtained by the fermentation is called vinasse, and it is a feedstock that has potential for fuel production [5].

As illustrated in Figure 1.4, vinasse can be produced both from molasses and from the juice and depending on the production process, its properties and characteristics vary [8].

The processing steps in vinasse production could be summarized as [9]:

- **Washing and cutting**  
The sugar cane stalks are at first subjected to hot water sprays to remove dirt and impurities. Then, they are passed under rotating knife blades to reduce the stalks in short pieces.
- **Extracting the sugar juice**  
Extraction can be realized in two possible ways: diffusion or milling. By the diffusion method, the cut stalks are dissolved in hot water or lime juice. In the milling process, the stalks are passed under several successive heavy rollers, which squeeze the juice out of the cane pulps. Water is sprayed throughout the process to facilitate the dissolving of the juice.
- **Clarifying the juice**  
The extracted juice is clarified by adding milk of lime and carbon dioxide.

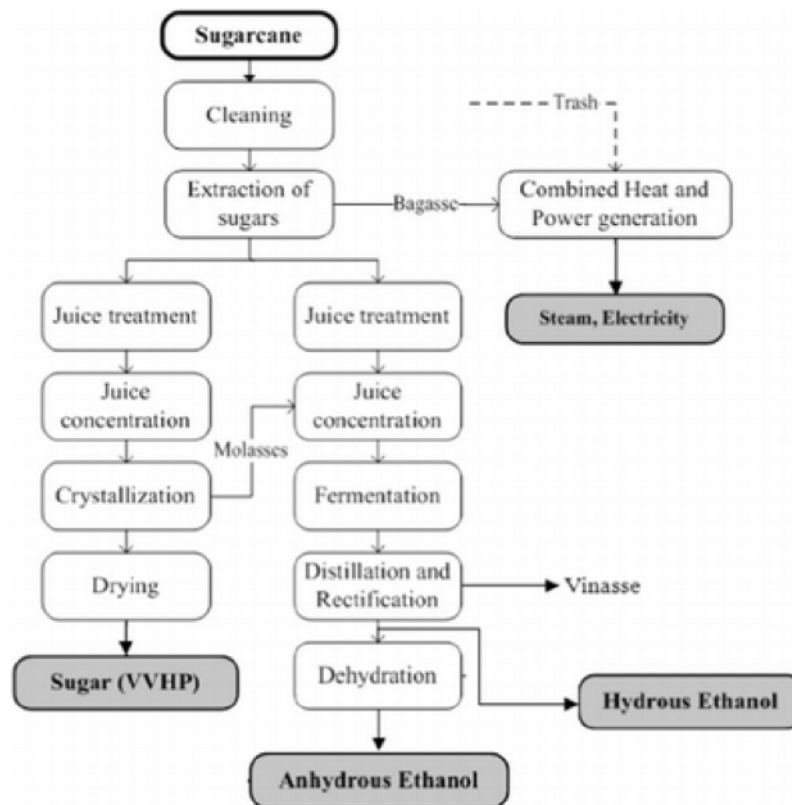


FIGURE 1.4: Vinasse production [8].

- Evaporating and concentrating the syrup

The juice is pumped into an evaporator that allows the evaporation of the water present in the juice. The syrup obtained is concentrated through several stages of vacuum boiling at low temperature. Eventually, the sugar crystallizes out of the syrup, creating a substance that is centrifuged to further separate the raw sugar crystals from the syrup. The remaining syrup is called molasses.

The molasses can be fed to a bio-ethanol plant where it is at first diluted with water and then is fed to the fermentation chamber where the fermentation process lasts from 30 to 70 hours based on temperature, yeast and sugar concentration [8]. After that, the process is fed to a distillation still where solid and slurry (the vinasse) is separated leaving a solution of ethanol and water. This solution is then fed to a distillation column which separates the aldehydes on the top, an ethanol mixture from the middle and a water stream at the bottom [8].

Vinasse from molasses generally presents higher values of COD (chemical oxygen demand) and BOD (biochemical oxygen demand) than vinasse from sugarcane juice. This means that the first one has a higher organic content, therefore his potential as fuel is higher [8].

### 1.1.3 Vinasse characteristics

The word Vinasse comes from the Latin word *vinaceus*, and originally it meant wine yeast. It is a residue obtained from alcohol distillation of fermented molasses and if properly treated it can be a

source of energy, because it can be converted into a valuable biofuel. It has a light black-reddish color when obtained from sugarcane molasses and it appears as a very viscous liquid. It contains between 50 and 100 g/L of suspended solids [10]. Generally, its pH ranges between 3,5 and 5 and it has a high chemical oxygen demand (COD) [11]. Its density has a value of  $1051 \text{ kg/m}^3$  at  $20^\circ\text{C}$  [12]. It is a mixture of water, organic and inorganic compounds, and its composition varies because of the different raw materials and different plants used for the ethanol production.

Vinasse is often used for irrigation due to its high organic content, although it needs to be treated before due to the presence of phytotoxic, antibacterial and recalcitrant compounds such as phenols, polyphenols and heavy metals, which cause negative effects on microorganisms and plants [11]. Vinasse is also used in on-site combustion.

Its high moisture content make vinasse a promising feedstock for hydrothermal treatments. In this study the supercritical water gasification (SCWG) of Vinasse will be evaluated. In spite of its large potential, the current studies made on the SCWG of vinasse do not lead to high energy and economic efficiencies, therefore more studies are needed [13].

#### 1.1.4 An overview on Hydrothermal treatments of biomass

The hydrothermal treatment is defined as “reactions occurring under the conditions of high temperature and high pressure in aqueous solutions in a closed system”[14], therefore a pre-drying step is not needed. Those include supercritical and subcritical treatments depending on the conditions of temperature and pressure.

When the process conditions are below the critical point of water ( $374^\circ\text{C}$  and  $22,1 \text{ MPa}$ ) Hydrothermal Liquefaction is used to produce bio crude. The reaction is carried out at a temperature that ranges between  $280^\circ\text{C}$  and  $370^\circ\text{C}$  and a pressure up to 200 bar. Under this conditions organic compounds have high solubility and thus are miscible. The process is essentially pyrolysis in hot water but it does not require a catalyst and the product obtained is a bio-crude. As well as pyrolysis, the HTL is the thermo-chemical decomposition of organic matter, but the working conditions are different. In the pyrolysis process the feedstock has to be dry (moisture content below 10%), it is treated at moderate temperature (between  $350$  and  $550^\circ\text{C}$ ) and atmospheric pressure; on the contrary, HTL is performed in the presence of water, in fact it can convert biomass with high moisture content, at lower temperature and higher pressure[15]. In the HTL process the feedstock is made of wet biomass and water (which needs to be added as auxiliary water in order to reduce the solid content) and the energy consumption in feedstock pre-treatment is limited compared to technologies in which biomass drying is required (pyrolysis). The low energy consumption is due to the possibility to recover the heat when the process is in liquid form, however in drying, the water is evaporated and the heat is dissipated to the atmosphere.

The product obtained, the bio crude, vary its chemical composition based on the HTL reaction conditions and also on the characteristics of the feedstock.

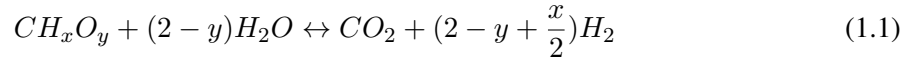
When the process condition are near and beyond the critical point of water ( $374^\circ\text{C}$  and  $22,1 \text{ MPa}$ ), biomass can be gasified with supercritical water (SCWG). As a result, the organic constituents of biomass decompose into synthetic gas.

At supercritical conditions the polarity of the water becomes negligible, and this characteristic allows a reasonably easy extraction of inorganic or polar constituents. Water under these conditions has high reactivity, and this makes it a good medium to improve the reaction rate. Reaction times are faster compared to other processes such as anaerobic digestion and as a result an optimal conversion of biomass can be achieved [16]. At lower temperature (from  $370$  to  $550^\circ\text{C}$ ) the tests can be taken both in catalytic and non-catalytic environments; the primary product are water soluble organics when

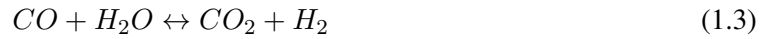
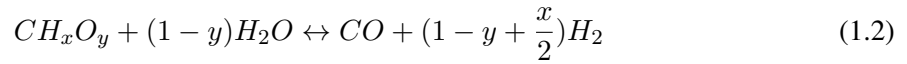
no catalyst is used, while it is observed the release of a carbon rich syngas when a metallic or alkali based syngas is used [16]. At higher temperature (beyond 550°C) the mainly product is a hydrogen rich syngas due to the endothermic reaction of  $H_2$  generation. According to Le-chatelier's principle when a change in temperature occurs, the system will no longer remain at equilibrium: the reaction will change the direction in order to restore the equilibrium. Effect of change in temperature is related to the nature of reaction whether it is an endothermic reaction or an exothermic reaction [17], consequently along with the hydrogen other compounds are formed.

Through SCWG the biomass is converted into multiple gaseous product such as  $CO_2$ ,  $CO$ ,  $H_2$  and  $CH_4$ . If the feedstock contains a large percentage of sulfur, the product contains  $H_2S$ , so it is necessary to include an acid gas removal step [16].

The syngas, can be produced from the material that contain carbon, so from a great variety of feedstock for example biomass, plastics and coal. The SCWG of biomass is a complex process, but the overall conversion for hydrogen production can be simplified as:



Where x and y indicated the  $\frac{H}{C}$  and  $\frac{O}{C}$  ratios in the feedstock [18]. Along with this reaction, some others can occur: steam reforming (1.2), water gas shift (1.3) and methanation (1.4) [18].



The methanation is an exothermic reaction while the  $H_2$  formation is endothermic; this is why it is possible to obtain a hydrogen rich syngas at higher temperature [18]. Differently from the thermal gasification, with the supercritical water gasification the syngas has a high content of  $CO_2$ . This is caused by both the oxygen present in the biomass and the one in the auxiliary water that allows oxidation.

## 1.2 Objectives and aim of the thesis

The thesis has the aim to design an innovative process for the production of valuable transportation liquid fuels from vinasse through supercritical water gasification (SCWG). The work wants to evaluate the possibility of this plant of being a sustainable alternative to the current solutions for the production of liquid fuels. Furthermore, it wants to find a new way of processing the vinasse, which is now only a by-product.

The main goals are:

1. Process modeling of a complete plant layout on Aspen Plus® for SCWG of vinasse and further syngas treatment for liquid fuel production
2. Evaluation of the economic feasibility and profitability of the process compared to other bio-fuels plants.

### 1.3 Methods

The complete process layout has been simulated through the commercial software Aspen Plus® V8.8 [19]. The first calculations of solvents and reactants mass flows were evaluated on Microsoft Office Excel and then implemented on Aspen Plus® for the model confirmation. The methods used for the Aspen Plus® simulation are the ones that better predict the evolution of the system, and were selected after a literature research.

The calculations and the choices made while building the model are illustrated in detail in the Chapter 3. Some of the physical parameters cannot be modeled on the software used, so they will be assumed based on the literature data, for instance the reactors working conditions when a catalyst is considered. The scenario was scrutinized with the pinch analysis method to define the auxiliary utility demand and evaluate the overall thermal performance (see Section 3.3). Finally, the techno-economic assessment (TEA) made establishes the commercial potential of the plant by determining the minimum selling price (MSP) and the net present value (NPV) relative to current commercial and market conditions (see Section 3.4).

## Chapter 2

# Literature review

### 2.1 Vinasse composition

Vinasse has a high moisture content which ranges between 40% and 50%. The solids include ashes, proteins and carbohydrate. The major compounds found in the vinasse are glycerol, acetic acid and lactic acid [20]. Akram et al. [21] analyzed sugarcane vinasse behavior during combustion in a fluidized bed combustor. For the study they used a vinasse with a moisture content higher than 50% (Table 2.1) and a low calorific value (CV) thus they showed the impossibility of combustion without the support of a fuel having a high CV (coal was utilized in the study).

TABLE 2.1: Proximate and ultimate analysis for the vinasse used by Akram et al. [21].

	Coal	Vinasse
<i>Proximate analysis</i>		
Moisture (%)	5.6	58
Ash (%)	4.9	9.7
Volatiles (%)	34.2	32.3
Fixed carbon (%)	55.3	-
<i>Ultimate analysis</i>		
Carbon (%)	74.4	17.3
Hydrogen (%)	4.06	2.11
Oxygen (%)	6.97	8.73
Nitrogen (%)	1.62	2.1
Sulphur (%)	2.01	2.1

One of the possible treatments for vinasse is gasification. It has been widely investigated because of its potential of providing higher efficiency cycles. Wide ranges of biomasses and waste products have been studied as potential fuels [22]. Dirbeba et al. studied the gasification of sugarcane vinasse in  $CO_2$  and the release of ash matters in  $CO_2$  and  $N_2$  atmospheres. As a result, they obtained an ash content on a dry fuel basis of 34.1 %wt on to 67.4%wt dry solids diluted samples of dried vinasse [22]. As said in the previous chapter, one of the uses of vinasse is its combustion on-site. Cortez et al. [23] investigated and developed the basic technology of on-site disposal of vinasse by combustion and they evaluated its heating value, composition, and flame characteristics through combustion tests. They tested the characteristics of the vinasse obtained from the Shepherd Oil Distillery, which has the

ultimate analysis reported in Table 2.2. In conclusion, even at lowered moisture content the vinasse combustion remains challenging and not a favorable process due to its low calorific value.

TABLE 2.2: Vinasse composition from Shepherd Oil Distillery on dry basis [23].

C	%	39.72
H	%	8.60
O	%	n.a.
S	%	0.12
N	%	1.65
Ash	%	18.95

## 2.2 Properties of water: comparison between subcritical and supercritical conditions

A fluid is considered to be in supercritical conditions when the temperature and the pressure of the system are above its critical point. The critical point of water is found at 374°C and 22.1 MPa (shown in Figure 2.1 where the water phase diagram is represented). Below that the fluid is considered to be in subcritical conditions. When a fluid is upgraded to supercritical conditions, its properties show important changes and the physical and transport properties are between those corresponding to the gas and the liquid state. The main advantage of reaching supercritical conditions is that the fluid has an optimal heat transfer and also an ideal mass transfer. [24].

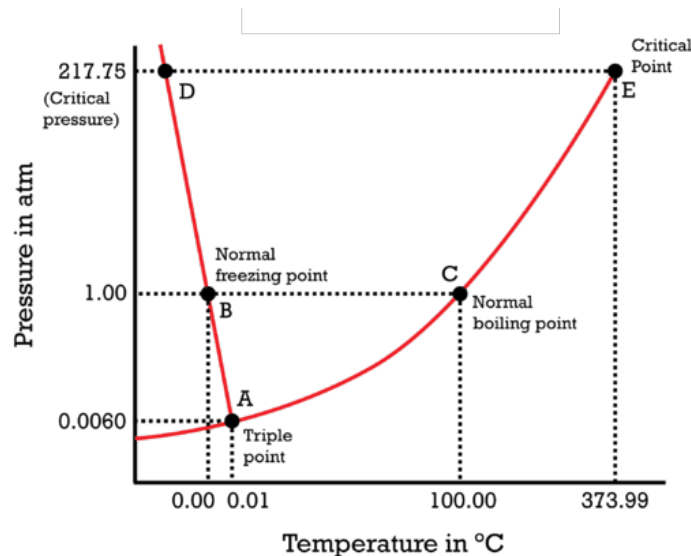


FIGURE 2.1: Water phase diagram [25].

One of the physical properties that can be evaluated is the density which in supercritical conditions is greater than the water vapor. During the transition, it decreases in a similar way as it happens during transition of phase in subcritical conditions, with a smoother reduction as pressure increase



above critical point, as shown in Figure 2.2. The SCW viscosity coefficient is similar and its diffusion coefficient lies between that of liquid and gas phases [24].

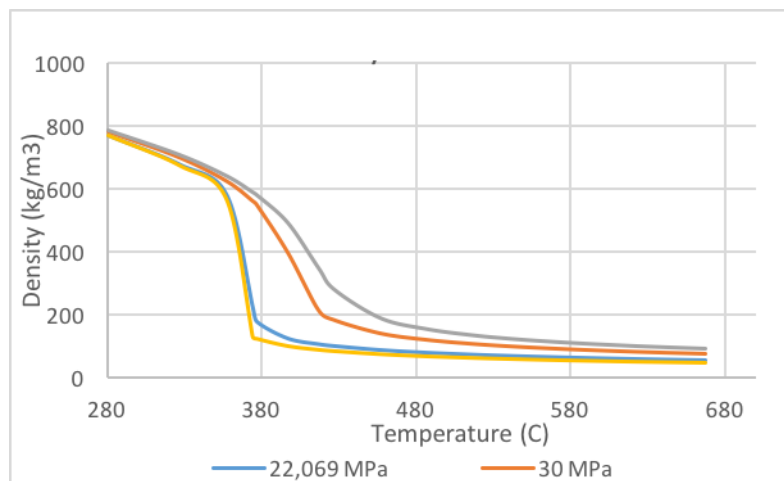


FIGURE 2.2: Variation of water density with temperature (diagram realized on excel based on the value of water density taken from the NIST database [26]).

When changing the physical conditions, water has different and often anomalous properties, which is due to changes in the hydrogen bonding [27]. As an example, when increasing the temperature, the dielectric constant ( $\epsilon$ ) of water decreases from 80 at standard conditions, to 6 at critical conditions as a result of a decreased effectiveness of hydrogen bonds. Above the supercritical point, the hydrogen bonds break down changing the water properties, which becomes less polar and it behaves more like an organic solvent [27]. The most interesting property of SCW is its low polarity; its high miscibility with gases as well as many organic compounds makes it a great solvent for reactions with organic slurries. The dielectric constant gives a measure of water polarity, as illustrated in figure 2.3. [27]

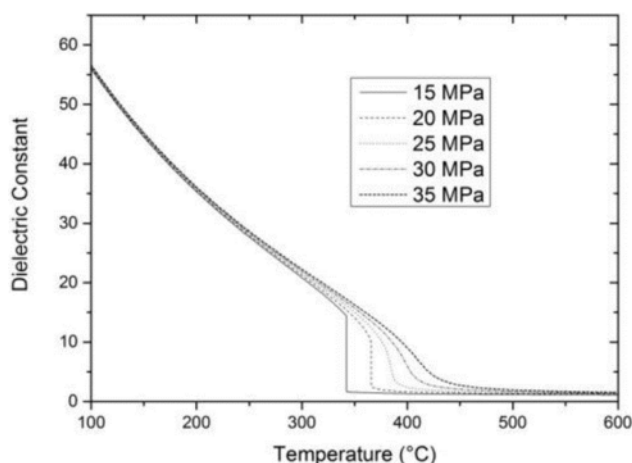


FIGURE 2.3: Variation of dielectric constant with temperature [27].

Above 300°C, water starts to behave as a near-critical liquid so the physical properties start to change significantly with the pressure. When the water is heated above 100°C the ionic products

increase due to a decrease of its dielectric constant, this indicates that water can act as an acid or a base catalyst [27]. As the temperature raises, the dissociation of water increases into  $H_3O^+$  and  $OH^-$  ions. In figure 2.4 is shown how the ionic product changes with temperature. The process is reversible and sufficiently rapid to be considered as an equilibrium reaction. The ionic product of water is significantly increased in the subcritical and supercritical state of water, and due to this property the supercritical and the subcritical water are highly reactive. This allows high reaction rates and leads to a better conversion and a shorter residence time with economic advantages and lower environmental mark. The amount of ionic products in subcritical conditions increase with the temperature because of the water dissociation, until they reach a maximum before the critical point. After that, the ionic products start to decrease because the decrease of the density of water and also as a result of the solvent change of polarity. With the reduction of ionic products, free radical gas phase reactions are promoted [27].

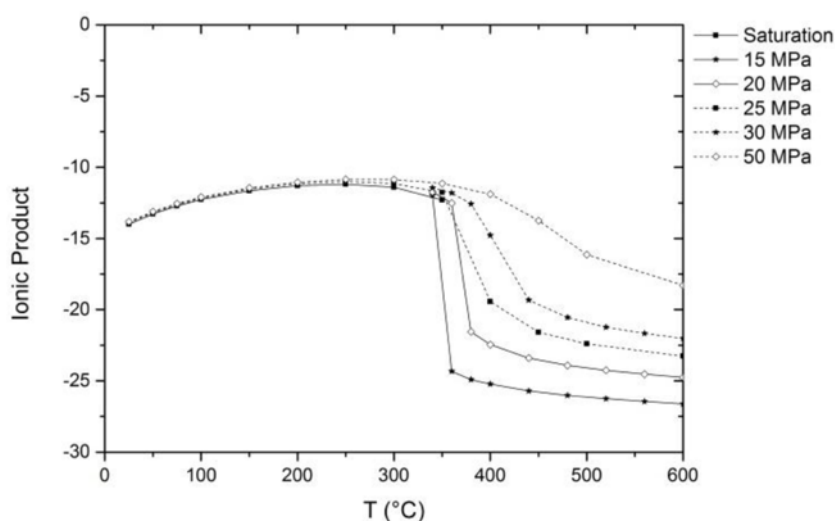


FIGURE 2.4: Variation of the ionic product  $[log_{10}K_w/mol^2kg^{-2}]$  with temperature [27].

At a fixed pressure, beyond the critical point, the solubility of organic compounds begins to increase quickly around 200°C and continues raising as it reaches the critical temperature. On the contrary, the solubility of ionic compounds, such as salts, decrease significantly in supercritical conditions, as it is possible to see in Figure 2.5.

To sum up, Table 2.3 shows the characteristics of water at different conditions.

TABLE 2.3: Properties of water in various conditions [28].

	Normal water	Subcritical water	Supercritical water
Temperature (°C)	25	250 - 350	400
Pressure (MPa)	0.1	5 -25	25-50
Density ( $g\ cm^{-3}$ )	1	0.8-0.6	0.17-0.58
Dielectric constant ( $F\ m^{-1}$ )	78.5	27.1-14.07	5.9-10.5
Ionic product	14	11.2 -12	19.4-11.9
Dynamic viscosity	0.89	0.11- 0.064	0.03-0.07

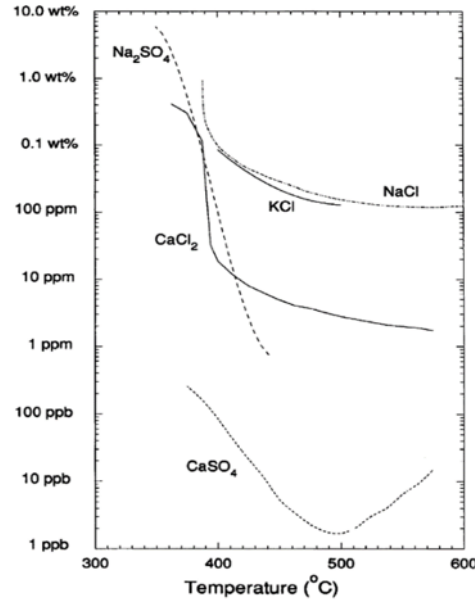


FIGURE 2.5: Solubility of various salts at 25 MPa ([27]).

## 2.3 Process designs

### 2.3.1 SCWG

The SCWG has been widely studied both for coal and for the biomass [16], [29], [24], [30]. Table 2.4 shows an overview of the studies based on SCWG of biomass.

Before modeling the process it is necessary to define the objective, so whether the desired product is a hydrogen rich syngas or a methane rich syngas. This choice allows to determine the optimal working conditions as the two reactions are one exothermic and one endothermic. The biomass feedstock is at first mixed with the solvent, water, which can be fresh or recycled in order to obtain a slurry with a solid content from 10% to 20% depending on the type of biomass. If the solid content is too high, the production of syngas is limited. Furthermore this leads to the possibility of obstructing the pipelines leading to an excessive pressure increase in the system when the solid content is too high.

After mixing, the slurry is pumped to the working pressure (above 221 bar) and it is pre heated. For the pre heating design it is necessary to understand the amount of heat needed and also the amount of heat that is released from the other sections of the plant in order to minimize the hot utility required and so the costs of the plant. This will be analyzed in Chapter 3.3.

TABLE 2.4: Overview of SCWG working conditions found in literature.

Feedstock	Reactor working conditions	Solid Content	Reference
Algae feedstock	600°C	5%	[16]
49 biomass	600, 700, 800 C	5, 15, 25%	[29]
Black liquor	600°C	5%	[24]
Sewage sludge	600°C	20%	[30]

Many useful products can be obtained from the SCWG of biomass: they include syngas, heat and power. These can be further treated to obtain bio-fuels and bio-char. Syngas can also be further processed into methanol, dimethyl ether and other chemical feedstocks. An evaluation of the thermal performance of the plant identifies the feasibility and the limitations of the study. Table 2.5 highlights the energetic efficiencies associated to different configurations for SCWG of biomass. Magdeldin et al. [16] studied the SCWG of an algae feedstock with different configurations and different products. What emerges from their study is that the energy efficiency related to the plant is higher in the BioSNG production cases. The thermal efficiency observed for the production of hydrogen is similar to the work of Lu et al. [31] whose study focuses on the thermodynamic analysis of biomass gasification for hydrogen production in supercritical water. Aziz et al. [32] proposed system for SCWG of a spirulina algae feedstock in a catalyzed fluidized bed for power production step in an integrated gasification combined cycle excess power integration configuration. Overall, the thermal efficiency of the plant ranges around the 40% because of the different systems of fluidized beds.

TABLE 2.5: Thermal efficiency associated with SCWG of biomass.

Feedstock	Product	Energy efficiency	Reference
Algae feedstock	Hydrogen	42-55%	[16]
Algae feedstock	BioSNG	63-66%	[16]
Lignocellulosis biomass	Hydrogen	55%	[28]
Biomass	Hydrogen	42%	[31]
Spirulina algae	Power production	40%	[32]
Sewage sludge	Hydrogen	56%	[33]
Corn starch and sawdust	Hydrogen and CO	44%	[34]

With the conventional treatments the efficiency decreases with the moisture content, mainly due to the heat load required for the drying step. On the contrary, this characteristic was not observed with SCWG. This makes SCW a good medium for gasification processes for feedstocks with high moisture content [28].

After the SCWG reactor, the water contained in the syngas needs to be removed along with the acid gas. For this reason, it is necessary a quench step, which allows the water removal, and then an acid gas scrubbing. There are multiple possibilities to quench the gas, for example it can be fed into an heat exchanger, or it can be diluted with cold water[35].

The acid gas removal (AGR) stage is a lot discussed in literature; hydrogen sulfide removal processes can be either physical-chemical or biological [36]. Swanson et al. [35] evaluated the acid gas removal from a syngas through amine scrubbing and subsequent LO-CAT<sup>®</sup> process. The advantages and disadvantages of the different possibilities for acid gas removal are analyzed in Chapter 3, based on the characteristics of the syngas.

As final stage, many solutions are possible. One option is, when obtained a hydrogen-rich syngas, to purify the syngas in order to rich a major content of hydrogen. This can be done through some special filter (Palladium) or with the pressure swing adsorption (PSA), as studied in [16]. This solution could be used for instance for chemical industry and fuel cells, but it is discouraging because it is needed a large scale consumer nearby due to the high transportation costs.

### 2.3.2 Liquid fuels production

Another possible route is the Fisher Tropsch (FT) process that can be carried out in order to obtain a bio-crude from the syngas, which is comparable with the one produced in hydrothermal liquefaction. The Fischer-Tropsch process allows the conversion of the syngas into liquid fuels. Several articles on the FT have been published from 1980, as shown in Figure 2.6. The increase of the studies on the FT process are due to the growth of the oil price, which makes the gas to liquid processes more interesting [37].

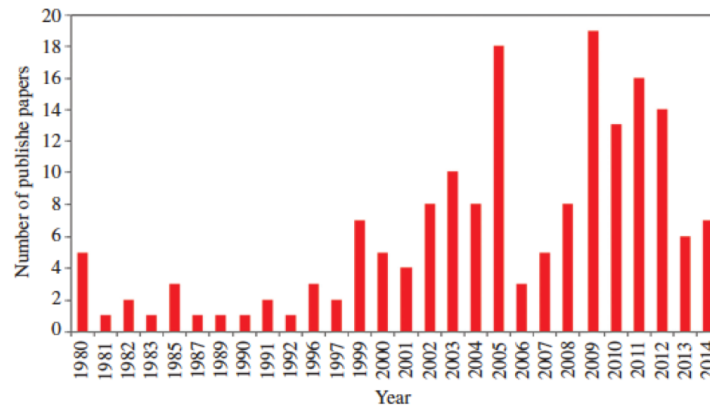


FIGURE 2.6: Number of published papers on the FT process from 1980 to 2014 [37].

The crude obtained from the FT reactor could be the end pipe product of the process, as studied by Pondini et al. [38], but it needs treatments to be used as transportation fuel. Many solutions are possible. For instance, Figure 2.7 shows possible routes for fuel production from Fischer Tropsch processes at different working conditions [37].

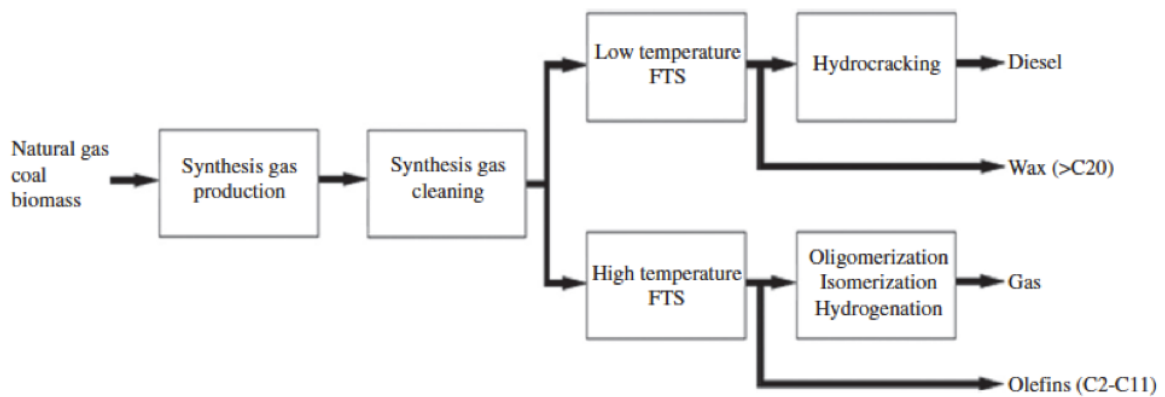


FIGURE 2.7: Processes for fuel production through FT [37].

The commercial FT processes can be carried out at lower or higher temperature. The type of reactor and the catalysts mostly depend on the working condition selected. Dancuart et al. [39] studied the development of the FT industry and their commercial expansion before 2007. In this paper is given a general information (see Table 2.6) on the commercial processes for FT at different temperature

(data referred to Sasol plants, which represent the most significant example of commercialized FT synthesis in the world).

TABLE 2.6: General information about the LTFT and HTFT processes of Sasol [39].

FT processes	Low Temperature (LTFT)	High Temperature (HTFT)
Temperature	220-260°	320-350°
Catalyst	Fe/Co	Fe
Traditional reactor	ARGE	Synthol
Type	Tubular	Circulating Bed
Advanced reactor	Sasol slurry phase distillate (Sasol SPD)	Sasol advanced synthol (SAS)
Type	Slurry (three-phase)	Fluidized bed

To adjust the  $CO-H_2$  ratio entering the Fisher-Tropsch, it is possible to design a reverse water gas shift process block that allows the production of carbon monoxide from  $CO_2$  and  $H_2$ . Saeidi et al. [40] studied the  $CO_2$  hydrogenation to hydrocarbons. The FT process was thus preceded by a reverse water gas shift (RWGS). The RWGS reaction is unclear and extensive studies have been performed to understand the RWGS reaction mechanism along with the characterizations techniques of the catalyst. The effect of catalyst types on RWGS reaction were evaluated along with the selectivity through CO, methane and methanol. Porosoff et al. [41] studied the catalytic reduction of  $CO_2$  evaluating the conditions for synthesis of CO, methanol and hydrocarbons. Each catalyst has different selectivity to the products, and therefore with the appropriate catalyst it is possible to pursue the required route.

As shown in Figure 2.7 the FT crude is not a transportation liquid fuel, consequently it has to be upgraded. Due to his characteristic and composition, the crude can be fractioned in a similar way as for the crude oil. Several studies are currently available on the fractionation of the crude oil. Haydary et al. [42] evaluated the separation of the crude oil into the different fractions with the use of Aspen Plus®. This unit allows the fractionation of the crude into naphtha, kerosene and diesel. The residuum of the fractionation step is the process by-product. As an alternative it can be treated through fluid catalytic cracking (FCC) riser reactor, as studied by Mahfud [43].

## 2.4 Technical analysis in literature

There are several method for a techno-economic evaluation of the plant. Generally, the capital investment and the present value of the plant is determined by finding all equipment costs and operating costs for the construction and operation a plant for 20 years. Total capital investment is based on the total equipment cost with the additional installation costs and indirect costs (such as engineering, construction, and contingency costs) [35]. Generally, the costs related hydrothermal treatments are higher than other major biomass conversion technologies, such as pyrolysis or indirectly-heated gasification.

Magdeldin et al. [44] studied the hydrothermal liquefaction of lignocellulose residues. They obtained a minimum selling price of the crude which varies from 94 to 169 €/MWh depending on the study case. The production cost obtained from a study on the bio-oil integrated gasification and Fischer-Tropsch [45] is ~75 €/MWh. Hence, it is possible to notice that the minimum selling price related to the gasification and subsequent FT process is lower than the hydrothermal liquefaction case.

Biocrude yield is one of the most important factors driving economics of the HTL plant. A maximum yield of 45% is thought to be achievable, therefore the motivation for this research is to

look into the SCWG with 98% carbon recovery into the syngas compared to the liquefaction route with only 45% and even lower.

Hrncic et al. [46] gave an overview of the current high pressure processes for treatment of biomass for production of energy and chemicals. They made an analysis of the economic viability of hydrothermal processes. Table 2.7 shows the economic evaluation of scale up hydrothermal processes reported in ref [46].

TABLE 2.7: Economic evaluation of hydrothermal processes [46].

Process	Capacity (t/h)	Feedstock	Total Capital Investment (TCI) (M€)	Product cost (€/Gj)	Environmental impact
<b>HT carbonization</b>				Hydrochar	
Plant 1	5.7	Empty fruit branches	9.01	9.67	Carbon neutral
Plant 2	13.8	Empty fruit branches	16.47	7.49	Carbon neutral
Plant 3	48.6	Waste food	20.719	5.27	Carbon neutral
<b>HT liquefaction</b>				Biofuel	
Plant 4	0.468	Microalgae	1.6	65.9	Carbon neutral
Plant 5	83.3	Microalgae	468	/	Carbon neutral
<b>HT gasification</b>				Syngas	
Plant 6	5	Sewage sludge	2.915	7.1	Carbon neutral

What emerges from the table is that the production cost related to the HT liquefaction is higher than the other hydrothermal processes. The capacity of the plants shown in table 2.7 varies from 0.468 t/h up to 83.33 t/h depending on the feedstock type and operating conditions. As illustrated in the table, the main advantage of these processes is that they have a low environmental impact due to the fact that they are carbon neutral [46]. Generally in a hydrothermal industrial plant the reactor represents between 8.5 and the 21.3% of the total capital investment (TCI) of the plant depending on plant capacity and feedstock used [46]. The cost contribution of the reactor to the TCI may vary significantly with increasing the plant capacity and with a hypothetical product upgrading. The design of a product upgrading requires a higher number of equipments and therefore a greater TCI. The competitiveness of the process can be assessed in comparison to conventional technologies of hydrogen production based on the techno-economic assessments reported in literature [30]. The working conditions of the equipments effect the investment cost of the plant, hence materials with suitable characteristics are required.

The current technology based on biomass to liquid production through the Fischer Tropsch processes are based on the treatment of a syngas obtained from conventional gasification. Trippe et al. [47] carried out a techno-economic assessment of Fischer Tropsch synthesis as alternative process steps within biomass-to-liquid production. They concluded that the production costs related to biomass fuels are higher than the current price markets (between 76 and 93%). However, this technologies are close to becoming competitive to conventional refineries since the production costs are slightly higher than the current production costs when coal is used as main feedstock (between 7 and 18%) [47]. A similar approach was carried out by Meerman et al. [48]; they evaluated the economic potential of integrated gasification co-generation facilities with carbon dioxide capture and storage. As a result they obtained that the current production costs of transportation fuels are above market prices, nevertheless, future improvements and plant upgradings could make gasification facilities economical. The main process blocks for the biomass to liquid conversion of their review are shown in Figure 2.8.

As the reader can see, the conventional gasification happens through addition of auxiliary  $O_2$  to the biomass feedstock differently from SCWG where the auxiliary feedstock is pure water. In both cases, however, the syngas obtained contains impurities, and therefore it needs to be purified (Figure 2.8).

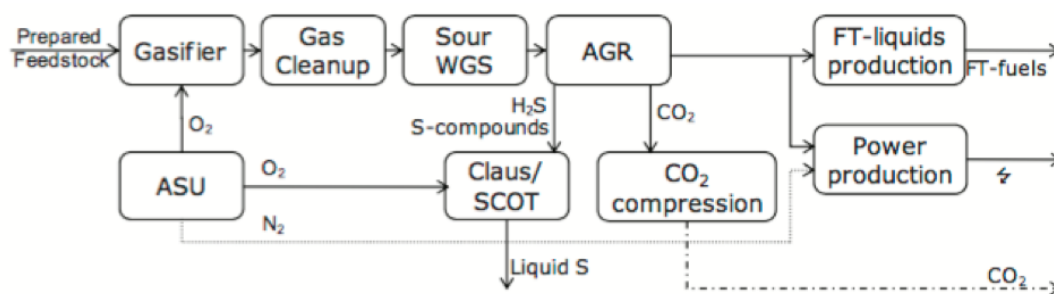


FIGURE 2.8: Biomass to liquid production through FT in ref [48].

As far as vinasse is concerned, it was not possible to find in literature values of the minimum selling price of the bio crude produced from it.

## 2.5 Experimental set-up in Åbo Akademi

Currently, experiments are being carried out in the Åbo Akademi which include the supercritical water gasification of model compounds for vinasse. In particular glycerol, which is used in small quantities to study the production of syngas. The experimental set up is shown in Figure 2.9.



FIGURE 2.9: SCWG reactor in Åbo Akademi.

The core of the experimental set-up is a tubular reactor which works as a semi-batch. This type of reactor was chosen to simulate an industrial plant, where feed is sent continuously. In the laboratory scale, reactor dimensions are limited in order to reduce costs and increase safety, because of use of high temperatures and pressures. However its dimensions are enough to understand SCWG potentiality. Nitrogen is fed first to pressurize the reactor and the temperature is reached and maintained by a tubular oven, made of insulation material and heated by electrical resistors. There are then three thermocouples to control the temperature inside the reactor. Feedstock does not actually pass directly



into the pump in order to avoid corrosion problems, but it is stored into a cylinder put between pump outlet and reactor. At the end products are cooled down and thanks to a apposite valve the products are collected to be analyzed. The experiments were conducted with two different types of reactors: one in Stainless steel and one in Inconel.

The solid content used is around 10% in order to avoid the obstruction of the piping system leading to an excess increase of the pressure. Some of the data obtained from the experimental plant in Åbo Akademi have been used as input for modeling the system. Further information are given in Chapter 3.



## Chapter 3

# Modeling

### 3.1 Introduction

As explained in the previous chapters, the need of alternative sustainable solutions is rising due to the limited source of fossil fuels. This thesis has thus the aim of designing an innovative process for the production of liquid fuels from a by-product: the vinasse. As a matter of fact vinasse is a potential source for fuel production due to its characteristics, such as high organic and moisture content. Currently the researches on vinasse are mainly focused on its gasification, both conventional and supercritical, or on its combustion. Nevertheless, the further processing of the syngas for liquid fuel production is not analyzed, and this is what makes this plant innovative. In this chapter will be described the decisions made and the problems faced while designing the plant.

The process was designed and simulated on the commercial software Aspen Plus® V8.8 [19]. The equation of state selected for this study is the Peng Robinson Boston and Mathias (PR-BM), due to its ability to describe gas processing. However, for the units containing aqueous solutions the ENRTL-RK was chosen due to its ability to predict the behavior of aqueous electrolytes [49]. For the fractionation and product separation step the BK10 method was used because it is strictly applicable to heavy hydrocarbon systems [19].

### 3.2 Process design

The flowsheet of the plant is reported in Appendix B. The amount of vinasse that arrives to the plant was chosen by literature data as 230 ton per day. This value was obtained considering that the amount of vinasse generated by the bioethanol production plant is around 23 kg/l of ethanol. Moreover it was considered that on an industrial scale the capacity of the ethanol plant is around 10  $m^3$  of ethanol per day [50].

To model the behavior of the vinasse, three model compounds were selected: lactic acid, acetic acid and glycerol [12] and their concentrations were calculated as average of the values found in literature [21], [22], [23], [12] for vinasse from sugarcane.

As lactic acid, acetic acid and glycerol are found to be the major compounds but not the only ones, a non conventional compound (CHOSN) was defined for modeling purposes to balance the organic content which was evaluated through an iterative calculation. Knowing the molecular weight of lactic acid, acetic acid and glycerol and their amount contained in the vinasse, the moles of carbon, hydrogen and oxygen were calculated. Based on the average of the elemental analysis of the vinasse found in literature [21], [22], [23], [12], it was evaluated the quantity of C, H and O needed in the non-conventional compound in order to compensate the quantity left, and therefore to simulate

the presence of other compounds other than lactic acid, acetic acid and glycerol. The solid components (ash and organic content) were defined as non-conventional compounds and from a thermodynamic point of view they were simulated based on the built-in enthalpy and density coal correlations, HCOALGEN and DCOALGEN, respectively. Table 3.1 shows the composition of the feedstock.

TABLE 3.1: Concentration of compounds in the feedstock (Vinasse) for Aspen simulation.

Compound	%wt
Glycerol	6.46
Lactic acid	14.44
Acetic acid	4.87
Ash	15,33
Moisture	49.16
$C_v H_w O_x N_y S_z$	9.75

As previously said, the elemental analysis used for the simulation was defined as average between all the values reported in literature for vinasse from sugarcane.

The heating value of the biomass was evaluated through the Boie correlation (Equation 3.1) based on its elemental composition.

$$\Delta H = 100(a_1 x_C + a_2 x_H + a_3 x_S + a_4 x_O + a_5 x_N) + a_6 \quad (3.1)$$

The values of the constants of Boie correlation are shown in Table 3.2.

TABLE 3.2: Constants of Boie correlation [51].

Constant	Btu/lb	kJ/kg
$a_1$	151,2	351,69
$a_2$	499,77	1162,46
$a_3$	45,0	104,67
$a_4$	-47,7	-110,95
$a_5$	27,0	62,80
$a_6$	-189,0	-439,61

Therefore the estimated higher heating value of the biomass was 14,66 MJ/kg on dry basis.

The process modeled can be divided in five main sections (Figure 3.1 shows a block diagram of the process):

- Super critical water gasification (SCWG) (1)
- Gas cleaning (2)
- Fischer Tropsch process (3)

- Product separation (4)
- Utility (5)

For each unit of the process, a parametric analysis has been made in order to find the optimal working conditions. Figure 3.1 shows the block diagram of the process (the numbers on the streams are used to direct the reader to the focus streams in Appendix B). The utility unit will be explained in Section 3.3.

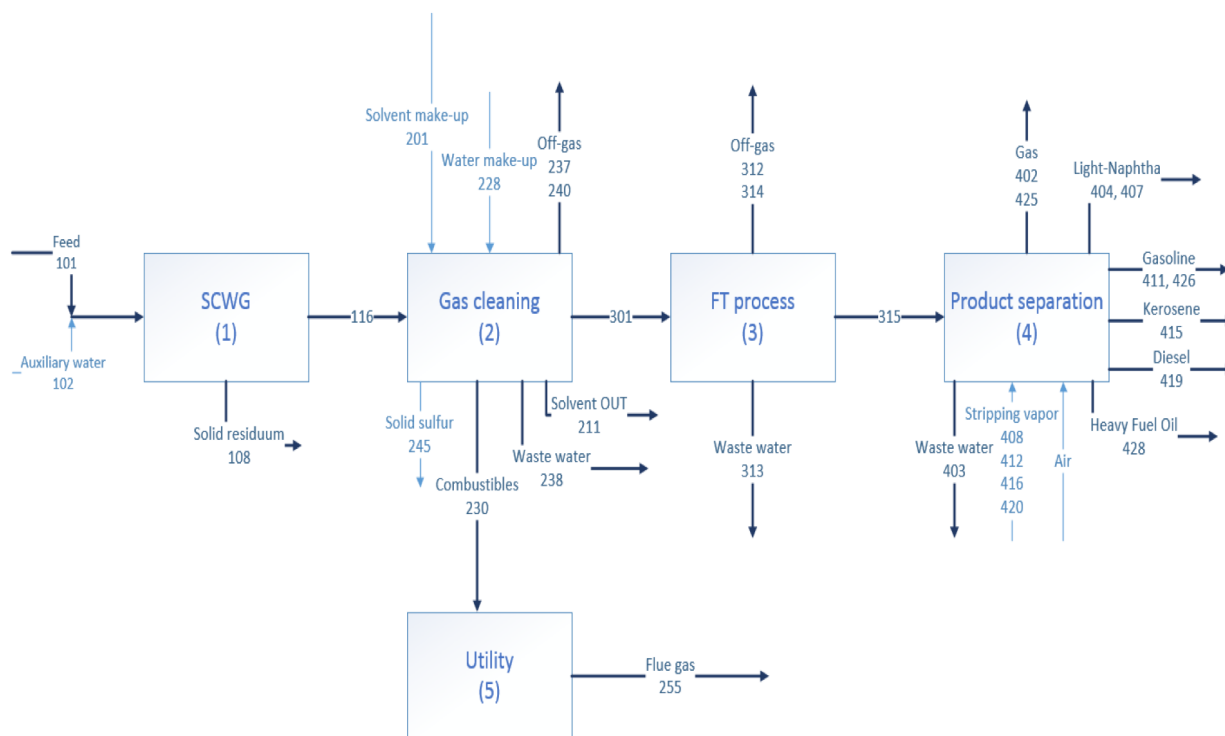


FIGURE 3.1: Block Diagram of the plant (the thicker line shows the Carbon Input and output).

### 3.2.1 Supercritical water gasification

This step has the objective of converting the solid biomass into a syngas through reactions 1.1, 1.2, 1.3 and 1.4. The main parameters that affect the yield and the characteristics of the product required are temperature, pressure and solid content. These conditions were selected based on the actual experiments that were carried out on SCWG of model compounds for vinasse in the Åbo Akademi in Turku with the aim of producing hydrogen. The solid content can vary between 5 and 40 wt%. For this study a value of 10% of solid content has been selected; in this way the probability of obstructing the inlet pipe is limited. If the pipe is blocked from the biomass entering the system, this can lead to an excessive pressure buildup and consequently to an unsafe and dangerous environment.

Before entering the reactor, the water can be preheated and pumped before being mixed with the biomass feedstock, or it can be upgraded to supercritical conditions after being mixed with the biomass. For this study the solution preferred was the second one, therefore the feedstock is mixed with the water and then pumped and heated to enter the reaction system. This decision was made to

reduce the risk of obstructing the pipe with the solid biomass. Figure 3.2 shows the flowsheet of the SCWG unit.

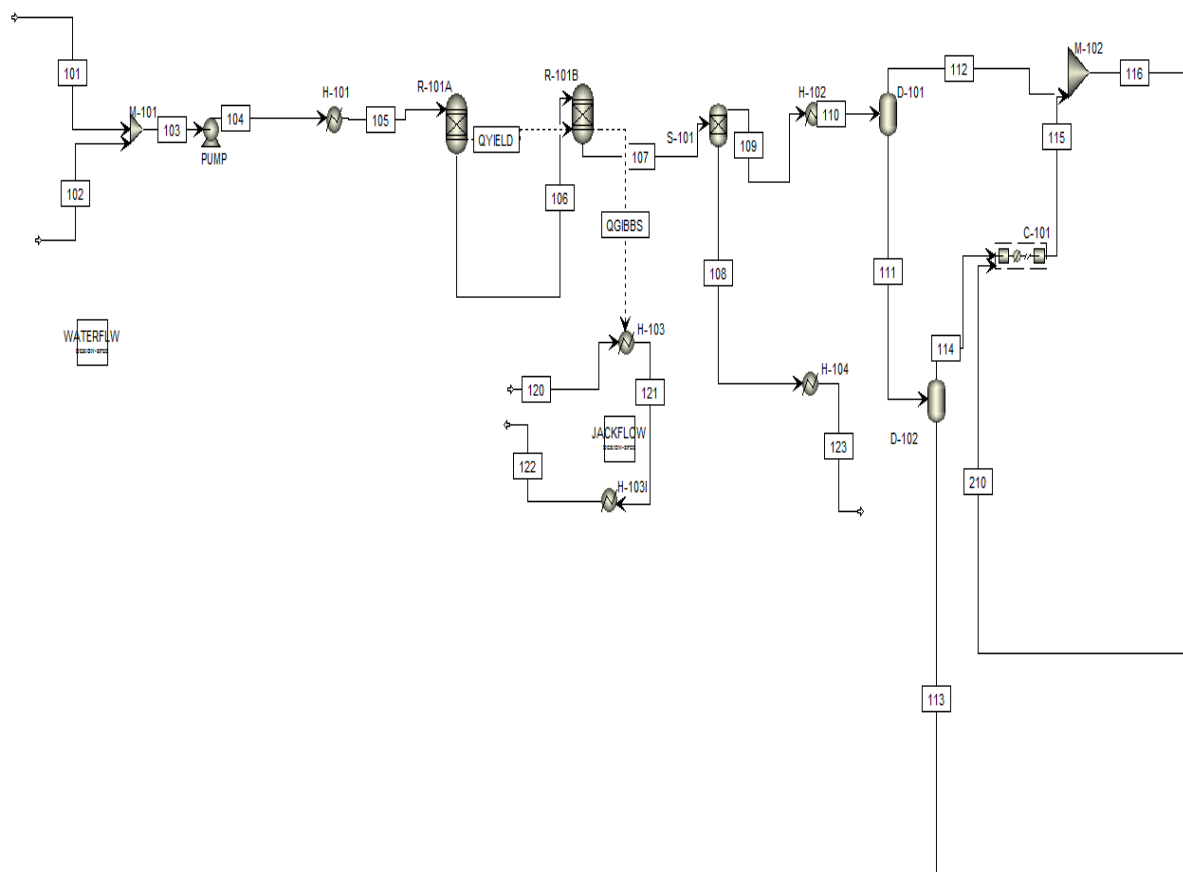


FIGURE 3.2: Snapshot for the SCWG section on Aspen Plus®.

The mixing step was modeled in Aspen Plus® as a Mixer. Afterwards, for the overall plant evaluation, an agitated tank is designed. The designing specifications are reported in appendix A. The reactor temperature selected was 600°C because in the experiments carried out in Åbo Akademi has been detected that the yield at this temperature is comparable to the one at 700°C, therefore it is preferred due to the lower heating load required. In each section of the plant the heating system was at first simulated with HEATER; afterwards an energy analysis has been made with Aspen Energy Analyzer and will be explained in section 3.3. The SCWG system has been modeled in two subsequently steps: a yield reactor (RYIELD) to decompose the slurry into its elemental constituents and a Gibbs reactor (RGIBBS) which calculates the product component formation. This design simulates an hydrolysis, which allows the cleavage of chemical bonds, and subsequent reactions (1.1, 1.2, 1.3, 1.4) [16].

The RYIELD reactor has the purpose to give as output the elemental analysis of the vinasse reported in Table 3.3. The yields in the reactor have been evaluated considering the elemental analysis and a solid content of 10%. Therefore the values inserted in the Aspen plus<sup>®</sup> simulation are shown in Table 3.4.

The Gibbs reactor calculates the formation of the products: hydrogen, methane, carbon dioxide, carbon monoxide water, hydrogen sulfide and ammonia. It was assumed that 98% of the carbon is transformed into products, thus the remaining 2% leaves the plant with the ashes. The working temperature of the reactor is fixed, thus a heating jacket needs to be considered in order to keep the temperature at a constant value. For modeling purposes the jacket was designed as two HEATER (H-103A and H-103B) that simulate the cooling of the auxiliary vapor through the heat load of the reactor.

TABLE 3.3: Elemental analysis of Vinasse for the simulation compared to the literature data.

C	18.86%	17.3%	14.6%	16.92%
H	2.58%	2.11%	3.16%	2.62%
O	20.86%	8.73%	11.38%	13.65%
N	0.57%	2.10%	0.61%	1.09%
S	1.55%	2.10%	0.04%	1.23%
Ash	22.98%	9.79%	13.31%	15.33%
Moisture	32.60%	58.00%	56.90%	49.16%
Reference	[22]	[21]	[23]	Model

TABLE 3.4: Yields of the RYIELD reactor (%) for the decomposition of the vinasse into its elemental analysis.

C	4.22
H	0.60
O	2.26
N	0.18
S	0.20
Ash	2.54
H <sub>2</sub> O	90

The reactor outlet is fed to a cyclone to separate the solid residue from the syngas. It was assumed that all of the solid residue exits within this step. As no information about the particles dimension were detected, a simple SEP was used assuming a complete solid removal. The specific design of the cyclone was not considered in this study. The solid exiting the separator needs to be cooled from 600°C to ambient temperature.

The syngas obtained is at first cooled to 90°C and then fed to a Flash separator for water removal (working pressure is 30 bar). The working conditions of this flash depend on the subsequent gas treatment. In Table 3.5 is reported the composition of the syngas after water removal.

TABLE 3.5: Syngas composition (%) after removal of 99.98% of water.

Compound	Volume%
$H_2O$	2.34
$H_2$	56.09
$CO_2$	21.96
$CO$	0.75
$CH_4$	18.69
$H_2S$	0.09
$NH_3$	0.09

From the composition of the dry syngas it is possible to notice that the working conditions selected allow the production of hydrogen as main product, with a lower production of carbon dioxide and methane. At this point of the process, it is important to focus on the acid gas content of the syngas and its removal. As shown in Table 3.5 the  $H_2S$  content is not negligible, therefore it needs to be removed.

### 3.2.2 Gas cleaning

This section was designed to lower the amount of  $H_2S$  entering the FT block, because it presents limit the activity of the catalyst and it is the cause of its deactivation. Before modeling, a research on the current technologies used was made.

#### Overview on acid gas removal technologies

Various solutions are possible to remove the  $H_2S$  and the  $CO_2$ .

- Claus process

This process is mainly used in oil and natural gas refining and it allows the  $H_2S$  removal by oxidizing it to elemental sulfur. It can be treated with different configurations, leading to a removal efficiency that ranges between 95% and 98% depending on the reactor configuration.

The main reactions that take place in the catalytic reactors are the following:



The Claus process is most effective in streams that contain a large amount of  $H_2S$ , greater than 15% [52].

Hence this solution was not modeled in this study because the  $H_2S$  content is lower (table 3.5).



- Chemical oxidants

Mainly used in wastewater treatment plants, they allow controlling both odor and toxic potential of  $H_2S$ . The most widely used chemical oxidation system is a combination of sodium hydroxide (NaOH) and sodium hypochlorite (NaOCl), which are selected for their low cost, availability, and oxidation capability [52].

- Caustic scrubber

It works similarly to the chemical oxidation system, with the difference that caustic scrubbers are equilibrium limited. This means that if caustic is added,  $H_2S$  is removed, and if the pH decreases,  $H_2S$  is produced [52].

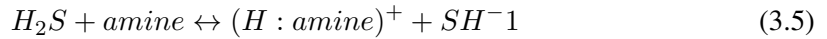
- Adsorption

With the adsorption process, the molecules of the compound that has to be removed are captured to the surface of an adsorbing material. This process can continue until the surface of the material is covered and then the material must be regenerated. The most common material used for adsorption is the activated carbon [52].

- Amine absorption

Amines have the ability to absorb acid gases due to their chemical structure [53]. They are able to remove  $H_2S$  by absorbing it and dissolving it into an aqueous amine stream. The desorbing of the acid components through a heating system allows the regeneration of the stream. Therefore, the acid gas absorption in amine solution is conducted using a two-column operation: the first column is used to absorb the acid gas into the absorbent amine, and the second column is used to regenerate the amine [53].

When  $H_2S$  is exposed to an aqueous amine, or any alkaline solution, it dissociates into the bisulfide ion,  $SH^-$ , by a proton-transfer reaction (equation 3.5) [54].



The  $H_2S$  reacts almost instantaneously with the amines by proton transfer (equation 3.6).



$CO_2$  reacts with primary and secondary amines to form a carbamate (3.7) [54].

Amines that are commonly used are monoethanolamine (MEA), diethanolamine (DEA), and methyldiethanolamine (MDEA).

### MEA

It is an organic compound with the formula  $HOCH_2CH_2NH_2$  generally used as a 10 to 20 weight % solution in water. Due to corrosion problems, the acid gas loading is usually limited to 0,3 to 0,35 moles acid gas per mole of amine for carbon steel equipment [36].

Since MEA is a primary amine, it has a high pH which enables MEA solutions to produce a sweetened gas product [36].

### DEA

It is an organic compound with the formula  $HN(CH_2CH_2OH)_2$  commonly used in the 25 to 35 weight % range. The total acid gas loading for DEA is also limited to 0,3 to 0,35 mole/mole for carbon steel equipment. The degradation products of DEA are much less corrosive than those of MEA [36].

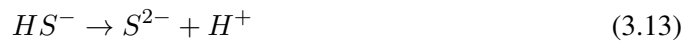
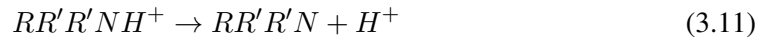
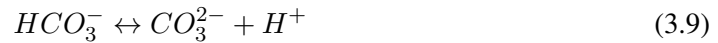
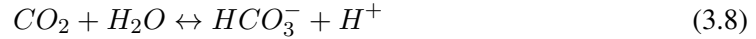
Since DEA is a secondary alkanolamine, it has a reduced affinity for  $H_2S$  and  $CO_2$ , although it is selective to  $H_2S$  under specific conditions [36].

### MDEA

It has the formula  $CH_3N(C_2H_4OH)_2$ . MDEA is commonly used in the 20 to 50%wt range. Solutions with lower content of MDEA are typically used in very low pressure, high selectivity applications. Due to considerably reduced corrosion problems, acid gas loadings as high as 0,7 to 0,8 mole/mole are suitable in carbon steel equipment.

The main advantage of this amine is its selectivity towards  $H_2S$  and  $CO_2$ . The higher selectivity of MDEA for  $H_2S$  results from the inability of tertiary amines to form a carbamate with  $CO_2$ , in fact MDEA does not have a hydrogen attached to the nitrogen and therefore it cannot react directly with  $CO_2$  to form carbamate. The reaction can only take place after the dissolution of  $CO_2$  in the water where it forms a bicarbonate ion [55].

The following chemical reactions occur in an aqueous MDEA solution when  $CO_2$  and  $H_2S$  are present:



Where R (3.11) represents a methyl group and R' an ethanol group [55].

The AGR through amine scrubbing allows a  $H_2S$  removal until 4 ppm, so if some other phases of the plant require a lower value of ppm, a second step of AGR has to be evaluated. If the Fischer Tropsch process follows, the  $H_2S$  content needs to be reduced below 1 ppm [35].

- Zinc Oxide Guard

Used to remove trace amounts of  $H_2S$  from gases at high temperatures (from 200°C to 400°C). Zinc oxide catalyst is used as a mixture of ZnO and alumina as a binder in addition to some fillers or binder materials. In general, 90%wt of ZnO is quite acceptable.

The main reaction in this process is:



The quantity of hydrogen sulfide being adsorbed by zinc oxide depends on its value in the feed and on the degree of contact between it and the ZnO bed [56].

Maximum sulfur loading is typically in the range of 30-40 kg sulfur/100 kg sorbent. When the zinc oxide reacts with the  $H_2S$  an insoluble layer of zinc sulfide is formed and approximately 40% of the  $H_2S$  is converted [52], [57].

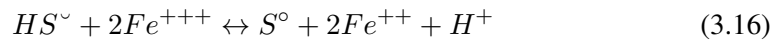
- LO-CAT<sup>®</sup> process

It is a process commercially used and it allows the conversion of the gas into solid sulfur. It is patented process that consists in a wet scrubbing, liquid redox system that uses a chelated iron solution to convert  $H_2S$  to solid sulfur. It mainly consists in the oxidation of the  $H_2S$  and it can be divided in five steps.

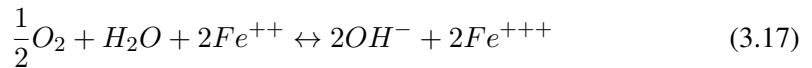
1. Absorption of  $H_2S$
2. Ionization of  $H_2S$



3. Sulfide Oxidation



4. Absorption of Oxygen
5. Iron Oxidation



Overall, the reaction can be expressed as:



This process, carried out at ambient working conditions (25°C and 1 bar) allows the conversion of the 99,99% of  $H_2S$  into solid sulfur [58].

## Design

The Gas cleaning was consequently designed: it is composed of four main sections, as illustrated in Figure 3.3.

Therefore the main sections are:

1. Amine scrubbing, to reduce the  $H_2S$  content in the gas fed to the FT;

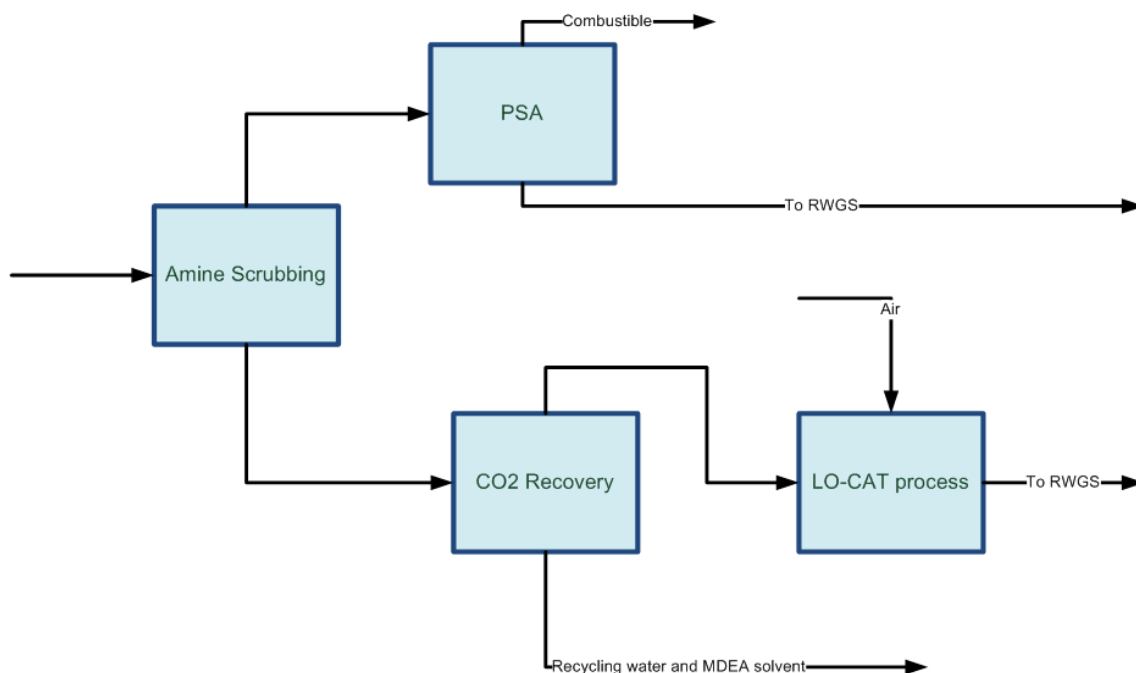


FIGURE 3.3: Block diagram of the Gas cleaning unit.

2. Pressure Swing Adsorption (PSA), in order to obtain a high purity hydrogen stream to send to the reactor;
3.  $CO_2$  recovery, to reduce the carbon losses of the plant;
4. LO-CAT<sup>®</sup> process, to further reduce the amount of  $H_2S$ .

Figure 3.4 shows the flowsheet of the gas cleaning unit.

1. **Amine scrubbing:** This step was evaluated in order to reduce the  $H_2S$  and  $CO_2$  content before the gas enters the Fischer Tropsch reactor because the content of sulfur tolerated by the FT catalyst is lower than 1 ppm [59].

The AGR was made through a MDEA absorption with a 30%wt MDEA solution. The MDEA is able to remove 0.4 mol of acid gas per mol of amine [38]. The amount of solution needed was thus evaluated. At first attempt, the stages needed for the extraction were evaluated with a simplified calculation on Excel for an ideal system. The design is reported in appendix C. The main assumption that has been made in the first extractor design is constant flow rates entering and exiting the extractor. With this, an iterative calculation has been carried out until the  $H_2S$  content was reduced below 4 ppm. The results obtained from the excel calculation were then simulated in Aspen Plus<sup>®</sup> and adjusted using a RADFRAC with 4 stages without a boiler nor a condenser. A sensitivity analysis allowed the evaluation of the optimal working conditions which were 90°C and 30 bar with a solvent mass flow of 67 ton/hr.

2. **Pressure swing adsorption:** Due to the composition of the sweet gas (Table 3.6) a pressure swing adsorption step was evaluated in order to recover  $H_2$  stream with high purity for the RWGS step leaving a combustible stream (219) that has a potential as hot utility for the plant (composition is shown in Table 3.7).

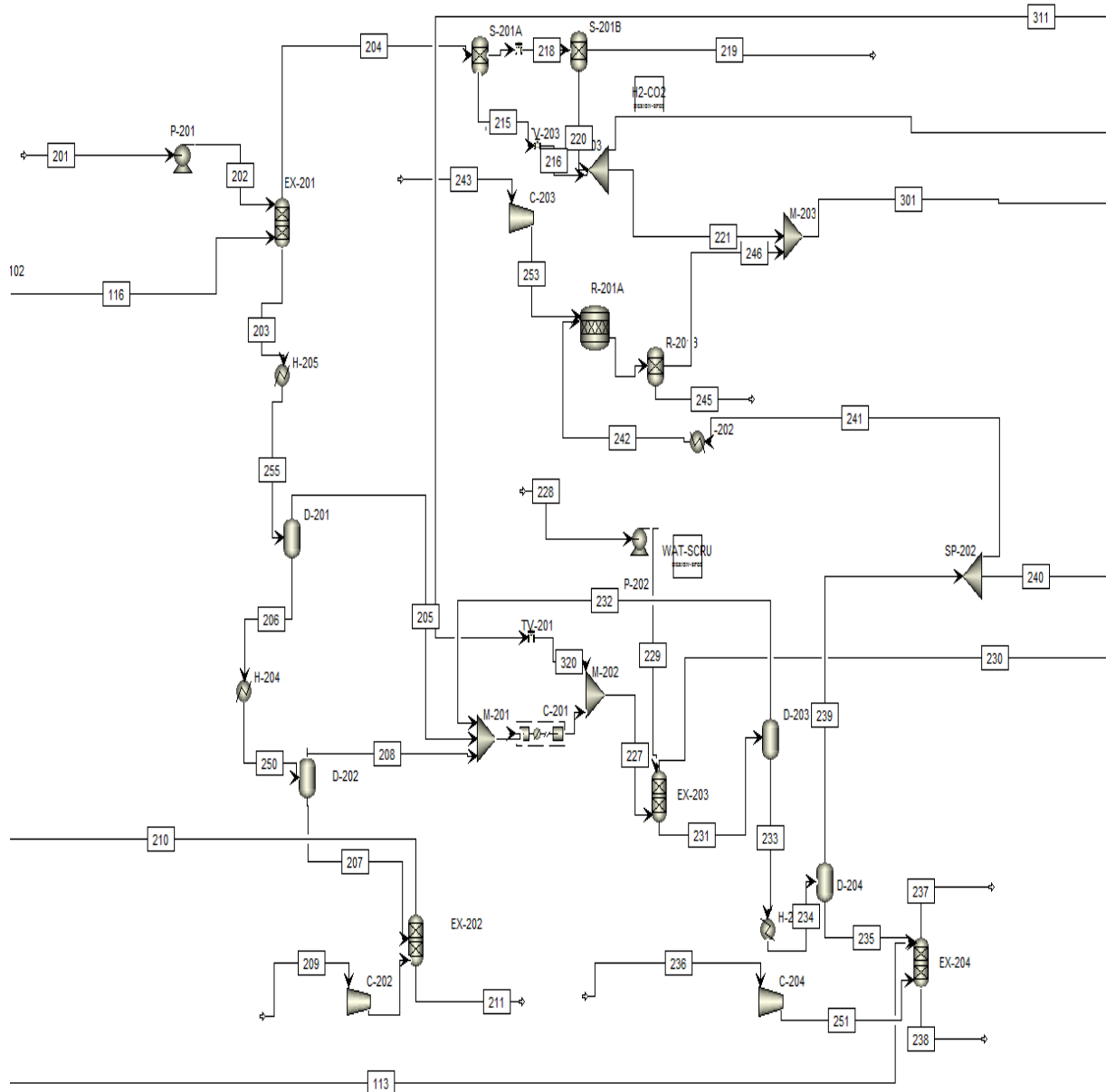


FIGURE 3.4: Snapshot of gas cleaning unit

TABLE 3.6: Sweet gas molar composition (stream 204 in Figure 3.4).

Compound	%
$H_2$	80
$N_2$	2
$CH_4$	18

Pressure swing adsorption (PSA) processes are generally used for the production of high purity hydrogen. It is technology based on a physical binding of gas molecules to adsorbent material.

TABLE 3.7: Major compounds in stream 219 (shown in figure 3.4).

Compound	vol%
$H_2$	39
$N_2$	5
CO	0.4
$CH_4$	54

The separation effect is based on differences in binding forces to the adsorbent material. The adsorption step works at high pressure typically in the range between 10 to 40 bar until the equilibrium loading is reached. In this condition, the maximum adsorption capacity is reached, and the adsorbent material needs to be regenerated. This is done by lowering the pressure resulting in a respective decrease in equilibrium loading [60].

The PSA unit was modeled as a two-stage separation with a recovery of 80% of  $H_2$  for each step. This two stages allow the recovery of a 99.99%  $H_2$  purity stream which is fed to the RWGS reactor. This allows the reduction of the amount of impurities that flow in the process.

### 3. $CO_2$ recovery :

This section was designed for two reasons: to recover of the MDEA solvent, and therefore to reduce the make up along with the cost of the plant, and to reduce the carbon losses of the plant. The  $CO_2$  recovered can be used for the RWGS step since it is one of the reactants.

The  $CO_2$  recovery can be divided in two steps:

- Solvent recovery
- Water scrubbing

As a result the  $CO_2$  (~83%) rich stream exiting this step contains a high amount of sulfur content, thus a LO-CAT<sup>®</sup> was modeled.

#### Solvent recovery

The solvent exiting the extractor (stream 203) was fed to a FLASH unit (block D-201) working at 1 bar and 25°C: this allows the recovery of the 99,99% of MDEA in the liquid stream, although a high amount of  $CO_2$  (~95% of it) exits with the liquid stream. This (stream 206) is heated to 45°C and fed to a second Flash step (Block D-202) where ~81% of the  $CO_2$  is removed from the liquid stream (207). Stream 207 is then fed to a desorption system where the 99% of  $CO_2$  and  $H_2S$  are recovered in the top stream through the air fed to the block. The gaseous stream (210) is then recirculated to the MDEA adsorption unit in order to reduce the carbon losses. The liquid stream (211) is composed by almost the 30%wt of MDEA and around 70% of Water (some impurities are present), therefore it could be recycled to the extractor lowering the solvent make-up (in the simulation the MDEA stream was simulated as pure solvent). Figure 3.5 shows a simplified block diagram of the solvent recovery step. Stream 210 is recycled to the amine adsorption due to its high content of  $H_2S$  (~4%).

#### Water scrubbing

The water scrubbing was evaluated due to the high content of methane present in the gaseous

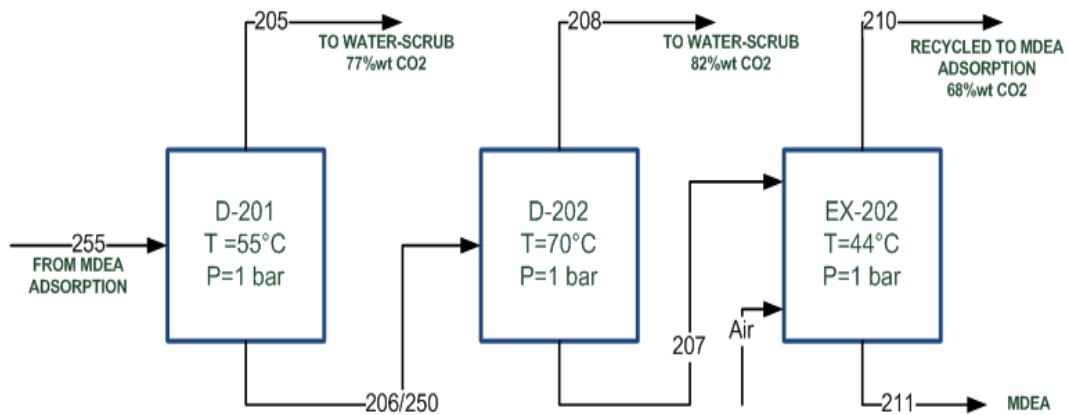


FIGURE 3.5: Block diagram of solvent recovery step.

streams. This step allows the separation of methane from  $\text{CO}_2$ . The methane stream (230) is mainly composed of combustible compounds (table 3.8 shows its composition), therefore it was evaluated as potential hot utility for the plant (more detailed explanation in Chapter 3.3 ).

A water scrubber is a physical scrubber that uses the property of carbon dioxide to have higher solubility than methane in water. The carbon dioxide is separated from the gas and dissolved into the water in the absorption column by using a pressure between 6 and 10 bar, consequently it is released from the water in the desorption column by addition of air at atmospheric pressure (schematic illustration in Figure 3.6) [61].

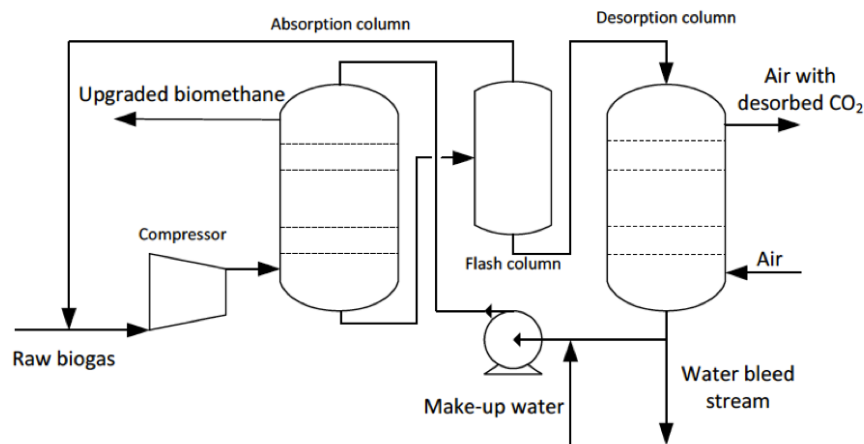


FIGURE 3.6: Schematic illustration of a common water scrubber [61].

In this study, the water scrubbing was modeled in four blocks: as a

- Adsorption column (block EX-203), where the separation of the  $\text{CH}_4$  stream happens;
- Flash column (block D-203), where the water is separated from the gas and the gas is recycled to the adsorption column;
- Gas-separation column (block D-204), for the separation of the remaining  $\text{CO}_2$  from the water;

TABLE 3.8: Major compounds of the combustible stream (230, shown in figure 3.4) (%wt).

Compound	%
$H_2O$	0,19
$H_2$	4,07
$N_2$	13,84
$CO_2$	0,19
$CO$	35,14
$CH_4$	37,53
$O_2$	4,91
$C_2H_6$	0,45
$C_2H_4$	0,23
$C_3H_8$	0,86
$C_3H_6$	0,39
$C_4H_{10}$	1,19
$C_4H_8$	0,46
$C_5H_{12}$	0,31
$C_5H_{10}$	0,16

- Desorption (block EX-204), for the recovery of high purity water in the bottom and off-gas in the top.

The adsorption step was modeled at 6 bar and the desorption step was simulated as atmospheric a flash (block D-204) for the separation of  $CO_2$  from the water. The optimal working conditions (3 bar and 75°C) were selected based on the  $H_2S$  content in the gaseous stream (239). Hence, the recovery of  $CO_2$  is limited by the content of  $H_2S$ . This configuration allows the recovery of the ~99,5% of water in the bottom stream, and the of ~70% of  $CO_2$  in the top stream. The stream fed to the LO-CAT<sup>®</sup> reactor is thus mainly composed of  $CO_2$  (~92%wt). Therefore differently from what reported in figure 3.6 in this study a second flash was modeled because of the high sulfur content which limits the amount of  $CO_2$  that can be recovered.

4. **LOCAT<sup>®</sup> process:** The LO-CAT<sup>®</sup> process for sulfur removal was selected to remove the  $H_2S$  content because compared to the zinc oxide guard, it allows higher conversions and therefore a lower  $H_2S$  content in the stream fed to the FT step. For modeling purposes, it was used a RSTOIC to model the conversion to elementary sulfur with a 99.99% of conversion and a separator to simulate the total separation of the solid sulfur from the gas stream. Air is fed to the reactor at stoichiometric conditions so that the equation 3.18 occurs. The possibility of feeding a stream with only  $O_2$  was evaluated, but the fraction of  $N_2$  that remains in the process after the LO-CAT<sup>®</sup> reactor was considered acceptable (below 1%). As a result, the ppm of  $H_2S$  leaving the unit are ~0.19.

At the end of the design of the plant, it was found a significant carbon loss, and therefore some streams were recycled to the  $CO_2$  recovery step. This caused several problems of convergence both because of the complexity of the flowsheet and also because the recycling changed the composition of the streams. This affected the working conditions previously decided, and thus several sensitivity analysis and design specification were made. For instance, due to the



characteristics of the process of RWGS and FT, which will be described in the next section, two design specification were used to control the ratio between the reactants in each reactor. Furthermore, the working conditions of the flash columns found at the bottom of the amine scrubbing were selected through sensitivity analysis in order to maximize the recover of  $CO_2$ , without allowing a simultaneous recover of  $H_2S$ .

### 3.2.3 RWGS and Fischer Tropsch

Figure 3.7 shows the flowsheet of the RWGS and FT unit.

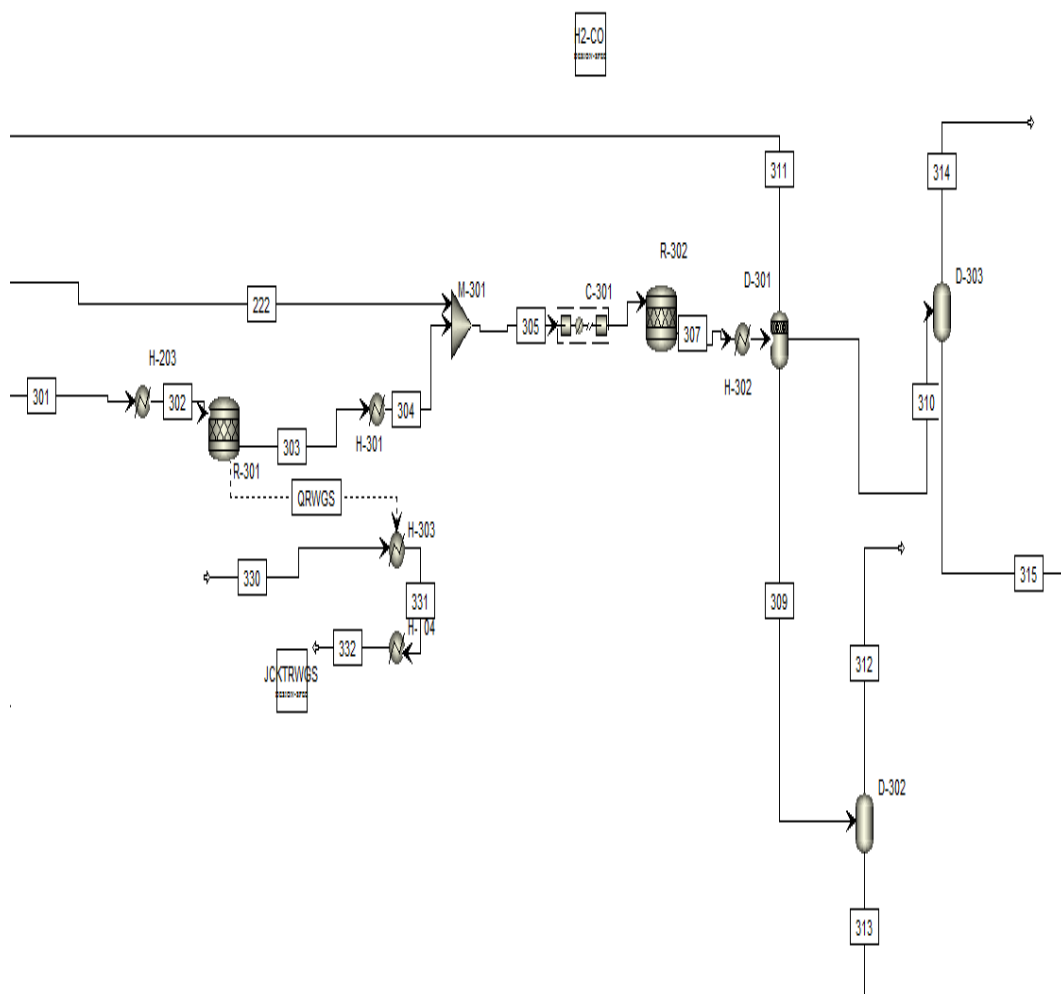
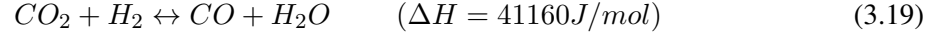


FIGURE 3.7: Snapshot of the RWGS and FT unit

#### RWGS

The ratio required from the Fischer Tropsch reactor is adjusted through a water gas shift (1.3) or a reverse water gas shift (3.19) depending on the stream composition. In this study the RWGS reactor was modeled for this purpose.

The ratio of  $H_2/CO$  happens to be high ( $\sim 84$ ), for this reason a reverse water gas shift is needed (Equation 3.19).

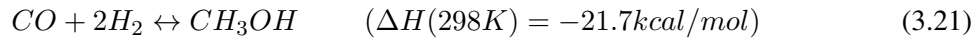


The reaction is mildly endothermic and therefore favored at higher temperature. It is necessary to select an efficient catalyst capable of bearing high temperatures and being selective to  $CO$  during the RWGS reaction.

The possibility of the formation of side products have been evaluated: with the same reactants it is possible the synthesis of methane (Equation 3.20) and methanol (Equations 3.21 and 3.22).



For this study the side reactions were not considered due to the catalyst selected.



As shown in the equations, both the methanol and methane reaction are exothermic, so higher methanol yields are obtained at lower temperatures and higher pressures, while the RWGS is endothermic. Thermodynamic evaluations at atmospheric pressure show that  $CO_2$  conversion in the RWGS reaction is greater with an excess of  $H_2$  flowing. The equilibrium composition changes with the temperature. It is possible to graph the molar distribution of the products with the temperature, as illustrated in Figure 3.8

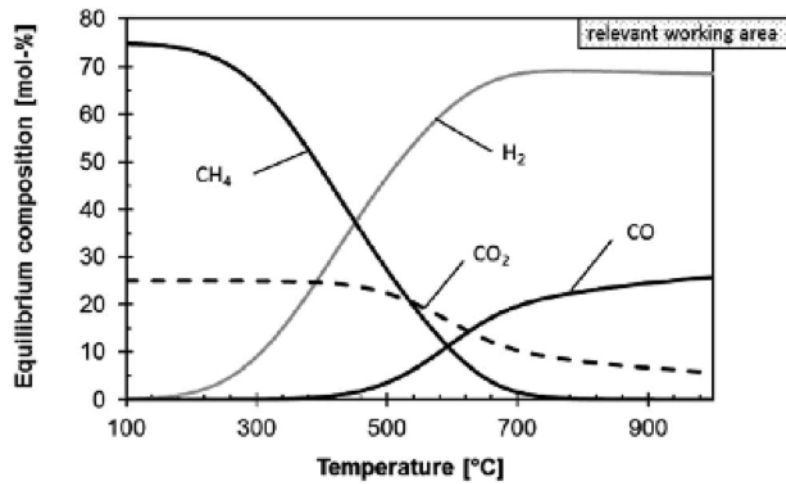


FIGURE 3.8: Influence of temperature on the thermodynamic equilibrium of the RWGS reaction at 1 bar and  $H_2/CO_2$  ratio of 3 ([62]).

To determine the working conditions of the RWGS reactor, an evaluation of the possible catalysts has been made in order to understand the conversion and the selectivity to  $CO$ . Porosoff et al.[41] reported a study of the catalyst used for the conversion of  $CO_2$  into  $CO$ , methanol and methane.

The selectivities and conversions of the catalysts are shown in Table 3.9. The catalyst selected for the simulation is the  $NiO/CeO_2$ ; it allows 40% of  $CO_2$  conversion and almost a 100% selectivity

TABLE 3.9: Summary of reaction conditions with conversion and selectivity to CO for RWGS catalysts [41].

Catalyst	$H_2:CO_2$ ratio	Temperature (C)	Pressure (MPa)	Conversion (%)	Selectivity (%)
NiO/CeO <sub>2</sub>	1:1	700	0.1	~40	~100
Cu/Al <sub>2</sub> O <sub>3</sub>	1:9	500	N/A	~60	N/A
Co/MCF-17	3:1	200-300	0.55	~5	~90
Pt-Co/MCF-17	3:1	200-300	0.55	~5	~99
Cu/SiO <sub>2</sub>	1:1	600	0.1	5.3	N/A
Cu/K/SiO <sub>2</sub>	1:1	600	0.1	12.8	N/A
Cu-Ni/ $\gamma$ - Al <sub>2</sub> O <sub>3</sub>	1:1	600	0.1	28.7	79.7
Cu-Fe/SiO <sub>2</sub>	1:1	600	0.1	15	N/A
Li/RhY	3:1	250	3	13.1	86.6
Rh/SiO <sub>2</sub>	3:1	200	5	0.52	88.1
Rh/TiO <sub>2</sub>	1:1	270	2	7.9	14.5
Fe/TiO <sub>2</sub>	1:1	270	2	2.7	73.0
Rh-Fe/TiO <sub>2</sub>	1:1	270	2	9.2	28.4
Fe-Mo/ $\gamma$ -	1:1	600	1	~45	~100
Mo/ $\gamma$ - Al <sub>2</sub> O <sub>3</sub>	1:1	600	1	34.2	97
Pd/Al <sub>2</sub> O <sub>3</sub>	1:1	260	0.1	N/A	78
Pd/CeO <sub>2</sub>	1:1	260	0.1	N/A	87
Pd/La <sub>2</sub> O <sub>3</sub>	1:1	260	0.1	N/A	70
CeO <sub>2</sub> -Ga <sub>2</sub> O <sub>3</sub>	1:1	500	0.1	11.0	N/A
Pt/TiO <sub>2</sub>	1.4:1	400	N/A	~30	N/A
Pt/Al <sub>2</sub> O <sub>3</sub>	1.4:1	400	N/A	~20	N/A
PtCo/CeO <sub>2</sub>	3:1	300	0.1	3.3	71.0
Co/CeO <sub>2</sub>	3:1	300	0.1	3.8	39.4
PtCo/ $\gamma$ - Al <sub>2</sub> O <sub>3</sub>	3:1	300	0.1	5.1	89.4
Co/ $\gamma$ - Al <sub>2</sub> O <sub>3</sub>	3:1	300	0.1	3.8	67.0
Mo <sub>2</sub> C	3:1	300	0.1	8.7	93.9
Mo <sub>2</sub> C	5:1	250	2	17	34
Cu-Mo <sub>2</sub> C	5:1	250	2	13	40
Ni-Mo <sub>2</sub> C	5:1	250	2	21	29
Co-Mo <sub>2</sub> C	5:1	250	2	23	24

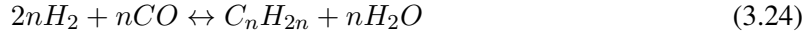
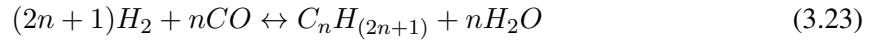
to CO, so no methane and methanol were considered as products. Therefore, the working conditions are 700°C and 1 bar. For modeling purposes a RSTOIC reactor with a fractional conversion of 0.4 of CO<sub>2</sub> was used.

As shown in Table 3.9, the  $H_2/CO_2$  ratio entering the reactor has to be equal to 1. For this reason, a split (SP-202) was added and it allows the monitoring of the ratio entering the reactor through a design specification.

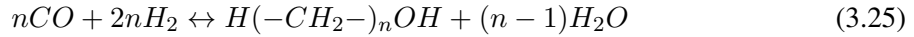
### Fisher Tropsch

The Fischer Tropsch reactor allows the synthesis of a crude from CO; the CO reacts with hydrogen to produce paraffins, olefins and water [59].

In the Fischer-Tropsch process, CO and H<sub>2</sub> are converted into hydrocarbons (alkanes and alkenes) of various molecular weights according to the Equations 3.23 and 3.24 thanks to the presence of a catalyst.



Along with the hydrocarbons synthesis, alcohol synthesis (3.25) and coke formation (3.26) take place as side reactions.



For this study the side reactions were not considered due to the catalyst selected because of the high selectivity to alkanes and alkenes. FT conditions are chosen in order to maximize the formation of higher molecular weight hydrocarbon, which form a higher value product [63].

The products and the hydrocarbons produced by this process vary with the  $H_2/CO$  ratio.

One of the main characteristic of FT synthesis is the liberation of large amounts of heat from the highly exothermic synthesis reactions [64]. The FT process, is analogous to the polymerization process, hence it shows a variety of reaction steps, such as chain propagation, chain branching, olefins and paraffins desorption, methanation and olefin secondary reactions of double bond shift and hydrogenation [64].

The product distribution is mainly affected by the chain growth probability parameter ( $\alpha$ ), so according to this value the quality and the characteristics of the biocrude differ. In order to produce a diesel fuel this factor should be between 0.85 and 0.9, as shown in Figure 3.9 [65].

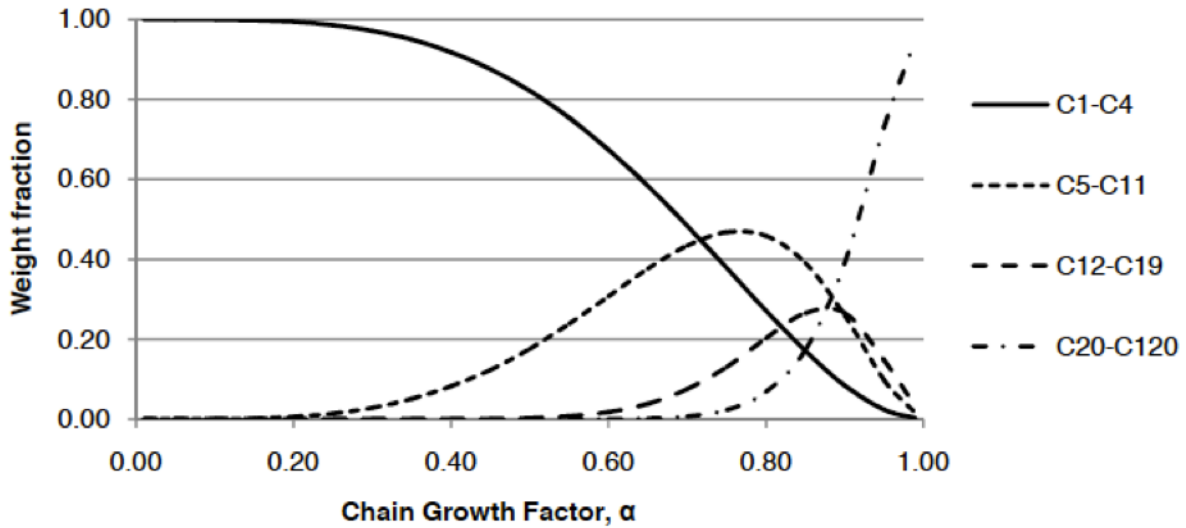


FIGURE 3.9: Fischer-Tropsch hydrocarbons distribution with the growth chain factor [65].

The FT process is catalytic, so different catalysts have been studied in literature [66], [67]. Usually the catalysts used are based on transition metals of iron, cobalt, nickel and ruthenium. The FT catalyst development has largely been focused on the preference for high molecular weight linear alkanes and diesel fuels production.

The activity depend on the  $H_2$  adsorption capacity and on the reducibility of the other metal oxide components [67]. Between the metals mentioned, Ruthenium is one of the most active catalyst for the FT process operating at low temperature. It allows the production of long chain hydrocarbons without the need of any promoters. However, in industrial applications other catalysts are preferred due to its elevated cost [68]. Iron and Cobalt are thus the most suitable catalyst for this application. Fe has a very low selectivity to paraffins, promoting the production of olefins; it is convenient from an economic point of view but it deactivates more quickly than the Cobalt-based catalysts. On the other hand, Co has a good selectivity to paraffins, low selectivity to olefins and oxygen, it is more resistant to deactivation, but it is more expensive than Fe [68].

For this study an iron based catalyst was selected [69] due to its higher tolerance to sulfur content [59].

The parameter used to understand the hydrocarbon formation with the FT process is the chain growth factor ( $\alpha$ ) which can be evaluated with the Equation 3.27. Figure 3.9 shows its influence in the distribution.

$$\alpha = (0,2332 \frac{y_{CO}}{y_{CO} + y_{H_2}} + 0,633)(1 - 0,0039(T - 533)) \quad (3.27)$$

In Equation 3.27,  $y$  represents the mole fractions of the compound and  $T$  is the temperature of the reactor expressed in Kelvin (K). This correlation is only dependent on  $H_2$  and  $CO$  composition in the gas [38]. The methodology for the FT reactor design was based on the method used by Pondini [38]. It is based on a defined temperatures ( $220^\circ$ ), which implies a specific  $\alpha$  value. The value of  $\alpha$  selected (0.82) is the most suitable for diesel fuel production. The distribution of the different hydrocarbons due to the reactions occurring, and therefore the  $CO$  conversion implemented in the software, are based on the hydrocarbon selectivities on a Fe-based catalyst at  $220^\circ$  C, 15 bar and 72% conversion of  $CO$  found in literature [69]. A detailed calculation of the FT reactor is reported in Appendix D. As a result it is obtain a crude which is mainly kerosene ( $C_{10}$ - $C_{16}$ ) and diesel ( $C_{17}$ - $C_{22}$ ).

### 3.2.4 Product separation

Figure 3.10 shows the flowsheet of the product separation unit.

The crude produced in the FT reactor needs to be treated in order to produce valuable bio-fuels. Therefore, a system to split the crude into the various distillates was modeled. Before entering the unit, the FT-Crude is fed to a three phases separator, where the crude is separated from the gas and from the water.

The composition of the FT-Crude allows to understand the distillates that can be produced (Figure 3.11). From the graph, it is possible to notice that the main hydrocarbons found are from  $C_{10}$  to  $C_{22}$ . This is due to the reactor design which, as previously said, allows the production of a crude of mostly kerosene and diesel. In fact, depending on the carbon number, each hydrocarbon can be distillate in a specific fraction, for instance hydrocarbons  $C_{10}$ - $C_{16}$  are found in the kerosene and  $C_{17}$ - $C_{22}$  in the diesel fraction.

The separation in multiple distillates is modeled in three steps:

1. Light naphtha and water separation
2. Production of gasoline, kerosene and diesel
3. FCC for gasoline production from the fractionation column residuum

For each step the appropriate parameters have been evaluated.

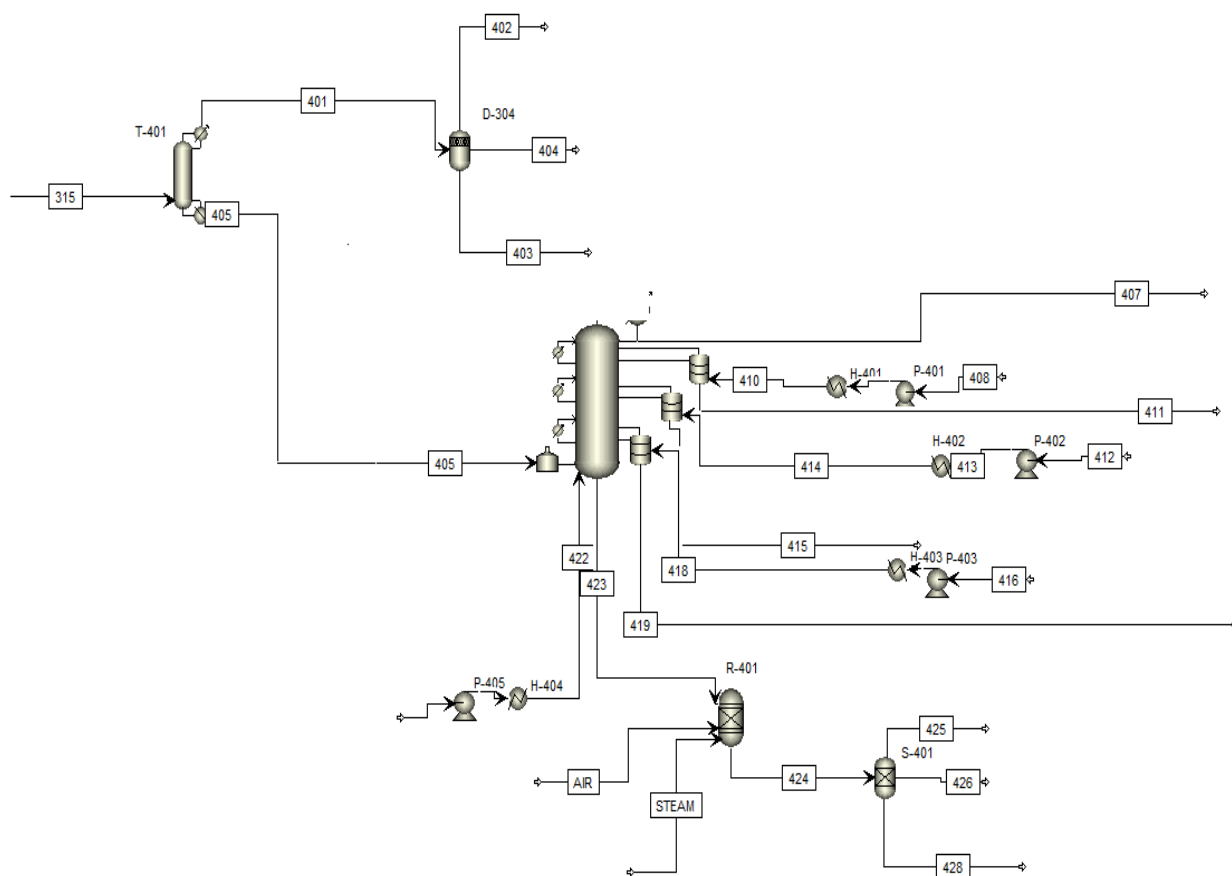


FIGURE 3.10: Snapshot of the product separation unit.

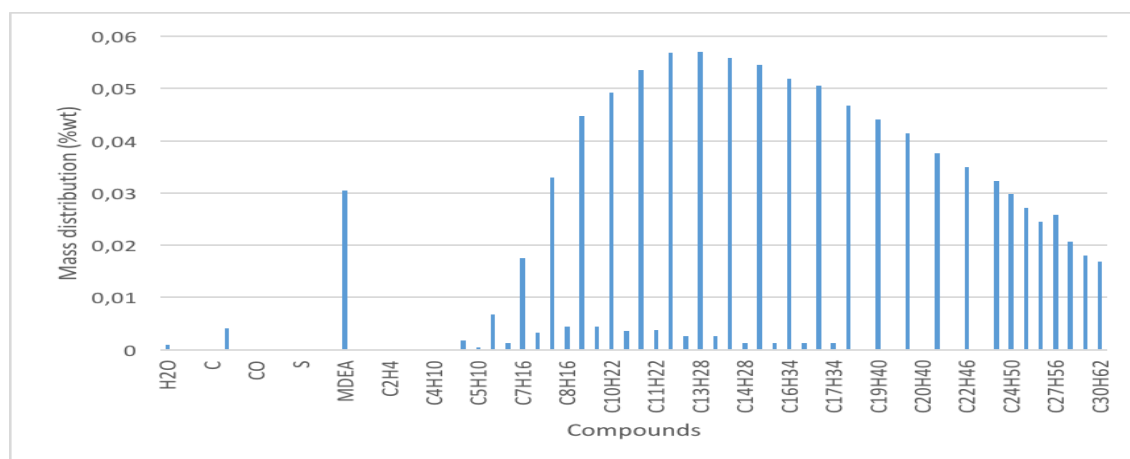


FIGURE 3.11: Hydrocarbons distribution in FT-Crude (%wt).

### Light Naphtha and water separation

In this step a distillation column (T-401) with total condenser and partial reboiler was modeled. A snapshot of the *Light Naphtha and water separation* section is shown in the left top part of Figure 3.10. The Distillate to Feed Ratio and the Reflux Ratio were varied until all the Light naphtha hydrocarbons were found in the top stream and the heavier ones remained in the bottom to be fed to the fractionation column. The column was modeled of 25 stages with a Kettle reboiler and a total condenser. After the column a separator was modeled to separate the light naphtha from the water and the off-gas (D-304).

### Production of Heavy Naphtha, Kerosene and diesel

In this step a atmospheric fractionation column (T-402) was designed to separate the distillates. The *Production of heavy naphtha, kerosene and diesel* section is shown in Figure 3.10. The distillate that can be extracted from the column are based on the carbon number.

- $C_1 - C_4$ : Light naphtha
- $C_5 - C_9$ : Heavy naphtha
- $C_{10} - C_{16}$ : Kerosene
- $C_{17} - C_{22}$ : Diesel
- $C_{22+}$ : Waxes

In petroleum refining the boiling points are used to characterize a fraction instead of mass and molar distribution. The curves used to describe the boiling point temperature for % of distillate are the true boiling point (TBP) and the ASTM D86. For modeling purposes, these curves, which were used in the streams specification for the fractionation column, were evaluated from the *Stream Analysis* tool of Aspen Plus®.

Both the fractionation and the FCC unit were simulated on a different Aspen flowsheet due to the characteristics of the distillates. The method to calculate the frictions is to break them into pseudo-components that have their own specific gravity (SG), API Gravity and molecular weight [42]. The crude fed to the column was then defined as Blend of the different pseudo-components knowing their mass fraction in the crude. For petroleum fractionation, usually it is specified along with the TBP, the sulfur content curve. In our case the sulfur content was reduced in the AGR unit to approximately zero, therefore its curve was not taken into account.

The column was modeled as a 25-stages column with a furnace, three strippers and one pump around. The furnace has the aim to partially vaporize the feed. It was located almost at the bottom, at the stage number 22 and at temperature of 370° that allows the vaporization of ~70% of the feed (data taken from literature [70]) The three strippers allow the removal of the distillates. To model the strippers it was necessary to evaluate the amount of each distillate that was possible to strip and extract from the column: this calculation was easily made knowing the hydrocarbons composition of the distillate and their content in the feed stream. To allow the extraction of the frictions, a stripping vapor stream is necessary for each stripper and also at the bottom of the column. The vapor entering the strippers is at 200°C and 4 bar.

The mass flows of the stripping vapor stream were evaluated knowing the quantity of each distillate that can be stripped in each stripper from the Graph 3.12 [70].

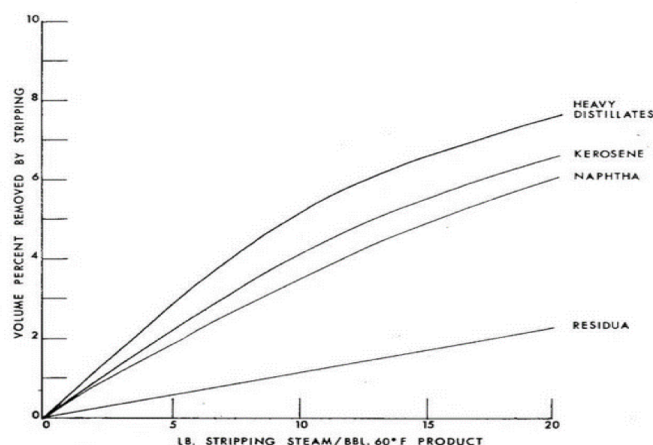


FIGURE 3.12: Stripping vapor [70].

The  $C_{22+}$  hydrocarbons, the waxes, are found in the bottom stream as residuum to be fed to the FCC step.

### Fluid Catalytic Cracking

Fluid catalytic cracking (FCC) is a process that allows the cracking of the heavy hydrocarbons present in the residuum into gas and gasoline. It is used in Petroleum refinery to convert heavy gas oil into more valuable lighter hydrocarbon products. The conversion happens because of the catalytic cracking of the heavy gas oil on the catalytic surface. The catalytic cracking reaction is carried out in a short contact-time vertical reactor called the riser: the reactor is a fluidized bed in which the residence time of the catalyst-hydrocarbon mixture is just for few seconds. A steam stream, normally around 5% of the oil, is fed along with the residuum to allow the conversion [71]. After the reaction, the catalyst needs to be regenerated and this step can happen by combusting the formed coke with air [43]. A schematic representation of the FCC unit is shown in Figure 3.13

To model this step, the reactor yields were taken from the literature. The snapshot of the block is shown in Figure 3.10. Various studies were made on how the yields change with temperature; the temperature condition chosen for this study was 221°C and 2.9 bar. As a result, the gasoline, gas, coke and residuum yield were inserted in RYIELD (Table 3.10)[43].

TABLE 3.10: FCC yields at 221°C [43].

Gasoline	0.56
Coke	0.07
Gas	0.18
Residuum	0.18

For the Aspen Plus® modeling, a RYIELD reactor was used with the yields shown in Table 3.10.

For modeling purposes, gasoline, gas, coke and residuum were inserted as pseudo-components in Aspen properties. For each pseudo-component the appropriate parameters were inserted: for gas and coke the molecular weight and the normal boiling point (NBP) and for gasoline and residuum the NBP and the API Gravity. In the Yield reactor were fed also the vapor stream needed from the reaction and the air stream for the catalyst regeneration in order to simulate the correct utility needed



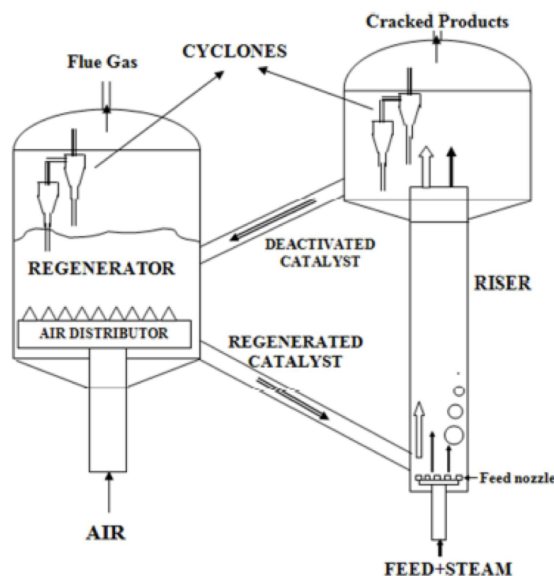


FIGURE 3.13: Schematic representation of FCC Unit [43].

from the block. Consequently the coke produced in the riser reactor does not exit from the block because it is combusted through air during the regeneration of the catalyst.

### 3.3 Energy Analysis

The energy analysis was made through the pinch point analysis that allows the understanding of the loads required from the plant. The Energy Analyzer tool gives an information on the amount of energy that can be saved, so the amount of heat that can be recovered along the plant. Aspen Energy analyzer is used to calculate the maximum energy recovery of the plant based on the temperature-heat duty profile. Due to the characteristic of the combustible streams exiting from the plant (streams 219 and 230) a boiler was designed to recover the heat. For modeling purposes the boiler was simulated as a RGIBBS that simulates the complete combustion of the compounds and it was assumed adiabatic working conditions ( $Q_{losses} = 0$ ). Table 3.11 shows the composition of the streams before and after combustion. To control the air to fuel ratio it has been used a *Design Specification*. For modeling purposes it has been designed in order to have a  $O_2$  %wt in the outlet stream of 6%. Furthermore, it has been assumed a preheated air stream at a temperature of  $450^\circ\text{C}$ . To recover at maximum the heat load available, the flue gas stream (502) is cooled to the pinch point and then to  $120^\circ\text{C}$ . Figure 3.14 shows the flowsheet of the utility unit.

For the endothermic reactors of the plant, for instance the SCWG and the RWGS, the temperature working condition needs to be maintained through a heating jacket. For modeling purposes, the heating jackets were represented as heat exchangers that simulate the heat load required by the reactor to maintain the correct working condition. It has been considered a  $\Delta T$  of  $15^\circ\text{C}$  for the stream crossing the heating jacket. Figure 3.15 shows a snapshot of the Heat Jacket modeled in Aspen plus®.

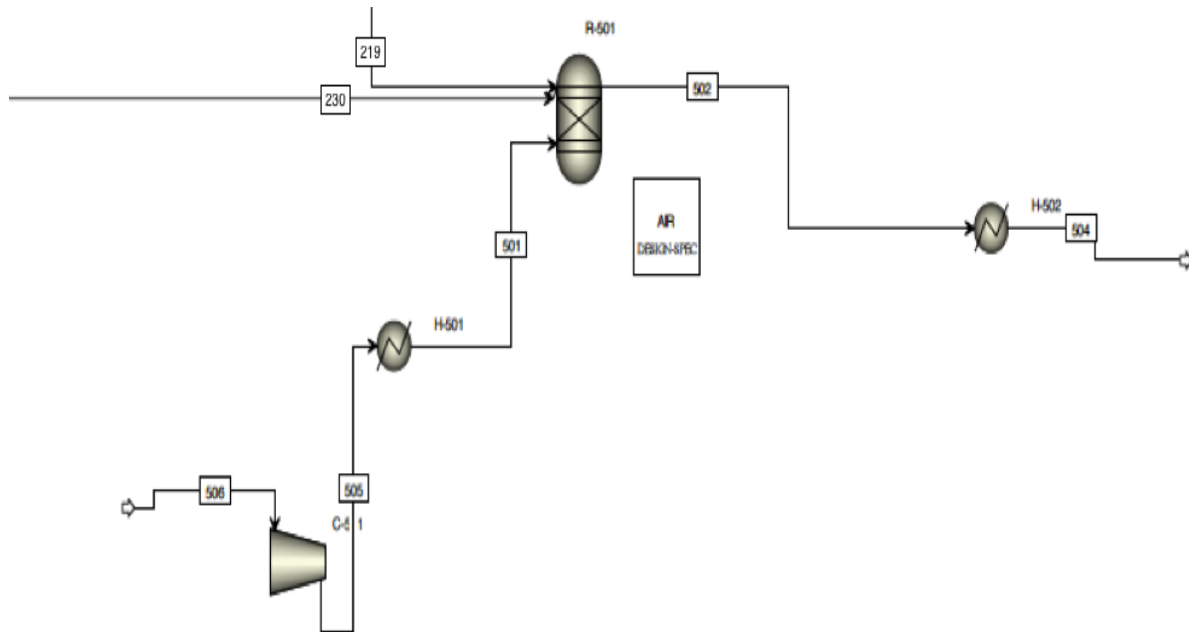


FIGURE 3.14: Snapshot of utility unit.

TABLE 3.11: Composition of combustible and combusted streams (%wt).

	219	230	502
$H_2O$	1.9	0.2	8.8
$H_2$	6.9	4.2	0
$N_2$	12.5	14.1	75.7
$CO_2$	0.2	0.1	9.6
$CO$	1.1	34.3	0
$CH_4$	77.3	38.4	0
$O_2$	0.1	4.7	6.0

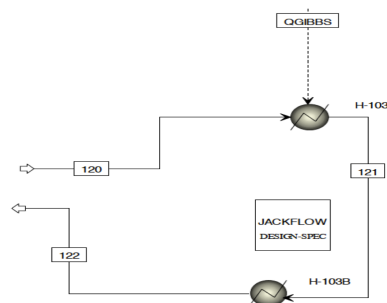


FIGURE 3.15: Snapshot of the modeled Heating Jacket in Aspen plus® for the SCWG reactor.

The overall plant conversion from solid biomass to liquid fuel product is defined on energy basis with biomass to liquid efficiency (BtL) as shown in Equation 3.28 (from Ref. [44]).

$$\eta_{BtL} = \frac{q_{HHV,product}m_{product}}{q_{HHV,biomass}m_{biomass}} \quad (3.28)$$

The global plant thermal efficiency can be evaluated through Equation 3.29 (from Ref. [44]).

$$\eta_{thermal} = \frac{q_{LHV,products}m_{product}}{q_{LHV,biomass}m_{biomass} + P} \quad (3.29)$$

In the equations,  $q$  represents the heating values (MJ/kg),  $m$  the mass flow (kg/s) and  $P$  represents the power consumed by the plant (MW). The efficiency of a fuel conversion to power was not taken into account because the power is purchased offsite (e.g. grid). The vinasse heating value was calculated based on the Boie correlation, as shown in Equation 3.1. The heating values of the products were evaluated through Aspen Plus<sup>®</sup> by simulating complete combustion in a Gibbs reactor. The results obtained were thus compared to the one found in literature (see Table 4.3).

The thermal efficiencies values were evaluated after the design of the heat exchanger network which was realized through the tool Aspen Energy Analyzer.

### Heat exchanger network

The Aspen Energy analyzer tool gives the recommended design of the heat exchanger network in order to meet the load targets. Thus the solution suggested represents an ideal system not always optimal from an engineering point of view. For this study the simulation was at first ran without restrictions. This means that aspen evaluates the recommended design allowing every combination of streams in order to reach the targets. Secondly, the simulation was run with some restrictions on the areas, therefore the stream combination was allowed only among same section (for instance a stream from unit 1 can give load only to streams of the same unit). The results obtained are shown and commented in Chapter 4.

## 3.4 Techno-economic assessment

The techno-economic assessment attempts to establish the economic feasibility and profitability of the plant. The evaluation was realized based on the methodology used is Ref [72]. It determines the minimum selling price of the final product taking into account operational and capital costs considering a plant lifetime of 20 years. The total capital investment (TCI) was evaluated through several assumptions as shown in table 3.13. It consists of the sum of fixed capital investment (FCI) and Other outlays. The FCI includes the direct costs (DC), which were evaluated as the purchased equipment cost (PEC) of the total units of the plant multiplied by a installation factor, and the indirect costs which depends on the total direct costs. The contingency is thus evaluated as a percentage of the FCI. As illustrated in Table 3.13, Other outlays consist of startup costs, working capital, costs of licensing, research and development which depends on O&M costs, DC, FCI and land cost. The PEC was evaluated in two different ways. For the standard units, for instance vessels, pumps, compressors, the Aspen plus<sup>®</sup>Economic tool was used. This gives an information of the purchase and installation cost in US dollars. The rest of the units, for instance the SCWG process, the RWGS, the FT, were adapted from literature data (Table 3.12). The purchased cost associated to the mixing tank at the beginning of the process (block M-101) was evaluated with the Guthrie method. The costs were brought to the start of construction year (2017) through the chemical engineering plant cost index (CEPCI) with Equation 3.30.

$$PEC_{2017} = PEC_{Ref.year} \frac{CEPCI_{2017}}{CEPCI_{Ref.year}} \quad (3.30)$$

The capacities of the reference were adjusted to the process ones through equation 3.31.

$$PEC_{process} = PEC_{Ref}(Capacity_{process}/Capacity_{Ref})^\alpha \quad (3.31)$$

$\alpha$  represents the escalation factor which varies between 0,6 and 1 depending on the equipment type [44].

TABLE 3.12: Capital cost of non-conventional equipment.

Equipment	Reference	Escalation factor	Year of Ref.	Capital Cost used (k€)
SCWG	[30]	0,6	2007	202
Cyclone	[73]	0,6	2014	0.71
Ash storage	[35]	0,6	2007	0.005
PSA	[74]	0,55	2016	55
LO-CAT	[35]	0,67	2007	153
RWGS	[75]	0,65		269
FT	[74]	0,75	2016	18
FCC	[76]	0,6	1998	16
Boiler	[77]	0,6	2016	186

The capital costs values reported in Table 3.12 are referred to Case 1.

A more detailed description of the TCI evaluation is shown in Table 3.13.

TABLE 3.13: Steps for the TCI evaluation [72].

<b>Fixed Capital Investment</b>	
<b>Onsite costs</b>	
Purchased equipment cost	PEC
Piping	35 %PEC
Instrumentation and controls	12%PEC
Electrical equipment and material	13%PEC
<b>Offsite costs</b>	
Land	6% PEC
Civil, structural and architectural work	21%PEC
Service facilities	35%PEC
<b>Total direct costs (DC)</b>	Onsite + Offsite
<b>Indirect costs</b>	
Engineering and supervision	8%DC
Construction costs and contractors profit	15%DC
Contingency	15% of Direct+indirect costs
<b>Other outlays</b>	
Startup costs	$\frac{Cost\ of\ fixed\ O\&M}{Month} + \frac{Cost\ of\ variable\ O\&M}{Month} + 2\%(FCI-Land)$
Working capital	$\frac{Fuel\ cost}{6} + \frac{Cost\ of\ variable\ O\&M}{Month} + Cost\ of\ fixed\ O\&M * 1.25$
Costs of licensing, research and development	2%DC
<b>Total Capital Investment</b>	TCI

The operational and maintenance costs (O&M) are divided into fixed and variable: the fixed include the employee salaries and benefits, overhead and insurance.

The variable cost depend on the annual operational time which was set at 8000 hours. This value includes the purchase of the material streams: feedstock (vinasse is considered a waste by-product with no associated purchased cost), solvent and process water (a 10% make-up was assumed for both), utilities and cost related to the replacement of the catalysts. The costs related to the O&M costs are detailed in Appendix F. The escalation rate was evaluated based on the changing price of the products between 2000 and 2017 in Finland. As shown in Table 3.14 it was assumed a straight line depreciation, which consists of computing depreciation and amortization by dividing the difference between an asset's cost and its expected salvage value by the number of years it is expected to be used (Equation 3.32). The salvage value of the plant was then assumed equal to zero.

$$StraightLineDepreciation = \frac{PurchasePrice - ApproximateSalvageValue}{EstimatedUsefulLife} \quad (3.32)$$

Table 3.14 shows all the economic assumptions made.

To evaluate the net present value (NPV), the selling prices of each product need to be analyzed which varies through the years as illustrated in the Figure 3.16.

For the kerosene and the LPG the prices selected were taken from year 2017 as average in Europe [79], [80].

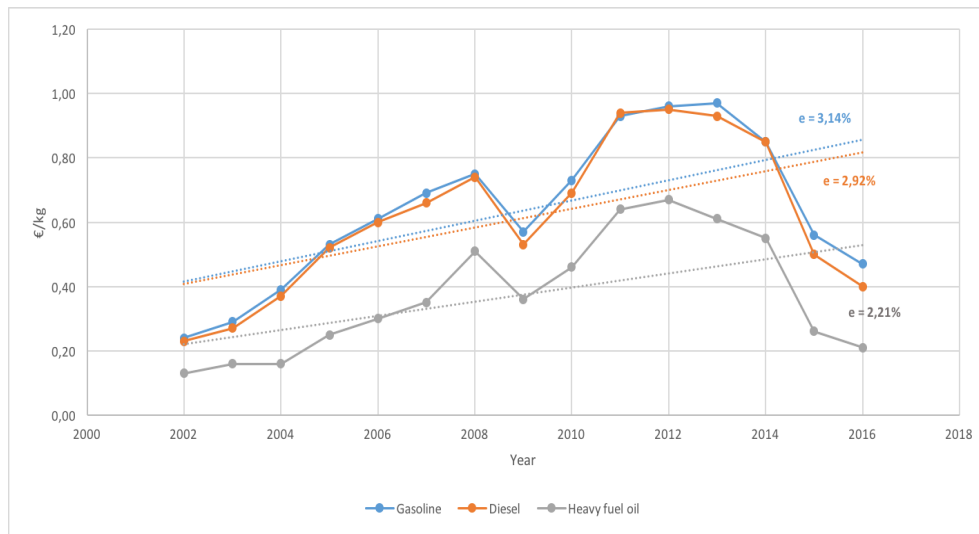


FIGURE 3.16: Prices of the major products over the years in Finland (€/kg) [78].

TABLE 3.14: Techno economic analysis assumptions.

Economic assumptions	
Return on investment	10%
Project life time	20 years
Capacity factor	91%
O&M escalation rate	3% except for electricity (0,34%)
PEC/TCI ratio	0.3
- Case 1	29%
- Case 2	30%
Contingency	15% of Total direct cost, Engineering and supervision and Construction costs
Equity	100%
Average labor cost	36000 euros per year
Number of employees	30
Overhead & insurance	1,5% TCI
Straight line depreciation	
No salvage value	
No fuel cost	
All construction costs are escalated to 1 <sup>st</sup> year operation	

## Chapter 4

# Results and discussion

### 4.1 Carbon recovery

Figure 4.1 shows the carbon distribution in the outputs of the plant. Only ~14% of the carbon is recovered in the product, therefore the carbon losses are high. However, ~42% of the carbon is recovered as combustible gas which covers the heating load required by the whole plant, leading to the absence of an external hot utility to the plant. One of the limiting factor for the carbon recovery in the product is the low  $CO_2$  conversion in the RWGS reactor. In fact, as shown in Figure 3.9, the catalyst selected allows a conversion of 40%.

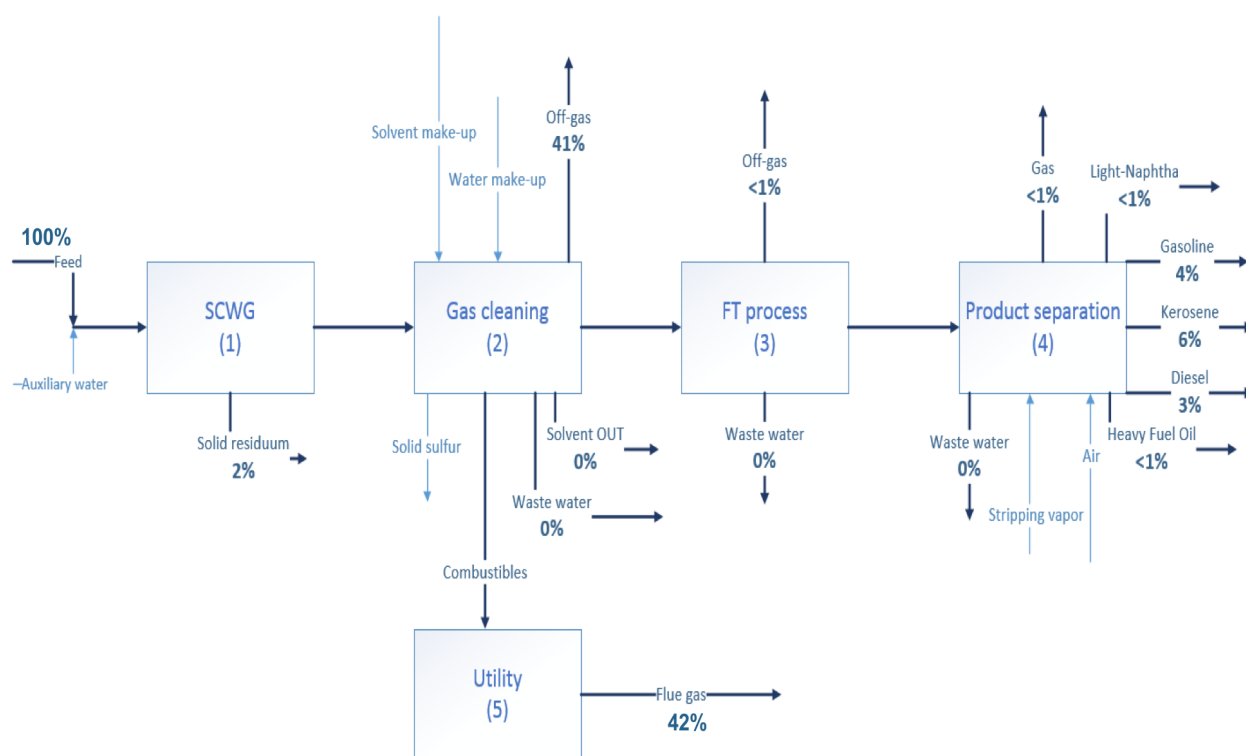


FIGURE 4.1: Carbon partitioning in the plant.

All of the combustible gas is used as hot utility. This stream is mainly composed of methane (~45%mol) and hydrogen (~35%mol) if we consider both the stream exiting the PSA system (stream 219) and the scrubbing system (stream 230). This was modeled so that the excess heat load can generate high pressure steam, and thus partially cover the duty (kW) required by the equipment of the

plant (pumps and compressors). As an alternative, the excess methane could be split before entering the boiler and then it could be evaluated as product of the plant (SNG) or it could be steam reformed in order to produce  $CO$ , one of the main reactants of the FT process. However this solution would lead to a higher cost of the electricity utility which in this case is partially covered by the HP steam generated. With the design selected for this study, the plant is self sufficient. This means that the heat recovered from the combustible gas is sufficient to cover the heat demand required by the plant and no external utility is needed.

The off-gas is ~41% of the carbon which exits the plant as  $CO_2$  mainly (~70%mol) with a minor content of water (~18%mol) and  $H_2S$  (~4%mol). The  $CO_2$  recovery is limited by the presence of  $H_2S$ : their solubilities are different but similar, therefore the more  $CO_2$  we manage to recover, the more  $H_2S$  ends in the stream going to the RWGS step.

The 2% of carbon going in the solid residuum stream, shown in the diagram (4.1), is the non-reacted carbon from the feedstock. In fact it was made as assumption that 98% of the carbon entering the SCWG step reacts to form the syngas, and the rest is assumed to leave the plant as solid residuum with the ashes. Sensitivity analysis conducted throughout the design of gas cleaning process units enabled the near elimination of carbon losses in the waste water or solvent recovery streams. For this reason they both were thought to be recycled with the evaluation of a periodical replacement of the make-up due to the impurities content. The carbon contained in the FT crude is then split into the different fractions where, as illustrated in the diagram 4.1, the kerosene is the main product. To have a more efficient description of the carbon partitioning within the products is necessary to analyze the product characteristics.

## 4.2 Product characteristics

The objective of the thesis is to produce transportation liquid fuels, therefore the product distribution should be evaluated. As shown in Figure 4.2 the major product obtained from the plant is the kerosene, which is around 40% of the FT crude. Furthermore, it is possible to notice that still a valuable percentage of the crude is fractionated as diesel (~22%) and gasoline (~28%), which are also the most valuables in terms of consumption. This characteristic is mainly caused by the catalyst of the FT process and the  $\alpha$  parameter which affects the hydrocarbons distribution.

As explained in the Chapter 3 chapter the way to evaluate the characteristic of a distillate is through the TBP curves.

The Graph 4.3 shows the temperature boiling point curves for each product. It is possible to notice that heavier fractions, and therefore the heavier hydrocarbons, have a higher TBP curve and thus the temperature necessary to extract a certain distillate percentage is higher.

Consequently, it is possible to compare the TBP of each fraction with the ones found in literature for the petroleum crude [70] shown in Figures 4.4, 4.5, 4.6. The graphs show the percentage of distillate that can be extracted at a specific temperature.

The difference found between the curves of the model and the one of the literature are due to the different crude fed to the fractionation unit. In fact, the values reported in literature are referred to the Petroleum crude which has different characteristics than the FT crude. In each graph (4.4, 4.5, 4.6), the blu line represent the value reported in literature for the petroleum crude. It is possible to notice that the TBP literature curves are higher than the ones evaluated for this study. This is mainly caused by to the higher presence of impurities in the streams, for instance sulfur and oxygen, which affect the boiling point. For example, in figure 4.5 is shown that a 90% distillate of petroleum crude can be extracted at ~270°C, while for the kerosene obtained from this study the value is lower (~240°C).



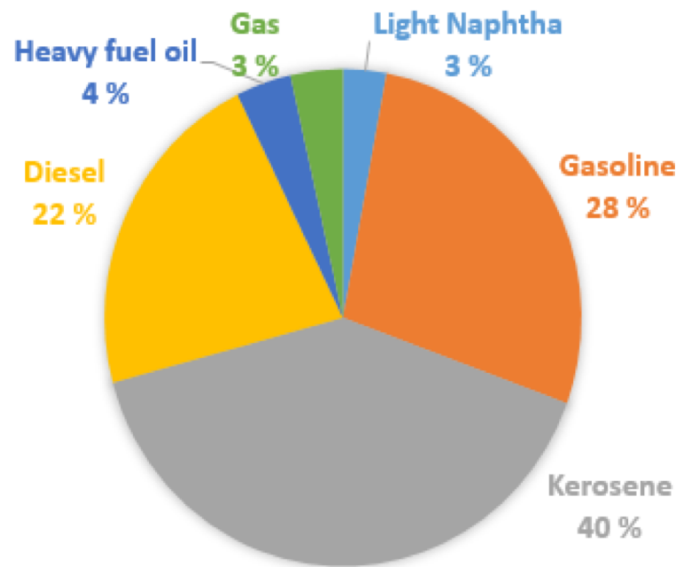


FIGURE 4.2: Product distribution.

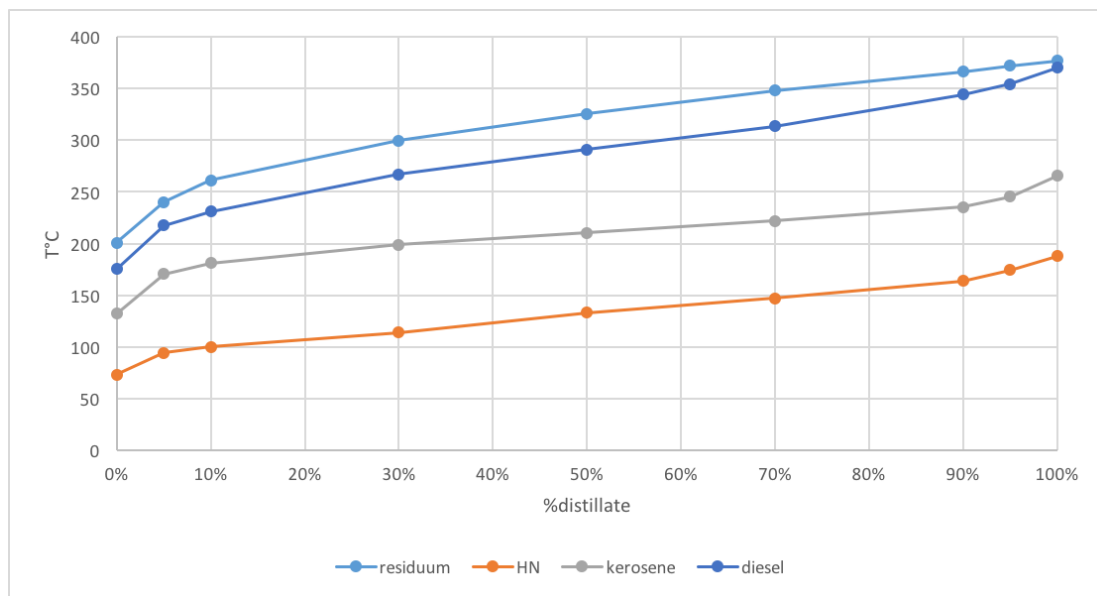


FIGURE 4.3: TBP curves of the fractions.

Therefore for the Petroleum fractions a higher load is required to obtain the same percentage of distillate.

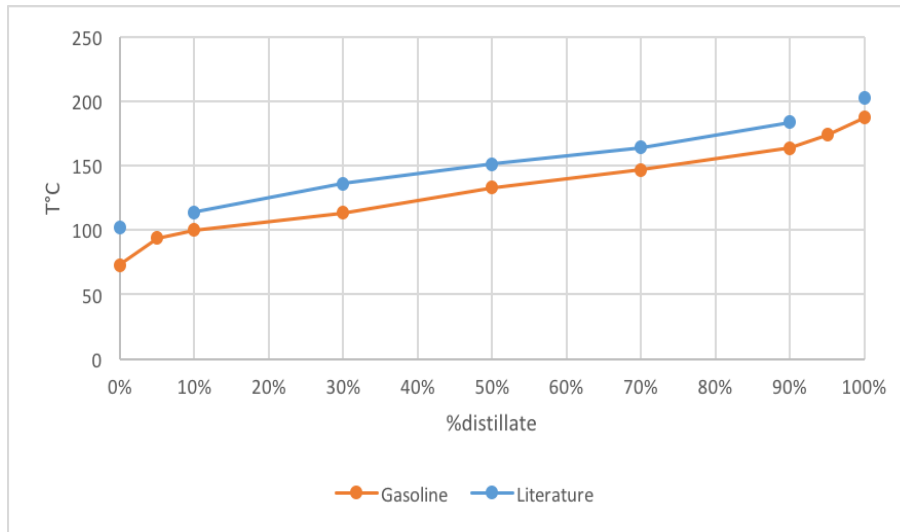


FIGURE 4.4: TBP of gasoline from simulation VS from literature [70].

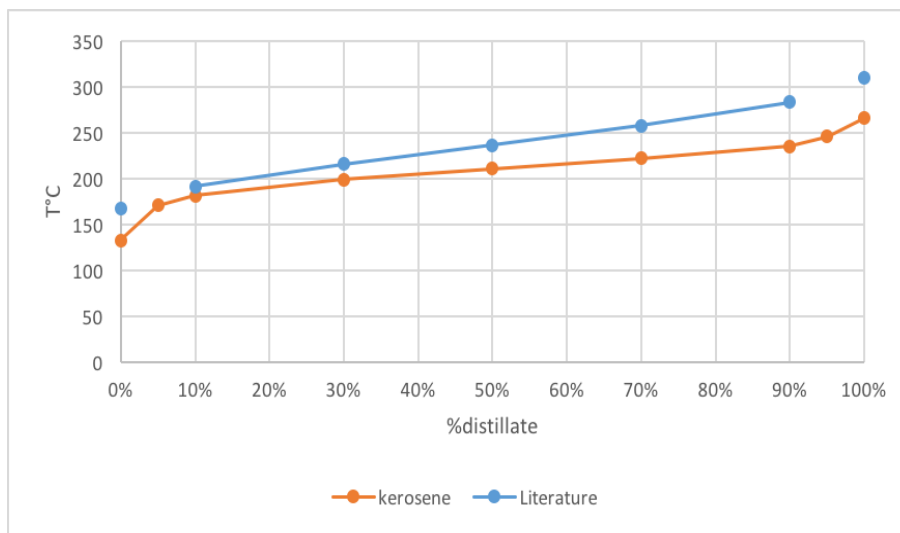


FIGURE 4.5: TBP of kerosene from simulation VS from literature [70].

## 4.3 Energy analysis

### 4.3.1 Thermal performance assessment

The grand composite curve (Figure 4.7) from the pinch analysis illustrates the maximum heat recovery and minimum utility demand (Table 4.1). In the figure the upper curve represents the flue gas from the utility unit. The other curves (highlighted in the box in Figure 4.7) represent multiple streams. It is relevant to notice that the main heat load, besides the flue gas, is given from the syngas, and the major demand comes from the feedstock pre heating. The heat exchanger network was designed with Aspen Energy Analyzer and, as a result, two different scenarios were selected. With this configurations, it is possible to cover all the heat load required, therefore no external hot utility is needed. On the other hands, the cooling load is high and thus a large amount of cold utility is required. What emerged

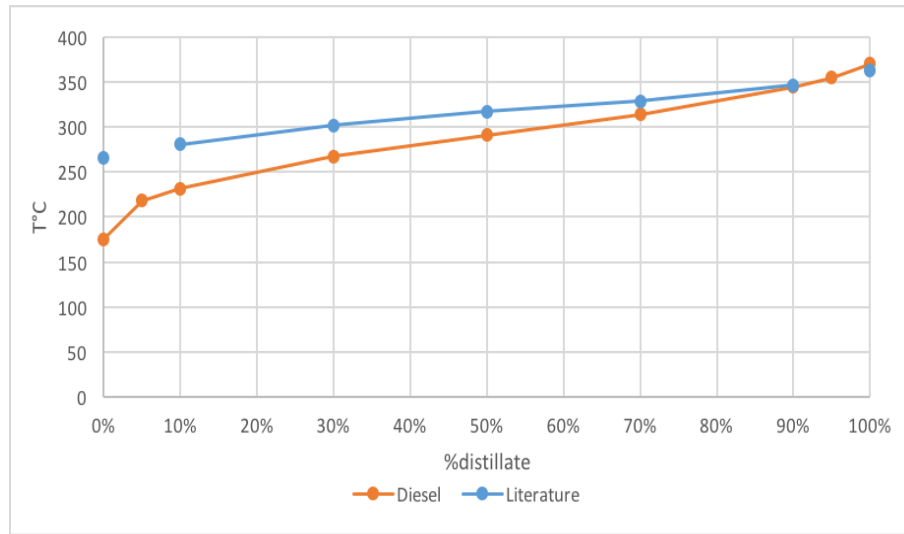


FIGURE 4.6: TBP of diesel from simulation VS from literature [70].

from the Aspen Energy Analyzer tool is that cooling water is sufficient, and there is no need for a refrigerant. At the same time it is possible to produce a certain amount of high pressure (HP) steam due to the high temperature of the syngas.

TABLE 4.1: Target loads of the plant.

Heating load	MW	0
Cooling load	MW	7.82

As illustrated in the Chapter 3, the HHV of the feedstock on dry basis was evaluated through the Boie correlation. From that the HHV on wet, on dry-ash free basis were calculated and the LHV of the biomass were calculated. To evaluate the heating value of the product it was used on Aspen Plus a RGIBBS reactor working at 25°C and 1 bar to simulate complete combustion. The heating values calculated are shown in Table 4.2. The heating values shown for the stream named *Product* represent an average between the heating values of the different products.

TABLE 4.2: Heating values of biomass and product.

Biomass		HHV	LHV
Wet basis	MJ/kg	7.45	5.68
Dry basis	MJ/kg	14.66	13.53
Dry ash-free basis	MJ/kg	20.99	19.38
Product		HHV	LHV
	MJ/kg	44.05	41.02

Table 4.3 shows the comparison between the heating values of the different products evaluated with Aspen plus® and the values reported in the literature [78]. In the table only the main products are present; the residuum and the light naphtha were neglected because their low amount in the product (as shown in Figure 4.2).

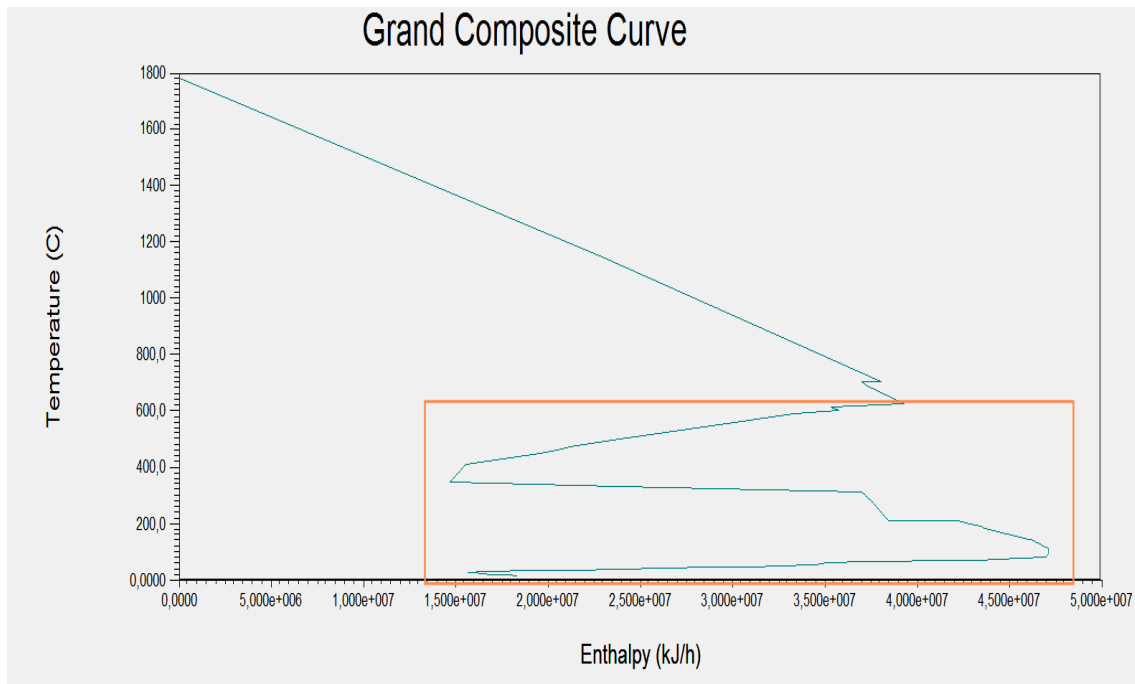


FIGURE 4.7: Grand composite curve.

TABLE 4.3: Lower heating values from Aspen Plus<sup>®</sup> vs values from literature [78] for the main products.

Product	$LHV_{Aspen}$	$LHV_{Literature}$
	MJ/kg	MJ/kg
Gasoline	43.2	41.9
Kerosene	42.6	43.3
Diesel	42.7	42.9

The thermal efficiencies of the plant were evaluated after the design of the heat exchanger networks.

### Heat exchanger networks

The heat exchanger networks design are reported in Appendix E. In the first case, illustrated in Figure E.1, the configuration requires the use of 33 units among which nine use external utilities.

The first solution has limitation from a practical point of view. This because the energy integration is realized between streams of different sections of the plant. Therefore a second simulation was run with some restrictions on the areas, therefore the stream combination was allowed only among same section (for instance a stream from unit 1 can give load only to streams of the same unit). As a result, in this case 34 units are required leading to a higher capital cost of the network. Figure E.2 shows the heat exchanger network of the second scenario.

The process streams that require the major amount of load are the hot syngas (109 to 110) and the feedstock stream that need to be pre heated (104 to 105). It is possible to notice (Figure E.1) that in the first configuration the syngas should go through ten exchangers to meet the requirements, and

TABLE 4.4: Case1 vs Case 2.

	Case 1	Case 2
Capital cost (M€)	51	77
Units	33	34
Number of syngas exchangers	10	6
Number of exchangers for feedstock preheating	8	4
HP steam production	Higher	Lower

stream 104 to 105 through eight exchangers. This makes the first configuration less feasible from an industrial point of view respect to the second case. In fact, in Figure E.2 it is shown that the syngas should cross six exchangers and stream 104 to 105 four. Hence, the amount of units needed from these streams is lower than in the first case. This represents a more realizable configuration on a practical point of view, even though it should be optimized to reduce the amount of heat exchangers for each stream. On the other hand, the second case leads to a lower production of high pressure steam, therefore the electricity required from external units is higher. This leads to a higher operational cost compared to case 1 (see section *Techno economic assessment results*).

At this point, the efficiency values were evaluated. Results are shown in Table 4.5. They refer two different cases based on the two heat exchanger networks evaluated for the study (see section *Heat exchanger network*).

TABLE 4.5: Thermal efficiencies (on biomass wet basis) results.

$\eta_{BtL}$	17%
$\eta_{Thermal-Case1}$	20.7%
$\eta_{Thermal-Case2}$	20.7%

The low efficiency values are mainly due to the low recovery of the carbon into the product (~14%). If, for instance, we compare this result to a Hydrothermal liquefaction of lignocellulose residues we can see that the thermal results found from Magdeldin et al. [81] are higher: the thermal efficiency of the plant is around 30%. The efficiency obtained are significantly lower than the one reported in Table 2.5, although it is important to notice that those efficiencies were associated with the SCWG of biomass. In this study, the product obtained from the SCWG process block has been further treated for liquid fuel production. Therefore it necessary to compare the results in Table 4.5 also with biomass to liquid production processes. Trippe et al. [47] studied the biomass to liquid conversion via dimethyl ether (DME) and via FT synthesis from a syngas obtained through conventional gasification. The energy efficiency associated to this process ranged around 38%. The value is higher than the ones in Table 4.5; one reason could be because of the working conditions of the supercritical water gasification system who require a great energy load. Moreover, in this study a RWGS reactor was designed before the FT process block differently from the work of Trippe et al. [47]; this step requires a significant energy load due to the high temperature working condition. Zhu et al. [82] made a techno-economic analysis for the thermochemical conversion of biomass to liquid fuels. They evaluated several pathways for fuel production from biomass and made a comparison among the results obtained (Table 4.6).

TABLE 4.6: Thermal efficiencies reported in ref. [82] for different biomass to liquid pathways.

Ethanol via Mixed Alcohols	41-45%
Methanol to Gasoline	44-46%
FT diesel	40-46%
Ethanol via acetic acid	55-58%
HTL	63%
Fast pyrolysis	64%

What emerges from the work of Zhu et al. [82] is that the processes with higher overall conversion, for instance methanol and DME, have better thermal efficiency than lower conversion synthesis cases, such as FT diesel. However, all the cases in Table 4.6 show a higher thermal efficiency than the one evaluated for this study. Many factor could be the cause of this result, first of all the low carbon recovery into the product. Secondly could be the complexity of the plant and the high energy demand due to the high pressure and high temperature working conditions for more than just one block.

#### 4.4 Techno economic assessment results

The annual costs associated to each case are reported in Table 4.7. From this table it is possible to notice that the cost related to the second case, the restricted one, is higher than the other. This is due to a higher number of exchangers and also to the lower amount of HP steam generated and therefore the higher power demand of the plant.

TABLE 4.7: Costs associated to the two cases.

<b>Case 1</b>			
TCI	M€		51.5
Fixed O&M	M€/year		3.5
Variable O&M	M€/year		4.7
<b>Case 2</b>			
TCI	M€		77
Fixed O&M	M€/year		3.5
Variable O&M	M€/year		5.1

The external power required is indeed lower in case 1 than in case 2 due to the HP steam generation. Average results are shown in Table 4.8.

TABLE 4.8: Power demand of the plant.

Case 1		
Power consumed	MW	2,03
HP steam	MW	-1,88
Power required	MW	0,14
Case 2		
Power consumed	MW	2,03
HP steam	MW	-1,26
Power required	MW	0,76

With the value of the amount of fuel produced every year, it was possible to evaluate its minimum selling price to reach the break even point. Results are shown in Table 4.9.

TABLE 4.9: Minimum Selling Price (MSP).

Case 1		
Total Price per kg of product	€/kg	4.28
Total Price per bbl of product	€/bbl	584.4
Total Price per MWh of product	€/MWh	360.2
Case 2		
Total Price per kg of product	€/kg	5.72
Total Price per bbl of product	€/bbl	780.72
Total Price per MWh of product	€/MWh	481.25

The cost contribution (%) to total product cost is shown in Figure 4.8 and 4.9. The first (4.8) allows to understand the contribution of each unit to the product cost. The second one (4.9) illustrates how much the capital cost and the operative costs effect the minimum selling price of the product.

The Graph 4.8 shows that the major cost contribution to the purchased equipment cost is given by the heat exchanger network (~42% for case 1 and ~62% for case 2). Secondly to that, the principal cost related to the purchased equipment is due to the SCWG unit. Therefore, although hydrothermal treatment offers efficient heat recovery compared to losses in evaporative process, the cost associated to the heat recovery is significantly high. Figure 4.9 highlights that in both cases the major contribution to the production cost is the capital carrying charges. This is due to the high cost of the units and the especially to the heat exchanger network, which affects the capital cost due to the difference in number of units. It is possible to compare the MSP obtained with the values found in literature for other processes of biomass conversion into liquid fuels (Table 4.10).

The values reported in Table 4.10 refer respectively to a hydrothermal liquefaction of lignocellulose residues and a Fischer Tropsch for liquid fuels production rescaled to year 2017. It is clear that the production cost related to the cases of this study are much higher than the others. This is due

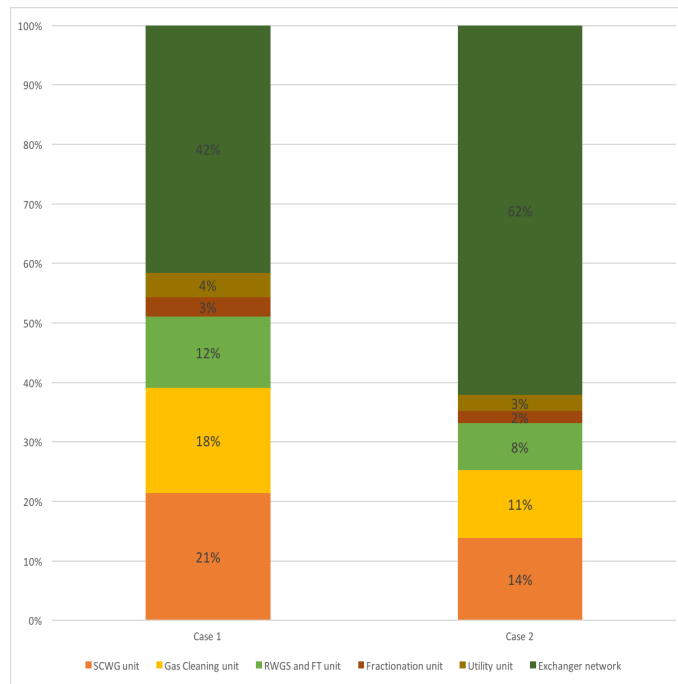


FIGURE 4.8: Cost contribution of each unit (%) to total product cost.

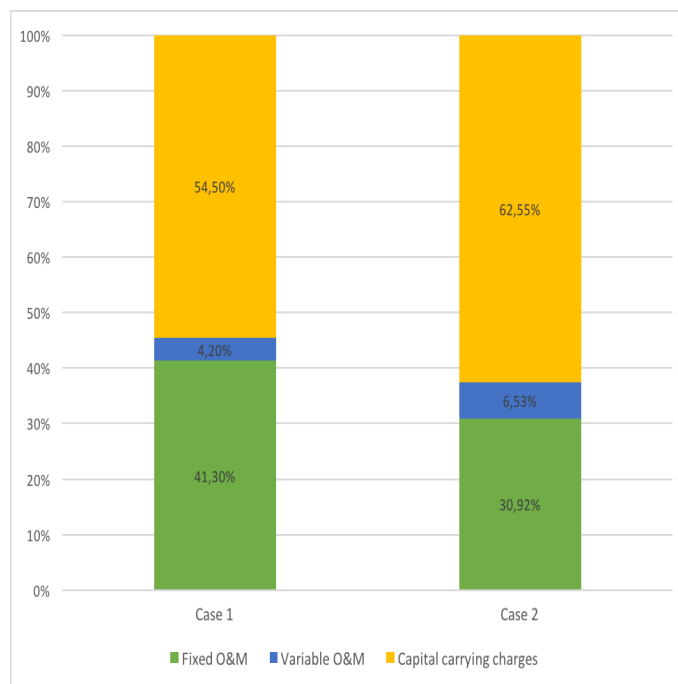


FIGURE 4.9: Cost contribution of capital cost and the operative costs (%) to MSP.

mostly to the low recovery of carbon into liquid fuel. One other reason could be the high cost associated to the heat exchanger network. The solution found from aspen energy analyzer tool represents the recommended design and thus the ideal network. Only through an optimization the process could



TABLE 4.10: MSP comparison.

Process	€/MWh	Ref.
Case 1	360	-
Case 2	481	-
HTL	191	[44]
FT fuels	74	[45]

lead to the production of a competitive product on the market.

The net present value of the plant was evaluated knowing the prices of each product in 2016 and their escalation rates from 2000. For both cases the incomes related to the products are lower than the total expenses. A sensitivity analysis was made to understand the amount of income necessary to breakeven. As a result, the incomes related to the process should be 7 times higher than the actual situation for Case 1 and around 9 times for Case 2. Table 4.11 shows a comparison between the current incomes related to the process and the incomes needed to breakeven.

TABLE 4.11: Current incomes related to the process vs incomes needed to breakeven.

	Current incomes	Incomes needed to breakeven
	€/year	€/year
Case 1	1.332.351	9.425.443
Case 2	1.332.351	12.591.915

## 4.5 Conclusions

The thesis proposes a design of an innovative process for the production of liquid fuel from vinasse, with the aim of finding an alternative and sustainable solution to current problems. The project wants to analyze the possibility of using the vinasse as feedstock for the production of liquid fuels and to understand the optimal path to it. The plant was designed and simulated on Aspen Plus®, based on the amount of vinasse produced from an ethanol plant from sugar cane. As a matter of fact, vinasse is a by-product obtained from the processing of molasses through fermentation and subsequent distillation.

As the plant is designed, what emerges is that vinasse has a good potential for the production of hydrogen through the use of supercritical water. However, the production of liquid fuel through the Fischer Tropsch process is limited by the low carbon recovery into fuel (only ~14%). As analyzed, this limit is mainly caused by the presence of a high sulfur content which affects the solubility of  $CO_2$  in water. One possible solution could be the pretreatment of the biomass before entering the SCWG reactor to remove the sulfur content and thus to reduce to the minimum the synthesis of  $H_2S$ . This solution, however, could effect the thermal analysis. The process modeled is thermally self sufficient but the use of external utility to cover the power demand is needed. It was possible to reach this self-sufficiency condition thanks to the presence of a high amount of combustibles which are recovered from the acid gas removal step. Hence, a different method for the sulfur removal could lead to a lower heat recovery and consequently to the need of an external hot utility. At last, the

production costs related to this process (in both case 1 and 2) are higher than the costs associated to other processes for conversion of biomass into liquid fuels.

The goal of the future studies is certainly the reduction of the carbon losses and the improvements of the current technologies. The conversion of waste biomass for biofuels production by applying supercritical fluids will play a significant role in industry development because of its sustainability and its environmental-friendly characteristics.

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## Appendix A

# Agitated tank

For the solid-liquid mixing system some simplified calculation can be carried out. At first the type of impeller has to be chosen for example the LIGHTNIN A310 (dimensions selected by catalog [83]). Based on that and on the physical properties of the feedstock, the diameter and the rotation speed can be evaluated through equations given from the literature [84].

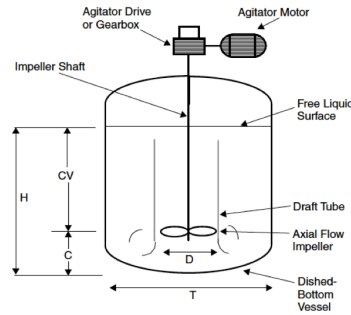


FIGURE A.1: Mixer design [84]

Therefore it is possible to evaluate the dimensions of the tank with equations A.1, A.2 and A.3.

$$T = 2,4 * D \quad (\text{A.1})$$

$$C = \frac{T}{4} \quad (\text{A.2})$$

$$H = T * 1,3 \quad (\text{A.3})$$

The rps required for a solid-liquid mixing can be evaluated through equation A.4 [84].

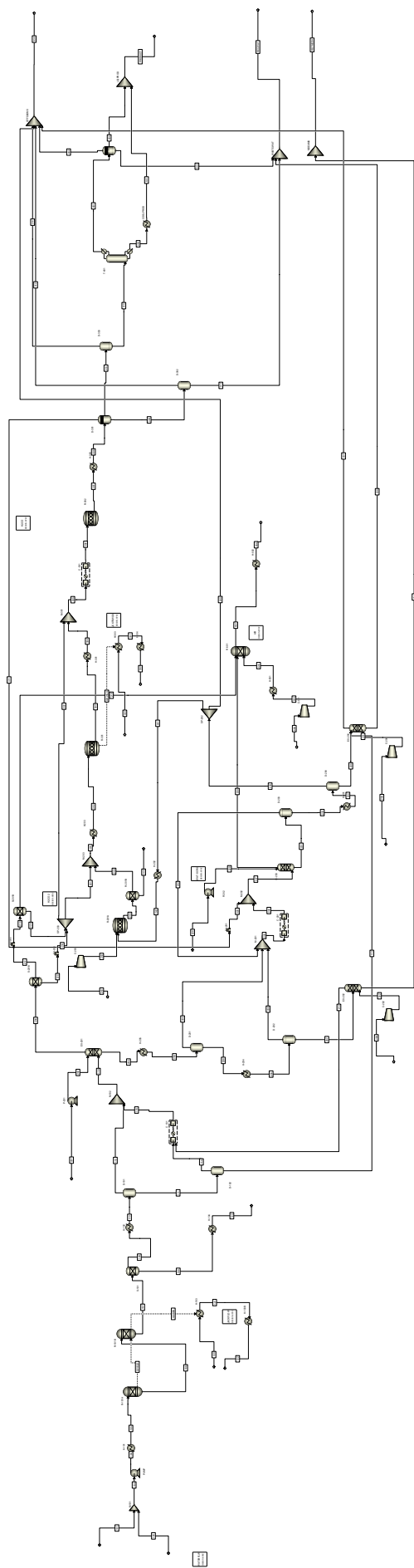
$$N_{js} = S v^{0,1} \left[ \frac{g_c (\rho_s - \rho_l)}{\rho_l} \right]^{0,45} X^{0,13} d_p^{0,2} D^{-0,85} \quad (\text{A.4})$$

In equation A.4,  $D$  is the impeller diameter,  $d_p$  is the mass-mean particle diameter,  $X$  is the mass ratio of suspended solids to liquid  $\times 100$  (kg solid/kg liquid) and  $S$  the dimensionless number which is a function of impeller type (6,9 for the impeller selected).  $v$  is the kinematic viscosity of the liquid ( $m^2/s$ );  $g_c$  is the gravitational acceleration constant;  $\rho_s$  and  $\rho_l$  the density of the solid and the density of liquid ( $kg/m^3$ ) [84].



## **Appendix B**

# **Plant Flowsheet and equipment table**



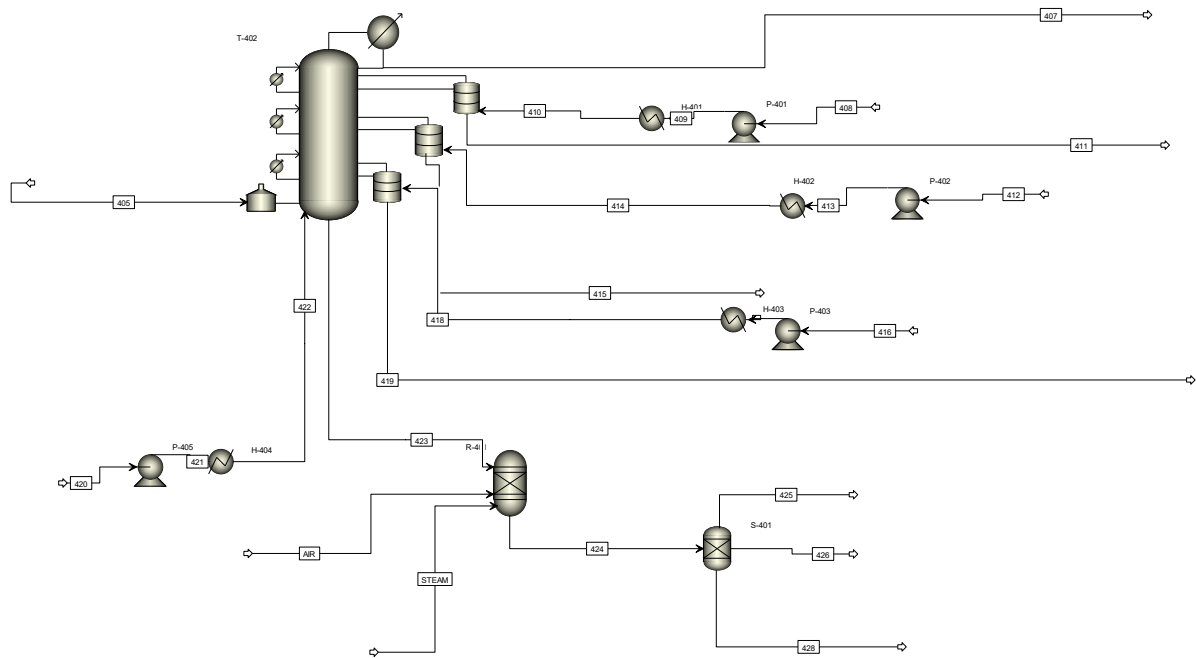


TABLE B.1: Plant equipments (Unit 1-2)

<b>Unit 1</b>	
M-101	Feedstock mixer
PUMP	Pump
H-101	Feedstock preheater
R-101A/B	SCWG reactor
H-103A/B	SCWG reactor heating jacket
S-101	Solid separator
H-104	Solid cooler
H-102	Syngas cooler
D-101	Flash for water removal
D-102	Flash for water purification
M-102	Mixer
<b>Unit 2</b>	
P-201	Pump
P-202	Pump
EX-201	MDEA adsorber
D-201	Flash
D-202	Flash
D-203	Flash
D-204	Flash
H-201	Heater
H-202	Cooler
H-203	Heater
H-204	Heater
H-205	Cooler
EX-202	Gas desorber
EX-203	Gas desorber
EX-204	Gas desorber
C-201	Compressor
C-202	Compressor
C-203	Compressor
C-204	Compressor
M-201	Mixer
M-202	Mixer
M-203	Mixer
TV-201	Trotting valve
TV-202	Trotting valve
TV-203	Trotting valve
SP-201	Split
SP-202	Split
SP-203	Split
R-201A/B	LO-CAT reactor
S-101A/B	PSA system

TABLE B.2: Plant equipments (Unit 3-5)

<b>Unit 3</b>		
R-301		RWGS reactor
R-302		FT reactor
H-301		Cooler
H-302		Cooler
H-303/H-304		RWGS reactor heating jacket
C-301		Compressor
M-301		Mixer
D-301		Flash for FT-crude separation
D-302		Flash
D-303		Flash
<b>Unit 4</b>		
T-401	Distillation column for Light naphta separation	
T-402	Fractionation column	
D-304		Flash
P-401		Pump
P-402		Pump
P-403		Pump
P-404		Pump
P-405		Pump
H-401		Heater
H-402		Heater
H-403		Heater
H-404		Heater
R-401		FCC reactor
S-401		Product separator
<b>Unit 5</b>		
C-501		Compressor
H-501		Heater
H-502		Cooler
R-501		Boiler





## Appendix C

# Amine scrubbing stages evaluation

This process block was at first modeled on excel in order to find the ideal number of stages for the acid gas removal and subsequently implemented o Aspen Plus.

The first assumption made is:

- $H_2S$  removal: 99%
- $CO_2$  removal: 90%
- Removal of 0.4 mol of Acid gas per mol of amine
- Solvent at 30% of MDEA

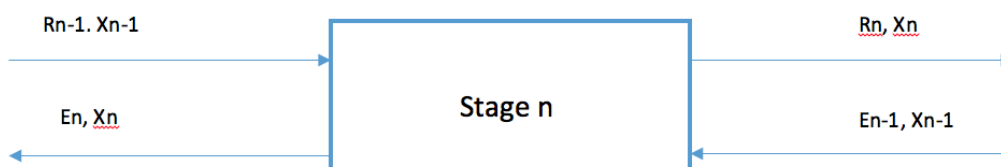


FIGURE C.1: Schematic representation of stage n of amine scrubbing

The calculation was iterative by assuming a certain amount of solvent and verifying the content of acid gas after each stage.

Stage 1			R1	R2	E1	En	w/w
SOLVENT							
CO2	kg/h	0	3841,679191		88231,7558	84390,0766	0,96
H2S	kg/h	3963,234	162,3760373		-233,9474	3566,9106	0,0405
Flow rate	kg/h	42,74777	1,926541517		1,4991	42,3203	0,0005
			4005,98177	4005,98177	87999,30754	87999,30754	
Stage 2					Extraction NOT verified		
			R2	R3	E2	E1	w/w
SOLVENT							
CO2	kg/h		3841,679191	4016,563486	88406,6401	88231,7558	1,00
H2S	kg/h		162,3760373	-10,6499573	-406,9734	-233,9474	-0,0027
Flow rate	kg/h		1,926541517	0,0682417	-0,3592	1,4991	1,70E-05
			4005,98177	4005,98177	87999,30754	87999,30754	
Stage 3					Extraction NOT verified		
			R3	R4	E3	E2	w/w
SOLVENT							
CO2	kg/h	0	4016,563486	4024,524722	88414,6014	88406,64013	1,00
H2S	kg/h		-10,64995733	-18,5265986	-414,8500	-406,9733573	-0,0046
Flow rate	kg/h		0,0682417	-0,01635346	-0,4438	-0,3592	-4,08E-06
			4005,98177	4005,98177	87999,30754	87999,30754	
Stage 4					Extraction verified		
			R4	R5	E4	E3	w/w
SOLVENT							
CO2	kg/h		4024,524722	4024,88714	88414,9638	88414,60137	1,00
H2S	kg/h		-18,52659863	-18,885166	-415,2086	-414,8499986	-0,0047
Flow rate	kg/h		-0,016353457	-0,02020447	-0,4477	-0,4438	-5,04E-06
			4005,98177	4005,98177	87999,30754	87999,30754	
					Extraction verified		

FIGURE C.2: Schematic representation of stage n of amine scrubbing

## Appendix D

# Calculations for the FT reactor

For the FT reactions usually the reactor works with a temperature range between 200 and 400°C. It was though evaluated the value and it was selected the optimal temperature in order to have a chain growth factor of 0,82 which is the most suitable for a diesel fuel production as illustrated in the table 3.9.

TABLE D.1: alpha values with temperature

Temperature [C]	$\alpha$ [-]
200	0,88
210	0,85
<b>220</b>	<b>0,82</b>
240	0,77
260	0,71
280	0,65
300	0,60
320	0,54
340	0,49
360	0,43
380	0,38
400	0,32

The main assumption made in this calculation are shown in table D.2

TABLE D.2: Main assumptions for FT conversions calculation

$H_2/CO$	-	2
$\alpha$	-	0.82
P	bar	15
T	°C	220
c	-	0.25
CO conversion	-	0.72

The conversions of  $CO$  into paraffins and olefins were evaluated with some equations. It is first possible to determine the molar ( $M_n$ ) and mass ( $W_n$ ) distribution (illustrated in D.1 and D.2) with equations D.1 and D.2.

$$M_n = \alpha^{n-1}(1 - \alpha) \quad (D.1)$$

$$W_n = \alpha^{n-1}(1 - \alpha)^2 n \quad (D.2)$$

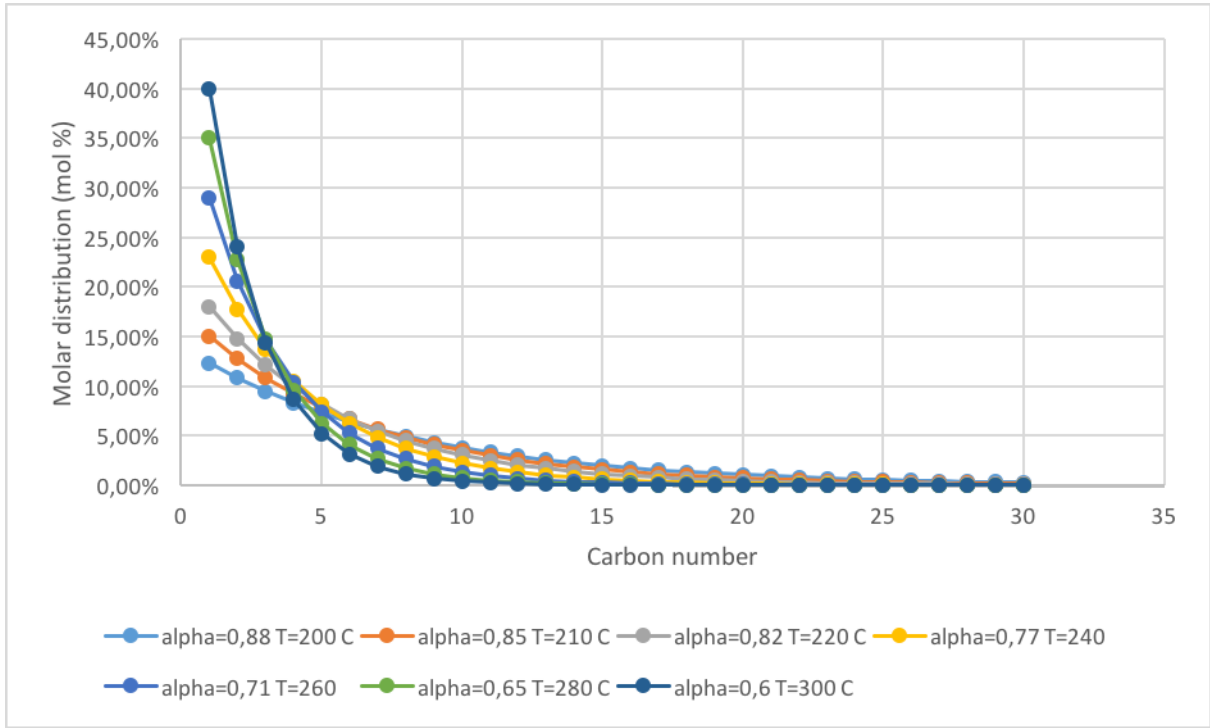


FIGURE D.1: Molar distribution (mol%) of hydrocarbon at different FT working conditions

In the graphs (D.1 and D.2), it is shown that for higher temperatures, and thus lower values of the chain growth factor, short chain hydrocarbons are mainly obtained while at lower temperature the greater amount of long-chained hydrocarbons is produced. It is possible to evaluate the Olefin to Paraffin ratio (O/P) for each carbon numbers. This parameter depend on the carbon number and on a constant value ( $c$ ) which varies between 0,19 and 0,49 [38].

$$\frac{O}{P} = e^{-cn} \quad (D.3)$$

The  $W_n$  value is thus adjusted based on the effective selectivities of the catalyst illustrated in table D.3.

As first step the selectivity of higher hydrocarbons should be evaluated as

$$\sum_{n=5}^{50} W_n \quad (D.4)$$

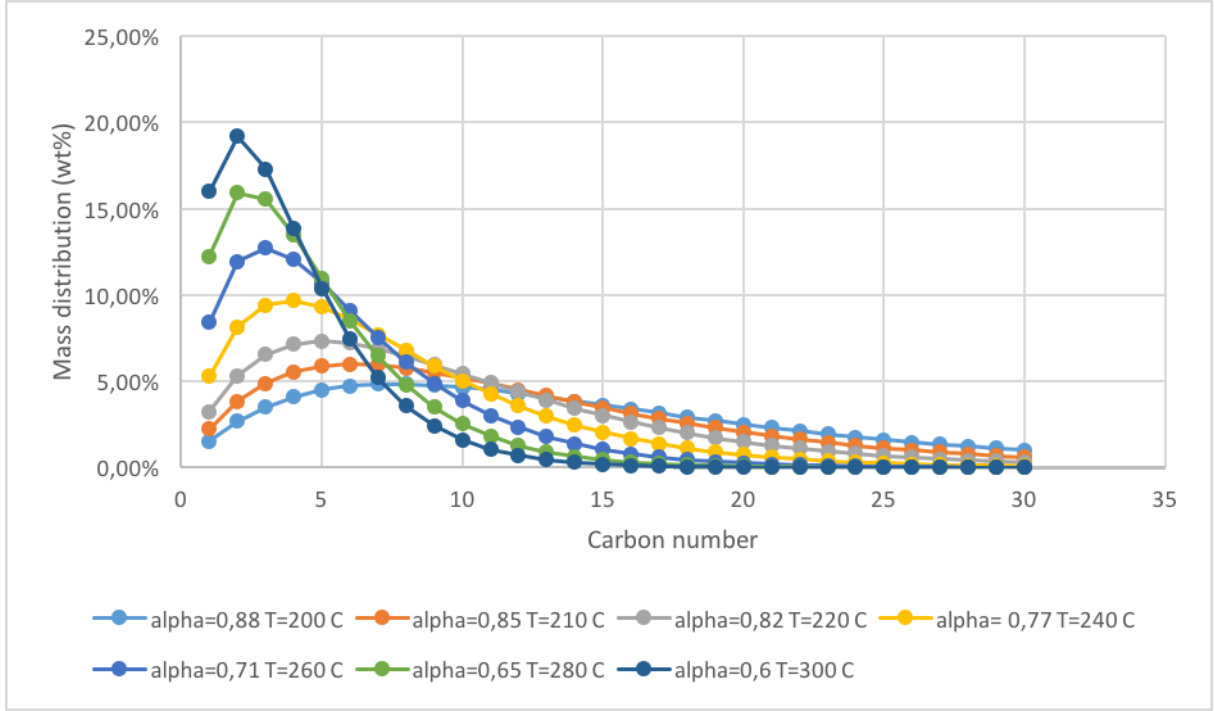


FIGURE D.2: Mass distribution (wt%) of hydrocarbon at different FT working conditions

TABLE D.3: Hydrocarbon selectivities on a Fe-based catalyst at 220°C and 15 bar and 72% conversion of CO [69]

$S_{C_{5+}}$	67,5%
$S_{C_{2-4}}$	25,2%
$S_{C_1}$	7,3%

Consequently the selectivity of  $C_{1-4}$  is calculated as  $S_{C_{1-4}} = 1 - S_{C_{5+}}$ . The weight distribution is thus calculated by rescaling the values.

$$W_n = \frac{S_{C_n}}{\sum_{n=1}^4 S_{C_n}} S_{C_{1-4}} \quad (D.5)$$

The selectivity is based on the carbon content, so the atom carbon content for each hydrocarbon is relevant:  $M_{C_n} = 12n$

Consequently, the  $M_{C_{tot}}$  can be evaluated as

$$M_{C_{tot}} = \frac{1}{\sum_n \frac{W_n}{M_{C_n}}} \quad (D.6)$$

Therefore,

$$M_{ncalc} = \frac{W_n}{M_{C_{tot}}} \quad (D.7)$$

With the  $M_{n_{calc}}$  values, it is possible to calculate the amount of  $CO$  converted for each reaction:

$$CO_{conv,n} = \frac{CO_{in}}{\sum_{n=1}^{50} nM_{n_{calc}}} nM_{n_{calc}} \quad (D.8)$$

At this point of the calculation, it is possible to determine the percentage of  $CO$  converted into olefin and paraffin per each carbon number knowing the  $O/P$  ratio. For this calculation it was chosen a medium value for the constant  $c$  ( $c=0,25$ ).

$$\%P = \frac{1}{1 + \frac{O}{P}} \quad (D.9)$$

$$\%O = 1 - \%P \quad (D.10)$$

Therefore, the flow of  $CO$  converted in paraffin and olefin are respectively the flow of  $CO$  converted times the percentage of olefin and paraffin per each carbon. The conversion value that was inserted in the RSTOIC used for the simulation is therefore the ratio between the amount of  $CO$  converted as explained divided by the flow of  $CO$  entering the reactor. Figure D.3 shows the main calculation for this block.

Carbon number	Mn	Wn	Wn_eff	Mcn	Wn/Mcn	Mn_calc	n*Mn_calc	FlowCOconv, (kmol/h)	O/P	%P	%O	flow Co		CONV P	CONV O
												Paraffin	Olefin		
1	17.89%	3.20%	0.98%	12	0.001	0.012 %	0.000	0.033	0.779	56.2 %	43.8 %	0.019	0.0146	0.0005	0.0004
2	14.69%	5.25%	5.58%	24	0.002	0.070 %	0.001	0.378	0.607	62.2 %	37.8 %	0.235	0.1428	0.0059	0.0036
3	12.06%	6.47%	6.87%	36	0.002	0.086 %	0.003	0.699	0.472	67.9 %	32.1 %	0.474	0.2241	0.0118	0.0056
4	9.90%	7.09%	7.52%	48	0.002	0.095 %	0.004	1.020	0.368	73.1 %	26.9 %	0.746	0.2743	0.0185	0.0068
5	8.13%	7.27%	6.44%	60	0.001	0.081 %	0.004	1.091	0.287	77.7 %	22.3 %	0.848	0.2429	0.0211	0.0060
6	6.68%	7.17%	6.34%	72	0.001	0.080 %	0.005	1.290	0.223	81.8 %	18.2 %	1.054	0.2353	0.0262	0.0059
7	5.48%	6.87%	6.08%	84	0.001	0.076 %	0.005	1.441	0.174	85.2 %	14.8 %	1.228	0.2134	0.0305	0.0053
8	4.50%	6.44%	5.70%	96	0.001	0.072 %	0.006	1.546	0.135	88.1 %	11.9 %	1.362	0.1843	0.0339	0.0046
9	3.70%	5.95%	5.27%	108	0.000	0.066 %	0.006	1.607	0.105	90.5 %	9.5 %	1.453	0.1532	0.0361	0.0038
10	3.04%	5.43%	4.81%	120	0.000	0.060 %	0.006	1.629	0.082	92.4 %	7.6 %	1.505	0.1235	0.0374	0.0031
11	2.49%	4.90%	4.34%	132	0.000	0.055 %	0.006	1.618	0.064	94.0 %	6.0 %	1.521	0.0972	0.0378	0.0024
12	2.05%	4.39%	3.89%	144	0.000	0.049 %	0.006	1.581	0.050	95.3 %	4.7 %	1.506	0.0750	0.0375	0.0019
13	1.68%	3.91%	3.46%	156	0.000	0.044 %	0.006	1.524	0.039	96.3 %	3.7 %	1.467	0.0569	0.0365	0.0014
14	1.38%	3.46%	3.06%	168	0.000	0.038 %	0.005	1.451	0.030	97.1 %	2.9 %	1.409	0.0425	0.0350	0.0011
15	1.13%	3.04%	2.69%	180	0.000	0.034 %	0.005	1.368	0.024	97.7 %	2.3 %	1.336	0.0314	0.0332	0.0008
16	0.93%	2.66%	2.36%	192	0.000	0.030 %	0.005	1.278	0.018	98.2 %	1.8 %	1.255	0.0230	0.0312	0.0006
17	0.76%	2.32%	2.06%	204	0.000	0.026 %	0.004	1.185	0.014	98.6 %	1.4 %	1.168	0.0167	0.0291	0.0004
18	0.63%	2.02%	1.79%	216	0.000	0.022 %	0.004	1.091	0.011	98.9 %	1.1 %	1.079	0.0120	0.0268	0.0003
19	0.52%	1.75%	1.55%	228	0.000	0.019 %	0.004	0.998	0.009	99.1 %	0.9 %	0.989	0.0086	0.0246	0.0002
20	0.42%	1.51%	1.34%	240	0.000	0.017 %	0.003	0.908	0.007	99.3 %	0.7 %	0.902	0.0061	0.0224	0.0002
21	0.35%	1.30%	1.15%	252	0.000	0.015 %	0.003	0.822	0.005	99.5 %	0.5 %	0.818	0.0043	0.0203	0.0001
22	0.29%	1.12%	0.99%	264	0.000	0.012 %	0.003	0.741	0.004	99.6 %	0.4 %	0.738	0.0030	0.0183	0.0001
23	0.23%	0.96%	0.85%	276	0.000	0.011 %	0.002	0.665	0.003	99.7 %	0.3 %	0.663	0.0021	0.0165	0.0001
24	0.19%	0.83%	0.73%	288	0.000	0.009 %	0.002	0.594	0.002	99.8 %	0.2 %	0.593	0.0015	0.0147	0.0000
25	0.16%	0.71%	0.62%	300	0.000	0.008 %	0.002	0.530	0.002	99.8 %	0.2 %	0.529	0.0010	0.0131	0.0000
26	0.13%	0.60%	0.53%	312	0.000	0.007 %	0.002	0.470	0.002	99.8 %	0.2 %	0.470	0.0007	0.0117	0.0000
27	0.11%	0.51%	0.46%	324	0.000	0.006 %	0.002	0.416	0.001	99.9 %	0.1 %	0.416	0.0005	0.0103	0.0000
28	0.09%	0.44%	0.39%	336	0.000	0.005 %	0.001	0.368	0.001	99.9 %	0.1 %	0.367	0.0003	0.0091	0.0000
29	0.07%	0.37%	0.33%	348	0.000	0.004 %	0.001	0.324	0.001	99.9 %	0.1 %	0.324	0.0002	0.0081	0.0000
30	0.06%	0.32%	0.28%	360	0.000	0.004 %	0.001	0.285	0.001	99.9 %	0.1 %	0.284	0.0002	0.0071	0.0000

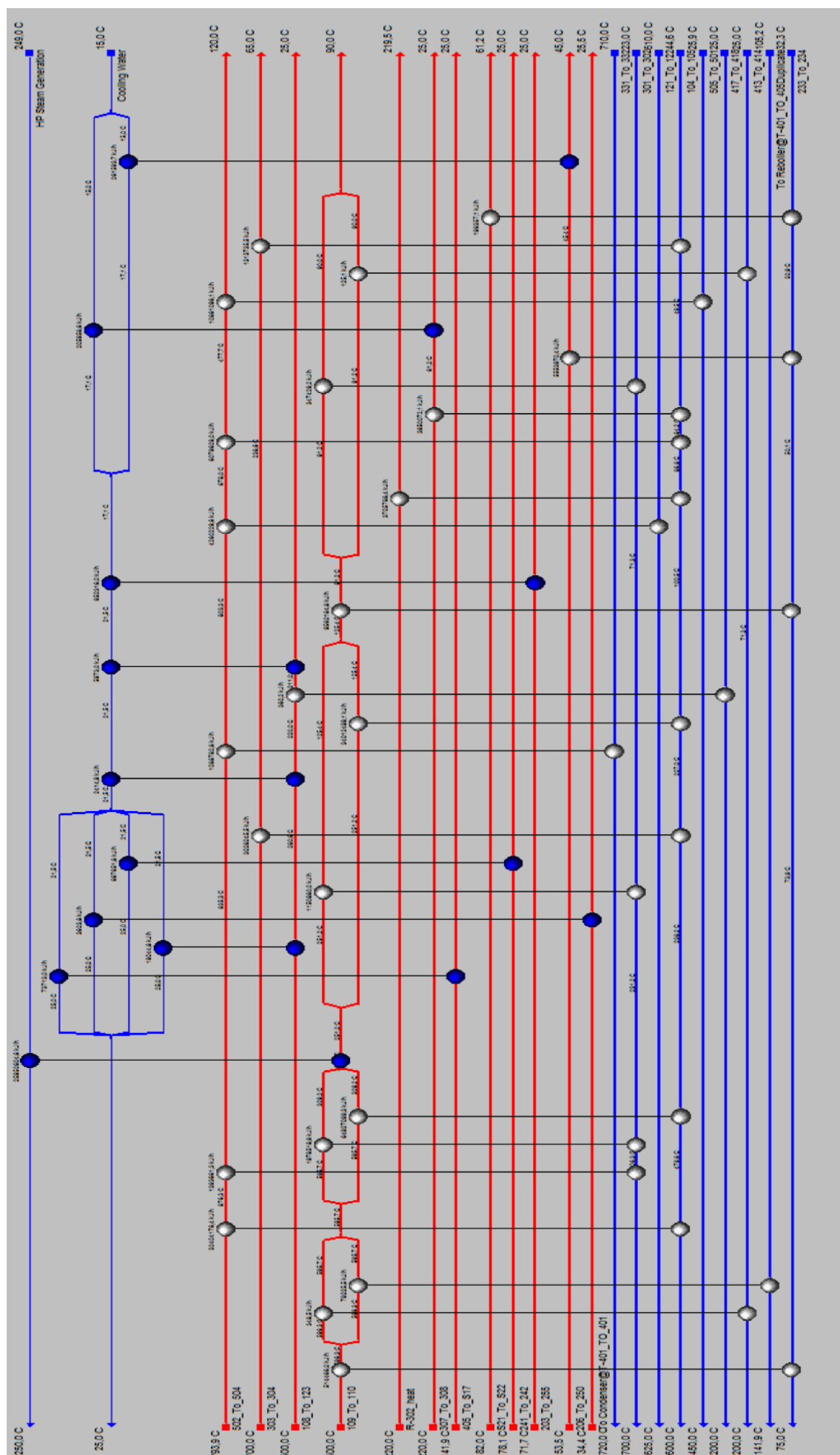
FIGURE D.3: Excel calculations for CO conversion into Hydrocarbons in the FT reactor

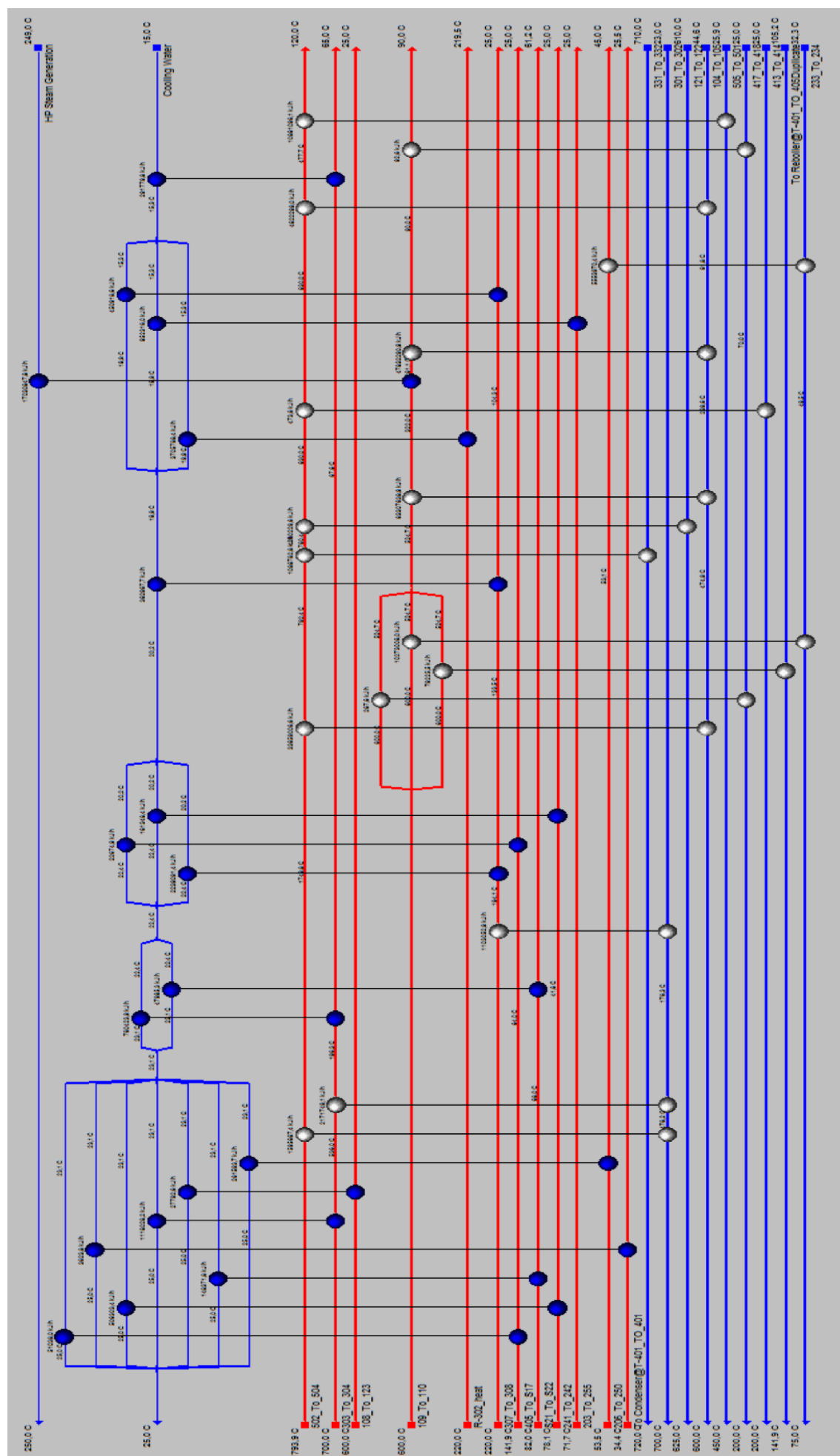




## **Appendix E**

# **Heat exchanger networks**







## Appendix F

# Calculations of Operative and Maintenance costs

TABLE F.1: Operative and Maintenance costs

<b>Annual Fixed O&amp;M</b>		
Working hours	h/year	8000
Capacity factor		91%
Labor position for operating and maintenance		30
Average labor rate	€/year	36000
Overhead expenses	% of Avg Labor	0.5
Personal insurance	% of Avg Labor	0.3
Cost of Plant Maintenance	1,5%TCI	722 k€
Cost of Property Insurance	1,5%TCI	722 k€
Total	k€/year	3.390
<b>Variable O&amp;M</b>		
	k€/year	Reference
MDEA replacement cost	96	[74]
Make-up water	130	[44]
Electric power	99	[44]
FT catalyst replacement	5	[74]
LOCAT catalyst	22	[85]
RWGS catalyst	40	-
PSA packing replacement	0.006	[35]
FCC catalyst replacement	5	[86]
<b>Total Utilities and Chemicals cost</b>	<b>4679</b>	

For the MDEA replacement cost it was assumed 5% of replacing solvent per year due to the recycle. The replacement of the RWGS catalyst was evaluated as 15% of the capital cost associated to the RWGS reactor divided by the working hours per year.



# Ringraziamenti

Vorrei prima di tutto ringraziare il professor Luigi Manna per avermi dato la possibilità di condurre questa tesi all'estero sotto la sua supervisione e per la sua costante disponibilità e cordialità.

I would like to thank my thesis supervisor Prof. Mika Jarvinen of the School of Engineering at Aalto University for giving me the chance of fulfilling this experience. The environment in which I had the chance to work in was friendly and stimulating. Moreover, I would like to thank my instructor D.Sc. Cataldo De Blasio for supporting me and giving me the chance to visit the Åbo Akademi laboratories for better understanding my research field.

A special thanks goes to my advisor M.Sc. Mohamed Magdeldin, who was always available and supportive. He was always willing to give me advices and help whenever I ran into a trouble spot or had a question about my research or writing and he steered me in the right the direction whenever he thought I needed it.

Finally I would like to thank the team of the building K4 for their kindness, availability and the nice moments we shared during coffe breaks.